THE U.S.A.I.D. DESALINATION MANUAL

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A Manual Prepared By

CH2M HILL INTERNATIONAL CORPORATION

CH2M HILL INTERNATIONAL

7201 N.W. 11th Place
Gainesville, Florida
USA 32601

200 S.W. Market Street
Portland, Oregon
USA 97201

NOTICE

This manual was prepared for the U.S. Agency for International Development (U.S.A.I.D.) to guide its officers and to aid foreign administrators and technicians involved in water resources development in formulating and evaluating desalination programs and projects for which United States Government assistance may be furnished or sought. The information contained herein is not sufficient (nor is it intended) for use as the sole basis for design or construction of a project at any particular location, or as a substitute for professional judgment, advice, and assistance as required for development of technical and economic feasibility, plans and designs, construction, and initiation of operation of such a project.

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Many features of the desalination processes described in this manual are covered by patents held by various organizations and/or individuals . No attempt has been made herein to explore the patent status of the various processes. Anyone who contemplates building a desalination plant should fully determine all proprietary rights relating to the particular process or design and should procure all necessary licenses, authorizations, etc. Desalination facility owners should consider obtaining a "hold harmless" clause from equipment suppliers as indemnification against liability or loss from patent infringements.

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PREFACE

It gives me great pleasure to introduce to you this desalination manual as part of IDEA's publication program. Though I might be considered biased by some by virtue of the fact that I have played a small part in its preparation, I feel that this is one of the best, if not the best and most lucidly illustrated volume presenting the subject of desalination on this level.

This is the second IDEA publishing venture and is much in the spirit of one of our association's objectives, namely to help in dissemination of desalination information and help build knowledge in the field and make desalination information available to as wide a circle of readers as possible.

As this book was prepared in the U.S. and by U.S. desalination professionals, it naturally takes a U.S. point of view and this includes predominant though not exclusive reference to U.S. desalination industry products. Speaking for IDEA and as its president of the international organization that we are, I must add that in addition to the manufacturers referred to in this text, many of our members around the world are equally competent to serve the field. We urge those interested in these manufacturers to write to us at Headquarters for a list of our members who serve as suppliers in this field. A list of these will be on the way to you by return mail.

In closing I wish to compliment Mr. Le Sieur of USAID* and the Agency for their wisdom, their effort, and the financial support which made the preparation of the original volume possible. We, IDEA, express our sincere thanks for having their permission to reproduce this volume and make it available to the world desalination community. My thanks to them also for the splendid cooperation with our association which made this IDEA publication project a reality.

It is my sincere hope that you out there interested in desalination will find this book as informative and worthwhile as the members of the IDEA's Board of Directors who authorized the publication project.

R. Bakish
January 19, 1981
Teaneck, N. J.

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*United States Agency for International Development

THE U.S.A.I.D. DESALINATION MANUAL

A planning tool for those considering the use of desalination to assist in the development of water resources.

O. K. Buros
Engineer, CH2M HILL International Corporation

With Technical Advice and Contributions By
Robert Bruce Cox, Desalination Consultant, Centerport, New York
Isadore Nusbaum, Environmental Consultant, San Diego, California
Ali M. El-Nashar, Associate Professor, Florida Institute of Technology
Robert Bakish, Professor, Fairleigh Dickinson University

Editor—David Dunagan, CH2M HILL
Graphic Artist—Mark Bradley, CH2M HILL

PREFACE

This is the third major publication by the U.S. Agency for International Development (U.S.A.I.D.) that covers the spectrum of desalination processes. The first, entitled A Manual on Water Desalination, was published in two volumes in 1967. A one-volume supplement to that work was issued in 1972 to update the original material.

This volume, published in 1980, reviews the state of the art of the four major desalination processes and discusses renewable energy systems for desalination, process selection, and desalination for developing countries. This is an entirely new book and is independent of the preceding two publications. The organization, text, and graphics have been designed to assist the reader in finding and understanding the material presented.

The purpose of this book is to serve as a planning tool for the staff of U.S.A.I.D. and others. It will assist readers in becoming acquainted with the various desalination processes, their performance, the overall state of the art, and the associated potentials, problems, development, and economics. Thus armed, the reader can better evaluate the applicability of desalination to a particular project. This book is not intended to be a design manual, nor will it answer every question on desalination; however, it will provide a better awareness and understanding of the subject, thereby enhancing the reader's dealings with professionals in the field.

This book is divided into nine chapters and five appendices. Chapters 1 and 2 constitute an introduction and a discussion of saline water sources, respectively. Chapters 3, 4, 5, and 6 cover the four major desalination processes: distillation, freezing, reverse osmosis, and electrodialysis, respectively. The internal organization of each of the four chapters is almost identical. Thus the major headings, Theory, Commercial Development, Engineering Considerations, etc., appear in each of the chapters, permitting easy comparisons between processes.

Chapter 7 discusses the use of renewable or alternate energy sources for desalination. It briefly covers solar, wind, wave, and hand energy and gives examples of applications where these sources are being used for desalination.

Chapter 8 examines the factors to be considered in process selection. Included in this chapter is information that will aid in gathering data and making some initial order-of-magnitude estimates on potential costs involved.

Chapter 9 is a discussion, with examples, of desalination in developing countries. It emphasizes some of the special factors that should be considered in the decision to use desalination and in the selection of the process.

Appendix A includes the names and addresses of a selection of U.S. manufacturers of desalination equipment. Appendix B contains the names and addresses of most of the U.S. manufacturers of reverse osmosis membranes. In addition, a table in this appendix lists the characteristics of many of the commercial reverse osmosis membranes made in the U.S. at this time.

Appendix C contains worksheets to assist in organizing information on desalination projects and making order-of-magnitude cost estimates. This appendix also includes some cost curves to assist in cost estimating. Appendix D contains an important table, "Summary of Guidelines for Desalting Process Selection," which contains information on the characteristics of a wide variety of desalination processes. Properly used, this table should be of great assistance in the evaluation and selection of an appropriate process. The final Appendix, E, contains a short discussion and references on source material for desalination for those readers who would like to remain current with developments in desalination.

Both English and metric units are used in the text. English units are shown first with a metric equivalent following in brackets, e.g., 5 mgd [18,900 m^3/d]. Where this arrangement was not appropriate, such as in complex tables, etc., a metric conversion factor is generally shown on the same page. The desalination industry is international in scope, but units used in practice are often a mixture of the English and metric system and opinion is divided on what units or equivalents should be used. Even in the matter of the basic measure of water, there is no uniformity in the local term used for water measurement: gallons (U.S.), gallons (Imperial), tons (U.S.), cubic meters, barrels (42 gallons), etc. are used--occasionally along with units of another system in the same sentence.

The purpose of using the dual system of units in this text is to increase the reader's understanding of the material. Accordingly, in some instances the metric units more commonly used in the industry are used rather than the strict SI (Systémè International d'Unites) metric equivalents. A conversion table has been included in the manual.

This book is not one which typical readers would want to read straight through from cover to cover in one sitting. It is a reference book to be used as needs arise to know more about particular topics. The table of contents directs the reader to the relevant subject where it can be studied to the degree necessary.

For anyone generally considering desalination in developing countries, a good, but brief, preparation for the subject can be gained by reading Chapters 1 and 2, the Summary, and Chapters 7 (if the use of renewable energy is being contemplated) and 9, in that order. This should take only about an hour and will provide considerable basic information.

The production of this book was a team effort led by the principle author, Dr. O. K. Buros of CH2M HILL. Other team members included R. Bruce Cox, Isadore Nusbaum, Dr. Ali El-Nashar, and Dr. Robert Bakish. Messers Cox and Nusbaum assisted with the writing and Drs. El-Nashar and Bakish provided valuable research input and review.

Dr. Buros specializes, for CH2M HILL, in the evaluation and development of water resources for water-short areas. He has been actively working in various Carribean Islands on projects of this nature for almost a decade. Dr. Buros has authored a number of papers on the economics of desalination and the integration of desalination with water reuse and other water resource development practices.

Mr. R. Bruce Cox has worked on desalination for almost 20 years. He holds basic patents in both the freezing and the horizontal-tube multiple-effect distillation processes. Mr. Cox currently is the principal of his own consulting firm in Centerport, New York.

Mr. Isadore Nusbaum is an independent consultant from San Diego, California, who has worked on the development, design, and application of the reverse osmosis process since 1967. For much of that time he was staff consultant for the Fluid Systems Division of UOP, a leading producer of spiral membranes and reverse osmosis systems. He has participated in design and consulting work in connection with numerous reverse osmosis plants in the USA and elsewhere.

Dr. Ali El-Nashar has served on the faculties of various universities in Egypt, the United Kingdom , and the USA. He has worked in the field of reverse osmosis and distillation and is currently on the staff of the Florida Institute of Technology.

Dr. Robert Bakish is on the faculty of Fairleigh Dickinson University, where he directs the University's research effort and B.S. program in desalination. He has worked on metallurgical aspects of desalination for about 10 years and is a founding trustee and current president of the International Desalination and Environmental Association (IDEA).

SUMMARY

Although the process of desalination, mainly by distillation, has been known and used for centuries, its widespread commercial application for potable water production did not begin until the 1950's. Since that time, its application has become worldwide and has proven to be useful in treating brackish water and seawater to provide freshwater in many water-short areas. Four major desalination processes have been identified. These include processes based on distillation, electrodialysis, freezing, and reverse osmosis.

Progress has been rapid in the field of desalination. In 1967, when the first U.S.A.I.D. publication on desalination, A Manual on Water Desalination, was published, the worldwide installed capacity of desalination units had grown to about 250 mgd [946,000 m^3/d]. Distillation was the most widely used process for the desalination of both brackish water and seawater. Electrodialysis for brackish water desalination had enjoyed some commercial success with a number of installations in North America and the Middle East. Reverse osmosis and freezing were still in their developmental stages.

By 1972, when a supplement to A.I.D.'s A Manual on Water Desalination was issued, the worldwide installed capacity had about doubled. Distillation still dominated both brackish water and seawater desalination, but electrodialysis and reverse osmosis were beginning to make significant advances into brackish water applications. Reverse osmosis was developing using four types of membrane configurations (plate and frame, tubular, hollow fine fiber, and spiral), and these were given equal attention in the second edition. The freezing process for desalting seawater was being tried in pilot plants, but had still not achieved commercial success.

At the time of publication of this manual, the worldwide installed capacity has again doubled to over one billion gallons per day [3,800,000 m^3/d]. Although distillation continues to have the greatest amount of installed capacity, economic constraints are restricting its utilization almost exclusively to seawater. Reverse osmosis and electrodialysis are widely used in hundreds of locations for the desalination of brackish waters. These two membrane processes are considerably more economical than distillation for the treatment of brackish water. The electrodialysis applications have been quite successful in treating low-TDS brackish waters (1,000-4,000 mg/l). In about 1972, a new process improvement, electrodialysis-reversal (EDR), was introduced and over 100 units have been installed since. This improvement reduced some of the problems experienced in the standard electrodialysis process, and plants with capacities of up to 4 mgd [15,140 m^3/d] have been built using EDR.

With regard to reverse osmosis, the spiral and hollow fine fiber membrane configurations have almost completely dominated the field. Large installations in the 1- to 3-mgd [3,790- to 11,400-m^3/d]range have experienced many years of successful operation on brackish waters, indicating that a predicted membrane life of 3 years is quite reasonable. Other types of membranes operated at higher pressures have been developed for seawater desalination. A number of seawater plants have been in service from 1 to 3 years, and experience is being gained in their operation. In some applications, seawater reverse osmosis installations appear to be quite competitive with distillation, although in very large facilities, distillation has many economies-of-scale and can benefit from the use of low-cost steam in dual-purpose (water-electricity) installations.

Although the freezing process has still not been commercially successful, work continues on it and many potential applications may emerge in the future.

Aside from solar stills, renewable energy systems, such as wind, solar, and wave power, have not received much commercial application for desalination. However, awareness of and interest in the subject has sharply increased, and a number of projects are being undertaken around the world. Many questions as to the design, feasibility, reliability, and economic viability of desalting with these renewable energy systems should be answered during the next 10 years.

There is considerable potential for using desalination to improve the utilization of water resources in developing countries. However, the production of desalted water is not inexpensive. Potential costs of $1 to $4/kgal [$0.26 to $1.06/m^3] for brackish water and $5 to $30/kgal [$1.32 to $7.93/m^3] for seawater usually encompass the overall unit costs (capital recovery plus operation and maintenance) of desalination. Items such as special feed-water development (well fields, intake structures, etc.), water storage, and brine disposal can add another $1 to $5/kgal [$0.26 to $1.32/m^3] to this cost depending on the site-specific conditions. Figures S-1 and S-2 show potential capital and unit costs for desalination facilities involving various processes.

In any country, no matter what the level of development, the best selection of a desalination or any water resource system is one which is more than just economically reasonable in the paper study stage. It is a system which works when it is installed and continues to work and deliver the desired product at the expected quality and quantity for the planned life of the system.

WARNING

1. Caution should be exercised when using these cost data.
2. These costs are computed for basic desalination facilities in the USA. These costs include no storage capacity and allow for only fundamental feedwater development and brine disposal. If storage capacity is added and/or more complex feedwater or brine facilities are required, overall project costs can significantly increase. This is especially true with brackish water facilities. Construction in locations outside the USA can also add to the overall cost. Actual costs of up to 300% (or more) of those shown on this figure (especially at lower capacities) would not be unusual for projects in many developing areas.

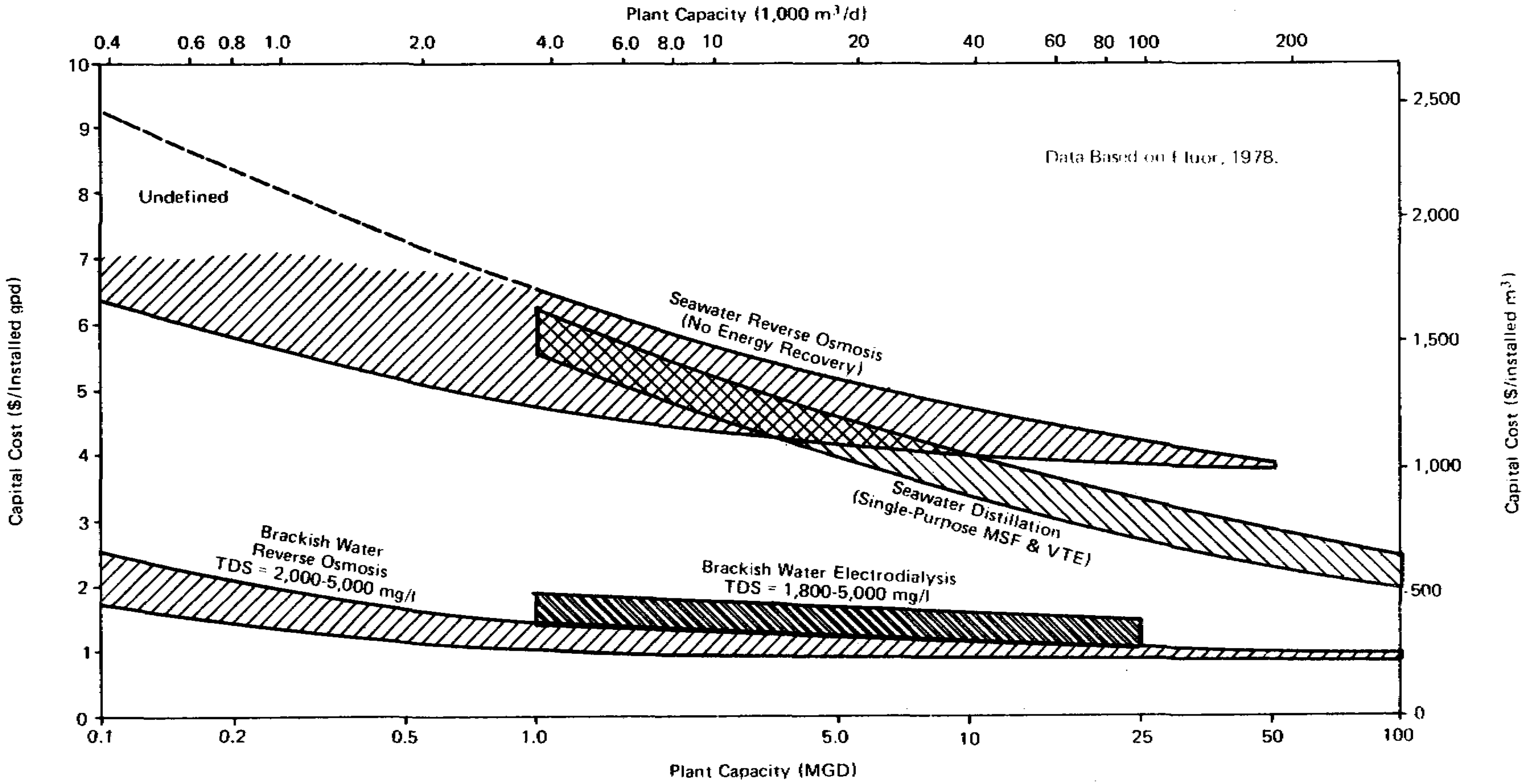

FIGURE S-1. Generalized capital cost estimates for desalination processes.

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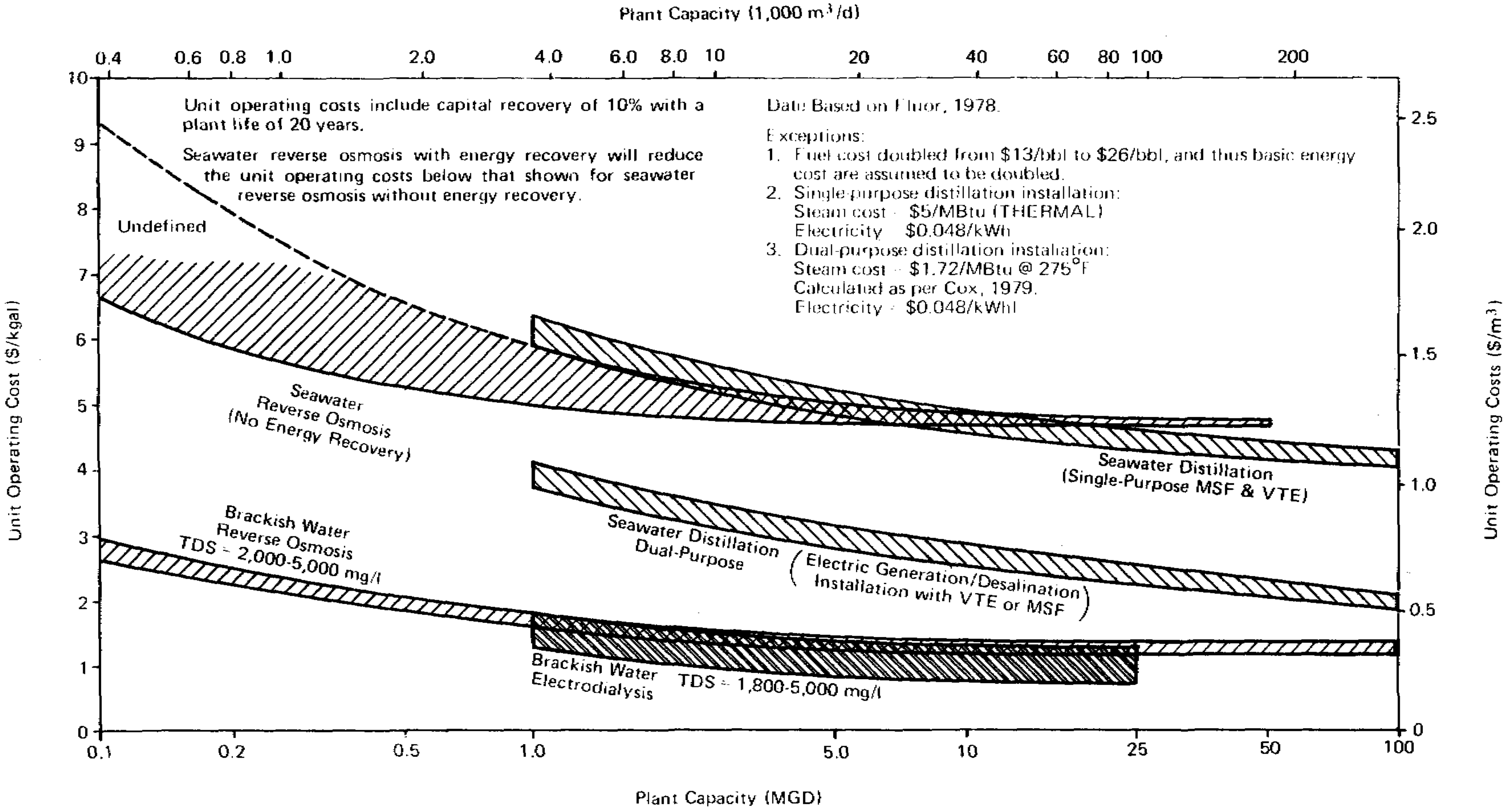

Unit operating costs include capital recovery of 10% with a plant life of 20 years.

Seawater reverse osmosis with energy recovery will reduce the unit operating costs below that shown for seawater reverse osmosis without energy recovery.

Data Based on Fluor, 1978.

Exceptions:
1. Fuel cost doubled from $13/bbl to $26/bbl, and thus basic energy cost are assumed to be doubled.
2. Single-purpose distillation installation:
   Steam cost = $5/MBtu (THERMAL)
   Electricity $0.048/kWh
3. Dual-purpose distillation installation:
   Steam cost = $1.72/MBtu @ 275°F
   Calculated as per Cox, 1979.
   Electricity = $0.048/kWhl

FIGURE S-2. Generalized unit production cost estimates for desalination processes.

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Figure

FIGURES--Continued

Figure

FIGURES--Continued

FIGURES--Continued

Figure

FIGURES--Continued

Figure

FIGURES--Continued

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Water Purification Systems
Water Services of America

ABBREVIATIONS

atm - Atmosphere
BC - Baja California
BCS - Baja California del Sur
BLS - Bureau of Labor Standards
Btu - British thermal unit
CA - Cellulose acetate
CEA - Commissariat à l'Energie Atomique
CFE - Multistage controlled flash evaporator
cm - Centimeters
Coah - Coahuila
CSMCRI - Central Salt and Marine Chemicals Research Institute
CTA - Cellulose triacetate
°F - Degree(s) Fahrenheit
DIGAASES - Direccion General de Aprovechamiento de Aguas Salinas y Energia Solar
DCC-MSF - Direct contact condensation multistage flash
ED - Electrodialysis
EDR - Electrodialysis-reversal
ENR - *Engineering News-Record*
FFNSF - Falling film multistage flash
fps - Feet per second
FRP - Fiberglass reinforced plastic
FW - Feedwater
G - Giga (1 billion)
G&A - General and administrative (overhead)
gm - Gram
gpd - Gallon(s) per day
gpm - Gallon(s) per minute
HFF - Hollow fine fiber
Hg - Mercury
HTME - Horizontal-tube multiple-effect
ID - Inside diameter
IDE - Israel Desalination Engineering, Ltd.
IDEA - International Desalination and Environmental Association
J - Joule
k - Kilo (1 thousand)
kg - Kilogram
kgal - Thousand gallons
kW - Kilowatt
kWh - Kilowatthour
l - Liter
lb - Pound
LTV - Long tube vertical
M - Mega (1 million)
m - Meter
MBtu - Million Btu's
MECO - Mechanical Equipment Company
MEMS - Multiple-effect multistage

meq - Milliequivalent
MES - Multiple-effect stacked
MG - Million gallons
mgd - Million gallons per day
mg/l - Milligram(s) per liter
mm - Millimeter(s)
MSF - Multistage flash
MSF/FBE - Multistage flash/fluidized bed evaporator
NL - Nueva Leon
NWSIA - National Water Supply Improvement Association
OD - Outside diameter
O&M - Operation and maintenance
ORNL - Oak Ridge National Laboratory
OSW - Office of Saline Water
OTEC - Ocean thermal energy conversion
OWRT - Office of Water Research and Technology
Pa - Pascal
PF - Performance factor
PR - Performance ratio
psi - Pound(s) per square inch
psia - Pound(s) per square inch--absolute
PVC - Polyvinylchloride
PW - Product water
RO - Reverse osmosis
SHMP - Sodium hexametaphosphate
S.S. - Stainless steel
ST - Submerged tube
SW ED - Seawater electrodialysis
SWRO - Seawater reverse osmosis
TDS - Total dissolved solids
U.S. A.I.D. - U.S. Agency for International Development
USBR - U.S. Bureau of Reclamation
VC - Vapor compression
VFEA - Vacuum-freezing ejector-absorption
VFRC - Vacuum-freezing refrigeration compression
VFVC - Vacuum-freezing vapor-compression
VIWAPA - Virgin Islands Water and Power Authority
VTE - Vertical-tube evaporator
VTFE - Vertical-tube foam evaporation
W - Watt

CONVERSION FACTORS

acre	x	4,047	=	m^2
atm	x	1.033	=	kg/cm^2
atm	x	760	=	mm Hg
bar	x	100,000	=	N (Newton)/m^2
atm	x	101,330	=	Pa (Pascal)
atm	x	14.7	=	psi
bbl	x	42	=	gallons (U.S.)
Btu	x	1,055	=	J
¢/kgal	÷	3.785	=	¢/m^3
cfm	x	0.472	=	l/s
$/kgal	÷	3.785	=	$/m^3
$/gpd	x	264	=	$/m^3/day
(°F-32)	x	0.556	=	°C
fps	x	0.3048	=	m/s
ft	x	0.3048	=	m
ft^2	x	0.0929	=	m^2
ft^3	x	0.0283	=	m^3
gpm	x	0.0631	=	l/s
gpm	x	5.45	=	m^3/d
gal (U.S.)	x	1.201	=	gal (Imperial)
hp	x	0.7457	=	kW
in	x	2.54	=	cm
Langley/min	x	221	=	Btu/ft^2/hr
Langley/min	x	1	=	cal/cm^2/min
Langley	x	1	=	cal/cm^2
Langley/min	x	697	=	697 W/m^2
kgal/day	x	3.785	=	m^3/day
kWh/kgal	÷	3.785	=	kWh/m^3
lb	x	0.454	=	kg
lb/kgal	x	119.7	=	g/m^3
lb/1,000 Btu's	x	0.429	=	kg/MJ
m^3	x	264	=	gal
mile	x	1.609	=	km
mph	x	1.609	=	kmph
MG	x	3,785	=	m^3
mgd	x	3,785	=	m^3/d
psi	x	6,895	=	N/m^2
tons (short)	x	0.907	=	Metric tons

CHAPTER 1

Chapter 1
INTRODUCTION

As the world's population has increased and/or achieved a better standard of living, the amount of freshwater needed and used has increased drastically. In some cases this has severely strained available freshwater supplies and has adversely affected development and citizens' everyday lives.

In the past, brackish water and seawater were not considered developable resources for potable or agricultural use. The high levels of salts in these waters made it unpalatable for drinking and impossible to use for normal agriculture. However, the use of desalination processes has now made these water resources available for man's use.

"Desalination" (also known as "desalting" or "desalinization") refers to the process of removing salts from water. Desalination does not create water where it did not exist before but does make possible the use of waters which were heretofore too salty. The concept of desalting is not new, having been discussed and experimented with for almost 2,000 years. What is new is the development of a number of processes over the past 50 years which has demonstrated their commercial viability. This work has been intensified in the last 20 years, and units using various processes have been built and installed around the world. Figure 1-1 shows the growth in desalination plants since 1950 and the distribution of desalination capacity around the world.

Five major processes have been used: ion exchange, distillation, freezing, reverse osmosis, and electrodialysis. Ion exchange has usually been used as a polishing process for industrial water processes. It is covered in most standard water treatment texts and is not discussed in this volume. The remaining four are covered in this text in some detail. Distillation, reverse osmosis, and electrodialysis have become commercially available and have been used rather widely during the past 10 years. The freezing process has not yet attained the commercial success of the other processes, but work is still going on and some knowledgeable people have predicted a substantial market for freezing desalination in the future. For this reason, this process and its variations are discussed in this manual.

Of the three commercially proven processes discussed in this manual, there is no one "best" method for desalination. Generally, distillation and reverse osmosis are used for seawater desalting, whereas reverse osmosis and electrodialysis are used for brackish water desalination. However, the selection and use of these processes can be very site-specific; they must be selected very carefully, especially in developing countries. Circumstances can arise which would alter the selection in order to provide the most effective method for a particular area.

This manual discusses the phenomena of saline water, the four desalination processes, the use of renewable energy sources for desalination, the selection of desalination processes, and the use of desalination in developing countries. The manual is intended to be used as a guide for personnel from the U.S. Agency for International Development (U.S. A.I.D.) in considering desalination applications around the world. It is not intended, nor should it be used as, the sole source of information and guidance for the selection, design, and/or installation of desalination facilities. Desalination and water resources are rather complex and can be quite site-specific in nature. Before any decision or commitment is made regarding desalination in a particular location, competent professional assistance should be secured to evaluate the application.

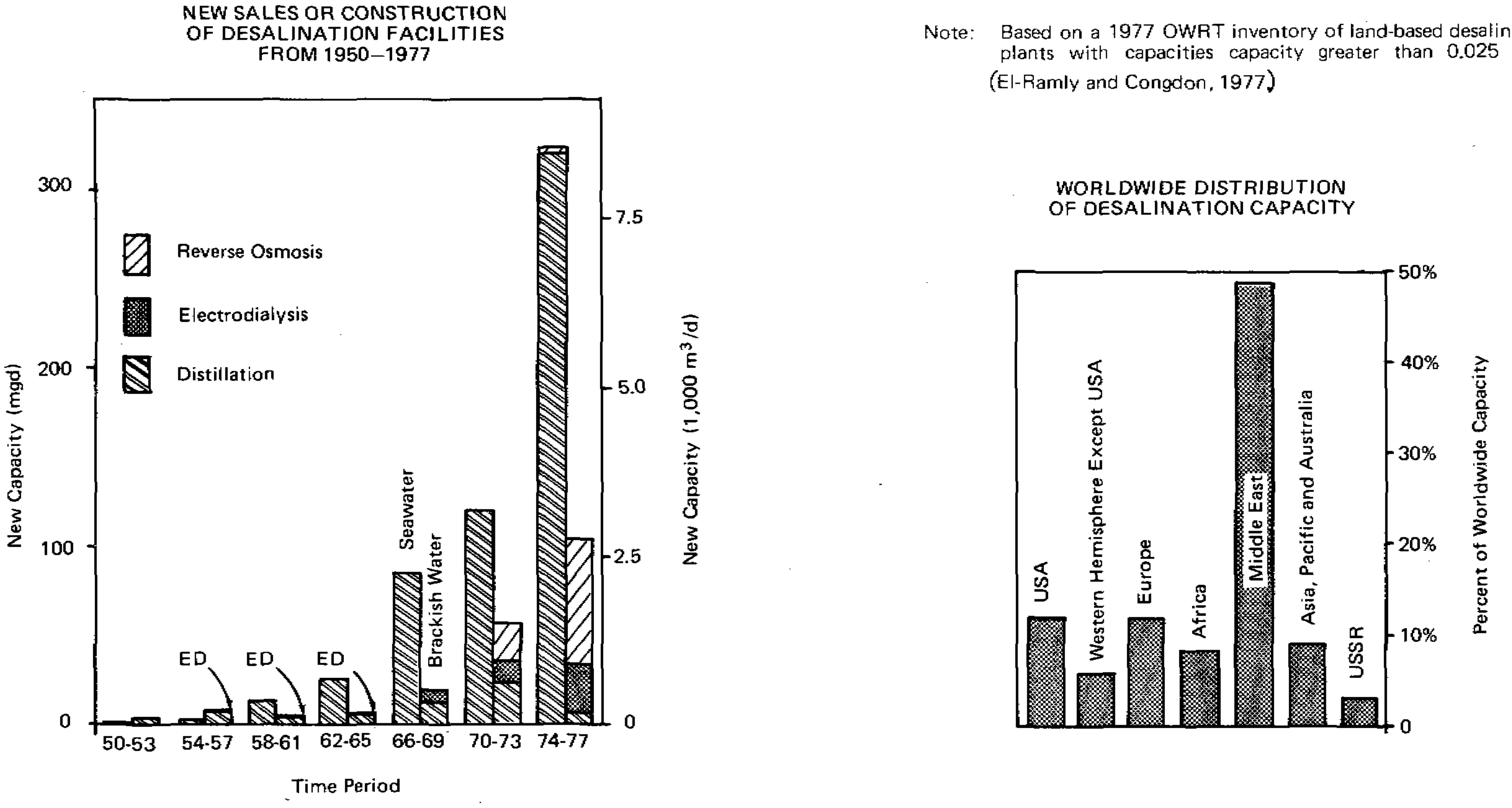
One of the major considerations in the use of a desalination process is its cost. In comparison to the costs of most water sources developed and treated in the USA by more conventional means, it is expensive--often from 2 to 50 times as expensive.

Although there have been considerable efforts toward the reduction of costs, desalinated water is still expensive. The process is inherently energy-intensive and improvements in equipment and efficiency have been offset by rapid increases in energy costs. However, despite the substantial costs involved, the availability of desalted water can be an economic boon for an area. Where water is scarce, it is often supplied by transporting long distances by truck or animal. If this water is sold, the unit price often approaches or exceeds that of desalted water. Hence, the economic conditions to support desalination already exist in many water-short areas.

Desalination is not the only method to effectively increase water supplies or sources in arid lands. Other techniques of water management, reuse, and conservation exist which, given appropriate circumstances, can also be useful. Nevertheless, the proper application and use of desalination can be a powerful tool in increasing the well-being of people in water-short areas.

Although this manual will assist the reader in the understanding of desalination, it is not intended to take the place of professional experience and advice, which should be used throughout projects.

Note: The left-hand column (or bar) refers to seawater installations and the right-hand column refers to brackish water installations.



Note: Based on a 1977 OWRT inventory of land-based desalination plants with capacities capacity greater than 0.025 mgd. (El-Ramly and Congdon, 1977)

Figures adapted from Buros, 1978.

FIGURE 1-1. Worldwide construction and distribution of desalination facilities.

CHAPTER 2

Chapter 2
SALINE WATER

GENERAL

Saline water, that is, water containing a significant amount of total dissolved solids (TDS), constitutes a resource which can be treated by desalination to produce freshwater for potable, industrial, and agricultural purposes.

Most saline waters include small quantities of organic material and dissolved gasses, but the majority of the dissolved materials are inorganic salts. Salts, by definition, are compounds made up of positive metallic ions or radicals (e.g., calcium, sodium, ammonia) bonded to negative ions (e.g., chlorides, sulfates, carbonates). Many of these salts are quite soluble in water and are generally derived from the weathering and eroding of materials on the earth's surface and/or from volcanic activity, which releases constituents such as hydrogen chloride, hydrogen sulfide, carbon dioxide, etc., from rocks in the earth's interior.

These constituents are dissolved in running or standing water, which generally flows to the seas or is trapped in inland locations. This continual load of dissolved salts is deposited in the oceans and seas, where the average level of total dissolved solids is about 35,000 mg/l.

The seas make up about 97 percent of the world's inventory of water, which is constantly recycled by evaporation and condensation to produce rain. Some of the dissolved salts within the ocean are processed by marine organisms into various materials and gases, thereby taking the salts out of solution. These materials include corals, algae, diatoms, etc. As the earth's surface relative to the ocean has shifted in elevation, large quantities of seawater have been trapped inland and have formed deposits of rock salt, gypsum, limestone, etc., which are eventually redissolved by rain and are returned to the ocean.

CLASSIFICATIONS

For convenience in discussions of desalination, water is divided into four classifications: freshwater, brackish water, seawater, and brine. Freshwater generally covers waters with a TDS of up to 1,000 mg/l, brackish water from 1,000 to 35,000, and seawater above 35,000 mg/l.

Brine is the water remaining in a salt solution after some of the pure water has been removed by desalination. The brine contains nearly all the salts that were in the original solution; in brine they are simply dissolved in a smaller volume of water.

In certain circumstances, brackish water is considered to be restricted to saline water with a TDS content of less than 10,000 mg/l, the range between 10,000 and 35,000 mg/l often being referred to as "difficult" brackish water.

Brackish Water

Brackish water can be found in lakes and streams and under the earth's surface (as ground water). The salts are usually added by (1) the process of water dissolving or eroding salt deposits present on or in the earth's surface and/or (2) the process of water mixing with other saline sources of water, including seawater.

Brackish water can have a wide variety and combination of constituents, and each combination can present a different problem for desalination. Table 2-1 shows a partial analysis of a selection of brackish waters around the world.

Brackish water usually occurs naturally due to four major processes: (1) surface water or ground water flowing through or over mineral deposits and dissolving them or mixing with saline ground water; (2) saline water having been trapped in a formation (connate water) when it was formed in an earlier geological period; (3) water containing dissolved salts enter a closed basin and begining to evaporate, leaving the salts behind dissolved in any remaining or newly entered water; and (4) freshwater and seawater intermixing where they interface in bays, estuaries, and/or subsurface formations along a coast.

Man's activities add to the quantity of brackish water in several significant ways. The first of these is the use of water for various domestic, industrial, and agricultural purposes, which add salts to the water. Typical domestic usage can raise the TDS by several hundred milligrams per liter and agriculture or industrial usage can add even more. As successive use and reuse of water occurs along a river, the salinity continues to build up. This is part of the salinity problem on the lower Colorado River in the southwestern USA, where a level of TDS has been produced that is unacceptable to Mexico, which receives the flow of the Colorado River near the town of Yuma, Arizona. In this case, desalination, in the form of reverse osmosis, will be used to reduce the salt content of the river.

A second means by which brackish water formation is stimulated is the contamination of ground water by interconnecting water-bearing zones of differing qualities. Such contamination can result from the construction of wells for the extraction of water, oil, etc., which do not seal off poor-quality water either during operation or after the well is abandoned.

Other means include contamination by discharge (disposal of saline solutions on or below the earth's surface) and ground-water contamination by seawater intrusion due to improper management of well fields located adjacent to the sea.

Table 2-1
IONIC COMPOSITION OF SELECTED BRACKISH WATER

Constituent	Location of Well Corfu, Greece Well[a]	 Yuma, Arizona Well[b]	 Virgin Islands Well[c]	 Riyadh, Saudi Arabia Well[d]
Chloride (Cl^{-1})	36	1160	330	300
Sodium (Na^{+1})	15	900	--	220
Sulfate (SO_4^{-2})	1147	916	--	500
Magnesium (Mg^{+2})	70	0.4	196*	48
Calcium (Ca^{+2})	476	226	204*	170
Potassium (K^{+1})	--	8	--	--
Bicarbonate (HCO_3^{-1})	329	437	--	--
Carbonate (CO_3^{-2})	--	--	620*	--
Iron (Fe^{+3})	--	0.24	--	--
Silica (S_1O_2)	--	33	--	25
Hardness (as $CaCO_3$)	--	--	400	625*
TDS	2073	3746	2100	1470

Notes: Values in mg/l, -- = Not given, * = as $CaCO_3$.

[a] Andreadis and Arnold, 1978.
[b] Van Hoek and Mavis, 1977.
[c] Buros, 1976a.
[d] Ghulaigah and Ericsson, 1979.

Seawater intrusion into the ground water is a major consideration in any coastal or island location that has an interface with highly saline water. Where well field development and/or pumping exceeds the safe yield of an area, seawater can be drawn into the well fields, causing a rapid deterioration of the water quality. (Some of the problems involved in seawater intrusion are shown on Figure 2-1). The result is that the ground-water TDS level increases, thereby restricting the beneficial usage of the ground water and often necessitating the use of desalination where none was needed previously.

Seawater

With desalination, seawater can be used as a source of freshwater. Since desalting seawater is generally more energy-intensive than desalting brackish water, seawater is usually considered for desalination only when sufficient brackish water is not available. All of the world's oceans are interconnected, and the general composition of seawater in the open ocean is relatively consistent, with a TDS of about 34,400 mg/l. However, in specific areas, seawater can vary considerably in salt concentration, even though the relative proportion of various constituents to the chloride ion does not generally vary significantly. Land masses, heating and cooling, currents, winds, and tides can all affect the concentration of a saltwater body at any location and time.

Table 2-2 shows the standard ionic composition of seawater as well as some of the variations in seawater in various places around the world.

Other Sources of Saline Water

Two other sources of saline water for desalination are occasionally considered as feedwater for desalination processes: wastewater and agricultural drainage water.

Wastewater. Unlike naturally occurring seawater and brackish water, discussed previously, wastewater tends not to be uniform in content. It contains whatever people dispose of down a drain. Today it may be dishwater and tomorrow it could be a highly potent pesticide or an industrial chemical. Although its average characteristics in an area can be estimated with a sampling program, its characteristics at any certain time cannot.

If any type of potable use is contemplated with the wastewater, treatment must be provided which will handle not only the average wastewater constituents but the surprises as well.

Wastewater is being used as a feedwater to a number of desalination plants, almost all of which use the reverse osmosis method of desalination. The purpose of these desalination plants is to reduce the salt content of the wastewater--they are not being used for general treatment. In fact, it is usually necessary to treat the water extensively before it is desalinated to avoid

Table 2-2
IONIC COMPOSITION IN SEAWATER (in mg/l)

Constituent	Normal Seawater[a]	Eastern Mediterranean[b]	Arabian Gulf At Kuwait[c]	Red Sea At Jeddah[d]	Atlantic Ocean At Wrightsville Beach, NC, USA[e]
Chloride (Cl^{-1})	18,980	21,200	23,000	22,219	19,102
Sodium (Na^{+1})	10,556	11,800	--	14,255	--
Sulfate (SO_4^{-2})	2,649	2,950	3,200	3,078	2,637
Magnesium (Mg^{+2})	1,272	1,403	1,765	742	1,204
Calcium (Ca^{+2})	400	423	500	225	356
Potassium (K^{+1})	380	463	--	--	--
Bicarbonate (HCO_3^{-1})	140	--	--	146	--
Strontium (Sr^{-2})	13	--	--	--	--
Bromide (Br^{-1})	65	155	80	--	--
Boric Acid (H_3BO_3)	26	72	--	--	--
Fluoride (F^{-1})	1	--	--	--	--
Silicon (Si^{-4})	--	--	1.5	--	0.7
Iodine (I^{-1})	<1	2	--	--	--
Other	1	--	--	--	--
Total Dissolved Solids	34,483	38,600	45,000	39,884	37,144

Note: -- = Means much less than 1 mg/l or not given.

[a] Normal seawater as per Sverdrup et al., 1946.

[b] Kaiser, 1966.

[c] Hickman et al., 1979.

[d] Harashina, 1977.

[e] Patel et al., 1976.

problems, especially if potable use of the resulting desalted water is intended. The desalted water from these plants has not been devoted to direct potable use, although investigations are going on for such use.

The most notable example of desalination being used in conjunction with wastewater is a 5-mgd [18,900-m^3/d] reverse osmosis plant which desalts a portion of the effluent from an advanced wastewater treatment plant in Orange County, California.

Utilizing wastewater as a feedwater involves not only the potential for additional expense due to the pretreatment needed but also involves added public health risks. If other sources of water are available for desalination it would be prudent to use them rather than wastewater.

Often, wastewater which has been properly treated by standard processes (not including desalination) can be used for irrigation and/or ground-water recharge. This can aid in alleviating overall water resource deficiencies. In addition, recharged water can be recovered from ground-water sources and then, if necessary, further treated by desalination. The ultimate treatment should match the application.

Agricultural Drainage Water. The drainage from irrigated areas also forms a potential source of brackish water for use as a feedwater for desalination. This has been discussed and/or experimented with in many countries, including Egypt, Mexico, Saudi Arabia, and the USA (Al Taweel et al., 1980; El-Nashar, 1976; Manjarrez and Gonzalez, 1976; McCutchan et al., 1976). Both reverse osmosis and electrodialysis processes have been used, and extensive experiments have been carried out as part of the design and component testing for the large (approximately 96-mgd [363,000-m^3/d]) brackish water desalination facility to be built at Yuma, Arizona, during the 1980's.

Agricultural drainage water varies in quality and quantity with the location, season, farming practices, etc. It is possible to reduce the salt level in the water by various desalination processes, but the water must first undergo some preliminary treatment to remove undesirable materials which could cause difficulties with the desalination process. The extra cost involved in this additional treatment generally makes other sources, such as ground water, more economical, but in special site-specific circumstances, irrigation return waters might be used as a source. In such cases, the effects on the various desalination processes pesticides, herbicides, and various other agricultural chemicals need to be carefully considered for the application under study.

SOURCE DEVELOPMENT

One key to the successful planning and operation of a desalination facility is proper characterization and development of the raw water source.

Characterization

In order to properly design and procure a desalination unit, the characteristics of the raw water supply must be known and specified. These characteristics usually include the chemical constituents, non-dissolved solids level, microbial content, and temperature. It is important to realize that these characteristics may change with time and/or quantity withdrawn. These changes could occur daily, annually, seasonally, etc., depending on the circumstances.

An adequate testing program is essential for many facilities to avoid mistakes in design and later problems in operation. For ground water, it is prudent to complete the production wells and to thoroughly test them at the expected production rates before drawing up the final specifications. Hydrogeologists also should conduct suitable tests to provide estimates of the future water quality and quantity available from the well field.

With seawater sources, samples should be taken at various times and under different conditions in order to properly characterize the feedwater. In one site on the Red Sea where a distillation plant was planned, the TDS of the seawater was found to change as much as 9,000 mg/l on a daily basis due to tidal action. A single measurement at the wrong time could have led to improper design of a unit in this area. In a Florida location, the TDS of brackish ground water in parts of a confined aquifer which served an electrodialysis plant almost doubled from 1978 to 1980. This has resulted in the expensive abandonment of several of the wells drilled for that facility (Watson and Derowitsch, 1979). These are just two examples of the many situations that can arise, pointing up the need for careful source characterization.

Development

As will be discussed in the following chapters, all of the desalination processes require pretreatment of some type to condition the raw water supply before the actual desalting step. The extent and effectiveness of this pretreatment step can be affected by the care taken in the development of the raw water source.

With proper planning and design, the amount of pretreatment can be reduced and the quality of the water delivered can be increased, providing long-term benefits in reduced maintenance costs and increased operating time of the desalting facility.

Wells designed to collect ground water are more than just holes in the ground. If they are of any significant size, they should be designed by competent engineers and carefully constructed. This may include the use of test holes, geophysical logging, placement of well screens, gravel packing, sealing of unsuitable zones, and/or development of the finished wells. The materials selected for construction of the well, pump, and water transmission line should be compatible with the water involved. For instance, steel-cased wells would be inadvisable in most brackish water or seawater locations because they can create problems due to corrosion of the casings and the pumping of loose corrosion products to the desalination facilities.

Although seawater can occasionally be obtained using seawells on the adjacent beach, the geology of the site or the quantity of water needed often does not favor this method of obtaining seawater for large plants. In these cases, a seawater intake is constructed which brings in seawater directly.

The intake must be located in an area where it will pick up high-quality water and will avoid silts, sand, floating, dissolved, and suspended wastes, as well as marine life. The process of selecting the intake location must take into account the soil's structural characteristics, the intake's effect on the marine environment, the effect on currents and navigation, as well as its potential for being damaged by storms or representing an aesthetic detriment. Proper location and construction can aid in obtaining suitable water in the desired temperature and salinity range.

An outfall must also be provided for the discharge of brine from desalination plants located by the sea. The outfall should be sited where it will cause minimal short-circuiting back to the intake.

In short, a sufficient effort and budget should be planned for the proper characterization and development of a raw water source for desalination facilities.

WATER QUANTITY

There is no minimum quantity of saline water necessary for desalination purposes. Desalination process selection is so flexible that a unit can be matched (although perhaps not economically) to any quantity of water available.

As will be explained in each of the chapters on the four basic processes, different processes (and variations within those processes) have different water requirements which must be met. These requirements and related economics must be considered in the ultimate selection of a desalination process design.

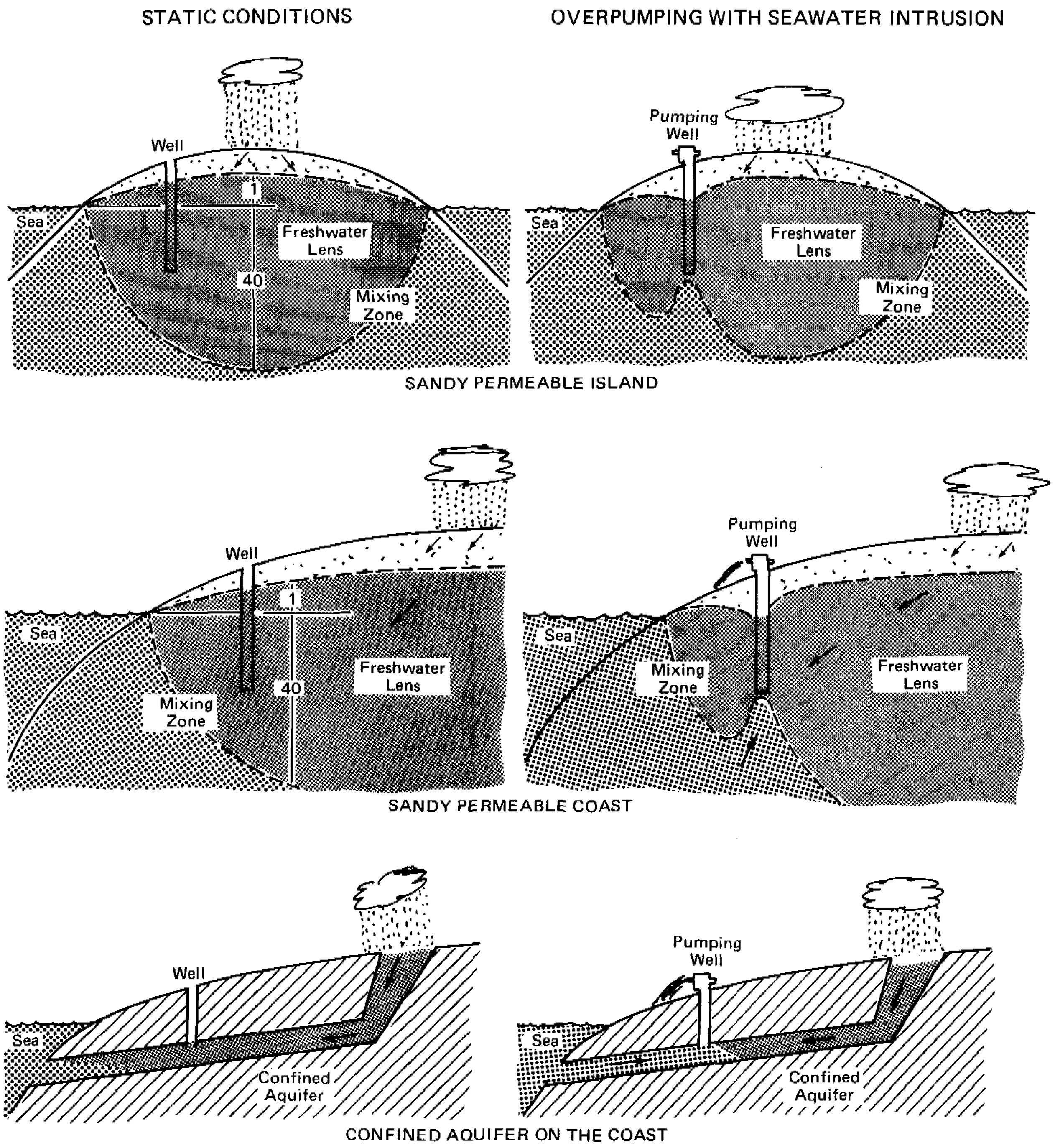
A wide range of commercially produced units is available to meet the varying water requirements both for feedwater and product water. These can range from a 1-gal/hr [4-l/hr] hand-pumped reverse osmosis unit to a 50-mgd [189,000-m^3/d] distillation plant.

DISPOSAL

The desalination processes discussed in this book treat saline water and produce two resultant streams. One is a product stream, which has a low level of TDS, and the other is the brine (or reject) stream, which has a higher level of dissolved salts than either the feedwater or the product water stream.

The brine stream must be discharged somewhere. For facilities located adjacent to the sea, disposal does not generally pose a problem, but it can be a serious problem for facilities inland. Improperly discharged brine can contaminate existing ground water or surface water due to its high salt content. It is generally too salty to be used for irrigation, so that disposal options are limited. Some methods that have been used include evaporation by solar (or conventional thermal) means, injection of the brine into existing zones of very saline ground water, and transportation of the brine to a saline water body.

Whatever the solution to the disposal situation, it should be investigated concurrently with any consideration of desalination. In some cases, the disposal problem can affect the selection of the desalination process as well as the overall viability of desalination as a water resource option.



The process has been greatly simplified for this illustration.

The lighter freshwater floats on the denser seawater, forming a lens. Each foot (or meter) of freshwater above sea level is balanced by 40 feet (or meters) of freshwater below sea level.

Since the fresh ground water is always flowing out to sea, the freshwater lens diminishes unless recharged by additional freshwater (such as rain).

When wells are pumped, there is a natural drawdown of water around the well. For each foot (or meter) of water that is drawn down, the bottom of the lens moves up 40 feet (or meters).

If the movement is too great, the saltwater moves into the well. Thus, caution needs to be exercised in locating and pumping wells on land adjacent to saltwater.

FIGURE 2-1. Seawater intrusion on islands and coastal areas.

CHAPTER 3

Chapter 3
DISTILLATION

THEORY

The theory of distillation is not new. It has been employed by alchemists, chemists, and the like for the separation of liquids (alcohol from water) or dissolved solids from liquids (salt production) for centuries. Distillation of saline water for potable use was of early interest to sailors on long sea voyages as a method of supplementing their stored water supplies. Patents were issued in the 17th century in England for commercial distillation units (Baker, 1948).

As necessity, technology, and economics have developed, distillation has come ashore and is now the most extensively used method of seawater desalination for the production of potable water. One of the earliest examples was the installation of plants along the Suez Canal in the late 19th century (Baker, 1948). The Weir Company of Scotland, using their experience with shipboard stills, installed a unit for the Safajah Phosphate Company in Egypt in 1912, which produced 16,800 gpd [64 m^3/d] and operated, with modifications of course, into the 1970's.

Essentials

There are three essential steps in the production of potable water by distillation:

Vapor production.

Transportation of vapor to a condenser.

Condensation.

Vapor Production. One of the reasons that distillation can be so successfully used to produce potable water from saline water is that natural saline water contains no other substances, in significant quantities, which will vaporize in the temperature range used in the process. This is not the case with fermented aqueous solutions (to make alcohol) or petroleum, both of which require considerably more care in the ultimate separation of two or more substances which readily vaporize at temperatures close to one another.

i) Evaporation--Evaporation (or vaporization) is a surface phenomenon in which water molecules accumulate sufficient energy to allow them to escape from the bulk liquid and enter the space above it. Evaporation occurs with any body of water which has a gas-liquid interface. The reason for this is that the energy contained in the water molecules is not distributed evenly but rather is distributed so that, although most molecules have near to average energy, some have less and others more. Those with sufficiently higher energies leave the water surface and enter the gaseous space above the liquid as vapor. If energy, such as heat from the sun, is continually added the liquid continues to slowly evaporate.

$$\text{Liquid} + \text{Energy} \rightarrow \text{Vapor}$$

Under any circumstances, the amount of vapor that can be produced is limited by the available energy, but evaporation can be accelerated by applying energy more rapidly or by increasing the surface area available for the molecules to escape. In any case, the molecules must acquire an energy during the process equal to the heat of vaporization in order to escape from the surface. This energy in the vapor molecule is lost by the liquid during evaporation, resulting in a cooling of the liquid. Hence, the cooling effect of perspiration evaporating from the body.

ii) Vapor Pressure--When water evaporates in a closed container the vapor molecules exert a pressure on the walls of the container and on the surface of the water. This pressure is referred to as vapor pressure. When water is originally placed in a closed container it continues to evaporate (and the vapor pressure increases) until the number of water molecules leaving the liquid equals the number returning to it. At this point an equilibrium between the liquid and vapor phase has occurred

$$\text{Liquid} \rightleftarrows \text{Vapor}$$

and an equilibrium vapor pressure exists. At this point the temperature of the liquid does not change due to evaporation, because molecules returning to the liquid give up their acquired "heat of vaporization" as an equivalent "heat of condensation" in the phase change from vapor back to liquid.

If something upsets the equilibrium of the system, such as an increase in energy or a change in the gas volume (which would change the pressure), there is a change in the number of water molecules entering or leaving the water surface in an attempt to restore equilibrium. The actual equilibrium vapor pressure is dependent on the nature and temperature of the liquid involved. For every water temperature there is a corresponding equilibrium vapor pressure.

iii) Boiling--When the vapor pressure of the liquid is exactly equal to the existing external pressure, any addition of heat causes the formation of vapor bubbles in the body of liquid. This action is called boiling. The boiling point is defined in terms of absolute pressure (psia or atm) and temperature (°F or °C), because for every pressure there is an exact corresponding temperature. For water at atmospheric pressure, the boiling pressure is 14.7 psia [1 atm] and the temperature is 212°F [100° C]. As the pressure on the liquid is increased (as in a steam boiler), the boiling temperature increases. As the pressure is reduced, the boiling temperature decreases. It is actually possible to reduce the pressure on water sufficiently that the water will boil at its freezing point, 32°F [0°C]. (This phenomenon is used in some freezing processes for desalting seawater, as discussed in Chapter 4.) The relationship between boiling temperature and pressure is shown on Figure 3-1.

An important phenomenon which occurs in the distillation process is boiling point rise. Boiling point rise is the elevation of the boiling point of a liquid (in this case water) due to the amount of salts in solution.

iv) Heat of Vaporization/Condensation--In order for water to undergo a phase change from liquid to vapor, it is necessary to acquire energy equal to the heat of vaporization. In numerical terms this is equal to 970 British thermal units (Btu's) per pound of pure water at 212°F [2,256 kilojoules (kJ's) per kilogram of pure water at 100°C].

This should be compared to the energy of 1 Btu necessary to increase the temperature of a pound of water 1°F [4.2 kJ/kg/°C], or 180 Btu's to raise 1 pound of water [418 kJ per kg] from the freezing point to the boiling point at atmospheric pressure. Hence, the heat of vaporization is a very significant energy requirement in the distillation of water.

In order for water vapor to condense to a liquid, it is necessary that the heat of condensation, 970 Btu/lb [2,256 kJ/kg], be removed. (The heat of vaporization and the heat of condensation are equal at any given temperature. They both increase as the boiling temperature is reduced.)

For desalination by distillation to work efficiently, it is necessary to recover or reuse as much of the heat of condensation as possible. This is often done by condensing water vapor on one side of a heat transfer surface while simultaneously utilizing the other side of the surface (such as a tube) for the heating and/or evaporation of a cooler liquid (such as seawater) so that the heat is recovered and productively reused. This is further discussed later in this section.

Vapor Transport. Vapor produced during vaporization must be transported to a place where condensation can take place. The boiling during the vaporization process can be violent, and salt droplets are often entrained in the vapor produced. These droplets must be removed to keep the salt content of the condensate low.

To promote vapor production the vapor produced must be removed rapidly, thereby reducing the vapor pressure above the boiling liquid.

Careful design is necessary to ensure sufficient space for vaporization to occur and to prevent salt carryover. Baffles or wire mesh are often used to remove entrained salt. The vapor itself is usually transported by pressure or temperature differentials or by mechanical means.

Condensation. Condensation generally takes place on the inside or outside of tubes made from material which is capable of a high degree of heat transfer. The heat of condensation, which must be removed from the water vapor for condensation to occur, is used either to heat or to vaporize saline water.

For efficient condensation to occur, the surface must rapidly remove the heat and allow the condensed liquid to flow to a collection point. If condensed liquid remains as a film on a surface, it acts as an insulator and reduces performance.

Heat Balance. The heat input into a distillation unit must be balanced by the heat output of the unit. The heat output includes (1) radiation and general heat loss (usually minor), (2) the distillate, (3) the brine, and (4) the cooling water. In general, this balance results in the temperature of the distillate and brine streams being elevated above the ambient feedwater temperature.

Methods of Vapor Production

In order to promote boiling, the distillation process takes advantage of the relationship between the boiling point and the applied pressure on the liquid in question. This relationship is shown on Figure 3-1. By successively reducing the pressure of a saline solution it can be made to boil successively many times without adding additional heat. This technique is used to one degree or another in almost all commercial distillation processes since reducing the pressure is less costly than adding heat.

There are three main methods that have been used for promoting vaporization in distillation units in the past 25 years: These are:

Submerged tube.

Flash.

Thin film.

Submerged Tube. Water is brought to the boiling point by the addition of heat in tubes which are submerged in a pool of water. This is illustrated on Figure 3-2. Various configurations that have been used include helical, curved, and straight tube bundles, with steam being condensed on the inside of the tubes to supply the heat.

When submerged in saline solutions, these tubes are subject to the formation of scale on the outside of the tubes, since certain compounds such as calcium carbonate and calcium sulfate precipitate out of solution at the high temperatures found at these tube surfaces. This scale can severely reduce the heat transfer of the tubes.

The submerged tube design was very prevalent in distillation units built in the 1950's but is now seldom used on large-scale plants. It is frequently used, however, in small single-stage units utilizing low-temperature waste heat as an energy source.

Flash. Vigorous boiling can be promoted by introducing water into a chamber through an orifice, thereby reducing the water's pressure below that of the equilibrium vapor pressure required for boiling at the water's temperature. This causes the water to immediately begin to boil vigorously when introduced into the chamber. This agitation and the manner in which the water is discharged from the orifice increases the surface area of the water exposed and therefore enhances vapor production.

This method, shown on Figure 3-3, is the method used in the majority of plants built in the past 20 years. One advantage to flash distillation is that once the flashing process begins, the saline water does not come in contact with hotter heat transfer surfaces and hence the chance for heat-induced scaling to occur is minimized. However, the flashing of brine is a violent process, with the brine explosively boiling when it enters the flash chamber. The turbulence caused by the boiling and the impact of the sprayed hot brine can cause considerable damage to improperly selected or installed materials in the chamber.

Thin Film. In this process saline water is applied as a film on the inside or outside of tubes which are being heated by the condensation of vapor on the opposite side. The heat is rapidly transferred through the tube walls and into the thin film of water. The thin film of water on the tubes permits rapid heating while exposing a large surface area for the promotion of vaporization.

Figure 3-4 illustrates the concept of thin film, and Figure 3-5 shows three methods used to apply films to both vertical and horizontal tubes. These methods include falling film, spray film, and climbing film. The film is usually applied on the inside of vertical tubes and the outside of horizontal tubes.

One drawback to thin film vaporization is that heat-induced scale can be formed on the surfaces of the tubes, since the heat transfer surfaces are hotter than the applied saline solution.

Major Distillation Processes

There are three major distillation processes being used in the industry today:

Multiple-effect evaporation.

Multistage flash.

Vapor compression.

In addition to these, there are waste heat recovery evaporators, which utilize principles of multiple-effect or flash distillation, and some hybrids which combine two or more of the above processes.

Multiple-Effect Evaporators. Conceptual diagrams and photographs of several types of multiple-effect evaporators are shown on Figures 3-6 through 3-13. In each effect steam is condensed on one side of a tube and the heat of condensation derived from this is utilized to evaporate saline water on the other side of the tube wall. Thus, the heat of vaporization imparted to the water to produce the initial vaporization is efficiently reused through the subsequent exchange of the heats of condensation and vaporization in later effects.

The subsequent use and reuse of the heats of vaporization and condensation to promote boiling is accomplished by reducing the pressure in each of the effects relative to the one preceding it. This allows the brine to boil at lower and lower temperatures as it flows through the plant. Because of this, the multiple-effect plant has a steam economy which is proportional to the number of effects. If the plant were perfect in insulation, heat transfer surface, etc., the steam economy in each effect would be close to 1. "Steam economy" is defined as the number of pounds [kg] of water produced for each pound [kg] of steam utilized. However, with general commercial units the steam economy is about 0.7 to 0.85 lb of water per effect for every pound [or kg/kg] of external steam introduced into the plant.

Multiple-effect distillation has been widely used in producing boiler feedwater for land-based power plants and ocean-going vessels, as well as in the sugar and salt industries since the 19th century. It was adapted to the potable water industry with few initial changes. The early designs, although only modestly efficient for the production of low-cost water for potable purposes, were proven by previous industrial experience and helped the industry begin.

i) Submerged Tube--The early multiple-effect potable water plants usually contained two to six effects, and the water was heated using submerged tubes. Figure 3-6 shows a submerged-tube configuration. One of the last major municipal multiple-effect submerged-tube distillation plants was built in 1958 for the public utilities system on the Island of Aruba in the Netherlands Antilles. It was a 2.5-mgd [9,460-m^3/d] facility consisting of five 0.5-mgd [1,890-m^3/d] units, each having six effects. Horizontal tubes submerged in the brine were used to heat the brine by condensation of the vapor produced in the preceding effect inside the submerged tubes. These units were operated for 22 years before being taken out of service in 1980.

The greatest problems with the submerged-tube units are that (1) the brine pool cannot be vaporized as efficiently as in other configurations because of the smaller relative surface area exposed, and (2) scale often forms on the hot submerged tubes and produces a coating which reduces the thermal efficiency. Thus the overall efficiency is low, and scale has to be removed frequently from the tubes. Scale was removed from the tubes in early plants by chipping the scale away with hammers. Submerged-tube plants utilizing waste heat for industrial, marine, and other special installations are still being manufactured and installed. These are basically one-stage units which run at temperatures in the range of 120-140°F [48.9-60°C] to minimize scaling. The thermal efficiencies of the plants are low but, since the energy cost is minimal, this is usually not critical.

ii) Vertical Tube--The vertical-tube evaporator (VTE) configuration was intended to resolve some of the problems of the submerged-tube configuration. The configuration is shown on Figure 3-7. Vertical tubes are suspended above a brine pool, and brine is allowed to flow on the inside of the tubes (tube side) while vapor is condensed on the outside (shell side). The heat of condensation is conducted through the tube wall and used to heat the water flowing on the inside. By reducing the atmospheric pressure inside the effect to the proper level, the liquid boils as it flows on the inside of the tubes. In general, the brine flow in the tubes is downward, although upward flow units are also used (see Figure 3-5).

A typical flow sheet and photograph of a vertical-tube multiple-effect plant are shown on Figures 3-10 and 3-11. These plants are forward-feed units, meaning that the feedwater enters the plant for distillation purposes at the high-temperature end of the plant. By convention, the highest-temperature effect (or stage in a multistage plant) is given the number 1, and the rest of the stages are numbered sequentially.

Units in which the feedwater begins to be distilled at the low-temperature end of the plant and moves through the units counter to the flow of the condensate and against the vapor temperature gradient are said to have a "backward-feed unit." Almost all potable water distillation plants have a forward feed whereas distillers (or evaporators) in the sugar industry have backward feed.

In the development of the present-day multiple-effect plant (both vertical- and horizontal-tube) a feed heater section is often added to each effect. This utilizes a portion of the heat recovered from vapor condensation to heat the feedwater to some degree before it arrives at the head of the plant, thus reducing the amount of external steam required to heat the feed to its boiling point. To avoid clutter, these feed heaters were not shown in the figures.

External steam must be supplied to the first effect to add sufficient heat to permit boiling and evaporation to take place in that effect. This external steam represents the major energy input to the process. This steam is generally condensed and returned to the boiler for reheating and recycling.

Much of the early research conducted on vertical-tube units was performed on the Office of Saline Water (OSW) Demonstration Plant No. 1 located in Freeport, Texas. One of the earliest large municipal seawater plants using this technology was the 1-mgd [3,790-m^3/d] vertical-tube multiple-effect plant installed on the island of St. Croix (U.S. Virgin Islands) in 1968. This plant was still operating in 1980, although at that time it was producing at a reduced capacity and had undergone considerable repairs and alterations during its 12 years of operation.

Compared to the submerged-tube configuration, the vertical-tube units have the potential for increased thermal efficiency and reduced scaling.

The vertical-tube plants are more complex and usually require considerably more external piping and pumps than the submerged-tube facilities. Pumps and associated piping are usually necessary to pump brine between effects. For efficient operation and the minimizing of scale formation on the tube surfaces, it is necessary that the brine feed be continuous and that the brine be evenly distributed during operation among all the tubes.

iii) Horizontal Tube (HTME)--Conceptual views and photos of the horizontal-tube multiple-effect configuration are shown on Figures 3-8, 3-9, 3-12, and 3-13. Although the principle of operation is the same as for the vertical-tube evaporator, the brine and steam are applied on opposite sides of the tubes in the two systems.

The brine is distributed as a film on the outside of the tubes, where it is partially evaporated by heat derived from condensation of vapor (to freshwater) on the inside of the tubes.

The effects are amenable to being stacked both vertically and horizontally rather than just horizonally as is necessary with the vertical-tube units. The configuration of this vertical arrangement permits greater use of gravity to move liquid between effects than in the vertical-tube units. Figure 3-12 shows a simplified flow diagram of a stacked horizontal-tube multiple-effect plant. The plants can also be constructed horizontally, and three such plants of 1.25 mgd [4,730 m^3/d] are currently being constructed in the U.S. Virgin Islands. These plants, produced by Israel Desalination Engineering (IDE), should be operational by about 1982.

Limited operational experience with horizontal-tube units in Belgium indicates that scale formation and removal is significantly less problematic in the horizontal-tube than in the vertical-tube units.

Multistage Flash. Multistage flash distillation units utilize the following phenomenon: when the pressure of the vapor above heated water is suddenly reduced below the boiling vapor pressure at that temperature, the water violently boils. This boiling occurs rapidly, releasing significant quantities of vapor until the temperature falls (due to the loss of the heat of vaporization) below the boiling point for that pressure condition.

The vapor produced by flashing is condensed on the outside of tubes which are conveying cooler brine to the hot end of the plant. This serves to recover much of the heat of vaporization. Figure 3-14 shows a conceptual diagram of the multistage flash process.

i) Single-Stage--A number of single-stage flash plants were built during the 1960's and 1970's. These were mainly used on brackish water in conjunction with power plant installations and generally were designed for less than 0.250 mgd [946 m^3/d]. The thermal efficiency of these plants is rather low, and hence they are seldom used for potable water production except for designs which utilize waste heat as an energy source.

ii) Multistage--The real potential for the use of flash distillation for potable water production developed in the 1950's with some basic design work and patents by R. S. Silver of Scotland. He differentiated the multistage flash (MSF) process plants from those plants using the flash principle with several stages by defining in the patent the multistage flash process as one used in a plant employing flash distillation in which the number of stages is more than twice the performance factor using the English units of measure, i.e., pounds per 1,000 Btu's (Silver, 1971). Hence, a plant with a performance factor of 8 had to have more than 16 stages; a plant with a factor of 10, more than 20 stages, etc.

In the late 1950's and early 1960's, the MSF plant was the only type which could be operated relatively scale-free for extended periods. At that time scale formation and prevention was not fully understood. Since multistage flash plants could be built in large capacities, they became very popular, and in the 1977 Office of Water Research and Technology (OWRT) survey they accounted for about 65 percent of the world's desalination capacity (El-Ramly and Congdon, 1977).

Figures 3-15 and 3-16 show a simplified flow sheet and a photograph of a multistage flash facility. The incoming seawater first passes through the heat rejection stage(s) and then is heated in the heat recovery sections in each subsequent stage. After passing through the last heat recovery section and before entering the first stage for flashing, the feedwater is further heated in the brine heater using externally supplied steam. This raises the feedwater to its top temperature, after which it is passed through the various stages where flashing takes place. The vapor pressure in each of the stages is controlled so that the heated brine enters each chamber at the proper temperature and pressure combination so that instantaneous and violent boiling occurs. Various types of orifices and baffles are used between stages to produce this pressure drop, to promote surface area exposure (to maximize vapor production), and to prevent brine droplets from being thrown into the demisters. In each stage the vapor passes through demisters and is condensed in the heat recovery section of the stage on the outside of tubes carrying the saline feedwater to the brine heater at the head of the plant. The cooler feedwater is heated by the transfer of the heat of condensation of the vapor.

The freshwater formed by condensation of the vapor is collected in each stage and is passed on from stage to stage in parallel with the brine. In each stage the product is also flashed so that it can be cooled and the heat recovered for heating the feedwater.

The percentage of brine actually vaporized in each stage is small, because the energy required to furnish the heat of vaporization is derived from the temperature drop of the brine in each stage (since, after the brine heater, no additional external heat is supplied to the process). A typical temperature drop in a stage of an MSF plant is about 5°F [2.8°C]. Under ideal conditions, only about half of one percent of the water flowing through a stage could be vaporized. Thus, the MSF process is characterized by high flows through the plant relative to the amount of water produced. An average high-temperature MSF plant recovers as freshwater only about 25-50 percent of the flow through the plant. This contrasts with a well-designed multiple-effect plant, which recovers from 40-65 percent of the feed. However, when the water required for rejecting heat from the plants is included, the total quantities of saline water used by the multistage or multiple-effect plants are nearly the same.

The number of stages in an MSF plant varies depending on the application, efficiency desired, temperature of the plant, etc. The number usually ranges from about 20 to 50. The number of stages is generally increased to improve the efficiency of recovering the heat used during flashing in each stage. Unfortunately, an increase in stages to improve efficiency also means an increase in the overall capital cost of the plant. By maximizing heat recovery, the amount of external heat required for the brine heater is reduced. The stages are usually designed for equal temperature drops (rather than pressure), as this results in about the same amount of tubing (heat transfer surface) and vapor production per stage, thus standardizing design and construction.

Due to the large amount of flashing brine required in an MSF plant, a portion (50-75 percent) of the brine from the last stage is often mixed with incoming feedwater and recirculated through the heat recovery sections to the brine heater and is then flashed again through all the stages. A plant of this type is often referred to as a "brine recycle," "recycle," or "recirculation" plant. This mode of operation reduces the amount of water-conditioning chemicals that must be added; this can significantly affect operation costs. On the other hand, it increases the salinity of the brine at the hot end of the plant, thereby increasing the boiling point rise and increasing the danger of corrosion and scaling in the plant. A plant which does not recirculate a portion of the brine is referred to as a "once-through" plant.

Once-through plants require a greater amount of water treatment chemicals than the recirculating plants. However, their operation, especially in startup, is considerably easier, because balancing the flows through the stages is not as difficult as it can be with a recycle plant. In addition, since the salinity is lower, there are potentially fewer problems with scaling. Due to its operational stability, this plant has considerable merit for use in areas where operation and maintenance may be a problem. This attribute, along with the newer, more effective, scale control treatments, may make this type of plant very desirable. A flow sheet for a once-through plant is shown on Figure 3-17.

The majority of the plants constructed in the past 10 years have been recirculation plants, although the advantages of the once-through plants may begin to reverse this trend.

In a modern MSF plant the stages are most often built contiguously in a horizontal arrangement. Adjacent stages generally are separated by a common wall, and 10 to 40 or more stages may be built in the same outer shell. The outer shells can be rectangular or circular in cross section depending on the manufacturer's preference. Flow is normally transferred by pressure difference between adjacent stages.

There are two basic configurations for the tubes in the heat recovery and heat rejection sections: long tubes and cross tubes. These configurations are illustrated on Figures 3-18 and 3-19, respectively. The long tubes run parallel to the flow direction of the flashing brine, whereas the cross tubes are oriented at right angles to the flashing brine flow. Each configuration has its advantages and disadvantages, and both are currently in use. The selection often depends on the preference and/or experience of the manufacturer or owner and whether the purchaser places a limit on the diameter of the tubes used.

The advantages of MSF plants lie in their ability to be constructed in large capacities; the fact that boiling does not take place on a hot tube surface (it flashes instead); the considerable design and operational experience accumulated over the past 20 years; and the widespread manufacture of the units throughout the world.

Its disadvantages, most of which are discussed later in "Engineering Considerations," involve the potential for scaling and corrosion (which has been alleviated somewhat by new anti-scale chemicals); the difficulty involved in startup; the inability to operate the plant below 70-80 percent of design capacity; the sensitivity of product water to salt contamination by pinpoint leaks in the heat recovery (condensing) tubes; and the necessity for pumping, treating, and heating large quantities of saltwater relative to the product.

Vapor Compression. The vapor compression process (also referred to as "vapor recompression") differs from the other distillation processes discussed thus far in this chapter in that it does not utilize an external heat source, such as steam, as its primary energy for distillation. It relies instead on compressing water vapor (by a compressor or steam jet) to increase the vapor's pressure and condensation temperature. The plant is designed so that the condensation of the vapor takes place on one side of a tube which acts as a heat transfer surface. Feed brine is applied to the other side of the tube (generally in the form of a thin film), and the heat of condensation that is released at the tube surface on one side is utilized as the heat of vaporization on the other side to boil the brine and produce water vapor. The principle of operation is illustrated on Figure 3-20.

Because the heats of condensation and vaporization are recycled across the heat transfer surface, the only major energy required during operation is that necessary to drive the compressor. As shown on Figure 3-20, the compressor serves a dual purpose: it compresses the vapor, raising its condensation temperature, and at the same time it lowers the pressure on the feedwater brine and hence reduces its boiling temperature.

i) Methods of Compression--There are two basic methods used to compress the water vapor. The first is to use a compressor which can be powered by any rotating source such as an electric motor, diesel engine, steam turbine, etc. Normally, either electric or diesel power is used. The second method is to use a steam ejector. This method is often considered feasible where a quantity of surplus waste steam exists; otherwise, producing steam solely for the purpose of operating an ejector would not be economical.

Both methods of compression produce the same result--potable water. The steam ejector plant is easy to install and start up, and requires considerably less care and attention than a plant using a compressor. However, it is not as thermally efficient in producing water as a compressor-operated plant. The performance factor (pounds of water per 1,000 Btu's of energy input) [kilograms of water per megajoule of energy input] of a plant with a steam ejector ranges from 2 to 6 lb/1,000 Btu's [0.86-2.57 kg/MJ], whereas one using a mechanical compressor is about 20 lb/ 1,000 Btu's [8.6-kg/MJ]. In spite of their relative inefficiency, steam ejector compression plants are widely used at construction sites because of their high reliability, low manpower requirements, and low attention requirement.

Plants using a compressor system have more moving parts than steam ejector plants and are therefore more susceptible to mechanical failure. The most vulnerable part is the compressor, which generally runs at high speeds at close tolerances, and is prone to corrosion, salt build-up, etc., on its internal parts, leading to unbalanced operation which, unless promptly corrected, can cause severe problems. Its operation also requires more attention than a steam jet unit.

Steam jets are often used in the chemical process industry and on off-shore oil rigs, where dependability is more critical than water cost.

ii) Configurations--Two configurations widely employed in vapor compression units are vertical-tube and spray film (horizontal-tube) types. These are illustrated on Figures 3-20 and 3-21, respectively. Photographs of three other vapor compression units are shown on Figure 3-22. These include an upflow vertical tube, a steam jet (thermocompression), and a low-temperature (135°F [57°C]) horizontal-tube unit.

Waste Heat Recovery Evaporator. This is a class of specialized distillation units which operate at relatively low temperatures. These are normally single-stage or single-effect units which use waste heat to heat the brine feed. Figure 3-23 shows one type of unit which is commercially available.

The waste heat can be obtained from hot process water, diesel engine jacket water, turbine exhaust, surplus steam, etc. The heat is usually transferred to the feedwater by submerged tubes. Low pressure is maintained in the evaporator so that the heated seawater feed boils at a low temperature (120°F-150°F [48.9-65.6°C]), thus minimizing feed treatment chemical consumption as well as the potential for scale formation.

i) Performance Factor--The performance factor is generally between 0.9 and 1.0 lb/1,000 Btu's [0.39 and 0.43 kg/MJ], which, compared to the multiple-effect, multistage, or vapor compression units, is rather low. However, the energy cost is very low, with the heat portion considered to be at no cost and the electrical power consumption averaging about 6 kWh/kgal [1.6 kWh/m^3].

ii) Simplicity--One of the major advantages of the waste heat recovery evaporators is their simplicity of operation and maintenance. The only significant moving parts are standard pumps which circulate the feedwater, cooling water, and distillate, and operate an eductor which reduces the pressure in the evaporator. Chemicals are generally not employed. In case of malfunctions the unit can usually be repaired by site maintenance crews, because there are no compressors or other close-tolerance equipment involved.

iii) Requirements--This unit requires three conditions for successful operation: (1) a continuous source of heat, (2) a source of electricity, and (3) an unrestricted source of seawater for feed and cooling purposes. The production of the unit is directly dependent on the amount of heat available and the temperature of the cooling water. Production increases as more heat and cooler water are available. Some installations have a cooling water requirement of 30 to 60 times the amount of product water produced.

iv) Applications--These units are often installed on ships, barges, offshore drilling rigs, etc. They could be installed to operate in a land-based facility where waste heat is available from some industrial operation or from the exhaust or cooling water from a diesel or gas generator operated almost continuously.

A unique waste heat recovery evaporator was tested by Alcoa for 3 years in Puerto Rico. This was a falling film vertical-tube evaporator in which the tubes and much of the other material were made from aluminum alloys. The source of energy for this installation was waste heat from a refinery.

Another unit apparently suitable for using waste heat for distillation is the Gadelius Nord-Aqua System. This system is a multistage flash plant which uses a siphonic principle to reduce the pressure through the stages. A 2,600-gpd [9.8-m^3/d] test plant in Japan operated on a temperature differential between the waste heat and the cooling water of about 13°F [7.2°C]. (Gadelius, 1977).

Hybrid Systems. A number of hybrid systems have been built or proposed which contain a combination of the basic units previously described.

i) Vertical-Tube Evaporator Multistage Flash (VTE-MSF)--For this unit, a multistage flash process with its heat recovery sections is used as a feedheater to efficiently increase the feedwater temperature. After passing through the recovery section in the highest-temperature stage, the feed goes to a series of vertical-tube evaporators. The product water is flashed through the MSF unit to recover its sensible heat. In addition, vapor from each effect passes to the appropriate (same-temperature) MSF stage to help heat the feed.

Two 2.25-mgd [8,520-m^3/d] plants of this type were built in the U.S. Virgin Islands in the mid-1970's. The plants had design performance factors of 12.2 lb water/1,000 Btu [5.23 kg/MJ]. There have been a number of problems during operation of these plants which are apparently related to mechanical details and materials of construction rather than the overall concept.

ii) Vapor Compression-Multiple Effect--Some compressors for large-capacity, single-effect vapor compression plants are very large and expensive. The capacity, and therefore cost, can be reduced in direct proportion to the number of series effects served by the compressor. However, the compression ratio also increases in direct proportion to the number of effects, and this increases the compressor cost somewhat. The heat energy developed by the compressor is sufficient to make up all heat losses and to keep the plant operating continuously. The Roswell, New Mexico, facility described later in the section "Process Experience" was a vapor compression multiple-effect unit.

iii) Horizontal-Tube Multiple-Effect Topping Unit--In order to make existing multistage flash (MSF) plants more efficient, it has been suggested (Cox, 1974) that a horizontal tube multiple-effect unit, which is available commercially, be placed before the hot end of existing MSF plants. The MSF plant would then be operated at lower than the design temperature, and vapor from the last effect of the HTME unit would provide the heat in the brine heater for the MSF plant.

iv) Multistage Flash--Vapor Compression--It has been proposed that a multistage flash unit be placed in series with a vapor compression unit (Tleimat, 1969). The multistage flash unit would function as a heat recovery process more efficiently than the liquid to liquid heat exchanger, which is normally utilized with vapor compression.

Others. There are a variety of combinations of distillation plants which have been discussed or tried on a small scale but none which have enjoyed great commercial success. Except in very large plants or for special conditions (such as solar power), increased energy efficiency is frequently offset by the difficulties caused by the increased complexity of operation and/or design.

COMMERCIAL DEVELOPMENT

History

Early processes for controlled evaporation of liquids with a high total dissolved solids content involved the production of salt and sugar. Solar evaporation was and still is used for the production of salt.

Both horizontal- and vertical-tube multiple-effect evaporators have been used since the beginning of this century in the production of sugar. Sugar is a high value product, and the technology was developed early. The process is employed to remove water which is derived from sugarcane itself as well as from water added during the process of grinding the cane.

The early applications of distillation for potable water came in ocean-going sailing ships. Seawater was distilled in small quantities on the ship's cook stove using a very basic still consisting of a pot, a condenser, and a collector. These were inefficient and inconvenient but augmented water carried in casks or collected during rains. As steamships were introduced, travel times were reduced and it was often more efficient to carry freshwater in tanks between ports than to produce water onboard. However, some marine stills of the single- and multiple-effect variety using submerged tubes were introduced in the late 1800's and appeared for the next few decades. In certain water-short areas these were adapted for use ashore. In 1915 in the Danish West Indies (now the U.S. Virgin Islands), a Danish physician, Dr. Viggo Christensen, considered the use of a shipboard distiller for the Town of Charlotte Amalie. He estimated that a 0.05-mgd

[189-m^3/d] marine unit would cost approximately $250,000 and would require 2,000 lb [909 kg] of coal to produce 9,000 gallons [34.1 m^3] of water, giving it a performance ratio of about 0.3 lb/1,000 Btu's [0.13 kg/MJ] (Christensen, 1915). Although potable water for ships refueling on St. Thomas sold for up to $4 to $5/kgal [1.06 to 1.32 $/$m^3$] during the next decade, desalination units were not installed on St. Thomas until the 1960's (Buros, 1978).

Submerged Tube. Curacao, another Caribbean island, installed a land-based submerged-tube desalination unit for potable water production around 1928. Others were built scattered around the world where high water cost was acceptable; however, the concept, due to economics, was not widespread and potable water desalination capacity installed by the 1940's was insignificant. These plants were, for the most part, submerged-tube single- or multiple-effect units operating at lower temperatures under vacuum and were built with high-quality materials. They had low performance factors, due partly to continuing shutdowns associated with removing scale from the tubes. The last major submerged-tube multiple-effect evaporator unit was installed in Aruba in 1958. During the Second World War considerable work was done in the USA on single-effect vapor compression units which could be used on-board ships as well as in isolated bases around the world. These units operated at higher temperatures, 212 to 220°F [100 to 104.4°C], to reduce their overall size and, consequently, had more scaling problems than the lower-temperature, submerged-tube, multiple-effect plants (Watson, 1976).

Flash Evaporation. The concept of flash evaporation was known from the early part of this century, and units with one to four stages were used for brackish water and seawater desalination from the 1940's into the 1970's both in the United States and elsewhere. In 1956, a four-stage, 2.5-mgd [9,460-m^3/d] flash plant was built by Westinghouse to supply potable water for Kuwait. This plant had a performance ratio of about 3.3 lb/1,000 Btu's [1.42 kg/MJ] and represented a milestone in the development of multimillion-gallon plants (Watson, 1976).

Multistage Flash. In 1957 the multistage flash concept was patented by Robert Silver of Glasgow. The first two significant plants of this type were built in Kuwait and on the Channel Island of Guernsey. The Kuwait plant had two units of 1.2 mgd [4,540 m^3/d], each with 19 stages and a performance factor of 5.7 lb/1,000 Btu's [2.45 kg/MJ]. The Guernsey installation of 0.75 mgd [2,840 m^3/d] had 40 stages and a performance factor of over 10 lb/1,000 Btu's [4.29 kg/MJ] (Silver, 1971). From this beginning, the multistage flash concept dominated the field of seawater distillation, and multistage flash plants presently make up about 85 percent of the world's installed distillation capacity (El-Ramly and Congdon, 1977).

Saline Water Act of 1952. The U.S. Congress passed the Saline Water Act of 1952 with the idea of accelerating the use of saline water conversion for potable and industrial purposes. The Office of Saline Water (OSW) was established and funneled millions of Federal dollars into research and development projects in the various desalination processes.

Many of the units, such as the Freeport, Roswell, Point Loma, Wrightsville Beach, and San Diego distillation plants, were OSW-funded projects undertaken to gain information on design and operation. In the field of distillation, there was a drive to reduce water production costs by increasing thermal efficiency and utilizing less expensive materials of construction.

Scale Control. The major problem in operating seawater distillation plants is the formation of scale caused by some of the dissolved constituents within the seawater. As temperatures and concentrations are increased, these constituents, such as calcium carbonate, magnesium hydroxide, and calcium sulfate, become less soluble and deposit scale which creates problems, the most serious of which are the reduction in heat transfer ability within the plant (which can severely reduce capacity and efficiency) and the plugging of tubes (which can render the plant inoperable).

Scale formation is promoted by the high temperatures and concentrations used to obtain higher efficiencies in many plants. Thus, the historical development of distillation has largely centered around methods of scale control.

In the submerged-tube plants, scale formed on the outside surfaces of the tubes was removed by either shutting down the units and cold shocking them or chipping the scale off by hand. Polyphosphate-based chemicals have been used since about 1950 to inhibit scaling but, although these are generally effective, they limit the upper operating temperatures to about 180-190°F [82.2-87.8°C] due to thermal decomposition of the polyphosphate above that point. This temperature limitation in turn limits the plant's thermal efficiency. A major advance was declared with the use of acid (generally sulfuric acid), which was added to the feedwater to break down the bicarbonate in seawater to prevent alkaline scaling. This allowed operating temperatures of up to about 235-250°F [112.8-121.1°C] before sulfate scaling (which cannot be inhibited by known additives) could occur. One of the earliest plants in which acid was used commercially was the 2.5-mgd [9,460-m^3/d] submerged-tube multiple-effect plant built in Aruba in 1958 (Watson, 1976).

Subsequently, the OSW tested acid scale control in several of its demonstration plants. On the basis of its testing, especially with the Point Loma multistage flash unit, in 1963 the OSW encouraged the use of acid feed in future plants.

Although the acid plants have been able to achieve high thermal efficiencies, with performance ratios of 10-15 lb/1,000 Btu's [4.29-6.44 kg/MJ], they require careful control and operation. Acid accelerates corrosion of materials, especially some of the less expensive materials such as steel, which were utilized in an effort to reduce material costs. The acid corrosion is not necessarily uniform in nature nor predictable in extent at the design phase.

Nevertheless, multistage flash plants using acid became a standard offering in the industry. Their high theoretical efficiency and often lower capital cost (due to the reduced heat transfer surface needed) made them extremely competitive in the world market. Acid was also used in multiple-effect and other configurations in order to allow higher operating temperatures.

The result of the extensive use of acid plants has been that a significant number (precise number is undocumented) of these plants suffered severe damage due to the use of acid combined with poor operation, design, and/or materials selected. Nonetheless, there are acid plants which have operated and are operating in top condition.

In the mid-1970's high-temperature additives (polyelectrolytes) were introduced into the market. These are meant to substitute for acid and supposedly will inhibit scale formation at high operating temperatures without causing the corrosion problems associated with acid. Operating experience is now being accumulated on their use. The initial results appear to be quite favorable.

Middle East Market. Since the 1970's the Middle East has dominated the market with respect to sales of distillation units. Over half of the world's desalination capacity is located in these countries. Whereas in the 1960's a 1- to 3-mgd [3,790- to 11,400-m^3/d] plant was considered large, the large installations built or planned in the Middle East during the late 1970's ranged from 10 to 30 mgd [37,900 to 114,000 m^3/d], with ultimate capacities of 200 mgd [757,000 m^3/d] being installed at Jubail and 50 mgd [189,000 m^3/d] at Al Khobar, both on the east coast of Saudi Arabia.

Brackish Water. Since the advent of electrodialysis in the 1950's and reverse osmosis in the 1960's, these processes have almost completely displaced distillation for the production of freshwater from brackish water. Distillation plants still being constructed are generally vapor compression units which are being used to concentrate brackish water derived from industrial cooling.

Extent of Usage

As of the latest worldwide inventory of desalination plants, conducted in 1977, distillation plants made up 77 percent of the world's capacity, compared to about 23 percent for the membrane processes (El-Ramly and Congdon, 1977). Among the three basic distillation processes, the percentage of capacity and number of plants were distributed as follows:

	Capacity			Quantity	
	(mgd)	(1,000 m^3/d)	Percent of Total	Number of Plants	Percent of Total
Flash	650	2,460	85	429	54
Multiple-Effect	95	360	13	263	32
Vapor Compression	15	56	2	115	14
Total	760	2,876	100	807	100

These data apply to plants with capacities of over 25,000 gpd. The majority of this capacity utilizes the flash process, primarily in the multistage flash configuration.

ENGINEERING CONSIDERATIONS

Seawater Chemistry

Although considerable attention in distillation is focused on thermal efficiency, it is ultimately the chemical constituents of seawater that control the process. The most significant of these controlling factors is the formation of alkaline and non-alkaline scales within the plant itself. This scale formation is associated with the operating temperature of the plant.

Dissolved gasses can also affect the rate of corrosion and the heat transfer within the units, and the concentration of dissolved solids affects the corrosion rate as well as the boiling point of the water.

Careful attention to the chemistry of the feedwater and its manipulation through the control of pH, gas concentration, etc., has a major impact on the longevity and efficiency of any distillation unit.

Major Costs

The major capital costs associated with large distillation plants themselves are heat transfer surfaces (usually tubes), the containing shell, pumps, and piping, generally in that order. Since the heat transfer surface is one of the most significant factors in plant capital cost, considerable effort has been directed toward reducing such costs.

The major costs in operation are steam (or another primary energy source for some vapor compression units), labor, chemicals, and other power (for pumping, motors, etc.), in that general order. Steam makes up at least 60 to 80 percent of the projected operating costs of moderate-sized multistage and multiple-effect plants (Fluor, 1978).

Thus the major capital and operating costs are heat transfer surfaces and steam. If one is decreased, the other must generally be increased. Each plant is generally a compromise, therefore, between the type and amount of heat transfer surface installed in the plant and the amount of steam necessary to produce a unit quantity of water.

Efficiency

Terms. The efficiency of a distillation plant is measured by the amount of water produced per unit of heat energy or steam required. This is expressed by the terms "performance ratio" and "economy ratio."

i) Performance Ratio or Factor (PR or PF)--The pounds of freshwater produced per 1,000 Btu's [kg/MJ] of heat input.

ii) Economy Ratio (also referred to as "Gained Output Ratio--G.O.R.")--The ratio of pounds [kg] of water extracted to pounds [kg] of heating steam supplied to the first effect, or to pounds [kg] of extracted water vapor from the previous effect (Kays, 1965).

$$\text{e.r.}_{(1)} = \frac{\text{lb [kg] water extracted in the first effect}}{\text{lb [kg] saturated steam fed to the first effect}}$$

$$\text{e.r.}_{(n)} = \frac{\text{lb [kg] water extracted in the } n^{\text{th}} \text{ effect}}{\text{lb [kg] vapor produced in the (n-1)th effect}}$$

$$\text{E.R.}_{(N)} = \frac{\text{lb [kg] water extracted from a plant with N effects}}{\text{lb [kg] saturated steam fed to the first effect}}$$

For vapor compression units requiring rotating power, a steam equivalent based on the Btu's [joules] required to produce the needed kilowatt hours is used in the performance factor calculations.

Of these standard measures of efficiency, the performance factor is most often used as an overall measure of an installation. For most distillation plants, however, the performance factor does not include the energy involved in pumping, and for multiple-effect as well as flash plants this can be a significant cost.

Temperature. In order to achieve high performance factors, the brine in a multiple-effect or multistage plant must be permitted to pass through a large number of effects or stages. Each effect or stage has a certain temperature loss associated with it based on the design, materials, brine concentration, etc. The difference between the highest temperature achieved at the head of the plant (usually in the brine heater or first effect) and the temperature of the final heat rejection stage is a limiting factor on the number of effects or stages. Within this temperature differential, only a certain number of effects or stages can be placed, since each effect or stage must have a discrete temperature difference to operate properly.

Since the temperature of the available feed (and cooling) water limits the lower temperature, significant efforts have been made to increase the upper temperature limit of distillation plants in order to increase the number of effects and/or stages and to improve the overall performance factors. However, these higher temperatures accelerate scaling and corrosion problems; this can adversely affect the life and operation of the plant.

Heat Transfer Surface. The stage or effect efficiency, which is directly related to the effectiveness of the heat transfer surfaces (primarily the tubes), is dependent on the design. The heat transfer efficiency depends on the type of material, amount of area, configuration of the surfaces, thickness of the tubes, brine velocity, and (when in operation) the degree to which heat transfer is hindered by scale formation and/or non-condensable gas blanketing.

The efficiency of heat transfer surfaces of the tubes can be improved by varying their shape. Two methods often used are fluting and roping of the heat transfer tubes as shown on Figure 3-24.

Fluted tubes are used in vertical-tube evaporators, in which the steam condenses on the outside and the saltwater evaporates on the inside. The fluting, which forms ridges and troughs on the tube wall parallel to the centerline of the tube, causes the condensate to flow from the ridge to the trough, leaving the ridges relatively dry and thus having a high film coefficient (Yorkshire, 1976). This improves the heat transfer.

Roped tubes are used in some multistage flash heat recovery sections in which the horizontal tubes condense steam on the outside while heating seawater feed on the inside. The tubes are formed with a helical groove. This groove causes some turbulence in the flow on the inside of the tube, breaking up the laminar flow regime of seawater adjacent to the inner tube wall. This increases the heat transfer coefficient.

Scale Formation

As mentioned previously in this chapter, scale formation is a limiting factor in the operation and efficiency of distillation plants. Scale formation in distillation processes falls into two classes: alkaline and non-alkaline. The alkaline scales of concern are calcium carbonate and magnesium hydroxide, and the major non-alkaline scales are various types of calcium sulfate.

Calcium carbonate, magnesium hydroxide, and calcium sulfate all have inverted solubility characteristics, i.e., they become less soluble as the solution temperature increases. When saline water is heated at the heat transfer surface (tubes), the temperature at the tube surface is higher than that in the bulk liquid. Since the scale compounds are less soluble at higher temperatures, if the concentration of the brine is already near saturation for some scale-forming constituent, that constituent will tend to precipitate on the wall of the tube when heated. If boiling occurs at the wall, as in a submerged-tube evaporator, brine near the point of bubble formation will be more concentrated than in the bulk liquid and will tend to form scale.

If scale is allowed to form it may have a number of detrimental effects on distillation processes. Normally, when scale forms it deposits on heat exchange surfaces, thus increasing the resistance to heat transfer. Also, it may accumulate in pipelines, orifices, and other flow passages, thereby impeding the flow of process fluids. Both the increased heat-transfer resistance and the fluid pressure drop increase the energy input necessary per pound of product water. At the San Diego, California, multistate flash (MSF) demonstration plant, scale compounds and organic sludge, which formed barely discernible scale deposits in the brine heater and condenser tubing, caused a 10 percent increase in specific fuel consumption (Mulford et al., 1965).

Alkaline Scale. Once the temperature of a saline solution rises above about 160°F [71.1°C], the bicarbonate ion tends to disintegrate, producing carbon dioxide, water, and a carbonate ion. The carbonate ions combine with calcium ions to form calcium carbonate, which is insoluble and precipitates. As heating continues, any remaining carbonate ions tend to react with water to form more carbon dioxide and hydroxyl ions. These hydroxyl ions combine with magnesium to form magnesium hydroxide. These reactions are shown below:

$$2HCO_3^{-1} \rightarrow CO_2\uparrow + H_2O + CO_3^{-2}$$

$$CO_3^{-2} + Ca^{+2} \rightarrow CaCO_3\downarrow$$

$$CO_3^{-2} + H_2O \rightarrow CO_2\uparrow + 2(OH)^{-1}$$

$$2(OH)^{-1} + Mg^{+2} \rightarrow Mg(OH)_2\downarrow$$

i) Removal--Alkaline scales can be removed by mechanical means such as brushing, hydroblasting, or thermal shocking, but the most common method is to dissolve them by reducing the pH of the feedwater with the addition of acid. The acid decomposes the alkaline scale. The use of corrosion-inhibited acid for cleaning aids in minimizing damage to plant materials.

ii) Prevention--Alkaline scale formation can be prevented in three major ways: controlling the temperature, controlling the pH, and introducing additives.

Controlling the temperature to restrict the upper operating temperature to be in the 160°F [71.1°C] range, means that, unless large amounts of heat transfer surface are used, the performance factor (thermal efficiency) will be low.

Lowering of the pH is often used to control scale. When the pH of seawater is reduced to about 4.3 the carbonates are converted to water and CO_2 (a gas) which can (and must) be vented. This elimination of the carbonate ion prevents calcium carbonate scale formation. At pH's below 7.5, hydroxyl ions are limited. This greatly reduces and/or eliminates the formation of magnesium hydroxide. Many plants are operated with a steady acid feed to reduce the pH of the brine stream from a pH of about 8 to a range of 7 to 7.5. This permits operation of distillation units with top temperatures up to about 250°F [121.1°C]. The disadvantages to acid feed include the cost of acid, the corrosion that can occur with incorrect pH, and the problems associated with handling acid. Approximately 2 to 4 lb of acid are required for each thousand gallons [239-478 g/m^3] of product water in a recirculating multistage plant. The precise control of pH is difficult and requires careful operation. Operation at the incorrect pH, due to over- or underfeeding acid, can be extremely detrimental to the materials (especially overfeeding acid) used in the plant even if it occurs for only a short period.

Two classes of additives are now in use: polyphosphates, which have been in use in distillation plants since at least 1950, and polymers, which were originally used in the 1960's in cooling water systems and then in the 1970's in various operating distillation plants.

Polyphosphates can be used up to an operating temperature of about 190°F [87.8°C], above which they tend to hydrolize and lose their effectiveness. The polyphosphates inhibit but do not prevent calcium carbonate and magnesium hydroxide scale; however, a soft type of sludge still forms which requires cleaning (generally with acid) periodically. A sludge cleaning process (Taprogge) which passes sponge balls through the tubes to remove the sludge during operation has been installed on a number of plants. The polyphosphate method has been used extensively in many large plants in the Middle East (Wade, 1979) and the Caribbean. Some plants in the Netherlands Antilles have used a type of polyphosphate treatment for several decades.

The major problems with the use of polyphosphate dosing are the cost of the chemicals and, most importantly, the restriction on the upper operating temperature of the plant. The operating temperature limits the performance factor possible and hence tends to raise the cost of water production. A potential problem with polyphosphates is the ultimate effect on receiving waters of the discharge of the phosphates with the blowdown (concentrated brine remaining after distillation) from the plant.

The use of organic polymers (polyelectrolytes) to control scale is relatively new compared to the other two methods. They have been undergoing operating tests at various plants. The polymers apparently work by inhibiting and distorting crystal formation. Good results were obtained with the use of organic polymers in a 150-day test run on a 2.25-mgd [8,520-m^3/d] multistage flash plant located in Curacao which normally used acid dosing (Stewart et al., 1979).

The outlook for the use of organic polymers in the future appears bright. Their potential for operating at high temperatures will allow high performance factors to be obtained without the associated severe corrosion and handling problems encountered in acid plants. Several large plants in the Middle East have now been designed around the use of these polymers.

The organic polymers are proprietary chemicals which can be expected to vary in effectiveness between brands and plants. Trial runs will probably be necessary to determine effectiveness and dosage required. Dosages on the order of 3 to 10 mg/l are being used in contrast to the 120 to 140 mg/l used with acid. In the USA these additives must be approved by Federal agencies for use in plants producing potable water. Many of the major additives have obtained this approval, but new compounds will need to obtain approval before usage.

iii) <u>Scale Control Cost</u>--The relative costs for each of these scale control methods at present are dependent on location, but they often result in approximately the same chemical costs. However, this could shift depending on the acceptance and competition in the polymer field.

<u>Non-Alkaline Scale</u>. Non-alkaline scale consists mainly of calcium sulfate compounds. The major scales that could crystallize are:

1. Calcium sulfate anhydrite ($CaSO_4$), generally the greatest problem.
2. Calcium sulfate demihydrate ($CaSO_4 \cdot \frac{1}{2} H_2O$), which is commonly known as "plaster of paris."
3. Calcium sulfate dihydrate ($CaSO_4 \cdot 2H_2O$), which is commonly known as "gypsum."

The formation of non-alkaline scales is dependent on the temperature and concentration of saline solutions. The theoretical scaling limits for seawater are shown on Figure 3-25.

Since recycling of brine, and hence increased concentration factors, are often used on high-performance (high-temperature) multistage flash plants, they can be subject to sulfate scaling. The most vulnerable part is the brine heater, where the highest temperatures occur. If the brine heater tubes become plugged with calcium sulfate anhydrite (the most common type), they generally require replacement.

i) Removal--Once formed, the non-alkaline scale is extremely difficult to remove. The scale is very hard and must be removed mechanically, as no known safe solvent exists that can be used in conjunction with copper alloys.

ii) Prevention--For distillation processes there are no additives or pH controls, such as those used for alkaline scale, to prevent non-alkaline scale formation. The major method of preventing scaling is to avoid operating in the concentration and temperature range where it can occur.

Other methods have been tried or suggested to minimize non-alkaline scale formation or its effect. The most experience has been gained on the use of a calcium sulfate seed slurry, which is circulated with the brine to provide preferential precipitation sites. This seeding technique was demonstrated on a long vertical-tube multiple-effect pilot plant at the Office of Saline Water's (OSW's) experiment station at Wrightsville Beach, North Carolina (Standiford and Sinck, 1961). A seed slurry system is used by the Resources Conservation Corporation (RCC) in the operation of their large vapor compression units which they manufacture for use mainly in the concentration of the blowdown water from cooling towers in which the level of calcium sulfate is very high (Kirchner, 1977). Other systems suggested have been the use of small fluidized beads (Veenman et al., 1978) and the removal of sulfate or calcium ions from the seawater feed by using ion exchange (DeMaio, 1979).

Plant Applicability. The forms of scale control discussed above are primarily applicable to multistage flash plants. For horizontal-tube, multiple-effect, vapor compression, and falling film types of vertical-tube evaporators, additives do not appear to be as effective as acid, because the turbulence of the film is too low (Wade, 1979); with additives other than acid, a sludge is deposited that interferes with the very high heat transfer coefficients.

Materials of Construction

The item which most significantly influences the capital cost and eventual life of distillation plants is the selection of the construction materials. Seawater distillation plants are exposed to high temperatures, concentrated brine solutions, and often corrosive chemicals. They are generally located on seacoasts, where they are exposed to a high-humidity salt environment.

Many of the materials which were used satisfactorily in shipboard units were found not to be suitable for land-based installations, due to differences in feedwater. The feedwater of the land-based units often contained silts and sands which removed the protective coatings on brasses and bronzes internal to the plant, severely diminishing the performance and life expectancy of these installations.

Material Selection. Construction materials for each major component of a distillation plant must be carefully chosen from a variety of materials which could be used. Material selection is generally based on:

Cost
Plant life desired
Design configuration
Operating temperature
Feedwater characteristics
Material compatibility
Customer preferences
Maintenance capabilities
Availability of spare and replacement parts

The material's resistance to corrosion and its heat transfer characteristics are extremely important. Although generalities are made in this section, it is imperative in the purchase and/or design of any distillation plant that expert advice be obtained in the selection of materials. Materials must be carefully matched to the process selected, the other materials used, and the environment. Table 3-1 lists materials commonly used in the past for the various components.

The ultimate selection of materials is often heavily influenced by the overall capital cost constraints for the project. However, whatever is saved in first cost by using less expensive materials is generally lost many times over in later repairs, replacements, and lost production.

Material Characteristics. The materials listed above have the following general characteristics.

i) Titanium--Titanium has high resistance to the corrosive and erosive environment found in the brine heater and the other high-temperature portions of a plant. Titanium is more expensive than some of the Cu-Ni alloys, and this has restricted its use in plants almost exclusively to the heat transfer surfaces (tubes). Although its thermal conductivity is lower than most commonly used tube materials, its other attributes such as smooth surface, thinner wall potential, high corrosion resistance, etc., compensate to some degree for this (McCue, 1975). However, care has to be taken in the selection of other metals in the unit to avoid setting up galvanic couples.

The first titanium-tubed multistage flash plant (1 mgd) [3,790 m^3/d] was built in 1965 for the Martin Marietta bauxite refinery on the island of St. Croix. Continued operation for 15 years has required only minimal repairs. Other titanium-tubed plants have been built since that time and have provided excellent availability. This includes a 0.1-mgd [379-m^3/d] all-titanium plant operated by Bacardi in the Bahamas.

Table 3-1
MATERIALS COMMONLY USED IN MULTISTAGE FLASH PLANTS (ADAPTED FROM AQUA-CHEM, INC., 1979)

Component	Material
Evaporator tubes	Titanium Copper-nickel Aluminum brass
Evaporator tube sheets	Copper-nickel Aluminum brass Carbon steel--cladded
Evaporator shell	Copper-nickel Carbon steel--lined with S.S. Carbon steel
Evaporator internals	Stainless steel Copper-nickel Carbon steel
Brine heater	Titanium Copper-nickel
Water boxes	Copper-nickel Carbon steel--lined with Cu-Ni Fiber glass
Interconnecting piping	Carbon steel Carbon steel--lined with Cu-Ni Fiber glass Copper-nickel Stainless steel

ii) Copper-Nickel--Copper-nickel (Cu-Ni) comes in various grades depending primarily on the relative percentage of each component in the mixture. The typical combinations contain copper with 10 percent (90/10) and 30 percent (70/30) nickel. The material resists corrosion by seawater but is susceptible to degradation by ammonia and hydrogen sulfide, which can be found in some sea wells and in water polluted by domestic wastewater or fertilizers. It is susceptible to corrosion by high-velocity brine flows. Due to the copper content, it has good anti-fouling properties against marine organisms. Copper-nickel is relatively easy to form and join to other materials (Gilbert, 1975).

Copper prices have, in the past, varied widely on world markets, and this can obviously affect the cost of copper-nickel relative to other materials.

iii) Stainless Steel--Stainless steel comes in various grades depending on the level of its chromium, nickel, carbon, and molybdenum content. It has high corrosion resistance to flowing seawater but in stagnant seawater it can experience severe pitting corrosion. This can have disastrous effects if stainless steel parts (such as pump impellers) remain submerged in brine during plant shutdowns. Special care must be taken in selecting the grade of stainless steel used in desalination applications.

iv) Carbon Steel--Carbon steel has been used for the shell and internals of many distillation plants, because it is a structurally strong, inexpensive, easily fabricated metal. It is, however, very susceptible to corrosion in the environment associated with a distillation plant, especially in the presence of hot brine and low pH's. In the 1950's and 1960's attempts were made to reduce capital costs of distillation plants by fabricating most or all components with carbon steel. Although generally the capital costs were low, maintenance was high and plant life short, especially when acid was used.

There are many places in a distillation plant in which carbon steel is acceptable. These include tankage and structural members. Steel often is coated or clad with other more resistant materials to increase its life and range of applications. The introduction of the use of high-temperature scale control chemicals (other than acid) may enhance the longevity of carbon steel in distillation plants.

v) Plastic--Usage of reinforced thermosetting resin materials (plastic) for water boxes, external low-pressure brine piping and pump cans, etc., has increased in recent years. This material, generally epoxy or polyester, is made from a variety of resins which are strengthened by reinforcement with glass fibers. These non-metallic materials are not affected by corrosion, but some can suffer from embrittlement caused by heat and/or ultraviolet rays from the sun. Although these materials can serve well in suitable applications, there can be variations in the quality of large size pipes, structures, and fittings produced since much of the production relies on hand work. Also, some thermosetting material must be carefully supported to prevent sagging.

Some thermoplastic materials such as PVC (polyvinylchloride) or polypropylene are used for low-temperature brine piping without being reinforced. Some engineers believe that the technology exists to utilize plastics more extensively in distillation fabrication. Indeed, the fabrication of almost an entire MSF plant (except the tubes) from thermosetting material (Morin and Johnson, 1979) was recently proposed. One problem with the increased use of thermoplastics is the lack of reliable data on longevity and/or potential failure areas for desalination applications. Such data will have to be built up through experience.

vi) Aluminum--Aluminum is a light, relatively inexpensive metal which is resistant to certain types of corrosion such as that caused by exposure to ammonia, hydrogen sulfide, nitric acid, etc. Although its thermal conductivity is high, its structural strength is low and it is easily eroded by high velocities. It can be formed in a variety of shapes, but requires special shop and field practices in fabrication due to its tendency to gall and the necessity to use special welding techniques (gas shielded arc) in joining materials. Aluminum is very susceptible to pitting corrosion (galvanic attack) due to heavy metals such as copper (Gilbert, 1975), and care must be exercised in selecting the proper alloys and in controlling exposure to these metals.

Several distillation plants have been, or are being, built in Israel, the U.S. Virgin Islands, and the Netherlands Antilles utilizing aluminum almost exclusively for construction. These installations should provide interesting information during the 1980's on the utility of aluminum for distillation plants.

vii) Other Materials--A variety of other materials have been suggested for use in shells, water boxes, tubes, and other internals in an effort to reduce capital costs and/or improve operation. One material which has been tested widely is concrete, which has been used in Japan and France for shell material. Japan carried out extensive research on the suitability of the material during the 1970's and concluded that concrete was suitable (Tazawa et al., 1977). However, this material has been utilized only sparingly in the industry for this type of application.

viii) Coatings, Linings and Cladding--In order to utilize less expensive materials or to combine qualities of two or more materials, coatings, linings, and claddings are used for some applications. They are most frequently used to protect carbon steel used for water boxes, shell material, and stage dividers.

Typical coatings used are epoxy- or rubber-based materials which are applied either in the factory or onsite. Although the materials themselves are generally corrosion-resistant, two problems can occur. The first is either an incomplete or a faulty coating which leaves small areas, even as small as a pin hole, unprotected. In the case of galvanic attack, the entire reaction is then concentrated on the one tiny area and the result can be a rapid localized failure. The second problem is that due to the differential coefficient of expansion, method of application, local environment, failure to exclude all moisture or to prepare the base surface adequately, or other factors, the coating sometimes does not adhere properly to the parent material. When failure occurs, large areas of the coating often come off, exposing parent metal to corrosion while the sections of coating disrupt flow through the tubes, gates, pumps, etc. Coatings can require repairs every few months; and when this is the case, they are usually not cost effective. They are extremely difficult to apply properly under the uncontrolled conditions which generally exist in the field.

Linings, if applied properly, generally are superior to coatings, but they have a higher first cost. Maintenance or replacement usually occurs at intervals of several years. Typical linings, for steel components, include copper-nickel alloys (generally 90-10 Cu-Ni), 316-L stainless steel, and Precrete (a proprietary aluminous cement which has the same coefficient of expansion as steel).

Copper-nickel sheet linings are made by cutting the thin sheet to the proper dimensions and then welding the edges together to form a loose, close-fitting, lining. The lining is then drilled full of holes (about 0.5-inch [1.3-cm] diameter) on 4-inch [10.2-cm] centers, and then, using the correct, recommended welding rod, the lining is plug welded through the holes to the steel base. The weak points in such a lining are the plug welds that unavoidably become diluted with carbon-containing steel, causing the plug welds to be brittle. Over a period of time, thermal stresses cause some of these welds to break, and then brine penetrates into the space between the lining and the steel, setting up a galvanic couple. Frequently, these weld breaks are not found until the steel section itself corrodes through. Repairs then may be fairly extensive.

316-L stainless steel linings have been very successfully used to protect the steel shells of desalting plants from corrosion, and some manufacturers now supply stainless-clad steel shells as their standard construction. But to protect existing unlined steel shells from corrosion, 316-L S.S. linings are field installed by welding. The lining is applied as strips 2 to 4 inches [5.1 to 10.2 cm] wide. Each strip is then welded to the steel, again using the electrode recommended by the welding supplier, so that the entire periphery of the strip is sealed. This is an expensive but permanent and effective method of stopping corrosion. There is, however, a precaution that must be observed. When the stainless steel lining is applied, some bare steel, either from the existing shell or from sacrificial steel plates, must be provided to prevent both chloride attack of the stainless steel and stress corrosion of the welds. The galvanic couple formed between the stainless steel and the steel will protect the stainless steel, but the steel will gradually go into solution and disappear. Therefore, the steel must be replaced at intervals. This replacement is very easy if the steel is made in the form of replaceable bars or plates that are bolted in place with stainless steel bolts. However, a tight connection between the sacrificial anode and the stainless steel to be protected is essential. The best way to obtain this is by welding.

As mentioned previously, Precrete is an aluminous cement having the same coefficient of expansion as steel. This material is sometimes used as a lining to protect steel from corrosion. However, its successful application is an art and should not be attempted by anyone inexperienced in its application. Since Precrete has no adherence to steel, it is applied over a steel mesh in much the same manner as Gunite cement is applied for swimming pools and similar structures. The wire mesh is attached to steel bars that are welded to steel studs which, in turn, are welded to the shell. Finishing-off the Precrete at corners and edges is especially critical. When misapplied, or after years of use, Precrete tends to come off in chunks which block brine gates and damage pumps and control valves. Skilled applicators can apply it to existing plants that are suffering from severe corrosion.

Cladding consists of metallurgically bonding, by the metal supplier, one metal to another. A typical example is a stainless steel lining bonded to a carbon steel water box. The first cost of clad material is more than that of coatings but less than that of field applied linings; and their use requires care in field welding, etc., but the combination appears to be permanent, to require little or no maintenance, and to be more cost effective in the long run. This type of construction is now largely standard with many manufacturers.

Corrosion. Corrosion is more than just rust. Corrosion is defined as "the deterioration of a substance [usually a metal] because of a reaction with its environment" (Bosich, 1970). There are many types of corrosion in a distillation plant and all of them cause operational or maintenance problems of one kind or another and shorten the life of the unit. Some of these types of corrosion which are important in distillation are discussed below.

i) Uniform--This type of corrosion refers to the general overall oxidation of entire metallic surfaces. This is the corrosion with which most people are familiar and is the easiest to design for, as its location and rate are predictable.

ii) Pitting--This a localized attack which results in holes in the metal. Pitting is usually associated with the chloride ion (very common in seawater) reacting with passive metals such as stainless steel. It can result in catastrophic failure. Once started, pitting is almost impossible to stop.

iii) Erosion--This is the increase in loss of a metal due to the relative movement of a fluid past a metal surface. This is quite common in distillation plants and affects the design in that it restricts both the flow through various portions of the plant and the materials which can be used. Table 3-2 shows the effect of velocity on the corrosion of metals under certain controlled conditions. As can be seen, velocity alone can have an extreme effect. Erosion removes the protective film from the metal, rendering it more vulnerable to attack. The results are frequently disastrous.

Table 3-2
CORROSION OF METALS BY AERATED SEAWATER AT AMBIENT TEMPERATURE MOVING AT DIFFERENT VELOCITIES[a]

	Typical Corrosion Rates, mg/decimeter^2/day	
Material	1 fps[b] [0.3 m/s]	27 fps[c] [8.2 m/s]
Carbon steel	34	254
Cast iron	45	270
Silicon bronze	1	343
Admiralty brass	2	170
Aluminum bronze (10% Al)	5	236
Aluminum brass	2	105
90-10 Cu-Ni(0.8% Fe)	5	99
70-30 Cu-Ni(0.05% Fe)	2	199
70-30 Cu-Ni(0.5% Fe)	(1)	39
Monel	(1)	4
Stainless steel (Type 316)	1	(1)
Titanium	0	0

Note: fps = foot per second; mg = milligrams

[a] International Nickel Co. data adapted from Bosich, 1979.
[b] Immersed in tidal current, 1 fps [0.3 m/s].
[c] Attached to immersed rotating disk, 27 fps [8.2 m/s].

Erosion corrosion can be an important factor in tubes where any blockage by scaling or partial plugging can reduce the cross section and increase the velocity. The erosion phenomenon is also important in pumps, water boxes, stage gates, tube sheets, piping, etc.

iv) Stress--Stress in a metal can accelerate general corrosion and can produce cracking (stress corrosion cracking). Stresses in desalting plants are often caused by field welding without proper stress relief. Stainless steel is susceptible to this type of corrosion, especially in unannealed field welded pipe seams.

Stress corrosion cracking is potentially very dangerous, as it can occur suddenly, without warning, and can result in the discharge of hot steam, brine, or product water.

v) Galvanic--Galvanic corrosion can occur when two metals of different composition are in contact with each other while exposed to a solution (such as brine) capable of conducting current. It can cause severe localized attack on metals, because the battery action causes one metal to act as the cathode and the other the anode. Metal is lost from the anode during the process, sometimes leading to structural failures or loss of the integrity of tubes, shells, tube sheets, etc. A copper-nickel tube rolled into a steel tube sheet can create the potential for severe attack of the tube sheet.

It is almost impossible to build a distillation plant without dissimilar metals; hence the opportunity for galvanic corrosion almost always exists. However, through good design and the careful selection of materials the effects can be minimized.

vi) Concentration Cells--These are similar to galvanic cells except that an anode and a cathode can be formed on a homogeneous piece of metal due to differences in environment between areas. Common concentration cells formed in distillation plants are due to differences in the dissolved oxygen content or metal ion concentraton. These often occur where stagnant conditions exist in the brine stream. The stagnant conditions can be small in extent and can be created by gasket surfaces, holes, scale deposits, bolt heads, corners, crevices, etc. A concentration cell can result in severe corrosion of the anodic portion of the surface.

Experience. Obviously, based on the above discussion of materials and corrosion, the design, selection, and fabrication of materials in a distillation plant are critical to its overall performance and longevity. Due to the complexities involved, it is very hard to predict how some variations in fabrication or materials will hold up under extended service. This is one reason that many purchasers of desalination plants insist that the manufacturer have prior experience in building plants of the type specified and that they have a plant of the same approximate capacity in operation for a certain period of time (usually 1 year).

These precautions are meant to minimize the risk of severe materials corrosion and/or performance problems. With each new plant the manufacturers, engineers, and owners can identify new or recurrent problems and try to rectify these in future plants. Occasionally these corrections have caused other problems, but generally they have assisted in the evolution of better design and materials selection.

Auxiliary Systems

There are several auxiliary systems which are important to the operation of a distillation plant: the deaerator, the vacuum system, and the chemical feed system.

Deaerator. The deaerator (and/or decarbonator) removes the dissolved oxygen and carbon dioxide present in all seawater. If these gasses were released inside the plant, they would interfere with heat transfer and cause corrosion. The deaerator is usually a tower containing packing over which the seawater is sprayed countercurrent to rising steam. The released gasses are corrosive. The tower is usually made of FRP (fiberglass reinforced plastic) or rubber-lined steel, the packing is PVC or polypropylene plastic, and the internals are stoneware and/or stainless steel.

Vacuum System. A vacuum system is required to remove the oxygen and carbon dioxide that were incompletely removed in the deaerator and were subsequently released in the evaporator. It is also required to remove the air that leaks into the large portion of the plant which operates below atmospheric pressure. These gasses, too, are corrosive, and the piping is titanium, FRP, or stainless steel. Steam jet ejectors have cast steel steam boxes and stainless steel throats and nozzles. Condensers are FRP or stainless steel.

Chemical Feed System. Chemical treatment is critical to plant operation. Because the cost of this metering equipment is comparatively small, it is generally made of the best materials available: FRP or stainless steel tanks and piping, Alloy 20 pump bodies, Hastalloy valves, Teflon diaphrams, and a good stock of spare parts. Because of their long life, these materials should be considered even when the chemicals are relatively non-corrosive.

Operational Problems

The most severe operational problems usually result from the purchaser obtaining a plant which was unsuited in design and/or materials for the level of operation and maintenance personnel technical capabilities, staff size, and/or budget.

As with plants in other technologies, successful operation of distillation plants requires trained personnel, excess capacity, adequate product storage, spare parts, tools, and maintenance equipment.

Distillation plants require careful operation and maintenance. Properly trained personnel must be available to perform operation and maintenance functions and to recognize problems and correct them before extensive damage is incurred. This becomes more important as higher-efficiency plants, which are generally complex in operation, are used.

Excess capacity, large storage tanks, or alternate sources of supply are necessary so that the plant can be shut down for preventive maintenance and/or repairs. A supply of spare parts and an ability to rapidly obtain them are also crucial to operational reliability. Small problems that are not corrected immediately can become major ones very rapidly in some plants. More than any other factor, preventive maintenance can keep a plant operating, minimize repairs, and cut the cost of the water produced.

The major problems that occur involve scale formation and corrosion and their effect on the plant's performance. Continual effort is required also to ensure that the many components of the facilities are properly functioning. These include intake structures, brine heaters, vacuum systems, pumps, chemical feeds, control instrumentation, etc. The failure or malfunction of any one of these can affect plant performance and lead to other failures within the system.

pH Control. In plants using acid feed, pH control is essential to prevent serious corrosion and/or scale formation. Too little acid results in the formation of scale, and too much corrodes the plant.

Scale Formation. Scale formation, which was explained earlier in this chapter, is a major operating problem. Scale formation blocks tube passages and tends to insulate heat transfer surfaces. The major operational causes for this are improper pH, improper dosage of scale control chemicals, and/or temperature excursions beyond the rated upper limits for the plant.

Scale formation can result in decreased efficiency, tube failures, and/or failure of the brine heater. These can require prolonged shutdowns to correct, with the concomitant lack of water production.

Corrosion. Corrosion in its various forms occurs whether or not the plant is operating, depending on the immediate environment, type of materials, etc. Certain types of corrosion can occur rapidly and lead to serious failures within the plant. Plants can often corrode faster when shut down than when operating.

Reliability

It is possible to build, operate, and maintain distillation plants which are reliable in operation. An example of this is a 1-mgd [3,790-m^3/d] titanium-tubed multistage flash plant built for an industrial complex in the U.S. Virgin Islands. Between 1966 and 1973 this plant had an operating factor (operating time as a percentage of total time) of 97 percent (Green et al., 1977).

Yet only 4 miles [6.4 km] away another plant, built about the same time, a 1-mgd [3,790-m^3/d] vertical-tube multiple-effect facility, had an operating factor of only about 73 percent

(Watson, 1976). This latter plant, run by a government agency, operated at only about 77 percent of its rated capacity. Differences in design, materials, and operation have combined to produce these variations. Table 3-3, which was published in 1976 (Watson) shows some typical variations in reliability recorded in various installations around the world.

The most important factors in ensuring reliability are the proper selections of a plant and materials for a given environment and application. Predicting the reliability of plants using new designs or materials is difficult and can be rather imprecise.

Operational and Installation Information

Operation of a distillation plant normally requires both steam and electricity, the former as the prime source of heat and the latter to operate the pumps, chemical feeders, etc.

Specific Fuel and Energy Consumption. Assuming an 85 percent efficiency in the boiler and utilizing a fuel with a high heat value of 19,100 Btu's/lb [44,400 kJ/kg], the approximate fuel consumption per unit of freshwater produced can be calculated for multistage flash and multiple-effect water-only plants based on the performance ratio of the particular plant (Watson, 1976).

Performance Ratio (PR)	lb [kg] fuel per lb [kg] water
4 lb/1,000 Btu's [1.72 kg/MJ]	0.0165
8 lb/1,000 Btu's [3.43 kg/MJ]	0.0083
12 lb/1,000 Btu's [5.15 kg/MJ]	0.0055
16 lb/1,000 Btu's [6.87 kg/MJ]	0.0042

Thus, within a multistage flash plant with a performance ratio of 8 lb/1,000 Btu's [3.43 kg/MJ], one pound [kg] of fuel will produce about 120 pounds [kg] of water.

For vapor compression installations, the specific fuel consumption varies according to the size of the plant, with the larger plants having the lower consumption. Vapor compression units can be diesel, electric, and/or steam powered (or combinations). Based on manufacturers literature, the energy requirements could vary as follows:

Diesel	0.0025-0.0055 lb [kg] fuel/lb [kg] water[1]
Electric	0.0112-0.026 kWh/lb [0.025-0.057 kWh/kg] water[2]

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[1] Computed from information in MECO Bulletin 117D.

[2] Computed from information in Aqua-Chem, Inc. Bulletin 750-1048.

Table 3-3
EXAMPLES OF ON-STREAM AND CAPACITY FACTORS FOR VARIOUS DISTILLATION FACILITIES
(ADAPTED FROM WATSON, 1976)

Plant Location	Process	Nominal Capacity (mgd)	Nominal Capacity [m^3/d]	Year of Startup	Years Assessed	Average Capacity Factor[a] (%)	Average On-Stream Factor[b] (%)	Average Plant Factor[c]	High or Low Temperature[d]
Key West	MSF(SP)	2.6	9,840	1967	1968-1970	85	91	77	high
					1969-1971	69	80	55	
Antigua	MSF(SP)	1.2	4,540	1970	1973-1974	77.5	60	47	high
Bermuda	MSF(SP)	0.16	606	1964	1968-1969	69	57	55	low
Guantanamo Bay, Cuba	MSF(DP)	1.5	5,680	1965	1965-1970	90	94	85	low
Aruba	MSF(DP)	0.8	3,030	1965	1968-1970	99	89	88	high
Kuwait	MSF(SP)	0.45	1,700	1969	1969-1970	100	85	85	high
Kuwait	MSF(DP)	2.4	9,080	1960	1971	103	--	--	low
St. Croix	VTME(DP)	1.0	3,790	1968	1969-1971	77	73	56	high
Curacao	MSF(DP)	3.4	12,900	1963	1969-1970	93	95	88	low

Notes: SP = Single purpose
DP = Dual purpose
VTME = Vertical-tube multiple-effect

[a] Capacity Factor = Ratio of actual production vs. rated (design) capacity.

[b] On-stream factor = ratio of days in operation to total days.

[c] Plant Factor = Capacity factor times on-stream factor. Annual plant factor is actual annual production divided by annual rated capacity.

[d] Temperature = High--up to about 250°F [121.1°C]; Low--up to about 190°F [87.8°C].

Another important consumable in the distillation process is electricity for pumping. This has been estimated by Cox (1979a) as follows:

Multistage Flash	6-10 kWh/kgal [1.6-2.6 kWh/m^3]
Multiple-Effect	4-6 kWh/kgal [1.1-1.6 kWh/m^3]
Waste Heat Recovery	15-35 kWh/kgal [4.0-9.2 kWh/m^3]

Physical and Operating Characteristics. Most of the vapor compression and the waste heat recovery evaporators are available from manufacturers as preconstructed, skid-mounted, package units. These can be shipped to the site and usually require only a foundation and hookups for feedwater, power, cooling water, product lines, and blowdown connections. Some of the physical and operating characteristics of some of these units are shown in Tables 3-4 and 3-5. The dimensions and shipping weights provide an idea of the size of the units.

The multistage flash and multiple-effect plants are larger in capacity and size than the vapor compression or waste heat recovery evaporators, and they are often custom-made for the locality and job.

Figure 3-26 shows the approximate land area required for the erection of a multistage flash distillation plant based on a number of plants constructed by a Japanese firm. The actual spacing of the units can be very site-specific, as can the definition of the battery limits of a plant.

Major Constraints

The major constraints to the various distillation processes are briefly discussed below.

Vapor Compression. Although the performance factors of these plants are higher (if a mechanical compressor is used) than those for the multiple-effect or multistage flash designs, vapor compression units are restricted in capacity to about 0.25 mgd [946 m^3/d] or less by the size of the compressors that are commercially available. Hence for larger-capacity installations, the capital cost economies of scaling up the units are not available.

Multistage Flash. The economies of scale work well on this type of design and units of about 25 mgd [94,600 m^3/d] may be built within the next 5 years. However, the performance factor is restricted to about 12 lb/1,000 Btu's [5.15 kg/MJ] due to the limited number of stages that can be installed between the hot and cold ends of the plant. Calcium sulfate scale formation sets the upper temperature boundary at about 250°F [121.1°C].

Multistage flash plants are often difficult to operate at partial capacity (below about 70 percent). Those plants employing recycling require considerable care during startup.

Table 3-4
PHYSICAL AND OPERATING CHARACTERISTICS OF VARIOUS VAPOR COMPRESSION UNITS

Flows				Power Requirements		Physical Size of the Unit	
Product Water Output (kgal/day)	Feedwater Input (kgal/day)	Waste Brine Output (kgal/day)	Compressor Motor Size (hp)	All Electric Unit (kWh)/(kgal)	Diesel Unit (gal diesel)[c]/(kgal)	Dimensions (L x W x H [in])	Dry Weight (lb)
Diesel-Driven Units[a]							
2.4	5.4	3.0	10	--	6.5	70 x 55 x 62	3,200
4.8	10.8	6.0	15	--	4.5	86 x 56 x 66	4,660
7.2	16.2	9.0	20	--	4.2	88 x 60 x 75	5,650
9.6	21.6	12.0	25	--	3.4	91 x 66 x 75	6,300
14.4	32.4	18.0	40	--	3.1	103 x 72 x 90	8,500
30	67.5	47.5	75	--	3.4	124 x 101 x 96	16,000
50.4	113.4	63.0	125	--	2.9	150 x 131 x 108	24,600
150	337.5	187.5	350	--	3.0	255 x 220 x 135	79,000
166.8	375.3	208.5	400	--	3.0	255 x 220 x 135	79,000
All Electric Units[b]							
2.4	4.8	2.4	10	220	--	96 x 60 x 71	4,200
4.8	9.6	4.8	20	140	--	126 x 60 x 71	5,100
7.2	14.4	7.2	30	118	--	126 x 60 x 71	5,300
9.6	19.2	9.6	40	110	--	126 x 72 x 77	6,900
14.4	28.8	14.4	50	95	--	172 x 72 x 78	9,400
28.8	57.6	28.8	2(50)	93	--	204 x 90 x 109	18,200

Note: Metric conversion factors are: kgal/day x 3.785 = m^3; hp x 0.7457 = kW; kWh/kgal ÷ 3.785 = kWh/m^3; in x 2.54 = cm; lb x 0.454 = kg.

[a] Data from Mechanical Equipment Company (MECO) Bulletin 117D.

[b] Data from Aqua-Chem, Inc. Bulletin 750-1048.

[c] The units l/m^3 could be used here.

Table 3-5
PHYSICAL AND OPERATING CHARACTERISTICS OF VARIOUS WASTE HEAT RECOVERY EVAPORATORS

Product Water[a] Output (kgal/day)	Condensing Water[a]		Approximate Waste Heat Requirements[a]			Electric Power[e] Required (kWh/kgal)	Physical Size of the Unit	
	Flow (gpm)	Temperature (°F)	Flow (gpm)	Temperature (°F)	Btu/hr		Dimensions (L x W x H [in])	Dry Weight (lb)
Single-Stage Submerged-Tube Units[b]								
2.4	53	85°	100	180	930,000	15	46 x 31 x 64	950
7.2	206	85°	350	180	2,800,000	7.7	69 x 43 x 67	3,650
14.4	600	85°	600	180	5,600,000	6.2	102 x 62 x 90	5,200
Single-Stage Submerged-Tube Flash Units[c]								
5	--	85°	--	170-200	1,950,000	14.2	70 x 40 x 85	2,100
7.5	--	85°	--	170-200	2,900,000	15.8	85 x 44 x 93	3,100
11	--	85°	--	170-200	4,250,000	11	109 x 60 x 111	5,100
15	--	85°	--	170-200	5,800,000	11.9	112 x 65 x 112	5,900
Single-Stage Submerged-Tube Units[d]								
3.7	--	--	--	165°	1,450,000	9.7	52 x 46 x 73	2,000
7.4	--	--	--	165°	2,200,000	6.5	75 x 51 x 78	3,750
14.8	--	--	--	165°	4,400,000	5.7	86 x 71 x 83	4,800

Note: Metric conversion factors are: kgal/day x 3.785 = m^3/d; gpm x 5.451 = m^3/d; (°F-32) x .556 = °C; Btu/hr x 1.0559 = kJ/hr; in x 2.54 = cm; lb x 0.454 = kg.

[a] Actual capacity and efficiency of all units are dependent on the amount and temperature of the waste heat stream and the cooling (condensing) water.

[b] Data from Mechanical Equipment Company, Inc., Technical Bulletin No. 130.

[c] Data from Riley-Beaird, Inc., Technical Bulletin No. 33175 M.

[d] Data from Riley-Beaird, Inc., Technical Bulletin No. 151775 HRC.

Multiple-Effect. There is considerable theoretical potential for building large plants similar in size to the multistage flash plants with performance factors as high as 17 to 20 lb/1,000 Btu's [7.3-8.6 kg/MJ]. The wide commercial use of this type of design (versus the multistage flash) in the next 5 years will be constrained by the lack of prototype units. Due to numerous problems with "new" designs in the past, most bid specifications require that the builder had designed and built a similar plant of the same size (or at least within 50 percent of the desired capacity) and that the plant had been operated satisfactorily for at least 1 year. Therefore, since the majority of plants built to date have been multistage flash, widespread construction of multiple-effect plants is likely to be delayed.

Potential

Distillation Versus Membrane Processes. There has been significant progress in the development of membrane processes for the desalting of both brackish water and seawater. Reverse osmosis and electrodialysis have virtually replaced distillation for the desalination of brackish water because they are more economical for that application.

With the installation of seawater reverse osmosis facilities in the 1- to 3-mgd [3,790- to 11,400-m^3/d] range, there is considerable discussion on whether reverse osmosis will also replace distillation for seawater treatment. As discussed later in this chapter on dual-purpose plants and in the chapter "Process Selection," there will probably always be, at least in the foreseeable future, a place for distillation in the spectrum of seawater desalination processes.

The most outstanding advantage of the distillation process is that it can utilize low-cost energy in the form of low-pressure steam from turbines or other waste heat. However, in many single-purpose installations, seawater reverse osmosis or a combination of reverse osmosis with distillation may become more economical (Sackinger, 1980).

High-Temperature Additives. The use of high-temperature polymer-based scale control chemicals as a substitute for acid offers a great potential for operating multistage flash and multiple-effect plants at high temperatures without the inherent (and often disastrous) disadvantages of acid. A body of experience is beginning to be developed on the use of these additives. Plants such as the 2.25-mgd [8,520-m^3/d] acid plant in Curacao have successfully substituted an additive during trial runs and more acceptance of additives can be anticipated in the future (Stewart et al., 1979). Their application to multiple-effect plants has not yet been tested.

If the polymer-based high-temperature additives continue to show these positive results in long-term operations, the change from acid to additives could bring some definite benefits to distillation.

Although the cost of unit production might not change, since the prices of the required additive and the acid dosage are apparently about equal, there is a great potential for savings through reduced maintenance costs and shutdowns (loss of production) and perhaps the possibility that the high-temperature plants will function for their design life.

DUAL-PURPOSE PLANTS

Definitions

A dual-purpose plant can be understood by first examining a single-purpose plant. A single-purpose plant is one in which fuel is burned to produce a single product. For example, in a power plant fuel is burned to produce electricity and generally nothing else. Likewise, many desalting plants burn fuel and produce only water. A dual-purpose plant burns fuel to produce two products: electric power and water. Lest this sound like something for nothing, or magic, a brief review of the nature of energy is in order.

Energy Concepts

The combustion of fuel produces energy in the form of heat, the usefulness of which depends upon its temperature. The ability to use heat can be compared to water in a lake on a high mountain. If the water is conducted from the lake down the mountain through a penstock or pipe to a water turbine, the turbine can generate power. On a high mountain, several turbines can be installed at successively lower altitudes so that, after the water flows through one turbine and generates power, it then flows further down the mountain through another penstock to another turbine and generates more power. In this manner, the water can continue to generate power until it reaches the bottom of the mountain. When it can fall no further, it can generate no more power. The amount of power that can be produced depends upon the quantity of water flowing and the distance it falls.

In heat energy, the number of Btu's [joules] generated can be compared to the quantity of water flowing down the mountain; the temperature at which the Btu's [joules] are generated corresponds to the height of the mountain. The greater the number of Btu's [joules] and the higher their temperature, the more work they can do. When their temperature drops to ambient (corresponding to the base of the mountain), the Btu's [joules] can no longer produce work.

Examples of Dual-Purpose Plants

A simple example can serve to illustrate the principle of the dual-purpose plant. Imagine a power plant in which fuel oil is burned at 2,000°F [1,093°C] to generate steam at 1,000°F [538°C]. The steam then flows through a turbine that drives an electric generator. As the steam passes through the turbine, both its pressure and temperature are reduced, but its quantity remains constant. When the temperature of the steam has dropped to 120°F [48.9°C], no more energy can be extracted economically from the steam. The steam is passed into a condenser, where it rejects its heat of condensation to cooling water and is converted to a liquid for return to the boiler. The boiler steam at 1,000°F [538°C] typically might have a heat content of 1,500 Btu/lb [3,489 kJ/kg], while the turbine exhaust at 120°F [48.9°C] might have 1,000 Btu/lb [2,326 kJ/kg]. Thus, 500 Btu/lb [1,163 kJ/kg] were converted into electrical power, whereas 1,000 Btu/lb at 120°F [2,326 kJ/kg at 48.9°C] were simply thrown away.

Now imagine a large office building adjacent to the power plant. In order to heat the office building, oil is burned at 2,000°F [1,093°C] to heat air to 90°F [32.1°C]. That air is then distributed throughout the building to maintain its temperature at 70°F [21°C].

In this illustration, it is possible to use the steam at 120°F [48.9°C], previously discharged by the power plant to waste, to heat the air to keep the office building warm while eliminating the consumption (and cost) of fuel to heat the building. Thus, both electric power and building heat are generated for the cost of oil for the power plant alone. (A very large percentage of the office buildings in New York City are heated by low-pressure steam obtained from nearby power plants.)

Various terms are used to describe this principle of using the same energy more than once at different temperature levels, e.g., dual-purpose, co-generation, cascading energy, total energy systems.

Heat at 120°F [48.9°C], even in large numbers of Btu's [joules], is at too low a temperature to be very useful except for heating buildings. Therefore, a modified dual-purpose system has been developed based on extracting steam at various points from a steam turbine.

A steam turbine is composed of a number of stages through which the steam flows in series. The temperature and pressure drop in steps in each stage from the high boiler temperature at the inlet to the low condensing temperature at the outlet. Steam turbines, therefore, are built with extraction points so that steam at any desired temperature can be extracted from the side of the turbine. These points are shown diagramatically in the Appendix on Figure C-16a. (It should be noted that once the turbine is built, these points cannot be changed.)

An extraction turbine is used where the quantity of steam required to generate power is large compared to the quantity of extraction or process steam, for example where the turbine generates electric power for a community. In some special cases, where a turbine can be operated at constant load and a large quantity of process steam is required, a back-pressure turbine is used. A back-pressure turbine can be considered a regular condensing turbine with the last several stages cut off so that the turbine, instead of exhausting at 120°F [48.9°C], exhausts at the required temperature of the process steam, e.g., 275°F [135°C] for a high-temperature distillation plant. A back-pressure turbine is less costly than an extraction turbine, but it is completely inflexible in operation, and all the exhaust steam must be used by the process in order for it to be condensed and returned to the boiler.

Although a dual-purpose plant generally can be installed anywhere steam is generated, it cannot be used for a plant powered by electricity brought in from a considerable distance or by wind or wave power.

Advantages and Disadvantages

Advantages. The one overriding advantage of the dual-purpose plant is that it drastically reduces the consumption of fuel compared to two separate plants generating two different products. Where electric power is generated by a steam turbine and extraction steam is used in a desalting plant, the cost of the extraction steam is reduced 60 to 80 percent (100 percent reduction or zero cost for a true waste heat evaporator) compared to that for a water-only plant (Cox, 1979a). The higher the temperature of extraction, the more the steam costs. But higher extraction temperatures allow the distillation plants either to (1) reduce capital cost, or (2) produce more water. The optimum extraction temperature is, therefore, an economic compromise between an increased cost for the extraction steam coupled with either a reduced cost for the desalting plant or an increased output.

There are practical limits to the lowest temperature at which the steam can be extracted from the turbine. As the steam temperature is reduced, its specific volume increases enormously. Consequently, at temperatures only a few degrees below the atmospheric boiling point, the volume of steam becomes so large that there is no room on the turbine to install an extraction valve that will let it properly pass through.

Some new distillation plants are being designed to use very low-pressure steam from a back-pressure turbine. To keep pressure drops within acceptable limits, enormous steam pipes are required, and the desalting plant must be located directly adjacent to the turbine exhaust. Such tight configurations can complicate plant layout, operation, and maintenance.

When steam is extracted at temperatures above the minimum turbine condensing temperature, there is some sacrifice in the quantity of power that can be produced per unit of fuel. Thus, instead of yielding "two for the price of one," the fuel produces two for the price of 1.2 or perhaps 1.4, which is still a bargain. In the case of a distillation plant that requires extraction steam at 275°F [135°C], the quantity of fuel required by the power plant is 140 percent of that for a condensing plant which has no associated distillation plant, but this is far less than the 200 percent required for two separate plants.

The reduced cost of process steam for a dual-purpose plant is actually based on a reduced consumption of fuel. Accountants are not agreed on how this cost savings should be allocated, but the method suggested by Cox (1979a) is one procedure. Another has been utilized by the Office of Saline Water (OSW) for their cost studies, but this is much more complicated. The OSW method requires that two separate, complete plants be designed and costs assigned to each. Then the two plants are combined into a dual-purpose plant and the costs for jointly used components (such as seawater intake and discharge, control rooms, piping and instrumentation systems, and operating and maintenance labor, etc.) are distributed as a function of their separate costs. Other methods of allocating joint costs, many of which are arbitrary, also are used.

Disadvantages. The principal disadvantage of dual-purpose plants is that the plants become interdependent. If the power plant does not operate, there is no steam for the desalting plant; if the desalting plant does not operate, there is no condenser for the steam. To keep both plants operating regardless of what happens to one or the other, it is therefore necessary to install auxiliary or standby equipment. The selection of the type and amount of such equipment is a matter of engineering design judgment and depends upon the sizes of the plants involved, whether there are alternative sources of supply for either power or water, the availability of spare parts and maintenance facilities, the possible penalties and dangers of temporarily losing these utilities, the magnitude of product storage facilities, the ability of the user to finance additional equipment, etc. The types of equipment most frequently installed to ensure continued operation of the most critical services consist of spare or standby boilers; steam bypasses with desuperheaters around steam turbines; diesel engines to provide an independent power supply (useful also for plant startup); spare pumps; an auxiliary steam condenser; product storage tanks; and a large supply of critical spare parts or complete components. It is also good practice to increase reliability by installing multiple units rather than one single large unit; e.g., two 2.5-mgd [9,460-m^3/d] desalting plants rather than one 5-mgd [18,900-m^3/d] plant. The number and size of units again are a matter of engineering judgment.

New Concepts

The preceding discussion of dual-purpose plants applies primarily to the most common type of installation, where both power in large quantities and water are required. But there are many areas where much water is needed but little or no power. This is a special case that thus far (mid-1980) has not been generally considered by consultants, manufacturers, or users. Such water-only plants have been primarily distillation units, but some of the newer installations now utilize reverse osmosis or vapor compression. However, there is a potential for the production of more product per unit of fuel by combining these plants than from either type of plant alone. Reverse osmosis plants (or electrodialysis or vapor compression) generally have electric drives.

It is therefore feasible to build a dual-purpose desalting facility utilizing (1) one desalting process that consumes electric power, and (2) another desalting process that consumes low-temperature heat. Such a combined facility would produce more water per unit of fuel than either process alone. The disadvantage is that, instead of only one plant, there are now two plants and two different processes to operate and control. Because of that complexity, such an installation probably is cost effective only in plant capacities on the order of 1 mgd [3,790 m^3/d] or larger, although such cost effectiveness can be determined only by an engineering study. The capital cost of using two different processes to produce a given quantity of product will be greater than for one process alone, and two different plants will cost more in operating and maintenance labor. The combined plants will, nevertheless, consume less fuel. The break-even point is the plant capacity at which the fuel savings of the dual-plant equals all the other increased costs. This point will vary depending upon the particular circumstances.

Developed processes which require primarily electric power include reverse osmosis, electrodialysis, and vapor compression. Plants which use primarily low-temperature heat include all of the distillation processes. A judicious combination of these plants will produce more product per unit of fuel than either alone, and this possibility therefore should be considered.

STATE OF THE ART

Major Technological Advances Since 1972

Aside from design improvements and better material selection in the fabrication of plants, two major technological advances, discussed below, have occurred since 1972.

High-Temperature Polymer-Based Additives. The use of these polymers instead of acid has increased considerably as various chemical manufacturers have aggressively moved into this market. Although additional long-term operating data will be necessary for widespread industry acceptance, early data indicate the potential for improved operation.

Large-Scale Units. Prior to 1972 the largest units built were approximately 2.5 mgd [9,460 m^3/d]. With the development of the desalination market in the Middle East, larger individual units (up to 10 mgd [37,900 m^3/d]) have been or are being constructed. This scale-up experience is vital to the reduction of future costs for large-scale distillation projects.

Major Problems

The major problems with the distillation plants continue to be corrosion, scale formation, improper material selection, and improper operation. The one area on which increasing attention has been focused is the numerous problems associated with acid dosed plants. The high hopes of the 1950's and 1960's of using acid feed to produce low-cost plants with high performance factors were dashed by the accelerated corrosion experienced in many of the plants. Although it has been shown that plants can successfully operate at high temperatures with acid, these plants are extremely delicate to operate. Many of those plants which were not fortunate in terms of the original design, materials selection, and/or operating staff suffered greatly, and the result was poor performance, low reliability, and a far shortened plant life.

In some cases there has been a failure by those entities acquiring plants to realize that properly trained and qualified personnel are required for successful long-term operation.

General State of the Art

Multistage Flash Design. The design of this process has matured to the point that MSF units are routinely manufactured in sizes from 0.1 to 10 mgd [379 to 37,900 m^3/d] by fabricators around the world. Materials, operating temperature, and method of scale control selected still vary, depending on the economics and the operating application desired by the owner. In 1977 over 85 percent of the world's distillation capacity was made up of multistage flash plants. Design performance factors of 8 lb/ 1,000 Btu's [3.43 kg/MJ] for low-temperature and 12 lb/1,000 Btu's [5.15 kg/MJ] for high-temperature plants are generally specified.

HTME Design. Commercial development of horizontal-tube multiple-effect (HTME) design occurred in the 1970's. The plants have been available since about 1975 under the names HTME, spray film, and/or multiple-effect stacked (MES). With the exception of the aluminum horizontal-tube plant in Israel, the largest operating unit, which is in Singapore, has a capacity of about 0.8 mgd [3,030 m^3/d]. This design potentially offers more operating flexibility and far less danger of scaling than a multistage flash design and develops performance factors ranging up to 20 lb/1,000 Btu's [8.6 kg/MJ].

The major difference between the aluminum-tube and the other HTME plants, aside from materials, is that the aluminum-tube plants are limited in their upper operating temperature to about 160°F [71.1°C]. The others can operate at higher temperatures and hence higher thermal efficiencies.

In 1977, the total installed capacity of horizontal-tube multiple-effect plants including those with aluminum tubes was about 10 mgd [37,900 m^3/d] compared to 638 mgd [2,410,000 m^3/d] for multistage flash plants. Consequently, the operating data base for these plants is relatively small.

VTE/MSF Design. Commercial development of the vertical-tube multistage flash (VTE/MSF) hybrid design was attempted during the 1970's. In 1975 a 3.0-mgd [11,400-m^3/d] test module was tested successfully in California (Bailie, 1977), but two 2.25-mgd [8,520-m^3/d] plants built in the Caribbean in 1974-75 have had considerable operating problems. These plants had a design performance factor of 12.2 lb/1,000 Btu's [5.23 kg/MJ] and an operating temperature of 225°F [107.2°C]. The problems seem to be related more to the specific plants and their operation than to the concept. During 1980 the two Caribbean plants had to be extensively retubed.

Vapor Compression Design. Units for seawater of this design in the capacity range of 0.005 to 0.25 mgd [19 to 946 m^3/d] are standard design and are readily available from manufacturers. Low-temperature units have been developed which reduce scale formation. Very little data have been published on the long-term performance of large (0.05-0.25 mgd) [189-946 m^3/d] low-temperature vapor-compression units.

Acid Dosed Plants. Although some plants of this design are performing well, others are experiencing heavy corrosion. There is a cautious attitude toward specifying new acid plants, especially in areas where operation and maintenance may be problematic.

Polyphosphate Plants. These have been the process of choice in many of the Middle Eastern countries and some parts of the Caribbean. They operate at up to 190°F [87.8°C] and thus their performance factor is limited to about 8 lb/1,000 Btu's [3.43 kg/MJ]. However, they have proved to be generally reliable in performance, although different feedwaters produce different results with the polyphosphates. Some concern has been expressed about the potential pollution aspects of the discharge of phosphate in the blowdown of the plants.

High-Temperature Additives. These polymer-based chemicals have been used instead of acid to control scale formation on high-temperature (up to 250°F [121.1°C]) MSF plants. Several lengthy trial runs have been performed at plants around the world during the past 4 years. Data are still being obtained under various feedwater and operating conditions. These additives appear to be effective in inhibiting scale formation, safer to handle than acid, and generally non-corrosive. Compared to acid, only small amounts of the polymer are needed (3-8 mg/l vs 100-150 mg/l for acid).

Titanium Tubes. Titanium tubes have been used in brine heaters and heat rejection sections of multistage flash and vapor compression units for over a decade. The tubes seem to resist corrosion rather well under adverse conditions. Although the cost of the tubes has been high, the cost difference between titanium and copper-nickel tube installations has decreased in recent years.

APPLICATIONS

Brackish Water

The development of membrane desalting processes has practically eliminated the use of the distillation process for desalting brackish water for potable use. The energy usage by distillation plants is nearly the same whether desalting brackish water or seawater. With membrane processes, on the other hand, there is a direct correlation between the concentration of dissolved salts and the energy required for desalting; as feedwater salinity decreases, so does the energy required.

One area where distillation could be used is for high recovery of potable water from a limited, extremely brackish water source. In such cases, some specialized vapor compression units are employed which recover over 90 percent of the feedwater stream. The cost might be justified in some inland areas where the source of raw water is extremely limited. In some cases, systems employing both reverse osmosis and vapor compression processes have been proposed (Kirchner, 1977).

There still may be steam or power generating stations which use distillation for the production of high-purity process water from a brackish source, but this does not represent a significant application for distillation.

Seawater

Seawater desalination is the major application for distillation processes. Illustrations of various applications are shown later in this chapter under the heading "Process Experience."

Other Uses

The other major use for distillation to produce potable water is the use of vapor compression units for the concentration of saline discharges from cooling towers or other desalination processes in areas where direct discharge to land or a body of water is uneconomical or environmentally unacceptable. Vapor compression units with seed-slurry scale control can achieve a high concentration factor which will reduce the brine output volume to about 5 percent of the feed. At the same time, potable water is produced and the amount of brine to be ponded or treated further is reduced.

HEALTH ASPECTS

Distillation is almost always used to process seawater, which generally contains only low levels of pathogenic microorganisms and other substances which are detrimental to health.

Microbiological

The phase changes from liquid to gas and back to liquid which occur in a distillation plant create a barrier to the passage of biological growth from the feedwater to the product water. The heat applied also acts as a sterilizing agent, although this is not as effective in the lower-temperature plants. Contamination could occur through leakage within the units, although major occurrences of such leakages are readily detected due to the rapid change in the dissolved salt level of the product water. For the sake of safety, product water from a distillation plant should be disinfected before distribution if possible.

Additives

The use of the new polymer-based additives in high-temperature distillation applications has prompted some discussion about long-term health effects on consumers. Concern centers around the toxicity of the portion of the additive that is carried over into the product water during distillation.

One additive manufacturer determined that the carryover under test conditions for their particular product was between 0.04 and 0.025 mg/l and that at that level a 132-lb [60-kg] man would ingest about 1×10^{-9} to 2×10^{-9} lb/lb/day [0.001-0.002 mg/kg/day]. Further studies by this manufacturer indicated that at this level, the use of the material produces no toxic hazard in the drinking water (Leimgruber, 1979). Other manufacturers are undertaking similar studies.

ENVIRONMENTAL CONSTRAINTS

General

On the whole, efforts to provide water for water-short areas have not devoted a great deal of attention to the environmental aspects of distillation facilities. The discussions below concern potential environmental constraints associated with seawater distillation facilities.

Intake Structures

Intake structures often must be built offshore in order to obtain clean water and/or allow for high tidal fluctuations. Thus an intake pipeline in a trench or on the bed of the adjacent coastal water must be installed, sometimes penetrating reefs or other coral structures. All of this activity can involve disruption of the marine environment.

Brine Discharge

The discharge from a distillation plant generally differs from the existing seawater in temperature, chemical composition, and microbial biota (Winters et al., 1979).

Changes in these parameters could cause some disruption of the marine environment, with the greatest impact occurring closest to the point of discharge before mixing occurs.

The concentrated brine is characterized by elevated temperature, the presence of pretreatment chemicals (such as scale control agents), altered (usually depressed) pH, and traces of heavy-metal corrosion products. There have been suggestions that eutrophication of relatively closed areas, such as the Arabian Gulf, might be caused by the heavy use of scale prevention chemicals such as polyphosphates (Leimgruber, 1979).

Concern is growing, especially with larger distillation plants being built, that more care needs to be taken with these discharges (Winters et al., 1979; Korte, 1979; Libert et al., 1979; Bakish, 1979).

Noise

Distillation plants can create noise due to activities associated with the production of steam and the running of rotating equipment. Obviously, a buffer area should be planned around such facilities to relieve this problem. For package vapor compression units, manufacturers can usually provide an estimate of the noise level involved.

FUTURE PROSPECTS

New Technology Being Marketed

Numerous developments have begun to be marketed or will be marketed in the distillation industry in the next 5 years. Many of these innovations or developments will enhance the practice of distillation. It should be recognized that segments of the distillation business tend to be very conservative about accepting innovations until they have been extensively proven in operation. This conservatism may stifle many good ideas, but it seems to be justified by some historical disasters in the industry.

Aluminum Plants. Aluminum has been used during the 1970's to construct several 1-mgd [3,790-m^3/d] distillation plants. Aluminum is relatively inexpensive, is easy to fabricate, and has a high thermal conductivity. However, it must be operated at low temperatures (below about 170°F [76.7°C]) to avoid some forms of corrosion, and the metal is very susceptible to galvanic corrosion, with trace amounts of heavy metals such as copper. Brine velocities within the plants must be kept below about 3 ft/sec [0.9 m/s] to avoid attack on the protective skin of the aluminum (Veenman, 1978).

A 1.0-mgd [3,790-m^3/d] multiple-effect distillation plant built in about 1974 has been operated in Eilat, Israel (Glueckstern, 1979). Three more plants of the same design, with a total capacity of 3.75 mgd [14,200 m^3/d], are being constructed for the Virgin Islands Water and Power Authority on the islands of St. Croix and St. Thomas.

In late 1979 the second of two 1-mgd [3,790-m^3/d] all-aluminum multistage controlled flash plants was completed for the Water Energiebedrijf (WEB) on the island of Aruba in the Netherlands Antilles. A very large 10-mgd [37,900-m^3/d] aluminum plant using a multiple-effect design is being designed for construction in Israel in the early 1980's. Aluminum has also been used on smaller, low-temperature, vapor compression units (Hoffman, 1977). Operating data from these plants should be available in the next few years.

Alcoa built and operated an aluminum waste heat recovery evaporator in Puerto Rico for 3 years during the 1970's with favorable results.

Concrete Plants. Work has been carried out in several countries on the use of concrete as a shell material. This material has corrosion, thermal, strength, and cost characteristics which indicate potential for future use. Test plants have been built in Japan and France (Norjiri and Fujii, 1976). Problems with concrete include attack by both hot brine and hot condensate and the attachment and sealing against both pressure and vacuum of metal penetrations through the shell.

Horizontal-Tube Multiple-Effect. Although this is not a new technology, there is only limited operating experience with large units. As larger units are built and operate successfully, this type of process can be expected to be marketed in direct competition with multistage flash plants. In addition to the low-temperature aluminum plants being built in Israel and the Virgin Islands, three 0.15-mgd [568-m^3/d] high-temperature plants are operating in Belgium, and two 0.4-mgd [1,510-m^3/d] plants are operating in Venezuela; also, a 0.8-mgd [3,030-m^3/d] is operating for the Shell Refinery in Singapore, and a duplicate is starting up at the Shell Refinery in Curacao.

Vertical-Tube Evaporators. The potential for a high performance factor using variations of vertical-tube configurations will make this process attractive in the 1980's. There have been a variety of studies and discussions using both rising and falling films in the process but, as with the horizontal-tube multiple-effect plants, long-term operating experience is limited (Kanal and Hughes, 1979; Nishimoto, 1977).

Although VTE plants have been built in several areas, including the Virgin Islands, California, and Japan (Satone et al., 1977; Bailie, 1977), no two plants are alike and documented experience is limited. If the installation of a large vertical-tube plant is being considered, the operating and performance experience of plants already installed should be reviewed to help determine the proper type, design, and materials.

Foam Ball Cleaning Systems. Foam ball cleaning systems (referred to as the Taprogge system) for use in distillation plants has increased in the past decade. In this system foam rubber balls are introduced into the tubes of the various sections of MSF plants along with the recycle flow. These balls help to scour out soft sludge and silt deposits. Some plants using the system have reported good results, including increased thermal efficiency and a reduced number of shutdowns (Romeijn and Eimer, 1978). Foam ball cleaning is not effective on all distillation plants nor on all scales. The systems can be costly and their application must be carefully selected.

Other New Developments. A number of other innovations are currently being developed and may be marketed. These include:

Vertical-Tube Foam Evaporation
Falling Film Multistage Flash
Direct Contact Condensation Multistage Flash
Multistage Controlled Flash
Multistage Flash/Fluidized Bed

These are described in further detail by Veenman (1978), and the following is a summary of these details.

i) Vertical-Tube Foam Evaporation (VTFE)--This process was developed at the Seawater Conversion Laboratory at Berkeley, California. A surfactant is added to the seawater feed to enhance evaporation. The process has been tried in a pilot plant with both up- and down-flow modes in the vertical tubes.

ii) Falling Film Multistage Flash (FFMSF)--Developed by W. L. Badger Associates, Inc., in Greenbank, Washington, this is a vertically stacked multistage flash unit with straight vertical tubes used for heat recovery. The brine feed enters the plant at the top and drops down the tubes which pass through the vertically stacked stages. The brine film passing down the inside of the tubes is inhibited from evaporating by the maintenance of atmospheric pressure at the center of the tubes. Flashing is enhanced by the use of rotating impellers on a common shaft which aid in increasing surface area exposure while they centrifuge the brine upward from stage to stage.

This type of unit was developed for small-capacity installation such as for ships and hotels.

iii) Direct Contact Condensation Multistage Flash (DCC-MSF)--This process was developed at the Israel Institute of Technology (Technion) at Haifa, Israel. The design eliminates the expensive heat recovery tubes and instead uses direct contact condensation which occurs between the hot brine and the cooling water stream, which flow countercurrently in open channels. A low-cost liquid-to-liquid heat exchanger employing plastic tubes was developed for recovering heat from the cooling water stream.

iv) Multistage Controlled Flash Evaporator (CFE)--This has been commercially developed by Aquanova, B.V., of Rotterdam, The Netherlands. Two of these units were placed in operation in Aruba in 1979-80. They are vertically stacked multistage flash plants made of aluminum and some plastic. The design lends itself to modular construction with compact flash chambers which reduce energy losses and have the potential for lowering capital and operating costs.

v) Multistage Flash/Fluidized Bed Evaporator (MSF/FBE)--This process was developed by Delft University of Technology and Esmil, The Netherlands. It utilizes a vertically stacked multistage flash configuration with straight vertical heat recovery tubes. In order to obtain the high heat transfer necessary in the heat recovery (and condensation) tubes so as to limit their length (and the height of each stage), solid particles (usually glass beads) are introduced into the feedwater. These fluidized particles disrupt the laminar flow along the inside of the tube walls to enhance heat transfer while inhibiting scale formation (Veenman, 1976).

Future Technology

It can be expected that future technological advances will be attempted in the following areas (Watson, 1976):

Increased Operating Temperature. In order to increase efficiency (performance factor) of operation, continued efforts to raise the upper operating temperatures are expected. This will require additional work with brine chemistry, pretreatment to inhibit scale formation, etc.

Increased Material Life. This is expected to center around studies of corrosion, heat transfer, non-metallics development, and material compatibility.

Increased Reliability. Improvements in the reliability of plants to increase the on-stream factor and reduce maintenance time and cost will involve improved design, materials selection, pretreatment, and operational control.

PROCESS EXPERIENCE

Previous A.I.D. Studies

Previous editions of the A.I.D. manual on water desalination published in 1968 and 1972 contained a number of case histories of various distillation plants. These case histories are summarized and updated in the following paragraphs.

Freeport, Texas (1 mgd [3,790 m^3/d], LTV). This demonstration and test plant was built for the Office of Saline Water (OSW) in Freeport, Texas. It began operation July 1, 1961 and by June 30, 1965, it had produced one billion gallons [3,800,000 m^3] of water. After a number of modifications, it was dismantled in the mid-1970's.

i) Process Description--The plant was a long tube vertical multiple-effect plant. It originally had 12 effects and used the falling film method of vaporization. Seawater was obtained from an adjacent seawater canal and steam from the nearby chemical plant. The product water was delivered to the City of Freeport. An acid-deaeration method of scale control was used, and the operating temperature at the brine outlet of the first effect was 265°F [129.4°C]. The temperature drop in each stage was about 12°F [6.7°C].

ii) Operating Information--The plant incorporated a number of innovations for testing the process and the materials used. One of these was the use of steel tubes in five of the effects. During operations these were excessively corroded and were replaced with aluminum-brass tubes. Later, other tube materials were tested, including copper-nickel alloys, stainless steel, and titanium. Additional work was done on acid pretreatment, pH stability, and increasing the capacity of the plant by adding five effects and using roped tubes in an MSF feed heater.

It should be remembered that in many of the OSW installations the operating personnel were specially selected and qualified and were able to give the experimental units the special attention which they might not receive at a typical installation.

iii) Current Status--Reduction of the OSW budget forced the dismantling of the plant in about 1975.

Point Loma, California (1 mgd [3,790 m^3/d], MSF). The Point Loma plant was placed in service in 1961 as the Office of Saline Water Demonstration Plant No. 2. It was shut down in 1964 for dismantling, elimination of four of the bottom stages, and then shipment of the modified plant to the U.S. Naval Facility at Guantanamo Bay, Cuba. As of 1980 it was still operating at Guantanamo furnishing potable water for the base.

i) Process Description--The plant was designed as a 36-stage multistage flash evaporator with a performance ratio of about 11 [4.72 kg/MJ]. When being used as an experimental plant at Point Loma, the plant was operated both with polyphosphate and acid pretreatment systems and top brine temperatures of 200°F and 250°F [93.3°C and 121.1°C], respectively, depending on the scale treatment.

At Guantanamo the plant operates at a top brine temperature of 180°F [82.2°C] using polyphosphate for scale control. At this operating temperature the production is 800,000 gpd [3,030 m^3/d].

ii) Operating Information--During operation at Point Loma, sand, silt, and marine growth in the seawater intake entered the intake pipe that had been extended into the ocean a distance of 700 feet [213 m]. This caused maintenance problems in the form of clogging of tube sheets (which restricted flow) and excessive wear on rotating equipment.

iii) Current Status--The plant is still being operated. At Guantanamo, the plant is operated by highly qualified personnel who observe a strict schedule for preventive maintenance, refurbishing, repairs, and operation.

Claire Engle Plant, San Diego (1 mgd [3,790 m^3/d], MEMS). The Claire Engle plant was built in 1967 under funding by the Office of Saline Water as a replacement for the Point Loma plant.

i) Process Description--This 1-mgd [3,790-m^3/d] combination multiple-effect multistage facility was designed for a performance factor of 21.5 lb/1,000 Btu's [9.22 kg/MJ]. It had three effects, each effect being a multistage unit.

ii) Operating Information--To obtain a performance factor of 21.5 lb/1,000 Btu's [9.22 kg/MJ], the plant had to operate at a recycle brine temperature of 300°F [148.9°C]. This high temperature, in turn, required that the seawater have a large portion of its calcium removed by the lime-magnesium-carbonate process in order to avoid hard scale. As it turned out, the bulk of the development efforts were involved in operating the seawater treatment plant. Also, the largely steel construction of the plant combined with the very intermittent operation to produce much corrosion so that considerable time was spent on repairs. It was in this plant that Precrete cement linings were first applied successfully. Just before the OSW financial support ended, the plant was operated at 300°F [148.9°C] for about 1 month.

iii) Current Status--When OSW funding ended in 1973, the plant was dismantled. Although its performance factor of 21.5 lb/1,000 Btu's [9.22 kg/MJ] was considerably higher than the usual maximum for MSF plants, its capital cost was 2.5 times higher. At that time, the low cost of energy would not justify a 250 percent greater capital cost and the concept was dropped.

Roswell, New Mexico (1 mgd [3,790 m^3/d], VC). This 1-mgd [3,790-m^3/d] vapor compression plant, no longer in operation, was built in 1963 as the Office of Saline Water's Demonstration Plant No. 4.

i) Process Description--The plant was a two-effect, vertical-tube, forced circulation, vapor compression unit. Because of the excessively high calcium and silica content of the brackish feedwater an ion exchange pretreatment feed system was installed. An oil-fired boiler provided steam to heat the unit up to operating temperature, above which it ran on energy supplied by the compressor. The compressor was designed to compress 175,000 pounds of vapor per hour [79,400 kg/h] from 17 psia to 24 psia [1.16 to 1.63 atm] (a 1.41 compression ratio) and utilized a 2,500-hp [1,864-kW] electric motor.

The brine was forced up vertical tubes which were heated by the vapor condensing on the outside (shell side). Upon reaching the top of the tubes, the heated brine was flashed in a large spherical chamber located on top of the tube assembly. This flashed vapor from the first effect passed to the shell side of the vertical tubes in the second effect where it was condensed so as to heat the brine passing inside. The condensate became part of the product.

The unvaporized brine in the first effect flowed to the bottom of the flash chamber, where it was circulated through an external downcomer to the recycle pump. A small part of the brine recycle stream was removed as blowdown from the first effect and flowed into the downcomer of the second effect as the makeup to the brine recycle stream of the second effect.

As the brine recycle stream passed through the vertical-tube bundle of the second effect it was heated by the condensation of steam from the first effect. This heated brine stream emerged into the second flash chamber, where part of it vaporized. The mixture of vapor and hot brine was separated in the flash chamber, and the vapor was pumped by the vapor compressor to the higher-pressure first effect shell, where it condensed to form product water.

The brine which was not vaporized in the second effect was recirculated through an external downcomer and recycle pump, as in the first effect. The brine blowdown from the process was transferred from the second effect recycle stream at the recycle pump discharge; it passed to the heat-recovery heat exchangers and then to the brine disposal pond. A portion of the brine blowdown was utilized to regenerate the ion exchangers. The condensed product water from the first and second effects was cooled in the heat-recovery heat exhangers, thereby heating the incoming feedwater, and was then stored in an onsite storage tank from which it was pumped to the Roswell water mains (Kaiser, 1967).

ii) Operating Information--The feedwater to the unit was uncommon because it was very high in calcium sulfate and silica relative to seawater, thus requiring the special ion exhange pretreatment step. The problems that occurred in the operations stemmed in a large part from failure to maintain the necessary feedwater quality. The problems with feed treatment led to the development by Resources Conservation Corporation of their highly successful calcium sulfate seeding technique for preventing scale even in boiling saturated solutions.

The plant was plagued from the outset by compressor problems associated with an inadequate foundation. These problems were never resolved. Toward the end, a high temperature water jet ejector was installed which permitted the plant to operate, but the water jet ejector was not energy-efficient and its development was not pursued further. The plant suffered severely throughout its brief life from corrosion and scaling.

iii) Current Status--When OSW financial support terminated in 1973, the plant was dismantled.

Fountain Valley, California (3 mgd [11,400 m^3/d], VTE/MSF). This facility, a 3-mgd [11,400-m^3/d] test module which was to be expanded to about 12.5 mgd [47,300 m^3/d], was built under joint funding by the Office of Saline Water and the Orange County Water District. The plant was designed to produce low-TDS water which was to be blended with effluent from an advanced wastewater treatment plant. The combined product was to be used for artificial ground-water recharge in the area.

i) Process Description--The test plant is a vertical-tube multiple-effect plant using a once-through multistage flash configuration for a preheater. The vertical tubes were fluted to increase heat transfer, and a falling film process was used to distribute the brine on the inside of the vertical tubes for heating.

An outline of the design data for both the 3-mgd [11,400-m^3/d] module and the full 12.5-mgd [47,300-m^3/d] facility is shown in Table 3-6.

ii) Operating Information--As Table 3-6 indicates, this module was designed to be an integral part of the 12.5-mgd [47,300-m^3/d] facility. This first 3.0-mgd [11,400-m^3/d] module was to be used for testing and evaluation of the design and the remainder of the facility was then to be added on. Unfortunately, the completion of the test module coincided with the reorganization of the Office of Saline Water into the Office of Water Research and Technology within the Federal Government. The continuation of funding by the Office of Water Research and Technology (OWRT) for the further expansion and operation of the VTE/MSF unit to the full capacity of 12.5 mgd was not possible due to changes in budget and priorities.

The test module was operated on an intermittent basis during 1975 and 1976, with the plant operating an average of 9 days per month over a 10-month period. Due to an irritating low-frequency boiler noise, the plant was only operated in the daylight hours (Fountain Valley is a residential community). The results of this testing program as interpreted by a consultant indicated that there were potential problems involved with the design using overly optimistic heat transfer rates, distributor nozzle plugging, brine flow limitations in the feed heater, cavitation with one of the brine recycle pumps, and corrosion of tube sheets (Bailie, 1977).

Unfortunately, due to the concept that the test module would be part of a larger plant, it was uneconomical (with a design performance factor of only 3 lb/1,000 Btu's [1.29 kg/MJ]) to operate the test module alone. Several suggestions were made on how the test module could be modified for expansion into a larger unit (Bailie, 1977). The unit was not utilized further but remains at the Fountain Valley location.

In 1977 a 5.0-mgd [18,900-m^3/d] reverse osmosis facility was built at Fountain Valley to reduce the salt content of the effluent from the advanced wastewater treatment plant before ground-water recharge. This facility is currently being used.

Table 3-6
Design Data for the Fountain Valley VTE/MSF Distillation Facility

	3-mgd [11,400-m^3/d] Test Module	12.5-mgd [47,300-m^3/d] Facility
Evaporator (VTE)		
Structure	Cylindrical shell	Cylindrical shell
Construction	Clad & concrete lining	Clad & concrete lining
No. of effects	4	16
Tube material	Al-Brass (CDA-687)	Al-Brass (CDA-687)
Tube type	Double-fluted	Double-fluted
Tube length	10'[3 m]	10' [3 m]
Tube count		
2-in. [5.1-cm] tubes	5,436	29,869
3-in. [7.6-cm] tubes	3,596	8,999
Feed heater (MSF--long tube)		
Structure	Cylindrical shell	Cylindrical shell
Construction	Clad & concrete lining	Clad & concrete lining
No. of stages	5	29
Tube material	90-10 Cu-Ni	90-10 Cu-Ni
Tube type	Spirally indented	Spirally indented
Tube diameter	1" [2.54 cm]	1" [2.54 cm]
Performance Data		
Performance factor		
lb/1,000 Btu's	3.3 Max.	14.0
kg/MJ	1.42	6.0
Steam		
lb/hr	344,000	344,000
kg/hr	156,000	156,000
Maximum Brine Temperature		
°F	162-265	267
°C	72.2-129.4	130.5
Concentration factor	1.2-2.9	3.0
Design life (years)	30	30

Source: Adapted from Bailey, 1977.

iii) Current Status--The Fountain Valley installation has not been operated since 1976.

Odessa, Texas (0.5 mgd [1,890 m^3/d], VTE). This industrial plant was used by the General Tire and Rubber Company's rubber producing plant in Odessa, Texas, to process waste effluent. Only limited information is available.

i) Process Description--The plant is a 0.5-mgd [1,890-m^3/d] 14-effect vertical-tube evaporator which operates on mildly brackish water of unknown composition.

ii) Operating Information--The plant was plagued with problems from an early date. The water feed composition is unknown, but since it was a waste stream from a rubber plant, it probably contained various sulfur compounds which caused the fluted 90-10 copper-nickel tubes in the first several effects to fail. These were replaced with fluted titanium tubes that were not anealed after fluting (like stainless steel, titanium work hardens rapidly) and the titanium tubes soon developed lengthwise cracks along the flutes.

iii) Current Status--Nothing concerning its current status is known with certainty. It is believed to be inoperative.

Additional Case Studies

Rosarito, Mexico (7.5 mgd [28,400 m^3/d], MSF). In 1968 the Mexican Government's Federal Electric Commission (CFE) started up the first of two units which comprise their 7.5-mgd [28,400-m^3/d] MSF distillation installation at Rosarito in Baja California (See Figure 3-27). The water produced at the plant supplies a portion of the needs of the area around the City of Tijuana, as the available natural sources of potable water are insufficient.

This plant, constructed by Aqua-Chem, Inc. (USA) and operated by the Federal Electric Commission, was one of the largest desalination facilities in the world at the time of construction. Information for this discussion of the Rosarito plant is outlined in a report by Gomez (1979) of the CFE.

i) Process Description--The plant is a multistage flash design with two 3.75-mgd [14,200-m^3/d] units. Each unit has 44 stages (40 heat recovery and 4 heat rejection stages) and uses recirculating brine. The top temperature of the feedwater in the brine heater is 235° F [112.8°C], and acid is used to control scale formation in the units. A flow diagram of the facility is shown on Figure 3-28.

The units were designed to utilize steam from the adjacent electric generating plant. Steam is extracted between the high- and low-pressure turbines which were modified to permit this extraction. After some modifications to the plant during the startup period, including the addition of the Taprogge sponge ball cleaning system for the heat exchange tubing, the units were accepted in 1970 (unit 1) and 1971 (unit 2).

At the time of construction the total cost of the faciity was about $10,800,000, including modifications to the steam generating facility.

ii) Materials of Construction--The following materials were used in construction of the facility:

Heat Exchanger Tubes--copper-nickel (90-10)
Shell--carbon steel
Pump Impellers--aluminum-bronze, stainless steel, monel, and bronze, depending on application
Piping--concrete, rubber-lined carbon steel, wrought iron, and asbestos cement

The copper-nickel tubes were manufactured in Mexico, but much of the other equipment was imported.

During the 1970's Mexico placed great emphasis on becoming self-sufficient in technical expertise and manufacturing capabilities in the field of desalination.

iii) Operating Information The design versus operating parameters are compared below.

	Design Parameter	Operational Condition
Production	7.44 mgd [28,200 m^3/d]	4.7 mgd [17,800 m^3/d]
Performance Factor	9.4 lb/1,000 Btu's [4.03 kg/MJ]	7.0 lb/1,000 Btu's [3.0 kg/MJ]
Product Quality (TDS)	25 mg/l	70-90 mg/l

Figure 3-29 graphically shows the plant factor and unit production costs in Mexican pesos for the period 1969-1979. Figure 3-30 shows the breakdown of the unit cost attributable to capital, labor, steam, etc., during the same period.

iv) Operating Problems--Some of the major operating problems during the past 10 years involved the heat recovery and reject tubes, flash chamber corrosion, and the brine heater.

Corrosion products, flakes, as well as copper and iron oxide sludge, accumulated in the tubes in the heat recovery stages, reducing heat transfer and hence plant production. Various methods of tube cleaning both during operation and shutdown periods were tried. The use of locally produced soft balls which were injected through the tubing proved to be one solution to the problem.

Erosion and corrosion of the heat rejection tubes were caused by sand, mud, and marine organisms in the seawater.

The interior areas and divisions between the flash chambers suffered severe corrosion of the carbon steel material. This corrosion caused air leakage into the chambers, thereby accelerating corrosion and reducing heat transfer efficiency. Many areas in the plant required protection of the carbon steel by lining with stainless steel.

Frequent startups and shutdowns have been very hard on the plant and resulted in considerable corrosion.

v) Current Status--The plant requires continued maintenance and repairs because of corrosion, but is still operating.

Ghubrah, Oman (7.2 mgd [27,300 m^3/d], MSF). In 1976 the Ministry of Electricity and Water of Oman commissioned this 5-mgd [18,900-m^3/d] multistage flash plant. Shown on Figure 3-31, this facility can supply 7.2 mgd [27,300 m^3/d] (when operated at 235°F [112.8°C]) for the area around the capital city, Muscat. The facility was built by Demag (West Germany) using a design by Envirogenics Systems Company (USA). Data for this summary came from a paper by Chalchal, 1979.

The plant has supplied water to the capital area since startup and utilizes a 1-month shutdown annually for maintenance.

i) Process Description--The facility is a multistage plant with a single unit of 7.2 mgd [27,300 m^3/d]. It has 20 stages with a long-tube configuration and uses recirculating brine. The unit was designed for flexibility in operating temperature, scale control, and production capacity. At an operating temperature of 195°F [90.6°C] using polyphosphate or any other low-temperature scale control agent, the plant can produce about 5 mgd [18,900 m^3/d]. At higher operating temperatures (e.g., 235°F [112.8°C]) using acid or other suitable high-temperature scale control agent, it can produce about 7.2 mgd [27,300 m^3/d].

The plant was designed with a performance factor of 6.8 lb/1,000 Btu's at 195°F and 7.2 lb/1,000 Btu's at 235°F [2.92 kg/MJ at 90.6°C and 3.09 kg/MJ at 112.8°C].

ii) Materials of Construction--The following materials were used in the construction of this facility.

Heat Exchanger Tubes
    Brine heater--70/30 Cu-Ni
    Heat recovery--90/10 Cu-Ni
    Heat rejection--70/30 Cu-Ni

Shell--Steel

Evaporator Internals
    Water boxes--rubber-lined steel
    Brine gates--high-molybdenum stainless steel

External Piping--rubber-lined steel

iii) Operating Information--Figure 3-32 presents a graph of the output of this unit. It should be remembered that the output reflects the demand rather than the capability of the plant. At startup in 1976, the water demand was low and the plant was operated at low temperature and production using pholyphosphate.

iv) Current Status--The facility is operating and the Ministry is currently evaluating bids on another long-term operating contract by several private firms.

APPLICABILITY FOR DEVELOPING COUNTRIES

The use of the distillation process in developing countries offers its greatest potential in the desalination of seawater. This potential is best realized when the process is used in conjuction with low-cost energy obtained from low-pressure steam as part of a dual-purpose (electricity-desalination) facility.

Multistage and Multiple-effect Plants

The greatest problem with the distillation process is the complexity of operation and maintenance. This is especially true with plants with high thermal efficiencies (performance factors) operating in temperature ranges from 200 to 250°F [93.3 to 121.1°C]. Associated with these plants are pumps, vacuum systems, deaerators, chemical feed systems, brine heaters, etc., and many of these components operate under conditions of high temperature in a highly saline environment.

This requires that meticulous attention be devoted to operation and that the facility have the capability to supply maintenance services involving skilled personnel, appropriate tools, and an adequate supply of parts and materials.

If facilities are to be built in areas where trained, skilled operators and maintenance personnel are not readily available, it is prudent to ensure that such facilities are designed and constructed conservatively. This means that care must be taken to select the proper materials of construction and a design which can withstand abuse. Immediate capital costs can therefore significantly increase, but the long-term operating costs will probably be much lower. It is speculated that the introduction of the new high-temperature scale control chemicals will ease many of the operating problems of multistage flash plants.

Vapor Compression Plants

Although vapor compression plants appear to have a higher thermal efficiency by virtue of their higher performance factor, these plants generally require high-cost energy, which increases their overall operating costs.

The three types of vapor compression units available include low-temperature (compressor), high-temperature (compressor), and steam jet ejector. Each offers different attributes for installation in developing countries. The units employing compressors have higher efficiencies but require the maintenance of a high-speed compressor. Conversely the steam jet ejector is not exceptionally efficient, but it requires minimal maintenance.

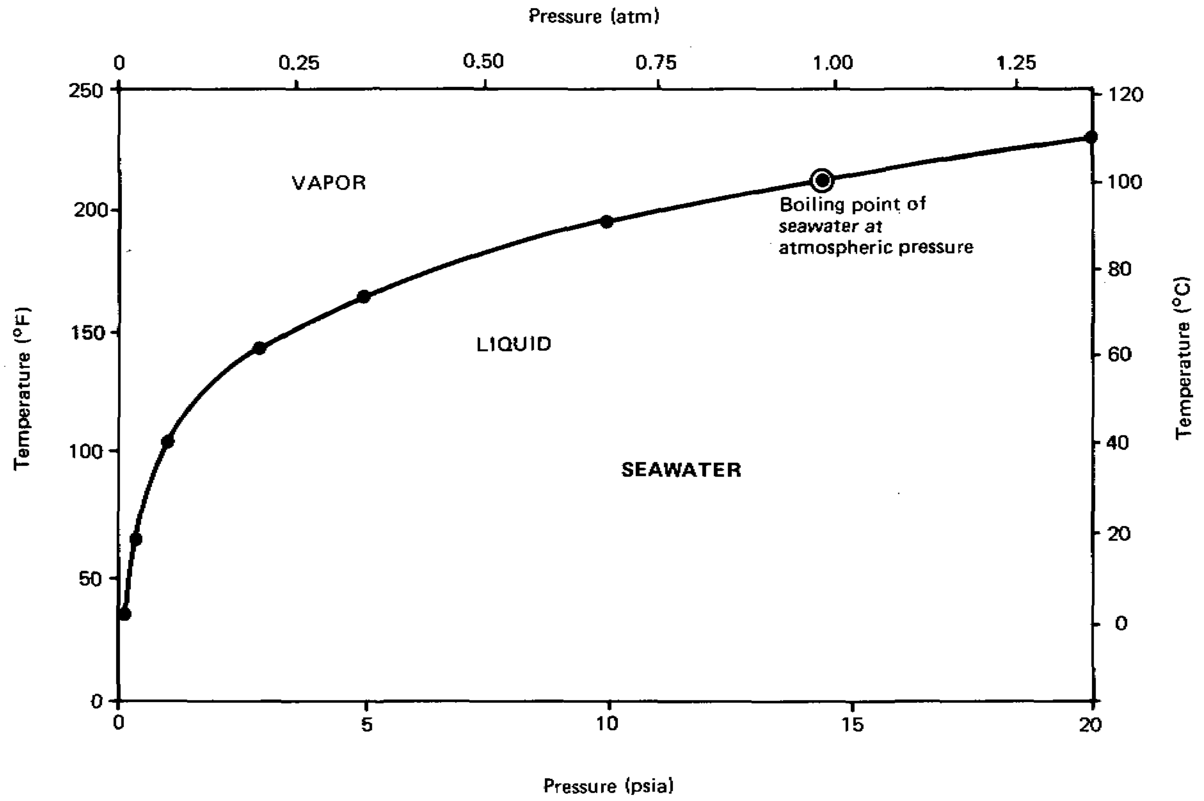
Waste Heat Recovery Evaporators

The simple, single-effect, waste heat recovery evaporator offers a potential for water production in areas with a continuous, dependable supply of waste heat and abundant cooling water. The evaporators are relatively easy to understand and operate, and the associated pieces of machinery (mostly electric pumps) are relatively easy to repair and maintain.

Summary

For successful long-term operation, distillation facilities, except for perhaps a steam jet ejector vapor compression unit, require that adequately trained and skilled operators and mechanics be available.

Further information is available in Chapters 8 (Process Selection) and 9 (Desalination in Developing Countries).



Data from Kellogg, 1975

FIGURE 3-1. Relationship between temperature and pressure at the boiling point.

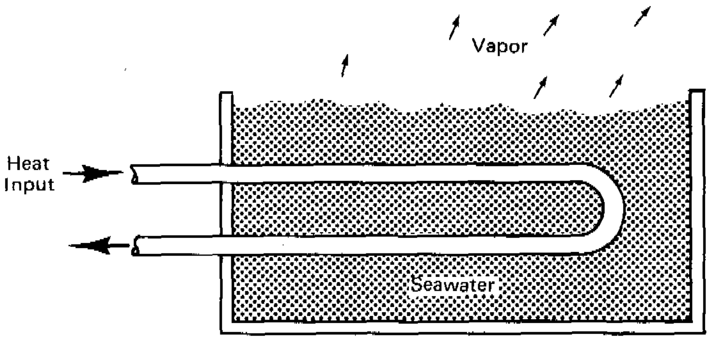


FIGURE 3-2. Submerged tube method of vaporization.

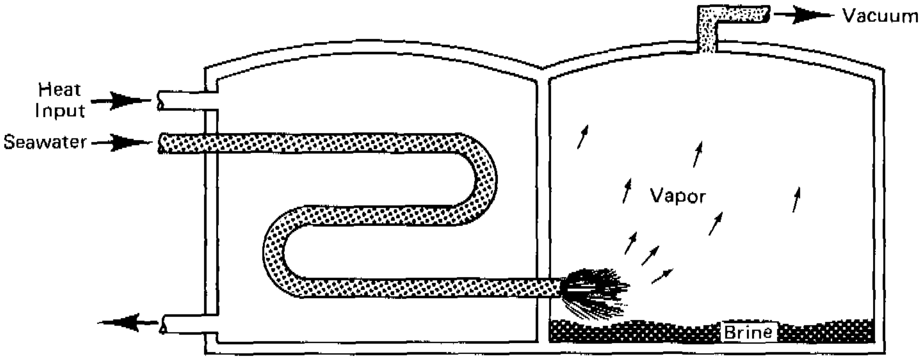


FIGURE 3-3. Flash method of vaporization.

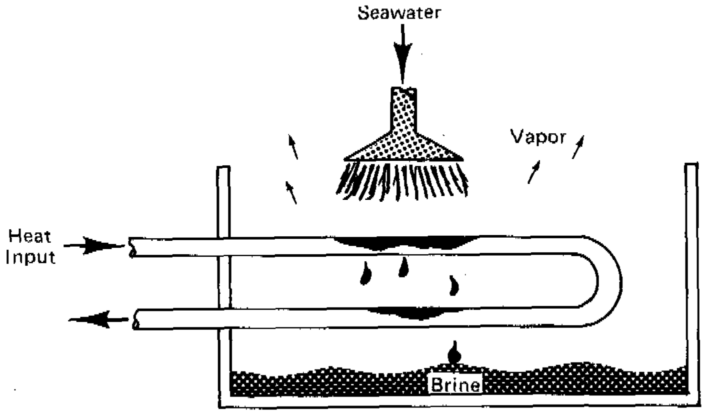
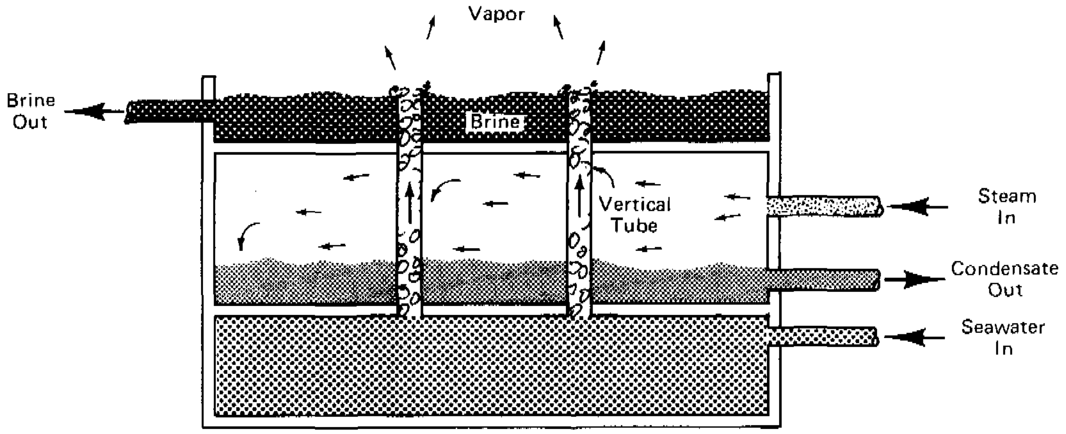
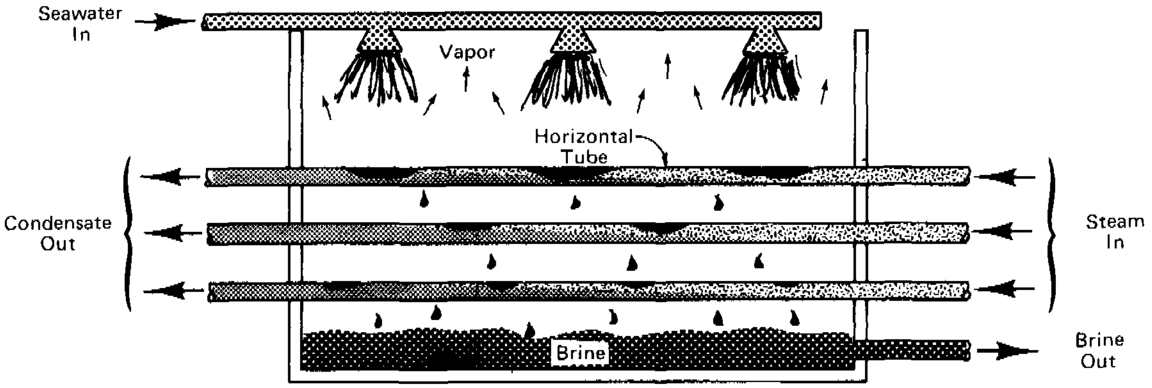


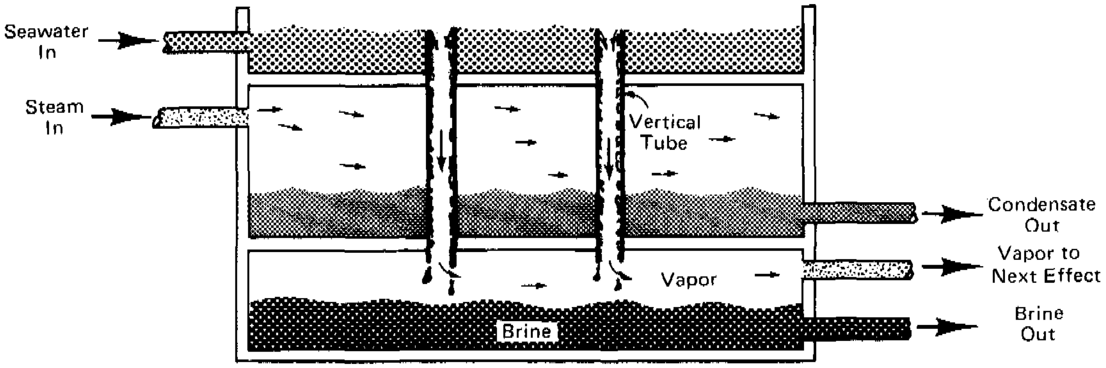
FIGURE 3-4. Thin (or spray) film method of vaporization.



CLIMBING FILM IN VERTICAL TUBES

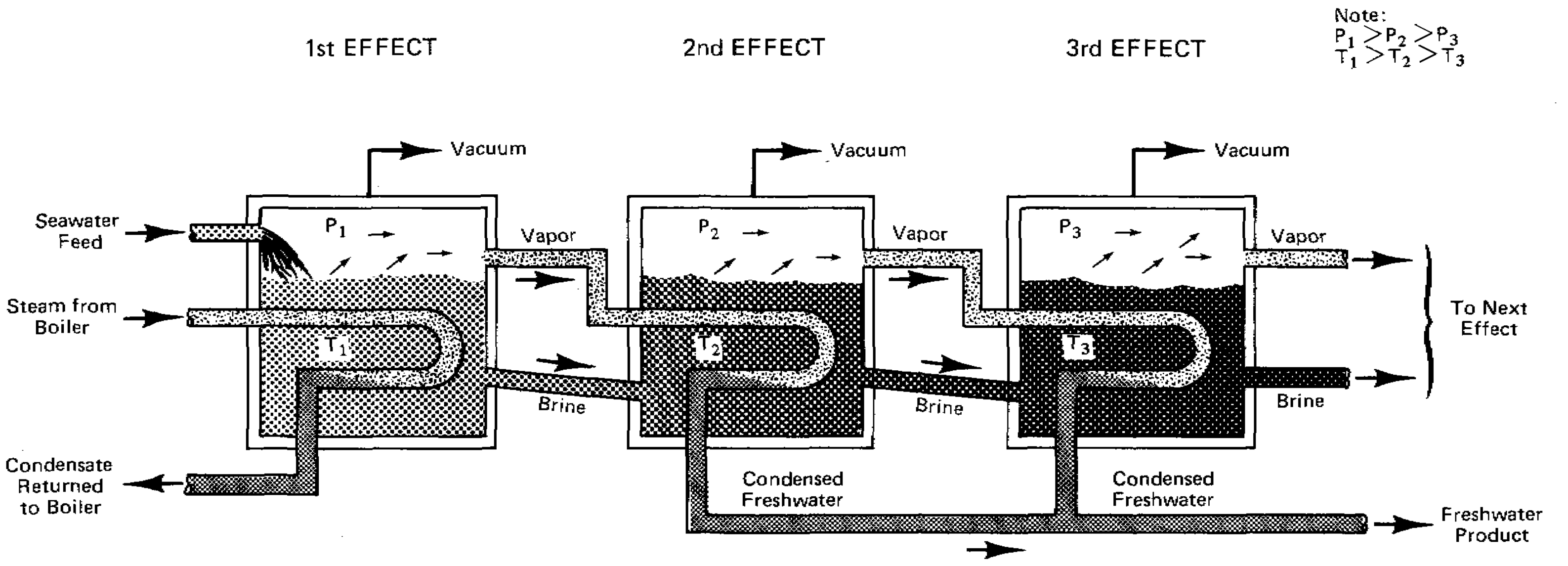


SPRAY FILM ON HORIZONTAL TUBES



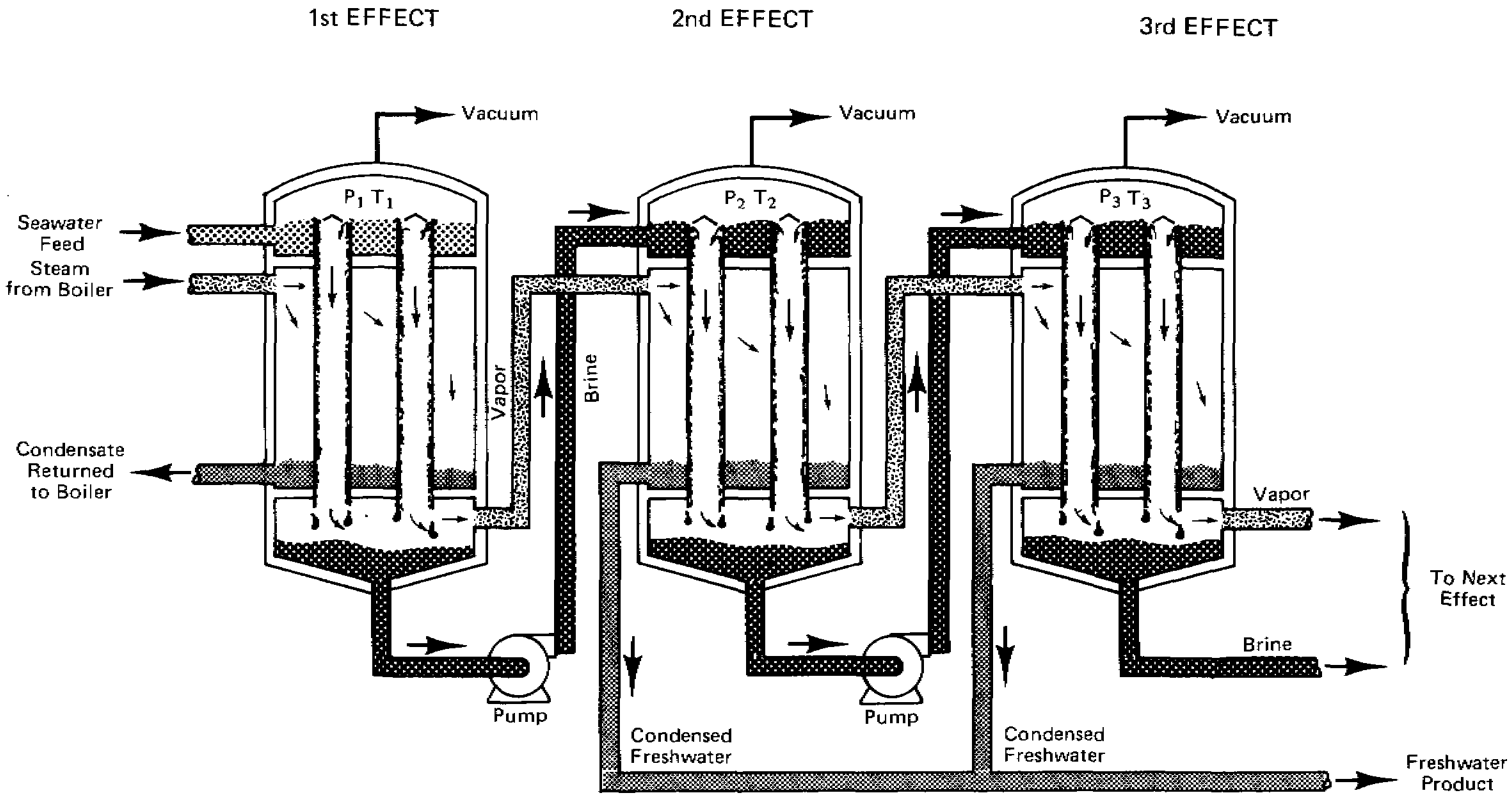
FALLING FILM IN VERTICAL TUBES

FIGURE 3-5. Three methods of applying thin films.



Notes: 1. This drawing is greatly simplified.
2. A final condenser such as shown on Figure 3-10 is necessary for operation.

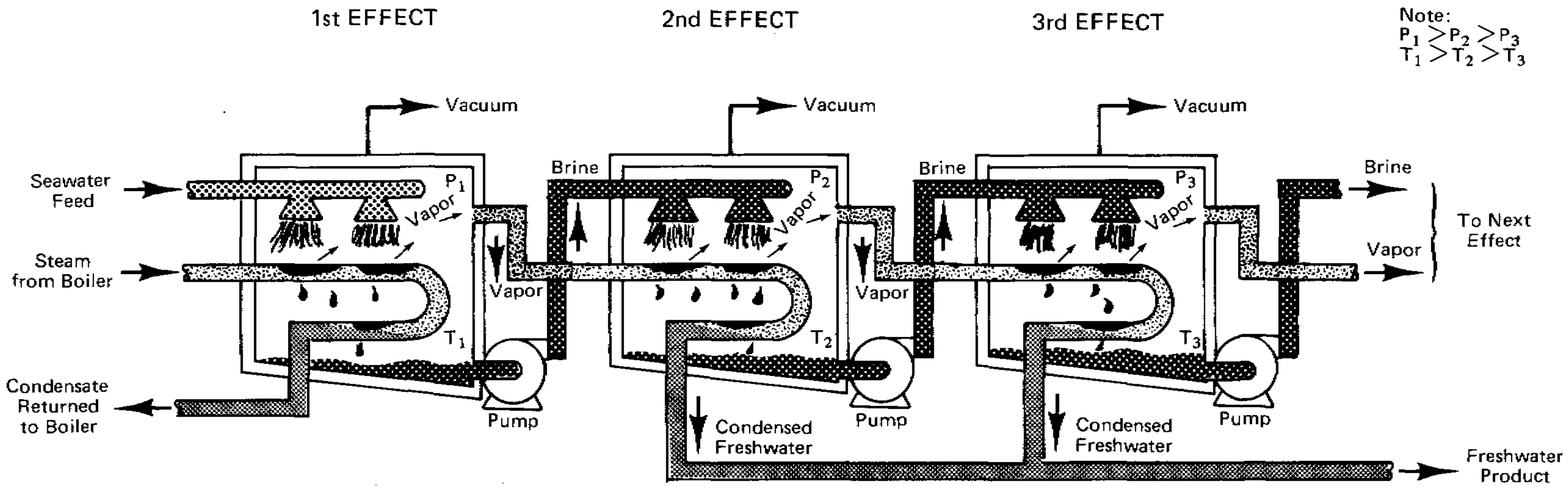
FIGURE 3-6. Conceptual diagram of a submerged tube (ST) multiple-effect distillation plant.



Notes: 1. This drawing is greatly simplified.
2. A final condenser such as shown on Figure 3-10 is necessary for operation.

$$P_1 > P_2 > P_3$$
$$T_1 > T_2 > T_3$$

FIGURE 3-7. Conceptual diagram of a vertical-tube evaporator (VTE).



Notes: 1. This drawing is greatly simplified.
2. A final condenser such as shown on Figure 3-10 is necessary for operation.

FIGURE 3-8. Conceptual diagram of a horizontal-tube multiple-effect (HTME) distillation plant.

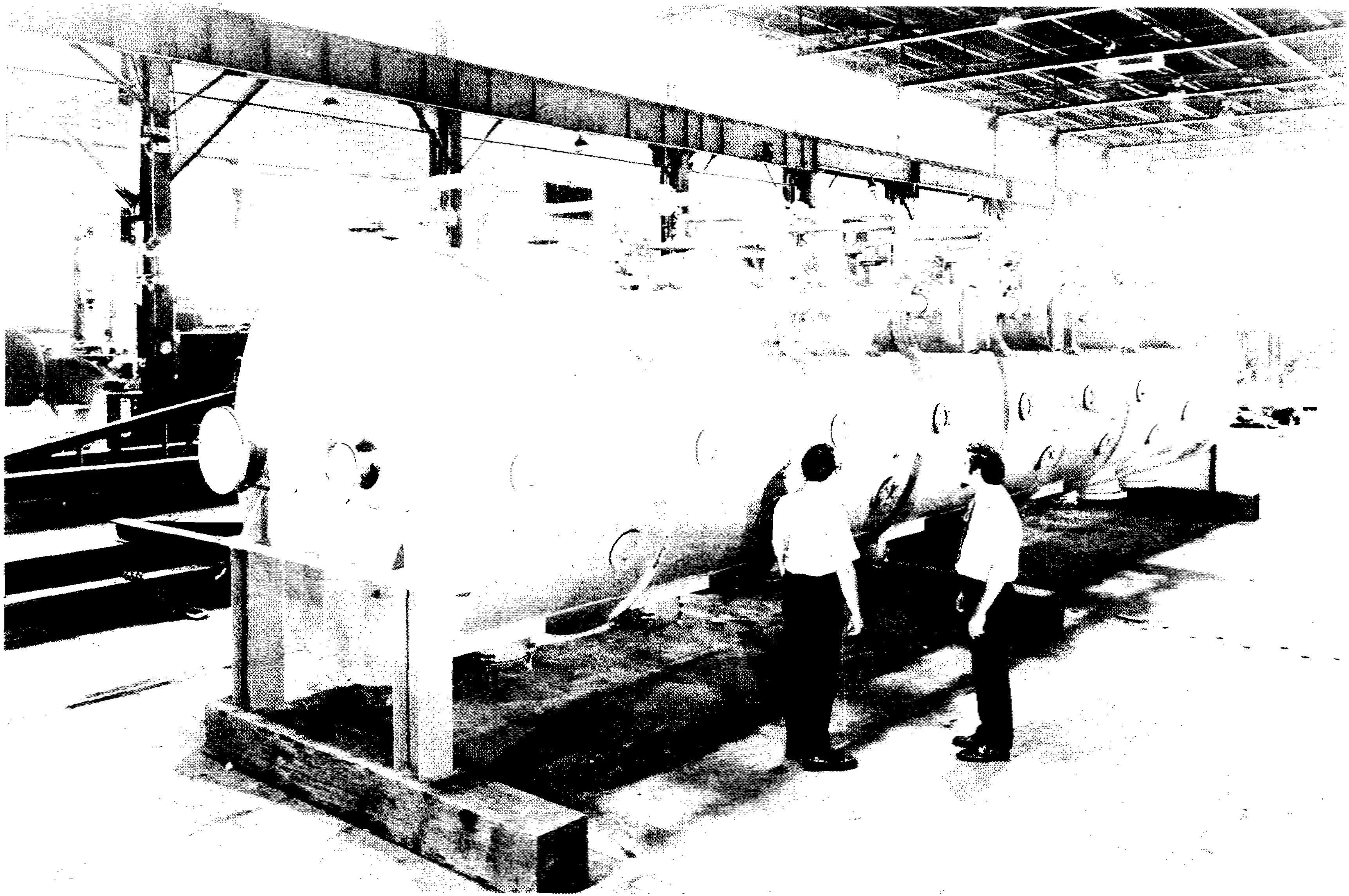


Photo Courtesy of Aqua-Chem, Inc.

FIGURE 3-9. Main unit of a multiple-effect, horizontal-tube, spray-film distillation plant.

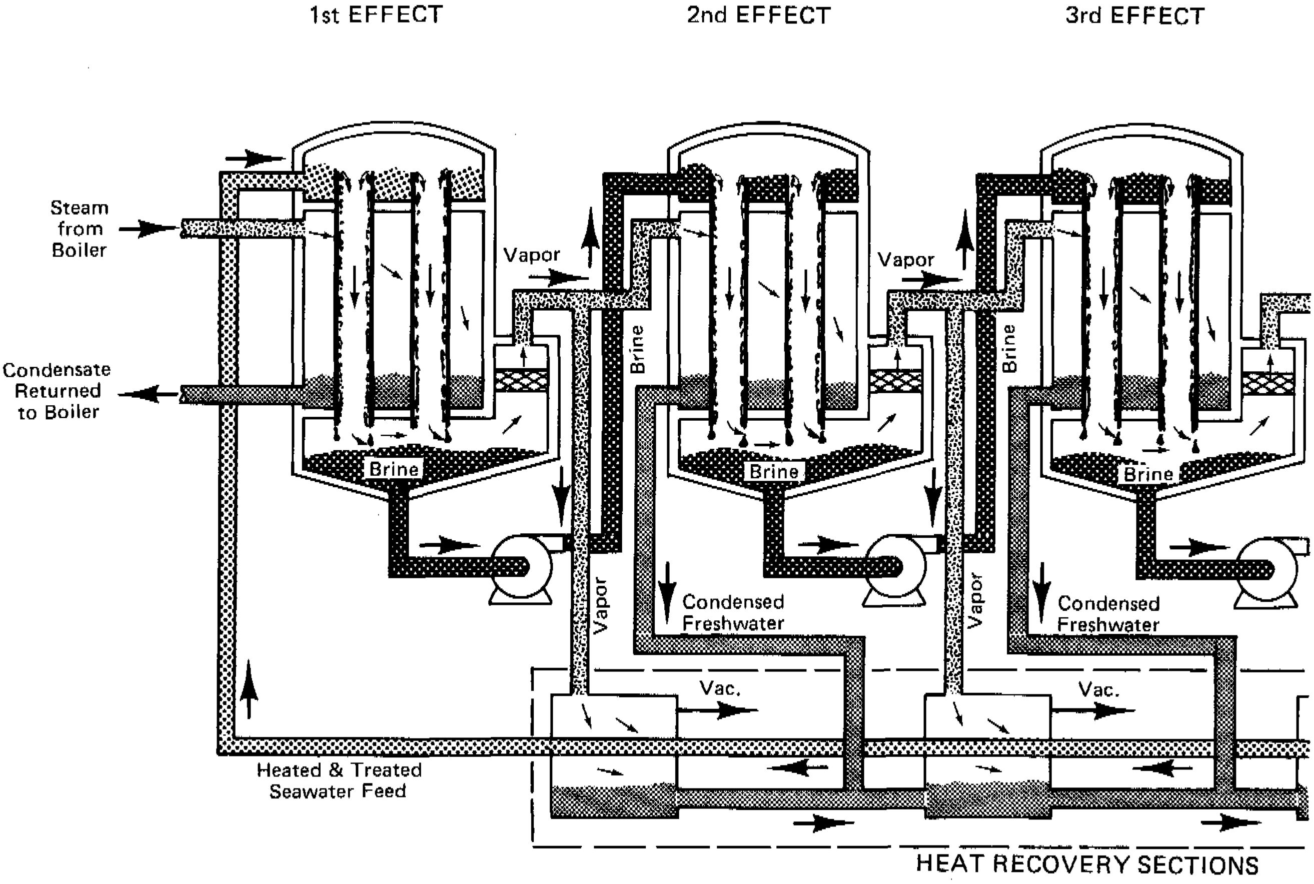


FIGURE 3-10. Simplified flow sheet for a multiple-effect vertical-tube evaporator (VTE).

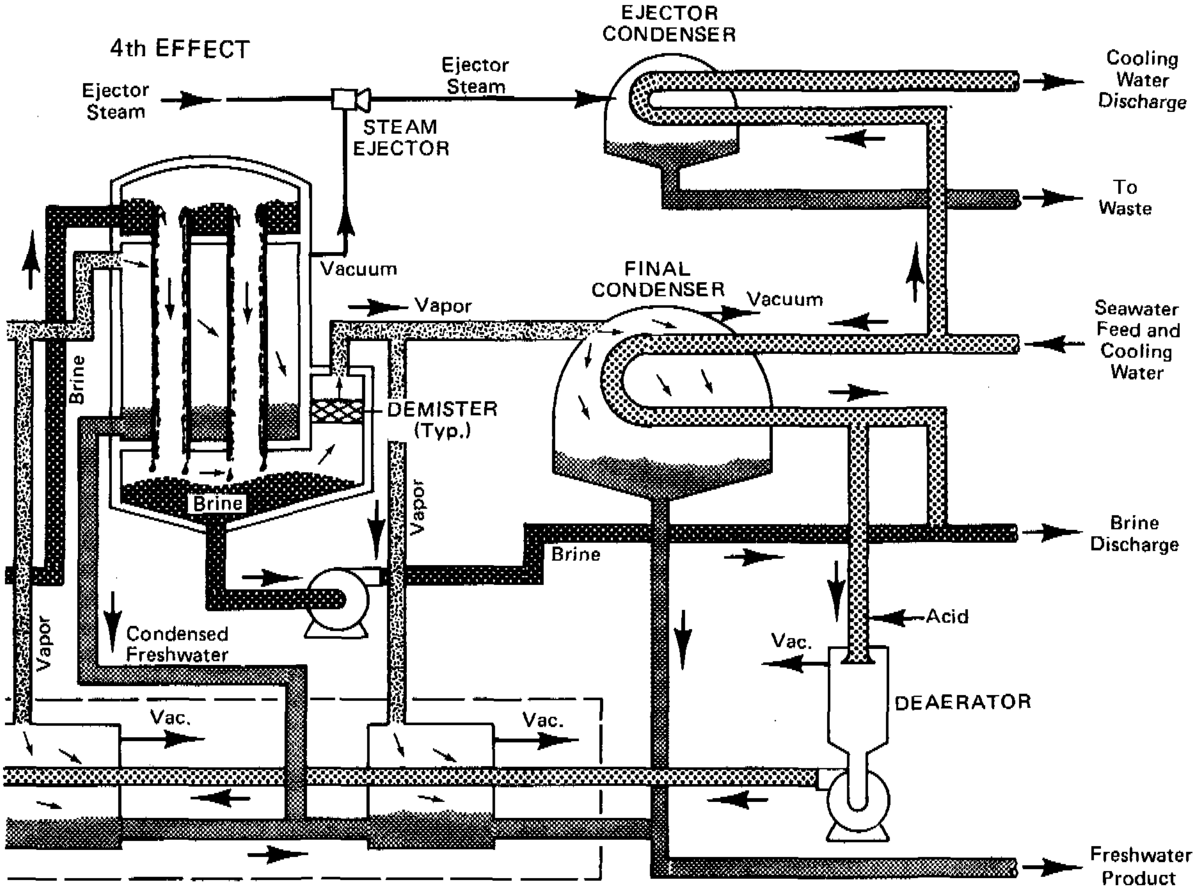


FIGURE 3-10. (Continued)

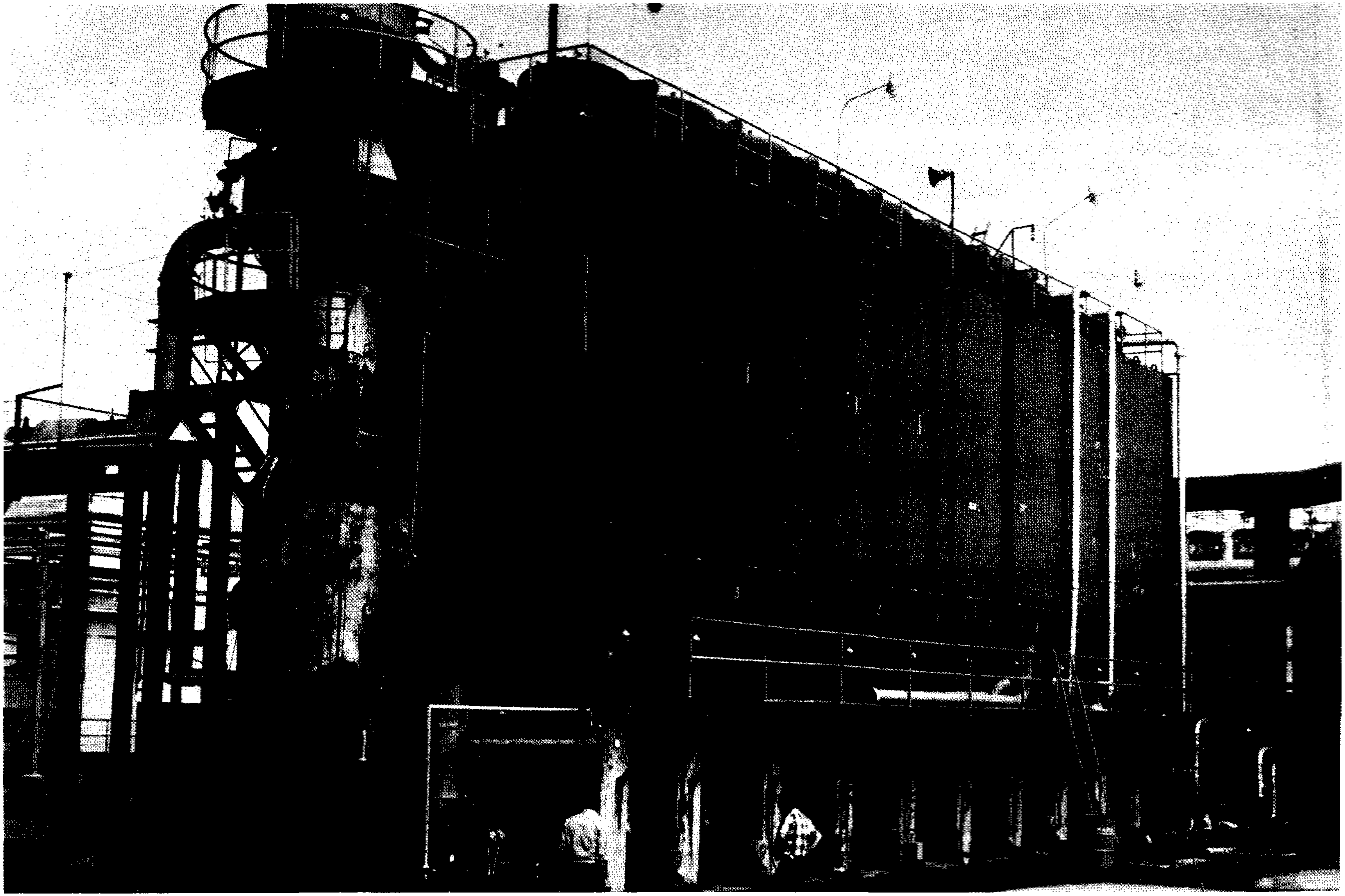
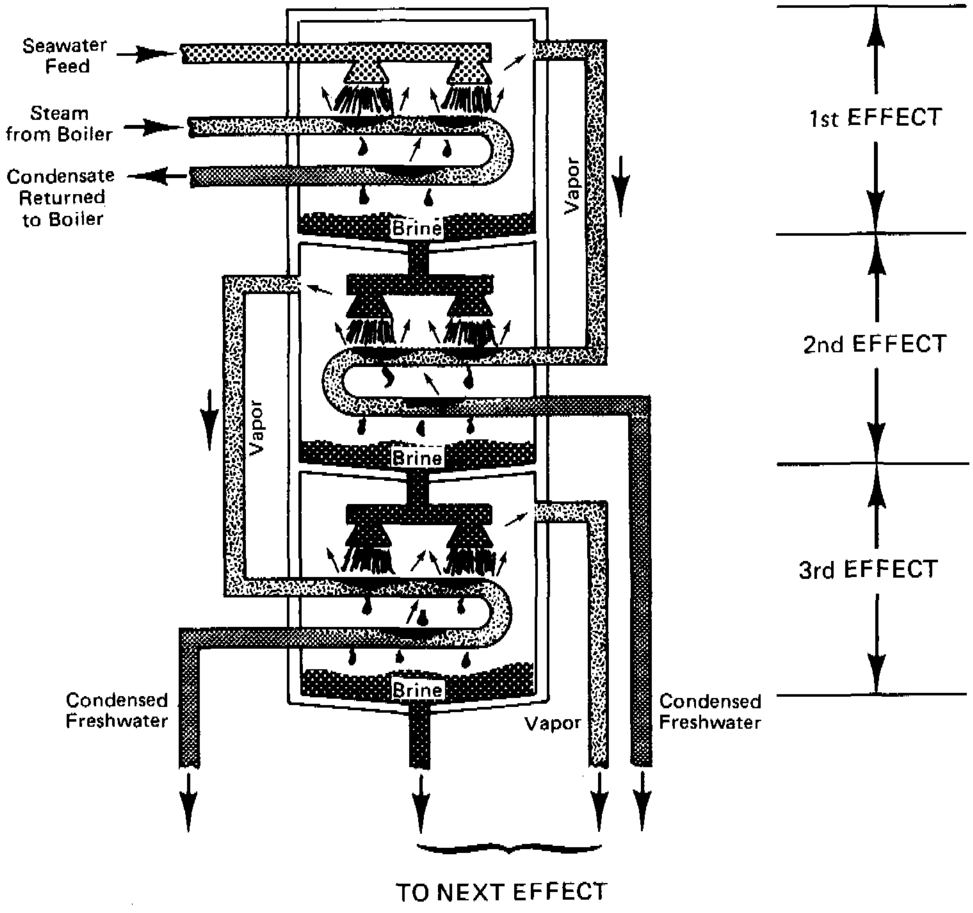


Photo Courtesy of O. K. Buros.

FIGURE 3-11. A 1-mgd [3,790-m^3/d] multiple-effect vertical-tube evaporator located on St. Croix in the U.S. Virgin Islands.



Notes:
1. This drawing is greatly simplified.
2. A final condenser such as shown on Figure 3-10 is necessary for operation.

FIGURE 3-12. Conceptual diagram of a horizontal-tube multiple-effect (HTME) distillation plant with vertically stacked effects.

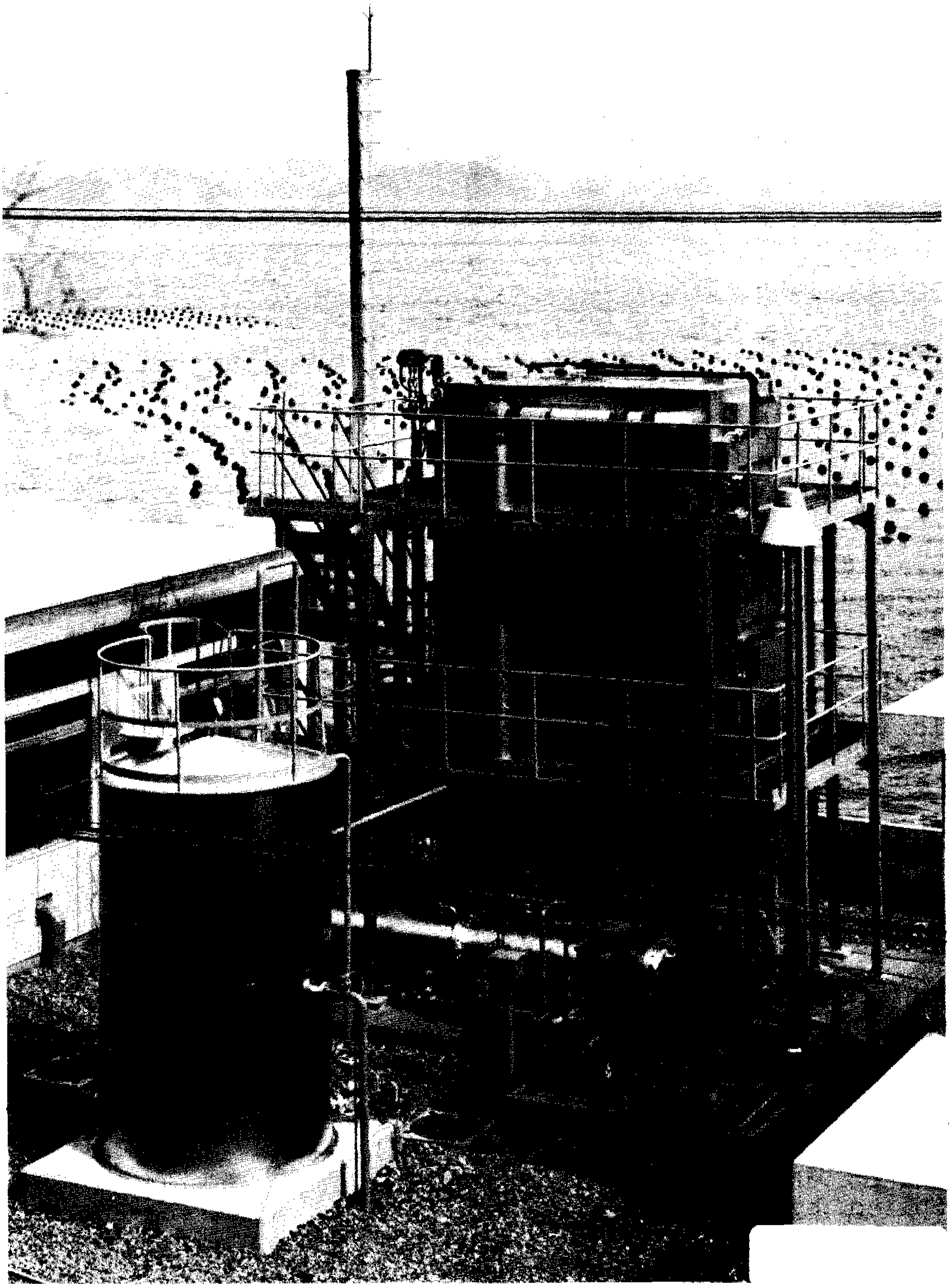
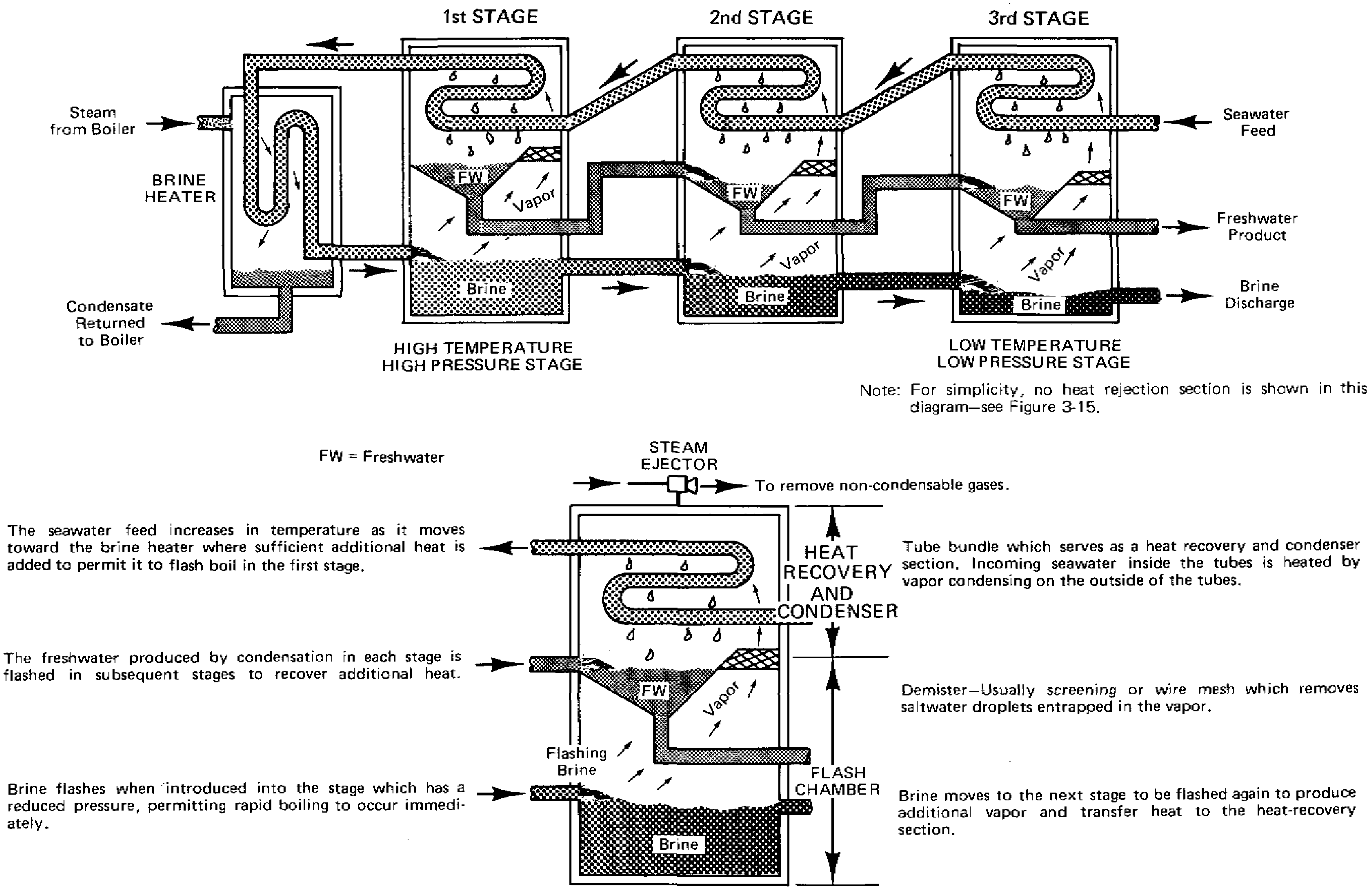


Photo Courtesy of Sasakura International American Corp.

FIGURE 3-13. A vertically stacked, horizontal-tube, multiple-effect distillation plant in Japan.



Note: For simplicity, no heat rejection section is shown in this diagram—see Figure 3-15.

To remove non-condensable gases.

The seawater feed increases in temperature as it moves toward the brine heater where sufficient additional heat is added to permit it to flash boil in the first stage.

The freshwater produced by condensation in each stage is flashed in subsequent stages to recover additional heat.

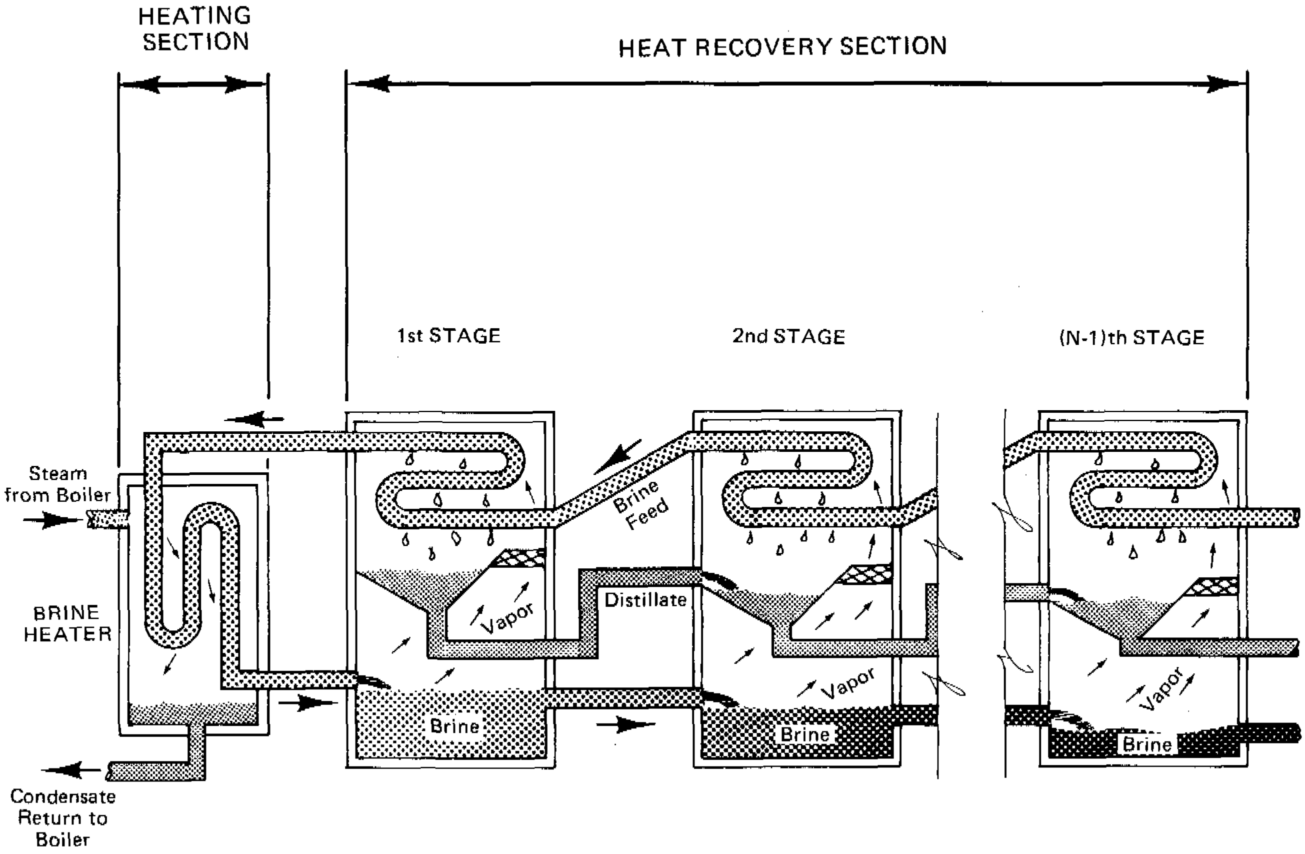
Brine flashes when introduced into the stage which has a reduced pressure, permitting rapid boiling to occur immediately.

Tube bundle which serves as a heat recovery and condenser section. Incoming seawater inside the tubes is heated by vapor condensing on the outside of the tubes.

Demister—Usually screening or wire mesh which removes saltwater droplets entrapped in the vapor.

Brine moves to the next stage to be flashed again to produce additional vapor and transfer heat to the heat-recovery section.

FIGURE 3-14. Conceptual diagram of the multistage flash (MSF) process.



Note: See Figure 3-17 for a once-through MSF plant.

FIGURE 3-15. Simplified flow sheet for multistage flash (MSF) recycle distillation plants.

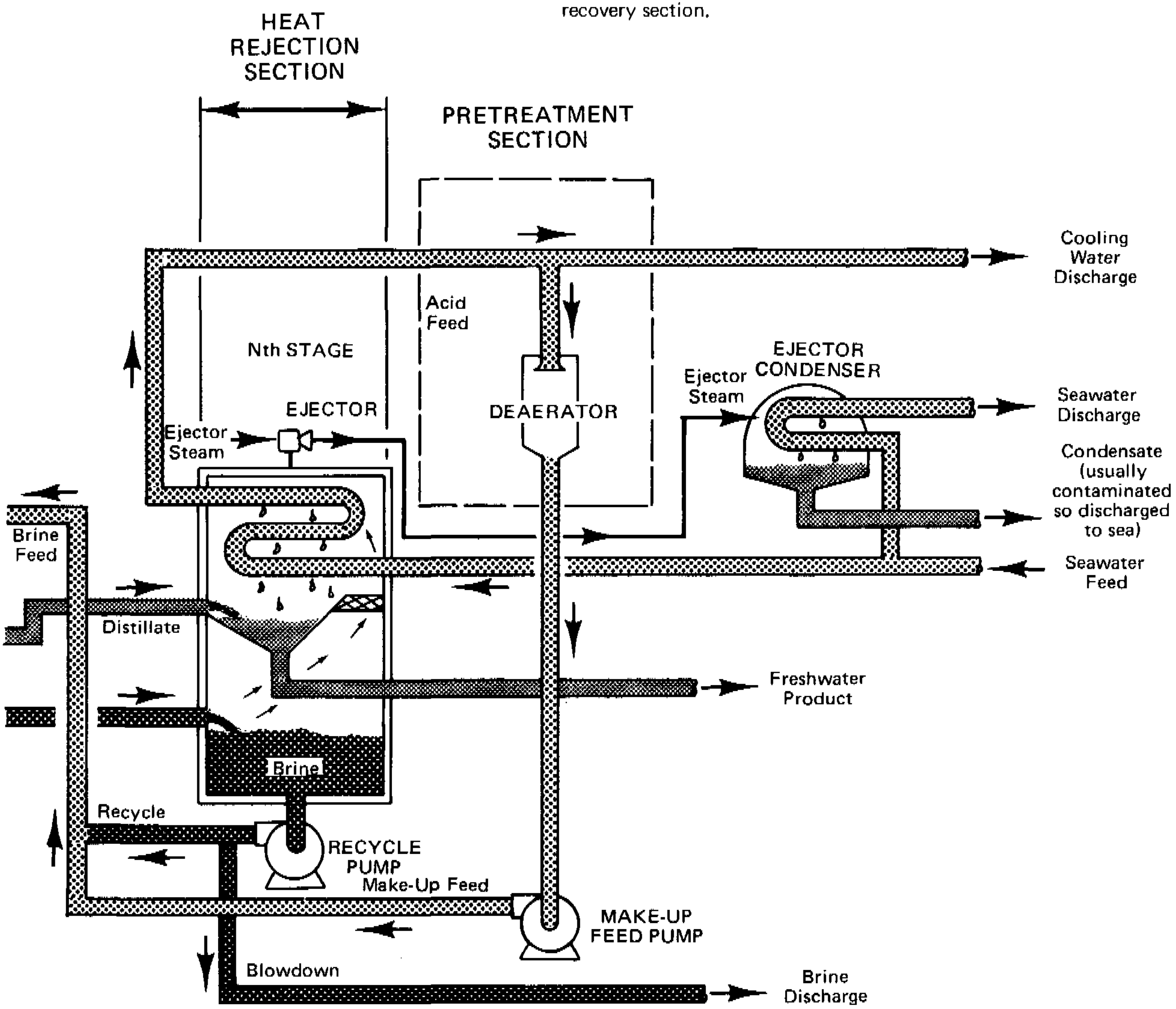

Note: Generally, there are several (often 4) stages in the heat recovery section.

FIGURE 3-15. (Continued)

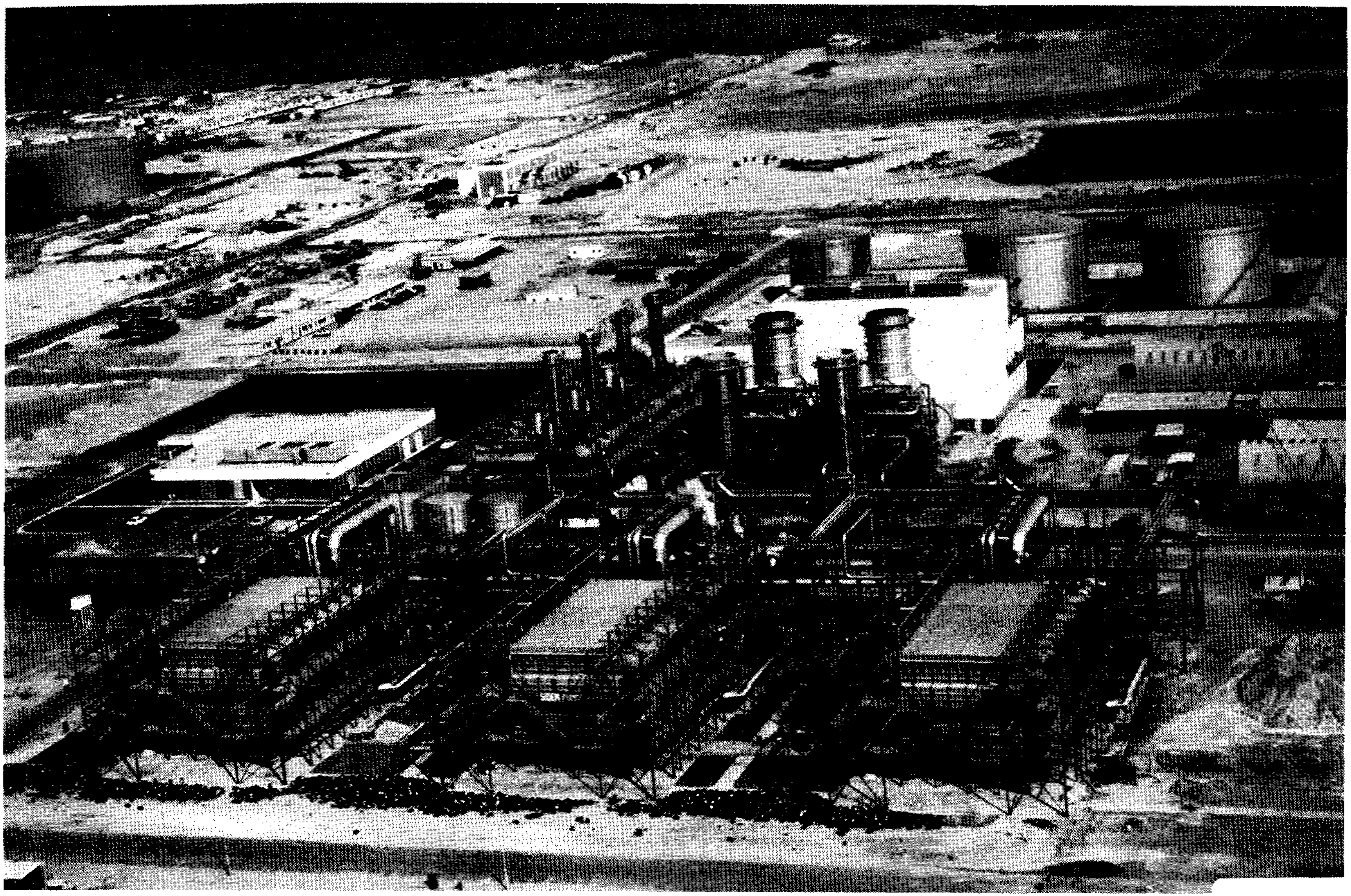
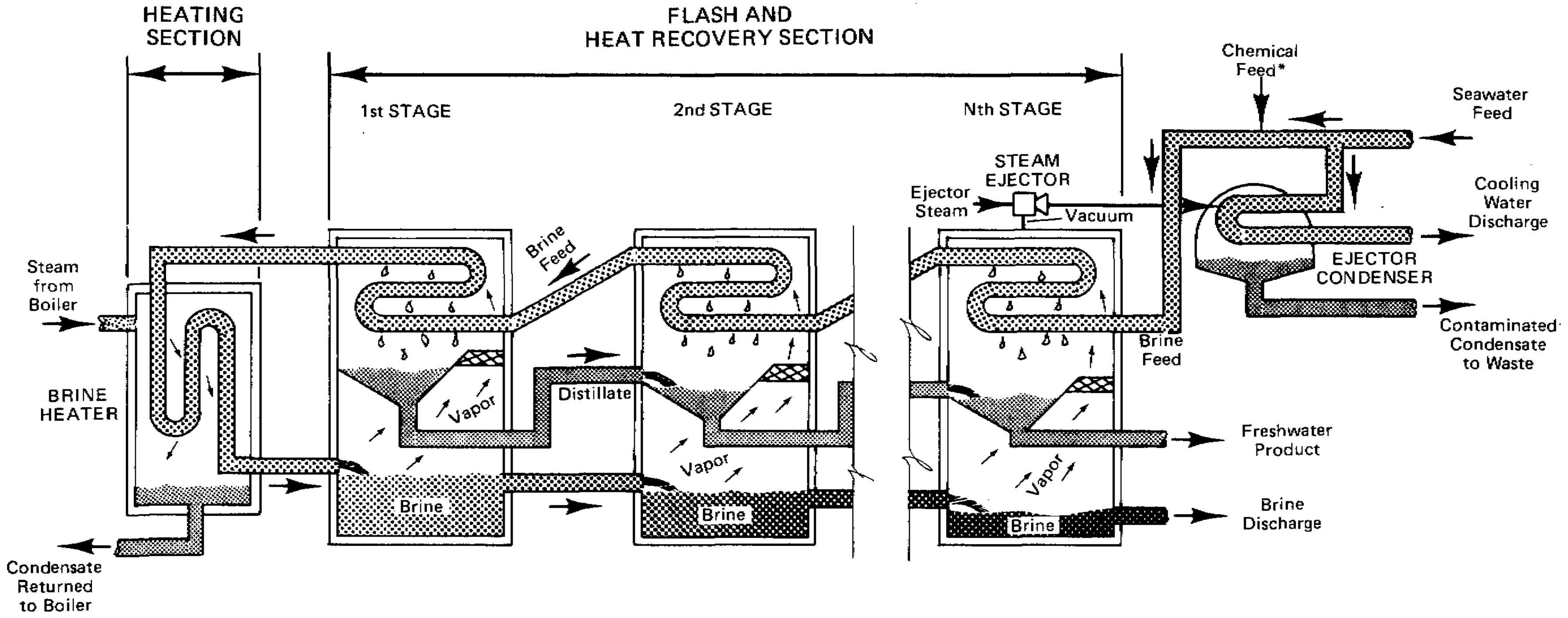


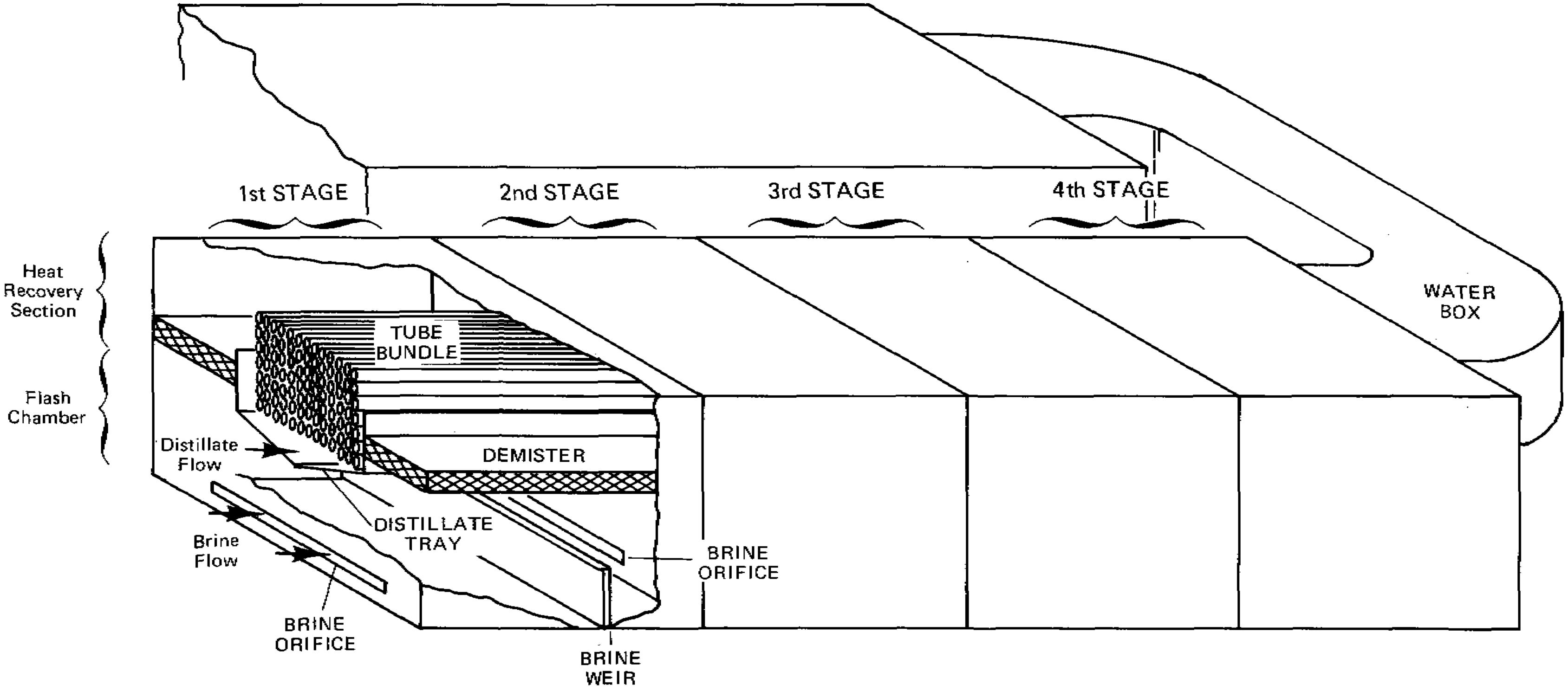
Photo Courtesy of Sidem, France.

FIGURE 3-16. Three 7.1-mgd [27,000-m^3/d] multistage flash distillation units in Abu Dhabi.



Note: See Figure 3-15 for an MSF with brine recycle.
*Polyphosphate up to 190°F [87.8°C]
Polyelectrolyte up to 250°F [121.1°C]

FIGURE 3-17. Simplified flow sheet for a once-through multistage flash (MSF) distillation plant.



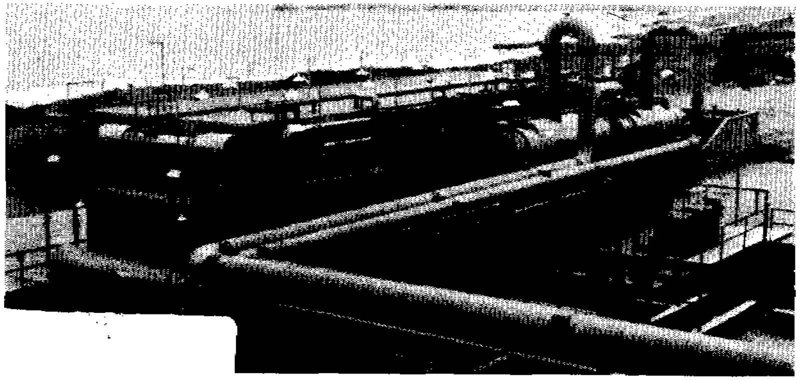
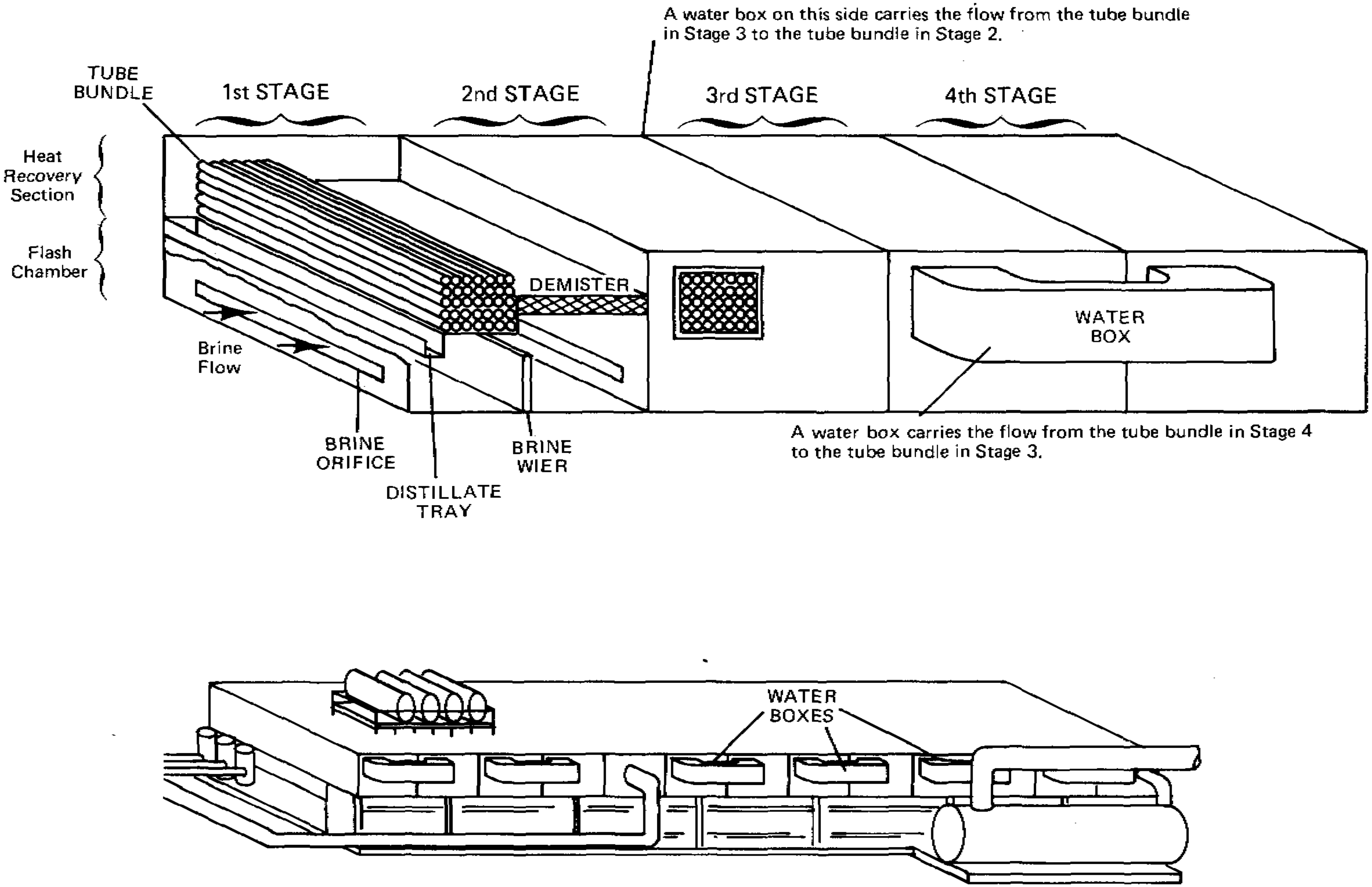


Photo Courtesy of Aqua-Chem, Inc.

A long-tube plant is identified by the location of the water boxes at the extreme ends of the first and final stages of the structure as shown in this 0.72-mgd (2,725-m^3/d) plant at Ilo, Peru.

FIGURE 3-18. Multistage flash (MSF) distillation plant with a long-tube configuration.

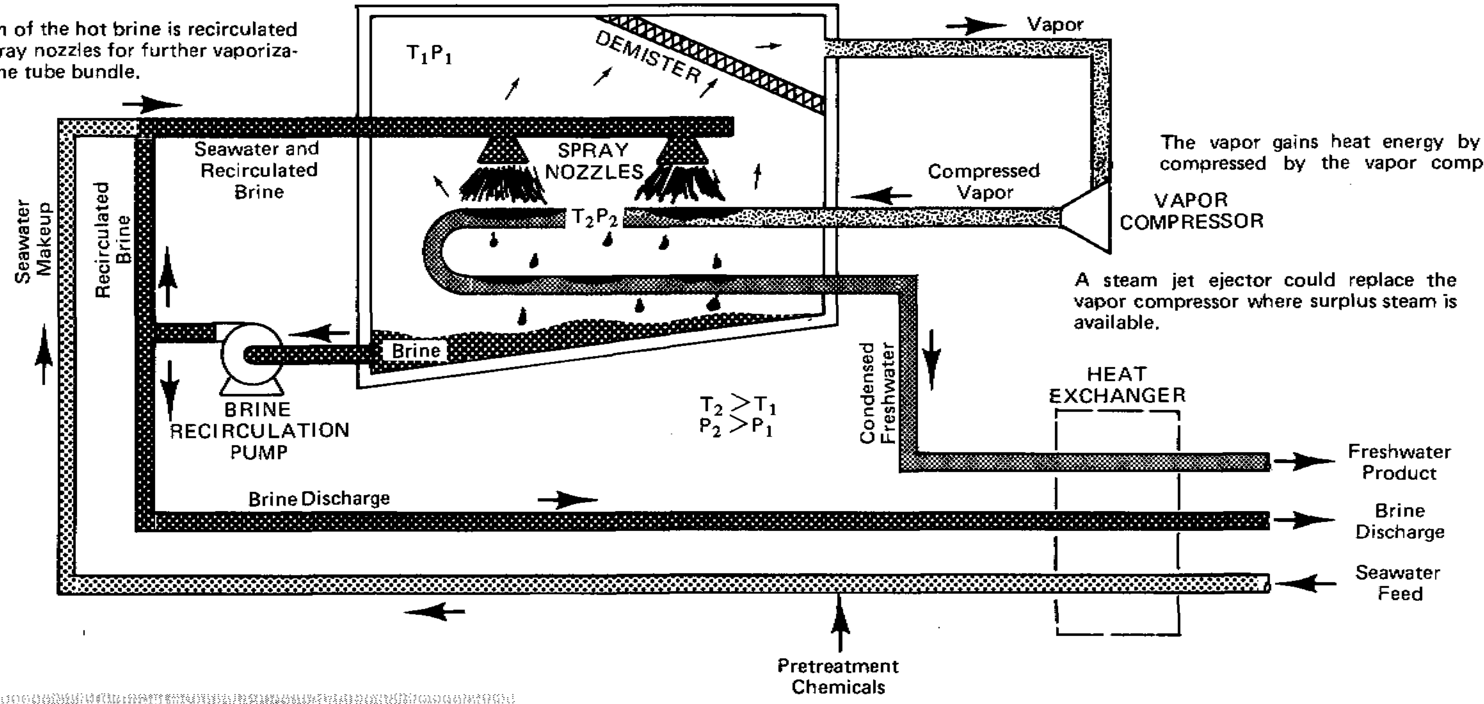


A water box on this side carries the flow from the tube bundle in Stage 3 to the tube bundle in Stage 2.

A water box carries the flow from the tube bundle in Stage 4 to the tube bundle in Stage 3.

A cross tube plant is readily indentified by the location of the water boxes on the sides of alternate stages of the structures.

FIGURE 3-19. A multistage flash MSF plant with a cross tube configuration.



A portion of the hot brine is recirculated to the spray nozzles for further vaporization on the tube bundle.

The vapor gains heat energy by being compressed by the vapor compressor.

A steam jet ejector could replace the vapor compressor where surplus steam is available.

Photo Courtesy of Aqua-Chem, Inc.

This type of electric-driven spray film vapor compression unit is used for facilities such as hotels, industrial plants, and power stations. It is generally available in capacities from 2,500 to 30,000 gpd [9.5 to 114 m^3/d]

FIGURE 3-20. Simplified flow diagram for a spray-film vapor-compression process.

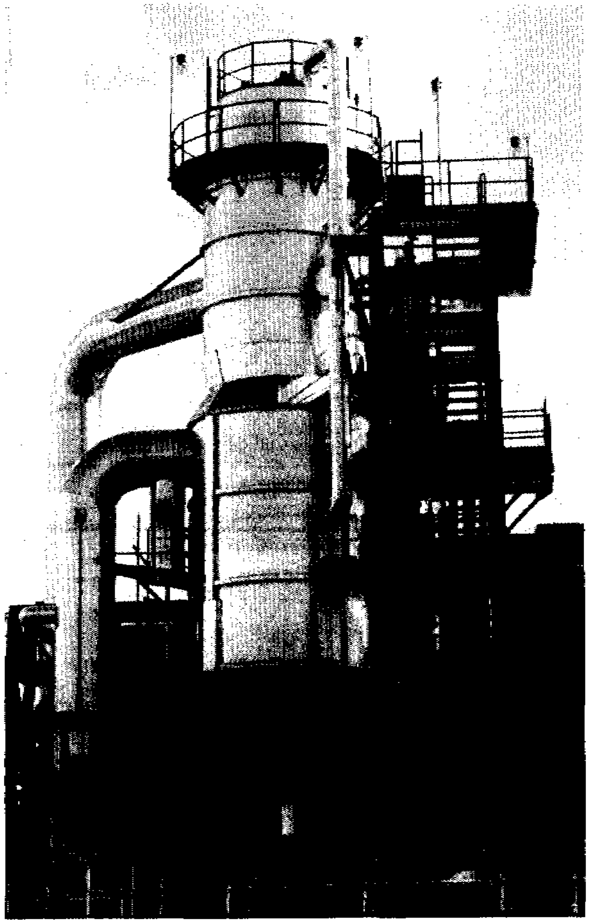


Photo Courtesy of RCC

A vertical-tube vapor compression unit to concentrate brine from a coal-fired generating station in southwestern USA. This unit processes about 0.225 mgd [852 m^3/d] of brine with a freshwater recovery of up to 98%.

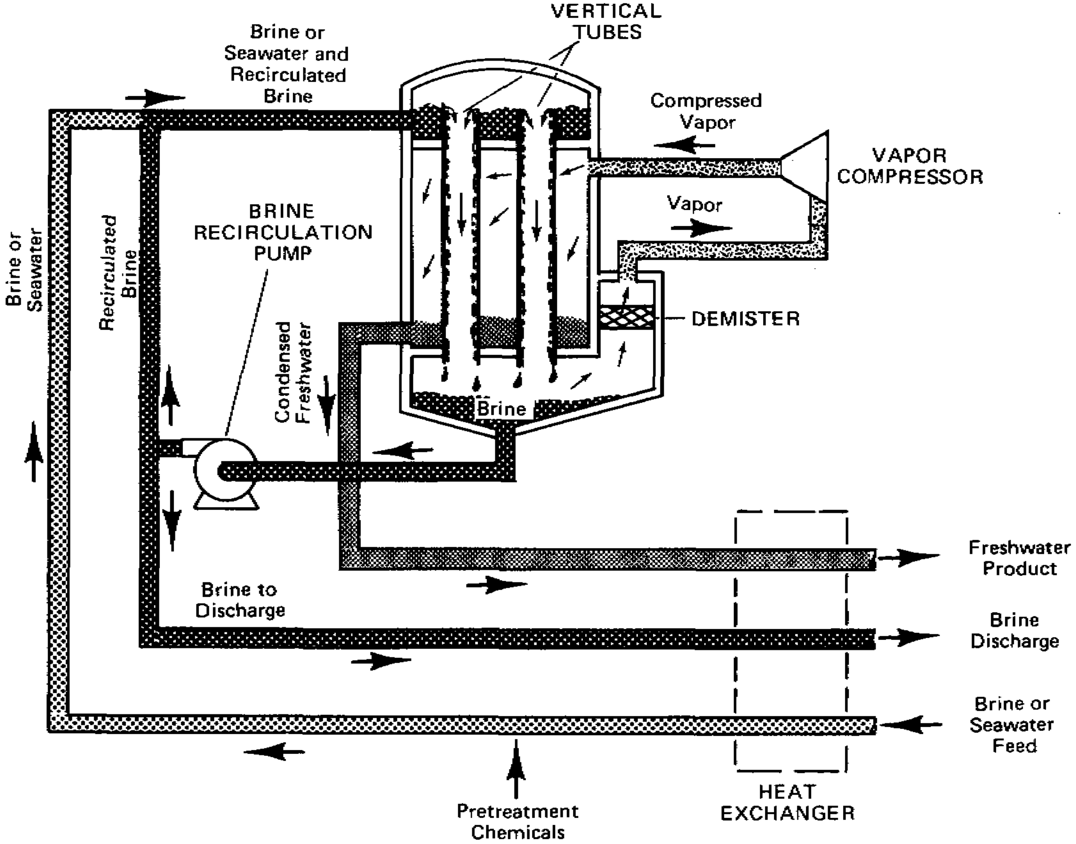


FIGURE 3-21. Simplified flow diagram for a vertical-tube vapor-compression process.

A 7,200-gpd [27-m^3/d] vertical-tube, upflow, vapor compression unit.

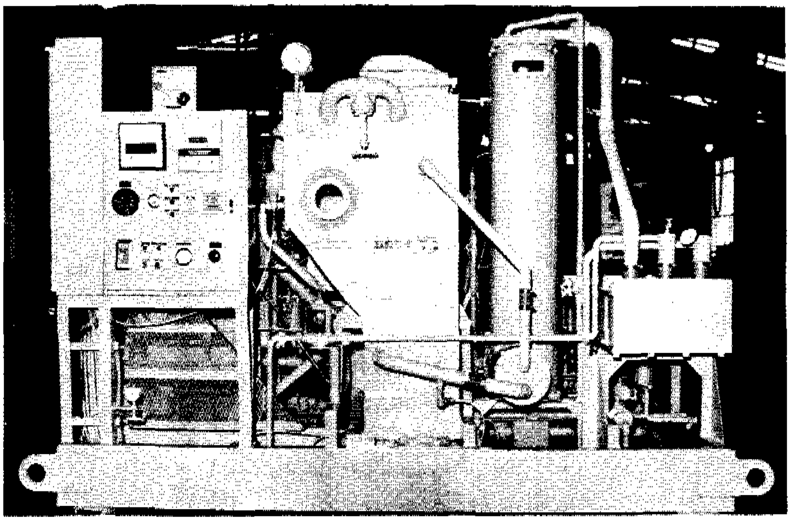


Photo Courtesy of Mechanical Equipment Co. (MECO).

A 92,400-gpd [350-m^3/d] steam-jet ejector vapor compression unit onsite in Jeddah, Kingdom of Saudi Arabia.

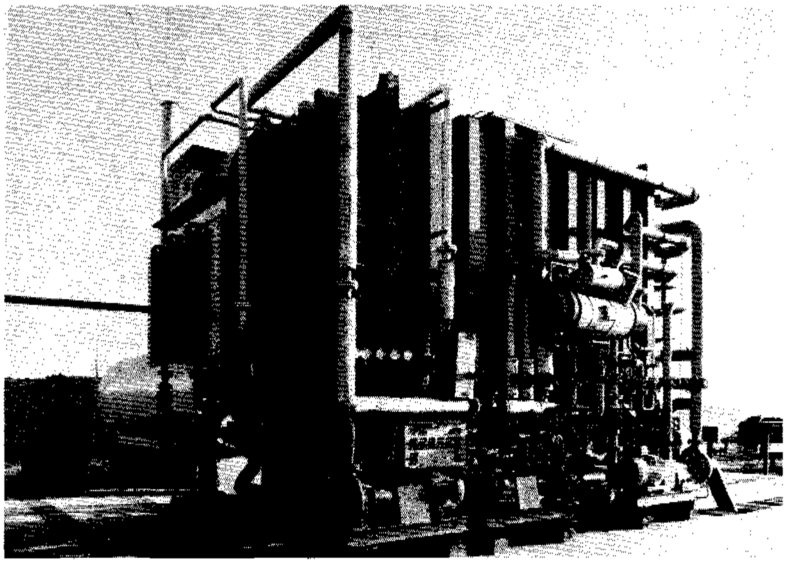


Photo Courtesy of Sidem, France.

A 130,000-gpd [500-m^3/d] horizontal-tube, low-temperature vapor compression unit. This unit has an operating temperature of about 126°F [52°C].

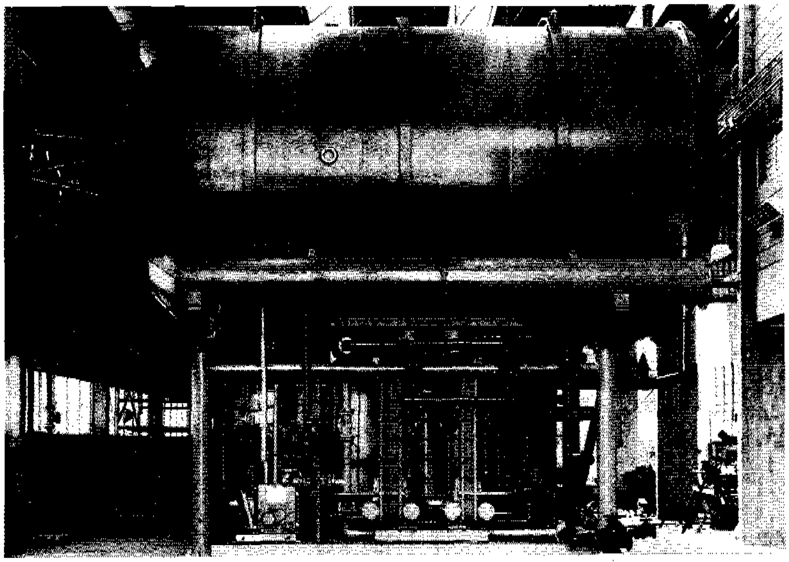
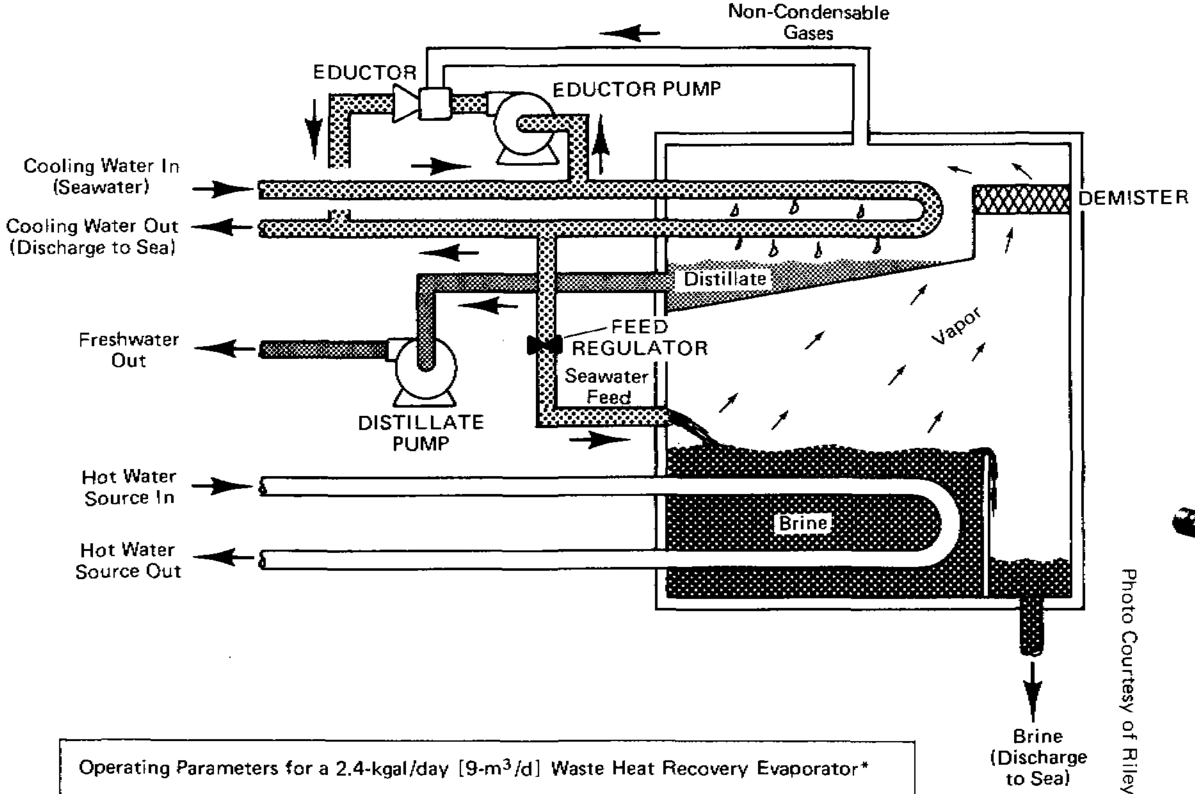


Photo Courtesy of Israel Desalination Engineering, Ltd., Israel.

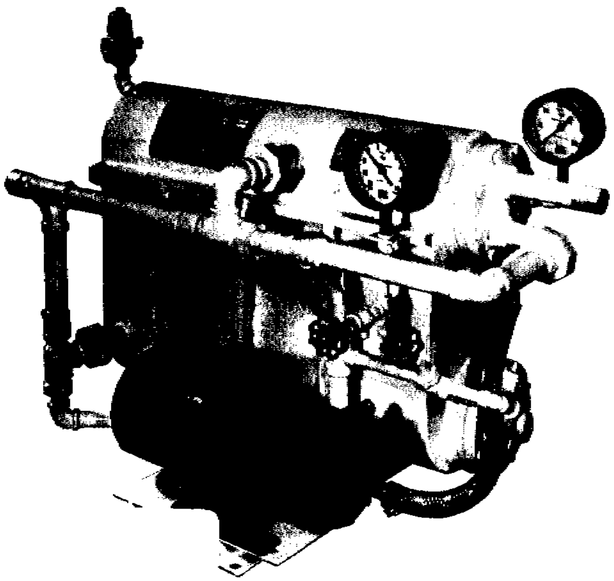
FIGURE 3-22. Various types of vapor compression units.



Operating Parameters for a 2.4-kgal/day [9-m^3/d] Waste Heat Recovery Evaporator*						
	Temperature		Flow		Power	
	(°F)	(°C)	(gpm)	(m^3/d)	(kWh/kgal)	(kWh/m^3)
Cooling water	85	29.3	53	289	--	--
Hot water	180	82.2	100	545	--	--
Feedwater	--	--	3.3	10	--	--
Freshwater	--	--	1.6	8.7	--	--
Electricity	--	--	--	--	15	4

*Data from Mechanical Equipment Company Bulletin No. 117C

Photo Courtesy of Riley-Beaird, Inc.



A small 480-gpd [1.8-m^3/d] heat recovery distillation unit.

FIGURE 3-23. A waste heat recovery evaporator.

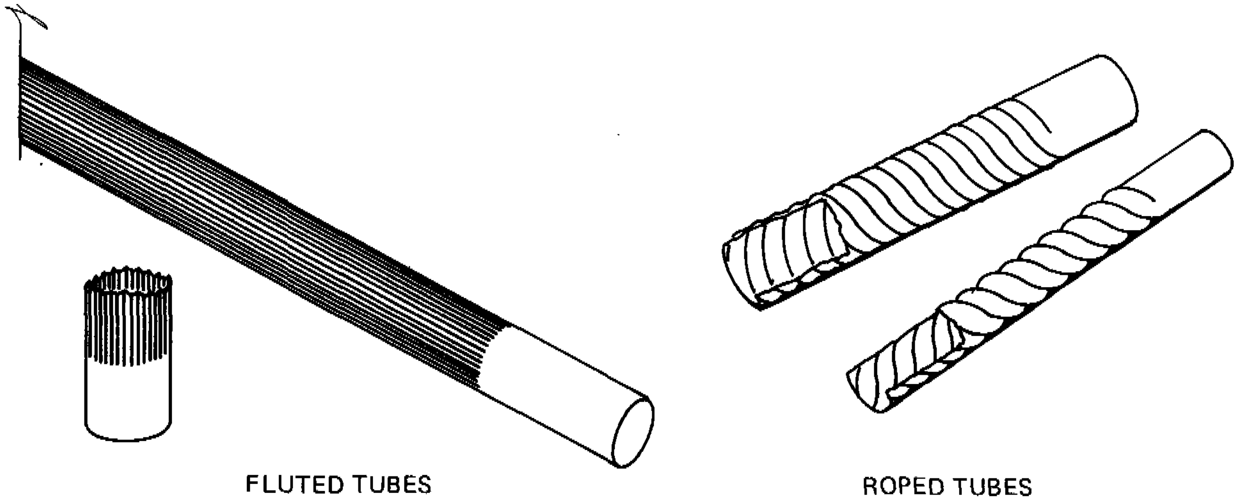
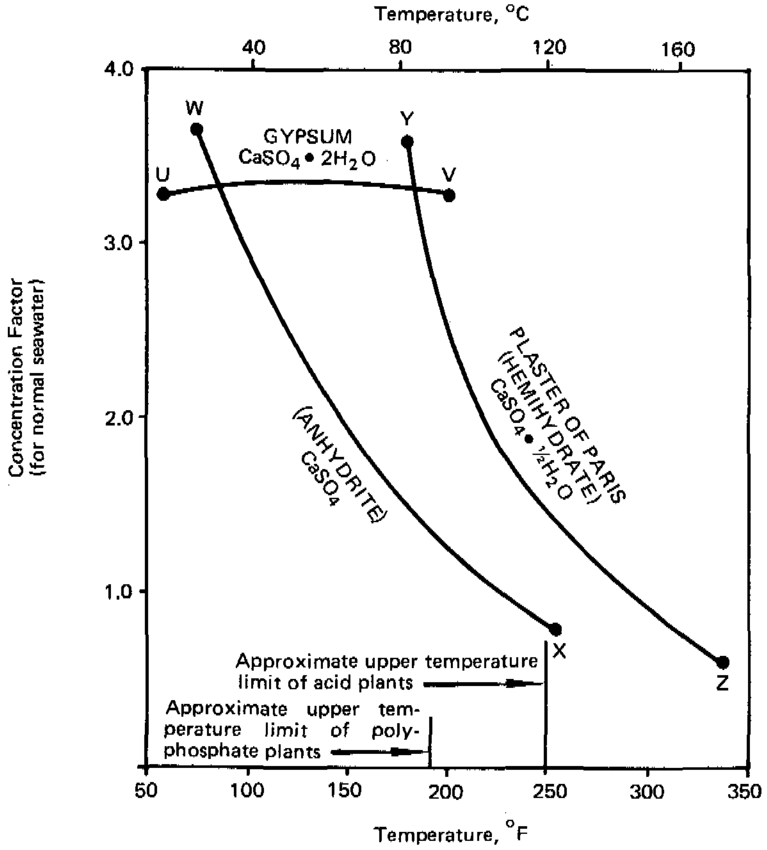


FIGURE 3-24. Fluted and roped tube.



Notes:
1. Solubility limits shown are approximate.
2. The area above curve UV represents supersaturation with respect to $CaSO_4 \cdot 2H_2O$.
3. The area above curve WX represents supersaturation with respect to $CaSO_4$.
4. The area above curve YZ represents supersaturation with respect to $CaSO_4 \cdot \frac{1}{2}H_2O$.
5. Adapted from Standiford and Sinek, 1961; Kellogg, 1975.

FIGURE 3-25. Approximate solubility limits for calcium sulfate in seawater.

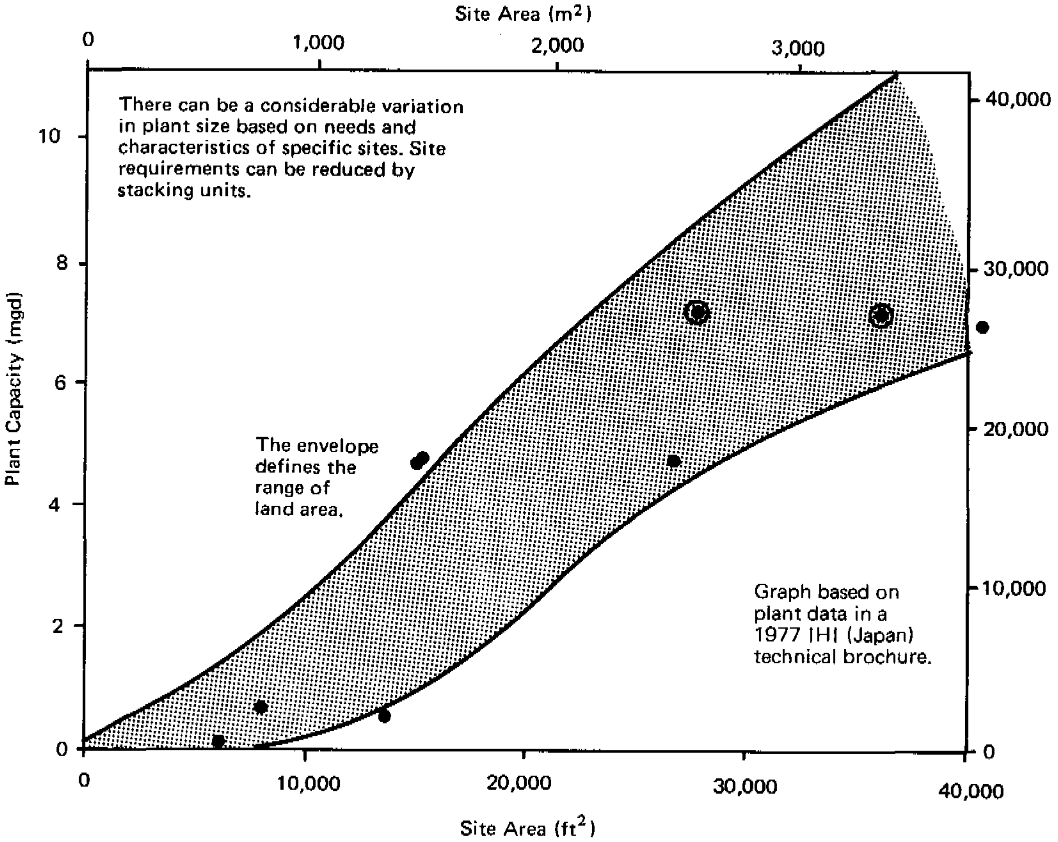


FIGURE 3-26. Approximate land area required for multistage flash (MSF) distillation plants.

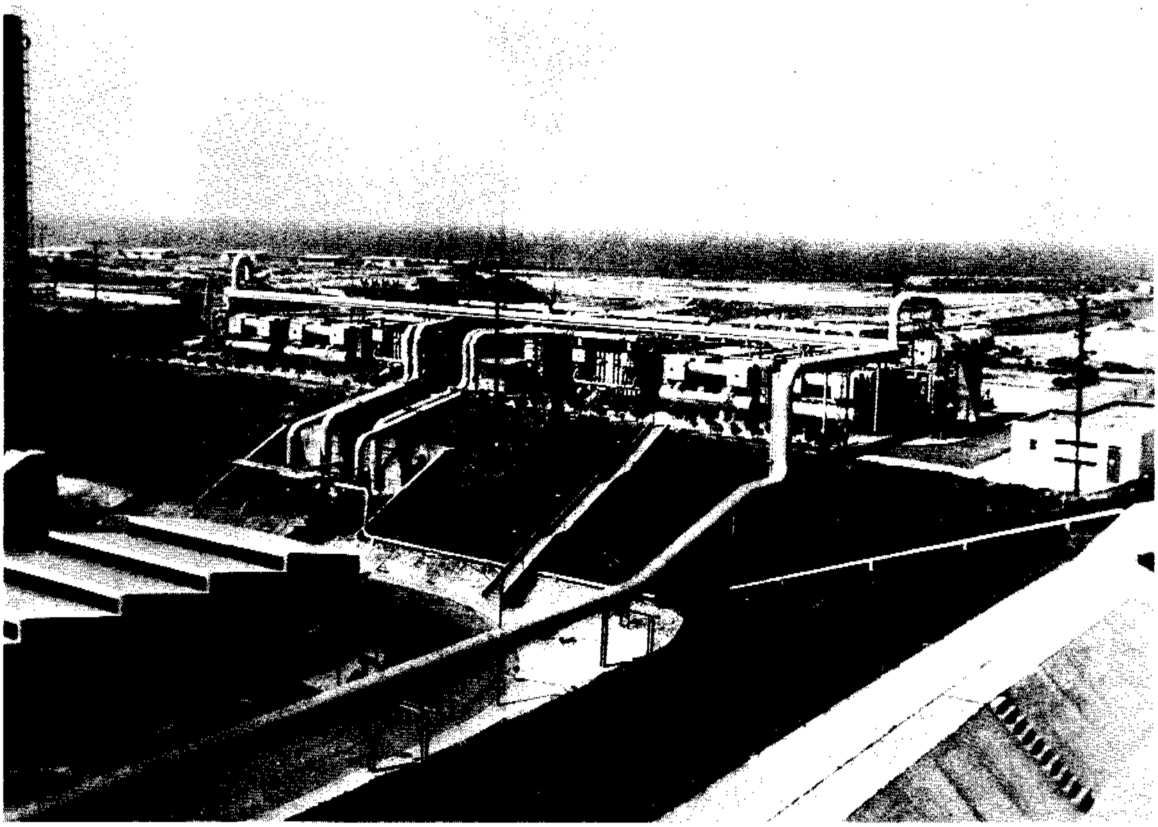
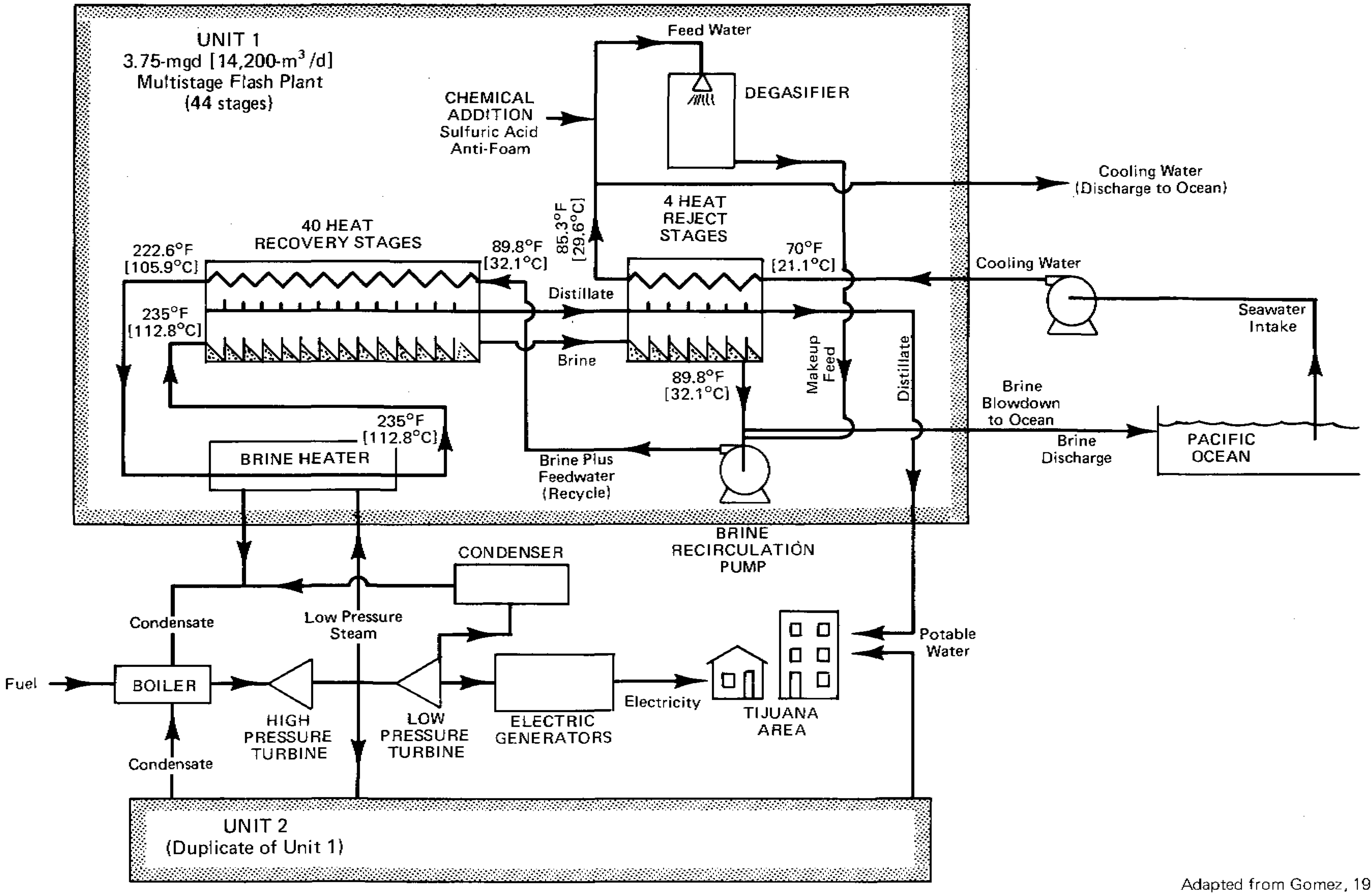


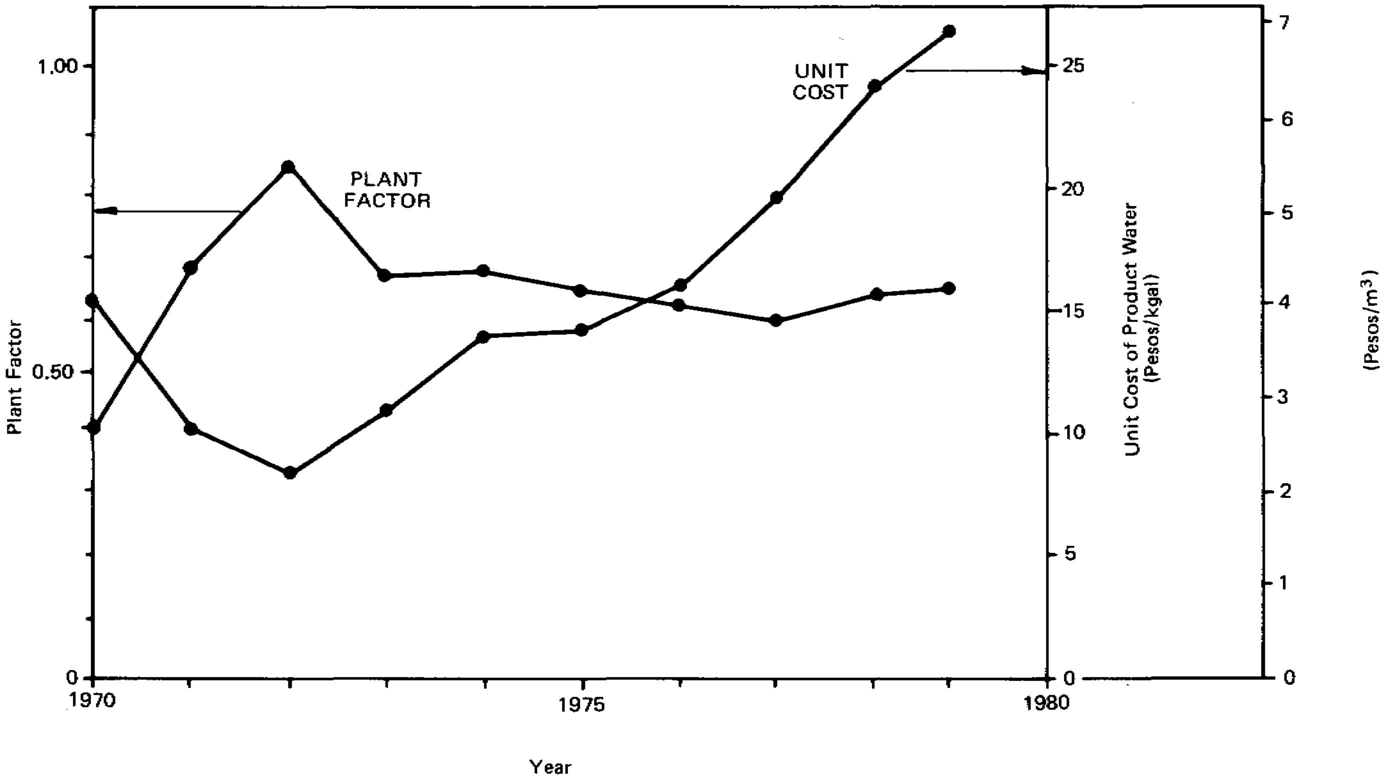
Photo Courtesy of Aqua-Chem, Inc.

FIGURE 3-27. The Federal Electric Commission's (CFE) 7.5-mgd [28,400-m³/d] multistage flash (MSF) facility at Rosarito, Mexico.



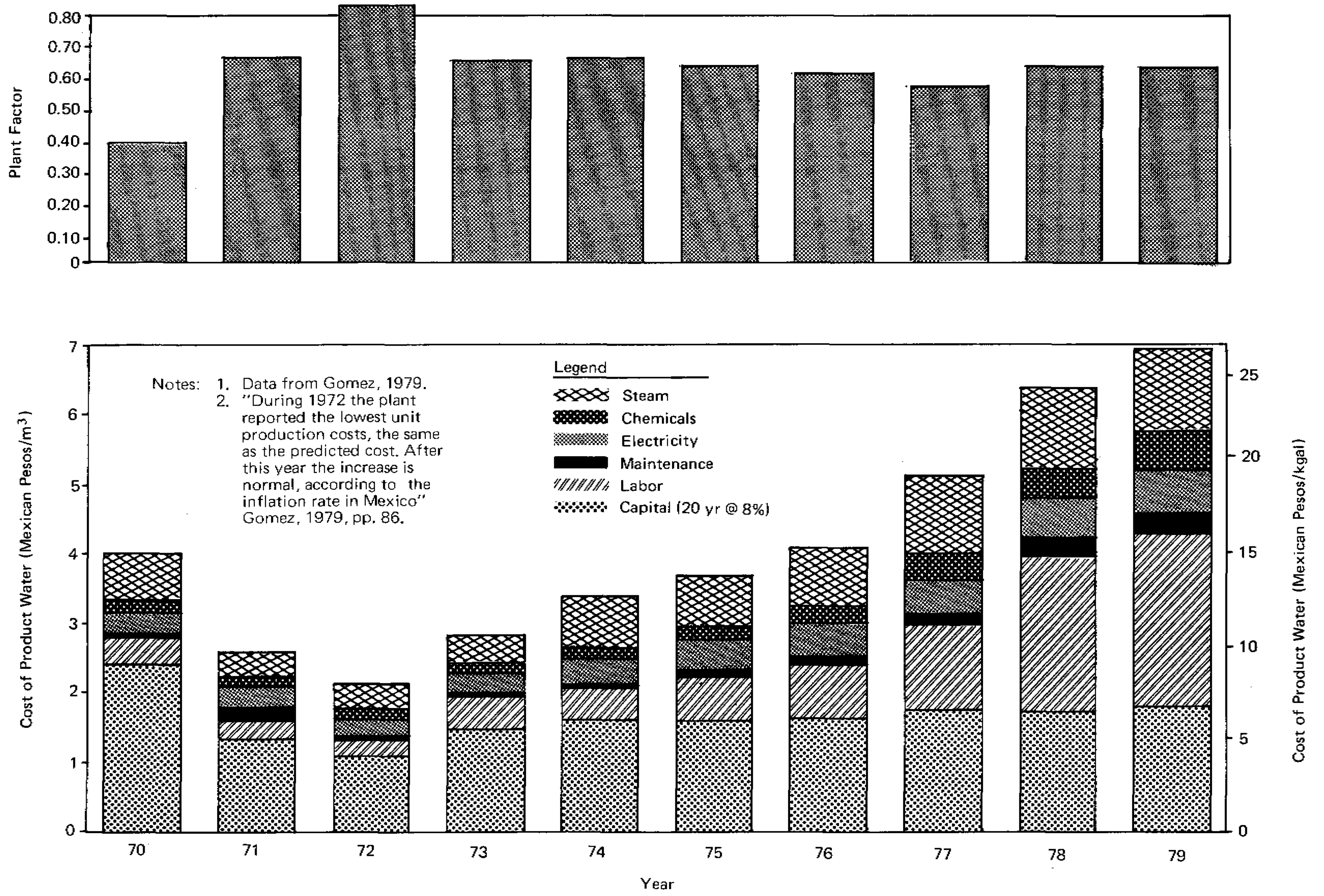
Adapted from Gomez, 1979.

FIGURE 3-28. Simplified flow diagram of the 7.5-mgd [28,400-m³/d] desalination facility at Rosarito, Mexico.



Notes:
1. Plant factor = Capacity factor x on-stream factor.
2. Capacity factor = $\frac{\text{Actual output capacity}}{\text{Design capacity}}$
3. On-stream factor = $\frac{\text{Days actually operated}}{\text{Total days in year}}$
4. Definitions— Watson, 1976
5. Data—Gomez, 1979.
6. Cost data in Mexican Pesos— Re-evaluation of the peso during this time period would unrealistically alter the data if converted to other currencies.
7. Data for 1979 are estimated by Gomez.

FIGURE 3-29. Unit cost of product water and plant factor data for the Rosarito, Mexico, desalination facility.



Notes:
1. Data from Gomez, 1979.
2. "During 1972 the plant reported the lowest unit production costs, the same as the predicted cost. After this year the increase is normal, according to the inflation rate in Mexico" Gomez, 1979, pp. 86.

FIGURE 3-30. Breakdown of the unit cost for desalted product water at the Rosarito, Mexico, desalination facility.

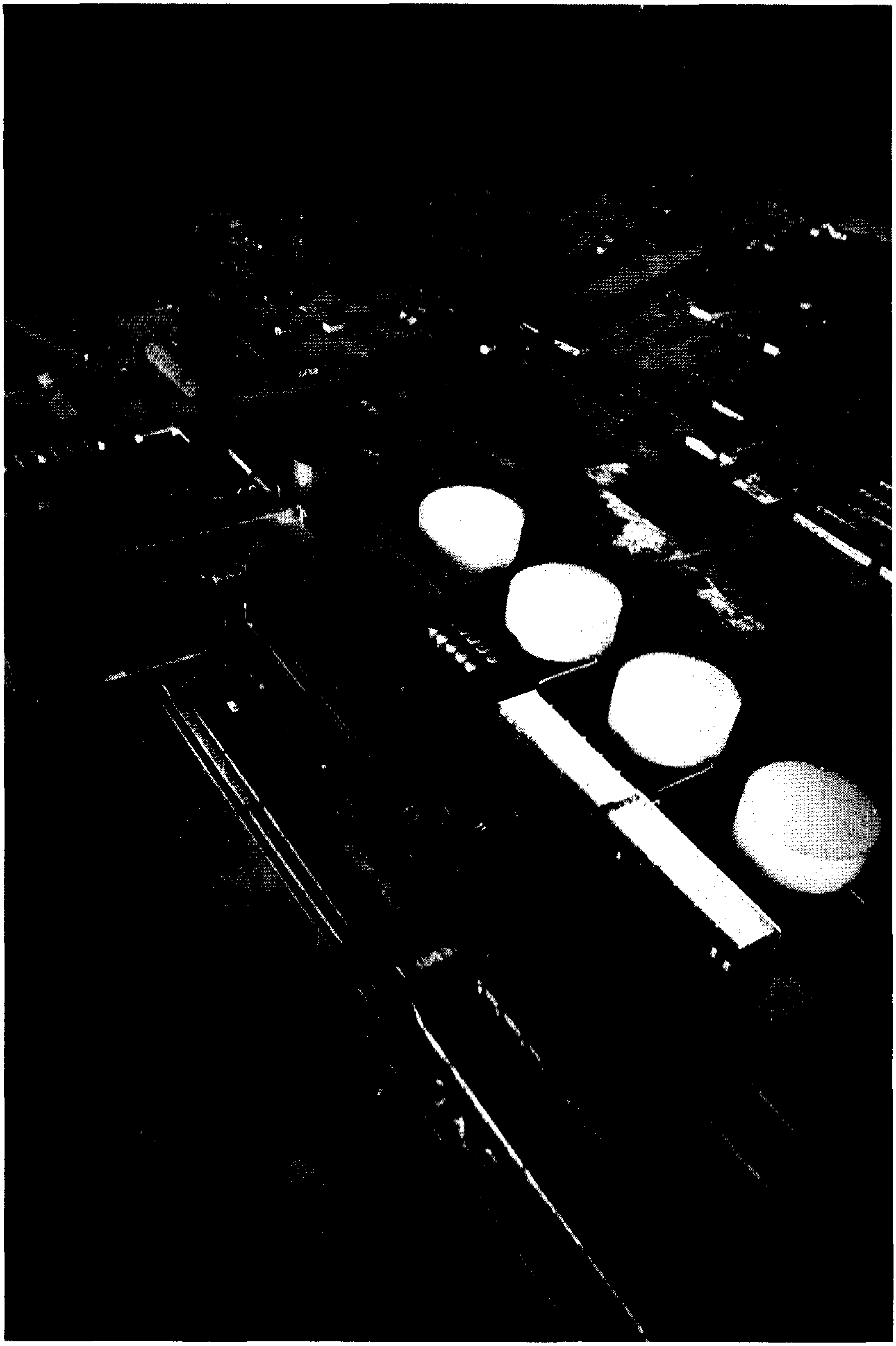
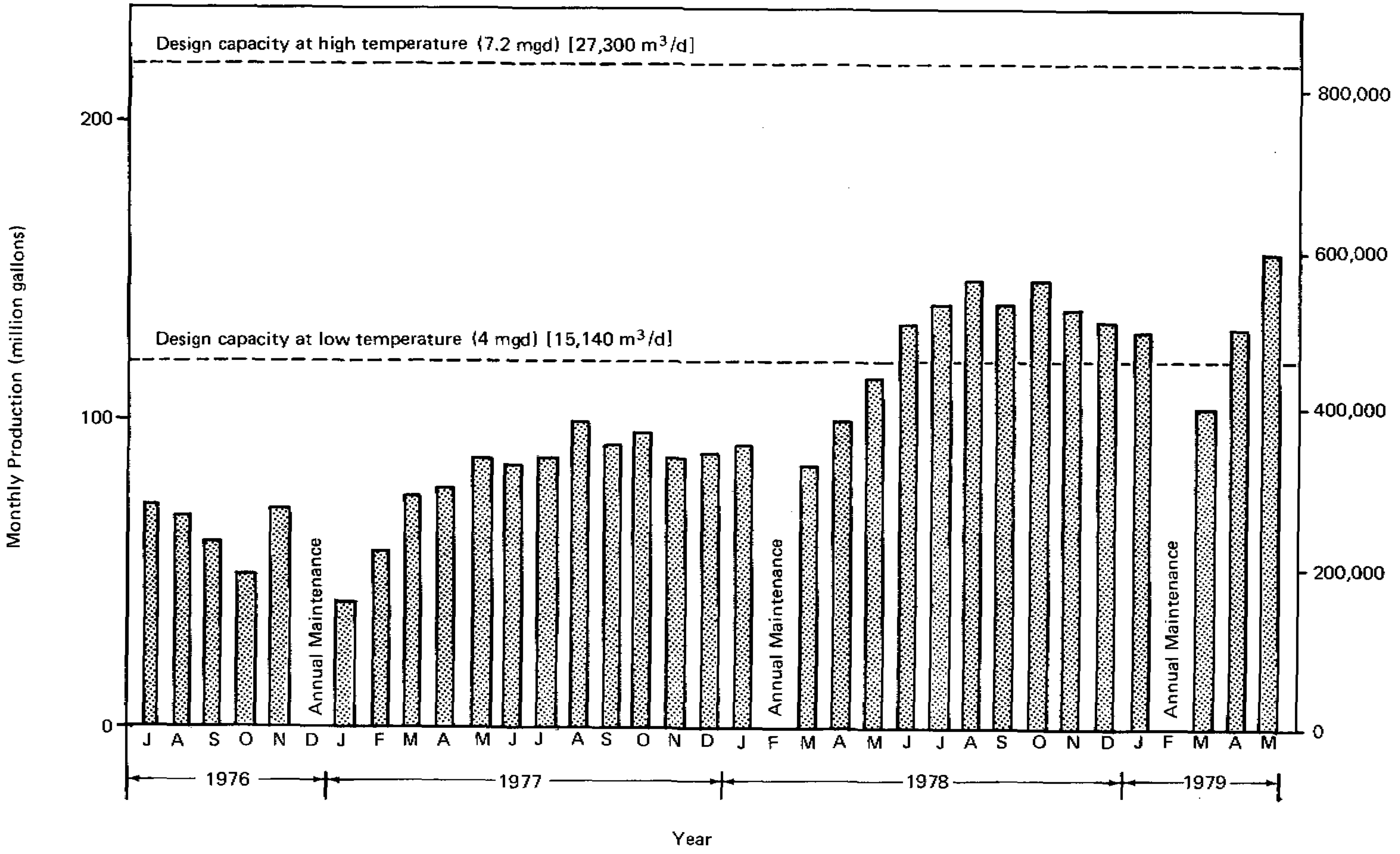


Photo Courtesy of Envirogenics (USA)

FIGURE 3-31. The 7.2-mgd [27,300-m^3/d] multistage flash (MSF) desalination facility at Ghubrah, Oman.



Notes: 1. Data adapted from Chalchal, 1979.
2. Level of plant operation was based on demand for water, not plant capacity.

FIGURE 3-32. Monthly production at the 7.2-mgd [27,300-m³/d] multistage flash (MSF) plant at Ghubrah, Oman.

CHAPTER 4

Chapter 4
FREEZING

THEORY

Introduction

The freezing process is based on the natural phenomenon which occurs when ice forms in a saline solution; the resulting individual ice crystals are made up of essentially pure water. In fact, it is the nature of all crystals to exclude impurities from their structure as they grow.

Although efforts have been undertaken, in the form of studies, pilot plant units, etc., since the 1950's to produce a commercial seawater desalination plant based on the freezing principle, the efforts have not been commercially successful. Only one commercial plant was sold in the western hemisphere, and this was to a condominium complex on the Caribbean island of St. Croix, in about 1967. For various reasons, including the financial problems of the purchaser, this installation was not a success, and the manufacturer had to reclaim the plant.

The primary development of the freezing process took place in the 1960's, when the only competitive process for the desalination of seawater was distillation. Compared to distillation, freezing theoretically has a number of advantages. It requires the transfer of less energy, needs almost no pretreatment, and has minimal corrosion and metallurgical problems. However, some of these advantages diminish when compared to seawater reverse osmosis.

Although freezing desalination is currently not a success, there is still potential for its development and use, especially in the desalination of difficult brackish waters and the concentration of brine streams from other types of inland desalting plants.

Fundamental Principles

Heat of Crystallization/Fusion. Ice is the solid form of water, and the phase change that takes place when water is transformed from a liquid to a solid requires the removal or release of energy. This heat of solidification, or crystallization, is equal to 144 Btu's per pound [335 kJ/kg] of ice. The temperature of the water must first be lowered to its freezing point (32°F [0°C]) for pure water and then an additional 144 Btu's [152 kJ] must be given up or removed without temperature change) for each pound of ice formed. Conversely, in order to melt ice which is at the freezing point, it is necessary to add 144 Btu's [152 kJ] per pound of ice to be melted. This heat required for melting is referred to as the heat of fusion. This heat, which is required to promote a phase change at the freezing point, is less than one-seventh of the heat necessary to transform water to vapor at the same temperature. This vast difference in energy required for phase changes for boiling (1,076 Btu's at 32°F [1,136 kJ at 0°C]) and freezing (144 Btu's [152 kJ]) has been one of the principal reasons for so much interest in the use of the freezing process for desalination.

Freezing Point. The freezing point for pure water at standard pressure (1 atmosphere) is 32°F [0°C]. As with the boiling point, the freezing point and vapor pressure for water change depending on the amount of solutes (e.g., salts) in the solution. Figures 4-1A and 4-1B illustrate these general relationships.

Crystal Formation. Ice is crystalline in nature (in contrast to amorphous solids, which have a completely random molecular arrangement) and, as water is cooled at the freezing point, individual crystals are formed. As the number of crystals in a solution increase, they tend to agglomerate unless kept separated by agitation or stirring.

Although water is very common in nature, the exact mechanism of ice nucleation is not known (Drost-Hansen, 1966). There appear to be two kinds of common crystal structures: dendrite and disc (see Figure 4-2). Crystal formation and growth are sensitive to the conditions of temperature, salt concentration, and agitation existing in the solution. The fragile dendritic crystals, unlike disc structures, tend to break up and act as nucleation sites for new crystal formation.

Brine Entrapment. The binding forces of ice crystals are such that only water molecules fit the crystal structure, and all other molecules, such as sodium chloride, are excluded. Evidence indicates that impurities found in partially frozen seawater are located on a layer adhering to the outside of the crystals and can be easily washed off. However, when the crystals form a larger and larger agglomeration, the brine layer on the individual crystals is trapped inside the clumps of ice and hence all of the impurities will not be readily removed by simple washing.

This natural purity of crystals during growth is the basis for the industrial crystallization of chemicals and compounds such as salt from seawater, potassium chloride from the Dead Sea, and silicon crystals for electronic circuitry.

Hydrates. Water combines with other substances to form crystals called hydrates. In this phenomenon a number of water molecules unite with a molecule of substances such as sodium carbonate ($Na_2CO_3 \cdot 10H_2O$) or copper sulfate ($CuSO_4 \cdot 5H_2O$) to form the hydrates. Each hydrate crystal has a definite proportion of water necessary for the formation of the crystal. As with ice crystals, the hydrate crystal formation excludes impurities in its structure.

Attempts have been made to use hydrates formed between water and organic substances such as Freon or propane as a method of seawater desalination. The hydrate crystals would be formed excluding salt impurities. Then the hydrate would be broken down by heat into its components of water and the unhydrated substances. These would be separated to produce pure water.

Unfortunately, although hydrate crystals have been formed, it has proved difficult to wash them to effectively remove the coating of impurities from their surfaces. Due to basic problems in the process related to crystal washing, hydrate pilot plants for the desalination of water generally have been unsuccessful (Johnson, 1976).

Basic Components

In all freezing desalination processes there are four basic components:

Freezer
Washer
Melter
Heat removal system

Figure 4-3 presents a block diagram of these components and their relationships in the freeze desalting process. The freezer, washer, melter, and heat removal system can be seen in the schematic diagram of a developed process on Figure 4-4.

Freezer. The freezer consists of a vessel called a crystallizer (or boiler) in which ice crystals and vapor are formed simultaneously. The remaining portion of the freezer is made up of equipment necessary to remove the heat from the brine in the crystallizer to promote freezing. The various freezing processes generally differ in the apparatus which removes heat from the brine in order to produce a slurry of ice crystals which can be easily removed, separated, washed, and then melted. The heat which is removed in the freezing process is usually transferred to the melter, where it is utilized for melting the ice.

Care must be taken in the operation of the freezer to produce a high proportion of discrete ice crystals rather than clumps of ice so that the amount of brine trapped between crystals is minimized. The size of the ice crystals formed is very important as fine crystals are difficult to wash, whereas very large crystals are expensive to produce because a large, costly crystallizer is required. Tests have shown that an agitated brine pool in the crystallizer which allows a retention time of 4 to 5 minutes will produce disc ice crystals that are 0.008 to 0.012 inches [0.2 to 0.3 mm] in diameter and are easy to wash (Carrier, 1964).

Washer. The ice is pumped as a slurry from the freezer to the washer, where the brine is drained from the slurry and the remaining ice is washed with freshwater to remove the brine adhering to the crystal surfaces. The efficiency and ease of the washing operation are especially sensitive to the size of the ice crystals.

The washer usually contains a rising rectangular or cylindrical column of porous ice to which the new ice from the freezer is added at the bottom of the column. The washwater is sprayed on the top and trickles down through the porous ice pack to drains on the sides. Washed ice is removed from the top of the column. Product water must be used for washing purposes, representing a loss in productivity for the system; hence the overall production efficiency is tied to the efficiency of the washer.

Melter. In the melter the ice from the washer is usually melted by transferring the heat of crystallization (about 144 Btu/lb [335 kJ/kg]) removed from the brine in the crystallizer to the melter. This is done by discharging the refrigerant (which removes the heat from the freezer) into the melter, where the ice picks up the heat (144 Btu/lb [335 kJ/kg]) and melts. The freshwater from the melted ice usually contains about 10-300 mg/l of total dissolved salts depending on the operation of the freezer and washer.

Heat Removal

Because all freezing processes operate at temperatures below ambient, some means must be supplied for the continual removal of heat from the system. Heat enters the process (1) with the warm seawater feed; (2) as a result of the mechanical action of pumps, agitators, compressors, and scrapers; (3) from the warm atmosphere through insulation that can minimize but not eliminate heat leakage.

Heat removal can be accomplished by adding ice, made by an external source, to the system. An example of such an ice maker is the Pak-Ice Machine manufactured by the Vilter Refrigeration Corporation. A more economical means, because it operates at a higher suction temperature, is the installation of a commercial refrigeration system in which the evaporating (refrigerant section) coil removes the excess heat from the melter and the system then discharges it to cooling water. Still another method is the use of a compressor to remove the excess uncondensed water or butane vapor from the melter and to compress it sufficiently that it will condense, to transfer its heat of vaporization to the cooling water that should be at the lowest available temperature.

Other Conventional Components. Other conventional system components include heat exhangers and deaerators.

i) Heat Exchangers--In order to reduce the temperature of the feedwater entering the crystallizer and thereby conserve energy in the process, heat exchangers are generally utilized between the brine and feedwater streams as well as between the product and feedwater streams. Thus, the cold brine and product water are run through a heat exchanger in order to cool the incoming saline feedwater and reduce the cooling required in the crystallizer. This liquid-to-liquid type of plate exchanger requires a significant heat transfer area (and hence capital cost) to be effective. Generally, process economics favor a good heat exchanger to yield a feed stream with an approach temperature to the crystallizer of about 2°F [1.1°C]; that is, the temperature of the feed stream is only 2°F [1.1°C] above the temperatures of the cold streams exiting the process.

ii) Deaerator--This device removes the dissolved gasses in the feedwater which would otherwise be liberated later under vacuum or stripping conditions in the process. These gasses would affect the overall operational efficiency due to blanketing of heat transfer surfaces with non-condensible gases. In the secondary refrigerant process (described later) any oxygen liberated in the crystallizer could combine with butane (if used) and create a potentially explosive mixture.

Freezing Desalination Processes

One simple form of freezing process is the scraped surface freezer. It is similar to the old ice cream maker still used in some households and occasionally by small vendors in some of the developing countries.

The freezer is essentially a bowl or cylindrical container placed in a cold refrigerant. The heat from the solution in the bowl passes through the walls of the bowl and into the refrigerant. An internal rotating scraper promotes crystal formation by removing as rapidly as it forms the thin ice layer formed on the inner surface of the bowl. Although this is a viable commercial system for high-cost items such as ice cream or concentrated fruit juices, it is too inefficient for making potable water at prices competitive with other desalination processes. Hence other freezing configurations have been tried for this purpose.

Among the freezing processes studied and tested over the past 20-25 years, several major processes have appeared viable, and a number of others have been investigated. Among the major processes are:

Vacuum-Freezing Vapor-Compression
Vacuum-Freezing Ejector-Absorption
Vacuum-Freezing Refrigeration-Compression
Secondary Refrigerant Freezing

The following paragraphs outline the highlights of each of these processes.

Vacuum-Freezing Vapor-Compression (VFVC). This process was refined to the extent that a commercial unit was produced for a condominium complex on the island of St. Croix. The process diagram is shown on Figures 4-4 and 4-5. This process also is referred to as a direct freezing process since the refrigerant is a portion of the water in the brine stream in the crystallizer itself (Johnson, 1979).

i) Vacuum Freezing--An ice slurry is formed in the crystallizer by a series of events. The first is the maintenance of a vacuum of about 0.004 to 0.005 atm (0.06 to 0.07 psia) in the freezer. This is below the vapor pressure of the saline water at its freezing point, causing the water to boil and produce vapor. The vapor formed absorbs its heat of vaporization from the remaining liquid so that for each pound of vapor, 1,076 Btu's [1,136 kJ] are removed from the seawater. This cools the seawater below its triple point and ice crystals begin to form and release their heat of crystallization to reheat the seawater back up to its triple point so that more water can vaporize. The energy removal needed for the formation of each pound of ice requires that 144 Btu's [152 kJ] be given up (heat of crystallization). Thus the energy required to produce the vapor is derived from the heat of crystallization of the ice, the ratio between the two being about 7.5 pounds [kg] of ice formed to each pound [kg] of vapor (Cox, 1969).

These processes of boiling and freezing take place simultaneously in the freezer, with seawater being added and ice slurry and vapor being withdrawn. The ice-brine mixture, in a pool, is agitated or stirred to aid in crystal formation and exchange of the heat within the slurry. The whole process occurs around the so-called "triple-point" which is the condition of temperature, pressure, and salinity at which the solid, liquid, and gaseous states of a substance (in this case water) coexist simultaneously in equilibrium.

ii) Vapor Compression--In order to maintain the temperature and pressure in the freezer in the vicinity of the triple point, it is necessary that the vapor produced by the boiling be removed. This is done by the vapor compressor, which is the major energy consumer in the process. It maintains the vacuum by removing the vapor from the freezer as fast as it is produced so that the heat exchanges between the processes of crystallization and vaporization occur continuously.

The vapor compressor compresses the vapor produced at about 25°F (the vapor is essentially pure water) sufficiently so that, when discharged into the melter, it will condense at 33 to 34°F [0.5 to 1.1°C]. Condensation takes place in the melter, where the compressed vapor contacts the washed ice and where each pound of vapor gives up 1,076 Btu's [1,136 kJ] of heat. This heat is used to melt the ice to form product water. Thus, melting recycles the heat taken from the freezer during the freezing process. Inefficiencies and heat leakages into the process produce more vapor than can be condensed by the available ice, and the excess heat must be removed by the heat removal system to permit condensation of all the water vapor and to keep the process in continuous operation.

iii) Ice Washing--The ice slurry from the freezer is pumped to the washer, where a portion of the product water (normally about

4 percent) is used to wash off the salts adhering to the ice crystals. The brine, including the washwater, is drained and discharged to waste after passing through a heat exchanger to cool the incoming feed.

iv) Major Considerations--The high vacuum and the large vapor compressor required constitute two major considerations in the vacuum-freezing vapor-compression system.

At a vacuum of 0.06 0.07 psi [0.004-0.005 atm] the system must be carefully constructed and maintained as airtight as possible. The low absolute pressure results in the production of vapor having a large specific volume so that if a freezing plant produced 0.1 mgd [379 m^3/d] of product water it would need a compressor which could handle about 400,000 cfm [188,800 l/s] (Johnson, 1976). Compressors for this large a volume and low a pressure are not commercial off-the-shelf units and are not designed for low vacuum service. The most nearly applicable ones existing are designed for use at atmospheric pressure and would require a drive motor of about 10,000 kW (versus a requirement of 75 kW for the actual freezing application), and in this case the compressor would cost more than the whole freezing plant.

Under partial funding by OWRT, Colt Industries developed a unique compressor suited for this type of service. Its major feature was the use of large, thin, light-weight, flexible metal blades. A compressor for a 0.1-mgd [379-m^3/d] plant was built which had blades 7 feet [2.1 m] in diameter and 10 inches [25.4 cm] wide, which rotated at 3,600 rpm. The cost of this compressor was about one-tenth that of an available commercial unit. However, larger-size compressors of this type for larger plants would require additional development.

Vacuum-Freezing Ejector-Absorption (VFEA). This process (see Figure 4-6) was developed as one way to avoid the costs and potential problems of large compressors. It is basically the same direct type of freezing process as is used in the vacuum-freezing vapor-compression unit described above except that the vapor compressor is replaced by a steam jet ejector and absorber system. Although this looks a great deal more complicated, it has the advantage of having few moving parts and none with the mechanical complexity of the high-volume compressor.

i) Vacuum-Freezing--The formation of ice crystals and water vapor is identical to that of the vacuum-freezing vapor-compression process.

ii) Ejector-Absorption--The removal of the water vapor produced in the freezer so as to maintain the proper vacuum (0.06 - 0.07 psia) [about 0.004-0.005 atm] is accomplished by a steam jet ejector. Motive steam for the ejector at about 250°F and 5.59 psia [121.1°C and 0.38 atm] is created in the absorber-concentrator. The compressed vapor is discharged to the melter.

The excess water vapor from the freezer is absorbed by a concentrated caustic solution (49.6 percent sodium hydroxide) which acts as an absorber. When this solution has been diluted to 48.3 percent, it is reconcentrated to 49.6 percent by being heated and partially vaporized on the outside of tubes in which saturated steam at about 275°F [135°C] is fed. This steam at 275°F [135°C] from an outside source supplies the primary power for the entire process. Excess heat from the process is removed by passing the cold, discharged brine and product streams through the absorber, where they pick up heat.

When the diluted caustic solution is heated in the concentrator, superheated water vapor (at about 250°F and 5.59 psia [121.1°C and 0.38 atm]) is generated, leaves the concentrator, and is used as the motive steam to drive the steam jet ejector.

The compressed water vapor from the freezer, emerging from the ejector together with the motive steam, enters the melter, where it condenses at about 32-40°F [0°-4.4°C] on the ice, releasing heat to melt the ice delivered from the washer.

iii) Ice Washing--Essentially the same type of washer as used in the vacuum-freezing vapor-compression process is used here.

iv) Major Considerations--Since this process also relies on a high vacuum in the crystallizer, the crystallizer must be constructed and maintained as airtight as possible. When operating, a performance factor of about 14 [6] was obtained; that is, 14 pounds of product per 1,000 Btu's [6 kg/MJ] of steam (Johnson, 1976). It was believed that this could be increased to about 20 lb/1,000 BTU's [8.6 kg/MJ] or more by increasing the efficiency of the ejector and absorber system (Johnson, 1976).

Vacuum-Freezing Refrigeration-Compression (VFRC). This process is another form of a direct freezing process. It is shown in Figure 4-7. Here the vapor compressor or steam ejector-absorber system designed to remove and compress water vapor from the crystallizer, has been replaced by an absorber-compression system, called a triple-point heat pump.

i) Vacuum Freezing--This is the same unit operation as in the vacuum-freezing vapor-compression and the vacuum-freezing ejector-absorber systems.

ii) Refrigeration Compression--This system maintains the required vacuum in the crystallizer by removing the water vapor as it is produced and condensing it on a cold stream of antifreeze solution in the vapor condenser. The antifreeze solution which enters the vapor condenser has a concentration of about 15 percent salts and is kept cold by passing it over the evaporating coil of a standard, sealed, commercial refrigeration system. As the antifreeze solution absorbs water vapor, the solution becomes diluted. During the absorption of the water vapor, its heat of vaporization is released and is removed by the refrigeration system.

When the antifreeze solution is diluted to about 14 percent salts, it is transferred to the antifreeze regenerator, where it flows over the hot condenser coil of the commercial refrigeration system. The hot coil evaporates some of the water in the dilute solution, reconcentrating it to 15 percent salts. This refrigeration system uses a commercial refrigerant such as Freon or ammonia which is enclosed in the standard loop (compressor, evaporator coil, orifice, condenser coil) and never mixes with the antifreeze or product water.

The antifreeze regenerating chamber operates at a partial vacuum of about 1.32 psia [0.09 atm], permitting the diluted 14 percent antifreeze solution to boil at about 35-36°F [1.6-2.2°C]. The resulting loss of vapor reconcentrates the antifreeze, and the water vapor produced goes to the melter, where it condenses on the ice, melting it to produce fresh product water. Excess heat is removed from the regenerating chamber by a small auxiliary heat pump (refrigeration system).

The major energy user in this process is the compressor in the standard refrigerant circuit.

iii) Ice Washing--Essentially the same systems as used in the other vacuum-freezing systems are used with this one.

iv) Major Considerations--Since this process also relies on a high vacuum, the system must be constructed and maintained as airtight as possible.

Secondary Refrigerant. This freezing process is referred to as an immiscible or secondary refrigerant process (as opposed to the primary process described thus far) as it depends on the use of a refrigerant other than the seawater itself to cool the water to produce ice. A block diagram of the process is shown on Figure 4-8.

i) Crystallizer (Freezer)--In the crystallizer an essentially immiscible liquid, such as butane, is mixed with the seawater. The crystallizer is maintained at an absolute pressure below the boiling pressure of the refrigerant (but not of the seawater) so that the refrigerant vaporizes. In the process of vaporizing, the refrigerant removes its heat of vaporization (165 Btu/lb at 30°F [384 kJ/kg at -1.1°C] for butane) from the cold seawater causing a portion of it to freeze.

The advantage of using the secondary refrigerant is that it vaporizes at a much higher pressure than water and hence the freezer can operate near ambient pressures. The specific volume of the vapor produced, therefore, can be much less (about 0.15 percent) than that produced at the far lower pressures necessary in the direct freezing processes. This allows the use of standard commercial compressors, reducing the capital cost and operation and maintenance problems.

ii) Vapor Compressor--A standard gas compressor is used to maintain the lowered pressure in the crystallizer and to remove the butane vapor. The butane is compressed and then condensed on the ice in the melter, producing a mixture of water and liquid butane.

This mixture is allowed to separate in the decanter. The freshwater is removed as product and the liquid butane is recycled to the freezer.

iii) Ice Washer--Essentially the same ice washer used in the direct freezing systems can be used with this system, although the presence of liquid butane can cause severe washing problems.

iv) Stripping--The loss of any of the refrigerant can significantly add to operating costs as well as cause a possible explosion hazard. Therefore both the brine and product discharge streams are run through vacuum strippers to recover dissolved or entrained butane and recycle it into the process.

v) Other Considerations--Although a high vacuum need not be maintained, caution must be exercised to minimize the escape of refrigerant and/or its mixture with air. Some of the refrigerants used, such as butane, can burn and/or explode if ignited. All equipment must be explosion-proof.

Other Processes. There are several other freezing processes which can be used, among them the indirect, the hydrate, and the eutectic processes.

i) Indirect--This is a modification of the scraped bowl process discussed above. A standard commercial refrigeration system is used to freeze or crystallize water by transferring heat through either a plastic or metal heat transfer surface on which the ice freezes (Johnson, 1979).

In another system, ice is produced either by bulk ice formation on existing crystals, or by the formation of new crystals in a stream passing over the heat transfer surface. The surface is chilled by the cold (evaporator) portion of a standard refrigeration loop. The hot end of the loop (the condenser) is often used to melt the ice to produce freshwater.

ii) Hydrate--This is a modification of the immiscible refrigerant process discussed above. In the hydrate process, however, the refrigerant is selected to form hydrate crystals with water. Hydrates can form at temperatures considerably above the freezing point of water. These crystals can be separated, washed, and then broken down into the component parts--water and the refrigerant which can be recycled for reuse.

The process can take place above the freezing point of water, thus reducing the amount and cost of much of the cooling equipment (Johnson, 1976), but a feasible hydrating agent has apparently not yet been found.

iii) Eutectic--This proposed process involves driving the freezing process at a very high water recovery rate so that the quantity of ice formed is increased until the concentration of residual salts is high enough to precipitate some of the salts, resulting in the formation of ice and wet salts. The ice could be melted to produce freshwater while the salts could be separated by gravimetric means and processed for recovery or disposal. Unfortunately, this process has not been refined to the point of feasibility.

Iceberg Utilization

The utilization of icebergs as a source of freshwater was carried out as early as 1773, when Captain James Cook picked up several pieces of a berg while sailing off the coast of Antarctica. The crew melted the pieces and thereby added to the ship's supply (Cameron, 1978). Towing icebergs from Antarctica to Saudi Arabia, Chile, Australia, the United States, and other countries, and utilizing them for the production of freshwater has been proposed in recent times (Al Faisal, 1976; Weaks and Mellor, 1978; Burt, 1956).

A serious effort has gone into investigating the possibility of moving an iceberg from Antarctica to the west coast of Saudi Arabia on the Red Sea for domestic and agricultural purposes. It is also thought that the huge iceberg could act to modify the weather in the area (Al Faisal, 1976). The estimate that a typical iceberg suitable for economical long-distance towing could weigh approximately 100 million tons [90 million metric tons] indicates the magnitude of such a project.

Researchers from many nations are investigating various aspects of iceberg utilization, including the size, shape, movement and fate of icebergs; the methods of moving icebergs long distances during a period of 8 to 12 months; the losses in mass to be expected during transport; the methods of converting the ice into potable water; the potential for weather modification and ecological changes; liability associated with iceberg movement; and other related subjects (Benedict, 1978; Davis, 1978; Hammond, 1978; Goldman, 1978; Burton, 1978; Al Faisal and Ismail, 1978).

The technical complexities and unknowns of moving an iceberg to any area appear to be quite substantial (Weaks and Mellor, 1978; Bader, 1978). To launch simply a pilot project involving smaller icebergs will require a sizeable investment. Not the least of the problems is the substantial draft of a floating iceberg; this can easily be 100 to 400 ft [30.5 to 122 m], which restricts the choice of navigation lanes and presents a tremendous problem if the iceberg must be brought close to shore for conversion to freshwater.

Although pilot projects of this nature may be undertaken in the 1980's, considerable development and study will be required before iceberg utilization could be considered a viable method for dependably supplying water to developing countries.

COMMERCIAL DEVELOPMENT

History

Secondary Refrigerant. The processing of seawater by freezing appears to have begun in Japan for the purpose, not of obtaining potable water, but of concentrating seawater to augment the production of salt by solar evaporation. The Japanese process utilized a secondary refrigerant for freezing, and in this application it appears to have been successful. Washing of the ice crystals was not required.

When interest in obtaining potable water from seawater developed in the United States, a modification of this process appeared to be feasible. About 1960, Dr. Umano of Japan worked with North American Aviation Corporation, in California, and a small pilot plant was built and tested. The economics of the process apparently were not attractive to North American, and they sold their interest and knowhow to Struthers Scientific and International Corporation in New York (a Struthers-Wells subsidiary) who, under contract to the Office of Saline Water, constructed a 200,000-gpd [757-m^3/d] pilot plant at Wrightsville Beach, North Carolina. The pilot plant never operated properly and after a few years was dismantled. About the same time, OSW sponsored the construction of a pilot plant by Blaw-Knox for the Florida Power & Light Company.

In the early 1960's Israel Desalination Engineering built a secondary refrigerant pilot plant in Israel that used n-butane, and succeeded in operating it briefly. The economics were not attractive, however, and the process was dropped.

In the 1970's the process was again examined by several companies. Under contract to OWRT, the Dravo Corporation in Pennsylvania designed a 10-mgd [37,900 m^3/d] plant and examined its economics (Dravo, 1973). If this process could be made to function as designed, the costs appear attractive, but no hardware was built or tested.

The AVCO Corporation in Massachusetts, also under contract to OWRT, built and tested a pilot plant designed to concentrate waste streams from processing plants. So many difficulties were encountered that AVCO terminated the development.

Most recently, the United Kingdom Atomic Energy Authority, together with the firm of Simon Carves, operated a pilot plant using butane in England, apparently with success, and designed a 1-mgd plant. The project was terminated, however, shortly thereafter.

Mitsui Shipbuilding and Drydock Company in Japan reported their interest (Trade Times, 1978) in butane freezing, and Rohm and

Haas Company in Pennsylvania has been considering the process for industrial waste treatment.

Vacuum Freezing. The vacuum freezing vapor compression (VFVC) process was conceived in the early 1950's independently and almost simultaneously in the United States by C. M. Ashley of Carrier Corporation and R. B. Cox, and in Israel by Dr. A. Zarchin. Patents applied for in 1955 and 1956 led to the declaration of an interference by the U.S. Patent Office. The interference eventually was decided in favor of R. B. Cox, who was awarded the U.S. patent in 1969. Dr. Zarchin holds an Israeli patent, and in Israel VFVC is known as the Zarchin Process. Dr. Zarchin conceived the principle of the light-weight, low-cost, high-volume vapor compressor that made the process economical. Dr. Zarchin assigned his patent to the newly organized Israel Desalination Engineering (IDE), which signed a joint process development and commercialization agreement with what is now the U.S. firm of Colt Industries. Cox licensed his patent to Colt.

i) Early Work--The OSW began early to support work in freezing. Cox applied for a contract, but the contract was given to Carrier Corporation, which had applied at about the same time. Carrier developed equipment for freezing, washing, and melting the ice, but they did not develop a low-cost compressor, and in the middle 1960's dropped the development.

After much negotiating with their Israeli partners, Colt in 1961, also obtained a development contract with OSW which eventually resulted in the building and operating of a pilot plant at Wrightsville Beach, North Carolina. This plant incorporated their own low-cost compressor. Designed originally to produce 60,000 gpd [227 m^3/d], the plant actually produced as much as 120,000 gpd, [454 m^3/d] and by the end of the contract in the early 1970's operated steadily for 4 months, essentially unattended, and produced 100,000 gpd [379 m^3/d] of product water at a power consumption of 44 kWh/kgal [11.6 kWh/m^3]. The product was blended with the well water for the town of Wrightsville Beach to upgrade their supply.

ii) St. Croix Experience--In about 1969, Colt sold a 100,000-gpd [379-m^3/d] VFVC plant to a condominium on the island of St. Croix in the Caribbean. Because the heat removal system was apparently insufficient in capacity, the plant could not operate continuously. Before that problem could be rectified the condominium owners went bankrupt, and Colt removed the plant.

iii) Roswell, New Mexico, Experience--Under contract to OSW, Colt designed, built, and tested a 15,000-gpd [57 m^3/d], skid-mounted demonstration plant at the OSW test facility at Roswell, New Mexico. Its purpose was to demonstrate the feasibility of desalting well water saturated with scale-forming calcium sulfate. With this water as feed, all other processes require very extensive treatment to remove either the calcium or the sulfate to prevent scale. Also, because this VFVC demonstration plant had inadequate heat removal, it operated only for short periods. It did demonstrate, however, that the VFVC process would desalt this normally severely scaling feed without pretreatment. The tests confirmed that freedom from scale is one of the major advantages of freezing.

iv) Termination of Colt-IDE Agreement--IDE stopped their development of the VFVC process in about 1966 in favor of small, all-aluminum vapor-compression plants which they could manufacture and sell by themselves. These units used the low-cost compressor developed for the freezing process. Colt, on the other hand, pursued the VFVC process until about 1975, when they terminated their contract with IDE.

Extent of Usage

At the present time the usage of freezing plants for desalination is limited to pilot units.

ENGINEERING CONSIDERATIONS

Advantages

There are a number of advantages inherent in the freezing process relative to other processes, especially the distillation process (Johnson, 1979).

Resistance to Corrosion. The corrosion rate of many materials decreases exponentially with decreasing temperature (Johnson, 1979). This permits a wider selection of materials and construction methods without the problems usually associated with other desalination methods. Corrosion in the freezing processes generally is not a problem, even for ordinary materials, and low-cost plastic materials can be used for many applications.

Pretreatment. Eccept for deaeration, pretreatment is generally unnecessary in freezing.

Resistance to Fouling. Fouling of surfaces by organic growth or the formation of scale such as occurs in membrane or distillation units is not a problem.

Insensitivity to Concentration. The process is relatively insensitive to changes in the concentration or type of substances in the feedwater.

Low Energy Requirements. The energy requirement is reasonable when compared to single-purpose distillation units. The exact energy usage of a new generation freezing plant is not known, but it should be between the 44 kWh/kgal [11.6 kWh/m^3] reported for an operating vacuum-freezing vapor-compression unit and the prediction, in 1969, of about 25.5 kWh/kgal [6.7 kWh/m^3] for a fully developed 1- to 5-mgd [3,790 to 18,900-m^3/d] unit (Fluor, 1978; Colt, 1969). This advantage has to be weighed against potentially low cost for steam in a dual-purpose distillation plant. By comparison, seawater reverse osmosis plants have reported energy usages of about 32 to 38 kWh/kgal [8.5 to 10 kWh/m^3] (Buros, 1979, Larson, 1979).

Disadvantages

Aside from the fact that there are no commercial installations of freezing desalination plants, a number of other problems are associated with the freezing concept.

Solids Handling. All desalination processes involve the manipulation of liquid and/or vapor streams, but the freezing process involves the handling of small crystalline solids as well. These are much less predictable in their size and shape and are prone to coalesce and build up as massive solids at some points, and therefore crystal transfer and melting require special handling procedures. When the equipment is suitably designed, the handling of solids generally causes few or no problems.

Mechanical. The principal difficulty in vacuum freezing processes is the handling of the extremely large volumes of water vapor produced per unit of ice. It is the solution of this problem that has led to the development of the different freezing proceses (VFRC and VFEA) now undergoing pilot plant demonstration or study.

In the immiscible refrigerant processes, several problems have not yet been solved.

1. The ice washer sometimes does not wash the ice crystals free of salt.
2. The refrigerant sometimes forms emulsions with the brine, and these emulsions are stable even at ambient temperatures. This results in unacceptable loss of the refrigerant and in a high-salinity product.
3. Salt deposits in the butane compressor cause severe erosion.

Safety. Hazards in the VFRC and VFEA processes are essentially nil, but some of the immiscible refrigerants that can be used in the secondary refrigerant process such as butane are flammable and can explode under certain cicumstances. When butane is the refrigerant, all equipment must be explosion-proof.

STATE OF THE ART

Process Development

Only two processes have reached commercial development. These resulted in the construction of the 0.1-mgd [379-m^3/d] vacuum-freezing vapor-compression unit produced by Colt Industries and the design of a 1-mgd [3,790-m^3/d] secondary refrigerant plant for Ipswich, England, by the United Kingdom Atomic Energy Authority and Simon Carves. The latter plant was designed but not built.

Unit Operations

Although a commercial plant has not been in operation in the period 1975-1980, considerable work in the past has been undertaken on various unit operations involved in the different freezing processes. This progress has been reviewed by Johnson (1979) and his comments are summarized below.

Precooling. Approach temperatures (temperature of the feedwater stream relative to that of the streams exiting the process) of 2° to 3°F [1.1° to 1.6°C] have been achieved economically, utilizing aluminum sheet construction for the heat exchangers.

Crystallization. This has been accomplished successfully with the direct freezing processes. It is still in the laboratory study phase for the indirect process utilizing high-velocity flow over a cold surface.

Vapor Compressors. Some problems still exist here for the VFVC and the immiscible refrigerant processes, and further development is required in several areas.

Heat Removal. Standard equipment can be used.

Refrigerant Stripping. This is necessary in the secondary refrigerant process, in which the immiscible refrigerant is carried over in the product and brine streams. Dissolved refrigerant is easily removed in vacuum strippers, but the removal of suspended refrigerant from emulsions that form has proved more difficult.

Washing. Many schemes for washing the ice crystals free of brine have been investigated on a laboratory and pilot scale. The one with the greatest potential appears to be a device called the hydraulic piston, which works well for the vacuum freezing processes. The use of this device in VFVC has generated product water with a salt content as low as 4 mg/l. A satisfactory washer for the secondary refrigerant process has yet to be demonstrated.

Current Development

During 1980, work on freezing processes is being carried out by two firms in the USA with financial support by the Office of Water Resources and Technology (OWRT).

The first of these is the operation by Concentration Specialists of Andover, Massachusetts, of a 0.015-mgd [57-m^3/d] pilot plant at the U.S. Government Desalination Test Center at Wrightsville Beach, North Carolina. This pilot plant employs the vacuum-freezing refrigeration-compression process. The second is the investigation of the indirect freezing process by the Chicago Bridge and Iron Company. In 1981 the Chicago Bridge and Iron Company will build and operate a 0.006-mgd [23-m^3/d] indirect freezing unit at the Wrightsville Beach center to test out some of the recent developments in this process.

APPLICATIONS

Potable Water

Theoretically freezing can be used to produce water with a low salt content from either brackish water or seawater. Since there is only a minor energy difference between desalting either type of water, it would generally make economic sense to utilize freezing for seawater desalination. There are less energy-intensive methods suitable for brackish water processing unless the chemistry of the brackish water creates special problems, in which case freezing could offer a viable solution.

Industrial Wastes

A great potential for economic development of the freezing process lies in the use of freezing to concentrate industrial waste streams for disposal or the recovery of valuable products.

HEALTH ASPECTS

Although the levels of chemical constituents and microorganisms are reduced by the freezing process, they are not entirely eliminated. It is therefore necessary to treat the product water stream to ensure disinfection. Disinfection is a precaution that should be used with all desalination processes.

If the secondary refrigerant process were utilized, a portion of the refrigerant may be entrained in the product stream. The health effects would be dependent on the refrigerant used and the quantity ingested. At this time little is known about possible health effects.

ENVIRONMENTAL CONSTRAINTS

Potential environmental constraints for freezing plants could consist of noise and visual distractions the discharge of a concentrated brine stream at a temperature other than ambient and the possible discharge of entrained refrigerant in the brine stream from a secondary refrigerant plant.

FUTURE PROSPECTS

There appears to be a great potential for the development of the freezing process, especially in the field of concentrating industrial waste streams. In the 1978 report to the U.S. Government on the state of the art and future of desalination, the Fluor Corporation predicted that by the year 2000 the freezing process would capture up to 38 percent of the desalination market in the USA, which will be the largest market in the world. The expected feedwaters would consist of seawater, difficult brackish water, or waste streams. The advantages of the freezing process over other desalination processes for industrial wastewater concentration are numerous due to its relative immunity to fouling and corrosion. The market potential in industrial waste processing is so expansive that development work on the freezing process is expected to increase in the near future. A major spin-off from this work should be the refinement of the potable water freezing desalting process.

Another potential for the freezing distillation process is to produce water in conjunction with the revaporization of liquid natural gas (LNG). Where storage facilities for LNG exist, the large quantity of "cold" produced by the vaporization of the LNG before being distributed to pipelines could be utilized in conjunction with the freezing desalting process. Investigations have been conducted in this area by the Mitsui Shipbuilding Company of Japan (Johnson, 1976) and earlier (1969) by the Institute of Gas Technology in Chicago, Illinois.

PROCESS EXPERIENCE

Previous AID Studies

Wrightsville Beach North Carolina (0.1-mgd [379-m^3/d] VFVC). A 0.1-mgd [379-m^3/d] vacuum-freezing vapor-compression seawater desalination plant was operated at the OWRT Desalination Test Center at Wrightsville Beach.

i) Process Description--A vacuum-freezing vapor-compression (VFVC) plant as described above in the "Theory" section was designed, built, and operated by Colt Industries. The seawater feed was only filtered, deaerated, and then cooled in a plate-type heat exchanger before entering the crystallizer. The crystallizer, vapor compressor, and melter were built into one vessel called a hydroconverter. It used a compressor with large, flexible steel blades for the high-volume, low-pressure vapor compression needed for this operation. A diagram of the plant is shown on Figure 4-4.

ii) Materials of Construction--The materials of construction for the principal elements of the vacuum-freezing vapor-compression plant, together with those for other freezing processes, are listed in Table 4-1.

iii) Operating Information--For a 100,000-gpd [379-m^3/d] plant operating with 35,000 mg/l seawater feed at 75°F [23.8°C] and converting 50 percent of the feed to product water, the vacuum-freezing vapor-compression plant demonstrated a power consumption of 45 kWh/kgal [11.9 kWh/m^3]. The following tabulation shows how this energy consumption compares with that expected for other freezing processes (Concentration Specialists, 1977).

Table 4-1
MATERIALS OF CONSTRUCTION FOR PLANTS USING FREEZING PROCESSES

Component	Vacuum-Freezing Vapor-Compression	Vacuum-Freezing Ejector-Absorption	Vacuum-Freezing Refrigeration-Compression	Secondary Refrigerant (Butane)
Bar Screen and Rake	Steel and 316 S.S.	Steel and 316 S.S.	Steel and 316 S.S.	Steel and 316 S.S.
Deaerator	Steel lined with bitumastic PVC packing.	Steel lined with bitumastic PVC packing.	Steel lined with bitumastic PVC packing.	Steel lined with bitumastic PVC packing.
Feed Heat Exchangers	5052 aluminum sheets.	5052 aluminum sheets.	5052 aluminum sheets.	Carbon steel coated with epoxy.
Freezer	Epoxy-coated steel.	Epoxy-coated steel.	Epoxy-coated steel.	Phenolic-coated steel.
Ice Washer	Epoxy-coated steel.	Epoxy-coated steel.	Epoxy-coated steel.	Phenolic-coated steel.
Ice Scraper	Epoxy-coated steel. 304 S.S. cutter blades.	Epoxy-coated steel. 304 S.S. cutter blades.	Epoxy-coated steel 304 S.S. cutter blades.	Phenolic-coated steel. 304 S.S. cutter blades.
Melter	Epoxy-coated steel.	Epoxy-coated steel.	Epoxy-coated steel.	Phenolic-coated steel.
Vapor Compressor	Steel hub and shaft, titanium blades, FRP shrouds.	Steel steam-jet ejector.	Commercial refrigeration system.	Steel
Heat Removal	Commercial refrigeration system.	304 S.S.	Commercial refrigeration system.	Carbon steel clad with copper-nickel.
Vent Compressor	Ductile iron.	Ductile iron.	Ductile iron.	Ductile iron.
Pumps				
Seawater & Brine	Stainless steel.	Stainless steel.	Stainless steel.	Stainless steel.
Product	Ductile iron.	Ductile iron.	Ductile iron.	Ductile iron.
Vacuum	Cast iron.	Cast iron.	Cast iron.	Cast iron.
Butane				Ductile iron.

Table 4-1--Continued

Component	Vacuum-Freezing Vapor-Compression	Vacuum-Freezing Ejector-Absorption	Vacuum-Freezing Refrigeration-Compression	Secondary Refrigerant (Butane)
Absorber		304 S.S. (for NaOH)	5052 aluminum (for NaCl)	
Heat Exchanger		304 S.S. (for NaOH)	5052 aluminum (for NaCl)	
Concentrator		316 S.S. (for NaOH)	5052 aluminum (for NaCl)	
Condenser		304 S.S. (for NaOH)	5052 aluminum (for NaCl)	
Butane Storage				Carbon steel.
Butane Strippers				Phenolic-coated steel.
Butane Condenser				Carbon steel clad with copper-nickel.
Piping[a]	PVC, FRP, or steel, depending upon pressure, temperature, diameter, and service.	PVC, FRP, or steel, depending upon pressure, temperature, diameter, and service.	PVC, FRP, or steel, depending upon pressure, temperature, diameter and service.	Kynar, FRP or steel, depending upon pressure, temperature, diameter, and service.
Reference	Colt Industries, ____.	Colt Industries, 1972.	Concentration Specialists, 1977.	Dravo Corp., 1973.

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[a] PVC = Polyvinyl chloride.
FRP = Fiberglass reinforced plastic, usually epoxy or polyester.
Kynar = Polyvinylidene fluoride.

Process	Energy Consumption kgal	 m^3
Vacuum-Freezing Vapor-Compression	45 kWh	11.9 kWh
Vacuum-Freezing Ejector-Absorption[a]	14 kWh + 600,000 Btu (steam) at 270°F	3.7 kWh + 633 MJ (steam) at 132.2°C
Vacuum-Freezing Refrigeration-Compression	46 kWh	12.2 kWh
Secondary Refrigerant (Butane)	50 kWh	13.2 kWh

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[a] Vacuum-Freezing Ejector-Absorption may be economical if low-pressure extraction or waste steam is available.

iv) Operational Problems Encountered--Few problems were encountered in the 100,000-gpd [379-m^3/d] pilot plant at Wrightsville Beach. The plant was automated and operated with only intermittent attention for 4 months, until the expiration of the contract.

The major problems with the vacuum-freezing vapor-compression process have not been operational; rather, they have been legal, contractural, and (especially) financial.

Recent Studies and Experience

The previous A.I.D. manuals on water desalination (published in 1968 and 1972) described the then current developments. The status of the freezing processes at present can be summarized as follows.

Indirect Freezing. Although in the early days of freeze desalting the indirect freezing process was dropped due to various theoretical and mechanical problems, recent developments have revived this process. During 1979 to 1980 the Chicago Bridge and Iron Company has done extensive work, with OWRT assistance, on the development of a falling film system using a 6 percent brine solution. Improvements to tube surfaces have reduced undesirable ice formation which had previously caused problems. Furthermore, the process equipment is designed so that most components operate at atmospheric pressure. In 1981, a small 0.006-mgd [23 m^3/d] test unit will be built and operated in Wrightsville Beach.

Vacuum-Freezing Vapor-Compression (VFVC). Development and commercialization of this process was continued by Colt Industries until 1975, when it too was dropped. This process is described in detail under "Freezing Desalination Processes" and "Vacuum Freezing."

The vacuum-freezing vapor-compression is the only freezing process for which commercialization has been attempted. Other processes have been abandoned in the pilot plant stage or are still being developed.

Since Colt Industries terminated their work, the vacuum-freezing vapor-compression process has not been pursued, primarily because of the nonavailability of low-cost vapor compressors.

Vacuum-Freezing Vapor-Absorption. Carrier Corporation studied this process using lithium bromide as the vapor absorbent and an external steam supply as a means of regenerating the lithium bromide solution. Carrier concluded that the process is not cost competitive with vacuum-freezing vapor-compression, and development was not pursued.

Direct Freezing Using a Second Refrigerant. Also known as "Direct Freezing Using an Immiscible Refrigerant." Variations of this process are still being studied and tested. The concepts, together with interested developers, are discussed under "History--Secondary Refrigerant," and "Recent Studies and Experience."

A number of companies have pursued the butane process, primarily because of the availability of commercial equipment for building units of 10 mgd [37,900 m^3/d] or even larger. Although no commercial plants were built, a number of pilot plants were built and tested by:

North American Aviation Company, California.

Struthers Scientific and International Corporation, New York.

Israel Desalination Engineering, Israel.

Koppers Company, Pennsylvania.

AVCO Corporation, Massachusetts, using R-114, dichlorotetrafluoroethane as the immiscible refrigerant (Robertson, 1978).

United Kingdom Atomic Energy Authority, England.

Mitsui Shipbuilding and Drydock Company, Japan (current).

Rohm and Haas Company, Pennsylvania (current, for the treatment of waste streams).

Hydrate Formation Crystallization. Sweetwater Development Corporation tested the propane-hydrate process, and Koppers Company tested the Freon-hydrate process. Neither company was successful in developing a satisfactory, cost-effective method of washing the hydrate crystals. The Office of Saline Water withdrew its financial support.

In the 1960's, under contract to OSW, the Sweetwater Development Company of Texas, built and tested a propane hydrate pilot plant at the Wrightsville Beach test facility. The process was abandoned when financial support terminated before a feasible method was found for washing the almost microscopic, waxy, propane hydrate crystals.

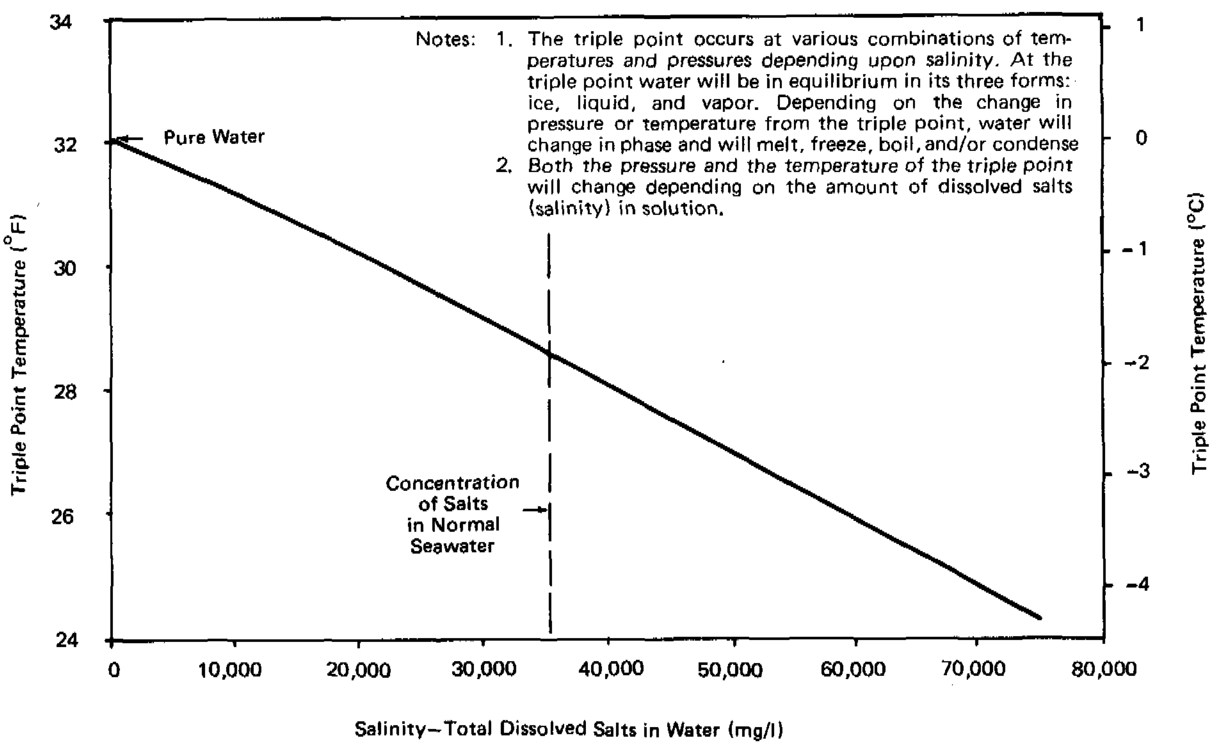
Vacuum-Freezing Ejector-Absorption (VFEA). This process is currently being studied by the Chicago Bridge and Iron Company of Illinois.

Vacuum-Freezing Refrigeration-Compression (VFRC). Under contract to OWRT, Concentration Specialists, of Massachusetts, have built and are operating a 25,000-gpd [95-m^3] vacuum-freezing refrigeration-compression pilot plant at the Wrightsville Beach test facility. The operability of all facets of the process and of the process itself, have been demonstrated, but the plant has not yet operated at its design capacity.

APPLICABILITY FOR DEVELOPING COUNTRIES

Although the freezing process has obvious potential for desalination of seawater and difficult brackish waters, the process has no commercially operating facilities in place in 1980. It is doubtful that full-scale 0.1-mgd [379-m^3/d] or larger size plants for producing potable water could be built and operating commercially before 1984-5. It could then take several years to "debug" the plants and build the second or third generation units, which would incorporate the necessary modifications derived from experience with earlier plants.

Until operating experience has been accumulated on commerical installations and plant design and construction have been greatly refined, extreme caution is advisable in considering the freezing process for applications in developing countries during the 1980's. There may be rapid developments, engineering breakthroughs, very attractive financial incentives, or other extenuating circumstances which could modify this judgment, but these would need to be carefully evaluated case by case.



Notes: 1. The triple point occurs at various combinations of temperatures and pressures depending upon salinity. At the triple point water will be in equilibrium in its three forms: ice, liquid, and vapor. Depending on the change in pressure or temperature from the triple point, water will change in phase and will melt, freeze, boil, and/or condense

2. Both the pressure and the temperature of the triple point will change depending on the amount of dissolved salts (salinity) in solution.

FIGURE 4-1A. The effect of changes in salinity on the temperature of the triple point of water.

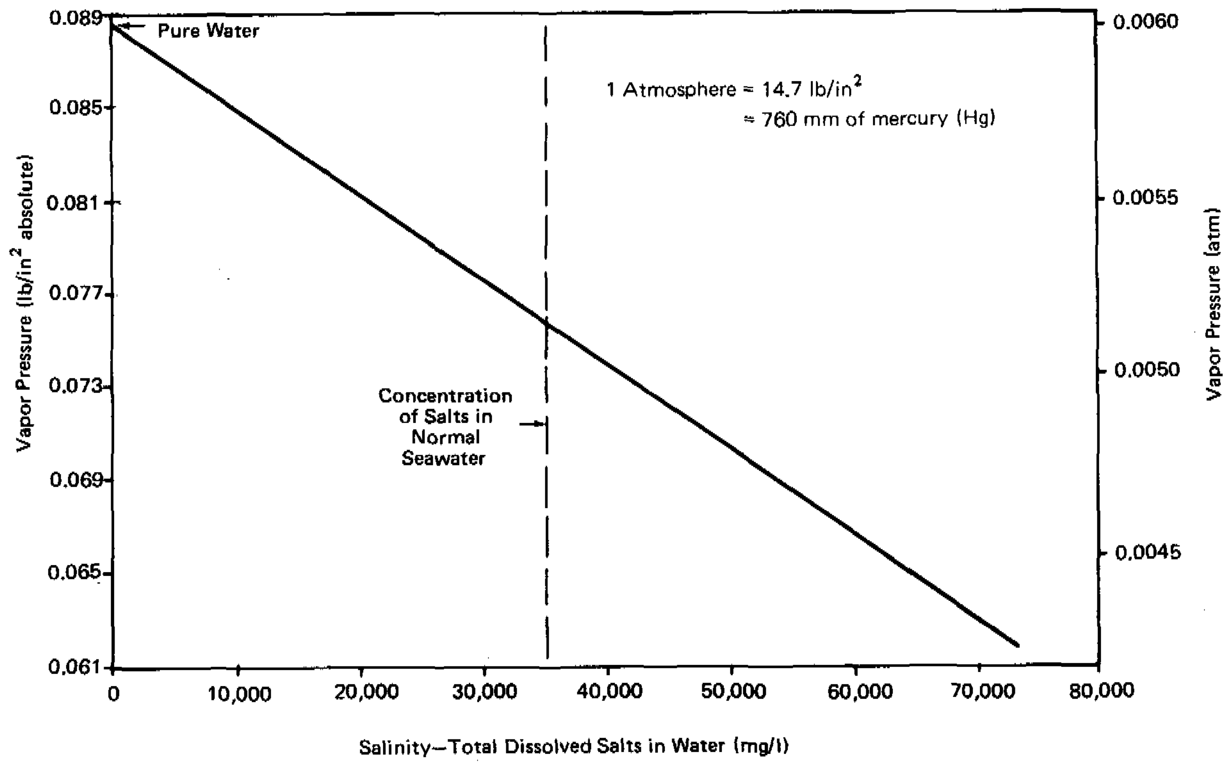
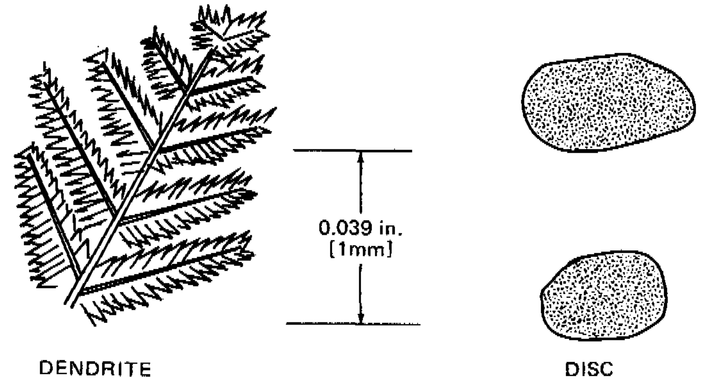


FIGURE 4-1B. The effect of changes in salinity on the vapor pressure of water at its freezing point.



Notes:
1. Dendrite crystals tend to be formed when agitation of the solution is minimal and the driving forces (temperature differential) are high. They are very fragile and can be easily torn apart by agitation. The resultant fragments will act as nuclei for new dentritic or disc crystals.
2. Disc crystals tend to be formed in the presence of good agitation of the solution and low driving force. Disc thickness is 10-20% of disc diameter.

FIGURE 4-2. Structure of ice crystals formed in saline solutions.

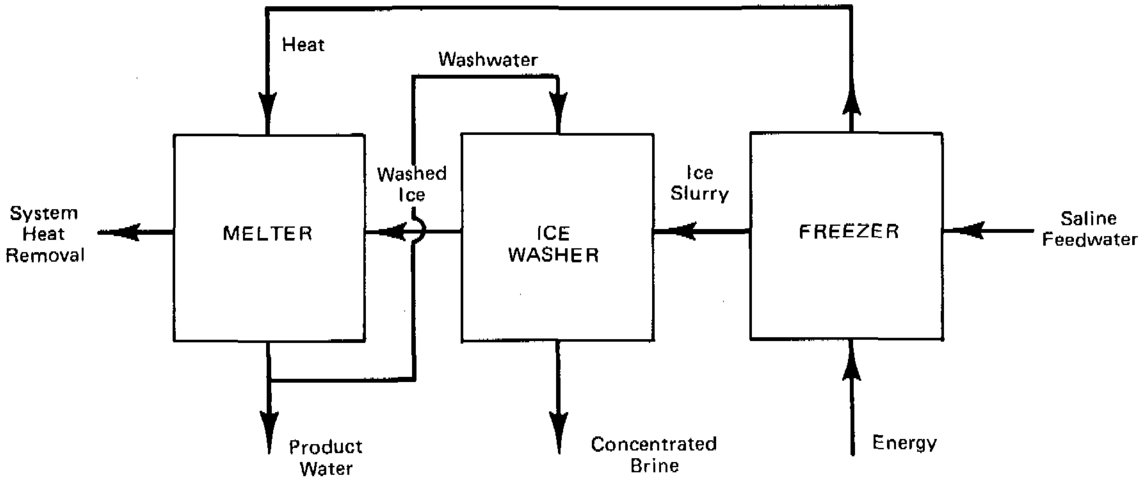
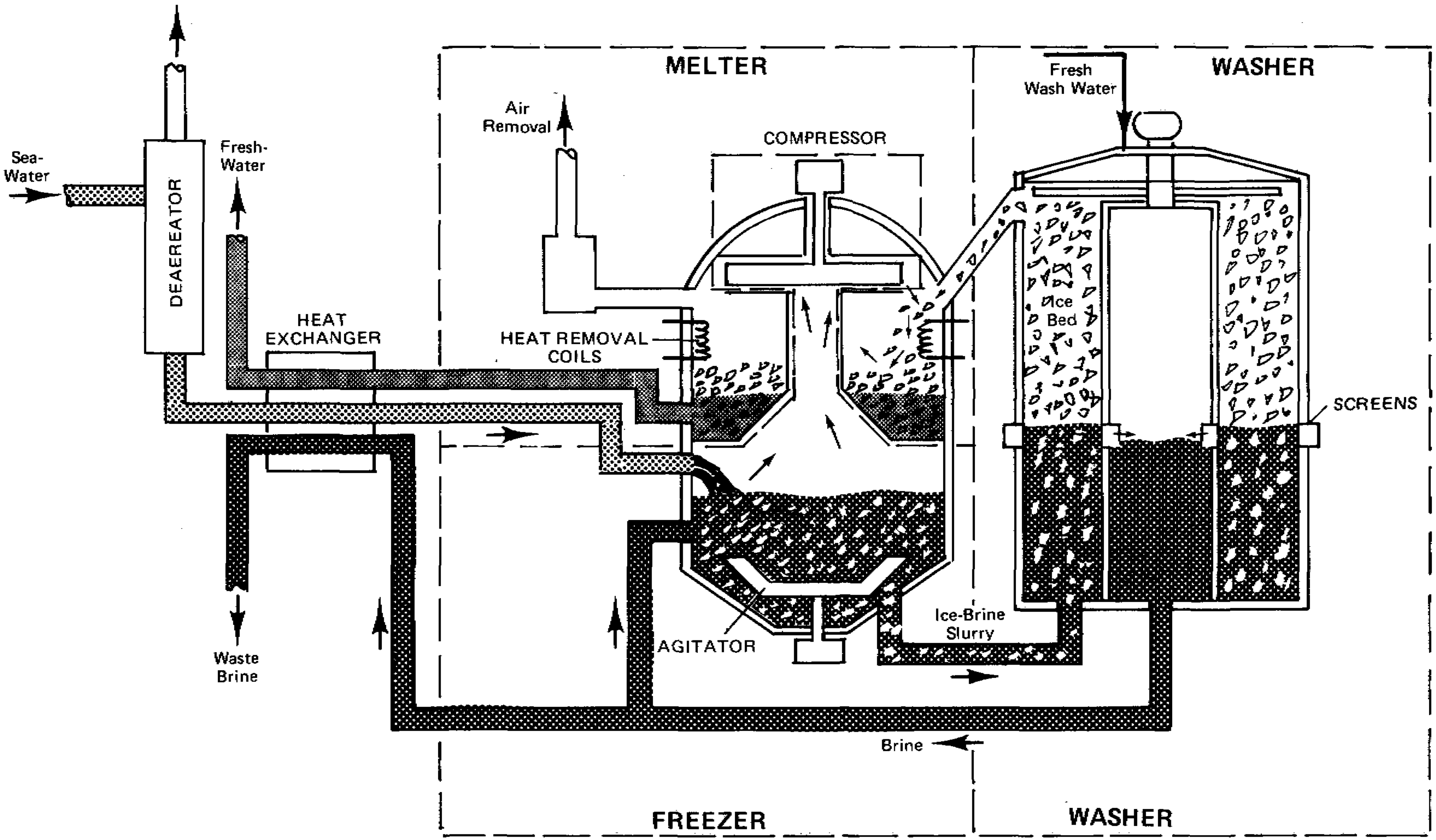


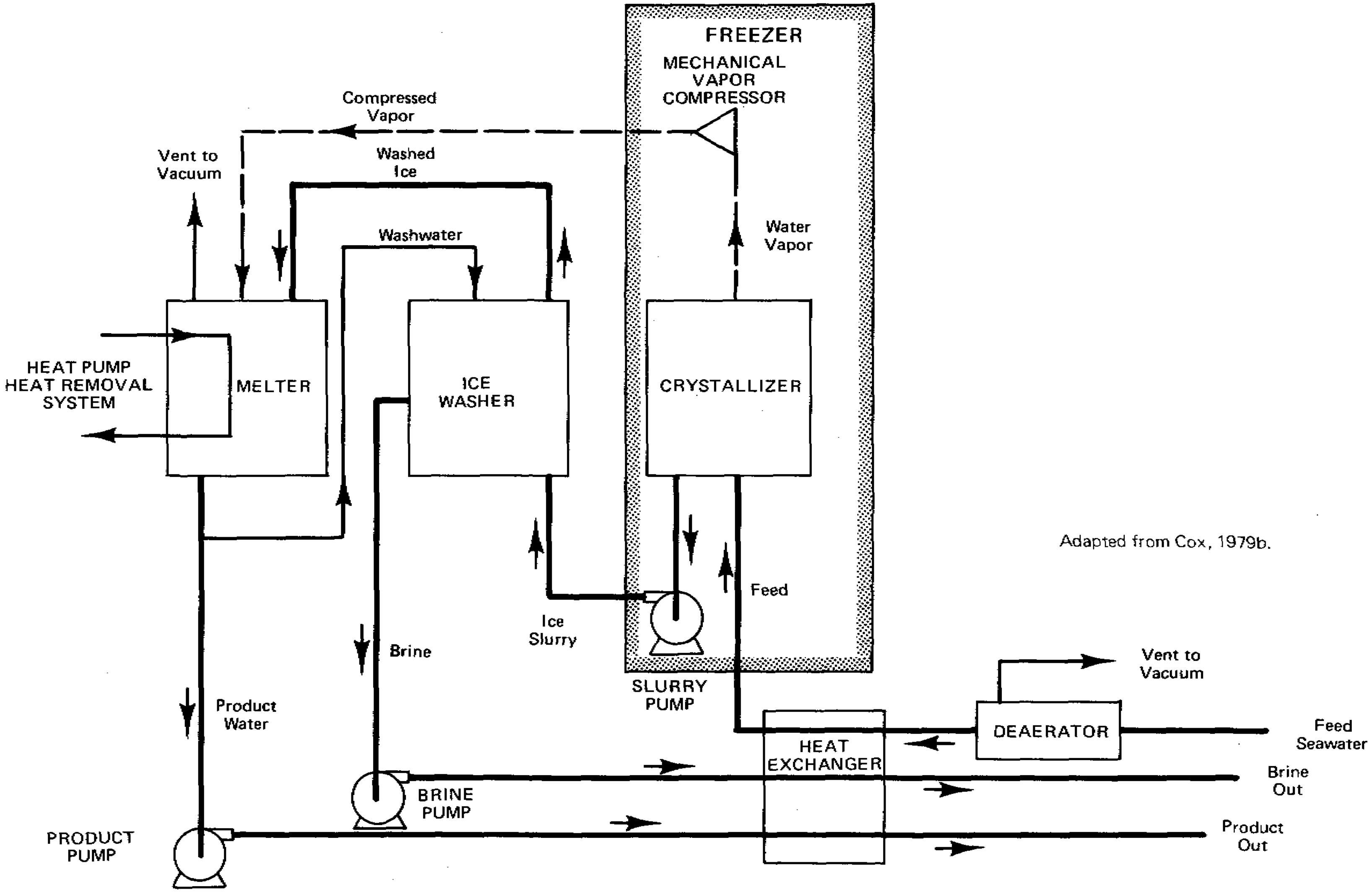
FIGURE 4-3. Basic components in the freezing desalination process.



Notes:
1. The combined freezer (crystallizer) and melter in one vessel is a unit called a hydroconverter.
2. The compressor utilizes thin flexible metal blades. It is built specifically for the high-volume, low-pressure conditions found in the VFVC process.

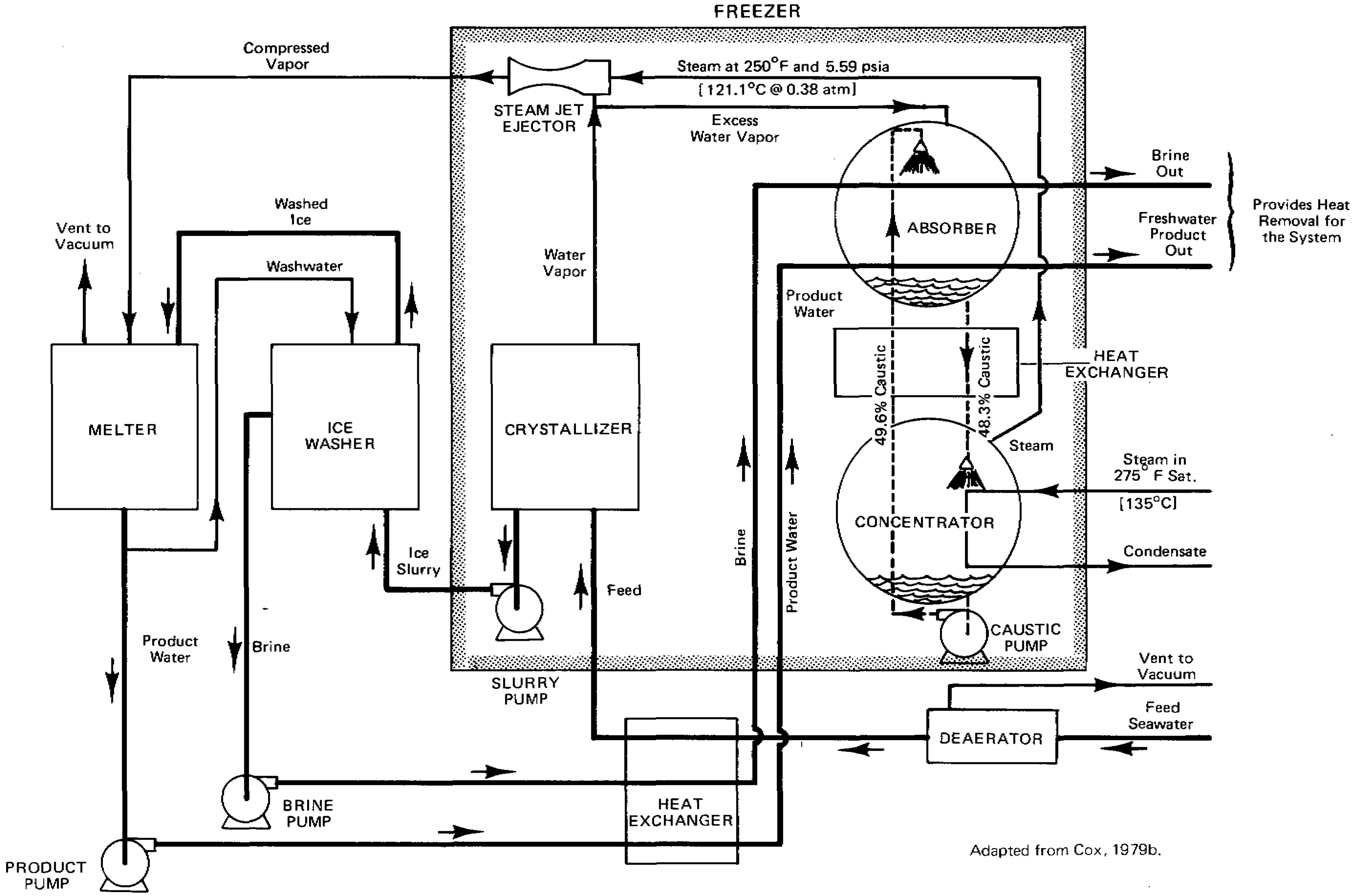
Adapted from The A-B-C of Desalting. OWRT, 1977.

FIGURE 4-4. Schematic diagram of a freezing desalination plant using the vacuum-freezing vapor-compression (VFVC) process.



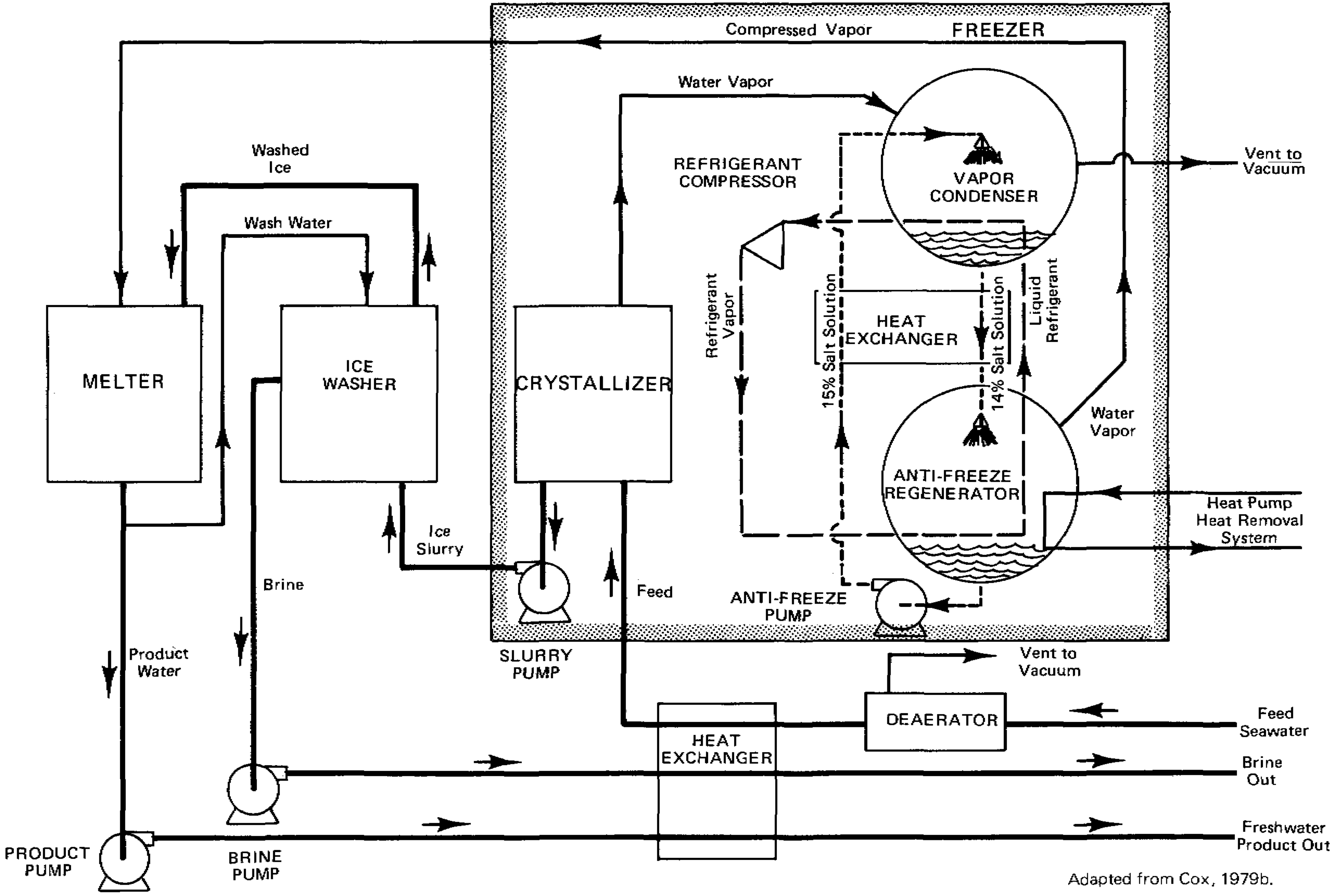
Adapted from Cox, 1979b.

FIGURE 4-5. Block diagram of a vacuum-freezing vapor-compression (VFVC) freezing process for desalting seawater.



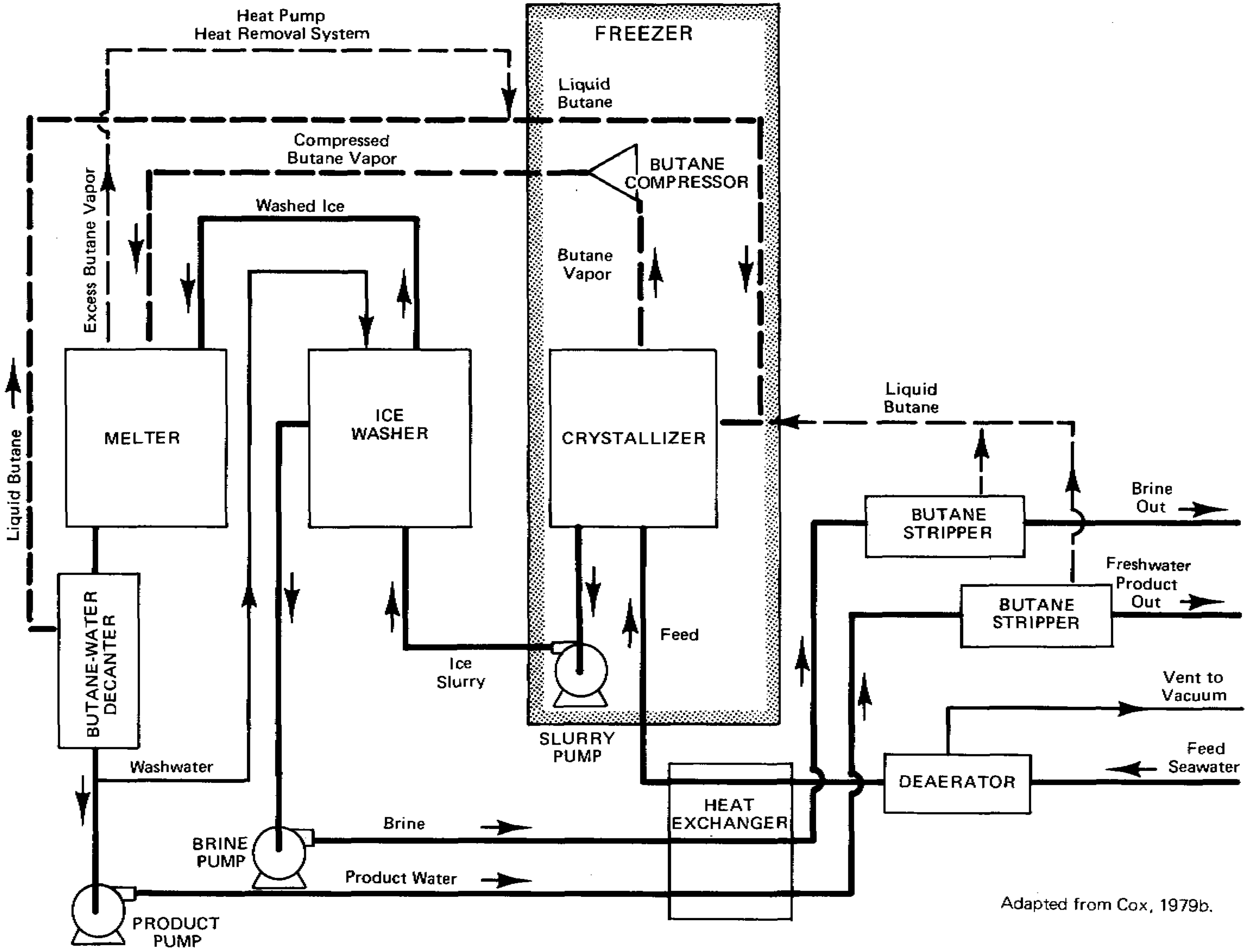
Adapted from Cox, 1979b.

FIGURE 4-6. Block diagram of a vacuum-freezing ejector-absorption (VFEA) freezing process for desalting seawater.



Adapted from Cox, 1979b.

FIGURE 4-7. Block diagram of a vacuum-freezing refrigeration-compression (VFRC) freezing process for desalting seawater.



Adapted from Cox, 1979b.

FIGURE 4-8. Block diagram of a secondary-refrigerant freezing process for desalting seawater.

CHAPTER 5

THEORY

Reverse Osmosis

The desalination of water by reverse osmosis is a membrane separation process in which the water from a pressurized saline solution is separated from the solutes and flows through an appropriate membrane. The permeate (the liquid flowing through the membrane), which generally emerges near atmospheric pressure, is reduced in salt content while the feed solution which is pressurized on the other side of the membrane concomitantly increases in salt content. As no heating or phase change takes place, the major energy usage in the process is that required to pressurize the feed. For brackish water desalination the operating pressure generally ranges from 300 to 400 psi [20.4 to 27.2 atm] and for seawater desalination, it generally averages from 800 to 1,000 psi [54.4 to 68 atm].

As illustrated schematically on Figure 5-1, the reverse osmosis process operates with three major streams. These are listed below:

Stream	Salt Content	Pressure
Feedwater	High	High
Brine	Higher	High
Product Water	Low	Low

The product water, which has passed through the membrane, is referred to as the permeate.

Although osmotic pressure effects are directly associated with the overall process, reverse osmosis is not the reverse of osmosis (Sourirajan, 1980). Over the years a number of theories have been advanced on the theoretical mechanism of reverse osmosis. These are basically two major concepts:

1. Absorption porous flow.

2. Solution diffusion.

Further information on the theoretical aspects of these two concepts can be obtained from the literature (Sourirajan, 1977; Sourirajan, 1980; Merten, 1966; and Lonsdale and Podell, 1972).

In the actual process, saline water is pumped to pressurize it against a membrane in a closed container. Figure 5-1 illustrates the basic components of a reverse osmosis system. As pure water from the feed solution passes through the membrane, the remaining solution becomes more and more concentrated. At the same time this, a valve (at point A) allows a portion of the feedwater to be discharged without passing through the membrane. Without this discharge (or blowdown), the concentration of dissolved salts in the feed solution would continually increase requiring the pump to add ever-increasing energy to overcome the increased natural osmotic pressure. In addition, precipitation of supersaturated constituents in the brine (and other complications) would occur. Reverse osmosis is often compared to filtration and, in fact, is sometimes referred to as hyperfiltration. Although reverse osmosis can be considered something of a molecular filter, experts disagree on the exact mechanism of differential passage between salts and water. Furthermore, several characteristics differentiate this process from conventional filtration: (1) the production of two output process streams, brine and product, in contrast to filtration, which has only product; (2) the lack of a build-up of rejected material at the separation surface (membrane); and (3) the existence of a flowstream parallel to the separation surface instead of perpendicular to it as in filtration.

Overall System

A reverse osmosis system consists of four major components/processes, as shown in Figure 5-2: (1) pretreatment, (2) high-pressure pump, (3) membrane assembly, and (4) post-treatment for stabilization.

Pretreatment. The incoming feedwater is treated so that it is compatible for use with, and the protection of, the membranes. This treatment usually consists of the removal of suspended solids, the adjustment of pH, and the addition of a threshold inhibitor for controlling scaling due to calcium carbonate and sulfate.

High-Pressure Pump. The pump energizes the pretreated feedwater to the pressure appropriate for the membrane and feedwater being used.

Membrane Assembly. The semipermeable membranes inhibit the passage of dissolved salts while permitting almost salt-free water to pass through. Feedwater applied to the membrane assembly results in a freshwater product stream and a concentrated brine reject stream. No membrane is perfect in its rejection of dissolved salts, so a small percentage of salts does move through the membrane and appears in the product.

Stabilization. The product water from the membrane assembly usually requires pH adjustment and/or degasification before being transferred to the distribution system for potable use.

Membranes

Membranes and membrane elements used for reverse osmosis usually operate in the pressure range of 250 to 1,000 psi [17 to 68 atm].

Several characteristics of membranes are important for commercial utilization. An effective membrane must be able to (1) withstand high pressures; (2) permit the flow of large amounts of water through the membrane relative to the volume they occupy; (3) reject or at least restrict the passage of dissolved solids; (4) remain physically and chemically stable in a saline aqueous environment; and (5) be readily manufactured with reproducible characteristics. Again, there is no perfect membrane, and the ones in use today vary in quality with regard to these characteristics.

The earliest work on osmosis used naturally occurring animal membranes. These were supplemented by artificially produced membranes and, by the late 1940's, cellulose and cellophane-based membranes were being used in osmotic experiments related to the studies of biochemistry.

Membrane Materials. Membranes are made from a variety of materials. Cellulose acetate was used in much of the original development work by Loeb, Reid, Sourirajan, and others in the 1950's and 60's. The early cellulose acetate membranes had some drawbacks due to their lack of hydrolytic stability and their tendency to lose flux by compaction. At the time, there was no recognition of biodegradability of membranes.

The Loeb-Sourirajan membranes were made of cellulose diacetate (actually about 2.5 acetate) which had an acetyl content of 38 to 40 percent. The important factor was incorporation of a "swelling" agent in the acetate-acetone solution like magnesium perchlorate (later formamide), jelling (fixing) the membrane in a low temperature bath and then heat treating to form a rejection layer.

The original cellulose acetate material has been largely displaced by various blends or derivatives of other cellulose acetates, polyamides, and other polymers, etc., which have been found suitable for commercial applications. However, no perfect membrane material has been found as each one has both advantages and disadvantages.

The early membranes of Reid and Breton were dense films (same structure all the way across) and had very low fluxes (Breton, 1957). Loeb and Sourirajan developed skinned (asymmetric) cellulose acetate membranes and actually led into commercial development of the process (Loeb and Sourirajan, 1961).

Most, but not all of the membranes now produced are asymmetric in cross section in that they have a variation in physical structure from one face to the other. These membranes are not reversible in action, i.e., salts are rejected efficiently only when a saline solution is applied on the surface layer and not the other way around.

Structure. There are two basic types of membranes used in reverse osmosis: the fixed membranes and dynamic membranes. The overwhelming majority of the membranes now being utilized in reverse osmosis have a fixed membrane structure. All membrane structures have two sequential components. The first is a surface (often called a rejection layer) exposed to the saline solution, which determines the characteristics such as salt rejection and water passage (flux). The other component is a porous support which supports the aforementioned surface and at the same time permits the transfer of water through it.

i) Fixed Membranes--With fixed membranes these two components are permanently joined together. In some instances, the surface layer can be made from the same material as the porous base, in which case the surface is usually altered during or after the forming process. Most of the polyamide and cellulose-based hollow fine fiber and sheet membranes are made in this fashion. The depth of surface which has been altered is a skin usually only about 0.2% to 0.5% of the thickness of the membrane itself.

Thin film composite membranes represent a different and relatively new technology which has been commercially available since about 1977. In making these membranes, a special, separate rejection layer is incorporated on the surface of the porous base (support) by polymerization or other means. In the case of Universal Oil Products (UOP's) PA-300 spiral membrane, this surface layer is only about 0.000002 inch [0.00005 mm] thick, compared to a 0.002-inch [0.05-mm] overall thickness of the membrane including the polysulfone base to which it is attached. Although thin film composite membranes can be more complex to produce commercially, they allow materials to be combined so as to optimize the rejection properties of the surface layer with the compaction-resistant characteristics of the base, rather than the compromise of both characteristics which occurs in the use of one material for both.

ii) Dynamic Membranes--With dynamic membranes the two components are produced separately and joined as part of the operating process. The membrane structure is produced by adding certain chemicals to the feed water as it flows through porous piping. These chemicals form the rejection layer on the permanent porous substrate. This type of membrane is mainly used for industrial waste applications.

Since the dynamic membranes have not been commercially developed to any extent and do not have widespread use at this time, especially for water supply applications, the remainder of this chapter is almost entirely devoted to discussion of fixed membranes.

Configurations. The fixed membranes considered for reverse osmosis have been produced in the four basic configurations discussed in the following paragraphs.

i) Plate and Frame--The plate and frame membrane configuration was offered for sale commercially in the USA about 1966. The membranes were placed between circular plates which were stacked in a large high-pressure vessel. The unit resembled a filter press oriented in a vertical position.

Extremely heavy, bulky, and inconvenient to operate, these membranes were not a commercial success and were soon displaced in the USA by tubular and other membrane configurations.

They are now produced by some manufacturers in Europe such as The Danish Sugar Company (DSS) in Denmark and GKSS in Germany, but with a different mounting arrangement. They are being used in brackish water and seawater applications but they do not make up a significant amount of the world's installed capacity. A diagram of a plate and frame membrane currently made by a German firm is shown on Figure 5-3. This new design has eliminated the need for a massive high-pressure vessel.

ii) Tubular--The development of the tubular membrane configuration solved some of these problems and permitted the production of membranes for commercial use, in about 1965. The membranes are tubular in shape with a diameter of about 0.3 to 1 inch [0.7 to 2.5 cm] and are formed or inserted on the inside of rigid tubes or pipes. Generally, the feedwater is pressurized on the interior of the tubular membrane and the desalted permeate moves through the membrane to be collected from the outer surface of the tube, as shown on Figure 5-4. Provisions, such as holes and/or porous supports, are made to permit the passage of water once it passes through the membrane. The tubes are normally in discrete diameters corresponding to tubing and piping used for plumbing. The most common commercial membrane is 1/2 inch [1.3 cm] in diameter. The tubular membranes have been arranged in a number of different ways in order to collect the desalted permeate. Arrangements vary from individual tubular membranes suspended above individual permeate collector channels (see Figure 5-5), to Fluid Systems' Osmotik® system where 18 tubular elements (about 7 feet [2.1 m] long) are placed in a larger-diameter (usually about 4 inches [10.2 cm] ID) pipe called a shroud, which collects the permeate as it weeps through the outside of the tubes. All of these configurations require a large physical volume relative to the surface area of membrane that can be incorporated.

Tubular membranes have not been a commercial success for large-scale water production in the USA due to, among other factors, the cost of membrane production and the large physical volume to production ratio. They are, however, being used for certain specialized water and wastewater treatment purposes, since they can be mechanically cleaned and can be operated with a turbulent flow to minimize scaling and fouling.

Tubular membranes are used only rarely in the USA for water production. However, because of their simplicity, ease of production, rugged construction, and ability to be cleaned after fouling, they may become feasible for use in developing areas. Tubular units have been built and operated in India, Mexico, and other countries for potable water production.

iii) Spiral--The spiral membrane configuration uses flat sheet membranes (such as those in the plate and frame configuration), but they are combined in a configuration which eliminates many of the problems that made the plate and frame membranes uneconomical. Spiral membranes were commercially available in about 1966-1967, and they rapidly displaced the tubular membranes for use in water production.

Figure 5-6 shows the construction of this type of configuration. The membrane is cast on a fabric support and then two of these fabric-supported membranes are glued together with a porous material. This porous material (usually a resin-impregnated Dacron tricot) provides a route for the product water, which has passed through the membrane. The membrane "sandwich" or leaf is glued together on three of its four edges. The fourth edge is joined to a tube which acts as a collector for product water. Usually 2 to 26 membrane leaves are attached to this central tube. Between each pair of leaves, a feedwater channel spacer is inserted. This spacer is a mesh screen designed to provide a passage through which the feedwater reaches the membrane surface and the remaining brine exits.

The leaves and spacers are then rolled to form a compact cylinder. Each of these cylinders forms a unit called an element or module, which ranges from approximately 2 to 12 inches [5.1 to 30.5 cm] in diameter and 12 to 60 inches [30.5 to 152 cm] long (see Figure 5-7) depending on the manufacturer and model. Two to six of these elements are usually placed in series in a long tubular pressure vessel to make up a single production unit.

Major producers in the USA of those membranes include Desalination Systems, Envirogenics, Fluid Systems, and Hydranautics.

iv) Hollow Fine Fiber--The hollow fine fiber (HFF) membranes were first put into commercial production in about 1969. They are produced as long capillary tubes with the diameters approximating that of a human hair. As shown on Figure 5-8, saline water is applied to the outside of the tubes and desalinated water is collected and carried away in the inner capillary. The capillary tubes are made with thick walls (the ratio of the outside to the inside diameter is about 2:1), giving the fibers the strength to resist the high pressures involved in reverse osmosis without additional physical support.

For commercial application, the fibers are arranged as a bundle of short loops tightly packed inside a sealed cylindrical pressure vessel. This assembly is called a permeator. Saline water is introduced into the vessel and circulates around the fibers while the product water is collected at one end of the vessel from the open ends of the fiber loops. Figure 5-9 shows a permeator.

Major producers in the USA of these membranes are DuPont and Dow.

Membrane Characteristics. Some of the important characteristics of membranes include flux, salt rejection, and recovery.

i) Flux--One of the most important characteristics of a reverse osmosis membrane is the quantity of water which can flow through the membrane. This is described by the following simplified equation:

$$F_w = A\ (P_m - \pi_m) \qquad \text{Equation 5-1}$$

Where:

F_w = Water flux [gal/ft²/day or cm³/cm²/sec]

A = Water transport coefficient [gal/ft²/day/psi or cm³/cm²/sec atm]

P_m = Pressure differential across the membrane [psi or atm]

π_m = Osmotic pressure differential across the membrane [psi or atm]

The water transport coefficient is determined by many factors, including the type and thickness of the individual membrane. The flux and salt passages for any membrane are measured at a reference temperature, pressure, and concentration. The flux and rejection can then be calculated under actual operating conditions.

By examining Equation 5-1, it is evident that for any given membrane, the flux will be proportional to the net applied pressure. The osmotic pressure for moderately brackish water is approximately 10 to 50 psi [0.7 to 3.4 atm], compared to 350 to 400 psi [23.8 to 27.2 atm] for seawater. Hence, in order to obtain reasonable fluxes, brackish water systems must be operated in the range of 250 to 400 psi [17 to 27.2 atm]. In order to decrease operating pressures, it will be necessary to develop membranes with more efficient transport coefficients; however, due to the actual solution osmotic pressures, there is a limit to how far pressures can be reduced.

In order to permit economical desalination of water, the flux must be relatively high so as to keep the cost of membranes at a reasonable level.

There is a considerable difference between the fluxes of the spiral and the hollow fine fiber membrane configurations, but this is offset to some degree by the amount of membrane surface area available in a given volume and the cost of producing the various types of membranes.

	Flux		Membrane Area Per Unit Volume
	gal/ft^2/day	m^3/m^2/day	ft^2/ft^3 or m^2/m^3
Spiral	15-30	0.6-1.2	300
HFF	2-3	0.8-0.12	5,000

Ultimately, the economics will be determined by the cost of the various membranes available and the power costs involved.

ii) Salt Rejection--Since the purpose of desalination is to separate salts from water, the ability of the membrane to reject dissolved salts is critical. Since no membrane is perfect in its rejection of solute, they all permit the transport of some salts through the membrane. This salt transport is defined by Equation 5-2.

$$Fs = Ks\ (Cs - Cp) \qquad \text{Equation 5-2}$$

where:

Fs = The salt flux [lb/ft^2/sec or g/cm^2/sec]

Ks = Salt transport coefficient [lb/sec or cm/sec]

Cs = Concentration of salt in the feed [lb/ft^3 or g/cm^3]

Cp = Concentration of salt in the product [lb/ft^3 or g/cm^3]

Each membrane and configuration has its own salt transport coefficient, Ks, which is measured at standard test conditions. As the equation indicates, there is no direct relation between pressure and salt passage.

Rejection is defined as follows:

$$\text{Salt Rejection} = \left(1 - \frac{\text{Product Concentration}}{\text{Feedwater Concentration}}\right) \times 100\%$$

Salt rejection is determined by measuring the total dissolved salts (TDS). Individual constituents in solution have different rejection levels depending on the membrane, constituent, and local environment.

iii) Test Conditions--Test conditions used by various manufacturers vary. A comparison of test conditions used by two different membrane manufacturers is shown below. Some of these are due to manufacturers' preferences while others are necessary due to the inherent differences between membrane configurations.

Test Condition	Spiral Element (UPO)	Hollow Fine Fiber Permeator (Dow)
Feed solution (mg/l of NaCl)	2,000	1,500
Pressure applied (psi/atm)	420/28.6	400/27.2
Solution Temperature (°F/°C)	77/25	77/25
Water Recovery (%)	10	75

Note: UOP = Fluid Systems Division of UOP; Dow = Dow Chemical USA.

A more complete list of the test conditions used by these and other membrane manufacturers is shown in Appendix B.

iv) Recovery--Recovery in a reverse osmosis system can be defined as:

$$\text{Recovery} = \frac{Qp}{Qf} \times 100\% \qquad \text{Equation 5-3}$$

where:

Qp = Product water flow (gpd or m^3/d)

Qf = Feedwater flow (gpd or m^3/d)

The higher the percentage of recovery, the greater the conversion of saline water into freshwater. Normally, the percentage of feedwater recovered in a membrane assembly[1] for a brackish water reverse osmosis system is about 45% to 55%. Thus, every 2 units of raw feedwater entering a plant membrane assembly will yield about 1 unit of product water and 1 unit of brine waste. The percent recovered with brackish water reverse osmosis can be increased to as much as 85 to 90 percent using multiple staging, which is explained in the following pages.

For a seawater system, the recovery factor is generally about 20 to 35 percent.

During operation, the recovery factor is determined by controlling the setting of the brine discharge valve. Limitations on recovery are governed by the concentration of the brine, the flux of the membrane, the available pressure, and the desired flow rates in

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[1] A membrane assembly means: (1) for hollow fine fiber systems, a single pressure element, and (2) for spiral systems, a single pressure tube containing 4 to 6 membrane elements.

various portions of the membranes. The recovery is normally increased by adding cascading stages to the system. Systems generally have one stage or two cascading stages and occasionally a third stage. The latter increases the overall recovery to about 85 to 90 percent.

Process Configurations

There are four basic process configurations in which membranes can be arranged for the production of desalted water: single, parallel, reject staging, and product staging. These are shown on Figures 5-10 and 5-11 and are discussed in the following paragraphs.

Single. The simplest configuration for a small system. It is limited in production by the capacity of the membrane assemblies available.

Parallel. Capacity is added to the process by adding membrane assemblies in parallel to increase production. This modification does not change the overall salt rejection, nor the percentage of recovery of the system.

Reject Staging. Also referred to as a multiple-stage, cascade, pyramidal, or tapered array configuration, it is used to increase the recovery of a system. The reject (brine) from one stage is used as feedwater to a following stage so as to recover additional water. Due to the high pressure in the reject stream, no additional pumping is necessary between stages. This configuration entails a slight sacrifice in salt rejection and an increase in overall power usage compared to a parallel system with the same feed rate, although the power per unit of product water is lower.

The quality of the feedwater must be such that the brine delivered as feed to the additional stages does not cause scaling or other problems with the membranes in the latter stages.

Product Staging. This system, which is well-suited to the desalination of seawater, is actually two separate process trains run in series. The product from the first stage is used as feed to the second stage, thus allowing the first stage (using seawater membranes and high pressures) to reduce the TDS of the water to a moderate level. The second stage then can use brackish water membranes to produce a lower TDS.

Operation of the second stage is simplified because passage through the membranes in the first stage prepares the water, so that very little pretreatment is necessary and thus high recoveries can be achieved. The reject from the second stage can be blended with the raw feedwater for reprocessing.

COMMERCIAL DEVELOPMENT

History

Although their osmotic properties were studied by various people during the past 100 years, membranes were hardly ever used for water treatment. One of the early indications of the use of both the name "reverse osmosis" and the concept for water treatment occurred in Pennsylvania, where a patent was issued under the name "reverse osmosis" for a softening process using a ferro-cyanide membrane on a porous porcelain support. Evidently, this process underwent no commercial development.

Working at the University of Florida in 1952, Reid and Breton demonstrated that a cellulose acetate membrane could be used to desalinate water by reverse osmosis. They produced a dense film type membrane which had good rejection characteristics but a low water flux (Breton, 1957). This was about the same time that electrodialysis was first being commercially introduced in the USA by the Ionics Corporation in Boston.

During the 1950's, Loeb and Sourirajan conducted research work treating Millipore membranes so that one of their surfaces was altered, producing a membrane which combined salt rejection and a high flux. Further work by Sourirajan and others culminated in the production of an asymmetric cellulose acetate membrane, which formed the basis for commercial development (Loeb and Sourirajan, 1961).

Between 1966 and 1968, desalination systems using the plate and frame, tubular, and spiral configurations were commercially available. Of the three, the spiral proved to be the most commercially viable and was largely developed by the General Atomic Corporation (now UOP) in San Diego under a contract with the U.S. Office of Saline Water.

Both Dow Chemical and DuPont worked on hollow fine fiber membrane development. In the early 1970's, DuPont introduced and began to commercially produce membranes using asymmetric polyamide material. Dow's efforts during this period were directed toward the commercial production of membranes based on cellulose triacetate.

Presently, only spiral and hollow fine fiber membranes are generally used commercially, and they are widely available for the desalination of water for potable purposes. Some work is still being carried out with the plate and frame configuration in Europe, but its use is relatively minor at this time. Tubular membranes are still produced in small quantities for special applications, usually involving wastewater treatment, but they are not produced widely nor aggressively marketed.

Early in their development, membranes were commercially produced only for brackish water desalination. Seawater membranes require much higher pressures and salt rejection capabilities, and took longer to develop. They were developed experimentally in the

1970's, and several small demonstration units were built. During the period 1977-1978, companies such as DuPont and Fluid Systems began to produce seawater membranes in large quantities. DuPont's seawater permeator uses a polyamide hollow fine fiber membrane, and UOP's uses a polyamide thin film composite membrane with a spiral configuration. Other companies such as Desalination Systems, Dow, and Envirogenics now also make seawater membranes.

The largest installations to date are a 3.2-mgd [12,100-m^3/d] plant in Jeddah, Saudi Arabia, which uses spiral modules, and a 0.8-mgd [3,030-m^3/d] plant in Punta Moron, Venezuela, which employs the hollow fine fiber design. The plants began operation in 1978 and 1980, respectively.

Extent of Usage

According to the U.S. Government's latest published inventory of desalting plants installed worldwide, as of January 1977 there were about 690 reverse osmosis plants with individual capacities greater than 0.025 mgd [95 m^3/d] with a total installed capacity of about 167 mgd [632,000 m^3/d] (El-Ramly and Congdon, 1977). This capacity represented about 17 percent of the land-based desalination plants at the time of the inventory.

ENGINEERING CONSIDERATIONS

Design

Although there generally is a standardized sequence of components for reverse osmosis systems, each situation must be studied separately to ensure that the pretreatment, membranes, and recovery desired are suited for the application.

Normally, reverse osmosis systems are designed, constructed, and installed by original equipment manufacturers. Systems generally include the pretreatment, high pressure pumps, membranes, and post-treatment portions of the facility, plus the interconnecting piping, instrumentation, valves, and power connections.

Many original equipment manufacturers, or their representatives, also design and/or construct the associated building, storage tanks, and distribution pumps, and develop the raw water supply. General contractors with standard plans and specifications are also used.

Frequently, especially with large projects, engineers are retained to assist in the design and procurement of reverse osmosis systems.

Regardless of who is involved with the design, it is important that the following major items be judiciously considered.

Quantity and quality of the raw water source.

Desired quantity and quality of the product water.

Pretreatment.

Post-treatment.

Brine disposal.

Recovery.

Pilot plants.

Safety factors included in plant design and operation.

Raw Water Source. Defining the characteristics of the raw water supply to be utilized is the most important item in the design of a reverse osmosis system. The chemical and physical characteristics and their variability should be thoroughly investigated. Where possible, the sources should be thoroughly tested and/or fully developed before design is completed. The proper design of the pre- and post-treatment facilities and the selection of the appropriate membranes is dependent on an accurate assessment of the raw water quality.

Whenever ground-water sources are to be used, existing wells, test wells, and/or the final production wells must be tested to predict long-term yield and quality.

Surface-water sources, including seawater can be subject to considerable variations in quality. Thus, it is essential that samples be taken under varying conditions with regard to such factors as tides, storms, wind directions, and seasons to ensure that all potential water quality conditions are determined.

Samples should be taken under conditions as close to potential operating conditions as possible. Although sampling and subsequent analysis may appear expensive, such procedures are usually much cheaper than unpleasant and potentially disastrous surprises after the system is installed and operating.

The information that should be determined in water samples is listed in Table 5-1.

Product Water. Information on the desired quantity and quality of the finished water is needed. The water quality characteristic most frequently used in referring to the quality of the product water is conductivity, which is generally measured using a conductivity meter. The conductivity measurement is then multiplied by a constant to estimate total dissolved solids, commonly referred to as TDS.

Table 5-1
RAW WATER SOURCE INFORMATION TO BE DETERMINED FOR A REVERSE OSMOSIS SYSTEM DESIGN

Temperature (maximum, minimum, and average)
Chlorine residual
Turbidity
Suspended solids
Color
Silt Density Index
pH
Coliform count (bacterial)
Total plate count (bacterial)
Calcium
Magnesium
Sodium
Potassium
Iron
Manganese
Barium
Strontium
Chloride
Sulfate
Nitrate
Ammonia
Phosphate
Fluoride
Silica
Total dissolved solids
Hydrogen sulfide
Carbon dioxide

Note: Variations in these parameters should be noted and odors, traces of industrial or domestic pollution, clay, sand, rust, or other unusual characteristics should be described.

The desired quality of the product water will determine the salt rejection required.

Other product water information has to do with the water usage requirements such as average daily and maximum daily flows, time periods when the water is needed, required pressure, and available storage.

Recovery. The selection of an overall recovery factor for the desalination system is a key design parameter. The normal recovery level for brackish water is about 45 to 55 percent of the feedwater per stage. With reject staging (cascade, tapered or pyramidal arrays) recoveries of 75 to 90 percent are possible with certain feedwaters using only a marginal increase in total energy compared to the energy requirement of a single-stage unit. Capital costs per unit of product water are also generally reduced.

Although theoretically there is an economic incentive to use a high recovery system to reduce energy costs, this advantage must be weighed against the disadvantages. The major disadvantages include: (1) the increased complexity of the system including pretreatment and, more importantly, (2) the increased potential for fouling and scaling of the membranes due to precipitation of sparsely soluble salts. The potential for the latter increases as the concentration of the brine increases for each successive stage. This is, of course, highly dependent on the chemical constituents in each individual water source and must be examined case by case.

Aside from economics, other factors contributing to the desirability of higher recoveries could include a restricted raw water supply or the need to minimize the quantity of brine being discharged.

The utilization of dependable and appropriate pretreatment procedures (usually including the use of threshold inhibitors) can permit markedly high recoveries, but planning and design should be carried out while carefully weighing the potential benefits of high recoveries against the risks.

Brine Disposal. The method of disposal of the concentrated brine stream which leaves the reverse osmosis system needs to be ascertained early in the conceptual planning of a project. The brine in a brackish water reverse osmosis plant can amount to 10 to 50 percent of the water produced and, where it can be discharged to the sea or other saline body, it is generally not a problem. However, since the brine has the potential for polluting the ground water and causing other environmental problems if discharged improperly, its disposal should be carefully considered. The necessity for special disposal techniques could make the system very costly.

Safety Factors. It is only prudent that a margin of safety be designed into a reverse osmosis system. This should be a concern when a high-recovery plant (more than 60 percent recovery for brackish water and 30% for seawater) is considered. Although high-recovery plants can be more energy efficient (and there are many examples of successful installations), they can pose the potential problem of precipitation of sparsely soluble salts in the membrane elements due to the high concentration factors that often exist in these plants. Examples of safety factors could include special pretreatment, reduced levels of recovery, and the addition of threshold inhibitors.

Once a heavy precipitation of material such as calcium sulfate or calcium carbonate occurs, more money can be lost due to membrane element failure than could have been gained through increased recovery.

Pilot Plants. Where conditions are unusual or high recovery is being considered, it is often worthwhile to run pilot-scale experiments using the membranes and pretreatment being considered. The pilot plant must be designed to test performance under conditions that would provide the basis for the actual design.

Design Review. Certain design information on available membranes and a general checklist for the review of reverse osmosis plants are shown later in this section under the heading "Design Information."

Major Operating Problems

Most failures in reverse osmosis plants occur because materials are deposited on the membrane surfaces or in the membrane elements, preventing the elements from functioning efficiently. Other problems occur due to mechanical failures, membrane failure, and poor operation. The most important factor in the successful operation of a reverse osmosis plant is the training and attitude of the operator.

Fouling. Fouling is the deposition of materials within the plant resulting in reduced performance of the system. These deposits are from four major sources:

Precipitates.

Colloids.

Particulates.

Microorganisms.

The membrane surface is especially sensitive to fouling, which can reduce the water flux to a major degree.

i) Precipitates--During the reverse osmosis process, the salts remaining in the water on the feed side of the membrane are concentrated as water passes through the membrane. The concentration can often be two or more times that of the feedwater. These concentrations often exceed the solubility limits of some of the dissolved salts and thus they begin to precipitate. Due to the molecular separation process taking place at the surface of the membrane, the layer of water adjacent to the membrane becomes even more highly concentrated than the general brine stream. This phenomenon is referred to as concentration polarization and can act as an initial cause of precipitation at the membrane surface. Once precipitation begins, the precipitates act as nuclei for formation of additional precipitates.

ii) Colloids--Colloids are particles less than one micron in diameter. The major colloids of concern for reverse osmosis are organic compounds, heavy metal colloids, and silica.

These colloids usually cause problems by coagulating during the reverse osmosis process and precipitating on the membrane surface and affecting the boundary layer.

iii) Particulates--Undissolved solids such as sand, rust particles, plant material, algae, etc., can clog the feedwater passages and the membrane surfaces. These types of particulates are often found in surface waters. Usually, they are minimal in ground water from properly constructed and developed wells.

Another source of particulates is the oxidation of constituents dissolved in the feed stream. Divalent iron and manganese, and hydrogen sulfide are characteristic of these substances. These can be oxidized under conditions in which dissolved oxygen and pH levels can result in the formation of precipitates before the feedwater enters the plant. The precipitates can coat membranes and cartridge filters and can clog the small flow channels in the permeators or membrane elements.

iv) Microorganisms--Most bacteria are 0.00004 to 0.00012 inch [1 to 3 microns] in size and can enter a pressure vessel but cannot pass through a reasonably perfect membrane. Certain species of bacteria can establish themselves on the surface of the membrane or in the other parts of the membrane assembly and utilize constituents in the feedwater and/or the membrane itself as a source of energy. Some microorganisms can cause biodegradation of some membranes and thus alter their flux and salt rejection characteristics, whereas other microorganisms and their byproducts can cause fouling and/or alteration of the membrane surfaces. Imperfections in the membrane can permit microorganisms to move through the membrane and establish themselves on the permeate side, where they can also create problems.

v) Other--Other causes of fouling can be oils, greases, organics, etc.

Mechanical Failures. Due to the high pressures needed for the transport of water across the membranes, the piping, supports, machinery, etc., can be subjected to major mechanical stresses such as high pressures and vibration. Pumps at these pressures tend to be high speed, especially for seawater, and can cause significant vibrations depending on the pumps, their mounting, and the design of the system. This vibration places stresses on joints, instrumentation, pipes, etc. A well-designed plant will have minimal vibration-related problems.

Corrosion. Corrosion of materials is a significant factor in reverse osmosis facilities. The feed, brine, and product streams can all be corrosive. Many of the chemicals used in the treatment process such as acids and bases can also be sources of corrosion.

Membrane Failure. Although operating problems may focus on the failure of a membrane to function properly, they often have other causes, usually a problem in the pretreatment system.

The manufacture of membrane assemblies requires a number of components, materials, chemical reactions, and production steps. This results in some variations between individual membranes from the same manufacturer. Overall, however, the quality of membranes has been good and most manufacturers are willing to stand behind their product. Guarantees on membrane life, of course, are based on certain operational requirements which must be met to in order for the guarantee to be valid.

Due to the different materials and methods used in fabricating membrane assemblies, their operational life and performance can be affected in various ways by the temperature, pH, existence of chlorine, storage conditions, etc.

Poor Operation. Poor operation can destroy even the best units. Operators must be trained in the performance of their duties and the capabilities of their plant.

Pretreatment

The proper preparation (pretreatment) of the water before it reaches the membrane is the key to successful operation of a reverse osmosis plant. As noted above, membranes can be sensitive to pH, temperature, the existence of certain chemicals, etc., and (like the feedwater passages) are highly sensitive to fouling and clogging. Proper design of the system and pretreatment of the water can greatly minimize these problems and hence protect the membranes, which are the heart of the reverse osmosis system.

Pretreatment and post-treatment (or stabilization) processes involve simply the application of standard water treatment and chemistry techniques. Since the flows in most reverse osmosis systems designed to date are not large, the pretreatment techniques and components are generally adaptations of those used on small package water treatment plants rather than full-scale unit operations seen in large water treatment systems. In many cases, especially with ground water, pretreatment may be unnecessary or at least relatively simple, involving only in-line chemical addition, filtration, etc.

The following paragraphs discuss the parameters affected by pretreatment and also discuss the safety filter used in virtually all systems.

Precipitates. Precipitation can be minimized by reducing the concentration of the constituents subject to precipitation such as calcium, iron, etc., that are exposed to the membrane. This can be done by prior removal through softening, ion exchange, and/or aeration-filtration or by controlling the recovery rate to reduce the concentration at the membrane. Prior removal of constituents is usually not a cost-effective solution due to excessive capital and operating costs, but reduction of the recovery rate is often employed where sufficient raw water exists.

The most common solution is to lower the pH by adding acid, thereby increasing the solubility of the dissolved substances or reducing their concentration by conversion into a more soluble form. A threshold inhibitor, such as sodium hexametaphosphate (SHMP), is almost always added to the raw water to reduce the chances for precipitation of calcium, barium, or strontium salts. Threshold inhibitors, unlike sequestering agents, do not act stoichiometrically, i.e., on a calculated mole-to-mole basis. The dosage of an inhibitor cannot be exactly calculated but is determined largely by experience. Concentrations used are normally in the range of 1 to 10 mg/l based on the feedwater characteristics. Threshold inhibitors are thought to work by preventing the formation of crystals.

Colloids. Colloids are particles which have their largest dimension less than about 0.00004 inch [1 micron]. They pass through standard filters and cannot be readily removed by ordinary sedimentation processes.

They are most often present in surface waters and wells improperly developed in clay strata and are usually removed by coagulation followed by filtration. In large systems, coagulants such as alum, lime, and polymers can be used in a standard coagulation, flocculation, sedimentation, and filtration configuration. In some systems, in-line coagulation and filtration can be used. With filtration, ample mixing and reaction time must be allowed after the addition of a coagulant before filtration.

The standard safety filter, which is composed of cartridge filters in the reverse osmosis pretreatment system, is not a reliable device for removal of colloidal particles.

Particulates. Particulates can be removed by sedimentation and filtration methods used in standard water treatment plants. Ideally, the best policy is to prevent particulate contamination altogether, so that removal is unnecessary.

This is most difficult when surface water is the raw water source, but particulates can be minimized in such cases by proper construction and placement of the intake structure and by proper selection of materials. For ground water, proper well construction techniques and materials selection can substantially reduce particulates in the feedwater.

Where dissolved substances such as hydrogen sulfide or ferrous salts exist, which can be readily oxidized into particulates (sulfur and iron oxides) by exposure to oxygen, they must be either deliberately oxidized and filtered out or prevented from oxidizing. The latter method is frequently employed and requires that the well and feedwater piping be constructed so that air or other oxidants are not introduced into the system.

Microorganisms. Where microorganisms cause problems in fouling membranes, elements, and permeators, disinfection is used to remove or reduce the microorganisms present. Chlorine has been the most frequently used disinfectant. To be effective, chlorine must be mixed with the water and allowed adequate contact time between itself and the microorganisms. Chlorine can be added using standard equipment used for water treatment.

The microbial population can be controlled in the reverse osmosis system in three major ways. The first is to prevent microorganisms in the feedwater to begin with. This is most feasible with a properly engineered and constructed well which has been thoroughly disinfected after construction and before use. The second is to disinfect the water before it enters the membranes, and the third is to disinfect the water after it passes through the membrane.

Only the first two methods will protect the system from microbial fouling. As the most common disinfectant used is chlorine and since membranes differ in their compatibility with chlorine, disinfection must be performed carefully. Cellulose acetate derivatives can tolerate chlorine in reasonable doses, but polyamides usually cannot. In this latter case, if residual chlorine is present, the feedwater must be dechlorinated before being applied to the membranes.

In some systems, a chlorine residual (combined) of about 1 mg/l in the feedwater to the membranes is continually maintained to minimize microbial fouling, and occasionally that level is raised to 9 to 10 mg/l as a shock treatment.

Another disinfectant that can be used is sodium bisulfite ($NaHSO_3$). It is an effective disinfectant and may be used intermittently on a regular schedule. The effective concentration range is about 500 to 2,000 mg/l. Hydrogen peroxide (H_2O_2) is also used for periodic treatment.

Safety Filter. In almost every type of reverse osmosis system, an in-line pressure filter is placed just before the membranes as part of the pretreatment system.

A series of fine cotton or polyethylene cartridge filters with a nominal size of 5 to 20 microns are usually used in the pressure filter.

The purpose of the pressure filter is to protect the membranes from any filterable matter that may have inadvertently passed through the pretreatment process. The safety filter should be equipped with a pressure gauge to indicate the differential pressure drop across the filters and thereby indicate the extent of their fouling.

If the differential pressure across the filters increases rapidly, it is an indication of possible problems in the raw water supply or in the pretreatment process. The filters provide some degree of short-term protection for the membranes while corrective action is taking place.

Replacing cartridge filters more often than every 1 to 3 months usually indicates a problem with the pretreatment. However, the safety filter is not meant to be major component for the removal of high amounts of filterable solids. This would be not only an inefficient use of rather expensive filters but it would probably lead to premature failure of the membrane systems due to the high probability that some of the unwanted material will break through.

Post-Treatment

The product water emerging from the membrane assembly generally needs some type of post-treatment before being distributed as potable water. Such post-treatment includes pH adjustment, usually by the addition of a base, removal of dissolved gasses such as H_2S and CO_2 by air stripping, and/or disinfection.

Materials of Construction

Materials of construction must be selected to be compatible with the pressures, potential for corrosion, and vibration that exist in the system.

Pressures. The reverse osmosis system is a dual-pressure system--the membrane section is operated at high pressure, and the pre- and post-treatment sections are operated at low pressures (generally below 70 psi [4.8 atm]). For the latter sections, low-pressure fittings and construction techniques can be used. Polyvinyl chloride (PVC) piping is often employed in these sections.

Care must be used with PVC and other plastic pipe to protect them from direct intense sunlight, vibration, physical damage, excess heat, and water hammer. The connection of metal to plastic at pumps or to other pipes can create problems especially where (e.g., at pumps) the metal portion may vibrate excessively or operate at higher temperatures. Water hammer caused by cycling pumps in ground-water delivery or other systems can cause problems, as can improperly supported large-diameter bends.

The section from the high-pressure pumps into the membrane pressure vessels must be designed and constructed to withstand pressures from 300 to 1,000 psi [20.4 to 68 atm] depending on whether brackish water or seawater is being processed. In this section, 316 (or better) stainless steel, high-pressure flexible hoses, and lined or coated piping have been used.

Corrosion. Due to the high TDS of the feedwater and brine stream, plus the potential instability of the product water, materials must be selected carefully to avoid both their damage by corrosion and damage to the membranes due to fouling by corrosion products. For these reasons, polymeric materials, stainless steel, and polymeric liners and coatings are widely utilized.

In most cases, the pH of the feedwater is reduced upstream of the high-pressure pumps and thus the pumps should be carefully selected for resistance to corrosion. Stainless steel (impeller and housing), Ni-resist in the proper grade, and other metals have worked satisfactorily, whereas for seawater both stainless and aluminum bronze have been used. In at least one case pumps made of conventional materials were used, and acid for pH control was injected into the high pressure feedline with good results.

Another factor to consider is the corrosive atmosphere in and adjacent to desalination plants, caused by (1) spills of feedwater, brine, sulfuric acid, caustic soda, or soda ash, etc., (2) salts transported by winds (affecting plants near seashores), and (3) hydrogen sulfide from ground waters. This corrosive atmosphere can be extremely hard on motors and instrumentation. Electric motors protected against moist and corrosive atmospheres should be used.

In the design of seawater plants in particular, care must be taken to avoid situations where stagnant seawater will remain in contact with stainless steel as serious corrosion can occur.

Vibration. Inherent in the use of high-pressure pumps is some vibration which can create stress on piping, machinery, control systems, and their directly or indirectly connected instrumentation. This can be especially noticeable on small installations where all the components are mounted on a single skid.

Long-Term Operation

Flux Decline. When operating at design pressure, the water passage or flux of the membrane can be expected to decrease with time, due to fouling and/or the densification (or compaction) of the membranes which occurs at elevated pressures. This decrease affects many membranes and steadily lowers production over the life of the membrane. Thin film composite membranes, however, are apparently not as seriously affected by membrane compaction.

Figure 5-12 shows the predicted loss of flux for a hollow fine fiber seawater membrane during a 3-year period at various temperature levels at a uniform pressure of 800 psi [54.4 atm]. Consideration of flux decline is important in specifying and designing the capacity of the system desired, because the initial capacity will be significantly higher than the capacity at the end of several years. It is usually prudent to specify the design capacity which will be required at the end of 3 or 4 years.

Records. Another key item in successful operation of reverse osmosis systems is the maintenance of performance records. Manufacturers generally specify data which should be collected by the operator. Much of this data collection is required by the manufacturer in order to keep the membrane guarantee valid.

Monitoring the operational data enables the operator to ascertain whether the system is performing as per specifications and to spot problems before they cause serious damage.

Membrane Cleaning. In almost all systems, no matter what pretreatment system is used, some membrane fouling occurs. The rate of fouling varies from system to system depending on conditions, pretreatment, etc. One remedial action that can be taken is the cleaning of the membranes to remove the fouling material, thereby restoring, or at least improving, the system's production capacity.

Cleaning procedures and reagents are recommended by membrane manufacturers and should be followed carefully. Since membrane materials and configurations vary between manufacturers, only cleaning instructions for the specific membrane should be used. In some cases, such as for DuPont's B-10 seawater permeator, certain chemicals must be added at startup and after cleaning to maintain the membranes rejection characteristics. Hence, the importance of following manufacturer's instructions.

In cases of well-designed systems with normal water sources, several years may pass before cleaning is required.

Flushing. To improve long-term performance of a reverse osmosis system, the membranes should be flushed with fresh water during periods of shutdown to remove raw feedwater and concentrate. If the raw water remains in the membranes, precipitation may occur. Although this precipitation may be minor, it leaves particles which may act as nuclei for additional precipitation at a later time.

Energy Sources

Requirements. The major energy requirement is for high pressure pumping to the membranes. Electricity is commonly used as the primary energy source, but other sources, such as diesel or steam engines with direct mechanical drives, have also been used. The 3.2-mgd [12,100-m^3/d] seawater reverse osmosis plant at Jeddah, Saudi Arabia, for example, uses direct coupled diesel engines to power the high-pressure pumps.

In addition to the high-pressure pumps, other machinery, such as intake pumps, chemical feeders, and instrumentation also require power. Electric power is the most convenient source for these applications. In general, overall power requirements are about 7 to 12 kWh/kgal [1.9 to 3.2 kWh/m^3] for brackish water reverse osmosis plants and about 30 to 40 kWh/kgal [7.9 to 10.6 kWh/m^3] for seawater plants. The relative costs of available fuel and energy determine the feasibility of energy sources for use.

Energy Recovery. The high pressures (800-1,000 psi [54.4-68 atm]) involved in seawater reverse osmosis have prompted discussion of the use of energy recovery systems to utilize a major portion of the energy in the brine reject stream. This stream is usually discharged at approximately 20 psi [1.4 atm] less pressure than the feedwater stream to the membranes.

The potential for energy recovery is high and recoveries of up to 10 kWh/kgal [2.6 kWh/m^3] of product water may be possible. A variety of energy recovery devices such as impulse (Pelton) turbines, and reverse running pumps have been used for other applications. These can produce rotating energy which can be used to augment shaft work in the plant or to operate a generator to provide electric power. A new type of turbine, called a biphase turbine, has also been suggested, but not yet used, for energy recovery for reverse osmosis on a commercial basis.

In a recent study (Singh and Cabibbo, 1980), the available power, in kilowatts, was estimated for both seawater and brackish water using energy recovery. Some of these data are shown in Table 5-2.

A theoretical analysis of power requirements for seawater reverse osmosis with energy recovery versus dual purpose multistage flash installations has been made by DuPont (Sackinger, 1980). This study is very optimistic about the potential for seawater reverse osmosis on an energy basis.

Despite the potential, there are few energy recovery systems operating with reverse osmosis units at present. It is expected that some installations will be equipped with energy recovery in the future, but practical experience for reverse osmosis facilities at present is practically nil. Since 1979 a power recovery unit based on a flow-work exchanger for reverse osmosis applications has been tested at the OWRT desalination test facility at Wrightsville Beach, North Carolina. Data should be available in late 1980.

Source Development

One of the best investments that can be made in a reverse osmosis system is the development of a good raw water source. The source should be capable of dependably delivering the required quantity of water within the expected quality limits.

Table 5-2
ENERGY RECOVERY FOR REVERSE OSMOSIS

Plant Size (mgd)	Product Water Recovery (%)	Brine Flow Rate (gpm)	Available Pressure (psi)	Available[a] Power, kW		
					Turbine	
				Ideal[b]	Shaft[c]	Equivalent Electric Motor
Brackish Water Reverse Osmosis[d]:						
1	50	694	360	109	96	101
	70	298	350	46	40	42
	80	174	335	26	23	24
5	50	3,470	360	545	480	505
	70	1,490	350	230	200	210
	80	870	335	130	115	120
Seawater Reverse Osmosis[e]:						
1	20	2,780	850	1,027	900	950
	25	2,080	850	770	675	710
	30	1,620	840	590	520	550
5	20	13,900	850	5,133	4,500	4,750
	25	10,400	850	3,840	3,375	3,550
	30	8,100	840	2,950	2,600	2,750

Notes: Table adapted from Singh and Cabibbo, 1980.
Metric conversion factors: mgd x 3,785 = m^3/d; gpm x 5.451 = m^3/d; gpm x 0.063 = l/s; psi ÷ 14.7 = atm.

[a] Recoverable.
[b] Ideal power at 100 percent efficiency.
[c] Shaft power at 88 percent efficiency.
[d] Feedwater pressure at 400 psi.
[e] Feedwater pressure at 900 psi, TDS = 35,000 mg/l and temperature = 77°F [25°C].

Whenever possible, the intake structures or wells should be constructed and tested before finalizing the design of the reverse osmosis system. Proper design and construction of the structure can often reduce the pretreatment required. Conversely, a poorly designed and/or constructed source can lead to inefficient operation and/or destruction of the plant. The two major methods of source development are the use of direct intake structures and the use of wells.

Intake Structures. These can supply raw water from intakes located onshore or offshore. The structures should be tailored to the individual sites to take into account shoreline characteristics, currents, marine growth, seasonal water changes, environmental considerations, offshore geology, maritime traffic, weather conditions, etc. The materials of construction should be compatible with the water source.

Wells. The use of wells for the supply can be very advantageous. Properly structured, a well utilizes natural filtering capacity to produce a water source low in microorganisms and suspended solids. Wells adjacent to surface-water sources such as lakes or the sea can often be used to supply the feedwater.

A small 0.01-mgd [38-m^3/d] seawater reverse osmosis plant at St. Thomas, Virgin Islands, has operated successfully since 1978 using a seawell and only a minimum of pretreatment.

Design Information

Addresses. Table A-1 in Appendix A contains the names and addresses of manufacturers in the USA of various types of desalination equipment, including reverse osmosis equipment.

Membrane Characteristics. Table B-2 in Appendix B shows the general characteristics of many of the membranes available commercially from manufacturers in the USA as of January 1980. This list is for general information only; the characteristics and availability of specific membranes should be checked at the time of design.

Capacities Available. The reverse osmosis process is amenable to a wide variety of capacities from 1 gallon per hour [4 l/hour] up to virtually any capacity. For example, the U.S. Government has under contract a 96-mgd [363,000-m^3/d] brackish water reverse osmosis plant at Yuma, Arizona.

It is primarily the wide selection of membrane capacities available (see Appendix B-2) which permits the construction of almost any capacity plant. The membranes must simply be matched with suitably sized components such as pumps, filters, chemical feeders, etc.

The reverse osmosis process is ideal for modular construction. Although unit costs decrease to some degree as larger plants are specified, reverse osmosis plants do not have the potential for the economies of scale that occur with distillation plants.

Installation. Installation is generally carried out by a contractor, although many small units come self-contained on skids. In order for these plants to be ready for operation, they generally need only (1) connections to sources of feedwater and power, and (2) provisions for product storage and brine discharge.

Others require more elaborate preparation, depending on the characteristics of the specific site and plant size. The area needed for reverse osmosis desalination facilities can vary widely depending on the configuration and the associated ancillary equipment, buildings, rooms, etc. In a large facility such as the 5-mgd [18,900-m^3/d] plant in Cape Coral, Florida, considerable space has been devoted to offices, laboratories, a machine shop, storage, etc. Other facilities are very basic, furnishing only a simple shelter for the unit. Figure 5-13 shows the approximate space required for both brackish water and seawater units produced by one manufacturer. This manufacturer has standardized and skid-mounted their equipment for ease of shipment and construction. Units from other manufacturers vary from this approximation, but it will give a rough idea of possible space requirements.

Table 5-3 shows some operating and installation information for some skid-mounted reverse osmosis units produced by one manufacturer.

CHECKLIST

A general checklist that would be helpful in reviewing a reverse osmosis plant design is shown below. Much of this list is adapted from the DuPont Technical Bulletin 690 (April 6, 1977) and may be specific for the hollow fine fiber elements.

This list does not cover all facets of the design but is meant only as a preliminary aid in review.

Pretreatment

1. Is the water analysis used for designing the system current? Is there a history of seasonal fluctuations?

2. Have sufficient Silt Density Index tests been run to determine whether additional clarification and filtration will be required on the raw feedwater if hollow fine fiber membranes are used?

3. If used, are the acid and sodium hexametaphosphate (SHMP) being introduced before the cartridge filter? (They normally should be.) Is the acid injection point

Table 5-3

OPERATING AND INSTALLATION DATA ON VARIOUS SKID-MOUNTED BRACKISH WATER REVERSE OSMOSIS PLANTS

Nominal Product Capacity (mgd)	Recovery (%)	High Pressure Pump (hp)	Power Usage[a] (kWh/kgal)	Outside Dimensions	Weight	
				Length x Width x Height (ft)	Shipping (lb)	Operating (lb)
0.015	50 - 80	15	16.1	12 x 3.5 x 5.5	1,715	2,181
0.022	50 - 85	20	13.9	12 x 3.5 x 5.5	1,975	2,550
0.030	50 - 85	25	11.5	12 x 3.5 x 5.5	2,235	2,925
0.037	50 - 90	25	9.6	12 x 3.5 x 5.5	2,495	3,300
0.045	50 - 90	30	8.7	12 x 3.5 x 6	2,755	3,672
0.055	50 - 80	50	14.0	15.5 x 4.7 x 6.2	4,040	5,020
0.073	50 - 80	60	11.5	15.5 x 4.7 x 6.2	4,530	5,810
0.092	50 - 85	60	9.5	15.5 x 4.7 x 6.2	5,020	6,580
0.110	50 - 85	75	10.8	15.5 x 4.7 x 6.2	5,510	7,370

Notes: Data from manufacturer's literature "ENRO Standard Packaged Reverse Osmosis Systems" Envirogenics Systems Co.

The equipment specified is a basic skid unit only--cartridge filter, high-pressure pump, membranes and pressure vessel, acid feed system, and connecting piping as well as basic instrumentation and valves. For each individual application additional facilities such as pumps, filters, chemical feed systems, building foundation, storage tanks, and instrumentation may be required.

Metric conversion factors: mgd x 3,785 = m^3/d; hp x 0.7457 = kW; kWh/kgal ÷ 3.785 = kWh/m^3; ft x 0.3048 = m; lb x 0.454 = kg.

[a] Usage per kgal of product. Could vary with recovery.

upstream of the SHMP injection point? (It normally should be.)

4. If plant operating conversion requires SHMP to prevent calcium sulfate precipitation, has a low-pressure flush been considered for shutdowns?

5. Is chlorine fed upstream of the RO unit? If so, are provisions required for removal? If dechlorination is required, has it been provided?

6. In the case of softening pretreatment, what precautions have been taken to ensure proper effluent quality?

Membranes

1. Does each permeator or pressure vessel have provisions for sampling product and reject? (The reject sampler should have capabilities for checking pressure directly at the outlet, thus indicating changes in pressure through the membrane assembly).

2. If two or more HFF permeators are operating in parallel, is flow balancing provided for?

3. Does branch piping to the feed, product, and reject have sufficient flexibility to allow easy field installation, future maintenance, and replacement?

4. Do the membrane pressure vessels or permeators have sufficient structural support? Are they accessible for removal, repairs, sampling, and maintenance?

Filters

1. Are the filter cartridges of the wound, "in-depth" type or the pleated paper type? Either is suitable for filtration purposes; however, the wound type fouls gradually allowing for programmed servicing.

2. Are the housings corrosion-resistant?

3. Have provisions been made to indicate or measure changes in pressure across the filter?

4. Have sample outlets been provided before and after the filter?

5. Has the value of an extra-fine cartridge been reviewed for the particular installation? Note: Generally, 5- or 10-micron cartridges are suitable for most feed streams with hollow fine fiber permeators.

6. Have the filters been sized for the expected flow? Generally, 2-4 gpm/ft^2 [5.7-11.4 l/s/min] is used.

Pumps

1. If the pump is the positive displacement type, have accumulator, relief valve, and back pressure valve on the recycle stream been incorporated into the discharge piping?

2. If pump is centrifugal, has an adequate throttling valve been installed directly in the discharge line?

3. If centrifugal pumps are installed in parallel, has a check valve been included upstream of the throttling valve? Note: The reason for upstream installation is that repairs can be made on the check valve without shutting down the entire system.

4. Also, in parallel installations, are pumps protected by a high-pressure or a high-temperature switch against accidental complete shutoff of throttling valve while the pump is running? Note: The instrumentation indicated should be installed between throttling valve and pump.

5. Are pumps protected against low suction pressure?

6. Are pumps readily removable or accessible for servicing?

7. If a recycle line with back pressure valve is used for pressure control, is the line sized small enough that bypass does not "hide" plant problems? Note: Neither the high-pressure switch nor the relief valve will function to indicate major system malfunction if a large bypass is installed.

Reject Flow Control Valves

1. Is a flow control valve being installed rather than a back pressure valve?

2. Are wetted parts all stainless steel or better?

3. Have the relative merits of automatic versus manual valves been considered?

Piping

1. Has the required feedwater pH been considered in the piping materials?

2. Is all piping between the cartridge filter discharge and the permeators of a noncorrosive material?

3. Have sufficient pipe supports been installed? Note: PVC pipe will probably receive more shock in shipping than in operation. Particular care should be taken at metal-to-plastic connections.

4. Does concentrated sulfuric acid enter into a turbulent or mixing zone? Is the header fabricated of suitable material in the "mixing zone?" Is piping installed to properly disperse the acid?

5. Have sufficient sampling valves or devices been included in headers and subheaders? In general, all collection headers for each stage of permeator product and each reject stage should have samplers.

6. Has galvanic corrosion of mixed piping materials been considered?

7. Assuming that all piping and flanges have been reviewed for pressure rating, has the difference in PVC rating between 75°F and 100°F [23.8 and 37.7°C] been considered?

Instrumentation

1. Are permeators protected by 50-psi [3.4-atm] relief valves in product subheaders terminating in isolation valves or pressurized equipment such as polishing filters?

2. Have pressure gauges been included before and after the filter, in the high-pressure feed header, and in all stages of the reject and product header?

3. Have sufficient meters been included to set the plant conversion with the reject flow control valve? Note: This requires that total product water and final reject water be metered in each control block. It is also preferable to have each stage of product water metered in a multistage control block.

4. Is it necessary to indicate and record product water conductivity or are the portable conductivity meters sufficient? Note: Portable units should be available, at any rate, for troubleshooting.

5. If acid is injected for calcium carbonate scaling control, have pH indicator-transmitter and Hi-Lo pH cut-off switches been integrated into the system? Note: If it is necessary to neutralize the product water, a decarbonator and soda ash feeder may be required.

6. If hexametaphosphate feed is required to control calcium sulfate scaling, has a pressure switch with time delay been installed in the SHMP feed line? Note: This switch would shut down the system if the SHMP feed stops.

7. In systems using a heat exchanger for heating or cooling feedwater, has a high-temperature cutoff switch been installed?

8. Are all cutoff switches "manual reset" to avoid on and off cycling of the plant during hours when the system is unattended?

Cleaning System

1. Have the membrane "control blocks" been designed for cleaning the membranes stage by stage in small enough increments to satisfy customer product water requirements? Note: Since an isolation valve will be necessary between first-stage reject and second-stage feed, two headers will be required though they serve the same process function.

2. Are hoses, valves, and connections rated at the same pressure as the cleaning pump? Note: Cleaning valves in the feed and reject headers must be designed for system operating pressure when closed. Also, the product (permeate) return hoses must be of a material which will not kink and thereby throttle the permeate flow to the cleaning tank or sewer.

3. Have general manufacturer's guidelines regarding pressure, flow, and materials, been observed in the design of the cleaning pump?

4. Does the cleaning tank have sufficient holdup time and does it require and have a cooling coil to retard heat buildup caused by recirculation (for hollow fine fiber systems)?

5. Can the top of the cleaning tank be exhausted to the building exterior in case cleaning agents such as ammoniated citric acid are used?

6. Does the cleaning tank have a mixer, charging hatch, operating platform, and dump valve?

7. Is a cartridge filter included in the reject return line with the same rating as the cleaning pump (for hollow fine fiber systems)?

8. Can cleaning system required flows be handled reasonably by removable hoses or should cleaning system headers and "jumpers" be permanently installed?

STATE OF THE ART

General

The commercial application of reverse osmosis has grown rapidly over the past 10 years. The spiral and hollow fine fiber membranes have emerged as the most commercially viable membrane configurations for large-scale production of potable water. There is now a body of experience with brackish water membranes; indeed, some plants have been operating 6 to 7 years, demonstrating that the membranes can have a sustained working life. Reverse osmosis plants treating up to 16 mgd [60,560 m^3/d] have constructed and operated and plants of up to 96 mgd [363,000 m^3/d] are being designed and constructed.

Currently, the use of reverse osmosis for seawater desalination has entered the commercial phase with the construction of large seawater plants in Saudi Arabia and Venezuela. Hollow fine fiber membranes for use with seawater were introduced in 1974 and have been used in a variety of desalting applications. There are insufficient operational data to confirm the projected life of seawater membranes under commercial conditions, but such data are being developed and from all indications seawater reverse osmosis is expected to be a viable process. Currently, it is the general policy of most manufacturers to offer at least a 3-year prorata guarantee on seawater membranes installed in properly designed and operated plants.

Thin film composite membranes for seawater applications have been commercially available since 1978. Membranes of this type for brackish water which will permit the use of lower pressures to produce high flux should be available in 1981. Their expected service life under commercial conditions has yet to be proven.

Major Technological Advances Since 1972

The major technological advances since 1972 include the following:

1. The commercial development of the hollow fine fiber seawater membrane (DuPont's B-10).
2. The construction and operation of a large-scale wastewater reclamation plant employing reverse osmosis. (Water Factory 21's 5-mgd [18,900-m^3/d] spiral-wound reverse osmosis facility.)
3. The commercial development of the thin film composite membrane for seawater and brackish water applications.
4. The construction and operation of a large-scale seawater reverse osmosis plant (the 3.2-mgd [12,200-m^3/d] facility in Jeddah, Saudi Arabia).

5. The development and use of low-pressure membranes in several plants in Florida for softening and color removal using membranes.

6. The development of pretreatment and cleaning procedures.

Major Constraints

1. The sensitivity of polyamide and thin film composite membranes to chlorine. All of these membrane types, at present, require the removal of residual chlorine.

2. The need for high pressures.

APPLICATIONS

Brackish Water

Reverse osmosis using moderate pressures up to about 400 psi [27.2 atm] is well-suited to the desalination of brackish water (TDS up to about 10,000 mg/l). Energy requirements for operation do not significantly change with differences in TDS within the brackish water range. The most significant variable cost factors between plants are based on the amount of pretreatment and the percent of recovery possible. Feedwaters low in suspended solids, colloidal material, organics, and heavy metals reduce the amount of pretreatment required. Normal recoveries can range from about 50 to 90 percent depending on the solubility limits of the constituents in the raw water, the concentration reached in the brine, and the desired permeate water quality.

Brackish water plants are readily available from manufacturers, and considerable experience has been gained in operation over the past 5 years. Off-the-shelf models are available from some manufacturers, although the pretreatment of most of these units is custom-fabricated for the customer's water source using standardized parts. Large plants (in the 2- to 16-mgd [3,790- to 60,560-m^3/d]) range are in routine operation around the world, producing water for both domestic and industrial usage.

Seawater

Reverse osmosis using high pressures of 800-1,000 psi [54.4-68 atm] has been shown to be a viable application for desalination of waters with a TDS between 10,000 and 55,000 mg/l. The most significant variable cost factors between plants are based on the amount and type of pretreatment necessary and the percent of recovery possible. The raw water source must be developed carefully. There have been some reports of fouling problems due to suspended siliceous materials from direct sea intakes. For small supplies, the potential for a sea well should be explored. Care should also be exercised in the construction of the well to ensure that water quality is not affected by organics or other material that may have been buried by dredging, filling, etc., in the area.

Normal recoveries range about 20 to 35 percent for seawater. Recovery using other highly saline waters must be analyzed on a case-by-case basis depending on, among other factors, the solubility of the dissolved constituents. With low recoveries, considerable water must be pumped during production.

Manufacturers are now gaining experience in the operation of commercial seawater desalination plants. Since much of the construction is the same as for brackish water plants, the major questions entail materials of construction, pumps, and the long-term performance of the high-pressure pumps.

Softening

Membranes with lower rejection characteristics and higher flux have been used for softening purposes. Cellulose acetate membranes of this type are prepared by annealing the membrane at lower temperatures. Monovalent salts are poorly rejected, but calcium and magnesium rejections are such that a significant reduction in hardness occurs.

HEALTH ASPECTS

The membrane in the reverse osmosis process theoretically provides a barrier to the passage of bacteria and other microorganisms. However, two health aspects of the process must be considered.

The first is that the membranes are not perfect. Holes, cracks, broken fibers, and/or imperfections in fabrication can permit bacteria and other microorganisms to pass the barrier and enter the product water.

The second health consideration is that the configuration of membrane elements and permeators provides a suitable environment for the propagation of some microorganisms. The membranes and spacers provide a substrate to which the microorganisms can attach themselves, and the water passing by brings nutrients for them to sustain metabolic activity. Although these are unlikely to be pathogenic bacteria, at least of the enteric variety, these colonies develop causing problems relative to the biodegradation and/or clogging of the membranes, changing of flow paths, etc. For this reason, especially where significant quantities of microorganisms are known to be present in the feedwater, the feedwater should be chlorinated, subjected to UV light, or otherwise disinfected. For sources which might have heavy bacteriological contamination, disinfection of the feedwater is necessary to ensure that the performance of the membranes will not be affected by microbial action. The permeate should always be disinfected with chlorine or other acceptable materials or methods to ensure that the water meets bacterial standards.

Thus, although the reverse osmosis process can provide a barrier to bacterial contamination of product water, it is best considered as a secondary rather than a primary protection for public health purposes. Reverse osmosis, of course, would never be used solely for purposes of reducing or eliminating microorganisms, since far cheaper methods are available. Cellulose acetate membranes in current use do not reduce the chlorinated hydrocarbon concentrations (such as trihalomethanes [THM's]), but there are indications that there is some rejection by some of the polymers in thin film composite membranes. However, some definitive research needs to be carried out in this area.

ENVIRONMENTAL CONSTRAINTS

Brine

The brine stream has a concentrated salt solution and often some levels of phosphates (depending on dosage) which must be disposed of safely. In coastal locations the brine stream can usually be discharged into the sea or ocean without problems, although the phosphate levels from some plants may, in special instances, cause some nutrient imbalance if discharged into estuarian areas.

Inland, brine discharges can be an important problem, since the brine can adversely affect the quality of the existing ground water if the brine is allowed to enter the aquifers. One of the first items that should be considered in an inland desalting project using reverse osmosis (or any desalting process) is the disposal of the brine.

Air Pollution

Degassification and/or aeration could be used in certain pre- and post-treatment applications of reverse osmosis, possibly causing minor air pollution problems. Some caution and planning should be employed if hydrogen sulfide removal is required due to possible odor problems.

Noise

Reverse osmosis facilities utilize high-pressure pumps which can create an unpleasant noise level. This noise level should be considered both in locating and designing the facility. It is usually a good practice to separate the general office and other work areas from the high-pressure pumps due to their high noise level.

FUTURE PROSPECTS

New Technology Being Marketed

Thin Film Composite Membranes. Thin film composite membranes are being marketed by a number of manufacturers in the USA. These have been manufactured commercially only since about 1977 and, as yet, there is only selected usage of the membranes. However, they offer great potential, since they apparently outperform cellulose acetate membranes in almost all respects with regard to water flux, salt and organic solute rejection efficiency, temperature tolerance, and stability in acidic and caustic feeds (Mattson, 1979).

Long-term data on the performance of composite membranes are still being accumulated from both commercial and test installations. The most significant applications to date are in the 3.2-mgd [12,100-m^3/d] seawater RO facility in Jeddah, Saudi Arabia; in a 300,000-gpd [1,140-m^3/d] wastewater application for a dyeing and finishing plant in Pennsylvania; and in a 2.5-mgd [9,460-m^3/d] wastewater treatment plant for a power utility in New Mexico. In both the latter two cases the permeate is reused in the plants. The Jeddah facility is described more extensively as a case study in the section on "Process Experience."

Dry Membranes. At least one manufacturer claims to have developed a spiral cellulose acetate membrane element which can be shipped and stored dry before being placed in service. This may have a number of advantages on certain projects, including a reduced shipping weight and a diminished potential for damage during shipping or storage due to freezing or overheating. The disadvantage is that the elements are not tested before shipping. Thin film membrane elements have been manufactured which are wet-dry stable.

Dynamic Membranes. Dynamic membrane modules using zirconium oxide-polyacrylate are currently available only on a limited basis in the USA. At least one experimental commercial unit has been built using these membranes to treat textile wastewater (Brandon, 1980), but their use for water supply purposes is essentially nonexistent.

Anticipated Technological Developments

Low-Pressure Membranes. Effective elements for low pressure are currently being developed. Advantages to operation at lower pressures include: (1) savings in energy cost; (2) use of lower cost materials (to withstand lower pressures) in the plant; and (3) a possible increase in effective membrane life and a reduction in flux decline (Mattson, 1979). Operating pressures of 200 to 250 psi [13.6 to 17 atm] are being investigated for brackish water applications. Some of these membranes are shown in Appendix B.

New Seawater Membranes. New seawater membranes are being developed by manufacturers and should be commercially introduced in the near future. Dow Chemical USA is currently field testing a homogeneous hollow fine fiber membrane made from cellulose triacetate. This would complement their line of brackish water hollow fine fiber permeators now in use.

Hollow Fine Fiber Composite Membranes. Efforts have been made in the past by various groups, including Albany International, to develop a commercially viable composite membrane using the hollow fine fiber configuration (Davis et al., 1977). Thus far, such membranes have not been offered commercially.

Energy Recovery

It is anticipated that energy recovery systems will be commercially available in 1980.

PROCESS EXPERIENCE

Previous A.I.D. Studies

No detailed case histories on reverse osmosis facilities were outlined in earlier editions.

Recent Process Experience

Four units have been selected as examples. Two are large plants (3 mgd [11,400 m^3/d] or larger) and two are small (10 kgal/day [38 m^3/d] or smaller). A large and small plant were selected for both seawater and brackish water applications. Three of them are discussed below in this chapter and the fourth, a brackish water plant in a Mexican village, in Chapter 9.

Cape Coral, Florida (5 mgd [18,900 m^3/d]--Brackish Water RO). The City of Cape Coral is located on a saltwater peninsula on the west coast of Florida. Rapid population growth in the 1970's produced water usage beyond the capacity of the existing high-quality ground water on the peninsula. After considering several water supply alternatives which included well field and surface-water development with long distance transmission to the site, an option using reverse osmosis was selected. This was based on economics and the desire to have the source located within the City's political jurisdiction. A locally available brackish ground-water supply was developed and reverse osmosis was employed to desalinate the water.

A 3-mgd [11,400 m^3/d] unit was manufactured by Permutit using Dow hollow fine fiber permeators and was put into operation in 1977. During 1980, an expansion was completed by Water Services of America using DuPont's hollow fine fiber permeators to increase its capacity to 5 mgd [18,900 m^3/d]. The facility is operated in conjunction with an existing softening plant which utilizes the limited low-TDS ground water as a supply. Information on this facility was derived from a recent report by Shannon (1979). In designing the facility, the ground-water quantity and quality were first determined by a test well program of drilling and pumping.

The result was the construction of a well field of six wells about 700 feet [213 m] deep. The first approximately 350 feet [107 m] in each well was cased with 12-inch-diameter [30.5-cm] plastic pipe and the remainder drilled as an open hole through sandy limestone.

Based on the water quality tests of the wells, the following raw water quality data were used for design (Shannon, 1979).

Temperature	85°F/29.3°C
Total Dissolved Solids	2,050 mg/l
Total Hardness (as $CaCO_3$)	604 mg/l
Sodium	365 mg/l
Chloride	743 mg/l
Sulfate	281 mg/l
Strontium	18.1 mg/l
Total Sulfide	3.25 mg/l

The total capital cost for the facility, including the well field, buildings, and entire equipment, was $4.5 million.

i) Process Description--Figure 5-14 shows the process flow diagram for the Cape Coral facility. The well field and raw water delivery system were designed so as to prevent or minimize the oxidation of constituents in the raw water to avoid formation of particulate matter such as elemental sulfur, etc. Sulfuric acid and polyphosphate are added to adjust the pH and provide a threshold inhibitor to the feed. Cartridge filters are located just upstream of the high-pressure pumps. The 5-mgd [18,900-m^3/d] plant uses 220 permeators, mounted and piped into ten individual banks with a capacity of 0.5 mgd [1,890 m^3/d] per bank, are used for desalination. The 22 permeators in each bank are in a cascade (two-stage) configuration, with the reject from the first 16 permeators being fed to the remaining six permeators. The design minimum recovery rate is 65 percent.

Following desalination in the membranes, the product water is stabilized by degassification and the addition of sodium hydroxide before being disinfected and transferred to a 5-MG [18,900-m^3] storage tank. A percentage of raw water is bypassed around the membranes to be mixed with the product water to produce a final blend that is used in the potable water system. The plant is producing a water with a TDS of about 65 to 70 mg/l, which is blended with up to 30 percent of the raw water, resulting in a final product with a TDS of about 375 mg/l.

The reverse osmosis facility, with the exception of the degassifier, is enclosed in a 12,000-ft^2 [1,115-m^2] building. The building also includes a large laboratory, shop, control room, offices, and storage. Acoustical concrete block was used extensively in the portion of the building containing the high-pressure pumps in order to reduce the high noise level associated with the pumps. Figure 5-15 depicts the use of an acoustical wall to separate the high-pressure pumps from the permeators and much of the rest of the facility.

ii) Operational Information--Figure 5-16 shows the monthly production of Cape Coral's reverse osmosis and softening plant. Table 5-4 summarizes the operating conditions for the plant for the period January 1978 to May 1979. Actual cost data published for this period are shown in Table 5-5.

The plant has performed well and after 2 years has a salt passage of less than 7 percent (the manufacturer's performance guarantee was for less than 7 percent salt passage during the first 3 years of operation). The quality of the raw water from the well fields has remained high, resulting in lower chemical usage of acid and polyphosphate than anticipated, i.e., 125 mg/l vs. 160 mg/l for acid and 3.4 mg/l vs. 10 mg/l for polyphosphate. The low fouling tendency of the water has been reflected in the need to change the cartridge filters only once in a 2-year period (Shannon, 1979).

No major problems have been experienced with the membranes, and it is anticipated that their average useful life will exceed 5 calendar years rather than the 3 years guaranteed by the manufacturer.

The major problem with the plant has been under-utilization. The average utilization during the period has been about 45 percent. This has the greatest effect on labor costs, which would not significantly change if the plant were operated at full capacity. Utilization will increase as population increases in the area.

Jeddah, Saudi Arabia (3.2 mgd [12,100 m^3/d]--Seawater RO). The area around Jeddah (on the Red Sea) is extremely arid and is experiencing a rapidly increasing demand for water. Several dual-purpose (electricity/distillation) plants have been and will be built in the area. By 1982, the installed capacity for seawater desalination facilities in the Jeddah area should be over 80 mgd [303,000 m^3/d].

Over 90 percent of this capacity will be in distillation plants, which require long planning and construction periods. In an effort to handle a shortage of water expected between the completion of distillation plants in the late 1970's, the Kingdom of Saudi Arabia decided to purchase a seawater reverse osmosis facility to be built and installed on an accelerated schedule.

The contract was awarded to the Fluid Systems Division of UOP in July 1977, with partial operation anticipated in 6 months and full operation in 1 year. Due to various problems regarding site preparation and transportation, the plant did not go into full operational status until January 1979, 1½ years later.

Table 5-4
SUMMARY OF OPERATING CONDITIONS FOR THE CAPE CORAL REVERSE OSMOSIS PLANT[a] (JANUARY 1978 TO MAY 1979)

Total water produced-product plus blend	1,117 MG [4.23 Mm^3]
Average daily water production	1.83 mgd [6,930 m^3/d]
Average raw water blend (24 percent)	0.4 mgd [1,510 m^3/d]
Potential water production at 4.5 mgd [17,000 m^3/d]	2,432 MG [9,21 Mm^3]
Overall plant utilization	46 percent
Feedwater pH range	5.8 to 6.0
Sulfuric acid dosage	125 mg/l
SHMP dosage	3.4 mg/l
Caustic dosage	21 mg/l
Chlorine dosage	4.4 mg/l
Free chlorine residual	1.5 to 2.0 mg/l
Feedwater pressure range	295 to 315 psi [20 to 21.4 atm]
Membrane Cleaning	None
Cartridge filter replacement	Once

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[a] Table adapted from Shannon, 1979.

Table 5-5
ACTUAL COST DATA FOR THE CAPE CORAL REVERSE OSMOSIS PLANTa
(APRIL 1977 TO MAY 1979)

Total water produced	1,368 MG [5.18 Mm^3]
Total plant capability for same period	
Product alone @ 3.0 mgd [11,400 m^3/d]	2,373 MG [8.98 Mm^3]
Product plus blend @ 4.5 mgd [17,000 m^3/d]	3,559 MG [13.50 Mm^3]

Item	¢/kgal	¢/m^3
Chemicals (sulfuric acid, SHMP, chlorine and caustic)	5.6	1.5
Electrical (entire plant plus well field)	29.0	7.7
Labor (salary plus fringes)	11.8	3.1
Membrane replacement[b]	12.6	3.3
Equipment maintenance (3.0-mgd [11,400-m^3/d] flow)[c]	4.1	1.1
Total operating cost	63.1	16.7
Amortized capital[d] ($4,495,355 at 7% for 30 years)	55.2	14.6
Total Cost	118.3	31.3

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[a] Table adapted from Shannon, 1979.

[b] Assumes that all membranes will be replaced after 4 years with a flow of 3 mgd [11,400 m^3/d] at a total cost of $550,000.

[c] Maintenance estimated at 1 percent of capital cost/year.

[d] All components of plant, excluding permeators and high-pressure pumps and including cartridge filtration, chemical feed systems, degassifiers, transfer and high-service pumps, ground storage, chlorination system, and building are sized to handle a product water flow of 5 mgd [18,900 m^3].

The facility is located on the shore of the Red Sea and obtains its raw water supply from a seawater intake located offshore. Brine and filter backwash water is discharged back into the sea adjacent to the site. A photograph of a model of the facility is shown in Figure 5-17.

The capital cost for the unit was approximately $30 million. A cost breakdown is shown in Table 5-6. The large capital cost differential between the Jeddah and the Cape Coral plants ($30 million versus $4.5 million) is due to several factors. Among these are:

1. Self-containment--The Jeddah plant had to generate its own power, provide living quarters for staff, store fuel, etc. Cape Coral has none of these requirements.
2. Feedwater--The Jeddah plant treats seawater versus brackish water for the Cape Coral plant.
3. Time of Construction--The Jeddah plant started construction about 3 years after the Cape Coral plant.
4. Time of Completion--The Jeddah project was performed on a rush basis versus standard construction time schedule for Cape Coral.
5. Location--The Jeddah plant was built at a site almost 12,000 miles [19,000 km] from the contractor's base of operation. No such problem existed at Cape Coral.

i) Process Description--The process flow diagram for the plant and the general layout of the facility are presented in Figures 5-18 and 5-19, respectively. The information on the process description and operational information comes from two 1979 papers, one by Al-Gholaikah et al., and one by Hickman et al.

The seawater, which has a maximum temperature, turbidity, and TDS of about 90°F [32.1°C], 0.5 to 0.6 NTU's, and 41,200 mg/l, respectively, is screened as it is brought into a seawater wet well and then pumped to 3 dual-media filters. Provisions have been made to allow injection and in-line mixing of a coagulant such as alum, ferric chloride, or a synthetic polymer.

Initial operation has shown that a coagulant is not required. The pH of the filtered water is adjusted to 6 by the addition of acid and the water is then pumped from the filter clear well into the reverse osmosis building, where polyphosphate is added. The water passes through the 25-micron cartridge filters to the suction side of the high-pressure pumps.

The first stage of the reverse osmosis section consists of nine units, each having 56 fiberglass pressure tubes 21 feet [6.4 m] long and 6 inches [15.2 cm] in diameter, as shown in Figure 5-20. Each of the tubes·containes six UOP Model 1501 polyamide thin film

Table 5-6
CAPITAL COSTS FOR THE JEDDAH SEAWATER REVERSE OSMOSIS FACILITY[a]

Item	Cost in in U.S. Dollars[b]
Filters	$ 1,412,000
Electrical Equipment and Diesel Generators	1,037,000
Piping, Valves, and Materials	4,830,000
R.O. Pressure Tube Stacks and Membrane	8,599,000
Pumps, Diesels, and Motors	5,202,000
Pretreatment Skids and Storage Tanks	302,000
Control Panels and Instruments	206,000
Miscellaneous Hardware, Structures, Bolts, Gaskets, Paint, Grout, etc.	1,018,000
	22,606,000
Civil Work	5,500,000
Electrical and Mechanical Installation	2,000,000
Total	$30,106,000

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[a] Source: Hickman et al., 1979.
[b] At the conversion rate of 3.5 Saudi Riyals = $1 U.S.

composite spiral-wound elements. The feed to each of these units is pressurized by two multistage vertical turbine pumps operated in parallel and directly driven by diesel engines. The pressure of the pumps is adjusted by regulating the speed of the engines; each pump has the capability of delivering 500 gpm @ 1,000 psi [31.5 l/s @ 68 atm]. The design recovery rate for the first stage is about 30.5 percent.

The second product stage reduces the TDS of the permeate from the first stage if the TDS of the water begins to rise significantly above the desired level of about 1,000 mg/l. There are three second-stage reverse osmosis units, which are also pressurized by a vertical turbine pump (500 gpm @ 600 psi) [31.5 l/s @ 40.8 atm] directly driven by diesel engines. The same number and type of pressure tubes and membrane elements are used in the second stage except that they are manifolded in triple reject staging arrangement (32-10-8) to allow a recovery of about 90 percent. A maximum of about one-half of the first-stage permeate can be treated in the second stage, but this is sufficient, when necessary, to produce a blended product capable of meeting the standard of 1,000 mg/l for dissolved solids. The reject from the second stage is returned to the filter clear well.

Although lower-pressure (and lower-cost) brackish water membranes could have been utilized in the second-stage units, duplicates of the first-stage membranes and tubes were used to allow interchange of parts and reduction of the spare parts inventory.

The blended output from the first and second stages is chlorinated for disinfection and stabilized by the addition of lime to bring the pH up to between 7 and 8.5 in the 18,500-gallon [70-m^3] permeate tank. Water from the permeate tank is then transferred to the distribution system for the City of Jeddah.

Other facilities constructed as part of the project include:

Electric power plant--three 440-kW generators driven by diesel engines.

Storage tank for sulfuric acid--50,000 gal [189 m^3].

Storage tank for diesel fuel--100,000 gal [379 m^3].

Control room and office building--1,100 ft^2 [102 m^3].

Warehouse and chemical storage building--6,000 ft^2 [557 m^2].

Dormitory for 15 operations personnel--3,050 ft^2 [283 m^2].

ii) <u>Operational Information</u>--From 1979 through 1980, the facility was being operated under a 2-year operation and maintenance contract by the original manufacturer and contractor, UOP. The value of this contract is approximately $4.7 million, exclusive of diesel fuel (Hickman et al., 1979).

Table 5-7 notes the average consumption of expendables at the facility during the 4-month period of February 1, 1979 to June 1, 1979.

Cowpet Bay, Virgin Islands (0.01 mgd [38 m^3/d]--Seawater RO). The island of St. Thomas in the U.S. Virgin Islands is located in a water-short area of the Caribbean. The potable water distribution system is limited in extent and much of the island is not served by the public distribution system.

Adequate ground water is found only in limited locations on the island so that people and businesses away from the distribution system must utilize cisterns for their water supply system. The cisterns are supplied by rainwater collected on roofs and water delivered by trucks.

The Cowpet Bay Condominiums are a complex of 107 apartment units located on a bay on the east end of the island. They have maximized their available water resources by installing special minimum water use plumbing and by treating and reusing their wastewater for sanitary flushing and landscape irrigation. Still, during the year the demand for potable water exceeds the amount of rainwater that the condominiums can collect and store.

In the past, this shortage was made up by purchasing water hauled 3 to 8 miles [4.8 to 12.8 km] in trucks at a cost of, in 1980, up to $23.00/kgal [$6.08/m^3]. To alleviate this problem, a 0.01-mgd [38-m^3/d] seawater reverse osmosis unit was purchased by the condominium, and it began operation in October of 1978. The unit was manufactured and installed by Polymetrics and employs three DuPont B-10 permeators for desalination. The total capital cost for the unit, including shipping, installation, and the construction of a seawell for a raw water source, was approximately $75,000 (Holden, 1980).

i) Process Description--The process flow sheet is shown in Figure 5-21. The raw water supply is derived from a 75-foot [23-m^3] deep seawater well which was constructed on a sand beach about 600 feet [183 m] from the unit on what was probably an old lagoon. It has performed satisfactorily thus far and delivers feedwater which is very low in suspended material and rather constant in water quality.

The raw water supply is pumped from the well into the unit, where it passes through a pressure sand filter, has sodium hexametaphosphate added to the stream, passes through cartridge filters, and then is pressurized by a positive displacement pump. The quality of the raw water has been sufficiently high that the sand filter has been bypassed during the first 1½ years of operation. Three DuPont B-10 permeators using the hollow fine fiber configuration are utilized for desalination. The plant is designed for operation at a recovery factor of about 20 percent. The unit's pumps, filters, etc., have been sized to permit expansion to 0.02 mgd [76 m^3/d] by the addition of three more permeators. The brine from the operation is returned to the sea.

Table 5-7
FUEL AND CHEMICAL CONSUMPTION AT THE JEDDAH SEAWATER REVERSE OSMOSIS FACILITY[a,b]

Item	Consumption (lb/kgal)[c]	Consumption [g/m^3][c]
Fuel (diesel)	24.61	2,945
Sulfuric Acid	2.85	341
Sodium Hexametaphosphate	0.11	13.2
Copper Sulfate	0.04	4.8
Lime	0.34	40.7

---

[a] Source: Hickman et al., 1979.

[b] During the period February 1, 1979 to June 1, 1979.

[c] Based on finished product water production.

ii) Operational Information--This plant is operated by one operator, who has several other responsibilities on the property, including general plumbing and electrical maintenance; operation of the wastewater plant; and operation of the water distribution system. The facility has thus far required minimal attention. It is, however, located where it can easily be checked several times during the day. Initial problems with the facility arose with the construction of the seawater well to service the plant. The initial attempt at locating the well was unsuccessful, but the final location on the beach proved to be viable. Production during the first 5 months averaged about 10 kgal/day [38 m^3/d].

In April of 1979 the unit was shut down because heavy rains during the spring and summer provided adequate freshwater supplies from the roof catchments alone. In December 1979 the plant was restarted after a shutdown of 7 months. The plant performed adequately, except that the flow meters were then out of calibration.

Power usage during the first 4 months of operation was approximately 25.6 kWh/kgal [6.8 kWh/m^3] excluding power for the well pump, which has been estimated at about 7-10 kWh/kgal [1.9-2.6 kWh/kgal] (Buros, 1979).

APPLICABILITY FOR DEVELOPING COUNTRIES

The principle of operation of reverse osmosis plants is relatively straightforward. However, their ultimate success depends upon proper development of the raw water source, selection of suitable equipment and pretreatment, and careful operation. If even one of these is lacking, severe problems can occur.

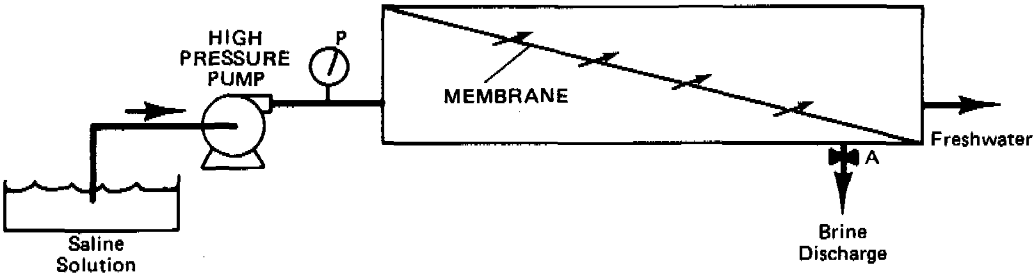
The operation of the plants requires personnel able to operate and maintain motors and pumps, mix chemicals, read meters and graphs, and make relatively simple process decisions based on these readings.

Reverse osmosis equipment is not always forgiving of errors or neglect associated with problems in the raw water source, addition of chemicals, adjustment of flows, etc. Failure to pay attention to the quality of the incoming raw water or to add chemicals or simply being overanxious in adjusting the equipment can create situations in which the membranes can rapidly foul and fail. This can occur in a matter of months, days, or hours depending on the transgression and it can require the replacement of all or a portion of the membranes.

Although the replacement of the membranes can be expensive it involves a portion of the plant which is anticipated to have a life of only 3-5 years under good conditions. Thus a disaster can possibly be financially absorbed. With distillation units on the other hand, if a disaster occurs it often means that a proportionally much larger cost must be incurred for repairs, and this is on equipment that probably should last 15 to 20 years.

Reverse osmosis equipment may not require much attention, but it cannot be expected to operate continuously unattended. Timely, knowledgeable adjustments have to be made, especially with plants with high conversion efficiency. The high pressures associated with the process provide the potential for leaks in the piping and fatigue or wear due to vibration inherent in the high-speed pumps.

As will be further discussed in Chapter 9, reverse osmosis plants cannot be expected to operate in isolated areas without some technical supervision. Although considerable experience has been gained in the field, even installations in developed countries have, on occasion, experienced severe problems.



A membrane assembly is generally symbolized as a rectangular box with a diagonal line across it representing the membrane.

FIGURE 5-1. Elements of a reverse osmosis system.

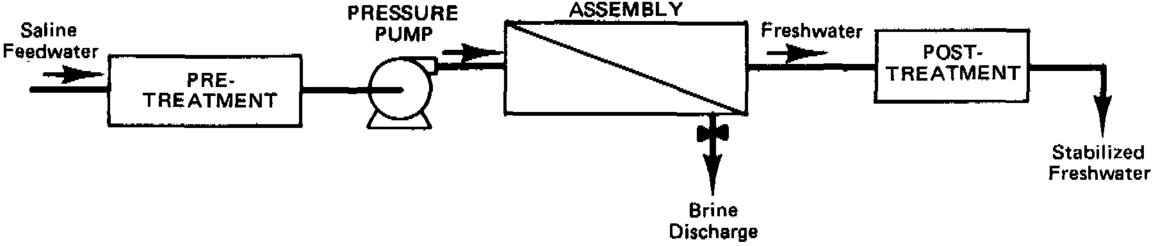


FIGURE 5-2. Flow diagram of a reverse osmosis system.

GKSS (Germany) PLATE AND FRAME MEMBRANE

Adapted from Boddeker et al., 1976.

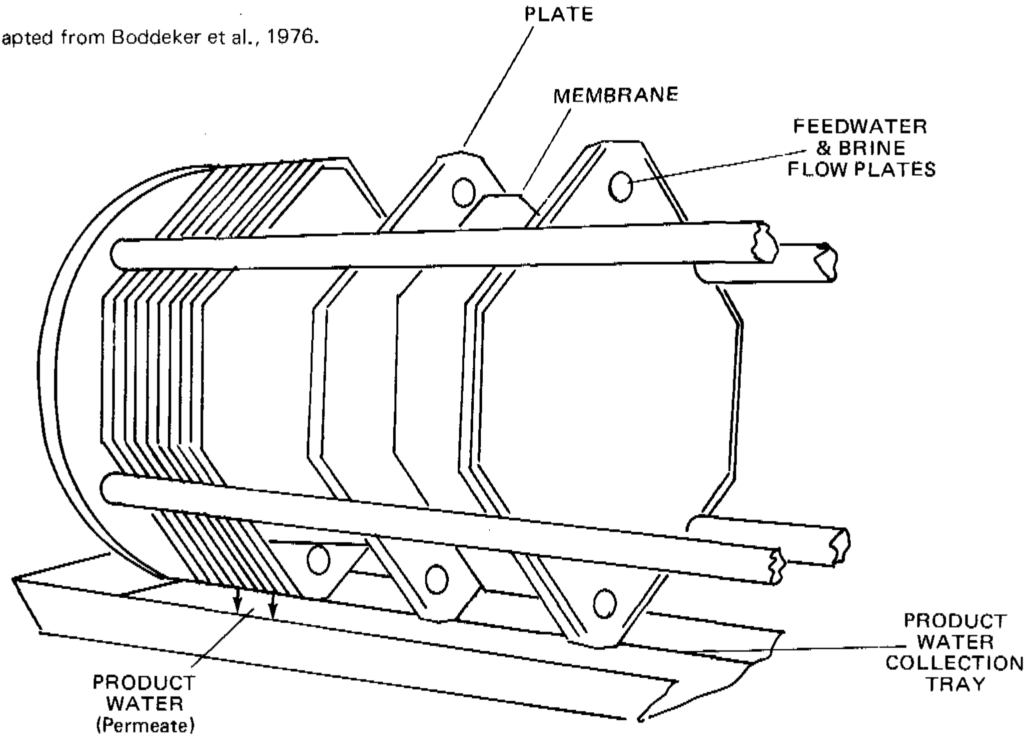


FIGURE 5-3. Construction of a plate and frame membrane.

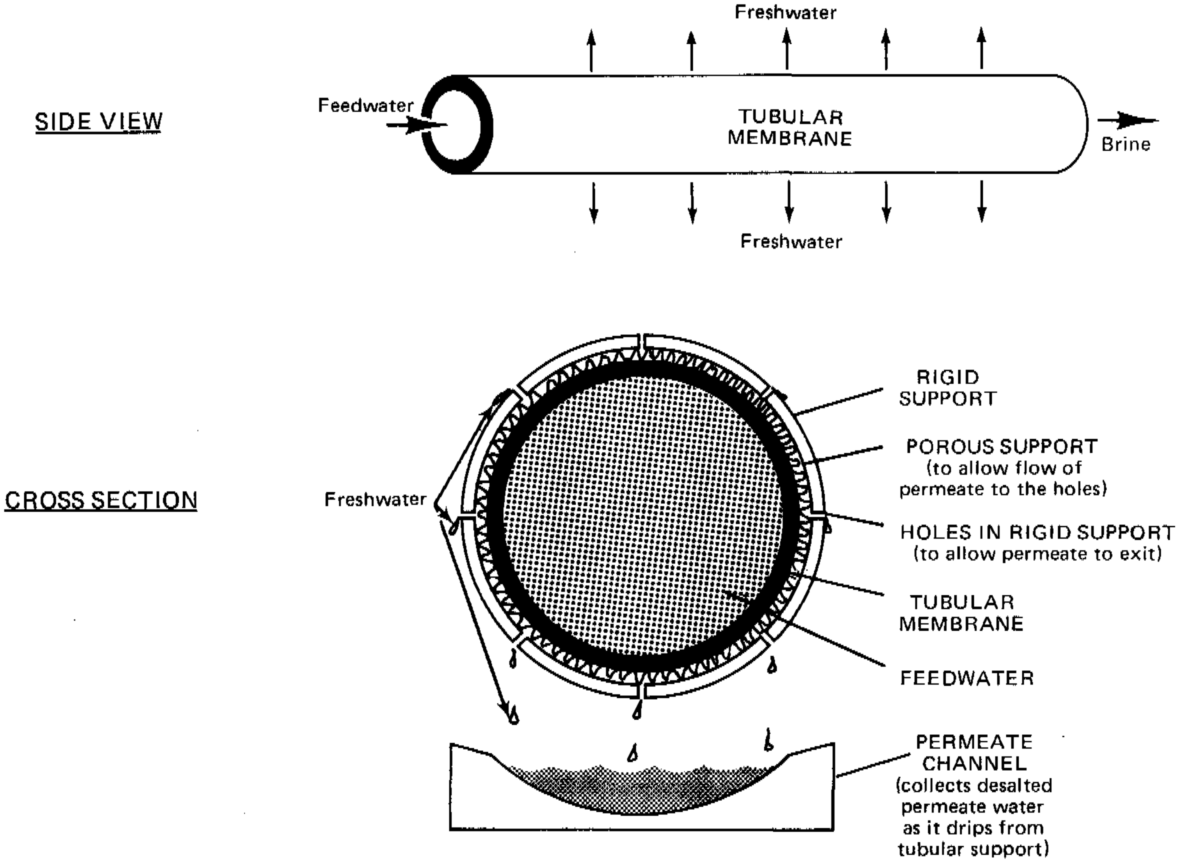
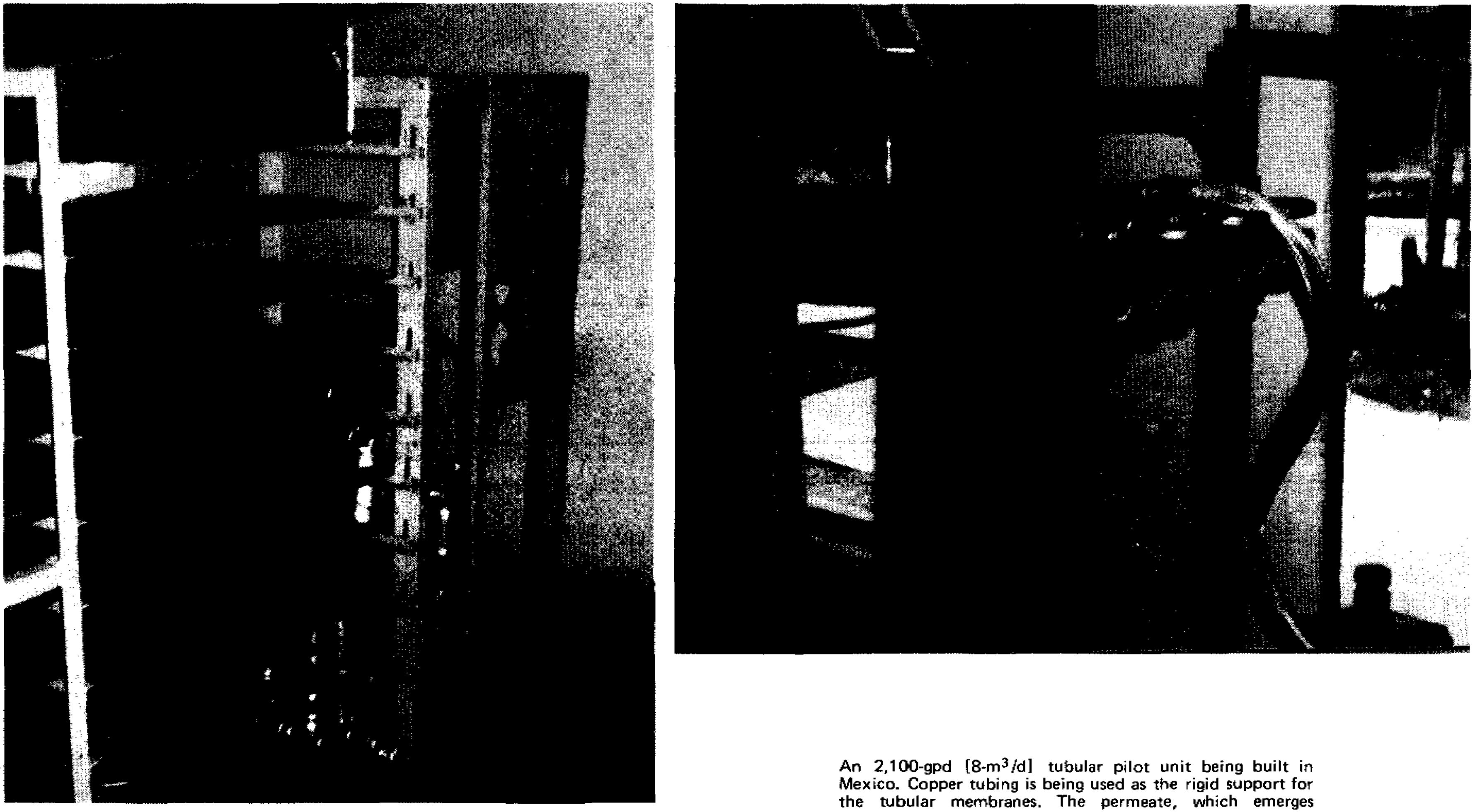


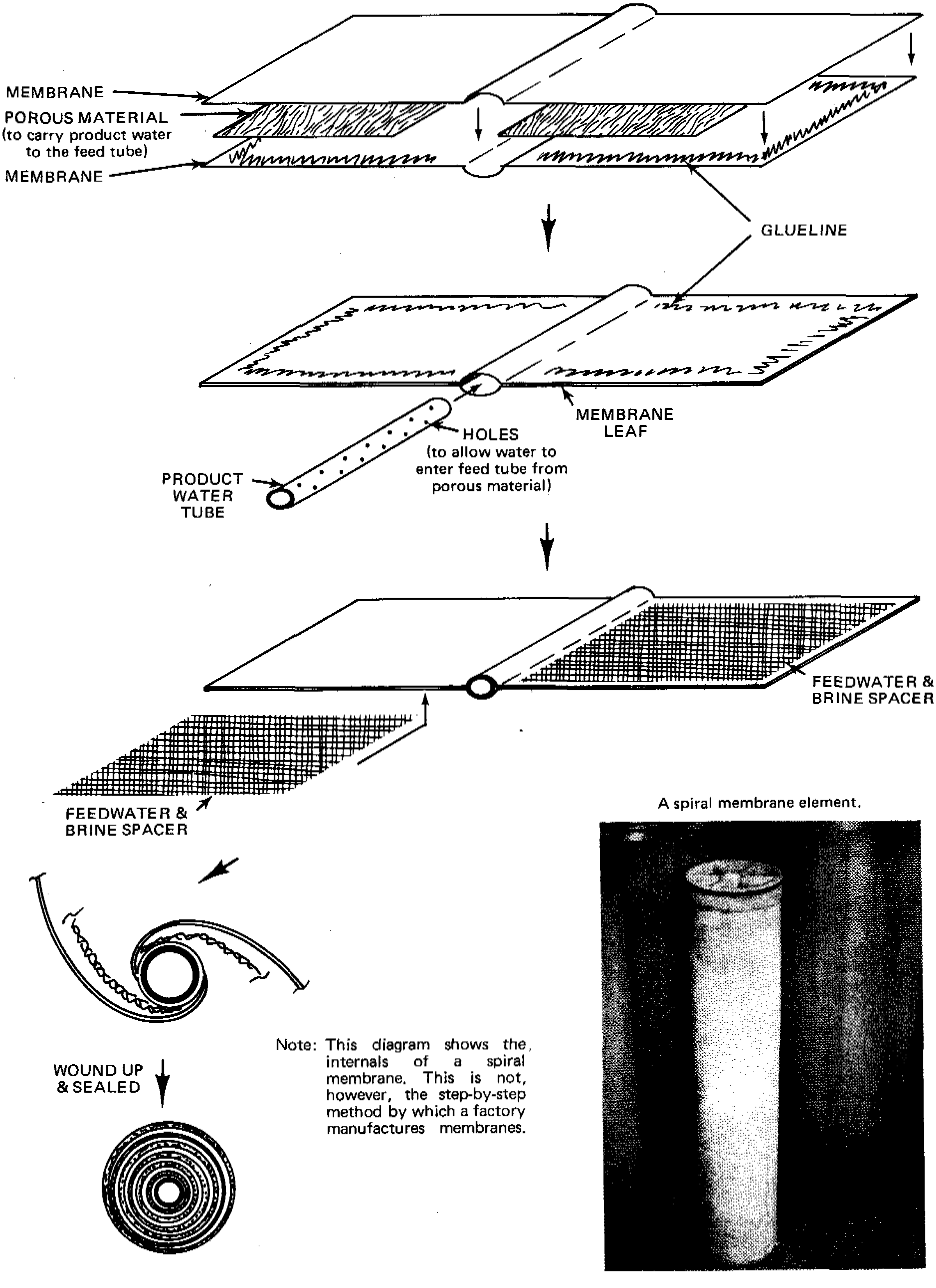
FIGURE 5-4. Construction of a tubular membrane.



Photos Courtesy of O. K. Buros.

An 2,100-gpd [8-m^3/d] tubular pilot unit being built in Mexico. Copper tubing is being used as the rigid support for the tubular membranes. The permeate, which emerges through the holes in the copper tubing, is collected in the corrugated metal sheets which act as a collector.

FIGURE 5-5. Tubular membrane pilot plant in Mexico.

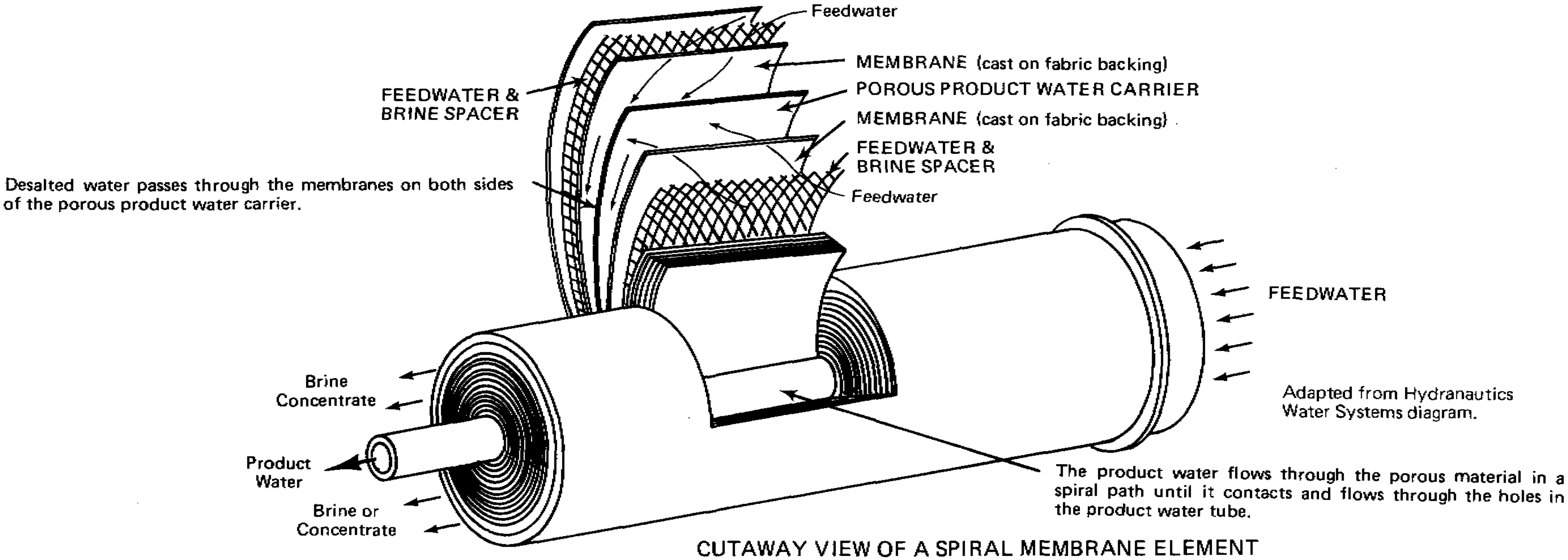


Note: This diagram shows the internals of a spiral membrane. This is not, however, the step-by-step method by which a factory manufactures membranes.

A spiral membrane element.

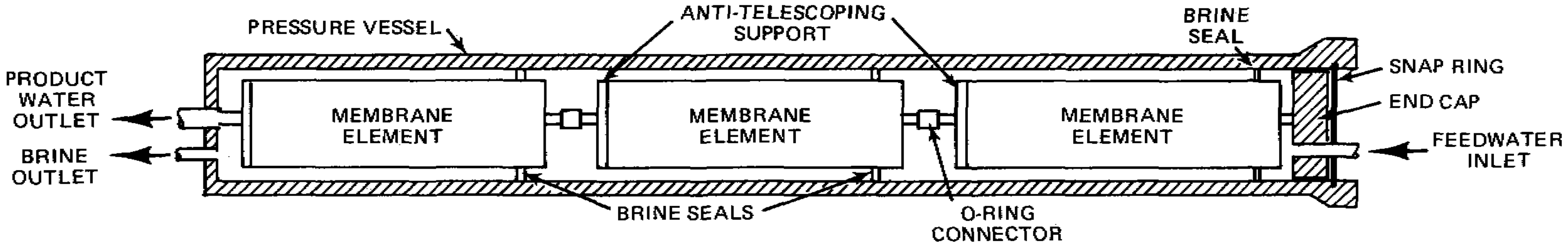
Photo Courtesy of O. K. Buros.

FIGURE 5-6. Internal construction of a spiral membrane.



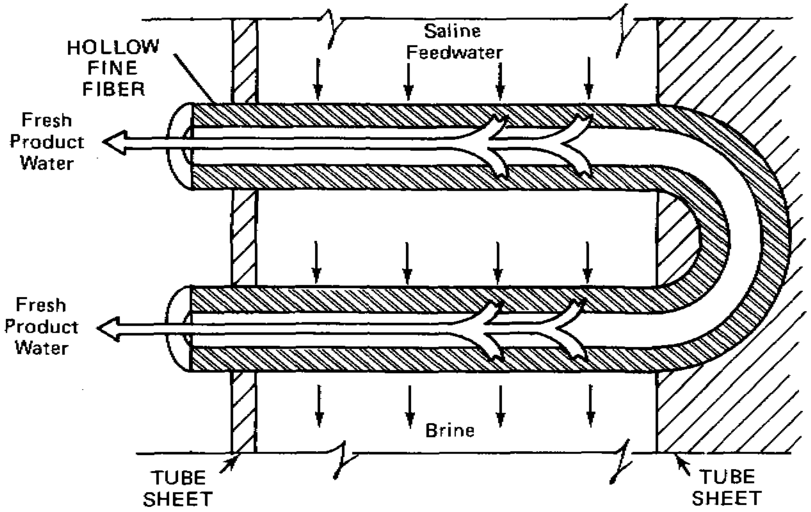
Adapted from Hydranautics Water Systems diagram.

CUTAWAY VIEW OF A SPIRAL MEMBRANE ELEMENT

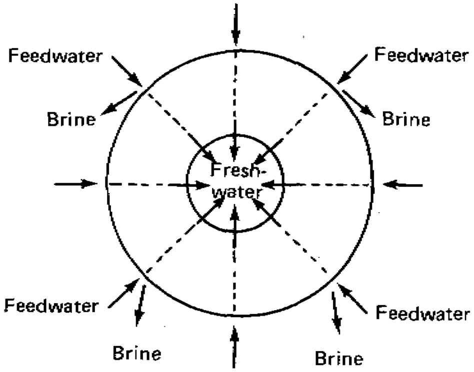


CROSS SECTION OF PRESSURE VESSEL WITH 3-MEMBRANE ELEMENT

FIGURE 5-7. Spiral membrane-cut-away view with elements in a pressure vessel.



Drawing adapted from Dow Chemical USA sales literature.



SECTION A-A
Cross Section of a
Hollow Fine Fiber-Enlarged

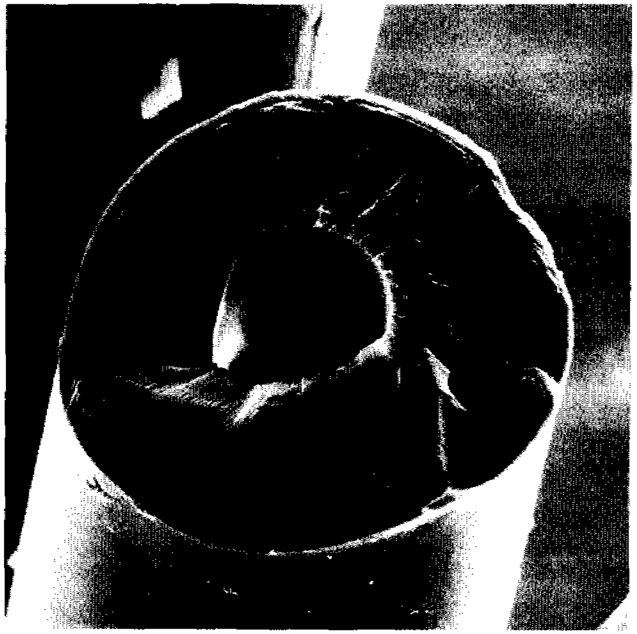


Photo Courtesy of Dow Chemical USA.

Photomicrograph of the cross section of an asymmetric cellulose triacetate membrane.

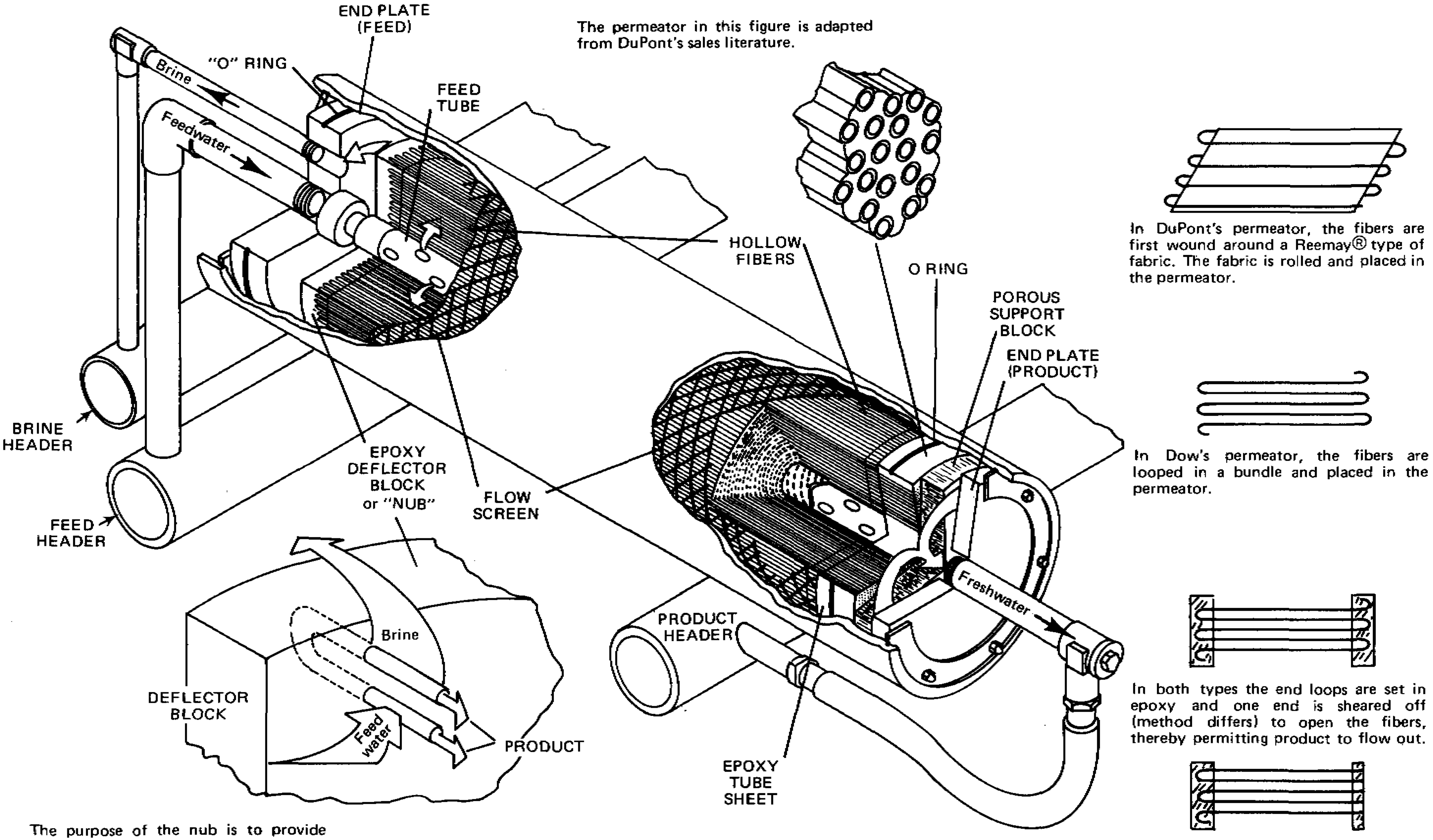
Diameter of Hollow Fine Fibers

	Fiber Diameter			
	Inside		Outside	
	(in)	(μm*)	(in)	(μm*)
DuPont				
Brackish	0.0016	40	0.0034	85
Seawater	0.0016	40	0.0034	100
Dow				
Brackish	0.0036	90	0.01	250

*μm = microns

Most hollow fine fiber membranes are asymmetric in structures, although development is taking place on thin fine composite hollow fine fiber membranes (Albany International), and Dow has introduced an experimental dense hollow fine fiber membrane for seawater applications.

FIGURE 5-8. Hollow fine fiber membrane.



The permeator in this figure is adapted from DuPont's sales literature.

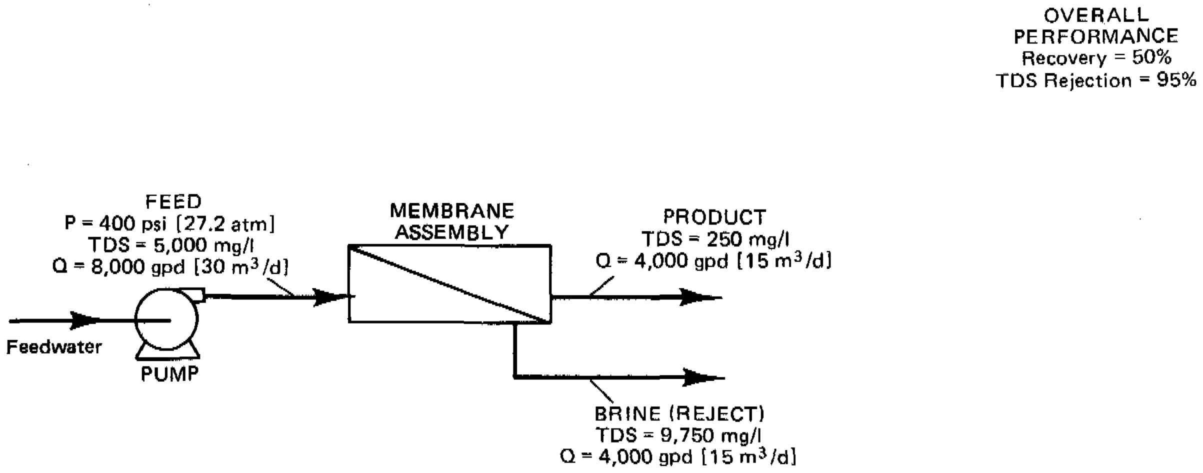
In DuPont's permeator, the fibers are first wound around a Reemay® type of fabric. The fabric is rolled and placed in the permeator.

In Dow's permeator, the fibers are looped in a bundle and placed in the permeator.

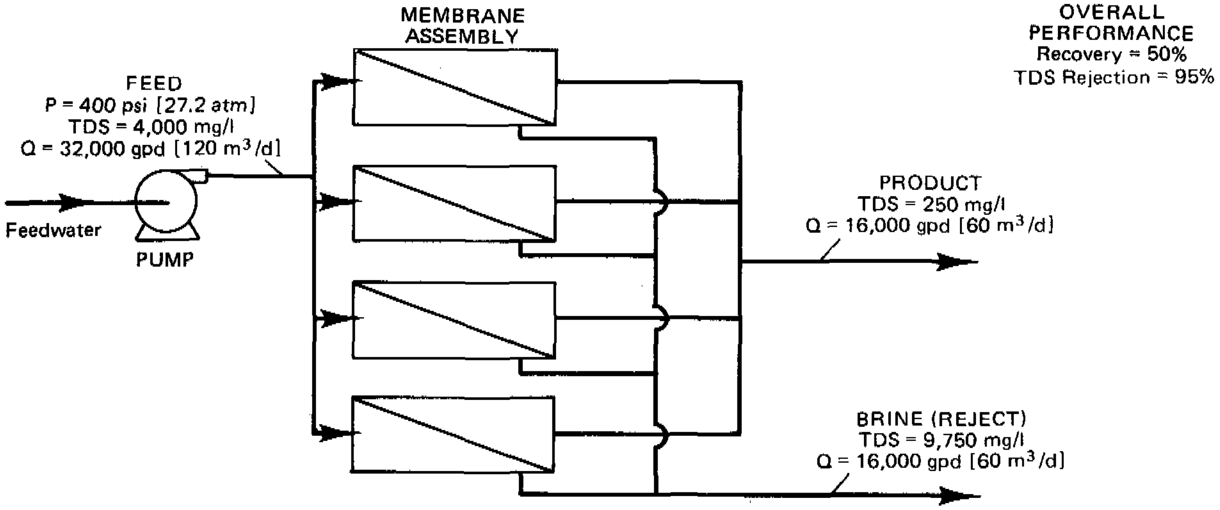
In both types the end loops are set in epoxy and one end is sheared off (method differs) to open the fibers, thereby permitting product to flow out.

The purpose of the nub is to provide mechanical stability to the bundle and to provide an annulus for the brine flow.

FIGURE 5-9. Permeator assembly for hollow fine fiber membranes.

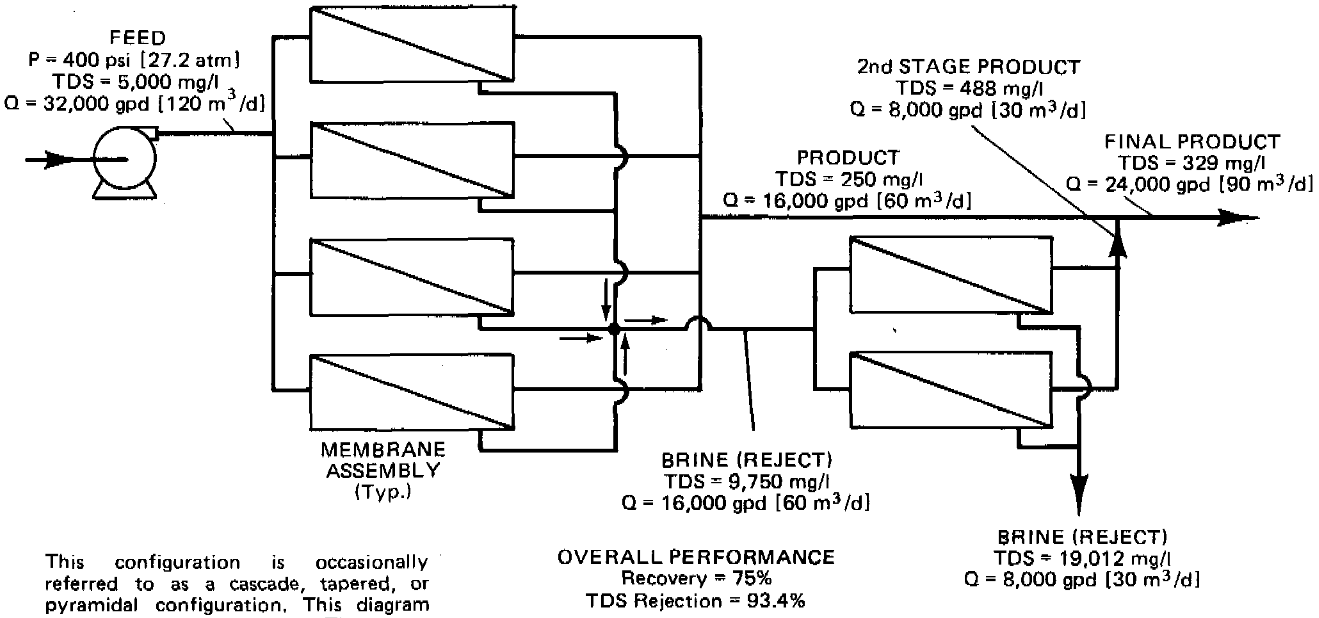


SINGLE STAGE WITH A SINGLE MEMBRANE ASSEMBLY



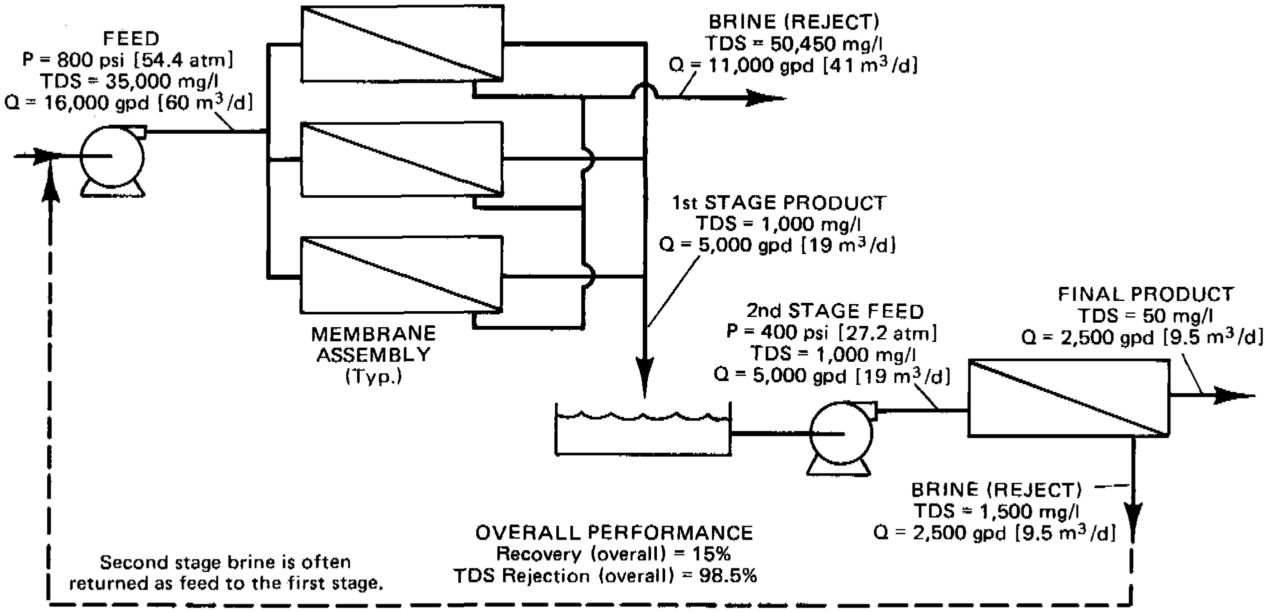
SINGLE STAGE WITH PARALLEL MEMBRANE ASSEMBLIES

FIGURE 5-10. Single-stage RO plant configurations.



This configuration is occasionally referred to as a cascade, tapered, or pyramidal configuration. This diagram illustrates a two-state unit. Three-stage units (with recoveries of 85 to 90%) are also used.

MULTISTAGE WITH REJECT STAGING



Second stage brine is often returned as feed to the first stage.

MULTISTAGE WITH PRODUCT STAGING

FIGURE 5-11. Multistage RO plant configurations.

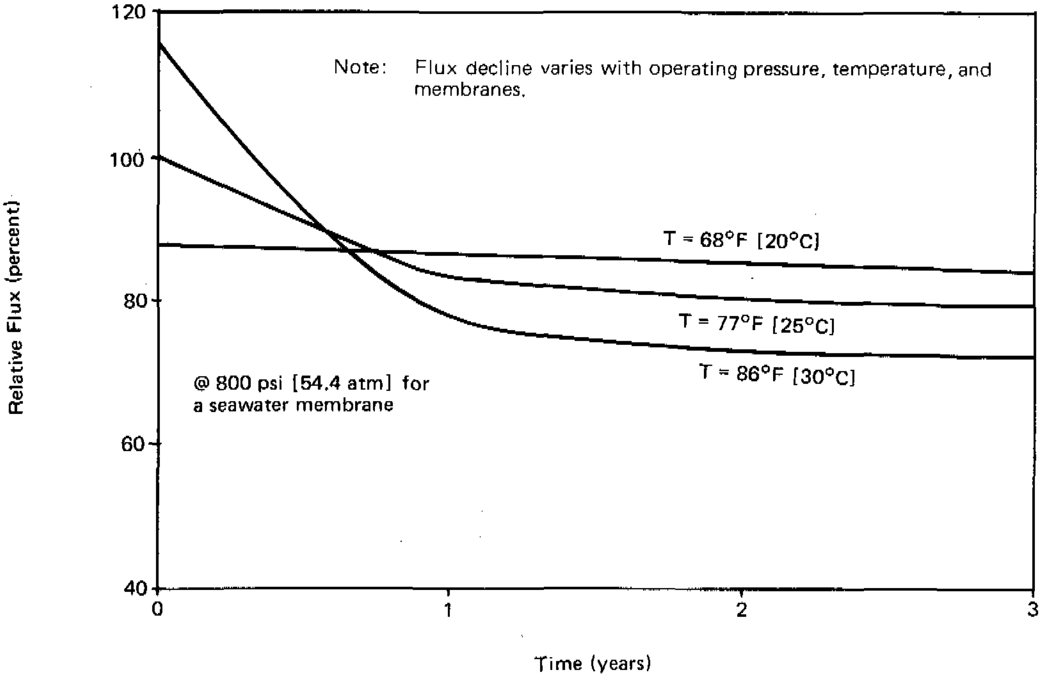
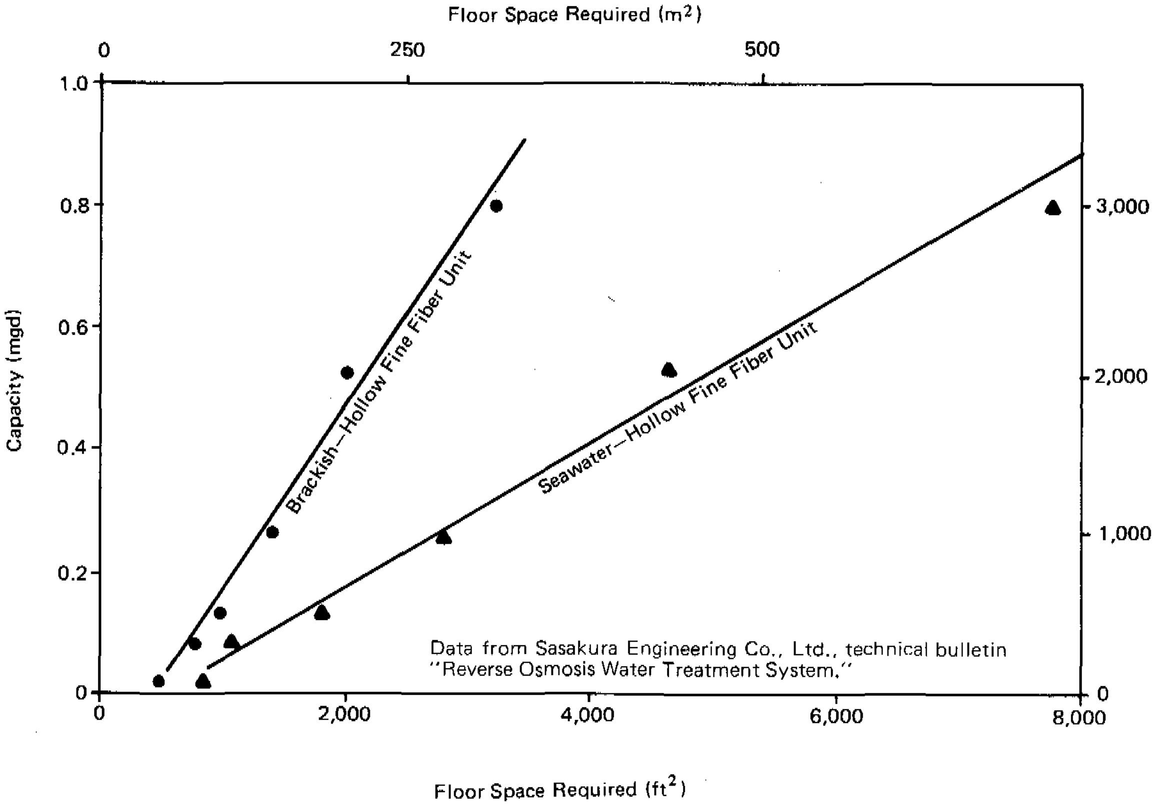


FIGURE 5-12. Flux decline in a hollow fine fiber reverse osmosis membrane.



Data from Sasakura Engineering Co., Ltd., technical bulletin "Reverse Osmosis Water Treatment System."

FIGURE 5-13. Approximate space required for skid-mounted reverse osmosis units.

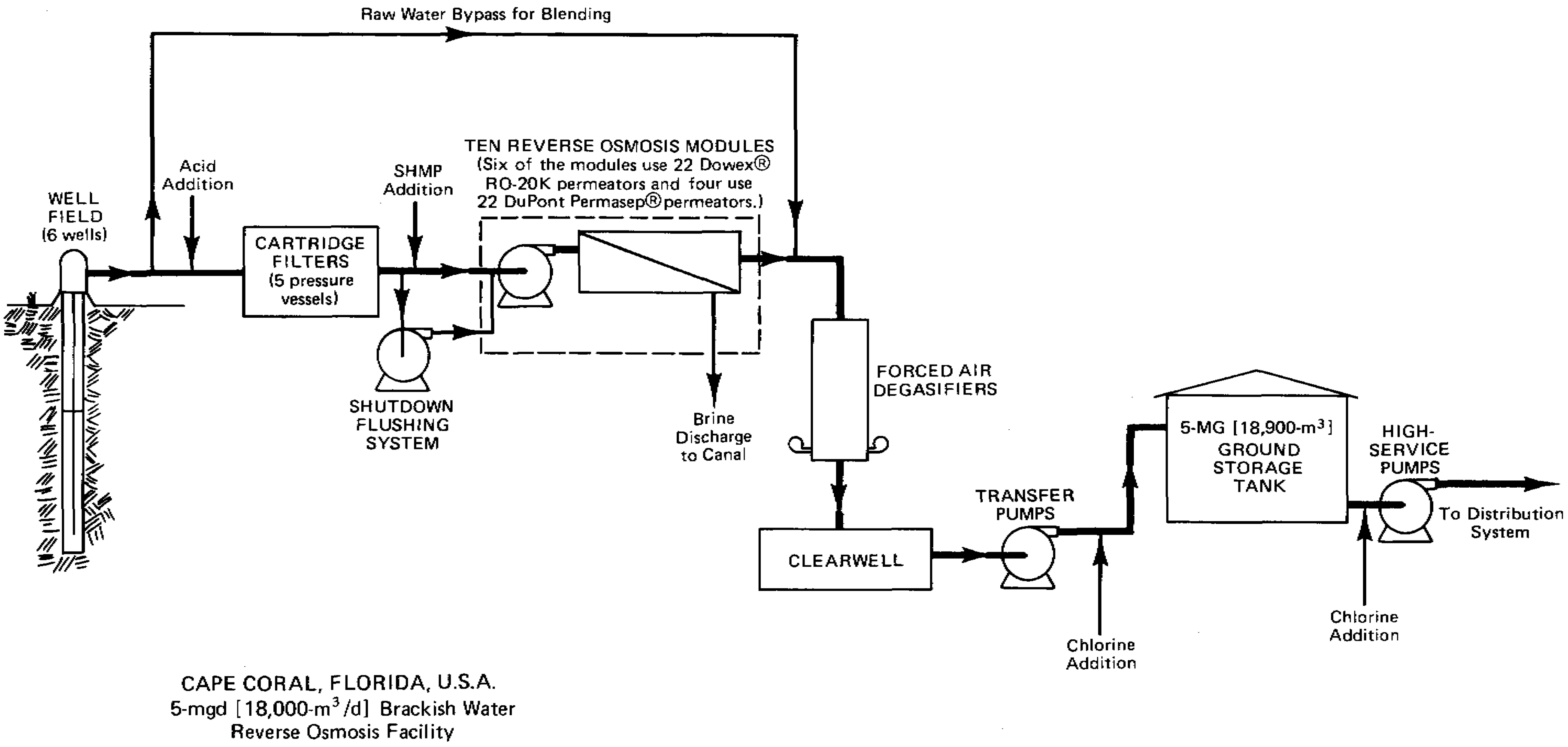
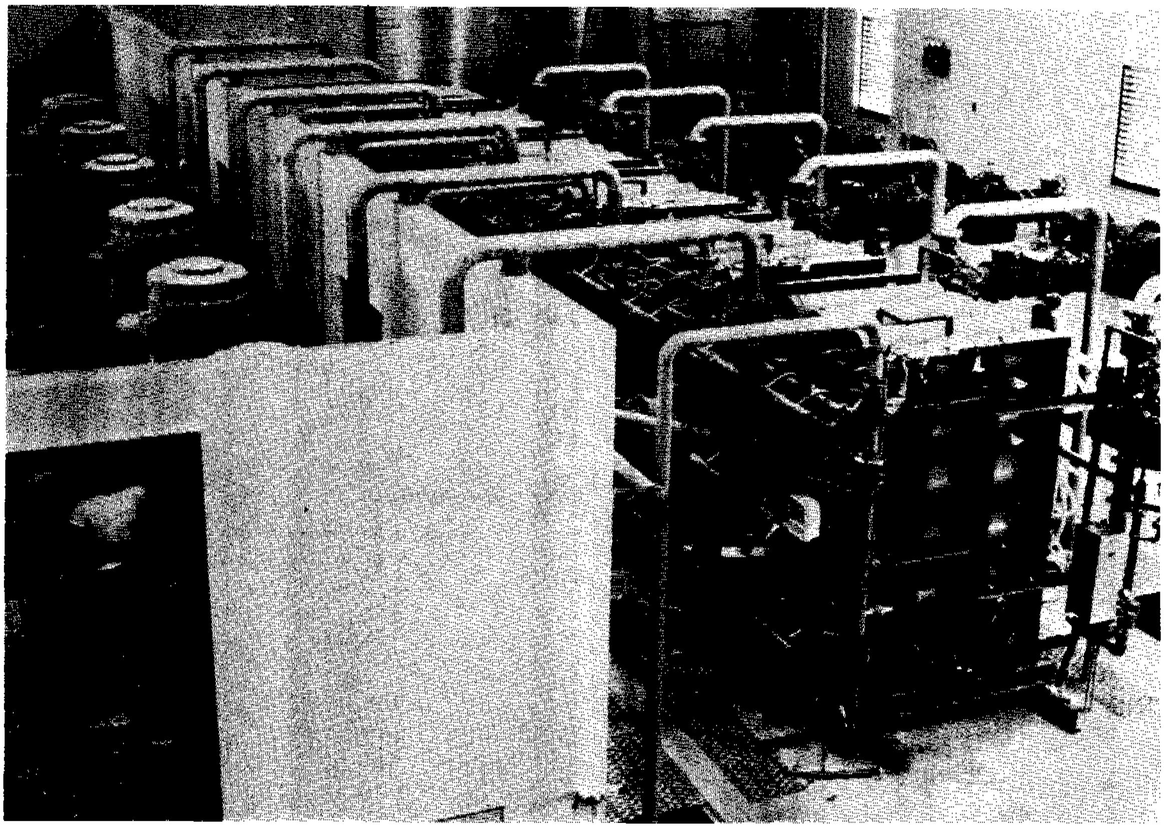


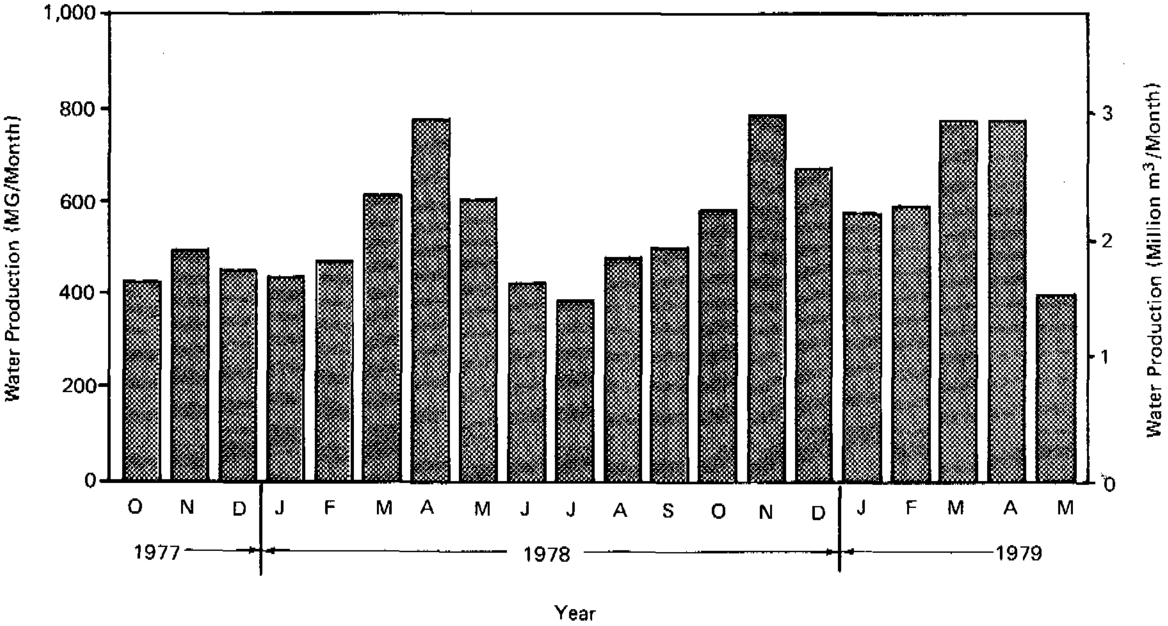
FIGURE 5-14. Process flow diagram of the Cape Coral reverse osmosis plant.



CH2M HILL, Inc., Photo.

The vertical turbine high-pressure pumps are on the left, separated from the permeator banks by an acoustical wall. On the far right are the high-service pumps.

FIGURE 5-15. Interior of the Cape Coral reverse osmosis plant.



Data to May 1979 from Shannon, 1979.

FIGURE 5-16. Monthly production of the Cape Coral reverse osmosis plant.

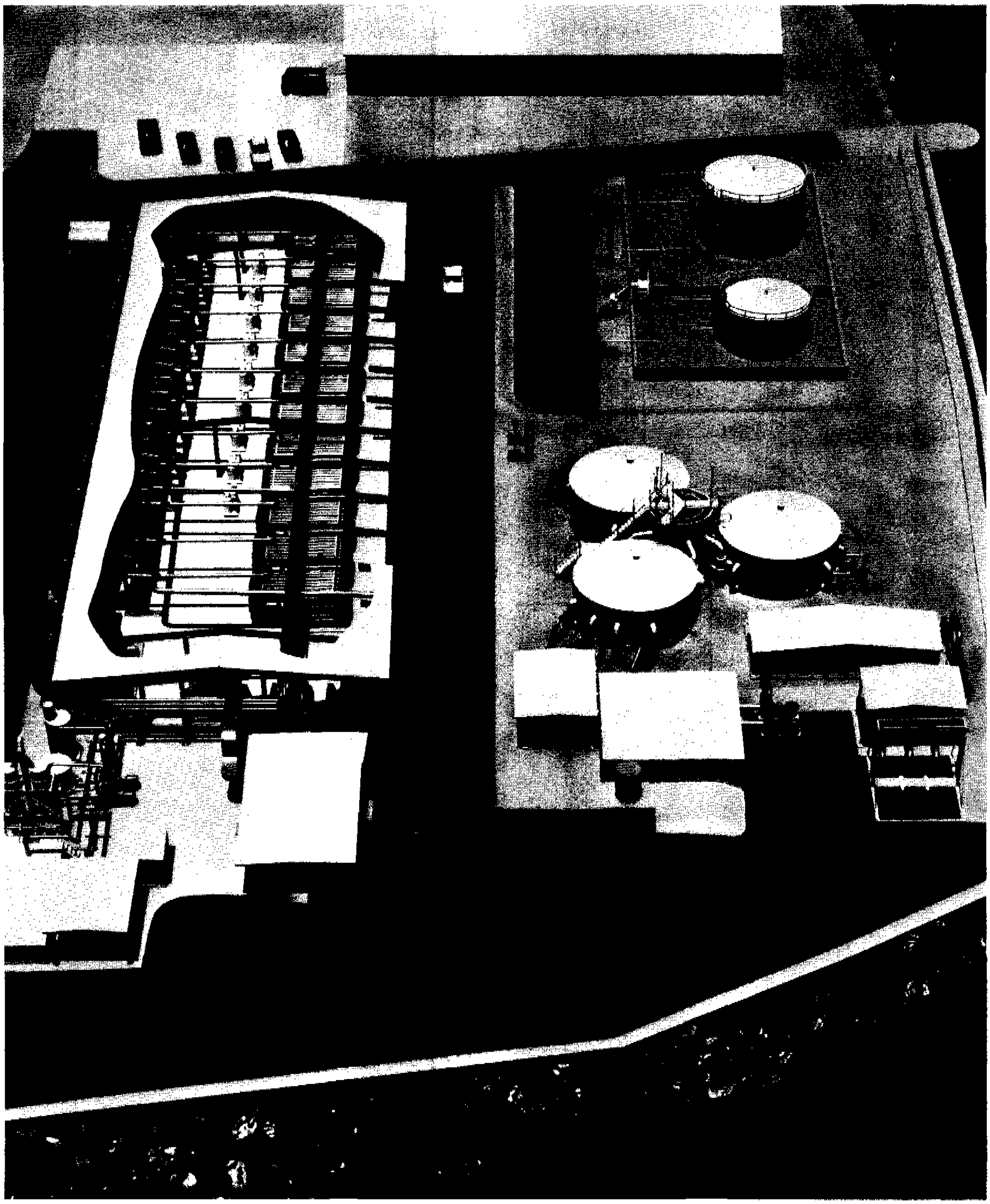
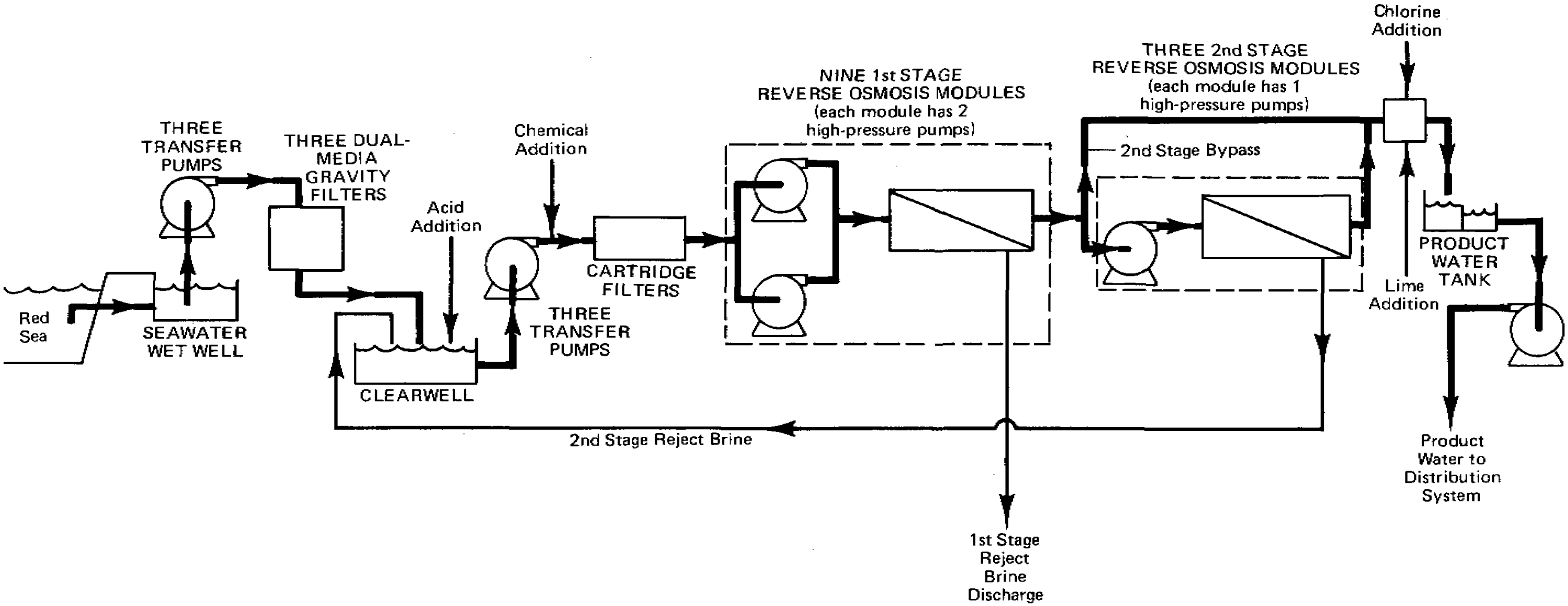


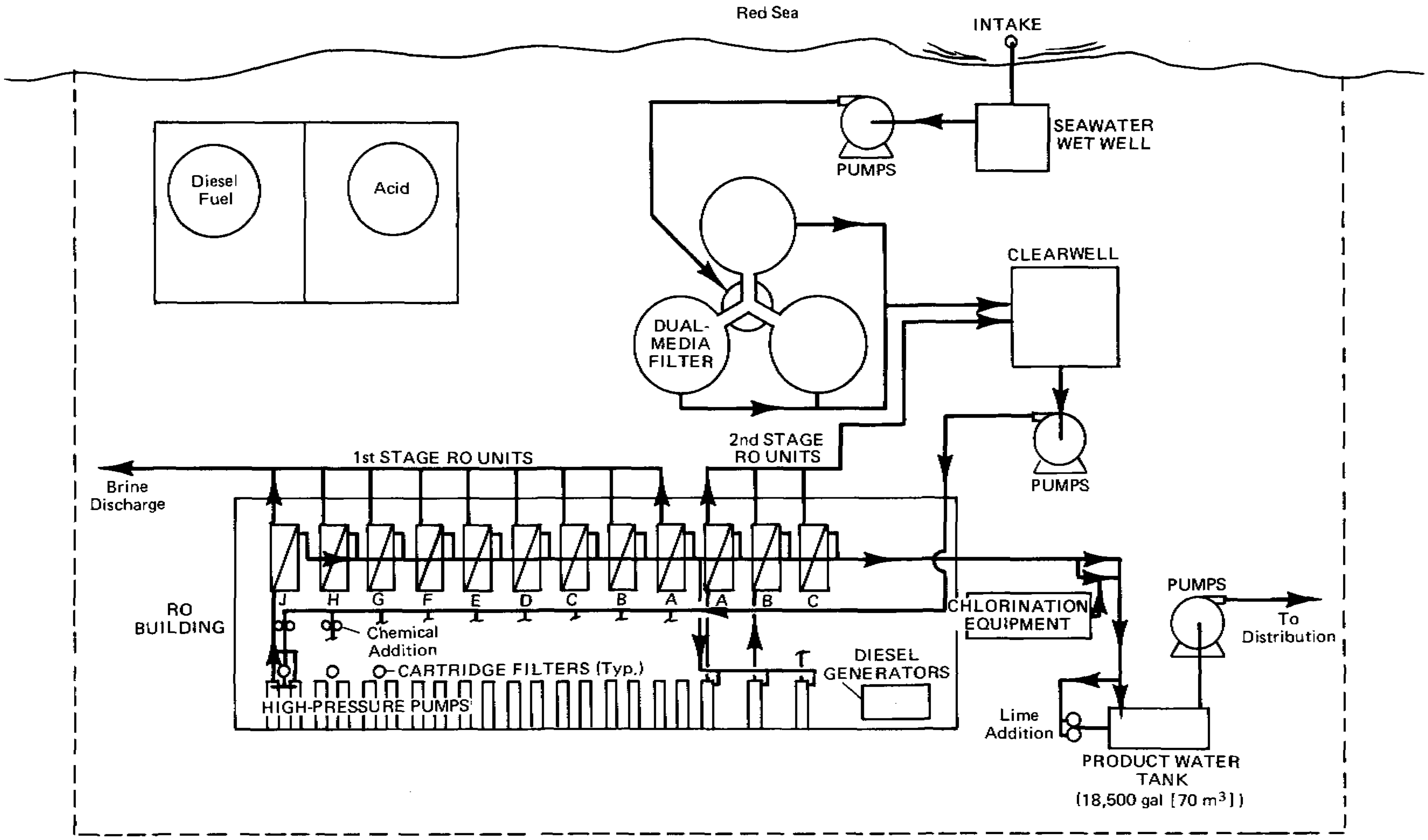
Photo Courtesy of Fluid Systems Division of UOP.

FIGURE 5-17. Model of the 3.2-mgd [12,100-m^3/d] seawater reverse osmosis facility at Jeddah, Saudia Arabia.



JEDDAH, KINGDOM OF SAUDI ARABIA
3.2-mgd [12,100-m^3/d]
Seawater Reverse Osmosis Facility

FIGURE 5-18. Process flow diagram for the Jeddah seawater reverse osmosis facility.



Adapted from Al-Golaikah et al., 1975
& Hickman et al., 1979.

FIGURE 5-19. *Layout of the Jeddah seawater reverse osmosis facility.*

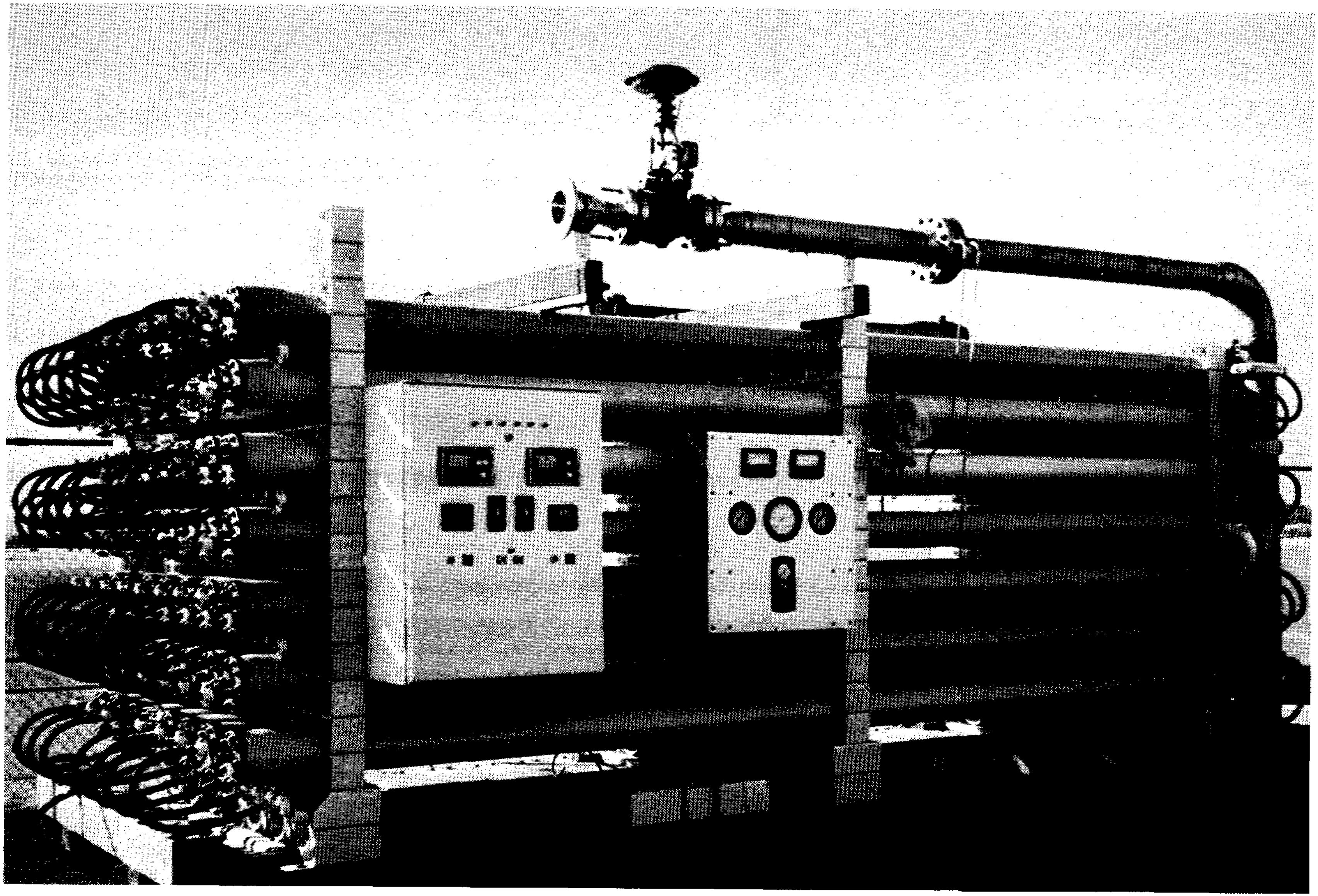


Photo Courtesy of Fluid Systems Division of UOP.

FIGURE 5-20. One of the nine first-stage units for Jeddah's seawater reverse osmosis facility.

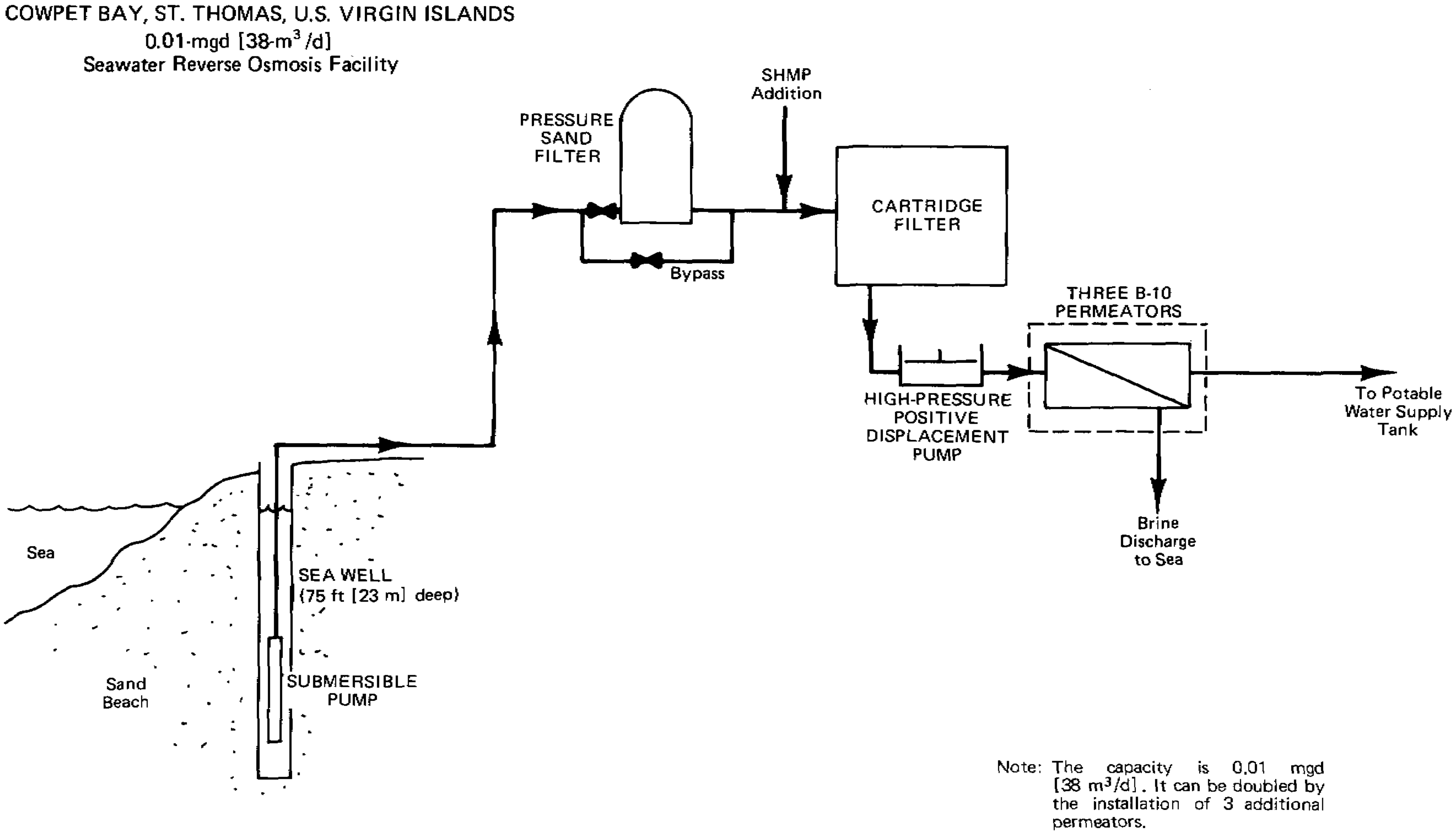


FIGURE 5-21. Process flow diagram for the Cowpet Bay seawater reverse osmosis facility.

CHAPTER 6

Chapter 6
ELECTRODIALYSIS

THEORY

Electrodialysis is the last of the four basic desalination processes discussed in this text. Desalination by electrodialysis is based on the following facts.

1. Most of the salts dissolved in water are ionic in nature.
2. These positively or negatively charged salts are attracted by an opposite electrical charge, i.e., positive ions (cations) will be attracted to a negative pole or electrode.
3. Membranes can be devised which will be selective in the type of charged ion which they will pass or reject, i.e., membranes can be made which allow negative ions (anions) to pass but reject cations. This type of membrane is referred to as an anion-permeable membrane, and one which rejects anions is called a cation-permeable membrane.

Most of the dissolved constituents in saline water are ionic. These constituents dissociate and are dispersed in water, effectively neutralizing their individual charges (see Figure 6-1). When electrodes, connected to an outside source of direct current, are placed in a container of saline water, the current is carried through the solution (a mild electrolyte due to the ionized salts), and the ions tend to migrate to the electrodes which carry the opposite charge. Thus anions, such as chloride (Cl^-), migrate toward the positive electrode, and cations, such as sodium (Na^+), migrate toward the negative electrode.

To utilize this phenomenon to desalinate water, membranes which will allow either cations or anions (but not both) to pass are placed between the electrodes, as illustrated on Figure 6-1. These membranes are arranged alternately, an anion followed by a cation-selective membrane. As the electrodes are charged, the anions are diverted from the main product stream and pass through the anion-selective membrane into the concentrate (or brine) cell. The anions are prevented from moving through the adjacent cell wall, as it is a cation-selective membrane and prevents their passage. Similarly, cations move from the dilute stream on the other side of the cation-selective membrane into the concentrate cell. Here they are prevented from moving further toward the negative electrode by the anion-selective membrane. By this arrangement, concentrated and dilute solutions are formed in the spaces between alternating membranes. These spaces, bound by two membranes (one anionic, the other cationic), are called cells. A cell pair consists of two cells, one from which the ions migrated (diluting cells for the product water) and the other in which the ions concentrate (concentrate cell for the brine stream).

A typical network (or stack) has several hundred cell pairs (one dilute and one concentrate cell) so that the proportion of ions removed from the feed stream relative to the current carried by the ions between the electrodes is very large.

In operation, feedwater passes simultaneously in parallel paths through all of the cells to provide a continuous flow of product water and brine stream, thus washing out the concentrated ions.

Elements of an Electrodialysis Unit

An electrodialysis unit is made up of the following basic components:

DC Power Supply (Rectifier)
Membrane Stack
Circulation Pump & Hardware
Pretreatment

These are shown diagramatically and in a photo on Figure 6-2.

DC Power Supply (Rectifier). The main element of the power supply is usually a rectifier which converts alternating current (AC) to direct current (DC). Direct current is applied to the electrodes on the membrane stack to remove the ions from the feed stream.

Membrane Stack. The stack includes the electrodes, membranes (both anion- and cation-permeable), spacers, plumbing necessary to transport water to and from the stack, and the hardware necessary to hold the stack together.

i) Membranes--The membranes are flat sheets, usually made of a plastic film formed on a fabric backing of dynel, glass, or other material to provide strength. Ion transfer sites are added to the membranes with the site charge differing between the anion- and cation-permeable membranes to give each type the characteristics to selectively pass either anions or cations. Manufacturers vary in how they incorporate the ion transfer sites into the membrane. Two types of membranes currently produced are the homogeneous and heterogeneous.

In the homogeneous membranes (such as made by Ionics), the sites are uniformly distributed through the membrane. In the heterogeneous membranes (such as made by Mitsubishi), the sites are distributed as discrete points within the membrane.

Membranes come in a variety of sizes, again depending on the manufacturer and the application. A typical size used in the USA is a rectangle about 18 by 40 inches [46 by 102 cm]. They are usually an unbroken sheet except for holes cut out to form flow channels and holes or slots to guide the assembly of the membranes into a stack. Figure 6-3 shows some typical membranes and spacers. During construction of the stack, the membranes and spacers are aligned in either a vertical or horizontal plane and held together by some type of press or clamping device.

The thickness of the membrane depends on the application, and its selection is a trade-off between membrane properties. Thicker membranes usually have greater strength, increased erosion resistance, and longer life, whereas thinner membranes have lower electrical resistance and hence reduced energy requirements. Typically, membranes are about 0.022 inch [0.56 mm] thick.

The structure of membranes can be tailored during fabrication to alter their characteristics by varying components such as their water content. The higher the water content, the "looser" the membrane, i.e., the easier it is for ions to travel through them. With a lower water content, the membrane becomes "tighter." Thus, manufacturers have a family of membranes, enabling them to tailor membranes to various applications and their particular water chemistry characteristics.

i) <u>Spacers</u>--The spacers separate the membranes and provide a pathway in the cell for the water flow. Three of the major pathway configurations provided by various spacers are shown in Figure 6-4. In each of the three spacers, the flow enters via the feed channel at Point A and then follows the pathway shown by the arrows until it leaves the cell by way of the discharge channel at Point B.

Cells are made up of two membranes with a spacer in between. Cells are stacked with alternating concentrate and dilute cells. In each stage the feedwater is exposed only to the electromotive force for the distance of the pathway in one cell (the hydraulic stage). Figure 6-5 shows different arrangements in which a four-stage configuration can provide the exposure to the electromotive force. By the use of special spacers, more than one hydraulic stage can be placed between a set of electrodes. The number of stacks, stages, and electrodes is determined at the time of design, based on site-specific information.

Spacers can be formed to provide different types of flow paths as shown in Figure 6-4. The sheet flow and tortuous path flow are two of the most commonly used designs, but work is now going on in the use of slanted strap spacers, a modification of the tortuous path (Mattson and Lundstrom, 1979).

ii) <u>Electrodes</u>--One pair of electrodes is required for each electrical stage with, typically, two hydraulic stages per electrical stage. Normally no more than two electrical stages are present in a single membrane stack whereas a pair of electrodes is needed for each electrical stage. The electrodes are generally constructed of niobium or titanium with a platinum coating.

A reaction occurs at each of the electrodes. Hydrogen ions and oxygen and/or chlorine gas are formed at the anode (positive electrode), and hydrogen gas and hydroxyl ions are formed at the cathode (Mason and Kirkham, 1959). Because a separate stream is generally used in the space adjacent to the electrodes in each stack, the byproducts of the reactions are confined to these streams. The anode stream is normally acidic due to the hydrogen ions, and the cathode stream is basic. The pH difference is partially neutralized by combining the streams. In most units, the electrode streams are kept apart and then usually discharged to waste, although in some instances this water is treated and recycled. Where the pH increase (at the cathode) could cause the formation of a calcium or magnesium precipitate, acid is often added to the electrode stream to keep the cathode stream acidic and maintain the precipitate in solution.

In the electrodialysis reversal process (described later), the anode and cathode are electrically reversed several times per hour, thus alternating the environment at the electrodes from acidic to basic on a regular basis. This acts to significantly reduce scale formation.

Circulation Pump and Hardware. In the electrodialysis process the water pump(s) is used only for circulation of the water through the stack. The head loss for this circulation varies with the construction of the stacks, number of stages, stacks, etc., but generally a pumping pressure of only about 50 to 75 psi [3.4 to 5.1 atm] is needed. This is considerably lower than the 250-400 psi [17.0 to 27.2 atm] required in the brackish water reverse osmosis process. Because of the reduced pressures involved, the connecting piping, valves, etc., can be made with material and tolerances suitable for the lower-pressure range. This has allowed the use of a great deal of standard plastic pipe and fittings. The use of plastic pipe produces benefits regarding lower cost (compared to stainless steel, etc.), high resistance to corrosion in a saline environment, and ease of construction.

Pretreatment. A certain degree of pretreatment of the feedwater supply is necessary in order to prepare it for desalination in the stacks. Pretreatment depends on the specific water being treated, but it usually includes the removal of suspended or dissolved solids which could adversely affect the surface of the membranes or mechanically block the narrow passageways in the individual cells. With the electrodialysis reversal process, the need for standard pretreatment is greatly reduced.

Major Processes

Two major processes are commercially in use today, generally referred to as the electrodialysis (ED) and the electrodialysis reversal (EDR) processes.

Electrodialysis (ED). This is basically the process which has been described above in the "Theory" section. Feedwater is pumped to the stacks, where it passes through the cells located between the electrodes. The electric charge on the electrodes provides the force to induce the ions to migrate through the membranes from the main feedwater stream into the concentrated brine stream. The two streams, product water and brine, are continually transported from the stacks in their respective pipelines.

This system often requires the addition of acid and/or polyphosphate to the brine stream in order to inhibit the precipitation of sparsely soluble salts in the stacks. To maintain performance, the membrane stacks need to be periodically cleaned to remove scale and other surface-fouling matter. In most electrodialysis units, this is done in two ways: (1) cleaning-in-place and (2) stack disassembly. Using a cleaning-in-place system, special cleaning solutions are circulated through the membrane stacks. These aid in cleaning but at regular intervals the stacks need to be disassembled and mechanically cleaned to remove scale and other surface-fouling matter. Those units without a cleaning-in-place system generally have shorter intervals between stack disassembly.

Electrodialysis Reversal (EDR). The reversal process operates on the same basic principles as the standard electrodialysis unit, except that both the product and the brine cells are identical in construction. At intervals of 3 to 4 times per hour, the polarity of the electrodes is reversed and the flows are simultaneously switched by automatic valves in the stacks so that the product cell becomes the brine cell and the brine cell becomes the product cell. The salts are thus transferred in opposite directions across the membranes.

Following the reversal of polarity and flow, the product water is discharged until the cells and lines are flushed out and the desired water quality is restored. This takes approximately 1 to 2 minutes. The reversal process aids in breaking up and flushing out scale, slimes, and other deposits in the cells.

This automatic cleaning action eliminates the need to continuously add acid and/or polyphosphate, and scale formation in the electrode compartments is minimized due to the continuous alternation of the environment from basic to acidic. Three methods of removing scale and other surface-fouling matter are used in the reversal process. These include: (1) the cleaning-in-place and (2) stack disassembly methods used in the standard electrodialysis units, plus (3) the reversal of flow and polarity in the stacks. This greatly extends the intervals between the rather time-consuming job of stack disassembly and reassembly. The overall result is reduced maintenance time.

COMMERCIAL OPERATION

History

The phenomenon of membranes having selective ionic permeability was investigated at least as far back as the 1890's (Curran et al., 1976). Much of the early work was done relative to the properties of living cells. Separation processes utilizing the principle of the attraction of charged particles to electrodes in an electrolyte have been developed, as electrophoresis, by the various life sciences.

By fabricating ion-selective membranes from various materials, several units were built to purify water in the 1910's and 1920's in Europe (Friedlander and Rickles, 1966). The development of membranes with both good ion selectivity and good mechanical characteristics was a major problem. In 1935 ion exchange resins were synthesized and in 1940 a multicell unit was developed having alternate product and brine cells similar to the present-day configuration (Curran et al., 1976).

The present-day electrodialysis industry came into being in the 1950's in the USA when sheet membranes made from ion-exchange resins were perfected and patented (Juda and MacRae, 1953). These were used to demonstrate, on a pilot scale, the desalination of brackish water and the membrane desalination industry was underway (Katz, 1977). At that time the only competing process being widely used for brackish water desalination was distillation, and compared to it electrodialysis offered the potential for considerable savings in equipment and energy costs.

The first significant sales for electrodialysis units were to petroleum companies operating in the Middle East. These sales started in 1954 and have continued to date. In 1958, electrodialysis was first used in the USA for a municipal application at the City of Coalinga, California. Other municipalities followed suit, and electrodialysis units were soon treating a variety of brackish waters around the USA and overseas.

Although much of the original development was done in the USA by the Ionics Corporation, other manufacturers in the USA and other countries such as England, Holland, France, and Japan, developed ion exchange membranes and/or began producing electrodialysis units in the late 1950's. Despite this competition, the market has been dominated by Ionics, which has produced about half of the world's standard electrodialysis units and all of the electrodialysis reversal units according to OWRT's 1977 survey (El-Ramly and Congdon, 1977). Their proportion of the standard electrodialysis market has diminished, primarily because they were switching to the production of reversal units from 1970 through 1975.

Extent of Usage

Usage, which is almost exclusively in brackish water desalination, began in the 1950's and has developed approximately as follows (Katz, 1977, and El-Ramly and Congdon, 1977):

Year	Installed Worldwide Capacity (mgd)	Installed Worldwide Capacity (m^3/d)
1952	0	0
1970	5	18,900
1977	51.7	196,000
1979 (estimate)	55	208,000

These figures include both standard and reversal units combined. In 1979 the approximate breakdown of installed capacity between these processes was estimated as follows:

Electrodialysis (ED)	39 mgd [148,000 m^3/d]
Electrodialysis Reversal (EDR)	18 mgd [68,100 m^3/d]

ENGINEERING CONSIDERATIONS

Specific Energy Usage

The distribution of energy usage in an electrodialysis unit is shown in Table 6-1.

Table 6-1
MAJOR ENERGY USES IN THE ELECTRODIALYSIS PROCESS

Unit	Energy Usage
Pumps	2 to 4 kWh/kgal [0.5 to 1.1 kWh/m^3] product
Membrane Stack	2.5 kWh/kgal [0.7 kWh/m^3] product/1,000 mg of TDS removed
Power Supply	5 percent of total energy consumption

The major energy usages are in the membrane stacks and the power necessary to pump the water through the process. Generally, overall energy usage is proportional to the amount of dissolved salts removed in the process. For any given water, the efficiency of the process depends on many factors, including:

1. Efficiency of motors, pumps, and rectifiers.

2. Stack losses.

3. Water temperature.

Efficiency in the motors, pumps, etc., is obtained not only by direct electrical and mechanical efficiency but by matching the motor pump, etc., to the application. Due to the need to use commercially available components, ideal matching is not always possible throughout the range of units available.

Stack losses are a factor which manufacturers are working constantly to reduce. In efforts to optimize various components in the stack, research is continuing in the development of such components as thinner, lower-resistance membranes; thinner water flow spacers; and engineered flow paths within the spacers.

Water temperature affects the conductance in cells such that as temperature rises, the conductance increases and hence the resistance (and in turn power consumption) decreases within the stack. Figure 6-6 shows a plot of the effect of temperature on TDS removal. Although there is a marked increase in removal rate (about 1 percent per degree F [1.8 percent per degree C]), this is only for the direct current used in the stack. The combined energy savings for the overall unit operation are usually somewhat less or even minimal, because the energy used by the pumps, rectifier, etc., is not significantly affected by the feedwater temperature change. For cases in which the energy utilization by pumping is high relative to the stack usage, there is a much smaller change than when stack power makes up a large percentage of energy usage. Thus the effect of temperature is more pronounced in larger, high-TDS plants than in smaller, low-TDS installations.

Pretreatment

Many of the techniques used in pretreatment for the electrodialysis process are similar to those used in reverse osmosis. Where suspended solids exist, sand filters are often used ahead of the membranes, as are cotton, polypropylene, or other cartridge filters. As in reverse osmosis, the cartridge filters are used more as a safety filter and are not meant to take on the major burden of filtration in water with excessive solids.

In the standard electrodialysis process, the likelihood of the precipitation of sparsely soluble salts is minimized by reducing the pH through acid addition and/or the addition of an inhibiting agent such as polyphosphate to the brine stream. Iron and manganese are removed by the use of green sand filters, aeration, or other standard water treatment methods, depending on the size of the equipment and the preference of the designer.

The overall requirements for pretreatment in electrodialysis are somewhat less rigorous than for reverse osmosis due to the nature of the salt separation and the larger passages provided. The electrodialysis reversal process reduces the necessity for chemical pretreatment for many applications, but removal of suspended solids, iron, and maganese is still critical to avoid fouling and/or other problems with the system. Manufacturers generally recommend pretreatment with the reversal process if the feedwater contains any of the following:

1. Free chlorine
2. Iron (over 0.3 mg/l)
3. Manganese (over 0.1 mg/l)
4. Hydrogen sulfide (over 0.3 mg/l)
5. Turbidity (over 2 JTU's)

In every case, of course, a careful examination of the prospective feedwater would be necessary to determine suitability and pretreatment.

Another function for which pretreatment is often necessary is to prevent the formation or deposit of organic growth (slime) on the inner surfaces of the membrane stack. As in the reverse osmosis process, these surfaces provide a place for organisms to grow if sufficient suitable material is present for them to metabolize in the feedwater. A chlorination-dechlorination step is used if the problem is of sufficient magnitude; otherwise, the organic slimes are removed in a cleaning step. The dechlorination step is necessary to protect the membranes from continuous exposure to free chlorine. Another method that has been used to remove the bacterial growth in the stack is the shock treatment of the solution for short periods of time. This has proved to be an effective method.

Post-Treatment

Post-treatment depends on the type of feedwater and pretreatment provided at the installation. With the electrodialysis reversal units, the amount of post-treatment can be usually reduced to only disinfection.

Operational Modes

There are two common operational modes used in electrodialysis: continuous and batch. In the continuous mode, which is most common in water treatment applications, the feedwater moves continuously into the unit and product water is always being produced and the brine recycled to maximize produce recovery.

With a batch operation a quantity of feedwater is processed and stored and then reprocessed through the same unit. This continues until the level of TDS in the product is suitable for use, at which point a new batch of feedwater is processed. The batch process is used to treat difficult feedwaters or in applications with insufficient stages available to reduce the TDS to an acceptable level.

Equipment Optimization

In optimizing the design and construction of a facility, a balance must be achieved between the power used and the equipment supplied. Since the amount of dissolved solids removed is proportional to the current utilized, the higher the current density, the more desalination that is possible with a given membrane area and the lower capital investment.

This maximization of dissolved solids removal per unit of power by increased current application is limited by polarization effects. Polarization occurs during electrodialysis in the dilute (or product) cells when a high enough rate of ion transport takes place to create a depletion of ions in the water adjacent to the membrane. This reduces the conductance value of the water (electrolyte), and the resistance increases sharply, resulting in higher energy usage in the stack.

In the absence of ions in the boundary layer, the continued high current density causes a dissociation of water molecules adjacent to the membranes and the diffusion of the hydrogen and hydroxyl ions through the membranes. As these ions enter the concentrating cell, they alter the pH and in the case of the anion transfer membrane this can result in a higher pH that will encourage the scaling of precipitates such as calcium carbonate and the formation of a high resistance gas layer on the membrane surface (Mason and Kirkham, 1959).

Optimization. The problems associated with polarization have limited the practical current densities obtainable. In an effort to increase the current density at which polarization becomes a serious problem, some manufacturers have incorporated turbulence promoters into the spacers. These create turbulence at the surface of the membrane and tend to break up the boundary layers thereby allowing higher current densities.

Operational Problems

A variety of operational problems can be experienced with electrodialysis facilities. Some of the major ones are discussed in the following paragraphs.

Scaling. Scale is formed in the membrane stacks due to polarization, the supersaturation of the brine stream, or other factors. Scale fouls the membrane surfaces, blocks passages in the stack (changing flow patterns), and creates areas of resistance. These areas of high resistance, called "hot spots," occur when the feedwater flow is stopped or slowed in its passage through the product cell. The slowly moving water then becomes highly desalted due to the longer period of exposure to the electromotive force. The highly desalted water has a low conductivity and offers a high resistance to current flow. Hot spots can consume excess power and reduce the efficiency of the stack.

Some scale can be removed by introducing special chemicals, such as acids, into the stacks in an attempt to dissolve or loosen the scales so that they can be washed out. In more severe cases the stack is disassembled and the membranes and spacers soaked in cleaning solution and/or scrubbed to remove the scale. Stacks are designed to be readily disassembled for this cleaning procedure. However, disassembly is a time-consuming job because of the large number of membranes and spacers involved.

Leaks. Operating and/or maintenance problems can result from leaks in two parts of the electrodialysis stacks: (1) between the stacked membranes and spacers; and (2) through the membranes.

Since stacks using membranes and tortuous path spacers are assembled much like a deck of cards, without a sealant or special gaskets, the ability of the stack to remain watertight is dependent on the material fitting tightly together, which, in turn, depends on the spacers and membranes being uniform in thickness and the stack being uniformly pressed together. This can be a problem, since some stacks contain a total of 1,200 to 1,800 membranes and spacers (300-450 cell pairs). Thus it is extremely important that they be manufactured within acceptable tolerances. In constructing the stack, the membranes and spacers must be assembled correctly and tightened carefully to apply uniform pressure and flatness all across the surface of the membranes where they join the spacers. Leaks can occur both between channels in the spacers and to the outside of the stack; the former are invisible from the outside of the stack. Figure 6-4, showing the tortuous path spacer, demonstrates the potential for path shortcuts and outside leakage in improperly assembled and/or aligned stacks.

Other leaks can develop through cracks or tears in the membranes or spacers as a result of manufacturing defects, improper handling, excessive tightening of the stacks, aging, and other causes. The immediate result is usually an intermixing of water between the dilute and concentrate cells, and the ultimate result is reduced product water quality. Normally stacks are operated with a slightly higher pressure on the product water side to prevent this intermixing.

Electrode Degradation. The electrodes at each end of the electrical stage within a stack distribute the current over the membranes. These electrodes are flat plates of metal (such as niobium or titanium) or carbon and are usually plated with platinum or (in the case of cathodes) stainless steel. In the electrodialysis reversal process, both electrodes are platinum-plated since both alternate in being the cathode and anode. The electrodes are exposed to separate rinse water streams and in the course of operation oxidation and reduction reactions take place at the anode and cathodes, respectively. These reactions create problems which result in (1) the degradation of the electrodes, especially at the anode, where oxidation can result in the erosion of the metal, and (2) the production of chlorine and/or oxygen, depending on the pH and constituents in the rinse stream. The effect on the cathode is not quite as severe and usually results in the formation of hydrogen gas. This in turn could create a high pH that may favor the formation of scale.

In operation, the rinse streams are usually acidified to inhibit scaling. The rinse streams from the anode and cathode compartments are often combined and then recycled again to reduce the amount of acid additive necessary. The electrodes must be replaced periodically.

In the electrodialysis reversal process, the electrodes alternate from being a cathode to an anode several times per hour. This serves to equalize the wear on the electrodes and chemically removes scale formed during the cathode operation by means of acid generated during its anodic cycle.

Cleaning. Cleaning is performed either by flushing the stack with suitable cleaning solutions or disassembling the stack to soak and/or scrub the individual components. Obviously, since the stack often contains 1,200 to 1,800 components, the option to disassemble the stack for cleaning is usually exercised only when absolutely necessary. Usually in-place cleaning is utilized whenever possible, and many units are now designed with automatic chemical cleaning systems. Cleaning methods and solutions employed include those containing sodium hydroxide, hydrochloric acid, sulfuric acid, and sodium chloride (Kishi et al., 1977, and Geishecker, 1977).

When in-place cleaning is insufficient, the stacks must be disassembled, cleaned, and reassembled. The individual spacers and membranes can be soaked in appropriate cleaning solutions and are rugged enough to be hand-scrubbed if necessary. In many cases when stack disassembly is necessary, only a portion of the membranes need to be cleaned. The membranes to be cleaned can usually be identified by taking resistance measurements (called stack probing) on individual membranes, or groups of membranes while the unit is operating.

Membranes

The two major types of membranes being manufactured for use in electrodialysis have the same basic principle of ion selectivity but some differences in construction and theoretical operation. The membranes made by Ionics consist basically of ion-transfer sites distributed through a homogeneous base which is made up of a cross-linked vinyl aromatic plastic (Curran et al., 1976 and Parsi, 1978). The active transfer sites of the cation membrane contain mobile counter ions (H^+) and fixed charged groups ($-SO_3^-$). The anion membrane is made up of fixed quaternary ammonium groups ($-NH_4OH^-$) with the hydroxyl (OH^-) being the mobile ion.

The heterogeneous membrane, produced by the Japanese firm of Mitsubishi, involves powdered anion or cation exchange resins uniformly dispersed throughout the film of polypropylene plastic that makes up the membrane. The film is then treated to create microcracks and cavities in the plastic (Kishi et al., 1977). The microcracks and cavities permit transport of the ions, while the ion exchange resins provide the selectivity.

The following characteristics represent features of an ideal electrodialysis membrane:

High thermal stability

High mechanical strength

High resistance to chemical and biological degradation

High dimensional stability

High ionic selectivity

Low cost

Low electrical resistance

Low diffusion of water

Since some of these properties are mutually exclusive, the commercial membranes produced are necessarily a compromise of these characteristics.

Operating and Installation Information

Table 6-2 contains information on operating specifications of various commercially available electrodialysis reversal units. With electrodialysis, it is very important that the unit be matched to the feedwater source and economic conditions at the site. The modular nature of electrodialysis permits the tailoring of a unit by varying stacks, stages, and number of cell pairs to fit the exact requirements of an application.

Table 6-3 provides some general information on the shipping and installation of some typical electrodialysis units. Again, the exact size, weight, and power requirements depend on the combination of stacks and stages provided.

STATE OF THE ART

General

The commercial growth of electrodialysis for brackish water desalination has developed slowly but steadily since its beginnings in the 1950's. Data are now available on long-term experience with facilities that have been operating up to 20 years. Continual improvements have been made on the equipment during this time, and sufficient experience has been obtained to receive firm guarantees from manufacturers regarding operating parameters such as water quality, power consumption, and chemical usage.

The commercial development of the electrodialysis reversal units during the 1970's has been a major step in the industry.

In the appropriate, proper applications, electrodialysis has proven to be a reliable and economical desalination process.

Major Technical Advances Since 1972

The major technical advances since 1972 have come about in two areas: (1) the commercial introduction of the electrodialysis reversal process, and (2) the increased reliability stemming from minor design changes based on process experience.

Electrodialysis Reversal Process. After development in the late 1960's and early 1970's, the reversal process was introduced commercially by Ionics, Inc. By 1974 this company had shifted almost all their production to this process. By the time of the OWRT inventory (1977) they had installed 86 reversal plants with capacities greater than 0.025 mgd [95 m^3/d] worldwide (versus about 170 for all electrodialysis processes) plus several times

Table 6-2
TYPICAL OPERATING SPECIFICATIONS FOR VARIOUS ELECTRODIALYSIS REVERSAL UNITS

Feedwater of 1,500 mg/l at 80°F[a]

Unit Model	(--)	V-2	V-2	X-2	X-2	XX	100
Capacity-Product	(mgd)	0.015	0.030	0.050	0.110	0.300	1.15
Feedwater TDS	(mg/l)	1,500	1,500	1,500	1,500	1,500	1,500
Product Water TDS	(mg/l)	70	160	80	210	130	140
TDS Removal[b]	(%)	95	89	95	80	91	91
Product Water Recovery[b]	(%)	73	76	74	80	80	90
Power Consumption[b]	(kWh/kgal)	11	8	8	7	7	5
Number-of Stages	(--)	6	4	4	3	3	4

Feedwater of 3,500 mg/l at 80°F[a]

Unit Model	(--)	V-2	V-2	X-2	X-2	XX	100
Capacity-Product	(mgd)	0.015	0.030	0.050	0.110	0.300	1.15
Feedwater TDS	(mg/l)	3,500	3,500	3,500	3,500	3,500	3,500
Product Water TDS	(mg/l)	170	390	175	525	320	380
TDS Removal[b]	(%)	95	89	95	85	91	89
Product Water Recovery[b]	(%)	72	75	73	78	70	85
Power Consumption[b]	(kWh/kgal)	17	165	15	14	13	10.5
Number of Stages	(--)	6	4	4	3	3	4

Source: Data from manufacturer's literature, Ionics Bulletins No. 101-M-1 (1979), 102-M-1 (1979), 103-M-1 (1979), and 105-E (1979).

[a] Assumes a feedwater with the TDS made up by 75 percent sodium chloride. Waters with lower dissolved solids can be treated at lower energy consumption, whereas waters with higher dissolved solids (up to 7,000-10,000 mg/l) require more energy.

[b] The recovery, removal, and power requirements can often be varied and optimized by using various configurations of stages and stacks to suit the economic and site-specific conditions.

Note: Metric conversion factor: kWh/kgal ÷ 3.785 = kWh/m^3.

Table 6-3
INSTALLATION AND SHIPPING DATA ON VARIOUS ELECTRODIALYSIS REVERSAL UNITS

		Typical Installation[a]			Shipping Information[a]	
Product[b] Capacity (mgd)	Feedwater[b] Flow (mgd)	Floor[c] Area (ft^2)	Overhead Clearance (ft)	Maximum Power Requirements[d] (kVa)	Weight (tons)	Volume (ft^3)
0.03	0.055	300	10	40[e]	2.6	390
0.105	0.175	410	10	75[e]	6.3	900
0.264	0.445	665	10	200[e]	11.2	1,300
1.150	1.715	1,250	12	375 to 755[f]	29 to 59	3,400 to 6,800

Note: Metric conversion factors: mgd x 3,785 = m^3/d; ft^2 x 0.0929 = m^2; ft x 0.3048 = m; tons (short) x 0.907 = metric tons; ft^3 x 0.0283 = m^3.

Source: Data from manufacturer's literature, Ionics Bulletins No. 101 (1976), 102-E (1977), 103 (1976), and 105-E (1979).

[a] Size, weight, and power vary depending on the configuration for the particular application.
[b] Maximum flow--actual flow depends on configuration (stacks and stages) required for the particular application.
[c] This area is for the unit and maintenance clearance only. Additional area will be required for the completed facility.
[d] Not necessarily the actual consumption, which normally will be lower.
[e] 480, 460, 380 or 220 volts, 3∅, 50/60 Hz.
[f] 480, 460, or 380 volts, 3∅, 50/60 Hz.

that many smaller plants with a capacity of about 18 mgd [68,100 m^3/d] (versus a capacity of about 55 mgd [208,000 m^3/d] for all electrodialysis processes).

The advantages of the reversal process have been outlined earlier in this chapter.

Reliability. The domination of the electrodialysis market during the past 30 years by one firm, Ionics, has produced some advantages in that they have worked almost exclusively on one standard model. This has developed into their present-day product line using many interchangeable parts between models.

Ionics has thus been able to concentrate on improving the basic components and their reliability. These improvements have come about through research and development as well as extensive field experience.

Other Developments. One of the most aggressive development programs in the electrodialysis field outside the USA is taking place in Japan, where a number of companies, including Asahi Glass, Asahi Chemical, Mitsubishi Industries, and Tokuyama Soda, have been involved in the development of electrodialysis. Japan presents a special case, since salt for industrial and food purposes is made within the country under the protection of a monopoly. This protected industry began to investigate and use electrodialysis to replace the expensive Japanese salt gardens for salt production soon after electrodialysis was commercially developed in the USA in the 1950's (Yambe, 1977). Along with the development of salt manufacturing, the companies have applied electrodialysis to the production of potable water as well as some industrial uses.

During the intervening years, the Japanese manufacturers have worked on developing vertically oriented membranes and spacers in contrast to the horizontally oriented membranes generally used in the USA and have developed sheet flow configurations instead of tortuous path. Since the beginning of 1972 six municipal water plants have been installed in Japan with a total capacity of about 0.630 mgd [2,380 m^3/d] (Yambe, 1977). The Japanese ED industry has undertaken some experimental and/or commercial developments in automatic chemical cleaning, heterogeneous membranes, seawater desalination (a 0.03-mgd [114-m^3/d] plant at Noshima in 1974), and large membranes (27.6 ft^2 [2.56 m^2]) among others fields (Kishi et al., 1977; Kawahara et al., 1977; and Seto et al., 1977).

Major Constraints

With the high reliability that electrodialysis has attained in much of its equipment, the major constraint to its water application is the high power usage required relative to the reverse osmosis process for the desalination of brackish water at the higher-TDS levels and seawater. Power requirements are expected to be reduced to some degree in the future through the development and improvement of various components within the stack.

A second major constraint is the need for additional work on pretreatment to permit economical treatment of feedwaters containing iron, manganese, and hydrogen sulfide so that they can be efficiently processed by electrodialysis.

A less serious constraint in the application of electrodialysis is partially psychological and partially educational in nature. The tremendous publicity that the reverse osmosis process has received during the past 10 years has seemed to overshadow the electrodialysis process in some people's minds. This, on occasion, may have resulted in a lack of adequate consideration of electrodialysis in situations where it might have been the best application.

APPLICATIONS

Brackish Water

The major application for electrodialysis is for the processing of brackish water. The power consumption and, to some degree, the extent of capital equipment required is directly proportional to the dissolved solids to be removed from the feedwater. Thus, as the feedwater TDS increases, the costs of removal also increase. In the reverse osmosis process, a cost-TDS removal relationship exists, but it is not as pronounced. Often, the variation in the scaling potential of the feedwater and its effect on the percentage of product water recovery can be more important than the cost-TDS relationship.

Thus, for standard applications with low TDS removals, electrodialysis is often the most energy-efficient method to use whereas with feedwaters which require a high level of TDS removal, reverse osmosis usually is less energy-intensive. The economical crossover point between electrodialysis and reverse osmosis based on operating costs is, however, difficult to define precisely and needs to be done on a site-specific basis. Aside from local power costs, other factors often must be considered as well in determining the overall economics. Among these, to the benefit of electrodialysis, are the high recoveries possible (up to 90 percent), the elimination of chemical feeding (with electrodialysis-reversal), and the reliability of performance that is characteristic of the electrodialysis processes.

Seawater

There is only limited use of electrodialysis for high-TDS or seawater desalination at this time. A small 0.03-mgd [114-m^3/d] installation has been operating in Japan since 1974 on the island of Noshima, and a 0.04-mgd [151-m^3/d] inland unit is operating on highly saline ground water, with a TDS of about 38,000 mg/l, in the Libyan desert (Yambe, 1977). With commercial equipment currently available, the energy usage is too high (in comparison to reverse osmosis) to be competitive except under specialized conditions. However, recent work under OWRT programs has indicated a potential for seawater desalination by high-temperature electrodialysis. Results indicate that the power consumption can be reduced to the levels used in seawater reverse osmosis and that a 50 percent recovery ratio can probably be attained.

Irrigation Drainage Water

Electrodialysis can be used to process irrigation drainage water, but the feedwater requires considerable pretreatment to remove suspended solids and other undesirable constituents.

Wastewater Reuse

Although electrodialysis could be used to reduce the dissolved solids in properly pretreated wastewater, the process does not provide the membrane barrier to suspended material, bacteria, etc., found in reverse osmosis. If such a barrier is desired, the use of a reverse osmosis process would be indicated.

HEALTH ASPECTS

Aside from reducing dissolved solids (and thereby improving the quality) electrodialysis does not significantly lower the bacterial level of the water processed except possibly as a byproduct of a particular pre- or post-treatment employed.

ENVIRONMENTAL CONSTRAINTS

Brine

Only mild constraints are involved in electrodialysis units, the most prominent of which is the necessity for disposal of the brine stream. Due to the high recovery rates possible (up to about 90 percent) with the units the volume of the stream is generally lower than with reverse osmosis units.

Air Pollution

Degassification and/or aeration could be used in certain pre- and post-treatment applications of electrodialysis, possibly causing minor air pollution problems. Some caution and planning should be employed if hydrogen sulfide removal is required due to possible odor problems.

Noise

Since electrodialysis operates with low-pressure pumps the noise level is very low compared to that of reverse osmosis installations.

FUTURE PROSPECTS

New Technology Now Being Marketed

The electrodialysis reversal process is a significant new technology which has matured and is now being marketed worldwide.

Other technological advances of a subtler nature, such as improved membranes, spacers, other materials and hardware, is being used to increase reliability and efficiency but is not readily apparent from a cursory look at the equipment or specifications. These developments are often a result of field experience (both good and bad) with new materials and innovations.

A new anion membrane based on aliphatic rather than aromatic chemistry has been made available (by Ionics) in 1980. This membrane will be interchangeable with existing membranes but the manufacturer indicates that based on field tests, the new membranes will have increased capacity and improved anti-fouling properties compared to the existing membranes.

Anticipated Technological Developments

Manufacturers and research agencies are working in a number of areas to make technological improvements in electrodialysis.

Among these are improvements in membrane, spacer, and stack design; solar applications for electrodialysis; and the use of electrodialysis for high-TDS desalination.

Design Improvements. In an effort to reduce capital costs, especially on large multimillion-gallon-per-day installations, certain design changes are being examined. These include the use of larger membranes with a higher effective utilization of membrane area. A contract has just been completed for the U.S. Office of Water Research and Technology in which the membrane size (area) was increased by three times the standard in the USA and the effective area utilized increased from 60 to 80 percent (Mattson and Lundstrom, 1979).

In addition, this same contract involved a study of methods to utilize the stack so that stages within a stack could be independently removed for maintenance without disassembly of the cell pairs in the stages above it. This could save considerable time during maintenance.

Other work is being undertaken in the development of low-resistance membranes with better anti-fouling properties, skin membranes, and improved spacers (Goldstein, 1979, Parsi, 1978, and Mattson and Lundstrom, 1979).

Solar Applications. Some efforts are being carried out in the USA using electrodialysis in combination with photovoltaic cells to desalinate brackish water. The most cost-effective application of this process would be in isolated areas where fuel is costly and difficult to obtain.

High-TDS Applications. Work is also going on to further explore the potential for the use of electrodialysis for high-TDS desalination. A 0.05-mgd [189-m^3/d] (expandable to 0.1-mgd [379-m^3/d]) seawater electrodialysis unit has been built and will be tested during 1980-1981 at the U.S. Government's desalination test facility at Wrightsville Beach, North Carolina. This unit employs a design in which seawater is desalted at a high temperature (in the 150°F [65.6°C] range) in order to take advantage of the reduced resistance of the membranes and the increased conductivity of water at elevated temperatures to significantly reduce the power consumption (McRae et al., 1977, and Goldstein, 1979).

Another possible use for this type of high-temperature high-TDS application would be in desalting geothermal waters. A pilot unit of this type has been tested at the Department of the Interior's geothermal test site at Holtsville, California (Boegli et al., 1977).

There has been some research for the U.S. Water and Power Resources Service (formerly the Bureau of Reclamation) on concentrating brine (and producing potable water) using electrodialysis in combination with ion exchange. Part of the concentrated brine is used to regenerate the ion exchange resins.

PROCESS EXPERIENCE

Previous A.I.D. Studies

Previous editions of the A.I.D. manual on water desalination published in 1968 and 1972 included a number of case histories of various electrodialysis plants. Those case histories are summarized and updated in the following paragraphs.

Buckeye, Arizona (0.65 mgd [2,460 m^3/d], ED). This plant was built in 1962 by Ionics, Inc., to produce potable water for the City of Buckeye. All of the City's water is treated by the plant.

i) Process Description--This is a 0.65-mgd [2,460-m^3/d], two-stage plant with 3 parallel units. It operates on ground water with a TDS of about 2,100 mg/l with a temperature of about 86°F [29.9°C]. A polishing filter is used ahead of the membranes, and sulfuric acid is added in order to control the pH and thus prevent scaling in the brine stream.

ii) Operating Information--The plant has operated quite well with routine cleaning of the stacks by disassembly, cleaning, and reassembly about once per month to remove built-up deposits of scale.

iii) Current Status--The plant is still operating, and an additional stack was added during 1980 to service peak loads.

Webster, South Dakota (0.25 mgd [946 m^3/d], ED). This plant was designed and manufactured by Asahi Chemical Industries (Japan) and began operation in 1962. It supplied the City of Webster with a portion of its municipal water. It was constructed as the Office of Saline Water (OSW) Demonstration Plant No. 3, and its operation was partially funded by OSW until 1972.

i) Process Description--This was a 0.25-mgd [946-m^3/d], four-stage unit with a single train. It used membranes mounted in a vertical plane and assembled on a type of filter press mechanism.

The water in two of the original wells was characterized by the following average analysis (Kaiser, 1967).

TDS = 1,385 mg/l
pH = 7.58
Iron = 1.2 mg/l
Manganese = 0.91 mg/l
Hardness ($CaCO_3$) = 788 mg/l
Temperature = 48°F [8.8°C]

The water was pretreated by ion exchange to reduce iron and manganese concentrations to less than 0.1 mg/l. After passing a final filtration, the feedwater was pumped separately between each stage.

ii) Operating Information--The biggest problem with the plant was in pretreatment of the feedwater--especially to remove iron and manganese. During the period 1966-1972, changes and improvements were made on both the pretreatment system and the stack operation. Originally, ion exchange was used to remove iron and manganese by the City. This was modified to use a potassium permanganate feed and then was modified to use partial lime softening. With improved pretreatment, the overall operation improved and the capacity per stack increased.

The plant was important in that it was one of the first plants to demonstrate the use of potassium permanganate to control iron and manganese fouling. It also showed that proper pretreatment could permit increased flow in the stacks and, in effect, increase the capacity of a unit.

iii) Current Status--Federal funding for the project was stopped in 1972. The pretreatment system, which included partial lime softening, was then turned over to the City of Webster and the electrodialysis plant was deeded to the State of South Dakota. The State used the electrodialysis unit as part of their water plant operation training program. It is no longer operated to supply water to the City. They use the partial lime softening plant for treatment. This reduces salinity about 10 percent, softens the water, and reduces the iron and manganese content.

Gillette, Wyoming (1.5 mgd [5,680 m^3/d], ED). This plant began operation in about 1972 to produce potable water for the City of Gillette.

i) Process Description--This is a 1.5-mgd [5,680-m^3/d] 3-stage plant with seven parallel units. Cold (52°F [11.1°C]) ground water with a TDS content of about 2,500 mg/l is pretreated by aeration followed by lime coagulation to reduce iron, hydrogen sulfide (sulfur), alkalinity, and hardness.

ii) Operating Information--This unit had problems with fouling of the stacks and reduced productivity. The cold feedwater required fairly high energy consumption.

iii) Current Status--The electrodialysis unit was shut down in 1979. The City currently uses the same feedwater for their system using the former pretreatment facilities (aeration and lime coagulation) as the full treatment process.

Additional Case Studies

Sanibel Island, Florida (2.1 mgd [7,950 m^3/d], ED). This plant was constructed for the Island Water Association , a member-owned utility, by Ionics. The initial facility, with a capacity of 1.2 mgd [4,540 m^3/d], was completed in 1973. In April, 1975 it was expanded to 1.8 mgd [6,810 m^3/d] and was further expanded in December, 1979 to 2.1 mgd [7,950 m^3/d]. The plant is located on Sanibel Island and serves consumers on the islands of Sanibel and Captiva. These islands are located off the west coast of Florida in the Gulf of Mexico. Although they are surrounded by saltwater, brackish ground water may be obtained on the islands by drilling into aquifers which are recharged on the mainland.

This facility is of interest for two reasons: (1) it is one of the last standard electrodialysis plants built by Ionics before they switched to the electrodialysis reversal process; and, (2) it has been intensively operated under difficult conditions since it was commissioned. The history, operation, and cost data on this facility have recently been summarized by the Island Water Association's general manager (Watson and Derowitsch, 1979).

i) Process Description--This 2.1-mgd [7,950-m^3/d] three-stage plant has 14 parallel banks arranged in 42 stacks. It operates on a ground water which has a TDS level between 2,400 and 2,700 mg/l and a hydrogen sulfide content of about 5-6 mg/l. A flow sheet for the facility is shown in Figure 6-7. The feedwater is first aerated to oxidize the hydrogen sulfide. Further oxidation of the hydrogen sulfide to elemental sulfur takes place in the intermediate holding tank by the addition of chlorine. The water from this holding tank is pumped through pressure dechlorinators (activated carbon) and cartridge filters before entering the stacks. The dechlorinators are constructed like pressure filters and act as roughing filters before the cartridge filters. These require backwashing to keep the plugging factors low but serve to enhance the operation of the stacks. Figure 6-8 shows a typical membrane stack with its associated instruments and controls. Acid is added to the cathode stream to minimize scaling in the stacks. The pH of the product stream emerging from the stack is adjusted by the addition of a base. Originally, soda ash was used, but caustic soda was chosen as a replacement in 1980 to reduce chemical costs.

ii) Operating Information--Table 6-4 summarizes the operating data from the facility from 1974 through mid-1979. The low load factor is a function of the demand of the system rather than the facility itself. The Sanibel plant serves a seashore resort which has a high population and water demand during the winter tourist season, but which has an additional source of water by pipeline from the mainland, which is often used as a base load for the system.

Operating problems have centered around fouling of the stacks due to insufficient pretreatment, corrosion of components, and a deteriorating ground-water supply. The latter problem has necessitated the construction of new wells using improved techniques.

iii) Current Status--The plant is operating and is expected to continue operation in the future. The facility has been, and continues to be, a site used by Ionics in a cooperative effort with the Island Water Association to conduct research and development work on evaluating various pretreatments, components, operating procedures, etc.

Corfu, Greece (3.9 mgd, [14,800 m^3/d] EDR). This plant was constructed for the Municipality of Corfu and began operation in 1977. The island of Corfu is located off the west coast of Greece in the Ionian Sea. The island's water supply is based on ground water but this water has a high mineral content and hardness, making it unsatisfactory for drinking and other uses. This situation is intensified each summer when considerable numbers of tourists come to the island, approximately doubling the daily demand on the ground water just when the lowest rainfall occurs.

i) Process Description--This 3.9-mgd [14,800-m^3/d] EDR plant is divided into six parallel modules, each containing 10 membrane stacks arranged in four parallel banks. Two of the banks have two stages and the other two have three stages. Each group can be operated independently with its own flow controls, rectifiers, and instrumentation. Overall operation is performed from a central control room, which contains all of the major instrumentation and controls.

Figure 6-9 shows a flow diagram of the facility. The influent feedwater to the plant is a blend from several ground-water sources. The blended water is filtered through 10-micron cartridge filters, processed through the stacks, and then stored in a reservoir before use. The reversal process does not require the use of chemicals, such as acid or polyphosphate, in the desalination process, although some muriatic acid is used in the in-place cleaning system.

Table 6-5 shows an analysis of the four major feedwater sources. The plant is located adjacent to the Chrysiis wells, which produce the water with the highest level of dissolved salts in the area. A portion of this water is taken before blending and utilized to supply water for the concentrate stream make-up. This stream is recirculated from a concentrate tank through the stacks and returned to the tank. The concentrate stream is kept at a TDS of

Table 6-4
OPERATING DATA FOR THE SANIBEL ISLAND ELECTRODIALYSIS FACILITY[a]

	Year					
	1974	1975	1976	1977	1978	1979 (6 months)
Production						
Design Capacity (daily) (mgd)	1.2	1.8/2.1	2.1	2.1	2.1	2.1
Annual Production (MG)	276	335	392	377	293	192
Load Factor (%)	63	53	51	49	42	50
Recovery Ratio (%)	82	74	78	76	85	89
Quality						
Feedwater TDS (mg/l)	2,470	2,570	2,440	2,700	2,460	2,500
Product Water TDS (mg/l)	517	445	479	489	479	482
Consumables (per unit of product)						
Power (kWh/kgal)	9.48	10.72	9.24	9.94	8.61	8.90
Acid (lb/kgal)	1.61	1.33	1.38	1.42	1.05	0.95
Chlorine (lb/kgal)	0.17	0.20	0.12	0.12	0.14	0.12
Soda Ash (lb/kgal)	0.45	0.42	0.31	0.67	0.69	0.36
Filters (Cartridge) (No./kgal)	0.014	0.012	0.009	0.015	0.012	0.006
Costs (cents/kgal)						
Well Field and Pretreatment	12.2	13.1	13.0	21.6	19.7	21.8
Power	27.5	31.1	30.3	33.8	32.7	35.6
Chemicals	5.0	5.5	5.0	7.7	7.0	5.1
Membranes	1.0	9.2	15.3	6.7	10.7	8.1
Parts (stack)	1.8	1.4	3.6	6.0	3.5	4.1
Labor	7.6	5.2	3.9	10.8	19.1	10.8
Total	55.1	65.5	71.1	86.6	92.7	85.5

Note: Metric conversion factors: MG x 3,785 = m^3; mgd x 3,785 = m^3/d; kWh/kgal ÷ 3.785 = kWh/m^3; lb/kgal x 119.7 = g/m^3; cents/kgal ÷ 3.785 = cents/m^3.

[a] Data in this table obtained from Watson and Derowitsch, 1979.

about 4,000 mg/l by discharging a portion to sea and adding water from Chrysiis. The facility is enclosed in a building and is operated from a centralized control room (see Figure 6-10).

Table 6-5
FEEDWATER SOURCES FOR THE
CORFU ELECTRODIALYSIS PLANT[a]

Quantity	Source			
	Gardiki	Nickolas	Karterion	Chrysiis
Feedwater[b] (mgd)	1.97	0.26	0.39	3.01
$[m^3/d]$	7,460	984	1,480	11,400
Constituents				
Sodium (mg/l)	18	55	21	15
Calcium (mg/l)	188	90	180	476
Magnesium (mg/l)	51	31	58	70
Chloride (mg/l)	36	35	36	36
Bicarbonate (mg/l)	281	207	268	329
Sulfate (mg/l)	422	244	446	1,147
TDS (mg/l)	996	662	1,009	2,073

---

[a] Data from Arnold, 1979.
[b] Including concentrate stream.

ii) Operating Information--With a feedwater flow of about 5.6 mgd [21,200 m^3/d], there is a recovery of about 70 percent, with 3.9 mgd [14,800 m^3/d] of product water and a concentrate discharge of approximately 1.7 mgd [6,430 m^3/d]. Table 6-6 shows the characteristics of the feedwater, product, and concentrate streams. The salt rejection is about 64 percent.

Table 6-6
CHARACTERISTICS OF THE FEED, PRODUCT, AND
CONCENTRATE BLOWDOWN AT THE CORFU FACILITY[a]

Constituent	Feedwater (mg/l)	Product Water (mg/l)	Concentrate Blowdown (mg/l)
Sodium	19	7	44
Calcium	295	105	873
Magnesium	58	21	153
Chloride	36	13	89
Bicarbonate	294	105	757
Sulfate	699	249	2,082
TDS	1,401	500	3,998

---

[a] Data from Arnold, 1979, and Andreadis and Arnold, 1978.

Some operating problems were encountered during the startup and initial period of operation, primarily with the stacks, hardware, and external factors (Arnold, 1979). In the membrane stacks there were some initial problems with minor variations in the spacers between the membranes, causing a poor hydraulic flow distribution and hence some blockage of cells. This was alleviated by the installation of spacers of proper and more uniform thickness.

Some hardware problems were encountered with pneumatically operated valves for the reversal system. These were corrected by resizing some of the pneumatic operators and adding equipment to improve the air drying in the pneumatic system.

Outside the system reduced production and disruptions have been caused by a lack of sufficient water and frequent breakages of old lines in the collector system for the feedwater. This latter has resulted in an increase in suspended solids and iron slimes in the feedwater after the repairs were performed and the pipelines flowed again. This caused plant upsets and premature clogging of the cartridge filters.

iii) Current Status--The plant has been operating about 9 months per year to satisfy the higher demands of the tourist season. During the winter the plant is shut down, thus providing an opportunity for complete plant maintenance.

APPLICABILITY FOR DEVELOPING COUNTRIES

The theoretical principles of operation of an electrodialysis unit are fairly straightforward. Its major application in developing countries in the foreseeable future is with the desalination of brackish water. As with the reverse osmosis process, it is important to properly select and develop the raw water source and the pretreatment for the unit. Because electrodialysis is the desalination process most energy-sensitive to the level of dissolved solids in the water, a key to its overall success is being able to predict (preferably by proper testing) the long-term TDS level of the water source and hence the unit's future performance.

The operation of electrodialysis plants requires personnel able to operate and maintain pumps, motors, rectifiers, valves, and automatic activating equipment. This would include the ability to disassemble and reassemble membrane stacks and the capability to read meters and/or graphs and make relatively simple process decisions based on these readings. Although the combination of associated pumps, piping, automatic valves, etc., is a bit more complex than that of a comparable reverse osmosis unit, the electrodialysis process offers several advantages for undeveloped areas. The first is that it is a low-pressure (50-70 psi [3.4 to 4.8 atm]) system which employs standard pumps (rather than high-pressure pumps as in reverse osmosis) and the associated piping and valves can be made of material such as PVC, which is more readily purchased and repaired than stainless steel, etc.

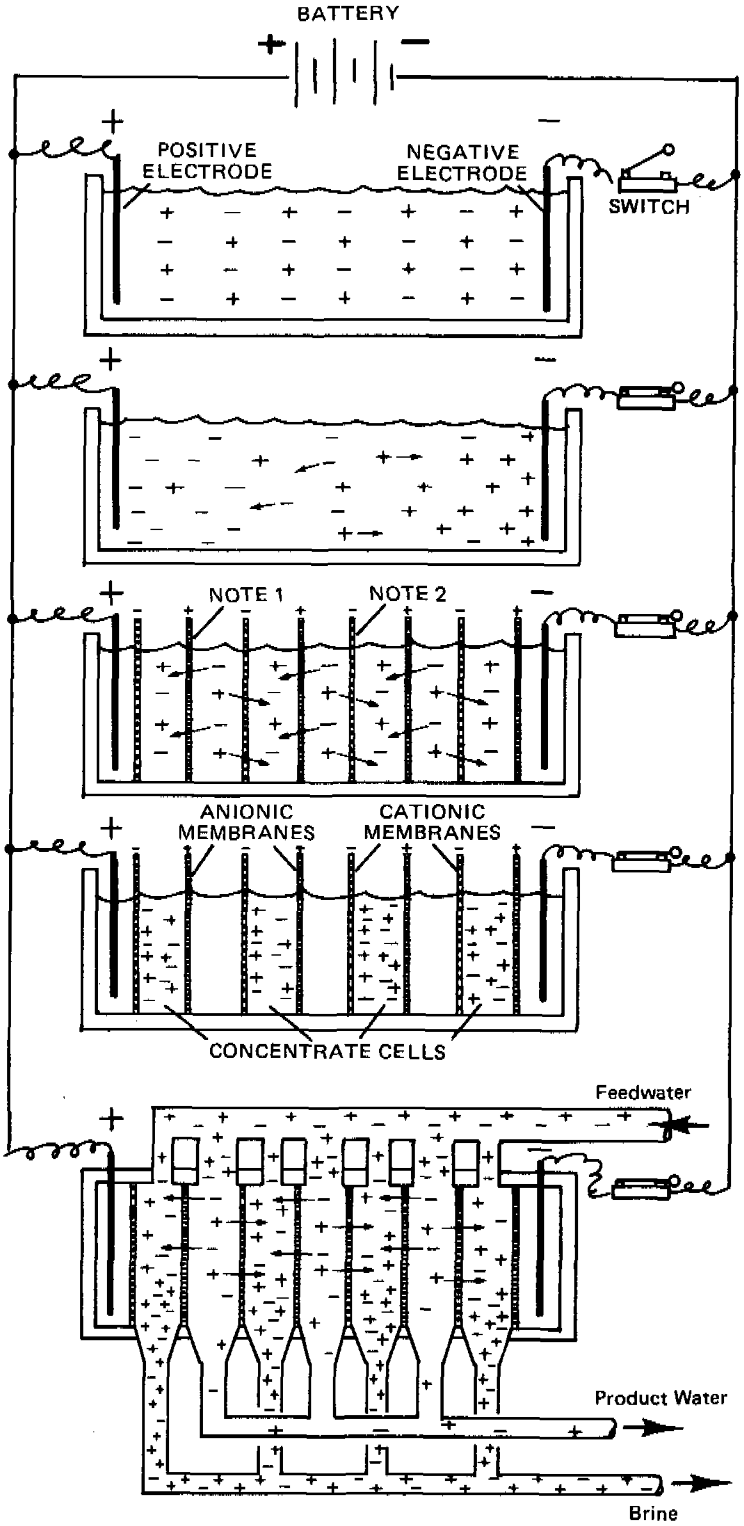
The second advantage is that in case of severe scaling or clogging problems which could not be relieved by in-place-cleaning, the stacks can be disassembled, hand-cleaned, and then reassembled. This can be a time-consuming, labor-intensive task, but in many areas labor is readily available. Furthermore, this disassembly procedure is far more satisfactory than the purchase of new units, which is a possibility with reverse osmosis units when they become heavily fouled with precipitates, sediment, etc.

The use of the electrodialysis reversal method can eliminate the use of chemicals during operation, reducing the necessity of buying, transporting, mixing, and adding chemicals on a continuing basis. This reduces the procurement and transportation problems that exist in many developing areas, although it does add some additional complexities with motor-operated valves, etc. The only consumables necessary on a day-to-day basis, therefore, are electricity and cartridge filters.

A properly designed and installed electrodialysis unit can run for hours at a time with little attention. The system can handle power losses fairly satisfactorily, but it does need trained technical supervision.

The 3,800-gpd [14-m^3/d] electrodialysis unit pictured in Figure 6-11 was built using components manufactured by the Central Salt and Marine Chemicals Research Institute in Bhavnagar, India, and was used in the village of Motagokharwala in the State of Gujarat for about 1 year. The operator was a villager trained in its operation, and supervision was provided by the Institute.

The source of water for this installation was the village well, which had a TDS level of 4,000 mg/l. The electrodialysis unit reduced this to about 1,000 mg/l. The water which was produced was well-received by the villagers, who used it for drinking and cooking purposes (Rao, 1980).



Many of the substances which make up the total dissolved solids in brackish water are strong electrolytes. When dissolved in water they ionize; that is, the compounds dissociate into ions which carry an electric charge. Typical of the ions in brackish water are Cl^{-1}, Na^{+1}, HCO_3^{-1}, Mg^{+2}, SO_4^{-2}, and Ca^{+2}. These ions tend to attract the dipolar water molecules and to be diffused in times, fairly evenly throughout a solution.

If two electrodes are placed in a solution of ions, and energized by a battery or other direct current source, the current is carried through the solution by the charged particles and the ions tend to migrate to the electrode of the opposite charge.

If alternately fixed charged membranes (which are selectively permeable to ions of the opposite charge) are placed in the path of the migrating ions, the ions will be trapped between the alternate cells formed.

Note 1: A positively fixed charge (anionic) membrane will allow negative ions to pass, but will repel positive ions.

Note 2: A negatively fixed charge (cationic) membrane will allow positive ions to pass, but will repel negative ions.

If this continued, almost all the ions would become trapped in the alternate cells (concentrate cells). The other cells, which lack ions, would have a lower level of dissolved constituents and would have a high resistance to current flow.

The phenomenon illustrated above is used in electrodialysis to remove ions from incoming saline water on a continuous basis. Feedwater enters both the concentrate and product cells. Up to about half of the ions in the product cells migrate and are trapped in the concentrate cells. Two streams emerge from the device: one of concentrated brine and the other with a much lower concentration of TDS (product water).

FIGURE 6-1. Movement of ions in the electrodialysis process.

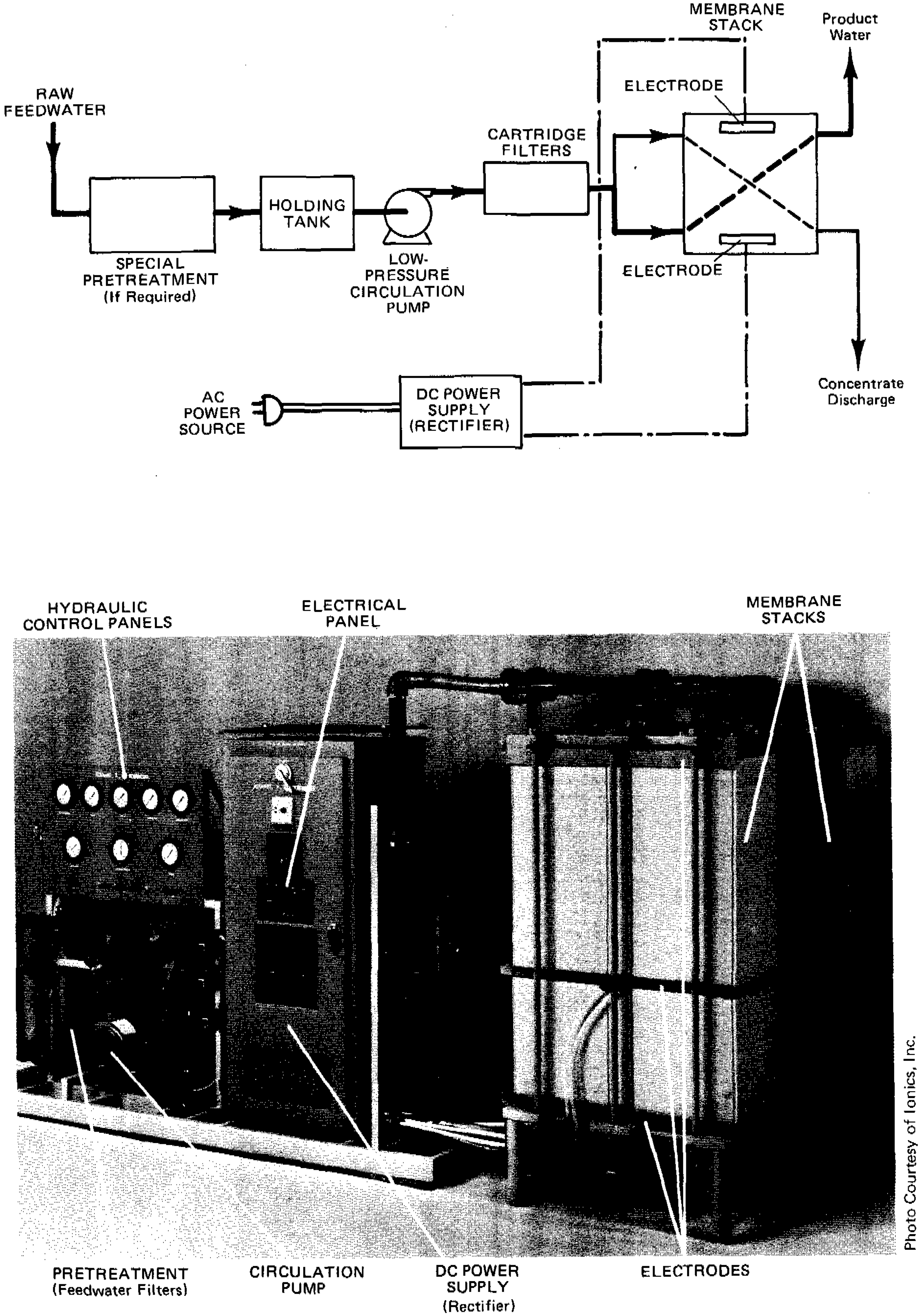


Photo Courtesy of Ionics, Inc.

FIGURE 6-2. Basic components of an electrodialysis unit.

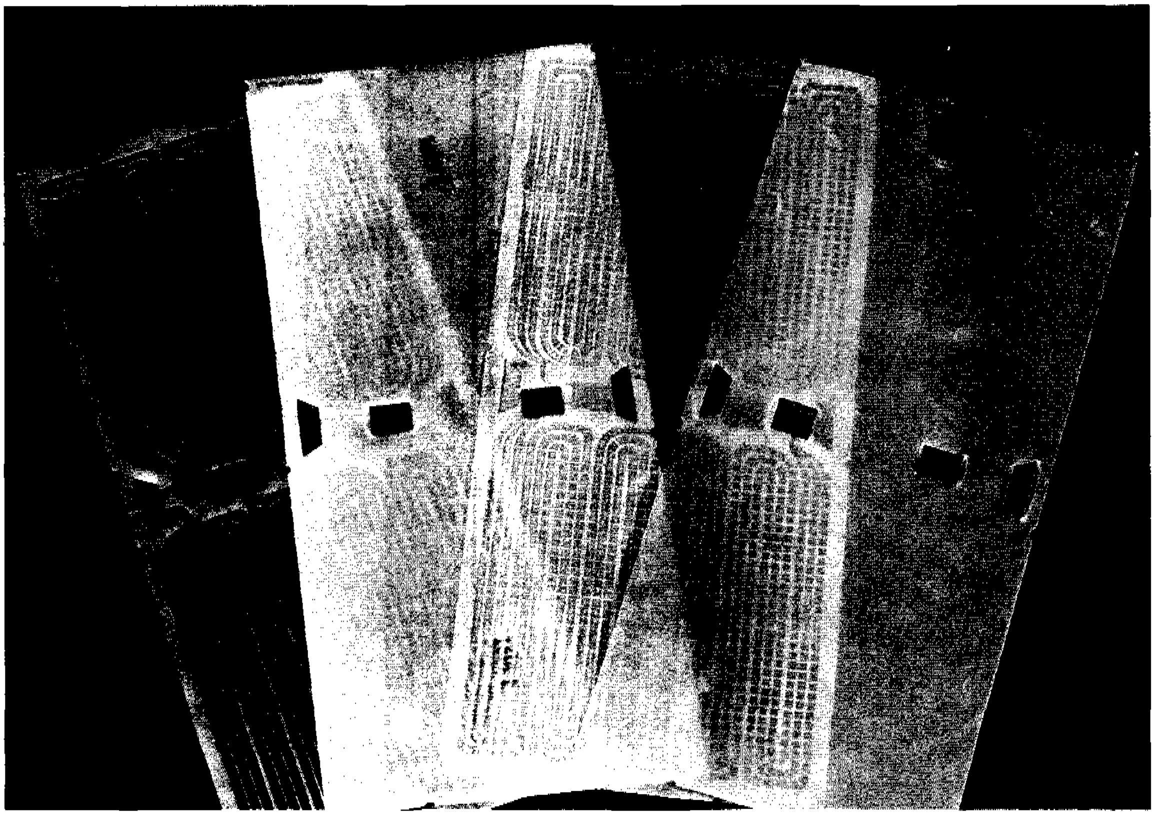
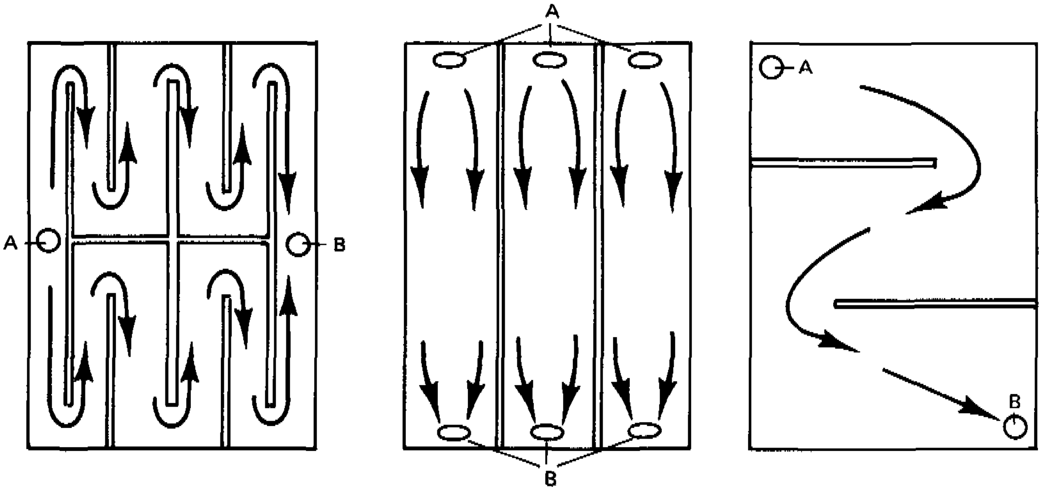


Photo Courtesy of Ionics, Inc.

FIGURE 6-3. Membranes and spacers for an electrodialysis stack.

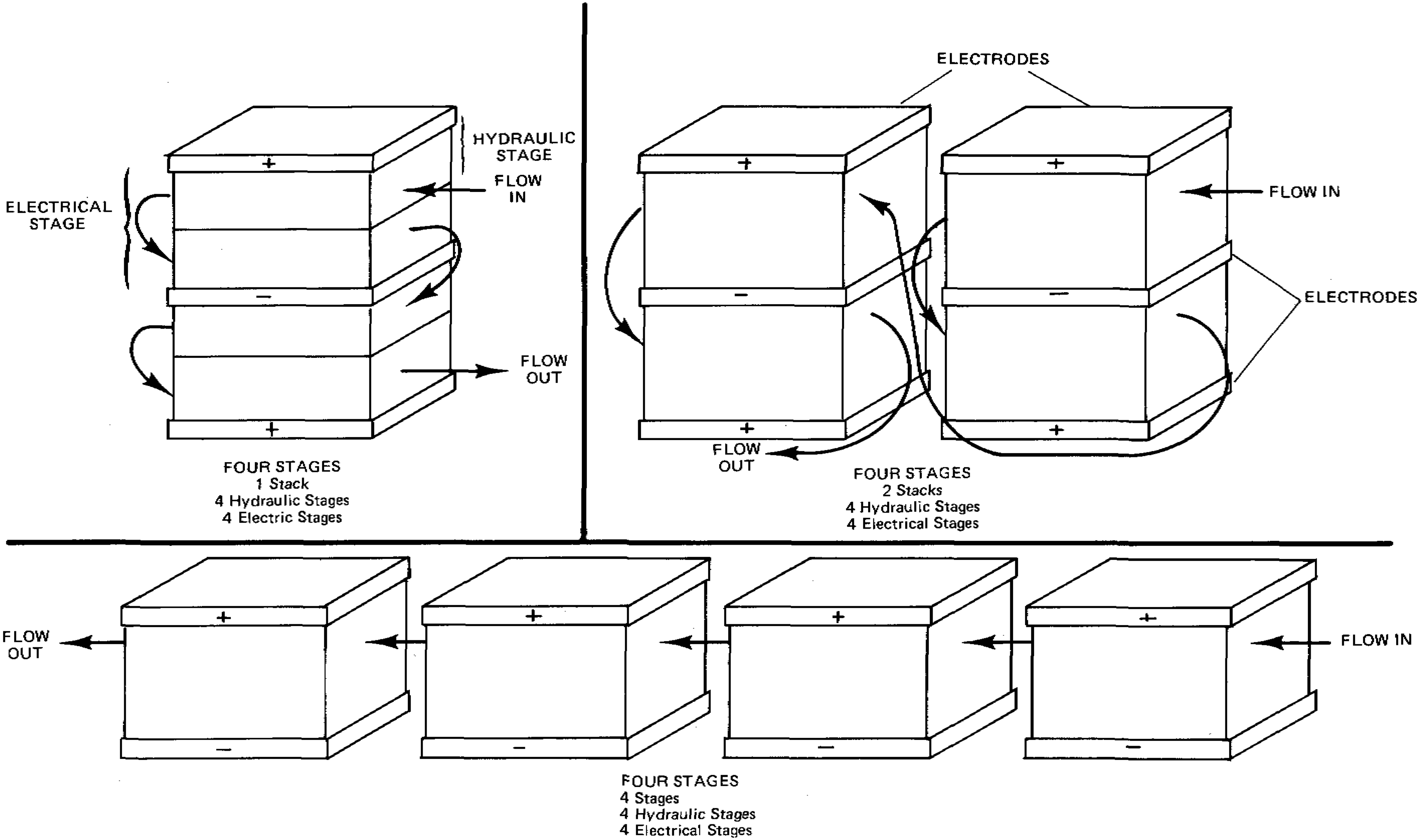


Tortuous Path

Sheet Flow

Slanted Strap

FIGURE 6-4. Three types of spacers.

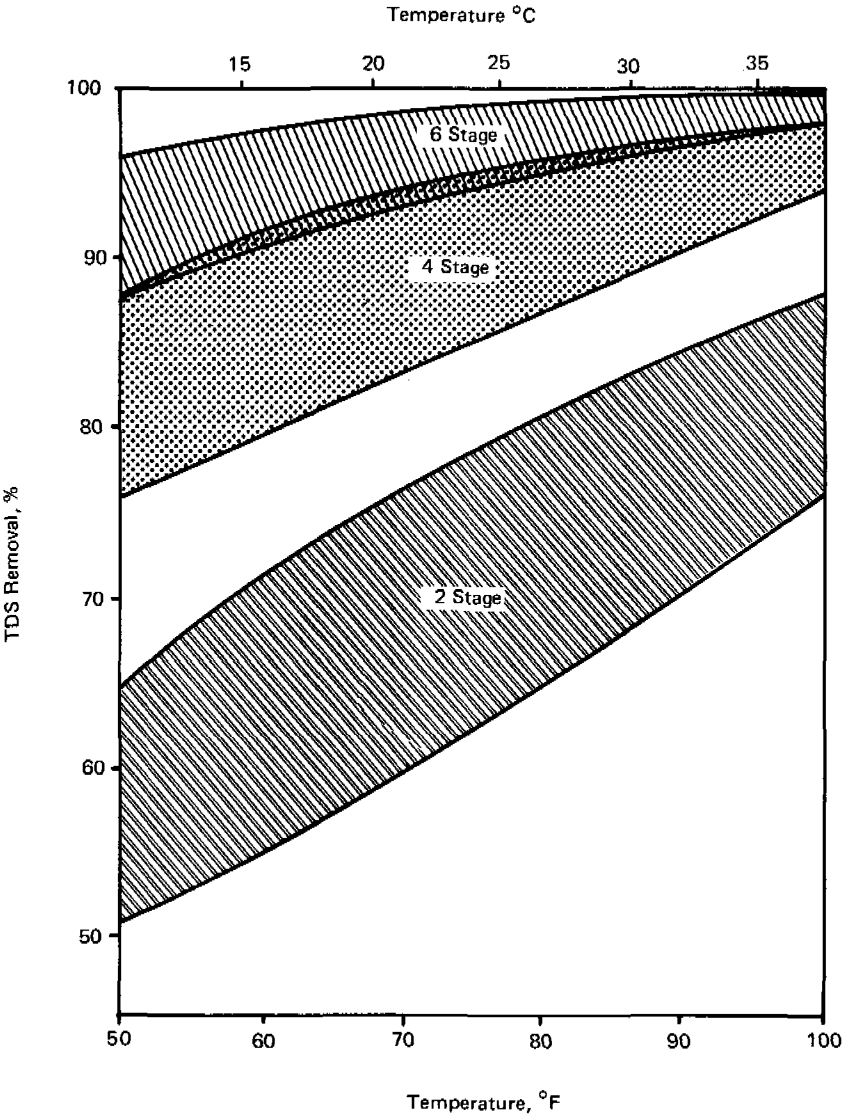


FOUR STAGES
1 Stack
4 Hydraulic Stages
4 Electric Stages

FOUR STAGES
2 Stacks
4 Hydraulic Stages
4 Electrical Stages

FOUR STAGES
4 Stages
4 Hydraulic Stages
4 Electrical Stages

FIGURE 6-5. Three variations in the flow through a four-stage electrodialysis unit.



Adapted from Katz, 1971.

FIGURE 6-6. The effect of temperature on TDS removal.

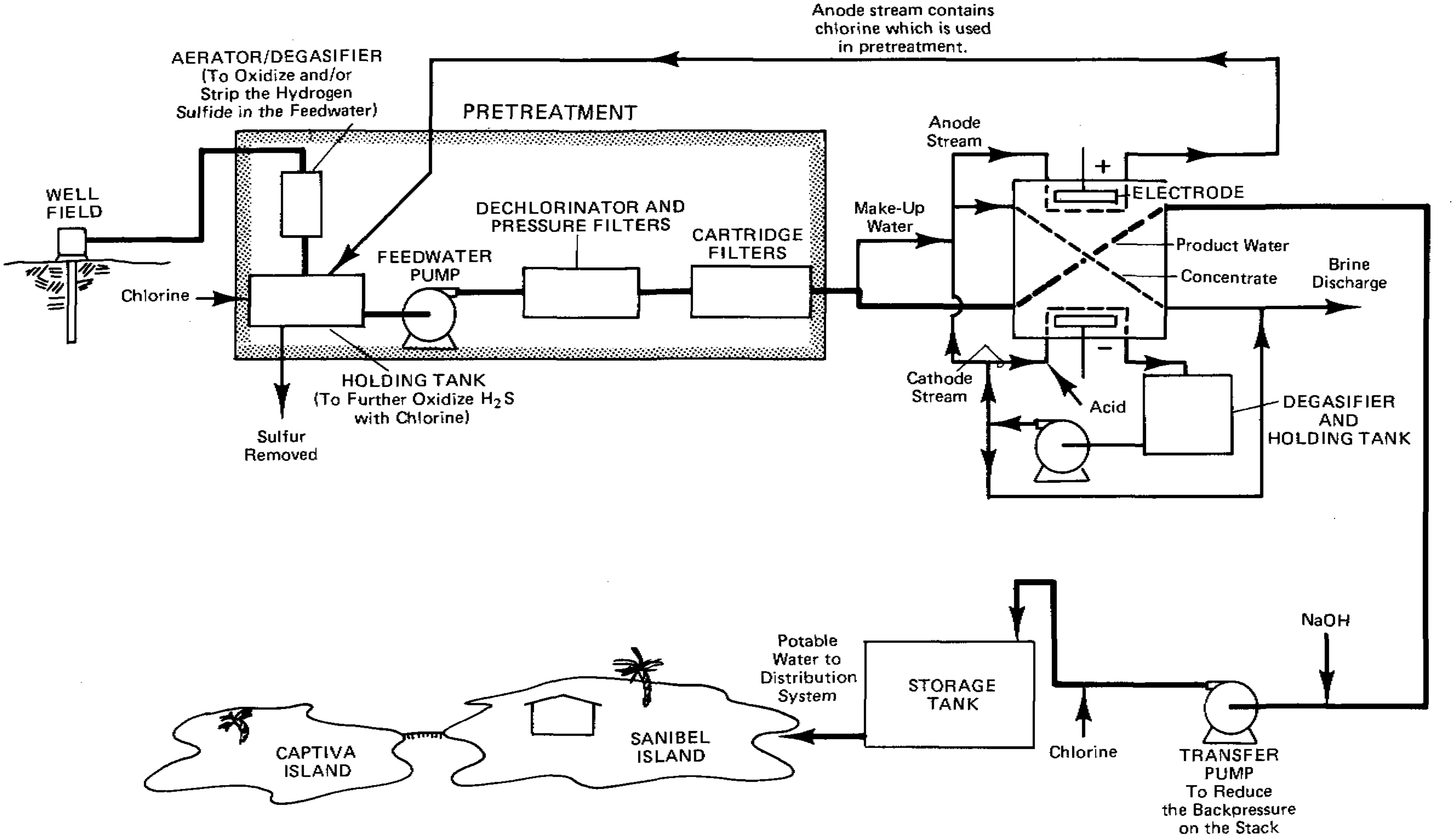


FIGURE 6-7. Flow diagram for the Sanibel Island electrodialysis facility.

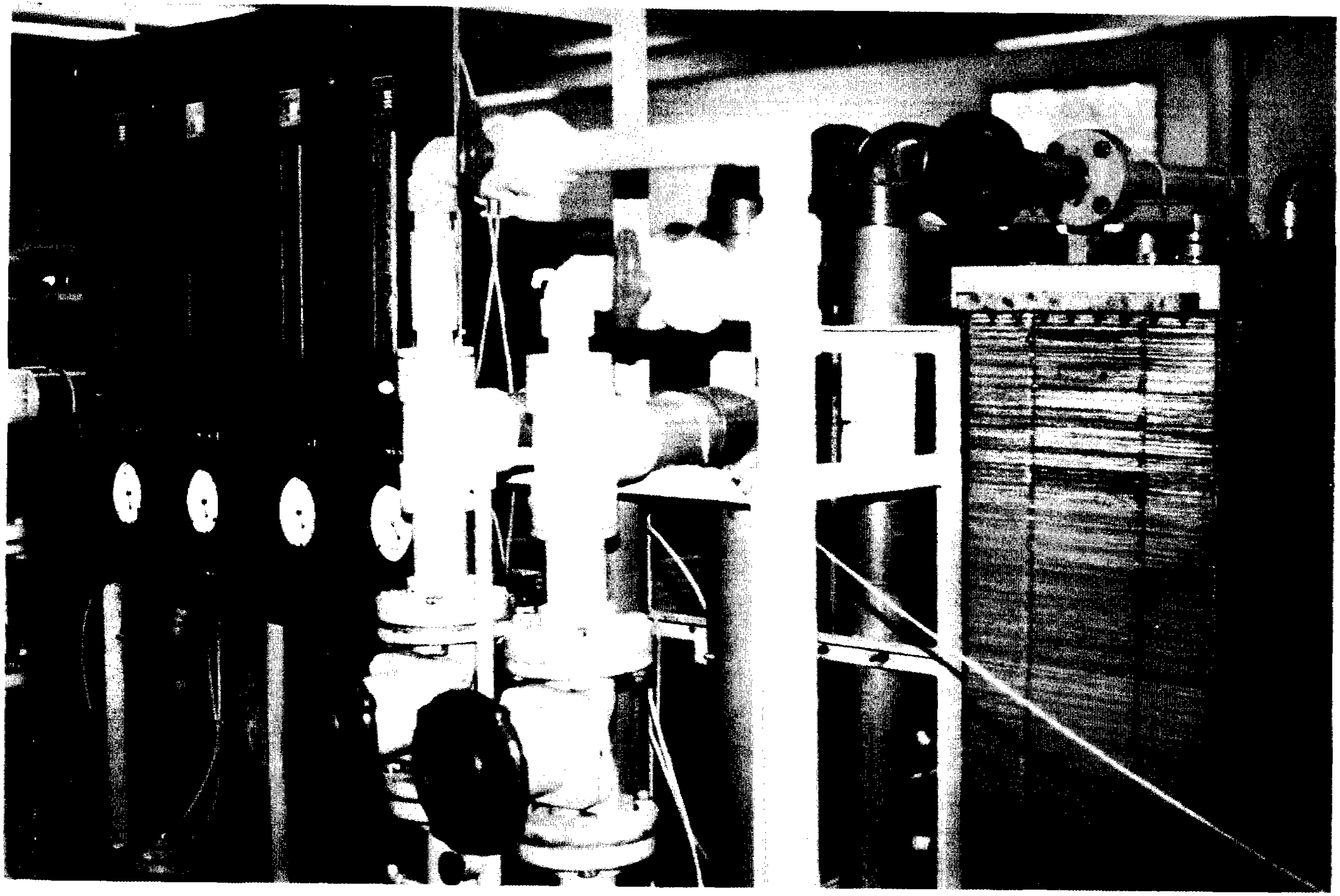
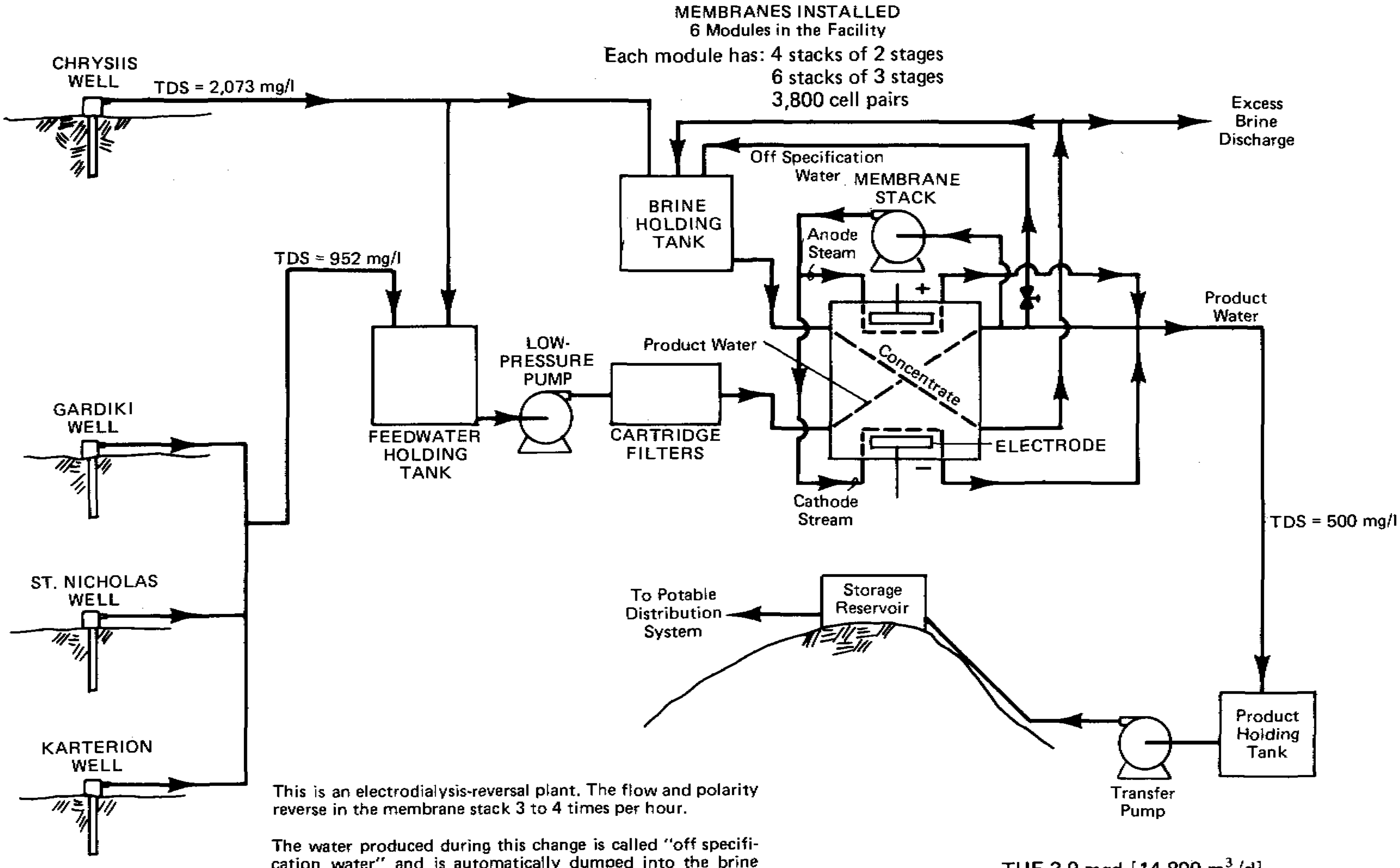


Photo Courtesy of O. K. Buros.

FIGURE 6-8. Membrane stack and instrumentation at Sanibel's electrodialysis facility.



This is an electrodialysis-reversal plant. The flow and polarity reverse in the membrane stack 3 to 4 times per hour.

The water produced during this change is called "off specification water" and is automatically dumped into the brine tank to be used in the concentrate circuit. Not all the valves and piping necessary for electrodialysis-reversal operation are shown.

THE 3.9-mgd [14,800 m^3/d]
ELECTRODIALYSIS-REVERSAL
FACILITY AT CORFU, GREECE

FIGURE 6-9. Flow diagram for the Corfu, Greece, electrodialysis-reversal facility.

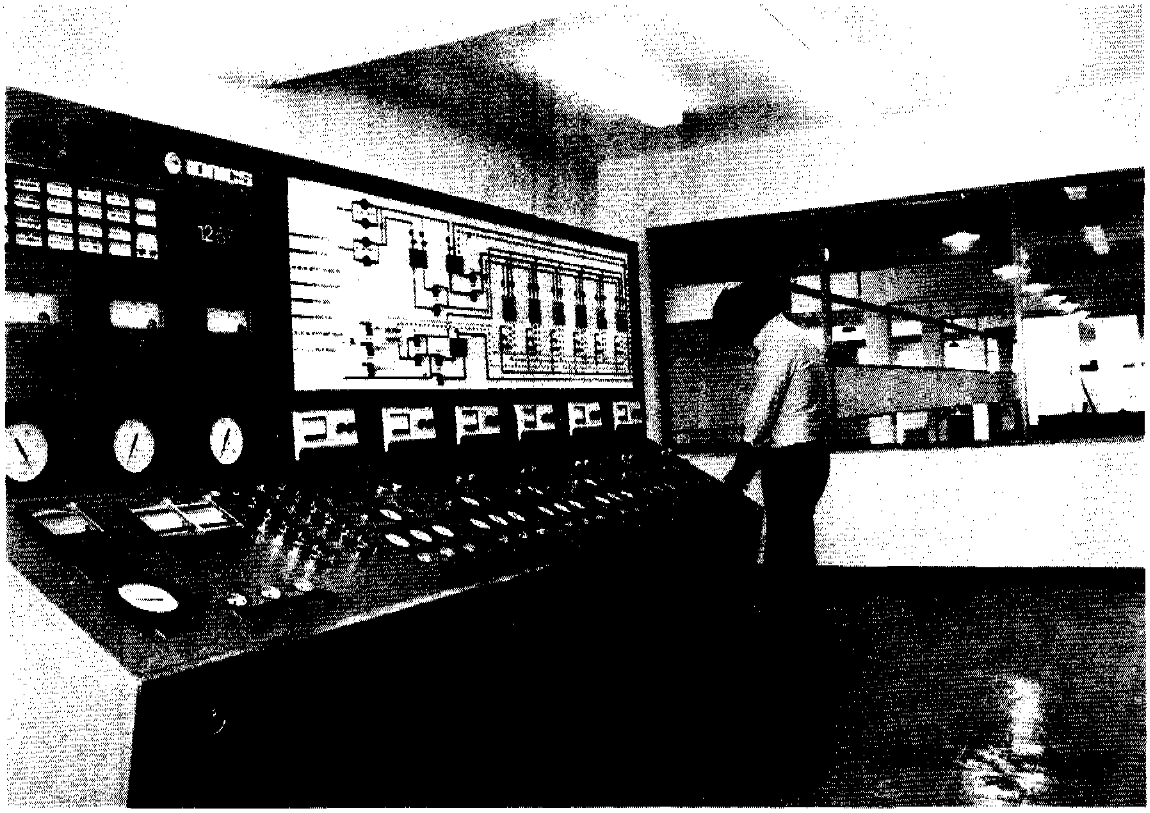


Photo Courtesy of Ionics, Inc.

FIGURE 6-10. Centralized control room for the Corfu, Greece, electrodialysis facility.

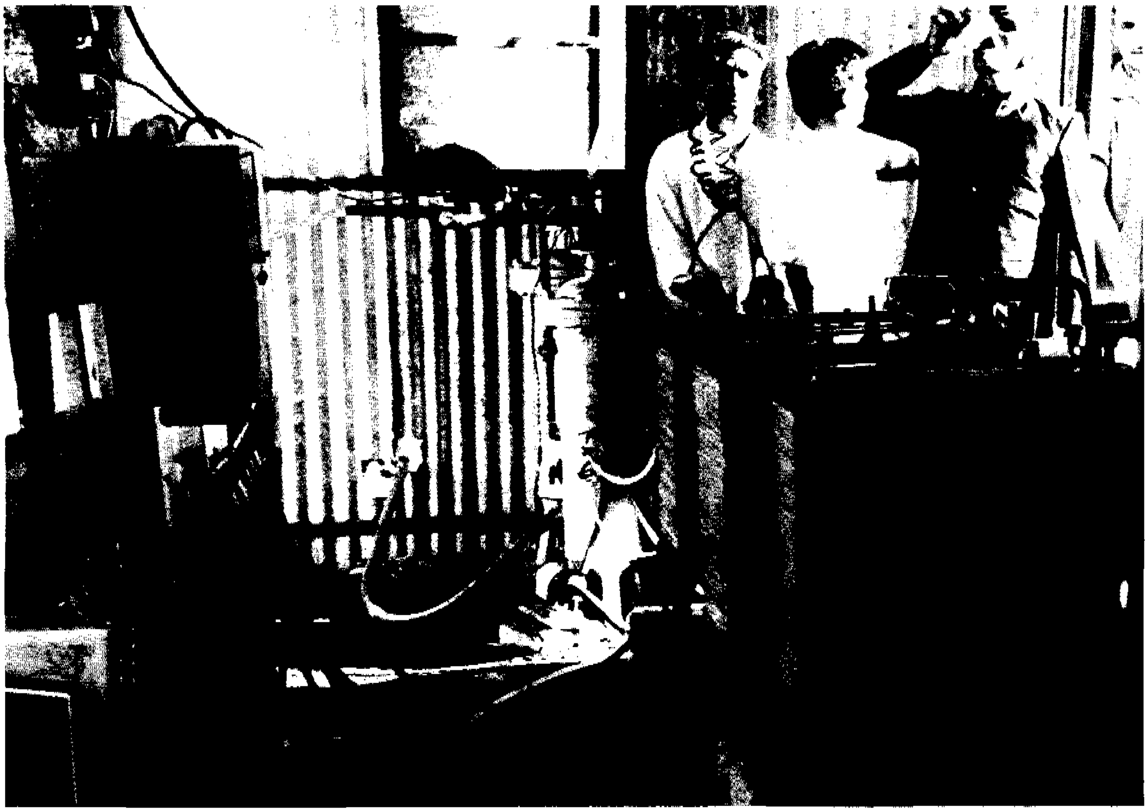


Photo Courtesy of Central Salt & Marine Chemical Research Institute, Bhavnagar, India.

FIGURE 6-11. An electrodialysis unit at the Village of Motagokharwala, India.

CHAPTER 7

Chapter 7
DESALINATION BY RENEWABLE ENERGY SOURCES

INTRODUCTION

"Renewable energy sources" as used in this chapter, refers to energy derived from sources other than fossil fuels or electricity generated from standard centralized locations. (These sources are sometimes referred to as "alternate energy sources.") This chapter discusses some of these energy sources and their application to desalination and only briefly covers many subjects which are sciences in themselves. Its purpose is only to present an overview. Specialized texts are available which examine the details and intricacies of the various subjects.

Power

The power considered in this chapter is derived from sun, wind, waves, and man's direct labor. Other varieties of renewable energy exist such as wood or biomass production, but these cover most of the major sources now being used, or investigated, for desalination purposes.

Need

There are three basic rationales for the use of renewable energy sources for desalination:

The high cost of standard fuels

The conservation of fossil fuels

The lack of power in isolated areas

The high, and continually rising, cost of conventional fuels is a dominant factor in directing attention toward the utilization of alternate energy sources. This is especially true of the renewable resources, such as solar and wind power, which require no further investment in power once the proper energy collection system is installed.

The use of alternate energy sources involves, in many ways, both a look backward and a step forward--a look backward in that it is prudent to examine some of the ideas, machinery, and practices that existed in the time before the widespread use of fossil fuels, particularly petroleum.

During the early 20th century, before rural electrification was widespread throughout North America and Europe, many reliable devices, such as windmills, were developed and sold to be used to pump water and generate small quantities of electricity.

The fact that many recent developments involve rather sophisticated devices, such as solid state circuitry and photovoltaic cells, represents a step forward.

Status

Aside from the basic solar still, the development of alternate energy sources for desalination is in its infancy. The need for these sources is recognized by many throughout the world, but their commercial development and large-scale application, especially in developing countries, can be expected to take time to implement. Most alternate energy sources take a considerable capital investment per unit of power to develop, and their interface with desalination equipment can be complicated.

During the decade of the 1980's considerable research (relative to the 1970's) into the development of alternate energy devices is expected. Currently, the alternate energy sources supply only a negligible amount of the water produced by desalting. In practical applications this is almost entirely in solar stills used in locations around the world. Although interest and experimentation have increased with regard to matching various other desalination processes, such as electrodialysis, distillation, and reverse osmosis, to alternate energy sources, this effort is really only beginning. When the match has occurred, it has generally been subsidized by some government and/or research and has seldom been commercially competitive at the same scale with desalination facilities powered by conventional means.

Efforts are taking place in Egypt, France, Germany, Italy, Jordan, Mexico, the USA, and other countries to work on matching alternate energy sources to desalination processes, but generally on a very small scale, with units of less than 5,000 gpd [19 m^3/d] due to the high capital costs involved. Although there is some independent effort (i.e., not funded by governments) by manufacturers, it is minimal compared to their efforts on conventionally powered apparatus, simply because there is almost no market for devices of this type. As (or if) a market develops, manufacturers can be expected to move into it. At this time, it does not appear profitable, and hence relatively little effort (and unsubsidized investment) has gone into alternate energy desalination.

Complexity

Although the concept of alternate energy may convey the idea of simplicity and moving "back to basics," its interface with desalination is actually not simple at all. The non-steady-state nature of many energy sources such as solar and wind power places special requirements on the desalination systems which are not present when using conventional power sources. This, in turn, adds complexity to a process which is often fairly complex already.

While the theoretical potential for renewable energy desalination can be readily demonstrated, it is extremely important that actual field experience be gained before making long-term projections and/or procuring major installations in any developing country. Small-scale plants, many of which are discussed in this chapter, are now being built for test purposes. These combine many of the energy sources and desalination processes and will aid in the developmental work necessary to produce units which will be viable in the future. For those interested in this field the alternative energy/desalination projects cited later in this chapter should be tracked as to their progress and success in later years.

SOLAR ENERGY

General

Of all the renewable energy sources available, solar energy has received the most attention as far as utilization for desalination is concerned. The earth's natural water cycle, which includes distillation, derives its energy from the sun and has gone on for millions of years. Using solar energy for desalination was mentioned as early as the 17th century by the Italian scientist Della Porta (Nebbia and Menozzi, 1966).

Energy Characteristics. The amount of solar energy which reaches the earth at any one time and place is dependent on the time of year, the time of day, and obstructions to the sun's rays.

Figure 7-1 shows a graph of the amount of energy received in a particular area during the day at different times of the year. The measure of radiation is in energy received per unit area and time. The units generally used include: Langley's/min, Btu's/ft^2/hr, cal/cm^2/min, and W/m^2 [1 Langley/min = 221 Btu/ft^2/hr = 1 cal/cm^2/min = 679 W/m^2].

The change in radiation intensity results from the change in distance (seasonally and daily) that radiation must travel through the atmosphere as well as the changing angle at which it strikes the receiver. As the angle from the vertical increases, the intensity per unit area of the earth's surface decreases. Clouds, dust, water vapor, etc., in the atmosphere tend to reflect or scatter part of the energy, reducing the radiation. The total energy received represents both direct and scattered (or diffused) radiation.

In some areas, for calculation purposes it is assumed that radiation is received about 8 hours (480 minutes) per day. Although there is a variation depending upon locality, a reasonable value for a cloudless day, radiation is received on a horizontal surface is 221 Btu's/ft^2/hr [1 cal/cm^2/min]. Charts are available which show the average radiation distribution for many areas of the world. These are often in Langleys/day or cal/m^2/day. These charts indicate that in December a horizontal area in the Sahara Desert might receive an average of 350 cal/m^2/day, whereas New York would average 125, and in June the Sahara would average 650 and New York 550 (Lof et al., 1966).

Naturally, for cloudy days the amount of energy received is severely reduced, but it does not drop to zero due to the diffused radiation that reaches the ground. Even on a clear day as much as 20 percent of the radiation is from diffused light.

As shown on Figure 7-2, as the receiver moves closer to the equator, the change during the year flattens out. This phenomenon is explained by the fact that the sun's angle to the earth does not change as much, on a seasonal basis, at locations near the equator relative to those at higher or lower latitudes.

In summary, solar energy varies during the day, from day to day, and depends on location and atmospheric conditions. Although general averages can be calculated, the exact energy available at any one place at a specific time is dependent on atmospheric conditions. Obviously, due to the diurnal nature of solar radiation, applications utilizing it must either gauge their operations to the varying energy supply or provide some type of storage to lessen the variations and deliver power for a longer period than provided by actual sunlight.

Energy Converters. A variety of mechanisms are used to convert solar energy directly to other forms of energy--either heat or electricity. Among the commonly employed devices for heat conversion which are commercially available now are flat plate collectors, focused collectors, and solar ponds. Examples of some of these are shown on Figure 7-3.

i) Flat Plate Collectors--These are made from flat metal plates which have been blackened to maximize their absorption of solar energy. The temperature of the plates rises as they are exposed to incoming solar energy. The heat generated can be transferred to another location by circulating a fluid (usually water or air) along the plates to pick up the heat energy. Usually, tubes or pipes are fastened to the plates to channel the fluid and provide good heat transfer from the plates to the fluid.

The flat plate collectors can utilize both direct and diffused sunlight. The plates are usually set at an angle relative to the horizon to maximize the amount of incoming solar energy received. The position of the collectors is usually fixed and hence is a compromise based on insolation data for the area. Flat plate collectors usually operate in the temperature range of 100 to 200°F [37.7 to 93.3°C]. For desalination applications, the heated fluid could be used to transfer energy to a heat engine, increase humidification, and/or provide heat for a low-temperature distillation unit.

ii) Focused Collectors--The focused collectors use reflective surfaces (mirrors) to concentrate solar energy on a fixed point, either by one or a whole series (sometimes hundreds) of mirrors redirecting the solar radiation. Figure 7-3B shows an installation with a single concave mirror, and Figure 7-3C illustrates a heliostat consisting of many flat mirrors focused on a stationary receiver mounted on a tower.

To maximize the utility of fixed focused collectors, especially large arrays, the mirrors are usually adjustable so as to track the movement of the sun. On a sophisticated installation this can be done automatically, as by optically following the sun or by using a small computer which is programmed with local data to track the sun's movements whether it is visible or not.

A fixed focus collector uses only direct sunlight and can operate in a wide range that includes very high temperatures--up to 6,500°F [3,600°C]. For desalination applications the range might logically include the area between 300 to 850°F [150 to 455°C]. This heat would be used to produce high-temperature liquids or vapor to operate a heat engine or to produce steam for a turbine and/or a distillation unit.

iii) Solar Ponds--There are a variety of solar ponds used to convert solar energy to usable heat. The solar ponds are usually 3 feet [1 meter] or deeper and can use salt concentration gradients to keep the heated water at the bottom. This reduces convection, conserves heat, and results in the production and storage of hot water in the ponds. The operating range of solar ponds is usually from about 80°F to 180°F [26.8 to 82.2°C], depending on the location and design of the pond. There has been some research in Israel on using these ponds. The heat output of solar ponds could be used for powering heat engines and/or use with low-temperature distillation units.

iv) Photovoltaic Cells--These devices convert light (the visible portion of the spectrum) from solar radiation directly into electricity. This conversion is based on the properties of specific crystalline materials to produce electricity when exposed to light. The conversion involves no moving parts, other energy, or special machinery. However, the manufacture of these cells is rather sophisticated and expensive. They have been used since about 1955 but mostly for space applications.

The photovoltaic cells produce direct current and can be arranged in flat panels or their unit output can be significantly increased by mounting them in a ribbon configuration so as to receive intensified radiation from a fixed focus reflector (see Figure 7-4). In the latter case the cells usually need to be cooled in order to reduce their operating temperatures. Current photovoltaic module prices range from about $7 to $12 per peak watt. The U.S. Government's Department of Energy (DOE) is working on reducing this cost to about $2.80 by 1982 and to $0.70 by 1986 (DOE, 1980).

For desalination applications photovoltaic cells could be used to produce electricity to operate motors, controls, and instrumentation in all the processes, as well as to supply direct current for the membrane stacks of electrodialysis units. This is, however, a very expensive energy source due to its high capital cost.

Energy Storage. Since solar energy is subject to fluctuations due to atmospheric conditions and its diurnal nature, some type of energy storage system is often required. Two basic systems which are often used are (1) insulated hot water (or other fluid) or rock bed storage for heat-producing systems and (2) batteries for photovoltaic systems. Other more complex storage systems involving molten salts, etc. are also available.

Applications. The following sections discuss the use of solar energy with various desalination processes. It should be clearly understood that although there is a long history of work and use of solar stills, the use of solar power in combination with other desalination processes is in the experimental stage and little or no long-term commercial use has occurred.

Solar Humidification

Solar stills operate on the principle of using solar energy to increase the relative humidity in a confined area and, in effect, distilling the feedwater without boiling. The classic still is a form of a solar basin in which solar radiation increases the temperature of the water to be desalinated. The resultant water vapor produced is allowed to contact a cooler surface, where it condenses (as freshwater) and is then collected for use.

History. Various experimenters have mentioned and tried solar distillation for at least the past four centuries (Nebbia and Menozzi, 1966). One of the first successful, well-documented stills in recent times was built in Chile's northern desert at Las Salinas in about 1872 by Carlos Wilson. It used plate glass, which was then a recent invention and is a necessity for most modern solar stills. Other stills were built at about that time and in the years that followed. They were used in various arid areas such as Chile, the Sahara, and other locales in North Africa. During World War II efforts were increased to produce a solar still that could be utilized on life rafts for ships and aircraft. Dr. Maria Telkes invented a small inflatable plastic unit for this purpose and hundreds of thousands of the units were produced. Her research efforts in this field continued after the war. Interest in this field increased so that more and more independent research and development was conducted around the world.

In the 1960's, the Office of Water Research and Technology (then known as the OSW) undertook an extensive program to investigate the potential use of solar energy for solar distillation.

A significant part of the OSW program consisted of construction and field testing at a station operated by Battelle Institute near Daytona Beach, Florida, various designs of solar stills to include glass-covered basins, inflated-plastic basins, tilted wicks and trays, and all-plastic double tubes. The OSW solar still program, conducted from 1952 to 1970, produced design data that have been used in numerous solar stills built around the world since 1960 (Eibling, 1980).

The OSW program was terminated in 1970, when it was concluded that the required reductions in capital costs could not be achieved. Based on this, it was concluded that the high fixed charges associated with the cost of still construction would not be offset by the savings resulting from free solar energy (Mattson and Lundstrom, 1979). Talbert, Eibling, and Löf summarized and reviewed efforts in the field of solar distillation in the Office of Saline Water's publication, Manual on Solar Distillation of Saline Water (1970). This extremely valuable reference text contains over 500 references and specifications, diagrams and/or photos of over 100 different solar stills. Although some of the cost information is outdated, the remainder of the information is useful and should be examined by any agency, group, government, and/or individual contemplating work on solar distillation. Another helpful resource on solar distillation is a paper by Lawand (1968), which contains an earlier, detailed literature review.

Design Considerations for Solar Stills. There are several factors which affect the performance of a solar stills (Rajvanshi, 1979). Among these factors are: solar radiation, depth of brine in the basin, cover material and its shape, ambient temperature, wind velocity, and temperature of the condensing surface.

Some of these are related to local conditions, while others are dependent on the physical design of the still. Although most solar stills are similar to the basic one shown on Figure 7-5, it seems that the subject has captured many a scientist's fancy and imagination. The result has been the production of a wide variety of designs and pilot plants (often just one-of-a-kind) which have been able to desalinate seawater or brackish water.

Solar stills are classified as either passive or non-passive. The passive still requires no outside energy except the initial conveyance of the feedwater to the still. Most operating stills are of this type and are very simple in nature.

Passive Still. A typical passive still is illustrated on Figure 7-5. This example has almost all of the basic components of a passive still. Other designs frequently utilized are shown on Figure 7-6. Most stills are fashioned in continuous bays so as to achieve adequate production with some economies of scale, but individual (or modular) units are also manufactured.

A careful and very practical review and analysis of design and operating considerations for family-size stills appears in another U.S. A.I.D. publication, entitled Fresh Water From the Sun (Dunham, 1978), which should be studied before any serious consideration of solar stills. A useful checklist on the elements of still design and construction was compiled by Lawand (1975) and is available from Brace Research Institute (Quebec, Canada), which has had a long history of work with solar stills. Other useful material on solar stills have appeared in the Journal, Solar Energy (Eibling et al., 1969 and Bloemer et al., 1965).

Non-Passive Still. The non-passive still requires some additional power for operation. Some features which might require this additional power are those designed to: (1) repump the water after the initial feed, as in a falling film multiple-effect still; (2) force air into the still to keep a non-framed plastic roof inflated; or (3) provide mechanical circulation of vapor through the still. Some of the features can increase still efficiency (and capital costs) but they also can increase the complexity of operation and maintenance and require an outside power source. In many cases the value of the additional water production is not commensurate with the extra complexity and capital cost. A number of these devices, as well as variations of the passive still, have been investigated by University of California researchers, who reviewed many of these efforts in a paper in 1974 by Howe and Tleimat.

Further discussions on solar distillation in this chapter are limited to passive stills.

Materials of Construction. A wide variety of materials can be used in building the basin, walls, frame, and condensate collector, and these portions lend themselves to construction with locally available materials.

The key material is the transparent coverings for the solar stills, which have been made from glass and plastics (both films and rigid or semirigid sheets). Glass has proved to be an excellent material but is often very costly and subject to breakage during transport, installation, and use.

Compared to the other coverings, films are lighter, usually cheaper, and easier to transport but appear to be subject to greater wind and rain damage during use. Dunham (1978) comments that "In spite of the considerable research and experimentation that has gone on with plastic films, no major still on which they are used as a cover is in operation today. Although research may continue to improve the suitability of plastic film, at present it does not seem appropriate to consider it for permanent applications in developing countries."

Several types of rigid or semirigid plastic sheets have been used as covers for solar stills, primarily on an experimental basis. Depending on their chemical composition and method of manufacture, their characteristics (strength, wettability, thermal conductivity, aging) for still applications vary. The advantages of plastic sheets include their light weight, ability to be formed into self-supporting shapes, and (compared to glass) their higher impact strength. Disadvantages include their low thermal conductivity, high thermal rate of expansion, high transmittance of infrared radiation, soft surface (scratchability), and high cost or absence in developing countries (Dunham, 1978). The use of plastic sheets may indeed be an attractive prospect for certain locations, but it should be approached with caution and field testing.

Performance. As a rule of thumb, a well-designed solar still located in an area of good insolation produces daily about 0.08 to 0.10 gallon per square foot [3.3 to 4.1 l/m^2] of basin area. This varies with still design, season, location, weather, etc., but is a good average rate. Stills may produce less, but claims that a still will produce significantly more than this should be regarded with caution. Such production is possible, but it warrants careful investigation and verification--preferably field verification.

Tleimat (1978) indicated that the following features seem to lead to high efficiency in stills: low heat capacity, low air content, vaportight cover, watertight basin, and good insulation around the basin.

Operational Problems. Aside from cultural and sociological problems associated with operating solar stills, discussed in general in Chapters 8 and 9, a number of technical problems can cause operational difficulties. These difficulties usually result in reduced efficiency rather than outright failure but, since the product of a solar still is already rather small, any reduction can be a serious matter.

Given a good basic design, there appear to be two major causes of problems: water vapor loss and reduced heat absorption capacity.

i) Vapor Leakage--In order to operate efficiently, passive solar stills must be reasonably airtight so as not to lose water vapor before it condenses. Any leaks in the unit can allow vapor to escape and production to drop. A major loss is created when the transparent cover is ripped, broken, or improperly sealed. Although adverse weather such as wind-, hail-, and sandstorms can create breakage, a great deal of breakage can be the result of human or animal intervention. Scientists from India's Central Salt and Marine Chemical Research Institute (CSMCRI) have built and operated stills in various village settings. Summing up their experience with a still in the village of Awania, they commented that "Glass breakages have been caused due to a number of factors like slippage of glass sheets at the lower edge, uneven thermal expansion and wind pressure, stone throwing by miscreants from villages, dogs making their way through fencing and by peacocks. The stone throwing, dogs, and peacocks accounted for more than 90 percent of the breakages. The glass slippage from the lower edge has been cared for and the instances of stone throwing have been appreciably reduced with the passage of time" (Natu et al., 1979a).

It is Daniel Dunham's view that any new solar still installation in many villages should be furnished (and included in the economic analysis) with one complete set of glass panes for replacement purposes. This is not only to recover from natural disasters but to utilize while the villagers determine the relationship between themselves (and their animals) with glass breakage and the production of water from the still.

Another serious cause of vapor leakage is improper sealing of the joints between the glass and the frames. Since glass expands and contracts with alternate exposure to heat and cold, it must be sealed with a material that will flex and still remain sealed. Unfortunately, the perfect sealing material, which will hold up forever under conditions of direct sunlight, heat, etc., has not been found. Those materials which do approach perfection (such as silicone sealing agents) have high first costs.

ii) Reduced Heat Absorption Capacity--The second major operational problem is a reduction in the efficiency of heat absorption by the water to be desalinated in the basin. It is usual practice to coat, paint, or otherwise color the bottom and sides of the basin black to increase heat absorption. However, if the water in the basin is allowed to dry out or overly concentrate, there is a tendency for the precipitated minerals to form a white coating on the basin's surface. This white coating can severely reduce the absorption of heat.

Drying out the basin can occur, all or in part, as most stills are designed for shallow basin operation. In shallow basin operation the depth of water is kept in the range of about 0.5 inch to 4 inches [1.3 to 12.7 cm] to maximize the heat transfer and temperature of the water. When operating with a shallow water depth "dry-outs" can easily occur if feedwater is not added regularly. Those stills which use dark rocks, wicks, etc., to promote rapid heat transfer in the basin essentially act as a very shallow basin. They also can develop problems with salt deposits, etc.

Stills with deeper basins of 2 inches [5.1 cm] or more water depth tend to avoid this problem of dry-outs, and the necessity to have flat and precisely level basins is minimized. However, the heat transfer, and hence efficiency, is lower. One method which has been suggested is to use a deep basin design but add a dark water-soluble dye to the saline water to be distilled. The dye enhances direct heat absorption by the upper layers of the water and results in an efficiency close to that of a shallow basin. Some recent research on proper dyes, dosages, etc., for solar stills has been done in the USA by Rajvanshi (1979). One problem with dyes is that they disintegrate under the effect of ultraviolet radiation, after a period of time, depending on the dye. Long-term actual field experience (published) using dyes is lacking.

A related problem is that of the growth of algae and other microflora on the surface of the brine and/or the basin. These interfere and generally reduce heat transfer to the brine. It is important that the basins be readily accessible to clean out these growths and wash off and/or repaint areas on which salt deposits have formed. Figure 7-7 shows a small solar still with an access hatch on one of the walls for that purpose. In some locations, the door and space in the still might be made large enough to permit a person (perhaps a small boy) to enter and clean. In other units some of the glass panels could be made readily removable to permit access for maintenance and cleaning.

Experience. Although there has been extensive discussion about solar stills, a number of pilot units and a scattering of installations around the world, there are not many documented cases of long-term operation. The often-referenced 48,000-ft^2 [4,460-m^2] basin type still constructed by Charles Wilson in Las Salinas, Chile, appears to have one of the longest operational histories--from about 1872 to 1912. Few, if any, large (or small) stills have equalled that longevity.

Improper selection, inconvenience, poor design, and/or availability of other water sources (sometimes water desalinated using fossil fuels or electricity) has usually resulted in the abandonment and/or destruction of many stills in the past. The Las Salinas still was abandoned when a pipeline was installed to bring in freshwater. A recent publication reported that the stills installed by CSIRO (Australian Commonwealth Scientific and Industrial Research Organization) in Australia in Darwin, Townsville, Coober Pedy, Muresk, and Griffith (totalling about 13,000 ft^2 [1,210 m^2]) have been demolished and replaced by conventional desalination facilities (McCarthy and Leigh, 1979). In Mexico two out of three of the solar still installations made by the DIGAASES (Mexico's Direccion General de Aprovechamiento de Aquas Salinas y Energia Solar) are not operating due to the availability of water from other sources. The 5,000-gpd [19 m^3/d] solar still on the Aegean Island of Symi (Greece) was dismantled and replaced with a small vapor compression unit. These are examples of some (and certainly only a portion) of the stills that were installed with high hopes but did not live up to expectations.

However, all of the experience is not bleak. Reasonable chances for success still exist in places where (1) the unit is suitable on a water resource basis; (2) there is a good organizational structure (social and political) for operating and repair; (3) there exists a local desire for the unit to work; and (4) the unit was well-designed, using appropriate technology. Some examples of working stills are detailed in the following sections.

i) Awania, India (1,300-gpd [4.9-m^3/d] Solar Still)--In 1978 the Central Salt and Marine Chemicals Research Institute (CSMCRI) of Bhavnagar, India, installed a 20,000-ft^2 [1,860-m^2] still at the village of Awania in the State of Gujarat. This village of about 1,400 people, which has no electricity, has brackish ground water as a source of potable water.

This is a bay type still, in which the brine pool is continuous beneath the glass covers. The facility has an average production of about 1,300 gpd [4.9 m^3/d] and is arranged as shown on Figure 7-8 and pictured on Figure 7-9. The stills are built in 15 blocks of 6 bays to each block. Each bay is about 40 feet [12.2 m] long and 5.6 feet [1.7 m] wide, and has its own brine pool. Several bays are seen in the photo on Figure 7-9, and a cross section of the bays is shown on Figure 7-10. The plant is operated on a batch basis. The bays are filled with brackish ground water using a diesel engine for power. When the level of water drops below a certain level in the pools more water is added.

Product water is distributed at the taps installed on the product water storage tank (see Figure 7-11). The monthly production at the Awania facility is shown in Table 7-1.

Table 7-1
AVERAGE DAILY PRODUCTION,
FROM THE AWANIA SOLAR STILL

	1978		1979	
Month	Gallons	Liters	Gallons	Liters
January	--	--	918	3,490
February	--	--	987	3,750
March	1,625	6,175	1,121	4,260
April	1,714	6,515	1,461	5,550
May	1,779	6,760	1,737	6,600
June	1,295	4,920		
July	758	2,880		
August	803	3,050		
September	803	3,050		
October	1,032	3,920		
November	1,218	4,630		
December	1,026	3,900		

Source: Data from Natu et al., 1979b.

CSMCRI reports that the plant operates quite well and that the villagers are able to operate and perform maintenance on the facility, although they do not readily volunteer to do so. Some problems were experienced in vapor leakage due to the sealant used and in lost efficiency due to algae build-up in the brine pools in the bays. There was a glass breakage problem caused by straying peacocks and other animals as well as stones thrown by some villagers. The latter was the greatest problem but has diminished (Natu et al., 1979b).

This still's design is based on pilot plant work on solar stills conducted by CSMCRI since about 1964. A similar plant, also designed and constructed by CSMCRI, with a capacity of about 35 gpd [130 l/d] per day, has been operating at the Navinar Lighthouse since 1968.

ii) La Gonâve (250-gpd [0.95-m^3/d] Solar Still)--The small village of Source Philippe is located on the Haitian island of La Gonâve off the west coast of Haiti. The area is semi-arid, with only 12 to 18 inches [30.5 to 46 cm] of rainfall per year. The tropical climate brings intense sunlight and constant trade winds, producing a high evapotranspiration rate, which further hinders the establishment of an adequate natural water supply. The villagers fish and raise cattle and depend for water on a well, which produced water (often polluted by the cattle) with a TDS content of 6,500 to 8,500 mg/l. Occasionally, this supply was supplemented by barrels of water brought in from the Haitian mainland at a cost, in the 1960's, of about $10/kgal [$2.64/m^3] (Lawand, 1970).

In 1969 a solar still was installed with the cooperation and assistance of the L'Eglise Méthodiste, OXFAM, the Brace Research Institute, and the Haitian Government. The design and construction was under the direction of the Brace Research Institute. They used local labor and materials wherever possible for completion of the still.

Figure 7-12 presents a picture of the still and Figure 7-13 shows the layout and cross sections. The still is located on the seashore, and brackish water with a TDS of about 8,000 mg/l is used as feedwater. Each morning the feedwater is hand-pumped to a 2,000-gallon [7.6-m^3] raw water tank. From there it flows by gravity into the bays. The still is operated on a batch (versus continuous feed) basis. The condensate flows by gravity to a product water tank which is divided into two parts, one holding the distilled water and the other storing rainwater collected from the roof of the still. When flushing of the bays is necessary, the brine is returned to the sea.

Since the solar still was installed in 1969, the water supply of the area has been supplemented by the construction of a hillside catchment and reservoir. In addition, some mineralized water is obtained from a nearby cave.

In the late 1970's a hydroponic garden was built adjacent to the still, and a portion of the water was used for the garden.

Both the rugged construction and practical design of the still have been demonstrated. In late 1977 the still, although needing some repairs and functioning at a reduced capacity, was reported to be in operation. In 1978 the installation was revitalized by the L´Eglise Méthodiste d'Haiti through consultation with the Brace Research Institute. This work consisted primarily of replacing butyl rubber lining to correct an original design fault. At the same time, the glazing was cleaned, replaced, or recaulked where necessary. Upon completion of the repairs, the average daily production of the still was about 250 to 300 gpd [0.95 to 1.1 m^3/d]. The unit serves a varying population which can amount to 250 to 1,000 persons.

iii) Puerto Chale, Mexico (260-gpd [1-m^3/d] Solar Still)--In order to provide water to villages or groups of families in the very arid areas of Baja California, the Mexican Government has used a variety of desalination processes, including solar distillation. The installation at Puerto Chale was completed in about 1973 and was still operating in 1980. This facility serves about 100 inhabitants living along the coast, many of whom work as fishermen. The Mexican installations use many small modular units connected together, whereas the other two examples of solar stills in this chapter feature units constructed in large bays. Figure 7-14 depicts these small unit stills. Each has an area of about 107 ft^2 [1 m^2]. The units are not normally mounted in the air, as shown on Figure 7-14, but are usually buried in the ground up to the top of their respective basins. The buried type of installation was used at Puerto Chale, as shown on Figure 7-15.

These units are used by DIGAASES (Direccion General de Aprovechamiento de Aguas Salinas y Energia Solar), which is the Mexican Government's agency for implementing the country's desalination program. Their solar distillation program is based on using this modular approach to all their installations. The basic still is manufactured from fiberglass which is lightweight, watertight, and corrosion-resistant. The units are designed for ease of rapid installation in isolated areas where no materials exist. To do this, the stills are constructed and shipped as four major components:

The one-piece fiberglass basin, including end walls.

The ridge pole made from aluminum.

The glass panes (usually 1/8-inch [3-mm] thick).

The miscellaneous piping.

The formed basins are designed so that when the ridge pole and glass are removed, the fiberglass basins will nest when stacked on top of each other. This allows a great many basins to be packed into a relatively small volume, greatly decreasing transportation costs and problems. The remaining glass, ridge pole, and piping are easily packed and shipped, to be assembled at the site.

The Puerto Chale facility uses a small gasoline-powered pump to deliver seawater, used as feedwater, to the stills. The stills are operated on a batch basis. A wall keeps out animals to help minimize breakage. The facility is visited at regular intervals by a staff member of DIGAASES for repairs, adjustments, etc.

Reverse Osmosis

General. The combination of reverse osmosis with solar power is a fairly new application. Energy must be furnished either as electricity or mechanical shaft power.

Where electricity is produced, it can be used to power a standard electric motor to drive the high pressure pump. Where rotating power is available, it can be used to drive the pumps directly; however, even some electrical energy is usually needed to operate controls, other pumps, instruments, etc.

Three ways have been proposed or used for combining reverse osmosis with solar power:

Photovoltaics--to produce electricity.

Flat plate collectors, solar ponds, focused collectors--to produce heated fluid to drive a heat engine.

Focused collector--to produce steam to drive a turbine.

In the latter case, the turbine could either drive the high-pressure pumps directly or produce electricity by driving a generator. When a heat engine has been used, it generally drives the pumps directly.

Due to the high pressures involved and the potential for membrane fouling by precipitation, etc., reverse osmosis installations do not perform well over the long-term if they are continually subjected to start and stop operation, unless they are specially constructed to do so. Thus, a workable system should have provisions to handle interruptions in energy generation. Whether to operate continuously or only during daylight hours is a decision that must be made based upon economics, site requirements, etc. In most units built to date, it has been easier (or more economical) to supply storage for heat energy than for electricity (batteries).

Howe (1980), who has worked with solar desalination for many years, comments that "The use of reverse osmosis with solar generated power suffers from the problem of having only a few hours of operation during the day. The power units are likely to operate only about 6 hours per day, thereby necessitating the start-up and shut-down of the RO unit after a very short run. Since the start-up and shut-down parts of the operating cycle are non-productive and require close and expert attention, the operational problems are severe. Also, since only 6 hours per day are available, the rating of the plant must be nearly four times that of a continuously operating plant. This means, in turn, that the capital expenditure will be very high per unit of water produced. If grid power is available, the cost of electricity to keep the plant running continuously should be balanced against the added capital cost to use only solar energy, with its short daily period of operation. In this case, the solar energy becomes an augmentation to the grid power and could be justified if it proves to be no more expensive than the normal grid power production units. These same comments would apply to the use of distillation and to electrodialysis, so that the problem reduces to the consideration of solar power rather than to desalination."

Experience. A number of experimental units have been built to date, two of which are discussed below:

i) Cadarache, France (11-gpm [60-m^3/d] Solar RO)--This is a prototype unit being operated at the French Atomic Energy (CEA) Center for Nuclear Studies at Cadarache.

Efforts were made to simplify the pretreatment system so that the complex chemical addition pumps, mixing tanks, etc., could be eliminated. Experiments were conducted using units which employed settling tanks, cartridge filters, sand, and/or activated carbon, electrochlorination, etc., in an effort to work out a suitable pretreatment system (Maurel, 1979a).

To reduce the energy required, several efficient pumps were tested. These included single cylinder types with pulse damping and a 3-cylinder 120° offset model. In addition, an energy recovery system using a Pelton turbine on the brine stream was incorporated in the design. The turbine was connected by a belt drive to the shaft of the high pressure pump to reduce the power required for operation.

The energy to operate the unit comes from flat plate collectors. There is about 2,400 ft^2 [223 m^2] of collector area in south-facing collectors set at 45° to the horizon. These heat water which either goes to storage or is used to power a heat engine using Freon as the expanding fluid. The engine's condenser is cooled by saline well water pumped from the same source as the feedwater.

Figure 7-16 is a flow diagram of the unit. The reverse osmosis unit, heat engine, and some of the flat plate collectors are shown in the photo on Figure 7-17 (Maurel, 1979a).

The unit has been operating on an experimental basis from 1978 to 1980 using brackish water with a TDS of about 2,000 mg/l. A larger unit is being constructed in 1980 by the CEA which will have a capacity of about 40 gpm [218 m^3/d]. It will be used by the Egyptian Government for desalination near the town of El Hamrawin on the coast of the Red Sea. This unit will use a collector area of about 4,100 ft^2 [381 m^2] (Maurel, 1979b).

ii) Concepcîon del Oro, Mexico (0.8-gpm [4.4-m^3/d] Solar RO)--In 1980 a reverse osmosis facility was installed in the mountain village of Concepcion del Oro in the State of Zacatecas in Mexico, as a joint research effort of the Mexican Government's desalination agency, DIGAASES (Dirección General de Aprovechamiento de Agua Salinas y Energía Solar), and two Federal Republic of Germany organizations, AEG-TELEFUNKEN and GKSS.

The facility is powered by electricity generated by photovoltaic cells and is designed to desalinate water 8 hours per day during the daylight hours using both the photovoltaic cells and batteries. Figure 7-18 presents the flow diagram of the facility and Figure 7-19 shows the layout of the facility. The batteries will permit power to be delivered at a more constant rate despite variations in insolation during the day. They are not meant to be used to operate the unit for night-time operation.

The photovoltaic system was manufactured by AEG-TELEFUNKEN and the cell arrays have an area of about 320 ft^2 [30 m^2], with a peak power output of 2.5 kW and expected daily energy production of 7.5 kWh at 24 volts DC (Petersen et al., 1979).

The desalination unit is unique in that it uses plate and frame membrane modules manufactured by GKSS. The membranes are pressurized at 588 psi [40 atm] by a piston pump. The recovery rate is about 37 percent, and the specific energy consumption is 15 kWh/kgal [4 kWh/m^3] (Petersen et al., 1979).

Distillation

General. The concept of using solar energy for the standard distillation process (multistage flash, multiple-effect, etc.) rather than solar stills (actually solar humidification) has been evaluated for a number of years. Most of the discussions, studies, and bench-scale experiments have centered around using solar energy as an energy augmentation to these units. That is, solar energy would be used to heat feedwater, to reduce overall fuel consumption, but the remainder of the motors, pumps, instruments, controls, etc., would be conventionally powered. Some of the many variations of this work are described in United Nations (1970), Howe and Tleimat (1974), Tleimat and Howe (1977), Klaxen and Pieper (1978), Mustacchi and Cena (1978), Gasparini et al. (1979), and McCarthy and Leigh (1979).

In recent years there has been an interest in distillation plants which are completely solar powered. Such arrangements would eliminate all need (except perhaps for standby facilities) for fossil fuels and/or outside electric power. Installation of a unit such as this might be appropriate for isolated, high-fuel-cost areas.

A recent Italian study described such a facility (Arazzini et al., 1979) and envisioned the use of solar energy to operate a 0.13-mgd [492-m^3/d] multistage flash distillation plant for the desalination of seawater. The plant would operate with a top brine temperature of 250°F [121.1°C] and would have 46 stages and a performance factor of 13.5 lb/1,000 Btu's [5.79 kg/MJ].

Three energy supply variations were proposed, each using parabolic focused collectors. Typical of these was the first variation, in which a single axis tracking concentration heated a pressurized water system that was used to raise the temperature of the brine feedwater in a heat exchanger in the brine heater. A second collector system heated oil which ran two heat engines (turbines) via heat exchangers. The turbines were used for direct pump drive and electrical power generation (Arazzini et al., 1979).

Many distillation processes are quite sensitive to the mechanics of starting and stopping; hence field work and testing are necessary to avoid problems in this area.

Experience. There has been almost no experience with completely solar-operated distillation facilities. An important pilot plant which will come close to achieving that experience began operation in Mexico in 1980.

i) La Paz, Mexico (2,610-gpd [10-m^3/d] Solar MSF)--This facility has been installed as an experimental unit in the City of La Paz in the arid peninsula of Baja, California. It is a joint research effort of the Mexican Government's desalination agency, DIGAASES, the Federal Republic of Germany, and the German firm, Dornier System GmbH. The facility is installed at the DIGAASES research station in La Paz.

The purpose of the project is to gain actual field experience with the interfacing of solar energy with a multistage flash distillation plant. Of special concern is the adaptability of the unit to operate in a stable condition with variations in the energy supply.

The facility is made up of a solar section designed by Dornier and a distillation unit designed by DIGAASES. The solar section consists of flat plate collectors (for low temperature) and parabolic concentrators (for high temperature). Water is the fluid used for heat collection and transfer in a closed loop. A heat storage system capable of storing sufficient thermal energy to provide a stable energy source for about 28 hours is provided. The plan is for the distillation unit to operate 24 hours per day.

The flat plate collectors are divided into three groups, one to provide energy for daytime operation and two to provide energy storage for night-time operation. Total flat plate collector area is about 5,600 ft^2 [520 m^2]. The parabolic concentrators will have a total collection surface area of about 1,700 ft^2 [158 m^2].

The distillation unit is a multistage flash plant made from 90/10 copper-nickel. It is a 10-stage plant using brine recirculation and has a design performance factor of 5 lb/1,000 Btu's [2.15 kg/MJ]. The stages are built with external condensers. This basic design has been used previously by DIGAASES on other distillation plants in the country. The upper design brine temperature is 190°F [87.8°C] (Manjarrez and Galvan, 1979).

Electrodialysis

General. Electrodialysis is just beginning to be seriously investigated for solar energy applications. The process itself has three major characteristics which lend themselves to use with solar energy:

1. Electrodialysis' major energy requirement is for direct current, which can be obtained directly from photovoltaic cells. Other processes which require alternating current lose 5 to 10 percent in the conversion systems.

2. The electrodialysis process is amenable to shutdown and startup.

3. Any heating of the feedwater by solar energy or other inputs improves the overall efficiency of the membrane stack operation.

Experience. Although there has been little or no field experience with solar-powered electrodialysis, the U.S. Office of Water Research and Technology (OWRT) has financed an interesting and detailed study on the subject.

The facility was designed to operate in a remote location in the southwestern USA. Brackish water with a TDS of 2,000 to 2,500 mg/l would be treated with an average production of about 4,000 gpd [15 m^3/d]. A water analysis for the Town of La Luz, New Mexico, was used for design. The plant was designed to operate using photovoltaic cells as the source of power. A flow diagram is shown on Figure 7-20. Due to the high calcium sulfate content of the proposed feedwater from Luz, the recovery is limited to about 50 to 60 percent without chemical treatment. To keep the unit operations as simple as possible the recovery has been limited to 50 percent, and no chemicals are added during pretreatment.

The plant is designed to operate 10 hours per day, with the batteries acting to smooth out transients rather than to provide long-term energy storage. To minimize the solar energy collection system and yet maximize output, the facility has two distinct modes of operation, full and half flow. During full flow the unit uses the main feed pump and produces at the rated capacity. For the half flow mode the main feed pump is shut off and a smaller, secondary feed pump is operated. Production is then reduced to about half and the power required is reduced

50 to 75 percent. The half flow mode is used during periods when sunlight is reduced by clouds or other atmospheric conditions and when the well pump operates.

In addition, the flow regulators and variable valves are to be instrumented such that the flow and recirculation through the membrane stack will be regulated so as to match the resistance in the stack (based on flow and TDS level) to the amount of incoming energy from the photovoltaic cells. The feedwater to the plant is used to cool the photovoltaic cells. This not only conserves pumping but heats the feedwater and improves the efficiency of salt removal in the membrane stacks. A preliminary layout for the facility is shown on Figure 7-21.

SOLERAS (Saudi Arabia--USA) Solar Desalination Project

In October 1977 the Kingdom of Saudi Arabia and the United States signed a project agreement for cooperation in the field of solar energy (SOLERAS). The objectives of the agreement were to:

Cooperate in the field of solar energy technology.

Advance the development of solar energy in the two countries.

Facilitate technology transfer between the two countries.

The Solar Energy Research Institute (SERI) in Golden, Colorado, is responsible for implementing the SOLERAS program. One area in which the SOLERAS program will operate is solar desalination.

In March 1980 SERI accepted proposals concerning studies on the technical and economic feasibility of large-scale desalination of brackish and seawater using solar power exclusively. These proposals were the initial part of a three-phase activity, consisting of:

Phase 1--Preliminary system design and cost analysis.

Phase 2--Detailed pilot plant design and construction.

Phase 3--Pilot plant operation and training of personnel.

Phase 1 will include preliminary system design and cost analysis of both a 1.6-mgd [6,000-m^3/d] base plant and a 0.025- to 0.1-mgd [100- to 400-m^3/d] pilot plant for brackish water and seawater. Phase 2 consists of detailed design and the actual construction of the pilot plant. Phase 3 is the operation and evaluation of the pilot plant.

Proposals for both brackish water and seawater plants were accepted with the supposition that a brackish plant might be built in the USA and a seawater plant in Saudi Arabia. The timing for the project is for phase 1 to start in 1980, phase 2 to start in 1981 with completion of the pilot plant by the end of 1982, and phase 3 to be completed by the end of 1983.

Aside from the actual project itself, which would be one of the largest solar projects in the world, there should be considerable information of interest on solar desalination in the proposals which were submitted to SERI.

WIND POWER

General. There is evidence that wind power has been used in windmill-type devices for several thousand years. Their major functions have been to grind, pump water, and (most recently) to generate electricity. In the USA and Europe, various types of wind machines were commonplace at the start of the 20th century but, with the advent of rural electrification and dependable gas engines, they were largely displaced.

Windmills have been constructed in many parts of the world, with three basic types of machines being developed. These can be categorized by the orientation of their rotating axis with relation to the wind and the earth:

	Axis Orientation	
Type	Relative to the Earth	Relative to the Wind
1	Horizontal	In-line
2	Horizontal	At right angles
3	Vertical	At right angles

These three types are illustrated on Figure 7-22. Almost all modern wind machines used today are Type 1 or 3.

No matter what type of machine is employed, the product is the same--rotary mechanical energy (shaft power). The machines can be coupled directly to a pump, saw, grinding wheel, etc., or used to generate electricity, which is usually an easier form of energy to distribute and store. However, as wind machines are used to generate electricity they become more complex, since they must maintain voltage, cycles, etc. As discussed by Brace Research Institute (1979), a wind-powered system consists of a windmill, transmission system, load, and usually energy storage, operating as a complete system.

Wind power actually is an indirect form of solar power, since it is the sun that causes the differential heating on the earth, creating winds. Wind is simply gas molecules moving in the air. These molecules impart energy to the blade of a wind machine as they strike them in passing. The power imparted is dependent on the area covered by the blades, the velocity of the wind, and the design of the wind machine. The basic formula for wind power is:

$$P = KAV^3$$

where

P = power, K = a constant, A = the blade area, and V = wind velocity.

Power is generally expressed in watts, and the constant depends on the efficiency of the machine. The blade area refers to the area which the blades sweep at right angles to the wind, which is generally a circle. The maximum power theoretically available for extraction by a wind machine is 59.3 percent. Usually wind machines can utilize from 40 to 70 percent of the theoretical power available, depending on design.

The power available is dependent on the cube of the wind velocity. Hence, velocity is an important factor which must be investigated before deciding what type of installation is possible. Wind, speed, duration, and direction are affected by the season, location, surroundings, height above the ground, etc. The design must take into account not only average winds but also the extremes.

In calms no power will be produced and during high winds, especially gusts, the structure could be physically damaged. Other extreme weather conditions such as hail, rain, snow, icing, and sandstorms can affect the machine's performance, its design, and its tower or other support.

Some improvements in wind machines have occurred in the past 50 years with attention being paid to their aerodynamic design. As with solar stills, the potential for wind power has intrigued man and has stimulated numerous variations on basic designs and materials of construction. Many of the commercial wind machines feature designs using a vertical axis such as the Darrieus and Savonius rotors.

Technological progress over the old style wind machines will probably continue to include the use of advanced rotor designs and improved solid state circuitry in the electric generation equipment to control voltage and/or cycle synchronization.

Some of the common types of wind machines in use today are illustrated on Figure 7-23. Table 7-2 contains information on the characteristics of several wind-powered electrical generating systems.

Efforts have been made by various organizations, such as the Brace Research Institute, the Indian Institute of Science, Volunteers in Technical Assistance (VITA), etc., to design efficient but simplified wind machines that can be built in developing areas.

Table 7-2
CHARACTERISTICS OF SOME WIND POWERED ELECTRICAL GENERATING SYSTEMS[a]

Trade Name	Blades		Rated Power		Wind Speed Characteristics			Weight on Tower[d] (lb)
					Operating[b]		Maximum[c]	
	No.	Diameter (ft)	Power @ (kW)	Wind Speed (mph)	Cut-in (mph)	Cut-out (mph)	Allowable (mph)	
Propeller turbine with tail:								
Aerowatt (France)	3	30.7	4.1	16	6	55	125	2,100
Dakota (USA)	3	14	4.0	24	10	60	--	550
Eleckro (Switzerland)	3	16.5	6.0	30	10	45	--	--
Millville (USA)	3	25	10.0	25	11	50	--	850
Windgenni (USA)	3	12	3.0	20	10	100	100+	250
Downwind propeller turbine without tail:								
Grumman (USA)	3	25	15.0	26	8	50	130	2,000
Zephyr (USA)	3	20	15.0	30	8	45	--	600
Bicycle turbine with tail:								
American Wind Turbine	48	16	2.0	20	8	40	125	520
Amerenalt (USA)	24	12	2.5	27	8	40	90	--
Darrieus rotor:								
DAF (Canada)	2	20 x 30	7.5	23	7	65	--	--

Note: Metric conversion units are: ft x 0.3048 = m; mph x 1.609 = kmph; lb x 0.454 = kg.

[a] Data selected from more complete and informative tables in Barnett (1978). Other information can be found in Brace Research Institute (1979).

[b] Cut-in is lowest wind speed at which machine delivers any electricity. Cut-out is wind speed at which machine begins to automatically protect itself against excessive rotational speed. In some machines this stops electric generation; in others it reduces electricity production, which may continue to survival speed.

[c] Maximum wind speed that machine can normally survive and still operate again at lower speed.

[d] Maximum weight of machine that has to be lifted to the top of the tower.

Any consideration of wind power for desalination must take into account the fact that wind speed and continuity are site-specific. Data need to be obtained on these meteorological variables so as to determine if wind power is possible, and if it is, to ascertain the economical energy level (mean wind speed) to use for design. Another important factor is that, although the wind is a free and inexhaustible source of energy, it does not always blow, and when it does, the velocity might be either too low to permit the wind machine to begin to produce power or too high for the mechanical design that is practical for the mean wind speed (NAS, 1976). Numerous good texts are available to assist in planning a wind energy project, including those by Golding (1955), Simmons (1975), and Justus (1978).

The economics of individual situations revolve for the most part around a balance of (1) the capital costs for wind machine capacity, (2) energy storage, (3) desalination plant capacity, and (4) water storage in order to satisfy the base and peak water demands.

The use of wind power for desalination is only beginning to be explored. As far as can be determined, there are only a few experimental units being built and none that are actually operating under realistic commercial conditions.

Brace Research Institute (Canada) conducted experiments over a period of years using a simulated windmill to run a standard reverse osmosis unit on brackish water. The varying flow rates had no significant effect on long-term performance.

Reverse Osmosis

General. The combination of reverse osmosis with wind power is a fairly new application. In theory, power would need to be furnished either as electricity or rotating energy. If electricity were produced it would power motors to drive the pumps, etc., while rotating energy would directly drive the pumps, although some electrical energy would usually be needed to operate controls, instruments, etc. Most often wind power has been used to produce electricity, which is then used to power the motors which drive the pumps, etc.

As discussed in the solar power section, reverse osmosis units need to be specially designed if they are going to be subjected to constant start-stop operation. Thus, a workable system should have provisions to handle transients in energy generation.

Experience. The use of wind power for reverse osmosis applications has been discussed (Cadwallader et al., 1977), but very few installations have been built. Two units are described in the following paragraphs.

i) North Sea, Germany (1.9-gpm [10.4-m^3/d] Wind Powered RO)--The unit in Germany constructed by GKSS is for use on an island in the North Sea. The facility is powered by a three-bladed, horizontal axis, head-on, propeller-driven wind machine manufactured by Allgaier/Hütter. The wind machine drives the generator, which produces three-phase alternating current at 380 volts. This is rectified to variable voltage direct current, which is used to operate the pumps in the reverse osmosis system.

The flow design for the system is illustrated on Figure 7-24. Photos of the system as it was set up at the GKSS Research Center are shown on Figure 7-25.

The reverse osmosis unit is a plate and frame system which operates on North Sea water with a recovery of about 40 percent and a membrane pressure of 1,160 psi [79 atm]. The specific energy consumption is about 42 kWh/kgal [11.1 kWh/m^3] of product. Pretreatment consists of mechanical filtration with the potential for backwash and the addition of acid (Peterson et al., 1979).

ii) <u>Planier Island, France (2.2-gpm [12-m^3/d] Wind-Powered RO)</u>-- This unit is being installed on the island of Planier in the Mediterranean off the coast of Marseilles. It has been designed and tested by the French Atomic Energy (CEA) Center for Nuclear Studies at Cadarache.

As with the CEA solar unit, this system has been designed for simplified pretreatment and the use of energy recovery to reduce the power consumption. The pretreatment consists of electrochlorination, a settling tank, a two-stage sand filter, an activated charcoal column, and a cartridge filter. It uses an efficient high-pressure displacement pump with three pistons at 120° and a Pelton turbine for energy recovery. The total energy usage, with recovery, is expected to be 28.2 kWh/kgal [7.5 kWh/m^3] of product including the low-pressure and high-pressure pumps (Maurel, 1979a). Hollow fine fiber reverse osmosis modules are used, and the design recovery is 25 percent.

The wind turbine is an AEROWATT Model 4100 FPX, using a two-bladed propeller with a rotor diameter of 30.2 feet [9.2 m]. The mast height is 49 feet [14.9 m]. The generator output is 4 kW, 3 phase, 380 volt AC, with a wind speed of 23 ft/sec [7 m/s].

The unit has been tested at the CEA Research Center at Cadarache, France and is expected to be installed at Planier during 1980.

<u>Distillation</u>

There is some potential for using the rotating energy from a wind machine for operation of a vapor compression unit. Careful design and consideration of the interface between the two systems, however, would be required.

Units with this type of application have received little or no field testing. The concept has been discussed in theory, however, in 1961 by Lawand, who performed extensive studies on a 500-gpd [1.9-m^3/d] vapor compression unit to test its performance with a varying energy source.

Electrodialysis

There is some potential for running an electrodialysis system using wind power. This type of system can tolerate the variable energy output of wind generators better than the other processes, but design work remains to be done to account for energy transients and power variability.

Little or no field work has been done with wind-powered electrodialysis, although it has been considered in theory in reports (Cadwallader et al., 1977).

WAVE ENERGY

General

The ocean's waves have long fascinated man with their potential as a source for free power. Investigations into recovering energy have been concentrated in two areas: (1) employing energy derived from tides and currents, and (2) recovering energy contained in wind waves. Wind waves are non-tidal in nature and are generated by winds sweeping across the sea's surface. These can be either local winds or winds from storms, hurricanes, etc.

The trapping of tidal water at high tide in a reservoir and releasing it through turbines at low tides has been discussed and investigated in various parts of the world. This requires a specialized site and massive structures.

More recently, efforts have been concentrated on harnessing the energy from wind waves by building devices to transform either the up-and-down or apparently moving wave motion of the sea into useful work. The waves only seem to move, as the relative position of individual water molecules does not significantly move--just the shape of the wave moves along the surface of the water. The energy in a wave is determined by $P = K H^2 L$, where K is a constant, H is the height of the wave, and L is the wave length. Half of the energy is kinetic and the other half potential.

The major problems with harnessing wave energy are associated with the changing wave patterns and the tremendous force which can be contained in some waves. A device needs to be designed to work with varying wave patterns and yet be rugged enough to withstand a massive wave that might only occur once every year or so. Research is continuing on the subject of waves and their inherent characteristics and potential use for energy, much of it focusing on offshore structures such as drilling platforms, which must withstand the force of waves. The loss of some of these massive rigs indicates that there probably is still much to be learned about waves and appropriate construction techniques.

Pleass (1978) has classified four of the commonly used methods for harnessing wind-wave energy as:

1. Interaction with wave orbits.
2. Utilization of a pressure field.
3. Use of accelerative devices.
4. Use of mass transport.

A good deal of work is going on in the United Kingdom on devices which interact with the wave orbits. Two basic types which have been experimented with are a raft hinged in at least two places and a device resembling a line of bobbing ducks, or cams, which rotate individually against a spine connecting them. The relative motion of one part of either of these devices versus the others is used to produce mechanical energy.

The pressure field principle is used commercially to generate electricity to light a small floating bouy. These devices, referred to as Masuda bouys, are manufactured in Japan. The device consists of an open can-like body placed with the open end down into the water. It has two holes in the top and as the waves move up and down in the can, air moves and is sucked or blown out, with the water acting like a piston. This air is used to drive turbines in the bouy, which are used to generate electricity. Pressure field devices such as the Masuda bouy rely on the relative motion between the device and the waves to create an air flow.

An example of an accelerative device is presented later in this chapter, in connection with a direct desalination application. This device is illustrated on Figure 7-26. The wave follower (float) has a general up-and-down motion and provides a force relative to the seabed to which the lower part is secured, thus creating mechanical work. Hydrodynamic reaction plates have also been used in earlier models to aid in providing relative motion to the device without depending wholly on the seabed anchoring device.

An application for a mass transport type of wave energy application was suggested by Mr. Bolton Botts for the island of Mauritius in the 1960's. This employed the energy produced from a differential head built up by waves which crash over a wall (in this case a built-up fringing reef) and then flow out of a lagoon. This general phenomenon can be seen in many of the islands with reef-enclosed lagoons. It has been suggested that this differential head be used to power hydraulic fans and/or turbines for purposes of energy conversion (Ross, 1979).

In wave energy, as with some of the other alternate energy technologies, there is still a great deal of research, development, and field testing necessary, both in providing reliable economic power and in combining it with desalination.

Experience

As far as is known, there is no commercially operating desalination unit employing wave energy as its power source, even though several proposals and/or experiments have appeared on a small scale. One of these is described below:

La Parguera, Puerto Rico (1-gpm [5.5-m^3/d] Wave-Powered RO).

During 1980 an experimental wave-powered seawater reverse osmosis unit will be installed near a small drying reef known as Media Luna, off the southwest coast of Puerto Rico near the town of La Parguera. The device, called the DELBOUY II system, is shown in Figure 7-26. The key to the system is a unique high-pressure pump made from PVC pipe, urethane elastomers, and natural rubber tubing (Pleass, 1979). The feedwater will be obtained from a seawell adjacent to the system, and no additional pretreatment is planned for the unit. The unit uses a small DuPont hollow fine fiber permeator for desalination.

Prior to deployment in Puerto Rico tests on both quarter- and 1/15th-scale models were conducted in the USA by researchers from the University of Delaware. The University is working on this project under the direction of Dr. C. M. Pleass using partial funding from the Sea Grant program. Special efforts are being made in this program to use techniques and materials which are readily available in developing areas, and components which can be fabricated and repaired easily. A distinguishing feature of this facility is that the wave follower is sacrificed in extreme storms. This lessens the need to use expensive structuralcomponents and hence reduces the capital cost involved.

HAND POWER

The direct use of hand power is probably the earliest alternate energy source used by man. Some attempts have been made to utilize hand power for desalination in reverse osmosis applications. One of the first applications used was in life rafts to desalinate seawater. This recalls the work of Dr. Telkes in World War II on devising solar stills for life rafts.

Seagold Industries Corporation of Canada has developed the unit, called Waterlever 25, shown on Figure 7-27. The unit operates using the hand lever shown on the figure. It can be used on both brackish water and seawater. The design specifications given by the manufacturer are as follows:

Operating Pressure:	
Seawater	1,000 psi [68 atm]
Brackish water	500 psi [34 atm]
Recovery	5 to 10 percent
Production (seawater)	1 gph [4 l/hr]
Salt Rejection:	
Seawater	98 percent
Brackish water	99 percent
Weight (dry)	13.2 lb [6 kg]

Salt rejection varies depending on the rate of pumping. A fine mesh suction strainer is used for pretreatment, but it is recommended that a special fine filtration device be used in addition to protect the unit in all but survival applications. According to the manufacturer the unit could be used to supply water while traveling in areas where water sources other than freshwater exist. The reverse osmosis membrane provides an added benefit in that it affords some protection to the product against microbial contamination.

Seagold has incorporated an energy recovery mechanism in their hand pump and are using the same principle in the development of a line of triplex pumps for use in seawater reverse osmosis.

OTHER ENERGY SOURCES

This chapter is limited in extent and has covered only some of the many alternate sources of energy that could be used with desalination. Some others include geothermal energy, low-head hydropower, biomass conversion, and ocean thermal energy conversion.

The desalination of geothermal brines is being investigated, and several proposals have been forwarded to use ocean thermal energy conversion (OTEC) for direct desalination of ocean water. One principle behind OTEC involves an attempt to utilize the temperature difference between the colder, deep ocean water and the warmer water on the surface. This requires an area where deep seawater is available. In some cases, the temperature difference can be increased by using waste heat, artificial solar heating, etc.

Both the Europeans and the USA are anxious to start an OTEC project which would include desalination and electric generation, and a project of this nature will probably be undertaken in the 1980's.

APPLICATION TO DEVELOPING COUNTRIES

Although the use of some renewable energy sources offers the potential for drastically reducing energy costs, its use for desalination in developing areas needs to be approached with caution and good judgment. The idea of alternate energy sources may convey the sense of going to a simpler way of life, but the fact is that there are still some very complex elements involved in selecting reliable units and properly interfacing them with desalination units. There has been very little experience in this regard and most of it is taking place under highly supervised and well-financed conditions.

Long-term commercial experience needs to be gained in this area to work out the problems in design, operation, materials, etc. This experience is not inexpensive, as units fail, need to be modified or repaired, and in general, require special attention during this phase of development. Whoever finances such projects must be able to afford unexpected expenses.

In the long term, there is certainly a potential for desalination using alternate energy sources in selected developing areas where conditions of climate, topography, technology, and economics are suitable. The test units described in this chapter will aid in clarifying some of the opportunities and problems that exist in the field. However, except for solar stills, it appears that each type of technology will require more work before units are ready for reliable commercial (versus research or experimental) operation in developing areas. At this time to routinely place these units in villages (where maintenance capability is lacking) would probably be counterproductive to both the concept of renewable energy and desalination. For any application involving desalination by renewable, or alternate, energy sources consulting someone with professional knowledge and experience in the field would be prudent.

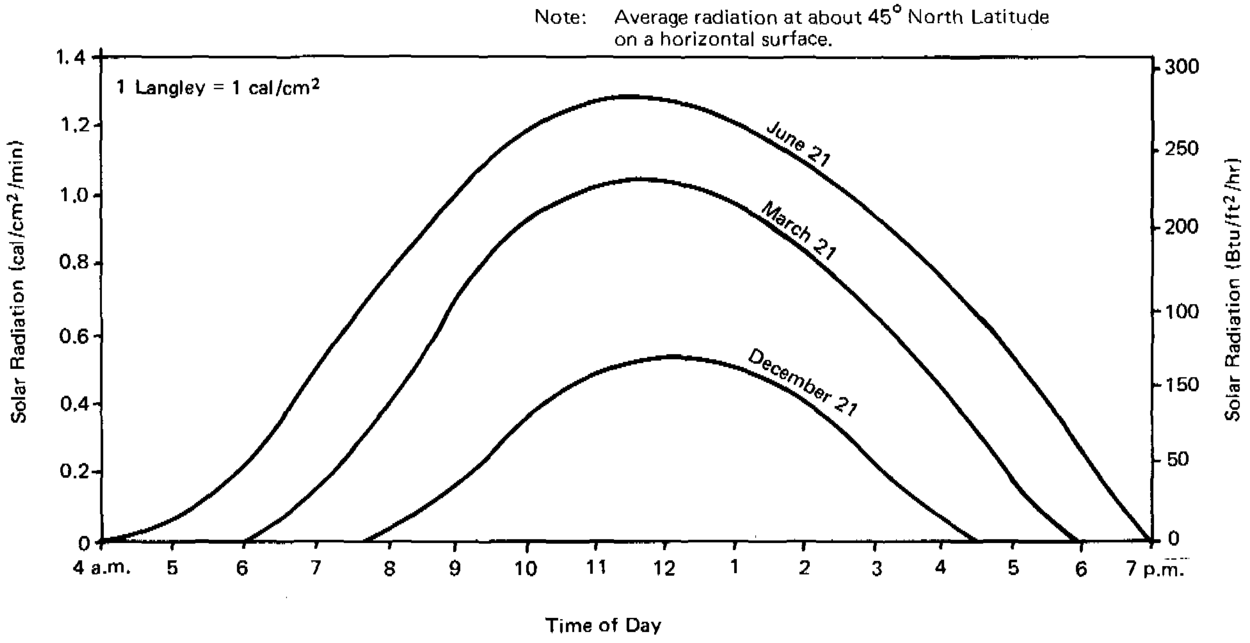

Note: Average radiation at about 45° North Latitude on a horizontal surface.

FIGURE 7-1. Variations in daytime solar radiation.

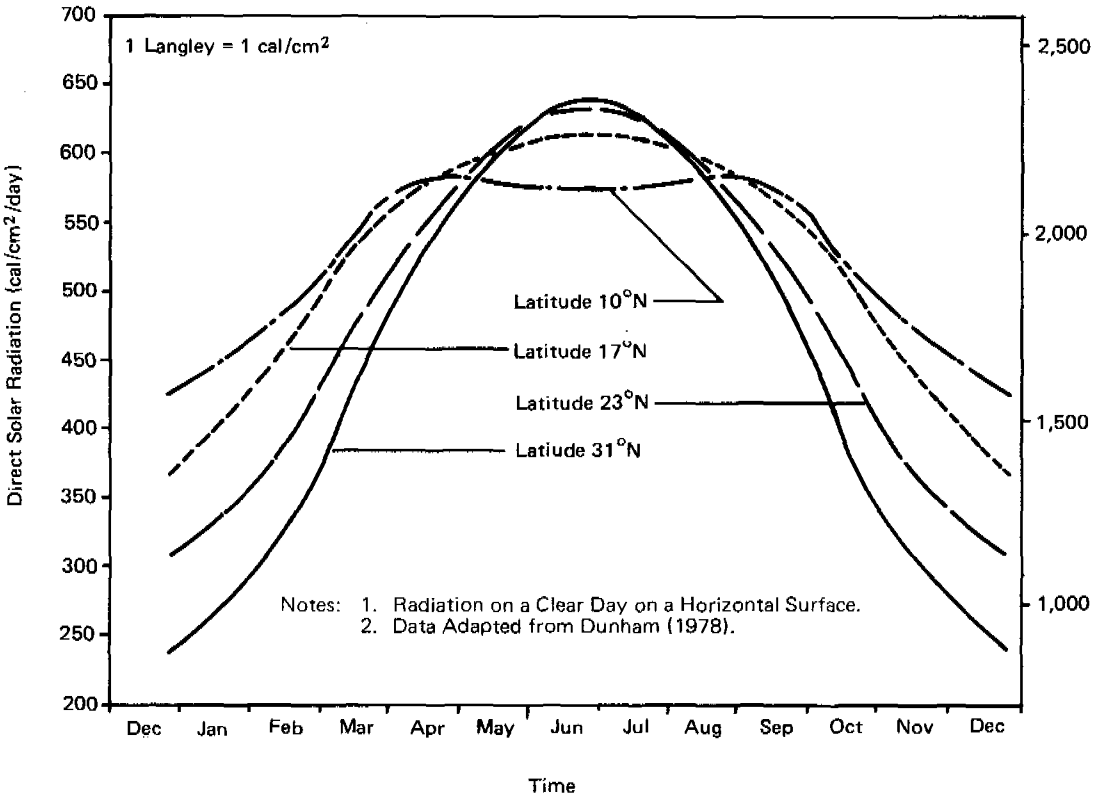

Notes: 1. Radiation on a Clear Day on a Horizontal Surface.
2. Data Adapted from Dunham (1978).

FIGURE 7-2. Variations in solar radiation with latitude.

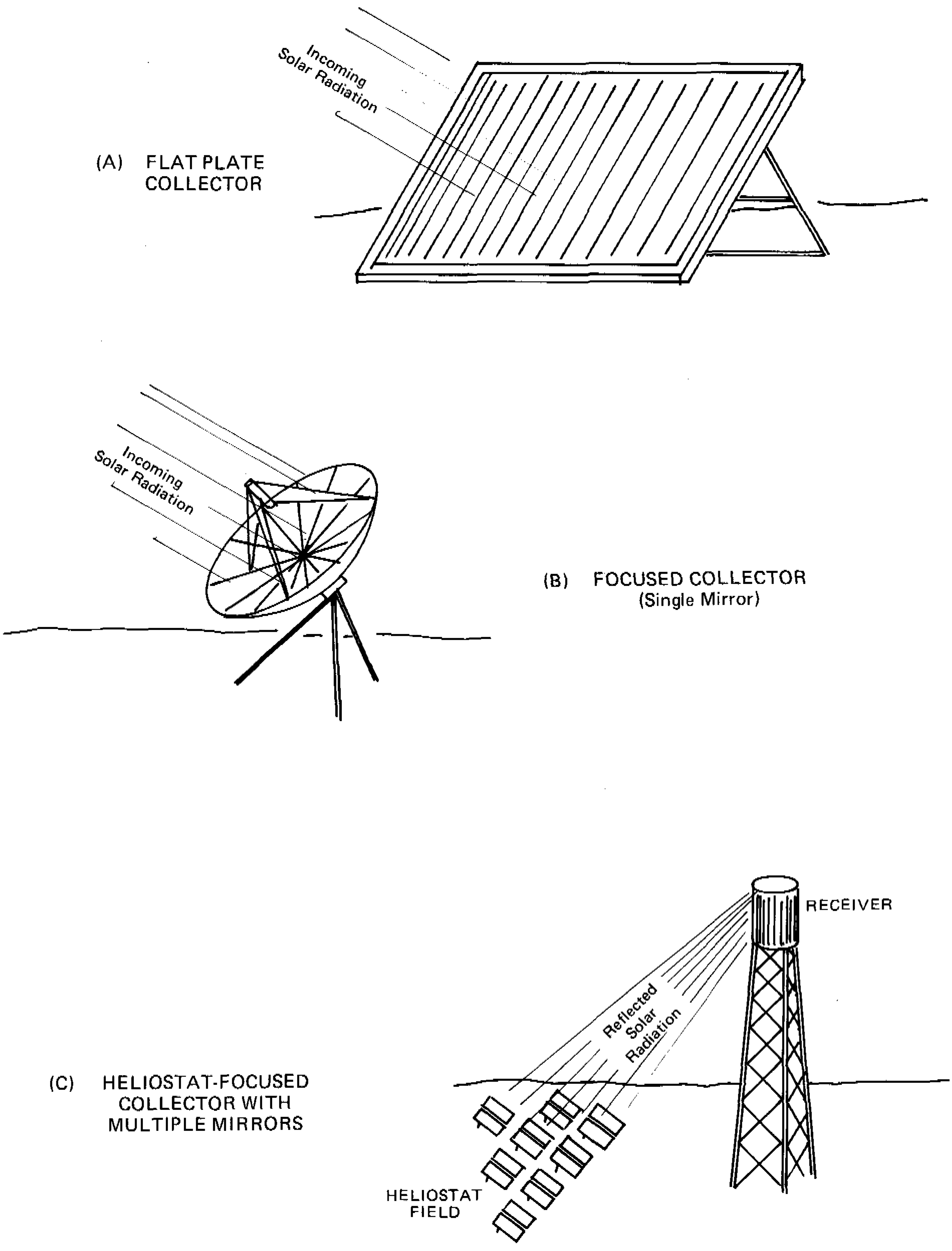
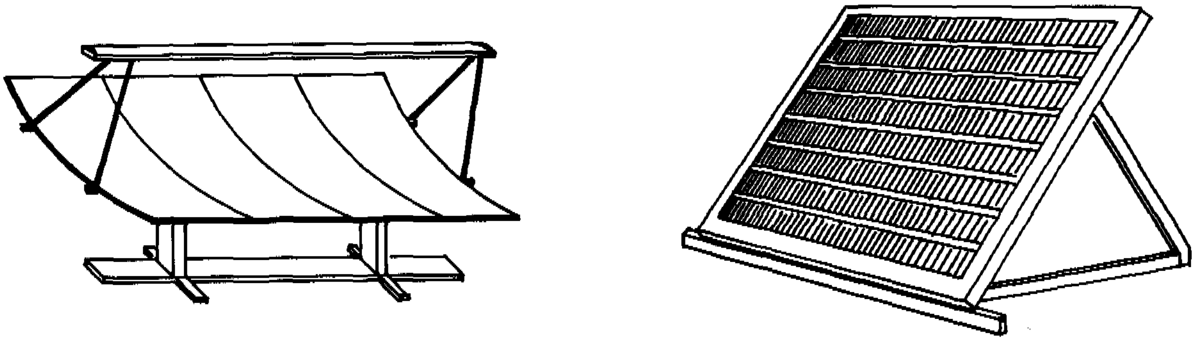


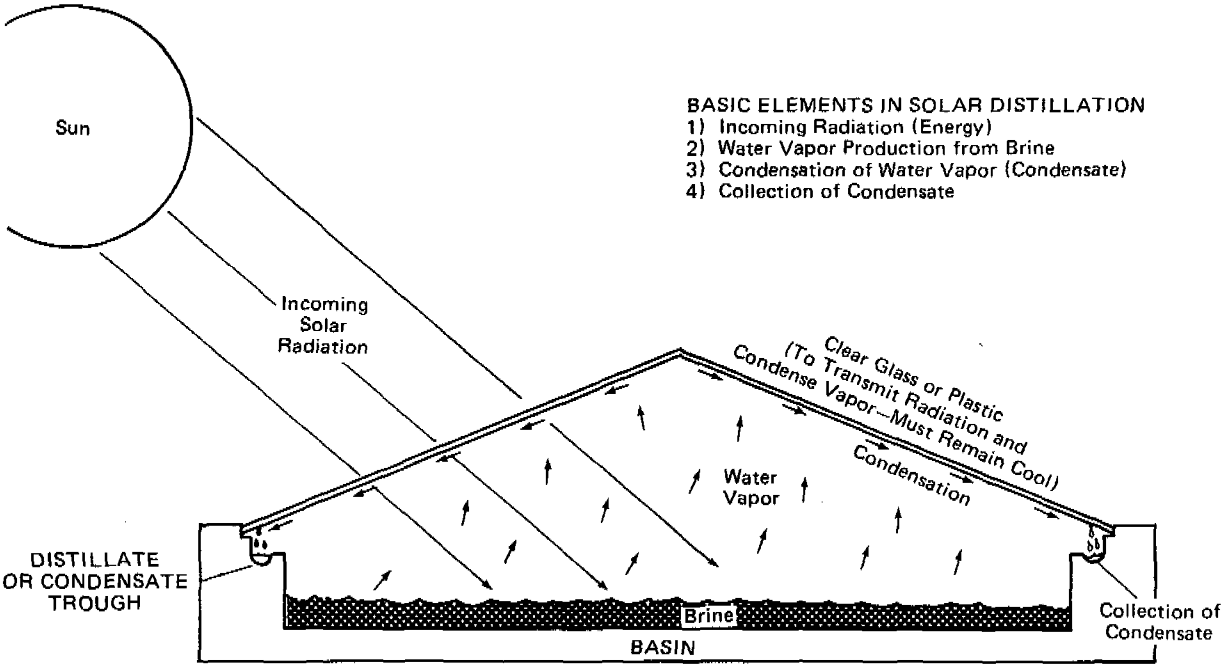
FIGURE 7-3. Representative solar-heat conversion devices.



Strip Photovoltaic Array Using a Cylindrical Parabolic Concentrator.

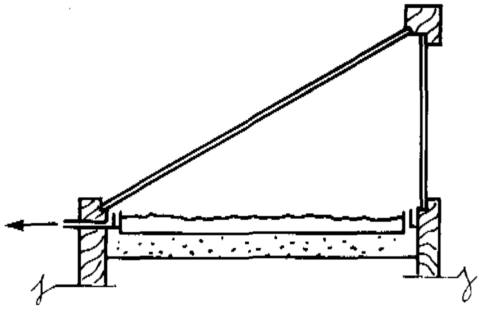
Flat Photovoltaic Array

FIGURE 7-4. Representative solar photovoltaic conversion devices.



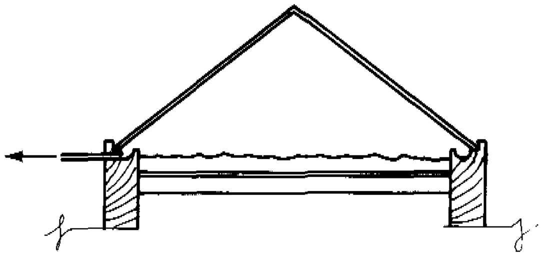
The inside of the basin is usually black to efficiently absorb radiation and insulated on the bottom to retain heat.

FIGURE 7-5. Basic elements in a solar still.

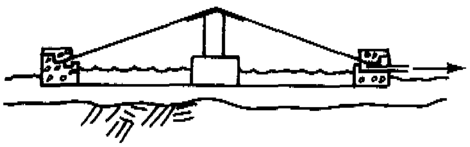


UNIV. OF CALIF. SOLAR STILL NO. 1

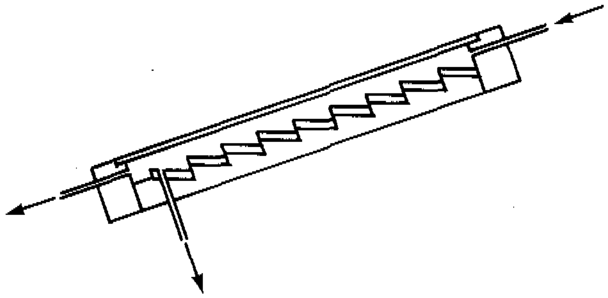
UNIV. OF CALIF. SOLAR STILL NO. 20



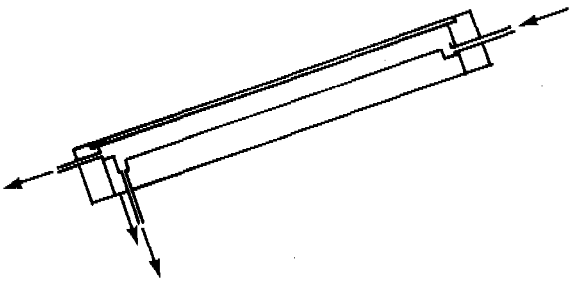
UNIV. OF CALIF. SOLAR STILL NO. 16



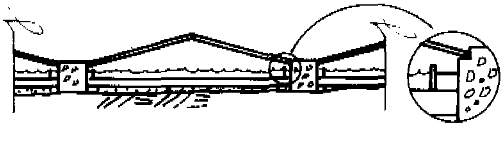
UNIV. OF CALIF. SOLAR STILL NO. 42



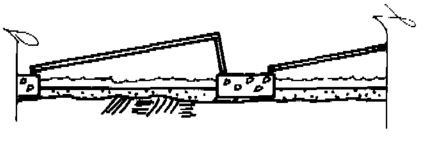
TILTED-TRAY STILL SCHEMATIC



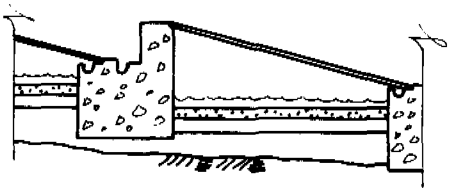
WICK-TYPE STILL SCHEMATIC



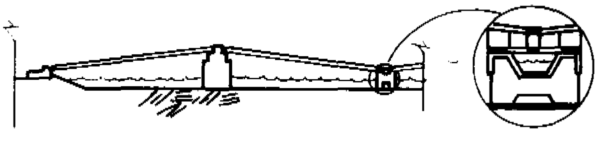
CSIRO DESIGN MARK VI
Australia



DELYANNIS DESIGN
Greece



McGILL UNIV. DESIGN
Haiti



MODIFIED DEEP-BASIN DESIGN
Las Marinas, Spain

Cross sections from Tleimat, 1978 in the United Nations Industrial Development Organization publication, Technology for Solar Energy Utilization.

FIGURE 7-6. Cross sections of some common types of passive solar stills.

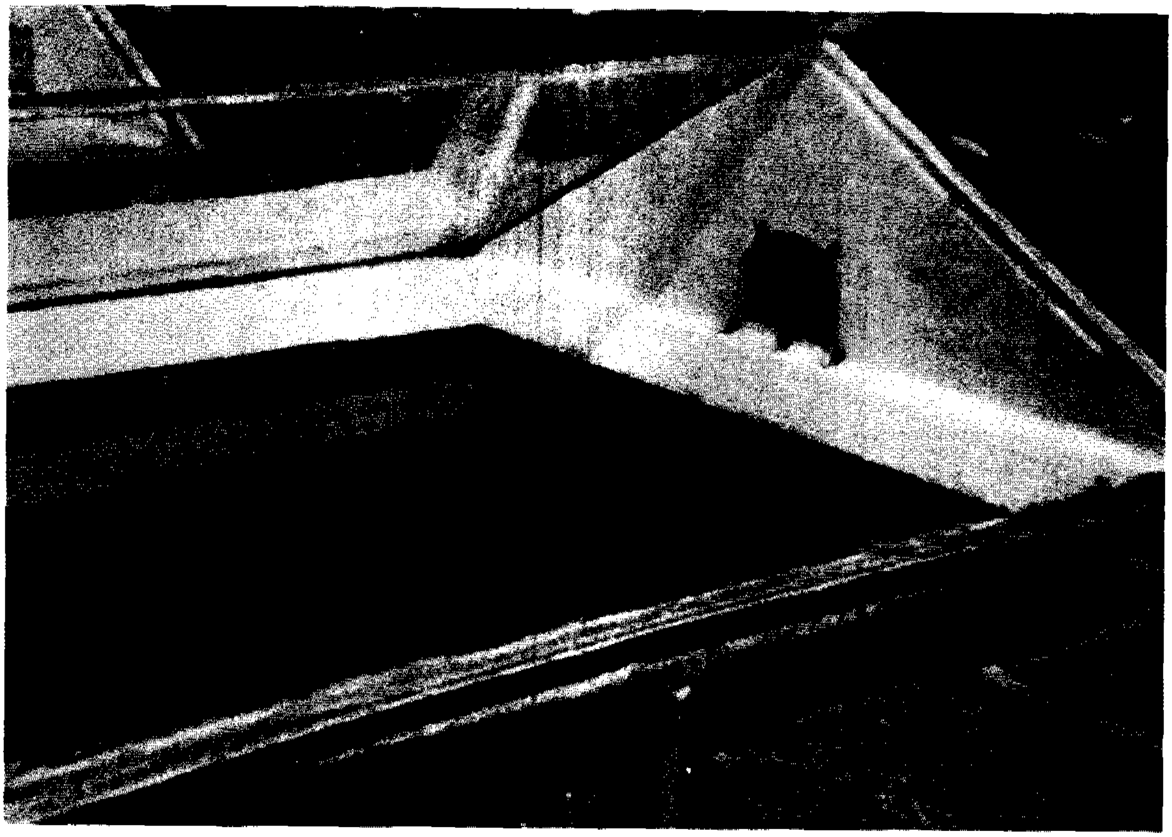
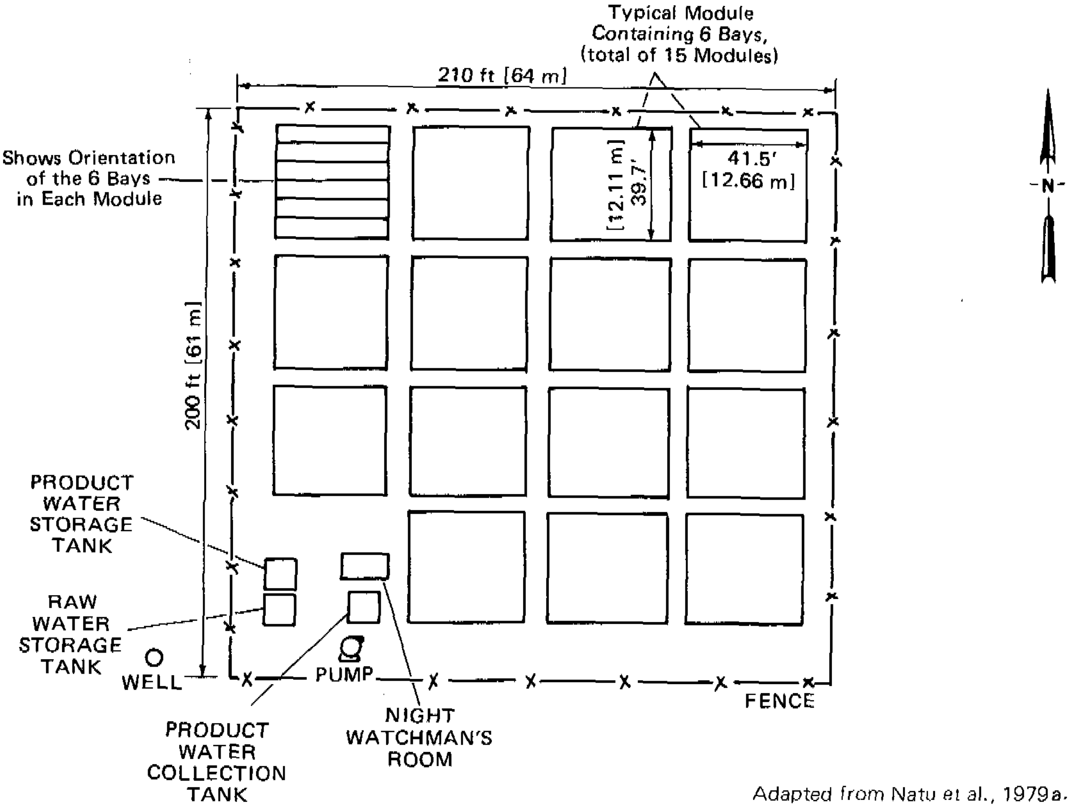


Photo Courtesty of O. K. Buros.

FIGURE 7-7. Access hatch in a small shallow-basin solar still.



Adapted from Natu et al., 1979a.

FIGURE 7-8. Layout of the solar distillation facility at Awania, India.

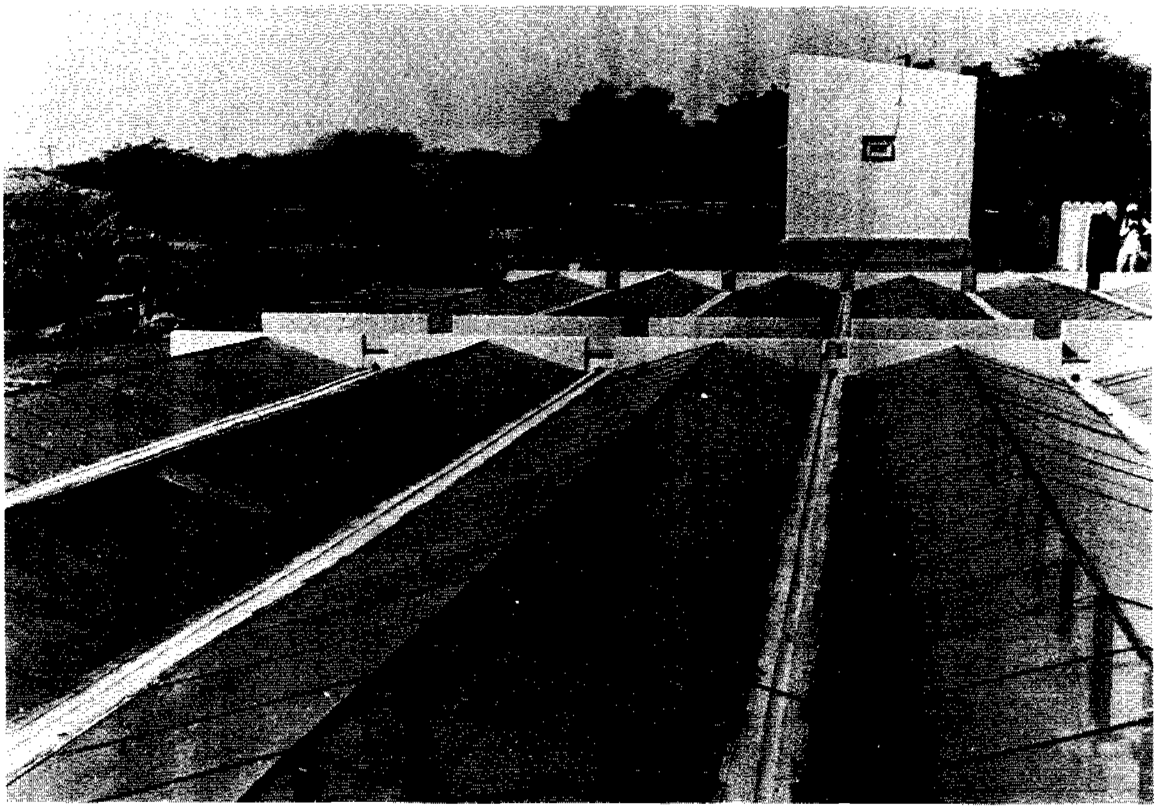
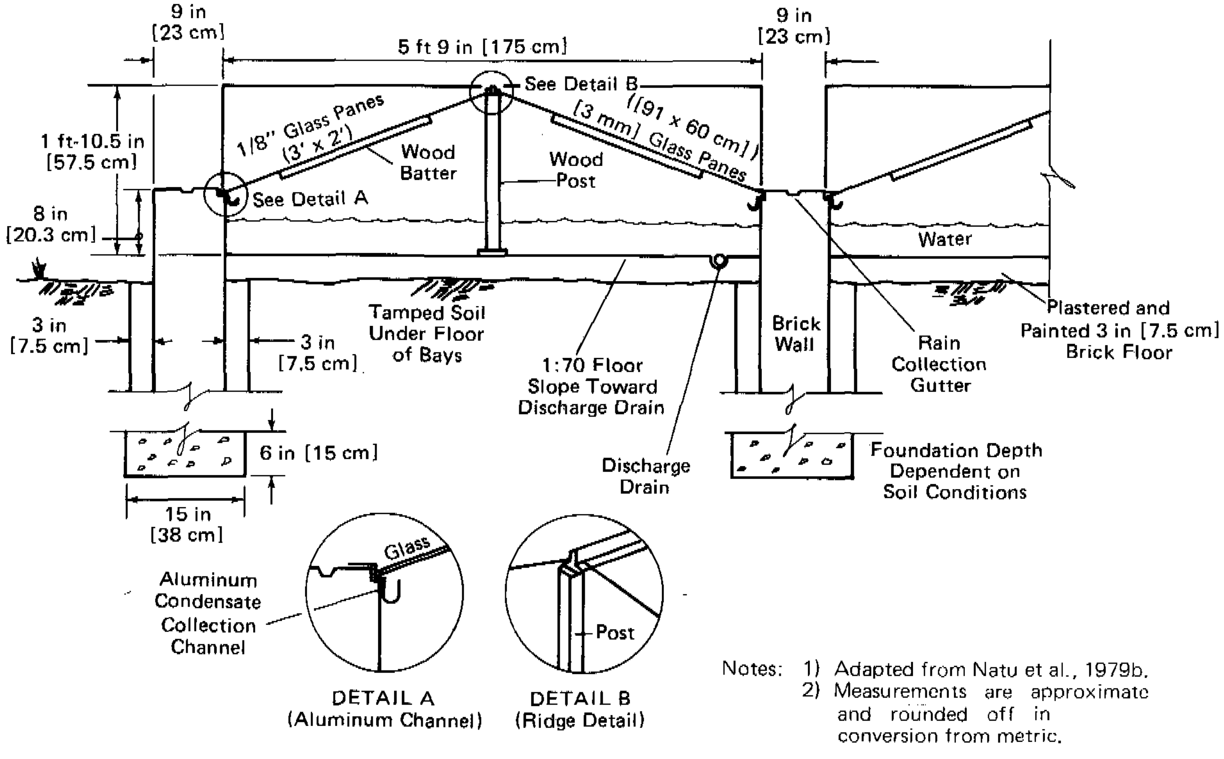


Photo Courtesy of the Central Salt and Marine Chemicals Research Institute, Bhavnagar, India.

FIGURE 7-9. View of some of the 90 bays that make up the Awania solar still.



Notes: 1) Adapted from Natu et al., 1979b.
2) Measurements are approximate and rounded off in conversion from metric.

FIGURE 7-10. Typical cross section of a solar still bay at Awania, India.

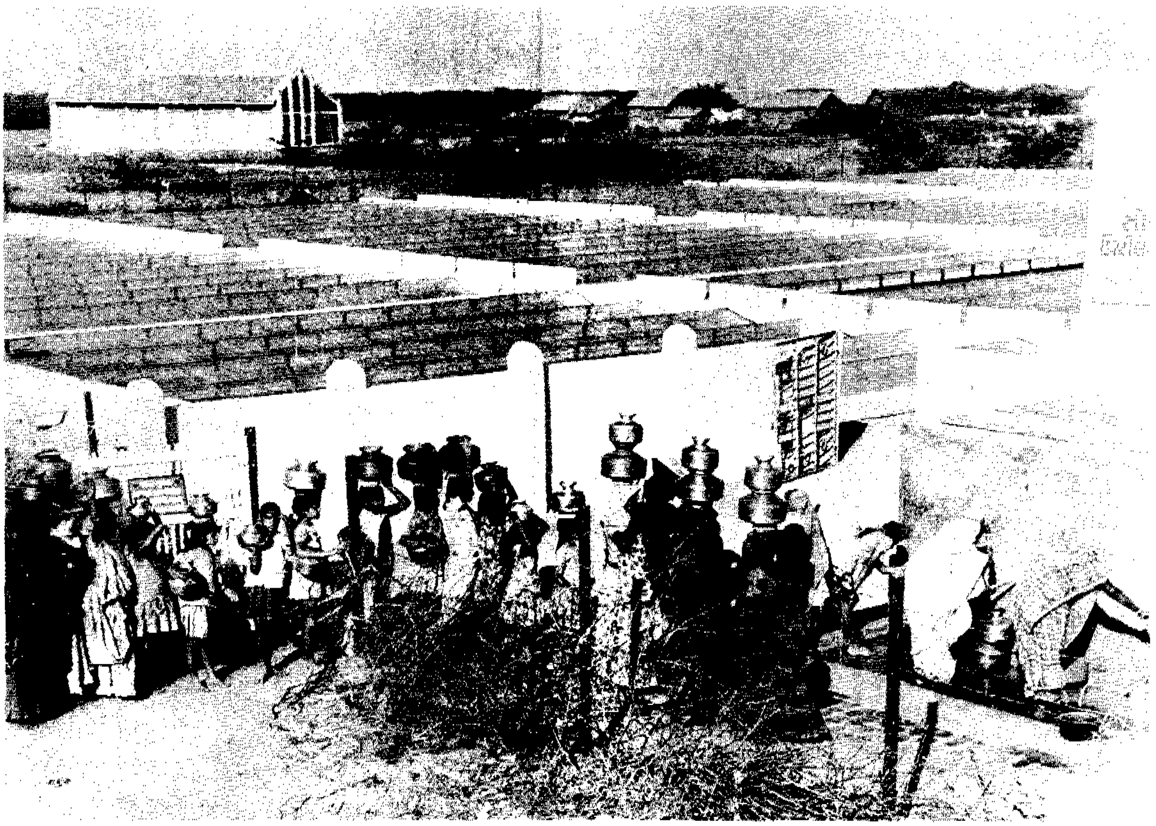


Photo Courtesy of Central Salt & Marine Chemicals Research Institute, Bhavnagar, India.

FIGURE 7-11. Product water being distributed from taps at the storage tank at Awania.

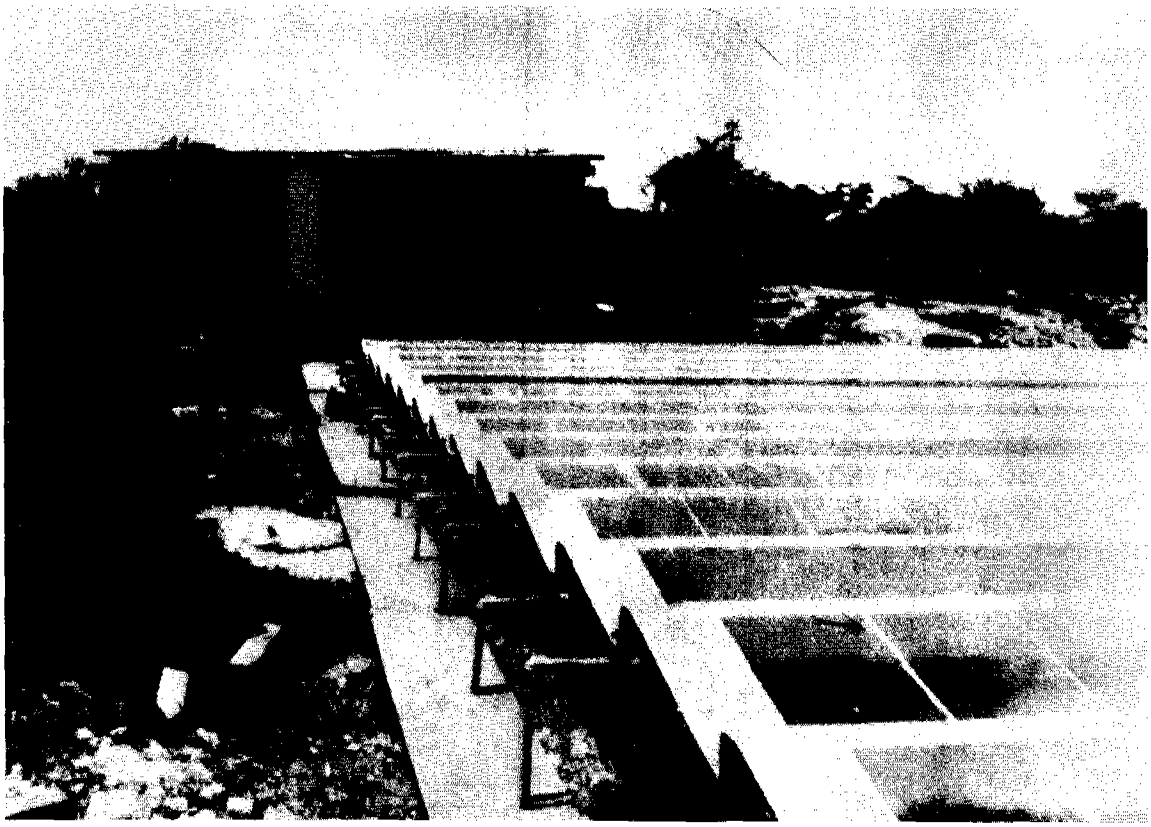
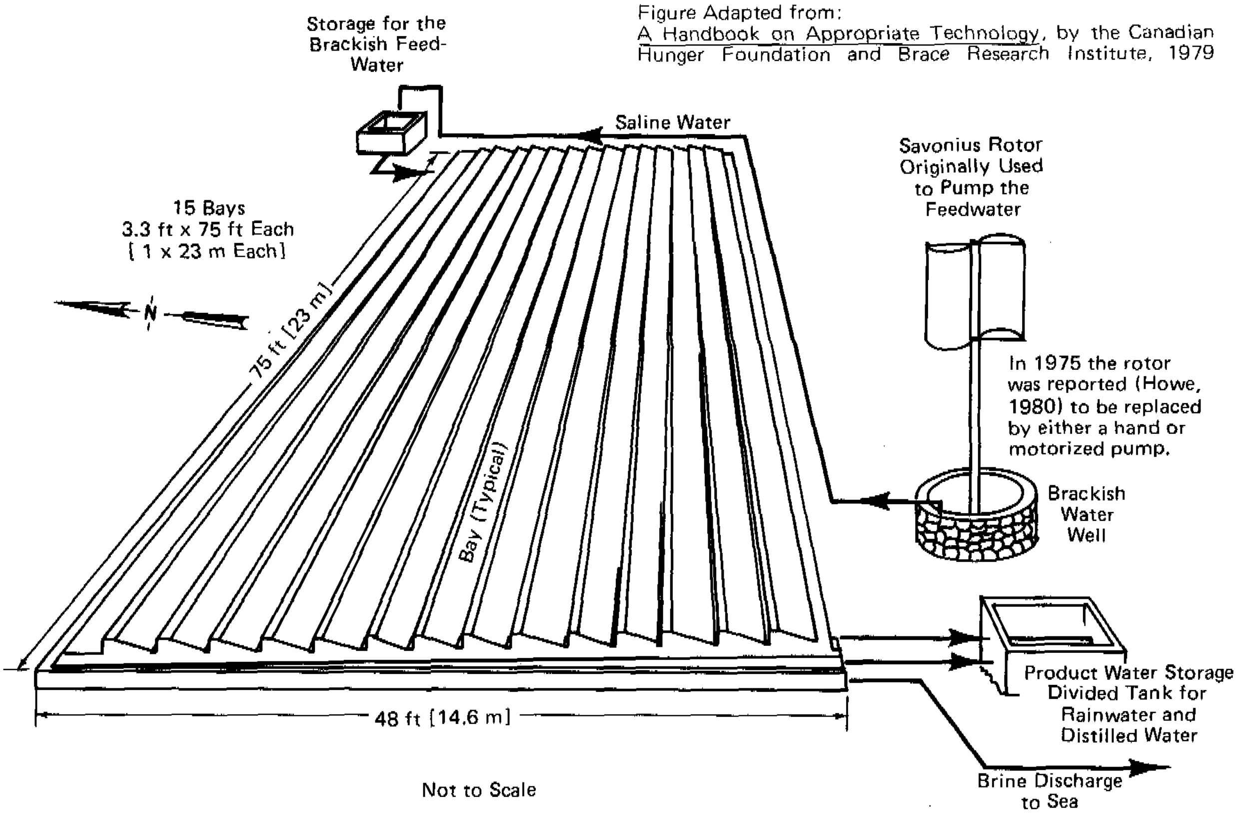


Photo Courtesy of Brace Research Institute, Canada.

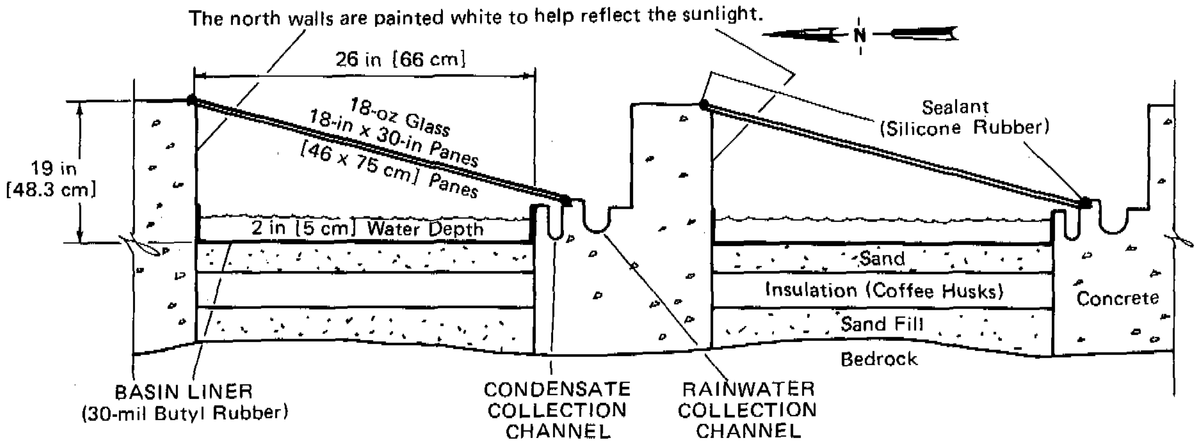
FIGURE 7-12. The solar still at La Gonâve, Haiti.

Figure Adapted from:
A Handbook on Appropriate Technology, by the Canadian Hunger Foundation and Brace Research Institute, 1979



GENERAL LAYOUT

The still is built on a gently sloping hill with the southern bays (shown to the right) lower than the northern bays. Gravity feed is used from the brackish water storage tank.



CROSS SECTION OF BAYS

Notes:
1. The floors of the bays slope at 1:75.
2. Both concrete block and poured concrete were used and finished with plaster.
3. Cross section drawing adapted from Talbert et al., 1970.

FIGURE 7-13. Layout and cross section of the solar still at La Gonâve.

Photo Courtesy of O. K. Buros.

An installation of several small stills. Note the collection piping below the stills. A Savonius windmill is seen in the background.

FIGURE 7-14. Modular solar stills used by DIGAASES in Mexico.

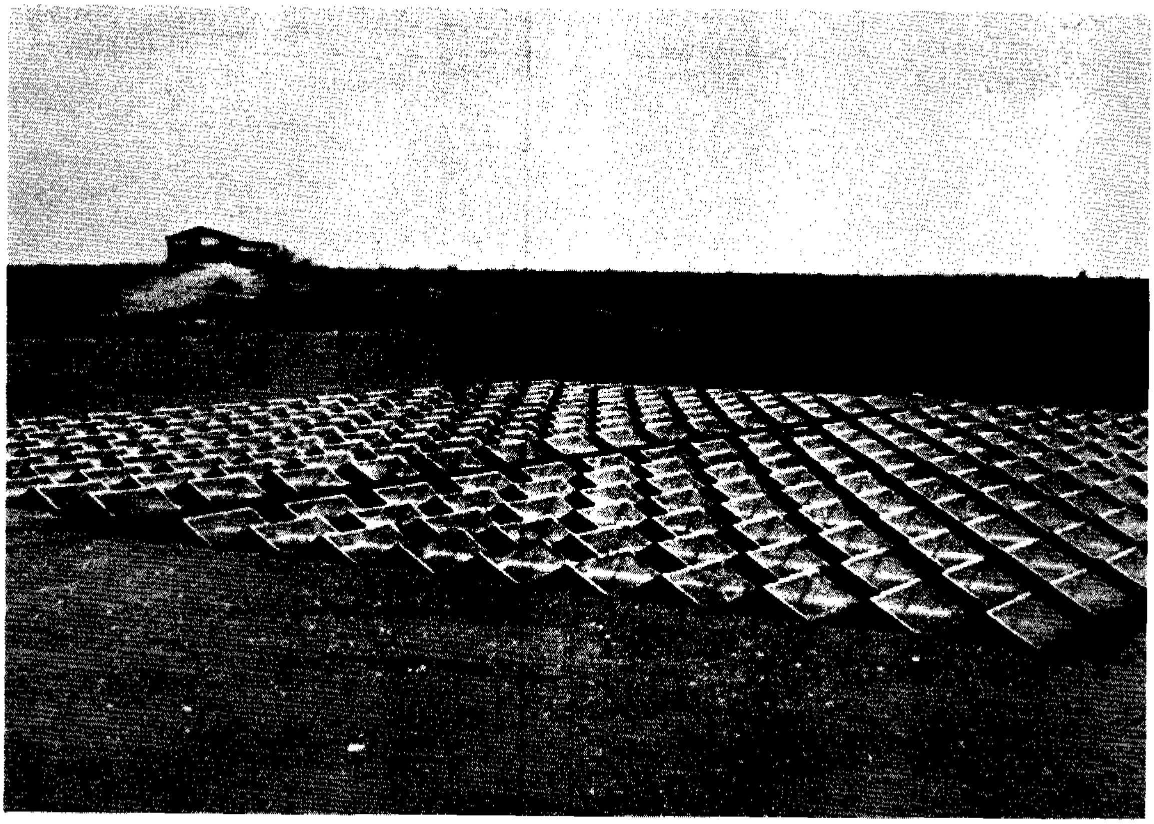
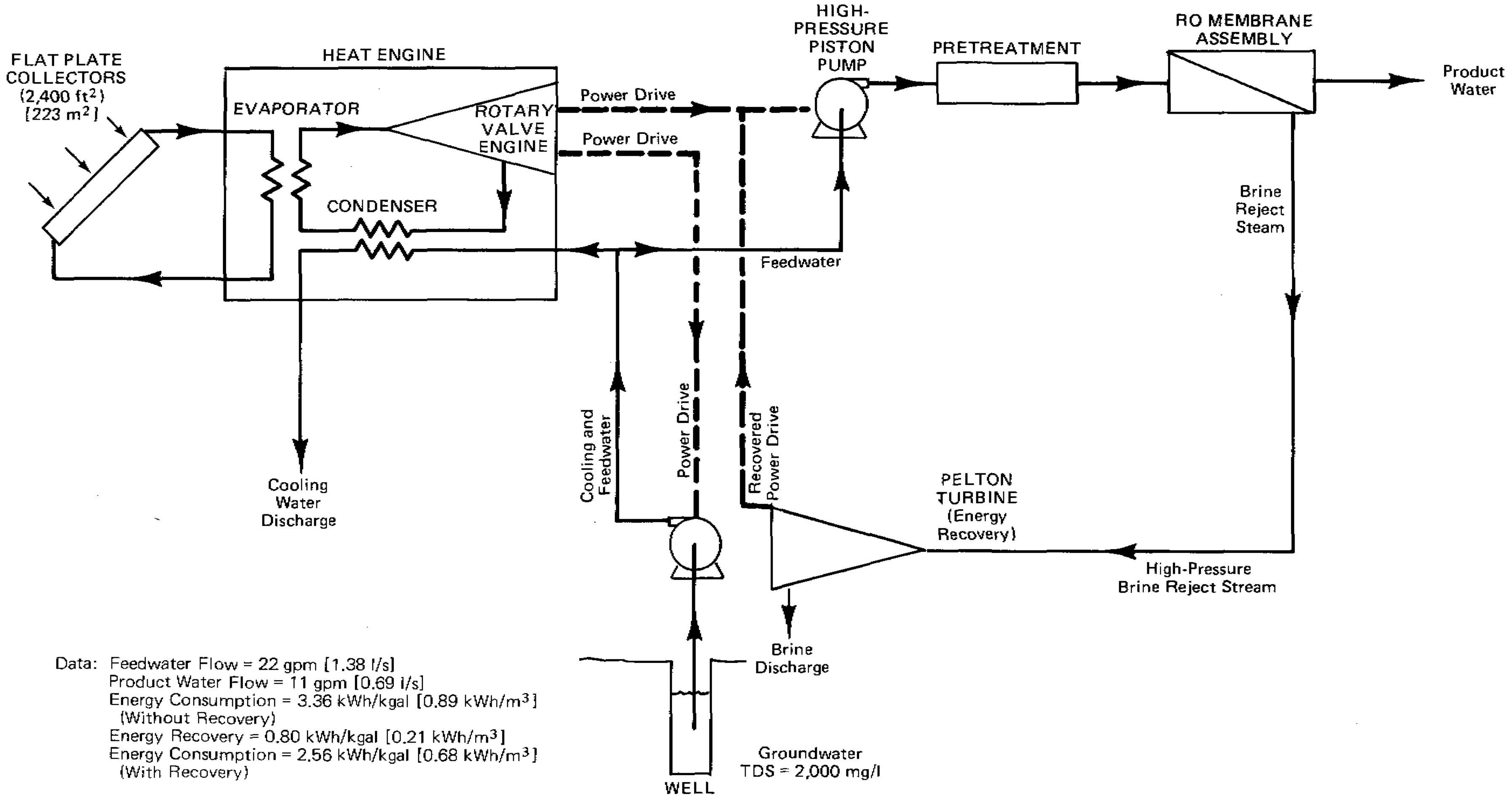


Photo Courtesy of DIGAASES, Mexico.

FIGURE 7-15. The solar still at Puerto Chale, Mexico.



Data: Feedwater Flow = 22 gpm [1.38 l/s]
Product Water Flow = 11 gpm [0.69 l/s]
Energy Consumption = 3.36 kWh/kgal [0.89 kWh/m³]
(Without Recovery)
Energy Recovery = 0.80 kWh/kgal [0.21 kWh/m³]
Energy Consumption = 2.56 kWh/kgal [0.68 kWh/m³]
(With Recovery)

Drawing and Data Adapted from Maurel, 1979a.

FIGURE 7-16. Flow diagram of the experimental reverse osmosis unit at Cadarache, France.

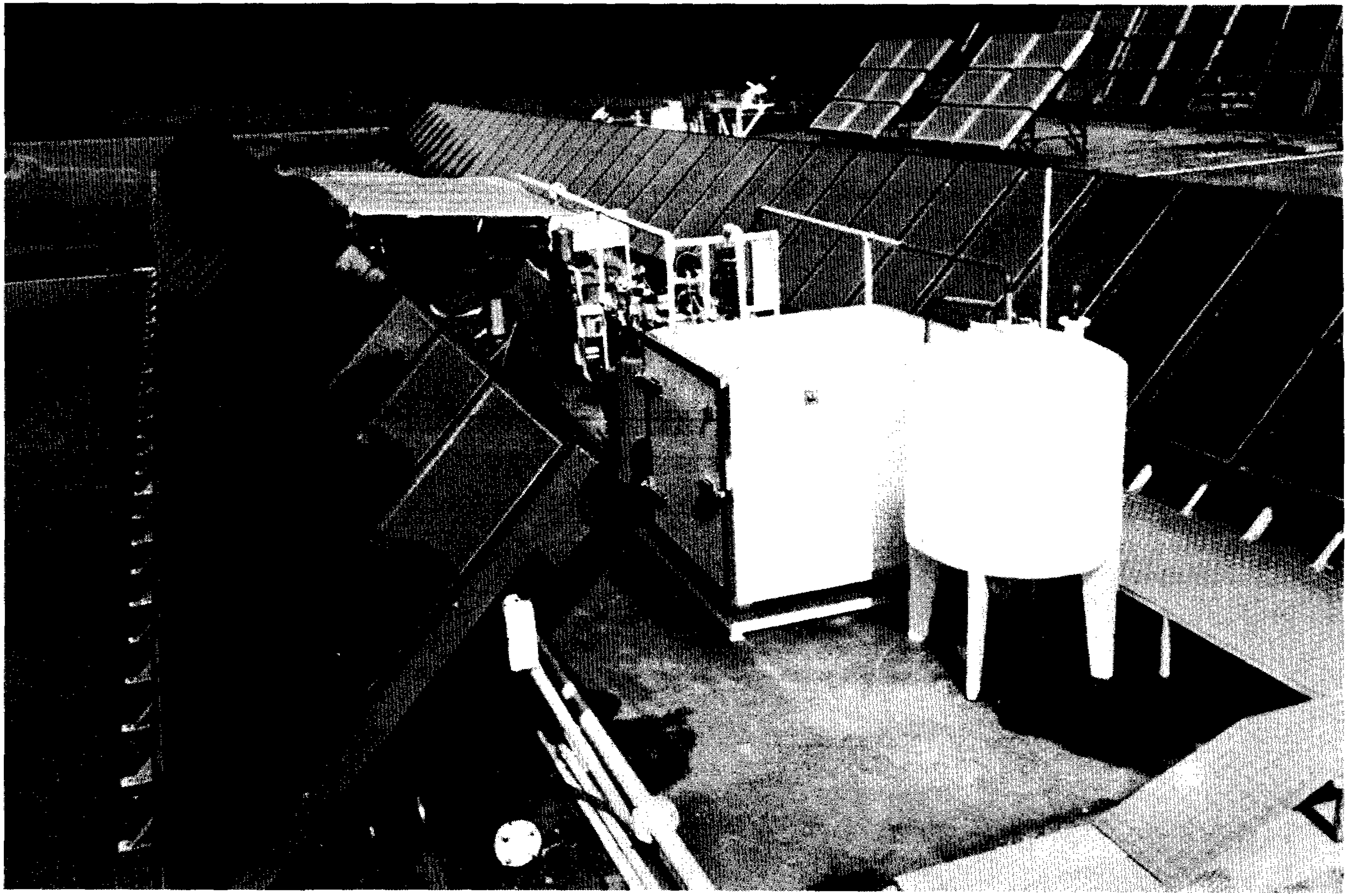
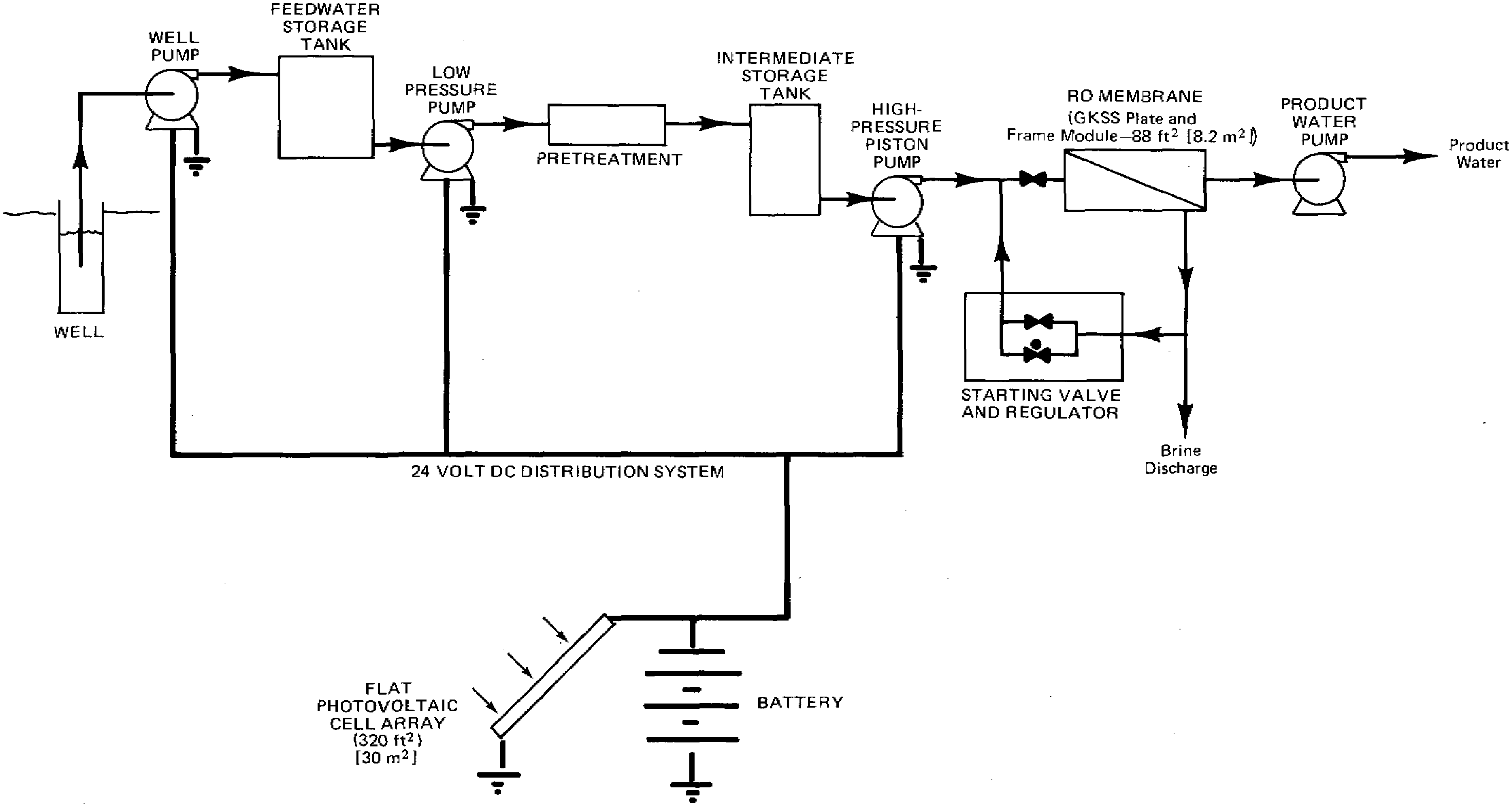


Photo Courtesy of the Commissariat a l'Energie Atomique (CEA) Cadarache, France.

FIGURE 7-17. The solar reverse osmosis unit at Cadarache, France.



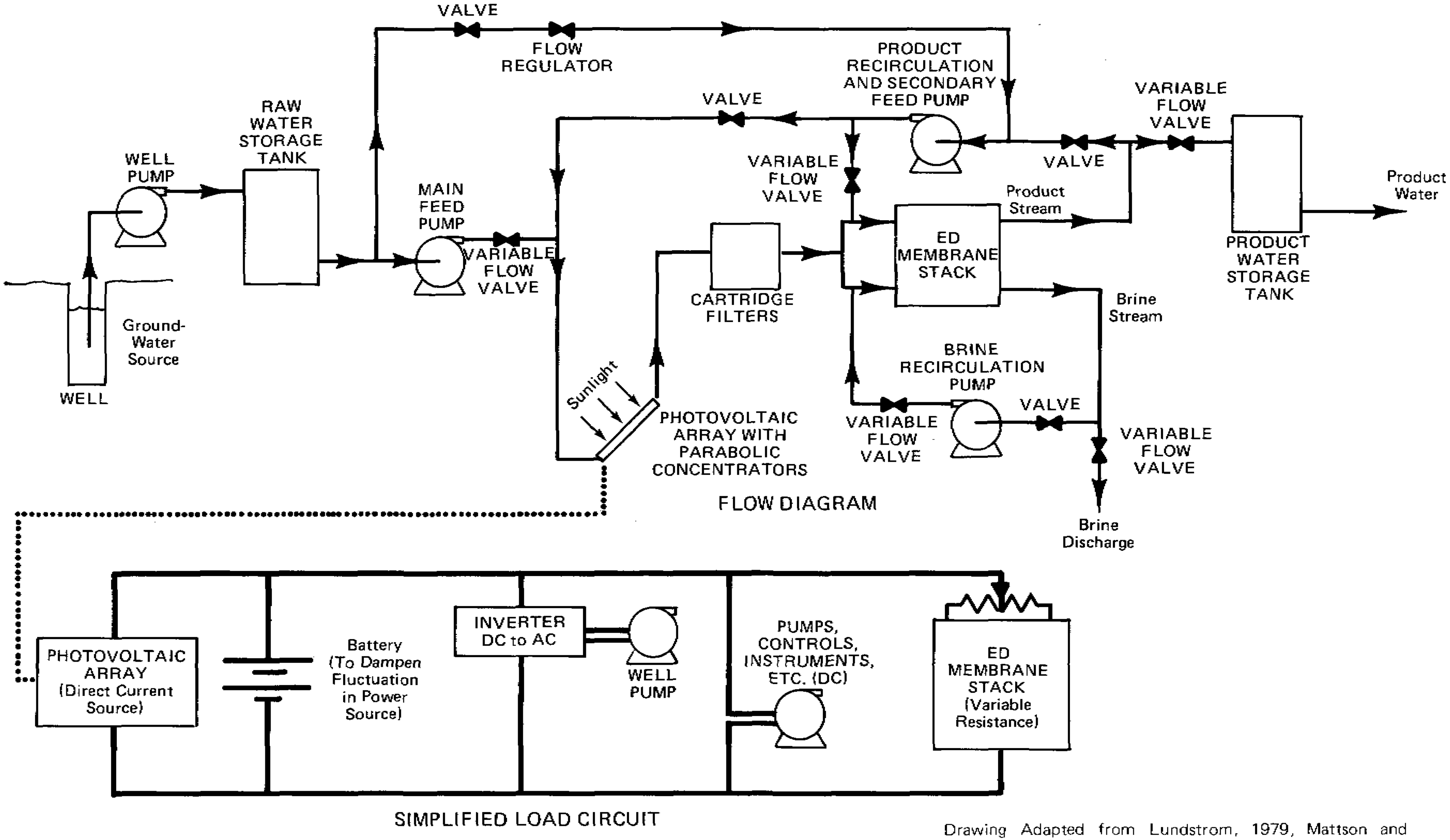
Drawing Adapted from Peterson et al., 1979.

FIGURE 7-18. Flow diagram of the solar reverse osmosis unit at Concepcion del Oro, Mexico.

Photo Courtesy of GKSS, Federal Republic of Germany.

Photos Courtesy of GKSS, Federal Republic of Germany.

FIGURE 7-19. Solar reverse osmosis unit at Concepcion del Oro, Mexico.

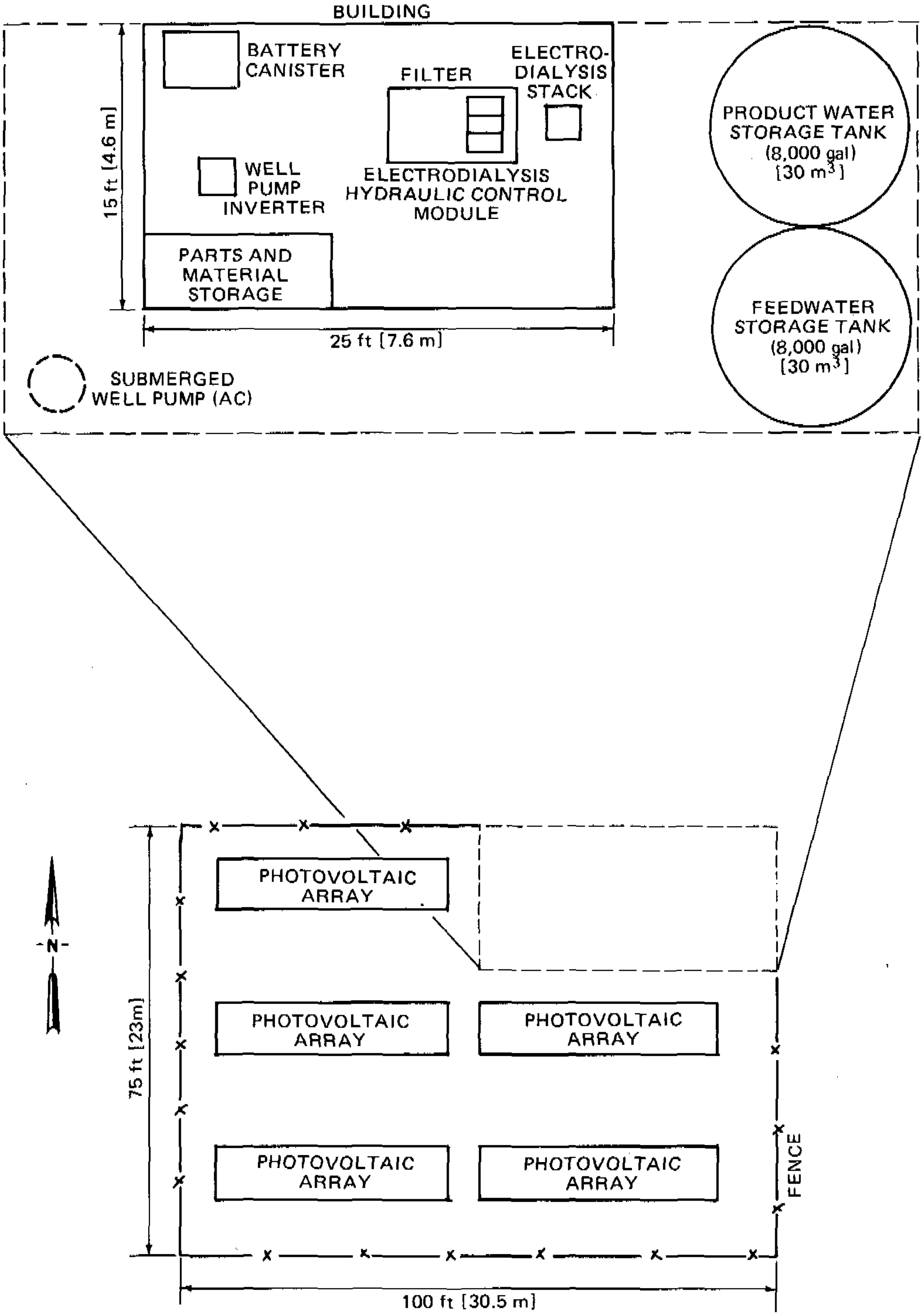


FLOW DIAGRAM

SIMPLIFIED LOAD CIRCUIT

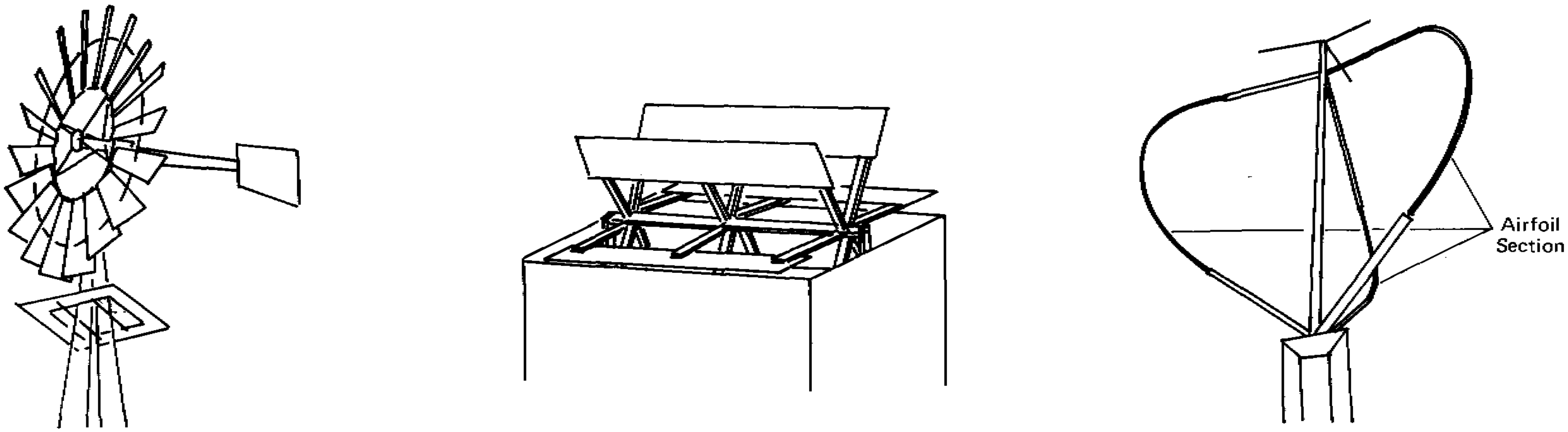
Drawing Adapted from Lundstrom, 1979, Mattson and Lundstrom, 1979.

FIGURE 7-20. Flow diagram and load circuit for a solar-powered electrodialysis unit.



Drawing Adapted from Mattson and Lundstrom, 1979.

FIGURE 7-21. Proposed site plan for a solar-powered electrodialysis unit.



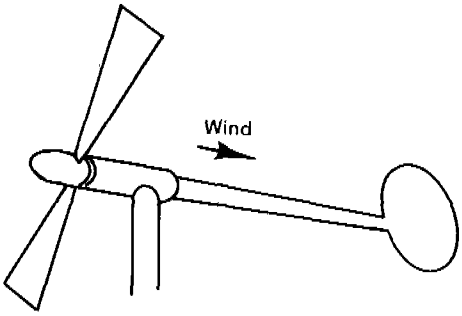
TYPE 1
Horizontal Axis
Inline with
the Wind

TYPE 2
Horizontal Axis at
Right Angles
to the Wind

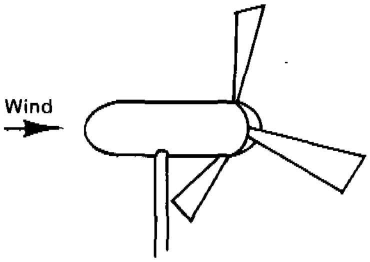
TYPE 3
Vertical Axis at
Right Angles to
the Wind

FIGURE 7-22. Examples of three basic types of wind machines.

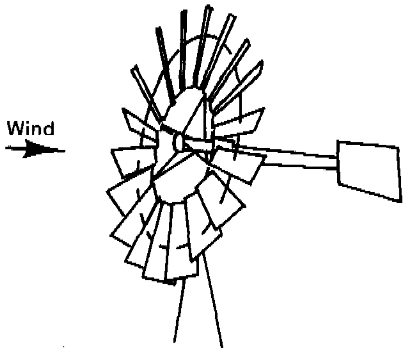
PROPELLER, UPWIND
Horizontal Axis with
2 or 3 Blades
Usually with Tail



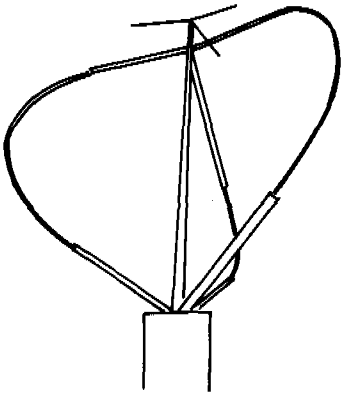
PROPELLER, DOWNWIND
Horizontal Axis with
2 or 3 Blades,
No Tail



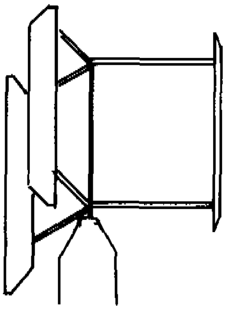
BICYCLE TURBINE
Horizontal Axis
Multibladed with Tail



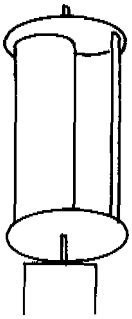
DARRIEUS ROTOR
Vertical Axis with
Semiflexible Curved
Blades



DARRIEUS ROTOR
Vertical Axis with
Rigid, Straight Blades



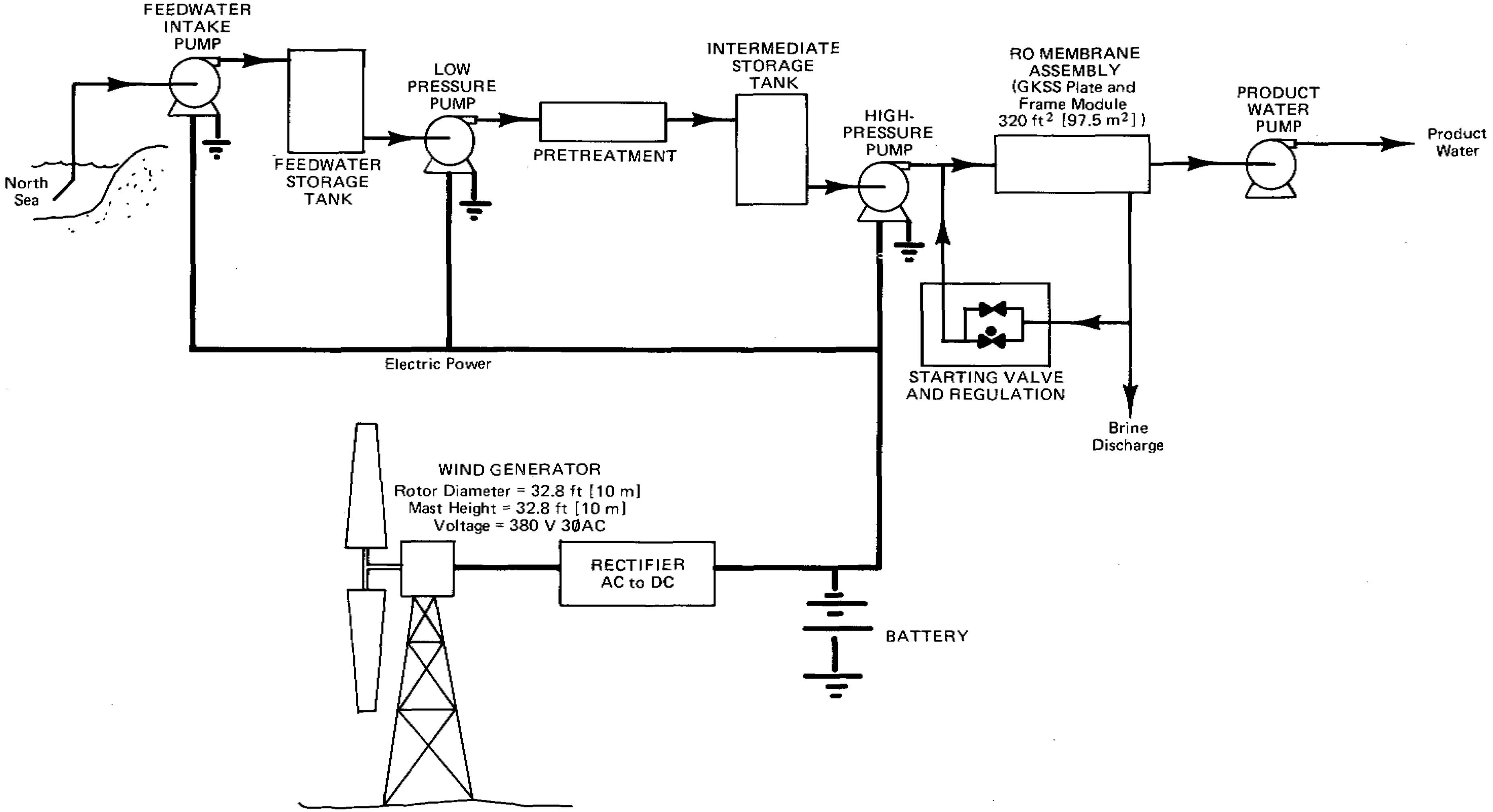
SAVONIUS ROTOR
Vertical axis



UNIDIRECTIONAL
WIND MACHINES
Machines must orient
themselves to the wind direction.
(Type 1)

MULTIDIRECTIONAL
WIND MACHINES
Machines can utilize
wind from any direction.
(Type 3)

FIGURE 7-23. Common types of wind machines.



Drawings Adapted from Petersen et al., 1979.

FIGURE 7-24. Flow diagram of the wind-powered reverse osmosis unit for an island in the North Sea.

Photo Courtesy of GKSS, Federal Republic of Germany.

Three-Bladed Horizontal-Axis Wind Turbine

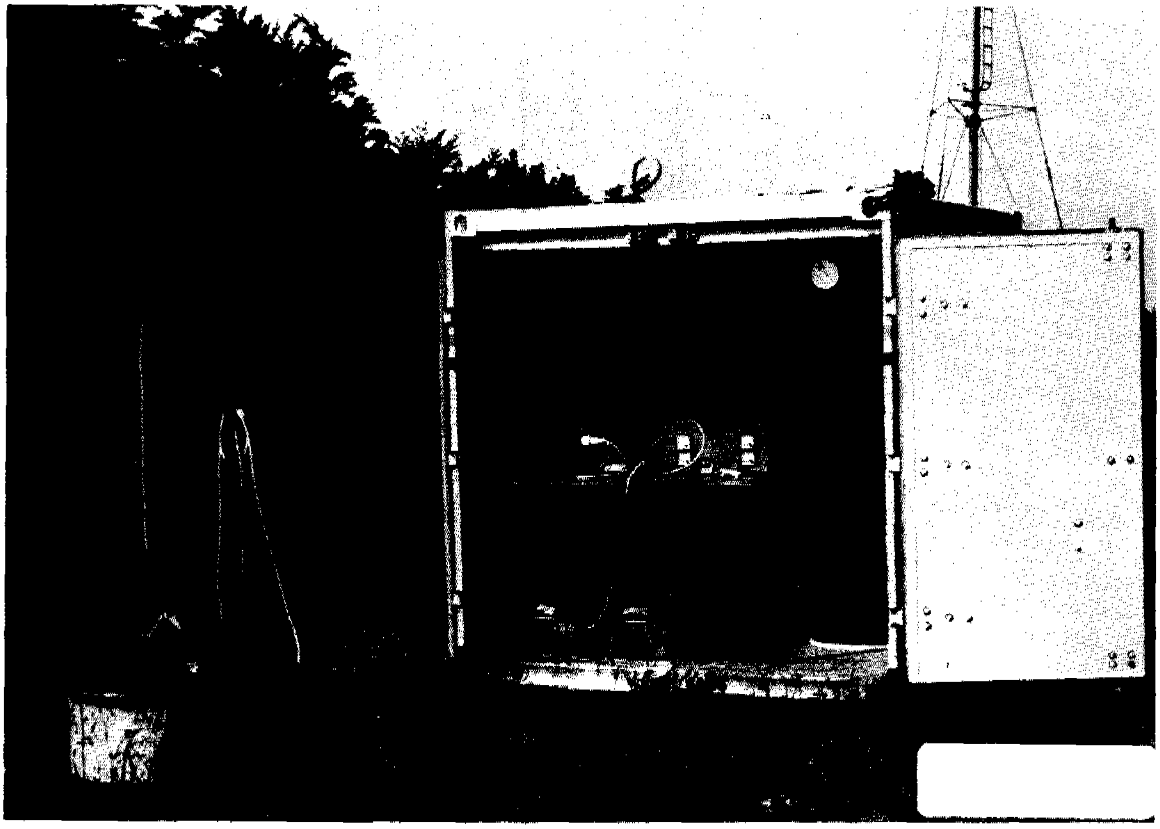
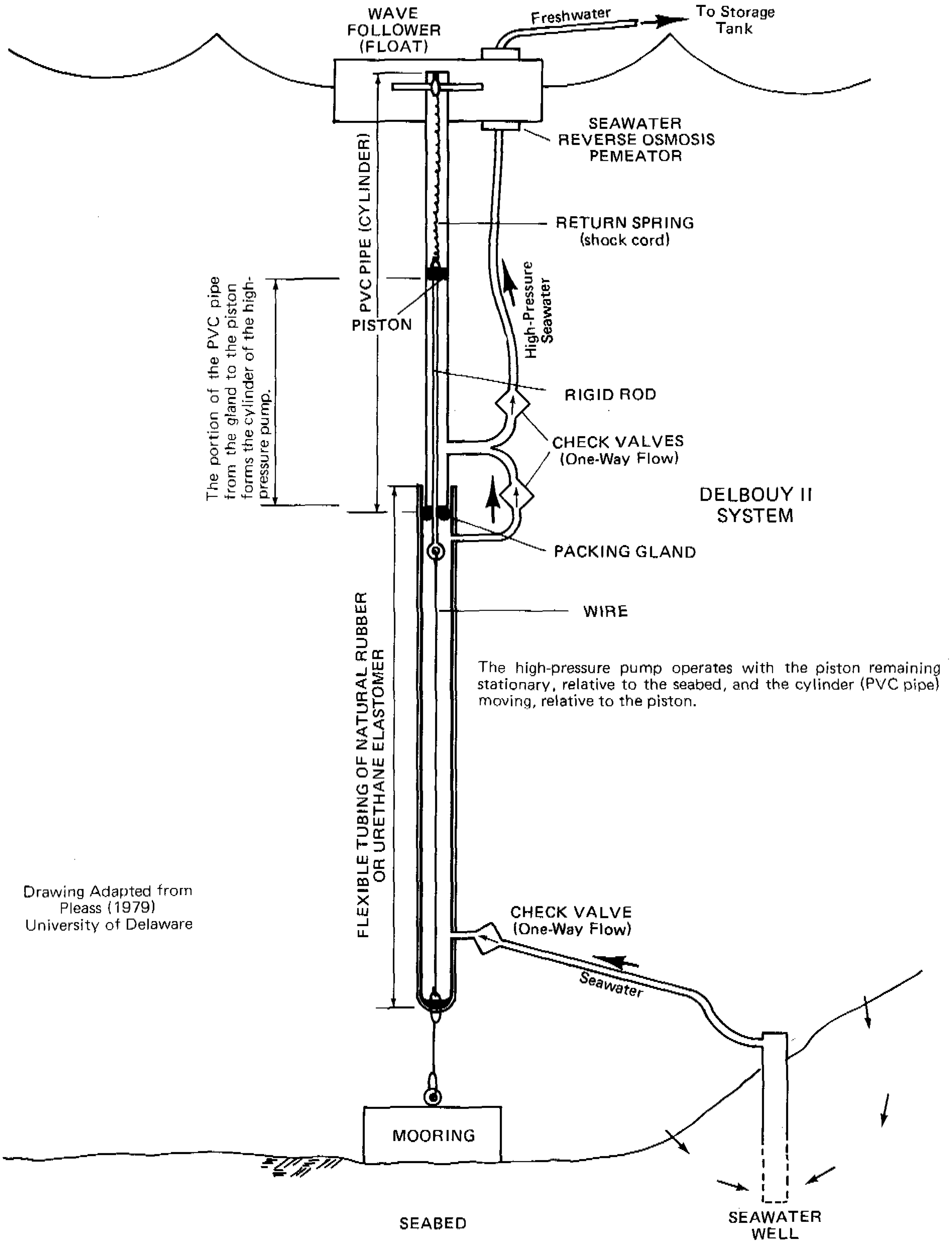


Photo Courtesy of GKSS, Federal Republic of Germany.

Test Unit and Instruments Used with the Wind Turbine

FIGURE 7-25. Wind-powered reverse osmosis system for an island in the North Sea.



The portion of the PVC pipe from the gland to the piston forms the cylinder of the high-pressure pump.

The high-pressure pump operates with the piston remaining stationary, relative to the seabed, and the cylinder (PVC pipe) moving, relative to the piston.

Drawing Adapted from
Pleass (1979)
University of Delaware

FIGURE 7-26. Generalized diagram of a wave-powered seawater reverse osmosis unit.

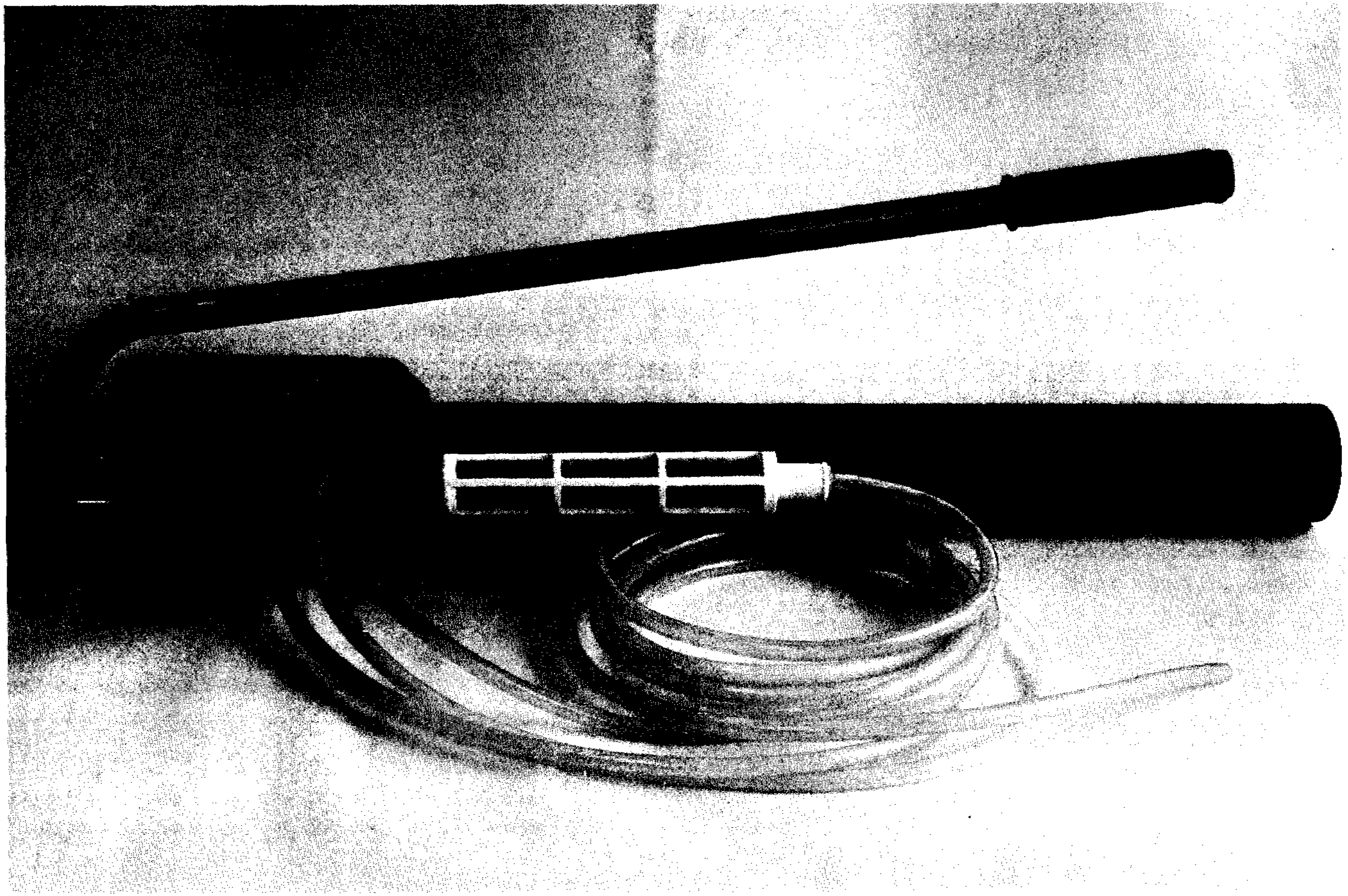


Photo Courtesy of Seagold Industries, Canada.

FIGURE 7-27. A hand-powered reverse osmosis unit.

CHAPTER 8

Chapter 8
PROCESS SELECTION

In any country, no matter what the level of development, the best selection of a desalination or any water resource system, is one which is more than just economically reasonable in the paper study stage. It is a system which works when it is installed and continues to work and deliver a desired product at the expected quality and quantity for the planned life of the system.

WATER RESOURCE EVALUATION

The desalination of water is an excellent water resource tool to enable an area to utilize water sources which in the past were too saline for potable usage. However, desalination is not an inexpensive process. A substantial capital investment is required and the operation of the system will continue to require expense for energy, chemicals, labor, repairs, replacements, etc. Thus it is important, and prudent, to determine whether desalination is needed in the first place and, if it is, to what extent it should be used.

Potential costs must be kept in mind, because they can affect consumption and the local economy. The curves on Figures 8-1 and 8-2 show estimated capital and operating costs for various desalination processes in the USA. Rough costs of $1 to $4/kgal [$0.26 to $1.06/m^3] for brackish water and $5 to $30/kgal [$1.32 to $7.93/m^3] for seawater usually bracket the overall unit costs (capital recovery and O&M) of desalination. Items such as special feedwater development (well fields, intake structures, etc.), water storage, and brine disposal can add another $1 to $5/kgal [$0.26 to $1.32/m^3] to this cost.

Whether or not consumers are asked to pay the full price is a local decision, but the cost must be paid, either by the consumers, the government, or someone else, requiring that monetary resources be diverted from other endeavors in the economy and used for this purpose for the life of the project.

Investment in a water-short area could lead to the development for the first time of an abundant supply of water for an area and, as such, it is important that the development be carried out carefully. The key to the success of any major water resource development of this type is proper water management based on a thorough water resource study and the formulation of a realistic development plan. Proper water management includes the entire spectrum of water resource activities ranging from source selection and development through water usage, to the ultimate disposal of the resultant wastewater.

Generally, in carrying out a water resource study, the existing and potential sources are examined from an economic and technical standpoint and analyzed along with the probable uses for the water. Based on the results, a plan for water resource development within the context of available water and financial resources is formulated. These water resource plans are usually very site-specific. They must be keyed to the economic development of the selected area, while exploiting the natural advantages and compensating for the locality's problem areas. It is unlikely that development plans for two different areas would be identical, due to the many factors involved (Buros, 1976b).

The most effective solution in many water-short areas may consist of the use of a variety of sources and methods of distribution and hence may not necessarily resemble the type of system common to cities in the USA. The following paragraphs outline some of the general principles to be considered in conducting an evaluation of this type.

Water Resources Available

The existing and potential water resources available, including all sources of fresh and saline water, need to be inventoried and evaluated. The quantity, quality, cost of development, and consequences of development and/or diversion should be determined for all potential sources.

Demand

In some water-short areas, it may be economical to develop a system with two or more grades of water distributed separately, providing distribution and usage of both potable and non-potable water. To analyze this potential, the demand for both potable and non-potable usage must be projected. The demand will depend on many factors including climate, culture, distribution methods, cost, uses (including agricultural, industrial, and public sector usages), accessibility, and availability.

Demand must be estimated carefully, because the potential added cost of desalted water (if passed on to the consumer) can greatly affect water usage. Not only can this water be expensive for domestic consumption, but high costs can affect the type of industries attracted and the economic well-being of industries using the water. Obviously, industries requiring large quantities of low-cost water cannot readily survive if forced to pay the true cost of desalted water, especially water desalinated from seawater. Since demand will be affected by the perceived cost to the consumer, the policy which will be used to charge consumers for water should be ascertained early.

Water Supply Alternatives

In studying water-short areas, the development of alternative and/or supplemental sources should be carefully examined. These could include: trucking, piping, or barging of water; reuse of water (wastewater reclamation); rainwater catchments; modification of existing runoff and recharge characteristics; improved development of ground- and surface-water sources and storage; etc.

Caution should be exercised if planning suggests the complete abandonment of other existing sources of water and full reliance on desalinated water unless a very careful study has indicated that this is, in the long-term, the best method for the particular locality.

System Integration

A desalination facility alone does not constitute a water system. Some method must be provided to distribute the water. Distribution can be as basic as hand-carrying water obtained from a tap on a storage tank or as complex as a full piped-in distribution system. The manner of distribution should be known or determined at the outset of the project.

Where the distribution will take place, using an existing piped system, it would be prudent to determine how much of the water entering the distribution system is actually delivered to a consumer. A poorly designed, constructed and/or maintained distribution system could lose a substantial portion of the water entering the system. Some distribution systems in the Caribbean employing desalinated water as a major source cannot account for 40 to 50 percent of the water in the system. This is due to a combination of leakage, unauthorized connections, and broken meters, and it results in lower revenues and increase unit costs.

A part of any distribution system should include sufficient storage to meet peak demand and/or permit shutdown of the desalination facility to permit adequate planned maintenance and unplanned repairs.

FACTORS RELATED TO PROCESS SELECTION

The decision to use desalination, and the selection of the appropriate process and its capacity, is one which should begin with an evaluation of the overall water resource picture (as discussed in the preceding paragraphs). The factors discussed below should then be examined and weighed. There is no set formula to arrive at the correct decision. As with any project that contains capital and operating expense, an economic evaluation can play a significant role but it should not be the only factor. Judgement based on a thorough review of the various factors, both objectively and subjectively, with a view to the long-term costs, benefits, and consequences of a proposed project, must be exercised in making many of the final decisions.

Brine Disposal

One factor to remember in all desalination processes is that three flows are involved: (1) an input of feedwater which is to be processed; (2) a product-stream with a TDS level lower than that of the feedwater; and (3) a brine stream with a TDS level higher than that of the feedwater. Provisions must be made for brine disposal in a fashion which will not cause any unreasonably adverse effects. For a seawater or brackish water unit on the seacoast, this is not usually a problem, as the brine can be discharged into the sea, but in inland locations, brine disposal can be a significant and very expensive problem. The economic evaluation of any desalination system is not complete without including the costs involved in brine disposal. When brine disposal is a serious problem, it has the potential to affect the ultimate process selection and often the viability of the project.

The most common method considered for safely disposing of brine from inland plants is to use evaporation ponds, which are costly and can take up considerable space. The cost of the ponds needs to be weighed against the expense of using and operating a high-recovery desalination system, which will minimize the quantity of brine produced (but not the total amount of salt) and the size of the ultimate disposal system. There has not been a great deal of long-term experience in safely disposing of brines from inland desalination facilities, and this should be kept in mind during planning. No project should proceed unless an acceptable method of brine disposal has been determined.

Raw Water Source

Identifying and characterizing the potential raw water source (or sources) is a basic part of process selection. The quantity and quality of a water source can change daily, seasonally, etc., and knowledge of whether this does occur, and the extent of variations, is important in process selection, design, and long-term operation of the facility.

Unfortunately, many desalination projects consist of development of the feedwater source and installation of a desalting unit simultaneously under the same or concurrent contracts. In many cases, it would be far better to develop the feedwater source first so as to clearly characterize it, and then move on to the final selection and/or design of the desalination facility, its procurement, and installation. This may not change the process selected, but it may modify the pretreatment and/or capacity required. Finalizing these changes before the original bidding is almost always cheaper than trying to make alterations and change orders in contracts at a later date. It has not been unknown in the past for desalination plants to have been purchased and installed only to discover that the source of feedwater was inadequate and/or unsuitable.

Product Water

Both the quantity and quality of the water to be produced by desalination need to be determined. The quantity needed and plant production capacity will depend on: whether the water is to be used to augment other water sources; the consumption demands on the system; the amount of storage in the system; etc.

Careful thought should go into specifying the quality of the water that is desired from the desalination facility. Automatically assuming a TDS content of up to 500 mg/l might not be appropriate. In order to specify the proper TDS level, many factors should be considered, including (among others) the following: the degree of blending with other water supplies, the taste preference of the populace using the water, the type of post-treatment to be used, the potential for wastewater reuse, and special uses for the water, such as for industries, agriculture, etc.

Based on currently available supplies, societies have adapted their tastes, food preparation, and digestive balance to the mineral content of the water to which they are accustomed. When the taste of water changes, displeasure and perhaps rejection of the water can follow. The taste of foods and beverages, especially tea, can be sensitive to changes in the constituents in water, and this should be considered in the initial planning to ensure that over- or under-demineralization will not cause social acceptance problems nor needless expense.

In an area where wastewater reuse is being practiced or considered for ground-water recharge and/or agricultural purposes, thought should be given to the level of TDS desired in the wastewater effluent. In cases where the resultant wastewater is to be treated and reused for ground-water recharge, benefits may be derived when distillation (instead of reverse osmosis) is used for seawater desalination to produce potable water due to its low TDS (Buros, 1980). However, a caution should be noted, as low-TDS water, if not subjected to suitable post-treatment, can be very corrosive and damaging to a water distribution system. There are also some concerns as to the desirability of using distilled water for potable purposes over a long time period, although this question has not been settled conclusively. Some plants blend the distilled water with a mineralized water to increase the overall TDS or the distributed water.

Operating Materials, Spare Parts, and Power

The availability, quality, and cost of power and materials necessary for operation of a desalination facility are also important considerations. These must be determined on a site-specific basis to evaluate the status of power, chemicals, spare parts, outside technical expertise, etc., for future operations. Site location, transportation facilities, taxes, customs duties and handling, climate, and local demand all affect the cost and availability of fuel, chemicals, and spare parts.

The need for locally available components is not confined to the specialized desalination equipment, but also covers fuses, bolts, screws, bearings, pumps, electrical systems, tools, etc. This can be something as basic as metric or English parts, threads, tools, etc., being used where the other system is usually found. If these are not available locally, either large inventories must be financed or long downtimes can be anticipated while even the most common parts are procured.

The source of energy is important as well. In some areas, there may be considerable merit in building a dual-purpose (electricity and water) plant as part of a national or local development program. In other areas, waste heat may be available for a low-temperature distillation process. Where existing steam and/or electrical power is to be used, the type, cost, adequacy, and reliability of these should be determined. For larger facilities that are to produce water only, there are some interesting dual-purpose combinations that could be considered. An example of this for seawater desalination might involve using a high-temperature boiler to produce steam. This steam could operate a turbine which directly runs a reverse osmosis unit while the turbine's back-pressure or extraction steam is used to operate a multiple-effect plant. A variety of these types of combinations is now being explored.

Financial Support

The financing of a desalination project is sometimes split between two agencies or organizations where one, such as U.S. A.I.D., purchases the unit and the other, for example an overseas government agency, operates it. In such cases, it behooves the purchaser to realistically assess the future financial support of the operating agency for the desalination facility before finally selecting the type of process and capacity of the unit.

An analysis of the operating agency and its proposed operating budget is necessary to estimate the level of technical supervision, spare parts, operating supplies, etc., to be allocated in the years to come and to select the plant wisely. Even the agency's philosophy on stocking spare parts and its procedures for procurring parts, technical services, and/or repairs (which could be time-consuming) should be considered in plant selection and specification.

Technical Skills

Where possible, the technical skill required for the operation and maintenance of a plant should be matched to that available (within the future operating budget) in the locality of the proposed plantsite. The presence of local personnel experienced in desalination plants (provided that the experience was positive) is a factor in favor of using desalination where technical experience is scarce.

Experience with boilers, steel work, steam, welding, etc., is useful with distillation plants, whereas the membrane processes usually require skills with plumbing, pumps, and electric motors.

Proper training for operating personnel is a necessity, but training will need to be repeated in the future as personnel are replaced or need additional emphasis in certain skills. A long-term training program should be planned and included in any acquisition of new equipment.

Site Conditions

The size, topography, and cost of suitable sites can be relevant factors, especially in urban areas. The land uses adjacent to the site can have an effect on a desalination facility and can, in turn, be affected by it.

Efficiency and Reliability

The desalination industry is experiencing a continuous trend to increase the efficiency of the various desalination processes in order to reduce operating costs. Examples of this trend include distillation plants with designs for higher performance factors and reverse osmosis facilities which are adding energy recovery units.

Higher efficiency plants, however, tend to be somewhat more complex in design and operation, thus requiring extra care in operation and maintenance. This extra care is not impossible to exercise, and many facilities exercise it without problems. However, if this extra care is overlooked, serious operational problems could result which would nullify any expected benefits.

The reliability of plant operation should be of paramount concern because, regardless of a plant's efficiency when it is operating, it is zero when the plant is idle. Reliability in developing areas is generally increased by proper design, good materials, and simplicity of operation.

Reliability becomes even more critical when an area depends almost wholly on the facility for its water supply. In such cases, alternative sources are usually difficult to obtain and very expensive.

A good indicator of plant reliability is the performance of plants of similar process and manufacture that have operated under similar circumstances. Their performance, capabilities, and problems can best be assessed by (1) actually seeing the plants in operation and examining operating data and repair records and (2) discussing the facilities' operation with operating personnel and manufacturers' representatives.

Table D-1, "Summary of Guidelines for Desalting Process Selection," in Appendix D contains many of the important characteristics of the four major desalination processes. This table is useful for comparing various methods of desalination during process selection and for providing information to be used in making and/or evaluating cost estimates.

PROCESS ECONOMICS

Information on the economics of various desalination processes can be of value to U.S. A.I.D. in two major areas. The first is in initially narrowing down the process options available while investigating the viability of a proposed project. The second is in reviewing proposals submitted by other agencies, vendors, consultants, etc.

A Desalination Cost Summary Sheet is shown on Figure 8-3, and some basic cost estimating material has been included in Appendix C--Cost Estimation of Desalination Processes. Whenever the cost estimating sheet and material are used, the Desalination Project Information Form in Appendix C should also be completed. These sheets will aid in the calculations and indicate data and/or information which are missing or need further investigation.

The data contained in the Appendix are intended for order-of-magnitude estimates only. They can be used to obtain a general idea of the expenses involved and to roughly compare process costs. However, the complexities of site-specific problems and market changes make the job of cost estimating for funding and specific planning one which is best undertaken by a consultant or someone else experienced in the area.

Cost predictions for desalination processes should be approached with considerable care and a realization that there are pitfalls involved. A number of studies sponsored by various agencies of the U.S. Government can be of assistance. Since 1975, the Oak Ridge National Laboratory (ORNL) has published three reports on desalination costs, (Reed 1975; Reed and Wilson, 1977; and Larson 1979). A report by another group, Fluor, Inc., (1978), brought together cost estimates from several U.S. Office of Water Research and Technology (OWRT) state-of-the-art reports as well as information from earlier ORNL reports. In 1980, the OWRT completed a new edition of the Desalting Handbook for Planners (Catalytic, Inc., 1979) which updated (to 1979 dollars) some of the Fluor costs and added cost curves for many additional items needed to estimate desalination project costs, such as: special pretreatment processes, brine evaporation ponds, cooling towers, injection wells, pipelines, and storage tanks.

In addition, the volume prepared by Catalytic, Inc., devotes about 120 pages to process selection and cost estimating, including example problems. It includes worksheets for cost estimating, computations, and supporting data. This planning handbook would be a useful companion document to this manual, since it contains considerable complementary information. This volume can be obtained from OWRT or the U.S. Department of Commerce, National Technical Information Service (NTIS), Springfield, Virginia 22161.

Included in Appendix C of this manual is a selection of essential material from the planning handbook, plus additional information, tables, and graphs to assist in estimating and tailoring some of the operating costs to local situations. Some of these additional graphs contain data on factory prices for smaller, skid-mounted units for desalination by reverse osmosis, vapor compression, electrodialysis, and heat recovery.

Appendix C should be used in conjunction with the remainder of the manual, which contains specific information on the processes and selected data on equipment specifications. There is sufficient information in this manual to make planning level estimates of the cost of desalination facilities but it should be remembered that cost predictions and comparisons are highly complex and cannot simply be gleaned from numbers on charts or curves, such as those contained in Appendix C and the reports mentioned previously. Although these are reasonable approximations, they are based on cost data supplied by vendors and other organizations and are predicated upon specific assumptions and conditions which are not always mentioned.

As these assumptions and conditions change, as they can when moving from the case in the report to a case under study, the costs change. The greatest difficulty and skill are involved in deciding what assumptions and conditions are appropriate for specific local conditions and how they will affect costs and operation. Each site represents unique problems of topography, climate, transportation, feedwater, brine disposal, power, labor, and availability of materials and equipment. In addition, each project requires consideration of schedules, risks, customs, taxes, delays, performance guarantees, spare parts, training, and numerous other factors. It is most important that competent, experienced, professional help be used in evaluating cost estimates to make them pertinent to the application being considered.

Aside from defining conditions and assumptions inherent in predicting final costs, it is necessary to remember that market forces affect the final bid price for equipment and/or construction. Among the factors which can affect the bid price, besides actual costs, are the desire to obtain the project, the financial climate, and the evaluation of what the project entails. An example of the variation in bid costs that can occur in desalination projects was the spread of 370 percent between the high and low bids (16 bidders) for a 0.3-mgd [1,140-m^3/d] reverse osmosis project bid during 1979 in Bermuda (Smith, 1980).

Along with Appendix C the large table in Appendix D on Process Selection Guidelines will be useful in providing guidance in evaluating particular processes, making a process selection, and/or formulating cost estimates. The table covers characteristics of 18 different types of desalination processes.

Cost Estimation

One effective way to compare cost estimates for desalination facilities is to break down the estimates in a form similar to that used in other desalination cost summaries. Such summaries are readily available, and they provide suitable data to allow for comparison of assumptions and/or prices.

The ORNL and Fluor reports, and Catalytic handbook can be helpful in this regard. The cost summary sheet (Figure 8-3) basically follows the format of similar tabulations found in these works and can be readily used for comparison purposes. When making a cost estimate, four steps are suggested. The first is to obtain information for definable capital and operational equipment or material costs, such as the desalting equipment, switch gear, intakes, membranes, etc. Estimates for these items can be made using the information in Appendix C, and/or other sources. The second step is to make assumptions, based on as much site-specific information as possible, on the costs for local items, such as power, chemicals, labor, etc. Information in Appendices C and D will assist in identifying and roughly quantifying the major items needed. In addition to this, a factor must be developed that can be applied to the prices in Step 1 to make them reasonably valid in the particular locality. The derivation of this factor is beyond the scope of this manual, but it needs to be developed for individual areas. This factor will reflect transportation, taxes, availability of labor and equipment, laws and regulations, climate, political stability, general business climate, etc.

Step three is to make assumptions about the factors to be used for indirect capital costs, capital recovery, plant factor, etc. Those used in other reports can serve as a guide.

The final step is to determine whether or not there are any serious problems which could cause delays, extra work, and/or added expense to the project. These could include problems of scheduling, brine dipsosal, source development, availability of fuel or power, permits, etc. These added expenses must be reflected in the cost estimate.

The data and assumptions should be entered into the project information and cost summary sheets and then, if possible, the summary sheets carefully compared to standard costs worked out for plants of a similar size and process. Table 8-1 presents a sample comparison between estimated costs in the Fluor and ORNL reports for four 1-mgd [3,790-m^3/d] desalination facilities. They cover the multistage flash (MSF) distillation, seawater reverse osmosis, brackish water (TDS = 1,800 mg/l) electrodialysis, and brackish water reverse osmosis processes. The basic assumptions for these costs are shown in Tables 8-2 and 8-3.

Table 8-1

A COMPARISON OF COST ESTIMATES IN DOLLARS FOR SELECTED 1-MGD [3,785 m^3/d] DESALINATION FACILITIES FROM THE ORNL AND FLUOR REPORTS[a]

Plant	1-mgd Seawater MSF		1-mgd Seawater RO		1-mgd Brackish Water RO		1-mgd Brackish Water ED	
Performance Details	PF = 12 lb/1,000 Btu Acid Feed--250° F Concentration Factor = 2		Recovery = 30% Feed Temperature = 70° F One Stage		Feedwater TDS = 2,000-5,000 ORNL = 80% Recovery FLUOR = Not Stated		Feedwater TDS = 1,800 ORNL = No. 2 Water Analysis	
Price Base	1979--1st Qtr.	1977--2nd Qtr.	1979--1st Qtr.	1977--2nd Qtr.	1979--1st Qtr.	1977--2nd Qtr.	1979--1st Qtr.	1977--2nd Qtr.
Plant Factor	0.85	0.90	0.85	0.90	0.95	0.90	0.95	0.90
Report	ORNL	FLUOR	ORNL	FLUOR	ORNL	FLUOR	ORNL	FLUOR
Price Base	1979--1st Qtr.	1977--2nd Qtr.	1979--1st Qtr.	1977--2nd Qtr.	1979--1st Qtr.	1977--2nd Qtr.	1979--1st Qtr.	1977--2nd Qtr.
DIRECT CAPITAL COSTS								
Desalination plant	4,676	4,416	3,190	4,000	851	850	920	800
Site development	250	222	125	110	125	111	125	110
Intake and outfall systems	221	197	221	200	75	50	75	50
Electric utilities & switchgear	125	111	316	110	125	111	80	111
Cranes	--	--	--	--	--	--	--	--
TOTAL DIRECT COSTS	5,272	4,946	3,852	4,420	1,176	1,122	1,200	1,071
INDIRECT CAPITAL COSTS								
Interest during construction	333	177	151	442	30	85	22	80
Project management, overhead, profit[b]	1,977	--	--	--	--	--	--	--
Working capital	379	247	193	221	59	56	60	53
Contingency & A-E Fee	796	859	692	813	202	202	205	171
TOTAL INDIRECT COSTS	3,485	1,283	1,036	1,476	291	343	286	304
TOTAL CAPITAL COSTS	8,757	6,229	488	5,896	1,467	1,465	1,487	1,375
UNIT CAPITAL COST ($/INSTALLED GPD)	8.76	6.23	4.89	5.89	1.47	1.47	1.49	1.38
ANNUAL COST								
Labor O&M--direct salaries	143	145	28	71	28	91	28	91
Labor O&M--G&A charges	57	73	11	36	11	45	11	45
Chemicals	48	42	97	230	42	50	3	3
Supplies & maintenance materials	30	22	15	20	4	4	5	4
Energy--steam	211	859	--	--	--	--	--	--
Energy--electricity	52	37	294	300	86	49	54	51
Membrane replacement--materials	--	--	161	267	56	57	25	21
Membrane replacement--labor	--	--	--	110	--	110	--	--
Filter replacement			17	--	7	--	10	10
TOTAL O&M COSTS	541	1,178	623	1,034	234	406	136	225
ANNUAL FIXED CHARGE[c]	1,445	340	807	494	242	133	245	143
TOTAL ANNUAL COSTS	1,986	1,518	1,430	1,528	475	539	381	368
UNIT OPERATING COST ($/kgal)								
As shown in reports	6.41	4.55	4.61	4.65	1.37	1.61	1.10	1.12
Plant factor 0.80[d]	6.80	5.20	4.90	5.23	1.63	1.85	1.30	1.26
Plant factor 0.85[d]	6.41	4.89	4.61	4.92	1.53	1.74	1.23	1.19
Plant factor 0.90[d]	6.04	4.62	4.35	4.65	1.45	1.64	1.16	1.12
Plant factor 0.95[d]	5.73	4.38	4.12	4.41	1.37	1.55	1.10	1.06

Note: Values in the table, except for the unit operating costs at various plant factors and the Labor G&A for the Fluor report were obtained from the original data and were not computed. Inconsistencies do exist, but the data were presented as in the original. PF = Performance Factor.
Metric conversion factors are: mgd x 3,785 = m^3/d, (°F-32) x 0.556 = °C; \$/kgal ÷ 3,785 = \$/m^3; \$/gpd x 264 = \$/m^3/d; lb/1,000 Btu x 0.429 = kg/MJ.

[a] ORNL (Oak Ridge National Laboratory) Report by Larson, 1979. Fluor Report by Fluor Engineers, Inc., 1978.

[b] Larson (1979).

[c] As shown in the report or obtained by subtraction--not computed using recovery factor.

[d] The cost for the plant factor used in the report is underlined.

Table 8-2
BASIS FOR THE CAPITAL COST ESTIMATES IN
THE FLUOR AND ORNL REPORTS[a]

Parameter	Fluor	ORNL[b]
Plant Location	USA	USA
Plant Life	20 Years	30 years
Time of Cost Estimate	1977--2nd Quarter	1979--1st Quarter
Interest During Contruction	Included[c]	9.5%[d]
Project Management, Overhead, and Profit Working Capital	Excluded[e]	37.5%[f,i]
Working Capital	5%[f]	5%[f]
Contingency and A-E Fee	16%[g]	16%[g]
Brine Disposal Costs	Included[h]	Included[h]
Land Cost	Excluded	Excluded
Taxes and Insurance Cost	Excluded	Excluded

---

[a] Reports by Fluor, Inc., 1978 and Larson, 1979 (ORNL).
[b] ORNL = Oak Ridge National Laboratory.
[c] Sixteen percent of all other direct and indirect capital costs.
[d] Exact procedure for calculation not given.
[e] Project overhead costs are excluded. Project management and profit costs are not mentioned; they may be buried in other costs.
[f] Percentage of direct capital costs only.
[g] Percentage of all other direct and indirect capital costs.
[h] Probably does not include difficult cases.
[i] These costs are excluded for the membrane processes. Many of these expenses, where they exist, are probably buried in other costs.

Table 8-3
BASIS FOR THE OPERATING COST ESTIMATES IN THE FLUOR AND ORNL REPORTS[a]

	Fluor	ORNL
Annual Fixed Charge (capital depreciation)		
Interest rate	10%	16.5%
Years of writeoff	20	30
Capital recovery factor	0.1174	0.165
Plant Factor--seawater systems	0.9	0.85
Plant Factor--brackish water systems	0.9	0.95
Type of Distillation Plants	Single-Purpose	Dual-Purpose
Miscellaneous Materials and Supplies	1/2% DC	
Membrane Replacement		
ED--life (years)	7.5	7.5
ED--annual cost	20% DC	20% DC
Brackish RO--life (years)	3	3
Brackish RO--annual cost	Varying % DC	Actual 1979 Price
Seawater RO--life (years)	3	3
Seawater RO--annual cost	Varying % DC	3 times BW RO
Energy Costs		
Fuel Oil (\$/bbl) [\$/m^3]	\$13.00 [\$81.76]	\$13.20 [\$83.02]
Steam--1,000° F ($/MBtu) [$/kJ]	2.50 [2.37]	3.62 [3.43]
265° F ($/MBtu) [$/kJ]	2.50 [2.37]	0.98 [0.93]
200° F ($/MBtu) [$/kJ]	2.50 [2.37]	0.62 [0.59]
Electricity		
membrane processes ($/kWh)	0.024	0.025
distillation processes ($/kWh)	0.024	0.036
Labor costs		
Staff size and wage	Estimated	Input[b]
General and administrative overhead	50%	40%

---

[a] Reports by Fluor, Inc., 1978, and Larson, 1979 (ORNL).

[b] System cost and operating cost obtained by direct contact with original equipment manufacturers, membrane suppliers, consultants, the U.S. Government, major A&E firms, etc.

BWRO = Brackish water reverse osmosis
DC = Direct cost
ED = Electrodialysis
MBtu = Million Btu's
RO = Reverse osmosis
kJ = Kilojoules
bbl = barrel (42 gallons)
m^3 = cubic meter

Table 8-1 lists data for costs involved in using distillation and reverse osmosis for desalting of seawater and electrodialysis and reverse osmosis costs for desalting brackish water. The processes are the most commonly used methods of desalination for these applications, although there are many variations available. The cost increase involved in desalting seawater compared to brackish water ranges from 300 to 500 percent, or more, as shown by the data in the table. Obviously, it is economically more advantageous to desalt brackish water but, unfortunately, brackish water is not always available.

Appendix C contains an example of cost evaluation for a hypothetical desalination application on a small Caribbean island. This example uses the cost curves and other information in Appendices C and D to complete both the "Desalination Project Information Form" and the "Desalination Cost Summary Sheet."

Cost Evaluation

Again, in evaluating cost estimates formulated by others, a good general approach is to take the stated costs and break them down into the components shown on the same worksheet used for cost assumptions made for each component such as plant life, plant factor, capital recovery, interest, labor, energy, etc. Components that are missing or not specified should be identified. The breakdown should be compared to those for equivalent equipment from information obtained from using the estimating information in Appendix C and/or data available in the ORNL, Fluor, and Catalytic documents. Items which are at obvious cost variance with those established by previous studies or by basic common sense should be carefully investigated.

The example of this type of evaluation in Table 8-1 places the data from the ORNL and the Fluor reports side-by-side. Obviously, with this type of format, deviations in assigning costs can quickly be identified. As an example, in Table 8-1 variances are readily apparent in estimating costs for the multistage flash distillation plant (which has almost a 300 percent difference in assigned indirect and capital costs) and the large difference (about 400 percent) in energy costs. The former variance is due to the cost attributed to project management, overhead, and profit. In the latter case, the energy costs estimated for steam differ widely, although the fuel price differs (see Table 8-3) by only 1.5 percent. This is due to the assumption by ORNL that the installation will be a dual-purpose facility, whereas Fluor priced the energy out for a single-purpose plant. With a dual-purpose plant, part of the cost of providing the steam has been assessed to an electric generating station, and the steam price is based on a formula which relates to the steam's capability of doing work or to some other method.

Certain variances may be justified, but the use of data comparison, such as in the example above, allows potential problems to be spotted and encourages investigation of the basic cost assumptions.

Some of the data presented in literature on operations and/or projections of costs are given as costs only, without the basic assumptions clearly stated. It is important to ascertain these assumptions, since they can affect the costs and possibly the conclusions.

COST COMPONENTS

As seen in Figure 8-3, there are two major cost components to be computed and considered: capital costs and annual costs.

Capital Costs

The elements which usually make up the capital costs of a desalination facility are divided into direct, indirect, and non-depreciable capital costs.

Direct Capital Costs. These are the costs of equipment and/or construction services and for desalination facilities. They generally include:

1. Installed desalination equipment, including piping, electrical, drains, etc.
2. Site development--buildings, roads, fences, grading, etc.
3. Development of the raw water source and its conveyance to the facility.
4. The equipment and/or structures associated with brine disposal.
5. Development of the power source and/or power conveyance to the site.

Indirect Capital Costs. These are capital costs which do not involve the purchase of equipment or construction services. They include:

1. Interest costs incurred during construction startup.
2. Contingencies.
3. Architect/Engineer project management and other consulting fees.

Non-Depreciable Costs. These expenditures must be made to begin the project and/or operation but the materials purchased remain (or are replaced) during the project's life so that they are on hand at the end of the project. These costs include:

1. Working capital which includes on-hand fuel, chemicals, materials, other supplies, and operating capital.

2. Land costs.

Evaluation of Capital Costs--In evaluating or estimating capital costs, the two major items of concern, aside from the basic equipment costs, are: (1) realistic definition of the assumptions for interest charges, contingencies, working capital, and (2) inclusion of all the needed major capital development costs. The latter could include special source development, brine disposal, special site work, etc.

Capital costs for dual-purpose (electricity--water) plants must be developed carefully so as to fairly allocate capital costs between the two components. Care needs to be exercised in interpreting or evaluating the cost of the desalination portion of an already built (or priced) dual-purpose facility. One method used to estimate the capital cost of the desalination facility (in dual-purpose plants) is to select a unit cost for the generating portion (MW of installed capacity) and to use this cost to arrive at the capital cost of the desalination facility by subtraction.

Obviously, the assumed unit cost for the generating facilities can greatly affect the desalination cost. If the unit power cost is high, the desalination cost is low; if it is low, the desalination unit cost is high. In such situations, it is prudent to investigate the premise on which unit costs are calculated.

Cost Curves

Figure 8-1 presents cost curves showing generalized capital costs for various desalination processes based on the price per gallon per day and $\$/m^3/d$ of installed capacity (a commonly used means of comparison in the industry). This is adapted from the data in the Fluor report. These are estimates for facilities in the USA. The actual cost of a facility can vary greatly (generally upward) from these costs depending on the circumstances of each individual application. These data are useful but should be used with prudence and discretion.

Annual Costs

The elements which usually make up the theoretical annual operating costs are: (1) recurring costs, (2) operating expenditures, expendables, and (3) the fixed charge to cover capital depreciation. These costs are combined with the annual plant factor to compute the annual unit operating cost in terms such as dollars per kgal or $\$/m^3$. A breakdown of the operating cost components, as a percent of the total operating costs, for four 1-mgd [3,785-m^3/d] desalination plants is shown in the form of per charts on Figure 8-4. These charts use data developed by Fluor Engineers (1978) on both brackish and seawater facilities.

Recurring Costs. These include costs such as insurance and taxes which continue year after year. In comparing processes for future installation, the recurring costs are often not considered since they often are equivalent for all of the options.

Operating Costs. Almost all of these costs occur steadily during the operation of the plant. They include:

1. Labor (both salaries and administrative costs).
2. Supplies and maintenance materials.
3. Energy.
4. Special repairs.
5. Chemicals and filters.
6. Membrane replacement.

The most significant of these are discussed below.

i) Labor--A guide to staffing of a desalination facility is presented in Appendix C. This is only a general approximation of labor needed, which can vary depending on local situations. Depending on wage scales, labor can often amount to up to one-third of the annual costs excluding capital recovery. This can be a critical investment of funds for a desalination facility. A good operating staff can often make the poorest plant perform to some degree and a decent plant perform excellently. An incompetent staff can usually destroy or severely damage almost any desalination plant produced.

The degree of investment necessary to obtain a good operating staff is a site-specific determination. The added production (or high plant factor) obtained by using a competent, and often highly paid, staff usually keeps the costs per unit of production reasonable and far more predictable.

ii) Energy--For most processes, energy costs make up almost 30 to 60 percent of the daily out-of-pocket costs involved in producing water by desalination. Energy costs need to be ascertained on a site-specific basis. True electric costs in many developing areas could easily be in the range of $0.10 to $0.30/ kWh and more, where small generators are utilized.

When comparisons are made between dual- and single-purpose thermal plants, and between processes using steam and those using only electricity, a fair value must be assigned to the energy utilized. Although the energy that a particular process requires to produce a given quantity of water can be estimated with some accuracy, the cost of that energy can vary widely between processes, depending on the energy source, temperature levels, previous use, and the method of accounting utilized.

This is explained in detail in an article by Cox (1979) in which he states that, "...the cost of energy for desalting usually has no relationship to the quantity of energy consumed by the process, nor to the process efficiency." Thus, when different processes are compared (especially between distillation processes or between distillation and membrane processes) it is important too that the comparison take into account more than just the quantity of energy utilized per unit of product.

"Most distillation plants do not use boiler steam directly, but rather extracted steam at a reduced pressure from a turbine after the steam has already performed appreciable work. There is not unanimity as to how extraction steam should be valued, but a frequently used method--because it is the simplest--is to value the steam according to the quantity of work it can still perform. On this basis, boiler steam that has done no work has a value rating of 100 percent, while turbine steam exhausted to the condenser when no more capacity for work remains has a rating of 0 percent.

Extraction steam at intermediate temperatures then is valued according to its heat content....To be valid, the turbine extraction must be designed for the temperature used by the desalting plant. This is an especially important consideration when installing a new desalting plant to use steam from an existing extraction turbine. Steam should be valued at the saturation temperature (determined by the pressure so as to eliminate possible superheat) at which it is extracted from the turbine, and not the temperature at which it is used by the desalting plant." (Cox, 1979.)

A general guide to energy cost evaluation is presented in Appendix C, Figures C-16A and B.

iii) Operating Chemicals--For some processes chemicals can constitute one of the major operating expenditures. More importantly, they are used continuously and hence are noticed since they are constantly being purchased. There can be a considerable variation in the cost and availability of chemicals in any area. Potential delays in transportation would require appropriate storage facilities to keep a sufficient stock available so as to permit continued operation.

The overall availability of chemicals can be a factor in process selection. One of the advantages of the electrodialysis reversal process is its minimal use of chemicals, which can be extremely important in reducing costs in many isolated areas.

iv) Membrane Replacement--Three years has been commonly used as the life of reverse osmosis membranes. Some manufacturers have extended this for brackish water applications up to 5 years. Which time estimate is used will have a significant effect on the cost allocated to this category. The assumption should be clearly stated and should correspond to the membrane life that the manufacturer guarantees. For electrodialysis membranes it averages about 5-10 years.

Capital Recovery. Standard tables are available for computing capital recovery. An interest rate must be set and a plant life estimated. A period of 20 years seems to be a realistic plant life for well designed and constructed desalination facilities. For poorly designed and/or constructed plants, 5 to 10 years might be more realistic.

Summary. The major items are capital recovery, energy, labor, and membrane replacement (for the membrane processes). The evaluation or estimation of these items should be performed carefully since they have a major impact on the annual costs. The assumptions behind these values should be clearly understood and stated.

Unit Production Costs

General. Unit production costs translate all the above cost elements into a cost per unit of production such as \$/kgal or \$/m^3. Such a parameter is useful in performing cost comparisons and enabling people to understand project costs. The method used to calculate the unit production cost is shown on the cost summary sheet on Figure 8-3.

Aside from the costs of the capital, operation, etc., developed above, the most important item in the calculation is the plant factor.

Plant Factor. The determination of the plant factor should play an important role in both the economics of a desalination facility and in the selection of an appropriate process.

$$\text{Plant Factor} = \frac{\text{Actual Production}}{\text{Rated (Design) Production}}$$

The plant factor is often considered on an annual basis. If during a year a facility is producing water continuously at the design capacity, it would have a plant factor of 1, or 100 percent. If it either did not produce at the design capacity or was shut down for some reason (such as general cleaning or repairs), the factor would fall accordingly.

Investigators and manufacturers often use plant factors in the range of 0.80 to 0.95 in making cost estimates. A 0.9 plant factor means that a plant producing at its design capacity would be shut down for about 36 days a year (or 10 percent of the time). At plant factor of 0.95, the shutdown would be for only 18 days (or 5 percent of the time). Many plants can maintain this type of performance, but others cannot. A major distillation complex in the Caribbean reported an average annual plant factor of about 0.30 over the 3-year period 1977-1979. This of course would have a profound effect on its overall economics of operation if the original prediction were 0.85 or 0.90.

The plant factor is used to determine the annual unit production cost (such as \$/kgal or \$/m^3) for water in the following manner:

$$\text{Unit Water Cost (\$/unit)} = \frac{\text{Annual Recurring Costs} + \text{Annual Fixed Charges}}{\text{Design Capacity (units/day)} \times 365 \text{ days} \times \text{Plant Factor}}$$

It is this cost that is usually used to select a process or design. Since the plant factor affects this cost directly, the assumptions behind its selection are extremely important.

Aside from the overall, generally predictable, technical reasons for plant shutdown for cleaning, inspection, and schedule maintenance, there are site-specific conditions which can greatly affect the plant factors. Familiarity with these local conditions is required to generate realistic predictions and to ascertain how they will affect plant operation, maintenance, and performance.

Innumerable factors can affect the actual operation of the plant and the plant factor to one degree or another: local residents; their desire and support for the project; climate; labor conditions; the degree of operator training; technological capability; economic conditions; transportation systems; availability of power, chemicals, spare parts, service, and tools; past experience; and the degree of planning and system integration carried out. A lack of consideration of these items will only change the economic evaluation and selection--it will not prevent them from ultimately affecting the project. As these factors are integrated into the economic evaluation, the theoretical cost figures will approach the actual number.

Unfortunately, this process requires that subjective information be translated into an objective number. This process can be open to question and criticism, especially by those whose interests may be affected.

Whatever the plant factor, its accurate prediction will play a major role in arriving at a final selection. It is a key number and its selection should be justified. As seen at the bottom of Table 8-1, a change in the plant factor can have a significant effect on the unit production cost.

Cost Curves. Figure 8-2 presents curves showing generalized annual unit production costs based on the data and assumptions in the Fluor report. These curves have been adjusted by doubling the cost of power to reflect the current price of fuel oil on the world market. A general review of the data from the two studies in Table 8-1 indicates that, overall, other costs shown have not changed substantially in the 2 intervening years between the studies.

SELECTION

The actual selection of a desalination process, its production capacity, and variations in its design should be undertaken with the help of a qualified consultant and/or other qualified assistance. However, the use of Table D-1, "Summary of Guidelines of Desalting Process Selection," can aid greatly in this effort. Discussions with manufacturing representatives can be helpful, but it should be realized that they can tend toward a process which they can furnish.

Table D-1 indicates the following process applications are currently viable in most cases:

- Seawater
    - Distillation
    - Reverse Osmosis

- Brackish Water
    - Electrodialysis
    - Reverse Osmosis

This may change for some cases involving difficult feedwaters, brine disposal, or other special local conditions. When the freezing process becomes commercially available, it will normally be used for seawater applications.

Economics always plays an important part in the ultimate process selection. In economic optimization between capital costs and operating costs, the changes in interest rates (for capital recovery) and energy (for operating costs) affect the ultimate solution. As fuel costs increase, the optimal solution (all other things being equal) tends toward a higher efficiency design.

When setting the capital recovery interest rate, a higher interest rate will (if fuel prices stabilize) optimize the selection at a lower energy efficiency or performance factor. As noted by the developer of the multistage flash process, Professor Robert Silver (1979), the rise in both interest and energy costs have tended to keep the optimum performance factors of distillation plants about the same over the past few years. Wherever the optimal solution occurs, it is important to keep in mind how it is derived so that when assumptions, such as those regarding interest rates or energy costs are made, the consequences are understood in advance.

Another aid to the selection process and evaluation is contained on the Desalination Project Information and Cost Summary Sheets in Appendix C. These sheets include a series of questions on conditions, assumptions, and data which are useful in approaching a project and determining whether sufficient data (with reasonable assumptions) exist in order to reach a conclusion. They should be used in project reviews and in making cost estimates of various processes as part of process selection and evaluation.

By completing the information for several processes (such as electrodialysis and reverse osmosis for a brackish water application), an estimate of relative costs can be obtained. This can furnish much of the initial objective input into the selection of the process.

As discussed in the next chapter, the selection and operation of desalination facilities in developing areas requires consideration of many additional factors which are essentially subjective in nature. They concern past experience, available technological expertise, availability of materials, cultural characteristics, special local conditions, etc. There are two ways to handle these considerations within the scope of an economic optimization of a solution. The first is to use a normal cost analysis and then to list the subjective conditions as a form of backup for the selection recommended (which may not appear to be the lowest-cost solution). The other method is to translate these subjective considerations into a reasonable plant factor to be used in the optimization process. There is little experience in doing this, but it can be important as it will require some thought on ways and means by which local conditions might logically (or illogically) affect the continued operation of a desalination facility in that area.

Unfortunately, without special consideration of these characteristics or situations of local significance, an economic analysis could lead to the selection of a process which might look good on paper, but be less than satisfactory in long-term operation. Thus, theoretically the most energy-efficient process may contain components or require operating and maintenance procedures which create difficulties in a particular local environment and result in excessive downtime, unexpectedly low plant factors, high unit production costs, and possibly a non-functioning plant.

This can be particularly problematic when introducing a process or design which has minimal realistic commercial operating experience. Without data (except those under controlled or theoretical conditions), it is difficult to accurately assess the the long-term plant factor. This then introduces a degree of risk or uncertainty which must be factored into the selection process.

In addition, there must be an evaluation of the future financial, manpower, and facilities support that a desalination project will have once installed.

Some of the financial risk associated with any of the uncertainties discussed above can be offset by requiring long-term performance guarantees and/or placing the operation under contract operation and/or management. The latter is a common practice in many areas of the Middle East, especially in isolated areas.

Generally, manufacturers are very reluctant to grant long-term overall guarantees on a facility unless they operate it. If they do not operate the facility, the guarantee is apt to have conditions and/or penalties which can negate much of its meaning. It is wise to carefully examine and evaluate any agreement concerning guarantees.

PROCUREMENT

Although desalination equipment can be purchased and installed through direct negotiations with the various vendors, it is usually good practice to utilize a qualified consultant or other independent agency to assist the owner in the transaction. The consultant can aid the owner in process evaluation and selection and then draw up appropriate specifications and bid documents.

The specifications and bid documents should be used to reduce the uncertainties associated with the future costs (and plant factor) of the desalination unit. This could include specifying the process, standards, materials of construction, performance, spare parts, guarantees, construction schedule, etc.

Evaluations of the bids, especially where exceptions are taken and alternates priced out, form an important part of the procurement process. The specifications should be written with sufficient care that the low bid is actually a good long-term investment and not a potential disaster. An ever-present danger in the bidding process, especially on large projects, is a bid which is too low to cover the actual reasonable costs of the contractor. This is generally no bargain, because all parties are likely to suffer in this type of situation.

WARNING

1. Caution should be exercised when using these cost data.
2. These costs are computed for basic desalination facilities in the USA. These costs include no storage capacity and allow for only fundamental feedwater development and brine disposal. If storage capacity is added and/or more complex feedwater or brine facilities are required, overall project costs can significantly increase. This is especially true with brackish water facilities. Construction in locations outside the USA can also add to the overall cost. Actual costs of up to 300% (or more) of those shown on this figure (especially at lower capacities) would not be unusual for projects in many developing areas.

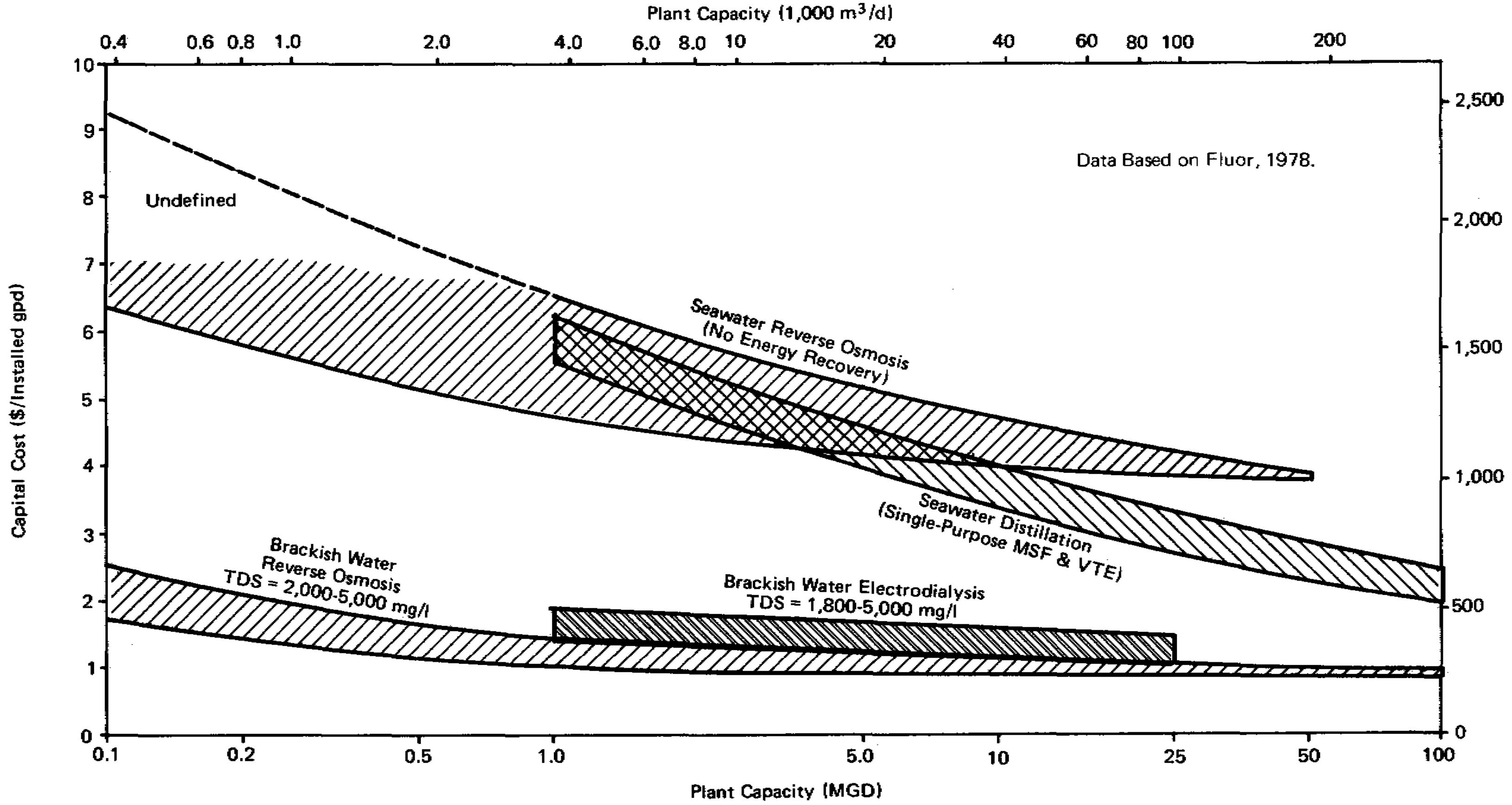


FIGURE 8-1. Generalized capital cost estimates for desalination processes.

WARNING

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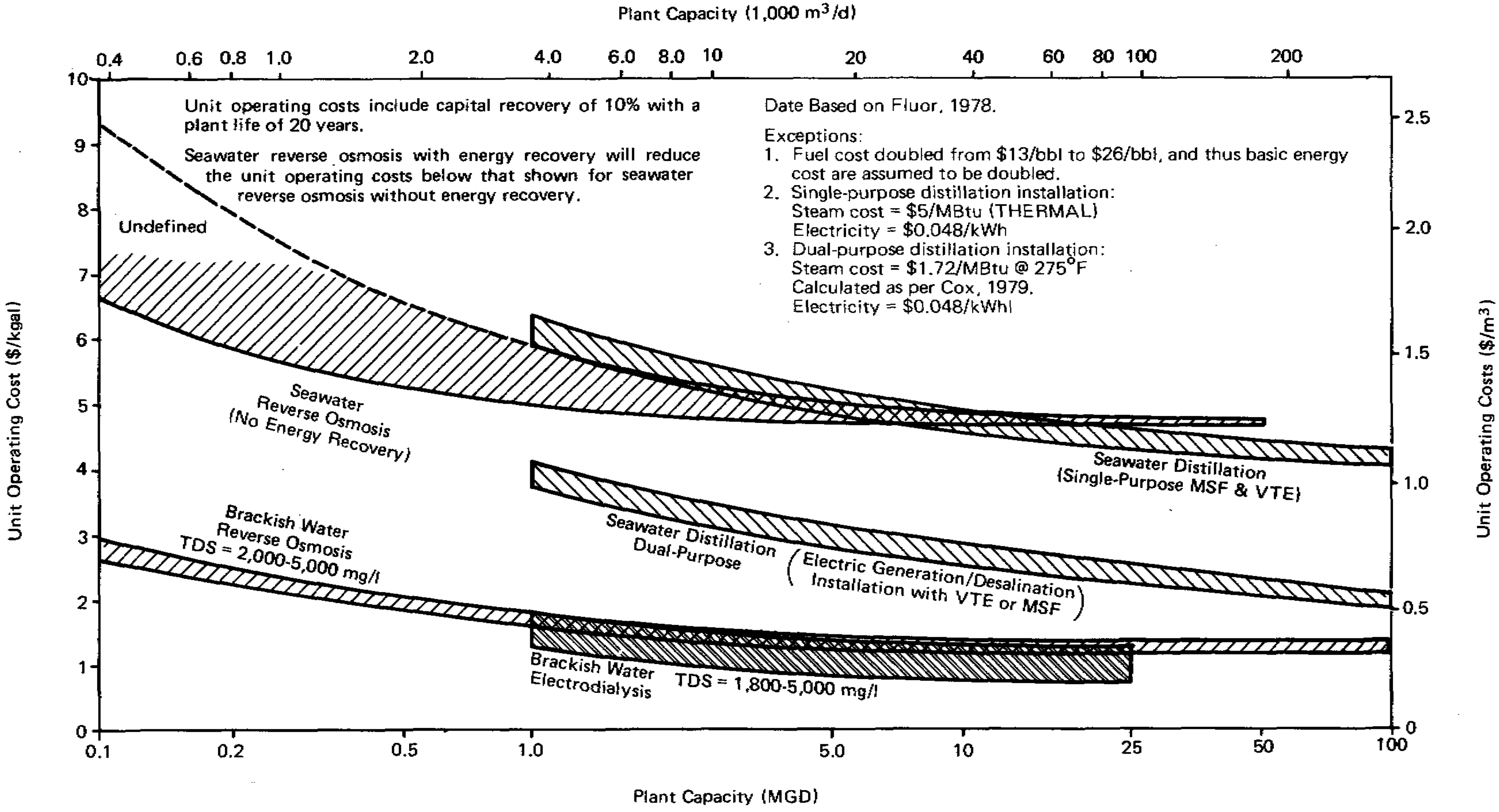
2. These costs are computed for basic desalination facilities in the USA. These costs include no storage capacity and allow for only fundamental feedwater development and brine disposal. If storage capacity is added and/or more complex feedwater or brine facilities are required, overall project costs can significantly increase. This is especially true with brackish water facilities. Construction in locations outside the USA can also add to the overall cost. Actual costs of up to 300% (or more) of those shown on this figure (especially at lower capacities) would not be unusual for projects in many developing areas.



Unit operating costs include capital recovery of 10% with a plant life of 20 years.

Seawater reverse osmosis with energy recovery will reduce the unit operating costs below that shown for seawater reverse osmosis without energy recovery.

Date Based on Fluor, 1978.

Exceptions:

1. Fuel cost doubled from \$13/bbl to \$26/bbl, and thus basic energy cost are assumed to be doubled.
2. Single-purpose distillation installation:
   Steam cost = \$5/MBtu (THERMAL)
   Electricity = \$0.048/kWh
3. Dual-purpose distillation installation:
   Steam cost = \$1.72/MBtu @ 275°F
   Calculated as per Cox, 1979.
   Electricity = \$0.048/kWhl

FIGURE 8-2. Generalized unit production cost estimates for desalination processes.

DESALINATION COST SUMMARY SHEET

Location ____________________ Capacity __________ mgd [m^3/d]
Type of Plant ____________________ Plant Life ________ Years Plant Factor __________
Source of Feedwater ____________________ TDS Level __________
Recovery or Performance Factor ____________________
Energy Source(s): Fuel (__________________ type)/Electricity/Steam (Circle Sources)

CAPITAL COSTS

DIRECT CAPITAL COSTS

1.	Feedwater Supply Development	$______	
2.	Feedwater Treatment	$______	
3.	Desalination Equipment	$______	
4.	Site Development	$______	
5.	Energy Source Development	$______	
6.	Electrical Equipment (switchgear)	$______	
7.	Brine Disposal	$______	
8.	Product Water Storage and Treatment	$______	
9.	Other	$______	
	SUBTOTAL DIRECT CAPITAL COSTS (1-9)	$ Ⓐ	

INDIRECT CAPITAL COSTS

10.	Interest During Construction	$______	
11.	A/E, Project Management Fees	$______	
12.	*Contingencies*	$______	
13.	Startup Costs	$______	
14.	Other	$______	
	SUBTOTAL INDIRECT CAPITAL COSTS (10-14)	$ Ⓑ	
	TOTAL DEPRECIABLE CAPITAL COSTS (A + B) →		$ Ⓒ

OTHER CAPITAL COSTS (NON-DEPRECIABLE)

15.	Land __________ acres [m^2]	$______	
16.	Working Capital	$______	
	TOTAL OF OTHER CAPITAL COSTS (15-16) →		$ Ⓓ
	TOTAL OF ALL CAPITAL COSTS (C + D) →		$
	UNIT CAPITAL COST ($/GPD [$m^3/d$] INSTALLED CAPACITY) →		$ /gpd [m^3/d]

ANNUAL COSTS

Recurring Costs

17.	Taxes	$______	
18.	Insurance	$______	
19.	Other	$______	
	TOTAL ANNUAL RECURRING COSTS (17-19) →		$ Ⓔ

OPERATION & MAINTENANCE (O&M) COSTS

20.	Labor—Salaries	$______	
21.	Labor—General & Administrative Overhead (_____%)	$______	
22.	Chemicals	$______	
23.	Supplies and Maintenance Materials	$______	
24.	Membrane Replacement (______ yr Life)	$______	
25.	Special Repairs or Overhauls	$______	
26.	Energy—Fuel/Steam (*Circle one*) cost/unit ______	$______	
27.	Energy—Electricity __________ $/kWhr	$______	
28.	Other	$______	
	TOTAL ANNUAL O&M COSTS (20-28) →		$ Ⓕ

ANNUAL FIXED CHARGE

_____% Interest & _____ yr Plant Life   Capital Recovery Factor (CRF) __________

$$\left(\underset{\text{(Depreciable Capital)}}{\$\underline{\qquad\qquad\qquad}}\ \textcircled{C} \times \underset{\text{(CRF)}}{\underline{\qquad\qquad}}\right) + \left(\underset{\text{(Nondepreciable)}}{\underline{\qquad\qquad\qquad}}\ \textcircled{D} \times \underset{\text{(CRF)}}{\underline{\qquad\qquad}}\right)^{*}$$

$ Ⓖ

TOTAL ANNUAL COSTS (E + F + G) → $ Ⓗ

*In many analysis the capital recovery is not computed for nondepreciable capital (D).

UNIT PRODUCTION COSTS

Annual Production __________________ kgal [m^3] ☐ Design ☐ Actual

Operating Units

$$\text{UNIT COST} = \frac{\text{Total Annual Cost } \textcircled{H}}{\text{Actual Annual Production}} = \underline{\qquad\qquad\qquad\qquad} = \$ \qquad /\text{kgal}\,[m^3]$$

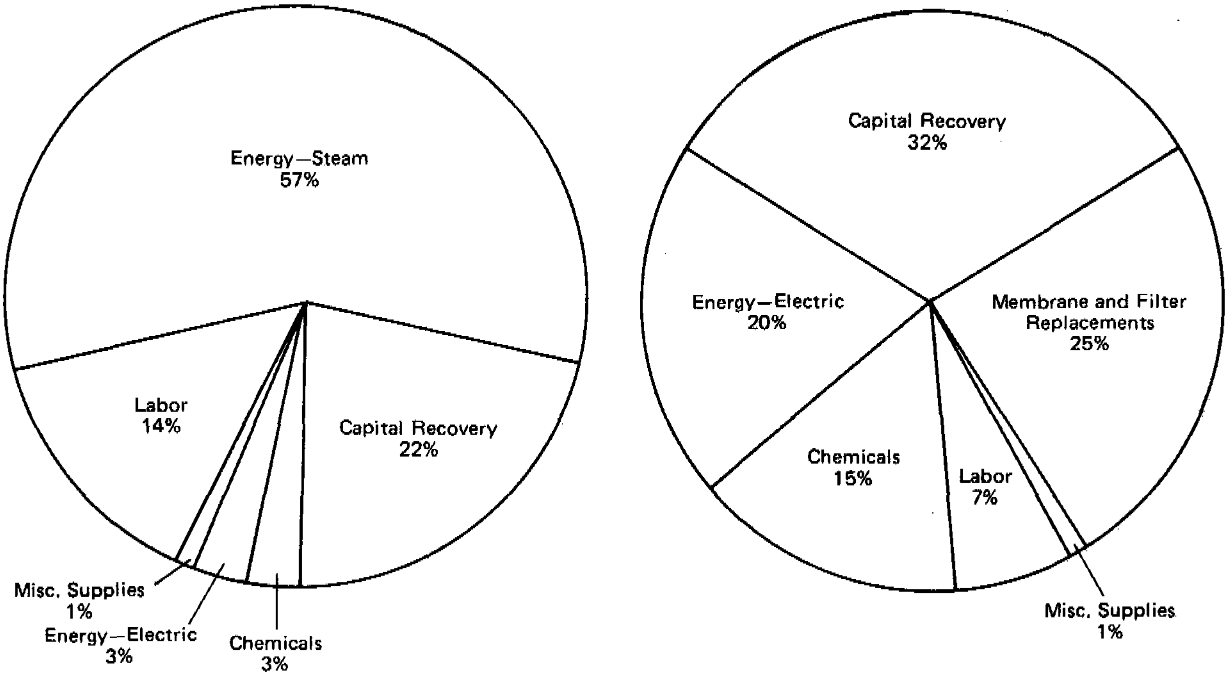
For Estimating Costs

$$\text{UNIT COST} = \frac{\text{Total Annual Cost } \textcircled{H}}{\text{Annual Design Capacity} \times \text{Plant Factor}} = \underline{\qquad\qquad\qquad\qquad} = \$ \qquad /\text{kgal}\,[m^3]$$

Cost Estimator ____________________ Date __________ Project __________

Attach the Desalination Project Information Sheets

FIGURE 8-3. Desalination cost summary sheet.



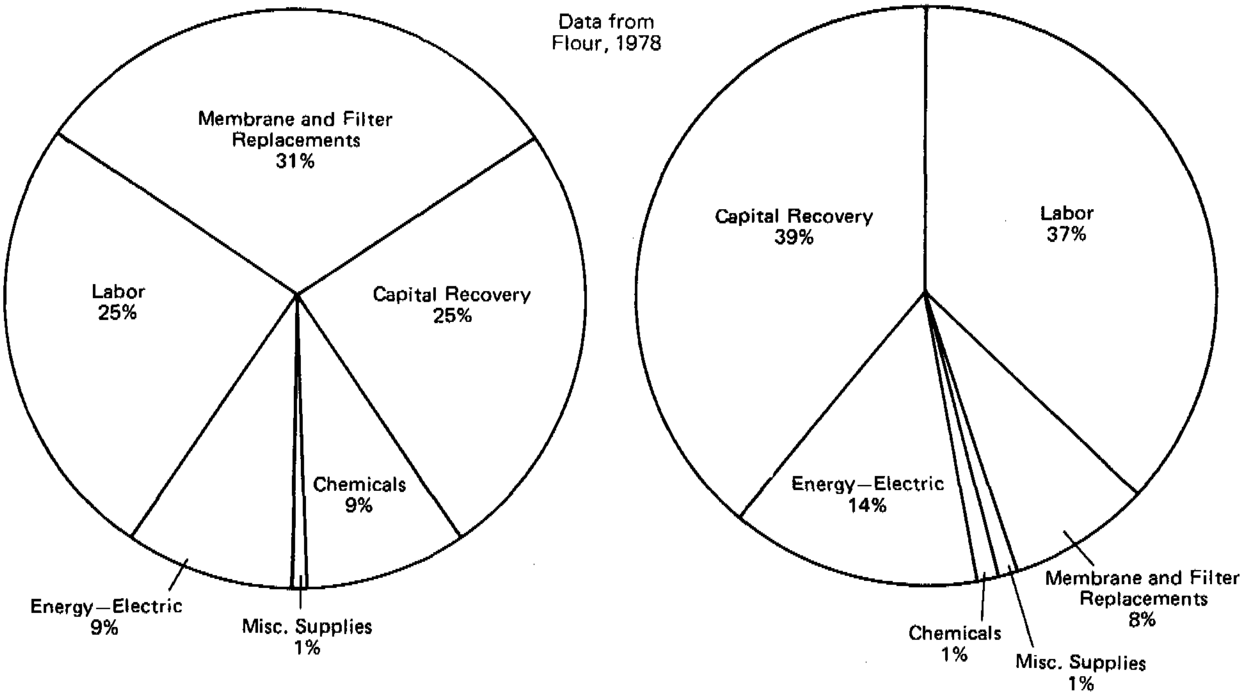
1-mgd [3,790-m^3/d]
SEAWATER MULTISTAGE FLASH PLANT—
SINGLE PURPOSE

Acid Feed—250°F [121.1°C]
Performance Factor—12 lb/1,000 Btu
Performance Factor—5.15 kg/MJ

1-mgd [3,785-m^3/d]
SEAWATER REVERSE OSMOSIS PLANT—
SINGLE STAGE

30% Recovery
Feedwater Temperature 70°F [21°C]

Data from
Flour, 1978



1-mgd [3,785-m^3/d]
BRACKISH WATER REVERSE OSMOSIS PLANT

Feedwater = 2,000-5,000 mg/l

1-mgd [3,785-m^3/d]
BRACKISH WATER ELECTRODIALYSIS PLANT

Feedwater TDS = 1,800 mg/l

FIGURE 8-4. Distribution of operating costs in various desalination processes.

CHAPTER 9

Chapter 9
DESALINATION IN DEVELOPING COUNTRIES

INTRODUCTION

As noted in the beginning of this manual, desalination does not create new water but only expands the potential for treating and utilizing water sources which were previously considered unusable due to their high salt content. This means that seawater and brackish water can now be considered as potential sources of freshwater.

For the developing countries which have shortages of freshwater, this offers an opportunity to expand their supply of freshwater, provided, of course, that a source of water to be desalted is available and that the country or locality is able to afford this type of treatment. As illustrated in Chapter 8, the capital and operating costs associated with desalination are high. With the possible exception of desalination with a reliable alternate energy source, these high operating costs continue as long as water is produced.

Standard of Living

The use of desalted water offers possible assistance in improving standards of living, as it provides a reliable source of freshwater for areas that in the past have relied on the uncertainties of nature to provide. It can also reduce the distance that water is transported from the source to the user, as desalination can make closer, previously unpalatable sources usable. Since this transportation is often by truck, barge, animals, and/or carried by people, desalination can often produce savings in fuel, money, and time.

Public Health

Both reverse osmosis and most distillation processes have the potential for dramatically reducing the bacteriological content of the water during the desalination process. However, it is recommended in both of those cases, as well as the other processes, that additional disinfection such as chlorination be used to provide a residual for disinfection in the distribution or storage system. If bacterial reduction is to be a major factor for using a desalination process, it will almost always prove more cost effective to simply use a standard water disinfection practice.

"In any case, in a rural village environment, the improvement of a public water supply will provide significant contribution to improved overall public health. Other conditions which may exist for exposing the populace to bacteria (especially the anus-tomouth path) often mask the true effects possible. A close positive link between public health and water supplies seems to be related to the care with which a householder can obtain water of a reasonably good quality" (Saunders & Warford, 1976).

Goals

As with any capital project in a developing country, the goal is to provide a long-lasting facility that will be utilized by the people of a locality to their general benefit.

Desalination facilities, in any country, tend to be technically complex. In addition, the business of supplying water to a community can be not only complicated--it can have a profound effect on the community in general. The combination of these in a developing country can present unique and difficult problems. This points up the need for a very careful approach to evaluation, planning, project execution, and follow-through so that the project is successful beyond the ribbon cutting and opening ceremonies and in fact will last for its projected design life.

CONSIDERATIONS

It is very difficult to generalize on the application of desalination to developing countries. Desalination is not new in some of these countries. Egypt, for example, has used desalination to produce water for municipal purposes since the early 1930's and currently has a wide variety of desalination units operating successfully in the country.

Most desalination units are much like any other piece of machinery. They will function well if there is proper equipment selection, installation, operation, and maintenance. When one or more of these elements are lacking, problems generally occur.

An organization such as U.S. A.I.D., when involved in a desalination project, has greater control over equipment selection and installation and very little control over subsequent operation and maintenance. Although the quality of the equipment and installation can be maximized by proper selection, supervision, and appropriate capital expenditures, this is of little value if proper operation and maintenance are not ensured.

There are four main keys to increasing the potential for proper operation and maintenance:

1. Proper Selection--A process must be selected which is suitable for the application.

2. Project Viability--The project must show strong indications of viability and economic soundness.

3. Appropriate Technology--A unit must be installed with design, materials, and technology appropriate for operation in the particular locality.

4. Organization--An organization which is able to effectively carry out proper operation and maintenance of the unit must exist.

These keys are applicable in both urban and village environments.

Process Selection

As noted previously in this manual, there is no one "best" desalination process for desalination in developing (or industrial) countries. Each application should be carefully studied so as to match it with an appropriate process. This manual provides a tool to assist in this selection process, but it must be augmented by experienced, professional help.

Generally, process selection is based on economic optimization of the estimated unit production costs. Aside from the need for an accurate, site-specific cost estimate, the prediction of plant factors for each process will be a key item. The problems associated with the prediction of realistic plant factors are discussed in detail in Chapter 8.

Project Viability

In view of the substantial capital and operating costs involved with desalting, the need for the project should be firmly established. This would involve a water resources evaluation as outlined in Chapter 8.

The desire of the local people to have a desalination facility is a very important factor. Compared to urban locations, village locations require more careful study. Local people must be involved and must want the project. This desire should be tempered with the knowledge of present and future costs and/or obligations of the community as far as the desalination facility is concerned. The community's reasonable desires should be respected during the project's planning and execution. Their continuing support will be needed for the project to succeed in the long run. If they are opposed to the project, it should be remembered that they will be there long after the project organizers and financiers have left.

The failure to solicit and establish local participation has led to the demise of a number of desalination units, especially solar stills.

The economic viability of a project will be tied to the water resources evaluation and the present and future development plans for an area. In some cases the benefits may be intangible in nature and difficult to quantify. For example, the availability of good water may reduce migration to urban areas.

Appropriate Technology

The type of technology employed in a desalination facility should be appropriate to the area where it will be operated. This technology could range from a simple one-family passive solar still to a multistage flash plant which would supply a large city or region. The matching of technology to the attributes and limitations of a particular area must be a vital concern if long-term successful operation is to be achieved. It can require more intensive planning and engineering considerations than is often needed in industrial countries. This, in turn, will add capital expense to the project, but this cost should be considered a necessary part of a project if it is to be a successful investment of a country's financial resources.

Desalination facilities can have a high percentage of parts and expendable materials which require the expenditure of foreign exchange for continued operation. These may include membranes, motors, gauges, electronic components, special metals, chemicals, etc. If foreign exchange expenditures become limited in the future, special efforts may be needed to include technology which will allow operation, repairs, and maintenance using locally available items and people wherever possible.

Design. Designing for energy-efficiency is important. However, for many developing countries reliability is even more important. The apparent sacrifice in efficiency will probably, in the end, represent no sacrifice at all, but rather the best long-term value. Even a high-performance plant is running at zero efficiency when it will not work.

The greatest difficulty comes in making the judgment between reliability and performance. For a developing area, it is probably better to err on the side of reliability.

Special consideration should be given to the ease of maintenance because effective maintenance will vary depending on the ability, training, and financial support (parts, tools, etc.) of the local staff. Automation should be evaluated carefully; although it simplifies operation for a technically less qualified staff, it can cause repair problems if a breakdown occurs due to the sophisticated electronics, sensors, motors, etc. Where the planned staff will be large anyway it may not be a cost savings and in many cases a less automated operation may promote a more in-depth process understanding of the unit by the operators.

The availability of spare parts is very important to continued operation. Where suitable inventories will not be possible to obtain or maintain and water production is critical, the installation of a number of smaller independent but duplicate units should be evaluated. This will result in higher capital costs but it will permit partial shutdown for repairs on one unit and the use of parts from other units for spares if necessary. In addition, the use of standard parts, motors, etc., even with some sacrifice in efficiency, might be warranted to reduce the number of spare parts required.

An example of this was the use of identical first- and second-stage pumps and membranes in the 3.2-mgd [12,100-m^3/d] seawater reverse osmosis facility installed in Jeddah, Saudi Arabia. Although brackish water membranes and lower-pressure pumps could have been used in the second stages, the use of a unit with components identical to those in the first stage allowed maximum flexibility in plant operation and repair.

Even though innovations in design may provide the opportunity to reduce capital and operating costs, they should be approached carefully. The risk and possible extra costs associated with new applications should be weighed, considering the fact that in many areas technical expertise routinely required for making field adjustments, alterations, and generally "fine tuning" the operation is often lacking. If there is a perceived benefit in utilizing some innovative technology, it would be prudent to try to have the expected operation covered by contractual guarantees. This will add some capital cost but will reduce risk.

Other problems in the design and specifications for desalination facilities can arise from the difference in standards and mechanical tolerances between regions and/or countries. These should be clearly stated and understood by all parties involved.

Operation. In any country after the plant is installed there is very little that can be done about the design and materials, and at that point it becomes the job of the operating staff to make the facility function as designed. As mentioned in Chapter 8, a very good operating staff can make almost any plant run to some degree, while an incompetent staff can manage to destroy almost any facility. In between lies a grey area in which many plants operate, but there is no question that a trained, motivated, and properly supported staff can be the best investment made in a desalination facility.

Desalination equipment has two characteristics which can cause problems in operation and/or longevity. The first is that virtually all the action takes place inside closed containers. The operator can determine what is happening and make process decisions based only on the interpretation of gauge readings for flow, temperature, pressure, conductivity, etc. The water will appear the same coming in and going out, and in fact in most cases the operator cannot even see the water. If an operator fails to understand the cause-and-effect relationship between the gauges and the plant's operation, the operator cannot make intelligent process decisions. As operating instructions cannot cover every possible situation, a poorly trained operator can easily make the wrong decision (which might be doing nothing at all).

The second characteristic is that most desalination equipment will continue to run even if it is slowly (or rapidly) destroying itself internally. Tubes will corrode, membranes will clog, scale will form, etc., and the plant will continue to produce water. The only indication might be the readings of the gauges (if they are functioning correctly) or perhaps warning lights or tripped safety switches on fairly automated equipment. Thus, for example, if the operator of a small village reverse osmosis unit ran out of acid, he would find that the unit would continue producing water. However, if within a short time, he did not correct the problem, the membranes could be ruined and water production would cease.

Many developing countries have a serious shortage of the type of skilled and/or trained labor necessary to operate a plant of this type. Without a nucleus of trained, skilled people, operation can be difficult and possibly disastrous depending on the type of plant. However, as Mexico has demonstrated, people with little formal education can become effective desalination plant operators if they are properly selected, trained, motivated, and technically assisted in their jobs. The Mexican experience is discussed later in this chapter.

Training must be considered for the long-term with a well thought out and adequately financed program. Desalination plant manufacturers can often assist in operator training, but this must be clearly specified in the contract. Training can be complicated by the lack of suitable texts and/or by the language problems which often exist.

Regularly scheduled visits by the manufacturer, or others, paid for in the original contract, to assist in training, troubleshooting, and maintenance over a period of years after installation can be a worthwhile investment. Because of the difficulty in finding or training suitable operating personnel in developing areas, the use of desalination processes or plants similar to those already existing locally is often advantageous. This operational advantage can often override possible extra costs, since the potential for providing proper operation is higher.

<u>Contract Operation</u>. In many developing areas where severe problems might be anticipated due to the lack of trained personnel and/or a suitable government organization to operate the desalination facilities, consideration should be given to the use of a private contractor to operate the facility.

This type of contractual arrangement is not uncommon for desalination plants. It is used in many areas of the Middle East such as the Sultanate of Oman, where the plants operated under contract range from the 7.2-mgd [27,300-m^3/d] plant at Ghubrah to a small 0.1-mgd [379-m^3/d] plant on the Island of Masira.

This type of operation can be expensive since many of the personnel are well paid and the operation must generally be paid for in foreign exchange. However, in competitive situations, this cost probably closely approximates the true cost (plus risk cost) of operations. A contract for this type of service must be written and executed carefully so as to result in the desired production and to protect the long-term operational capabilities of the plant.

Contract operation should generally be viewed as a temporary expedient for the operation of desalination facilities. It should include long-term planning for proper training of local personnel in these skills for eventual operation of the plant.

Organization

There must be an effective organization to operate the desalination facilities, including financial support for operating expenses, repairs, maintenance, etc. This type of organization should exist before the desalination facility is ready to operate. The organization's attributes and limitations must be appraised realistically in the process selection and design phases of the project.

Organization becomes especially important at the village level, since operation must often continue for long periods of time without outside assistance. Although Mexico is not classified as a developing country, the Mexican Government's performance in carrying out their desalination program is one which could serve as a model to many developing areas. This program is discussed in the next section of this chapter.

NATIONAL PROGRAMS

General

The technical aspects of desalination are rather specialized, and although most countries do not have many desalination facilities, where they do exist they represent a sizeable investment in funds as well as a vital part of the national economy.

In many countries, the functions associated with planning and/or operation of desalination facilities are grouped within one governmental department and/or agency. This centralizes technical expertise and aids in coordinating planning, purchasing, maintenance, training, etc. Occasionally, other agencies may share in this role as their activities warrant and/or overlap.

The following are three representative national governmental agencies which deal with desalination: the Saline Water Conversion Corporation (Kingdom of Saudi Arabia), the Ministry of Electricity and Water (Kuwait), and the Secretariat of Electricity (Libya).

One question which a desalination agency must deal with is the degree to which it will subsidize water sales to the public. If little or no subsidy is given then desalination facilities can be expanded as localized economic conditions warrant, since those regions will be financing their own operations. Where high subsidies are employed, future expansion (and in fact all operations) must depend on continued and increased financing from the central government.

Mexico

The Mexican Government established a central agency to direct its desalination programs in 1971. This agency, originally named CAAS, Comisión para el Aprovechamiento de Aguas Salinas (Commission for the Improvement of Saline Water) was shifted, as part of a Government reorganization in 1977 and renamed DIGAASES, Direccion General de Aprovechamiento de Aguas Salinas y Energia Solar.

Sixty-seven percent of Mexico's land area has a climate which is dry, ranging from semiarid to desert. Many areas in these dry regions have no readily available freshwater. However, in many of these areas ample supplies of brackish water and seawater exist, making desalination possible.

Figure 9-1 presents a simplified organization chart of the portion of DIGAASES concerned with desalination. DIGAASES's approach to desalination within Mexico is oriented toward the following goals:

1. Installation and operation of desalination units for potable water supply within the country.

2. Adaptation of desalination technology for local conditions.

3. Utilization of field experience to improve future design and operations.

4. Minimization of foreign exchange expenditures.

5. Increased investigation and research into desalination and its applications.

6. Development of trained personnel in the field of desalination.

DIGAASES has divided its efforts between supplying water to both urban and rural areas, expending particular effort toward developing desalination facilities in small villages. At the end of 1979, the agency was operating 33 desalination systems serving about 54,000 people (with a total installed capacity of about 1.15 mgd [4,350 m^3/d]) plus a number of experimental units. Most of the installations produce between 10,000 and 25,000 gpd [38 and 95 m^3/d]. To service its units and carry out experimental and research work, DIGAASES has centers in four areas outside of Mexico City, where the main headquarters are located.

DIGAASES's success with desalination has resulted from their approach, organization, follow-up, and ability to learn from experience in the field.

Village Program. DIGAASES has realized that a successful program of desalination in the village requires a special approach. They begin with a study of the village, in which they evaluate both the source of feedwater for desalination and any potential sources of potable water that could be used instead of desalted water. The study also includes the economic, social, and technical aspects to determine if there is a true desire for this type of facility, whether it will fit with the village social milieu, what its impact will be, and what type of local financial support, power sources, operator potential, etc., are available. Usually the local school teacher is an excellent source of information on many of the social and economic aspects. During this investigation period DIGAASES tries to maximize local involvement in the project, realizing that in the end the project will succeed only if the villagers want and support it.

A typical DIGAASES installation in a village includes: (1) a raw water source, (2) the desalination equipment in a building, (3) a power generating system (if power is not available), (4) a product water tank, and (5) a house for the operator (and family).

Figure 9-2 shows a typical DIGAASES installation in the village called Ejido 4 de Marzo (4th of March) in the State of Coahuila. This installation serves about 470 people and produces approximately 2,100 gpd [8 m^3/d]. The woman standing in the doorway of the desalination building is the operator. Reverse osmosis is used to treat brackish water with a TDS of about 3,000 mg/l. The water is distributed by villagers picking it up in buckets which are filled from taps on the product water tank.

Under present policy, DIGAASES tries to restrict its activities to desalination rather than becoming involved in subsequent water distribution within the villages. This enables them to concentrate their funding on their major mission, desalting. They do, however, make every attempt to encourage the villagers to organize in committees or cooperatives to handle the distribution of the water, and in many place, such as Ejido de 4 de Marzo, the most economical method is the use of buckets. In most cases, the local water committee supervises the collection of fees for the water. The fees are nominal and defray only a small portion of the total cost of the service. Success in fee collection varies greatly from region to region.

An important part of DIGAASES's success in the village program is their periodic assistance during operation. As part of their initial study, DIGAASES personnel select a local villager as an operator for the facility. The operator is given training to carry out the basic functions around the plant and is placed on DIGAASES's payroll. After the plant is placed into operation, an engineer and/or technician from DIGAASES usually visits the village about once a month to pay the operator; collect data on operation; assist in making repairs or modifications; deliver parts, chemicals, etc.; and generally assist the operator with the plant. DIGAASES personnel use the data collected to monitor the plant's performance, enabling them to spot and correct problems early and to use the information (especially on problems) to modify future designs, construction, and/or purchase specifications. Thus DIGAASES continues to profit technically from the experience gained in the operation of their field units while providing a high standard of service.

Mobile Units. In addition to the many small permanent units in villages, the Mexican Government has begun work on using mobile desalination equipment. This will enable them to service a number of smaller villages with one unit. The unit shown on Figure 9-3 has its own power generating facilities and is mounted on a trailer. It employs reverse osmosis for desalination, and the trailer includes all the necessary pretreatment equipment. The unit can be towed into a village and operated to fill up a freshwater tank with several days' supply. The unit then moves on to the next village, where it repeats the operation. By proper selection and design of the route, storage tanks, etc., a number of small villages can be successfully supplied with freshwater with a minimum number of operational personnel and problems with power, etc. As a result, all maintenance and repair work can be centralized.

Design and Construction. DIGAASES has played a continuing role in the design and manufacture of a variety of desalination systems. This has allowed them to: (1) adapt equipment for local operation and/or construction conditions (2) increase their knowledge of the desalination process, (3) incorporate information gained by field experience into new units, and (4) minimize foreign exchange expenditures.

This work has been concentrated primarily in four types of plants: multistage flash, vapor compression, reverse osmosis, and solar stills. In doing so, they have reduced the necessity for importation so that the percentages of parts for each process which can be fabricated in Mexico are as follows:

Type	Percentage
Multistage Flash	90
Reverse Osmosis	65
Vapor Compression	60
Solar Stills	100

Since 1971 considerable effort has gone into the design and manufacture of various multistage flash distillation plants by DIGAASES. Among the plants constructed have been the following:

Location	Capacity		Number of Stages
	(mgd)	[m^3/d]	
Punta Abreojos	0.026	100	12
Bahía Tortugas	0.08	300	16
La Paz (under construction)	0.26	1,000	20
La Paz (operated by CFE[a])	0.03	115	7
La Paz (operated by CFE[a])	0.03	115	7

Source: DIGAASES, 1979.

[a] CFE = Federal Electric Commission.

DIGAASES has developed a special design which allows them to utilize available fabricating machinery and a high percentage of locally made materials. In addition, their design allows easy transportation of fabricated components to isolated locations where cargo handling and assembly facilities are severely limited.

The photo on Figure 9-4 shows a part of DIGAASES's fabricating facility in Mexico City, where work is presently underway on the 0.26-mgd [1,000-m^3/d] plant for La Paz. This will be a once-through multistage flash plant to be built with the flash chamber and the condensers separated instead of combined (as is often the case). Figure 9-5 shows the field erection of the 0.08-mgd [300-m^3/d] plant at Bahía Tortugas and illustrates the advantages of the modular design for ease of assembly with a minimum amount of heavy equipment. The completed facility is shown on Figure 9-6.

After initially working with carbon steel, DIGAASES now uses 90/10 copper-nickel almost exclusively for their multistage flash plants. They have also assembled an experimental heat recovery evaporator and a vapor compression unit.

In the field of reverse osmosis, DIGAASES has assembled 19 plants using imported membranes. They are currently building a small (2,100-gpd [8-m^3/d]) experimental tubular membrane plant which utilizes membranes fabricated by DIGAASES.

Whether the construction of desalination plants by DIGAASES actually saves them money is not known with certainty, but it does provide them a chance to develop the technology and fabrication techniques within their own country; this alone can have considerable value.

Rural Filter

An outgrowth of DIGAASES's program to develop tubular membranes has been a device called a rural filter (Filtro Rural). DIGAASES attempted to make a porous support for their tubular membranes using baked clay. The support had certain properties which made it unsatisfactory for this purpose, but the research staff adapted it to work as a simple water treatment device that could be used in the rural areas to filter and purify small quantities of drinking water. A diagram of the rural filter is shown on Figure 9-7.

The porous clay body acts as a filter, and the activated carbon removes organic material and silver (a heavy metal), which is toxic to bacteria. The filters are placed in a large elevated jar of water and a siphon is started on the hose. The filtered water drips into another container placed at a lower level. The filter's components are constructed from locally produced materials at a cost of approximately 300 Mexican Pesos($14 U.S.).

Other Programs

As described in Chapter 7, DIGAASES is actively participating in several programs involving research into the use of alternate energy systems for desalination.

Summary

Figure 9-8 summarizes the organization of DIGAASES's activities which enable them to obtain and utilize feedback from their field operations to improve operation and design of both existing future facilities.

The national desalination program organized by DIGAASES has considerable merit. It has been accomplished with a moderate budget and has attracted a staff of competent, dedicated personnel. Developing countries which are interested in organizing a national program of this type would do well to study DIGAASES's program to see if portions of it are suitable for their situation.

Further information can be obtained from:

Dr. Javier Ibarra Herrera, Director General
DIGAASES
Boulevard del Pípila No. 1
Lomas de Tecamachalco
México City, D.F.
México

APPLICATIONS

The costs associated with desalinated water are sufficiently high that the applications for water are limited to those which can justify the economic considerations. Generally, desalination can be used for domestic purposes and some industrial uses, but only for very special agricultural usage. The latter generally requires water that is both abundant and inexpensive for normal cropping purposes. Since desalted water is generally neither, it is usually used only in association with special agricultural techniques and/or crops.

Thus, desalted water is generally used effectively in developing countries primarily for: (1) urban and industrial, and (2) village (or other rural) applications. These two major areas of application require somewhat different approaches.

Municipal/Industrial Desalting--General

These applications have many of the same general characteristics as similar uses in industrialized countries. The site-specific problems and limitations are discussed in both this chapter and in Chapter 8. Given proper planning and fulfillment of the four steps noted earlier, there is a reasonable chance for success.

Municipal Desalting--U.S. Virgin Islands

The U.S. Virgin Islands are made up of three major islands with a total area of about 140 square miles [363 million m^2] and a population in excess of 137,000 persons. The two largest islands, St. Croix and St. Thomas, depend heavily on the water produced by seawater desalination. The islands began to use distillation for desalination in 1964. Although the U.S. Virgin Islands is not classified as a developing country, its geographic setting, problems with water, and long experience with desalination make it a profitable case study for developing countries.

VIWAPA. Seawater desalination plants were installed for the production of potable water for the public and are operated by a government agency, the Virgin Islands Water and Power Authority (VIWAPA). As the name indicates, this agency also generates power.

i) Process Description--All of the plants built for VIWAPA are dual-purpose distillation plants operated in conjunction with the steam turbines used to generate electricity. The six plants, along with information on their supplier, design capacity, year of installation, and design type are listed below:

Plant No.	Location	Supplier	Design Capacity (mgd)	Design Capacity [m^3/d]	Year Installed	Type of Design
1	St. Thomas	Aqua-Chem	0.25	946	1964	MSF
2	St. Thomas	Westinghouse	1.0	3,790	1966	MSF
3	St. Thomas	BLH	2.5	9,460	1968	MSF
4	St. Thomas	Envirogenics	2.25	8,520	1975	VTE
5	St. Croix	Stearns Rogers	1.0	3,790	1969	LTV
6	St. Croix	Envirogenics	2.25	8,520	1975	VTE

MSF = Multistage Flash
VTE = Vertical Tube Evaporator
LTV = Long Tube Vertical

These are all high temperature plants and use acid for scale control. Until 1980 the total nameplate capacity of VIWAPA was 9.25 mgd [35,000 m^3/d], which made it one of the largest distillation organizations in the western hemisphere. A photo of the facilities on St. Thomas is shown on Figure 9-9. The four distillation plants on St. Thomas have been constructed around the electric generating station. The hillside behind the facility is covered with concrete and was used as a rainwater catchment to supply freshwater before the distillation plants were installed. In 1980 VIWAPA dismantled Plants No. 1 and 2 as part of a modernization program.

Distillation was selected by VIWAPA, as that was the only practical seawater desalination process in commercial use at the time of construction. Sufficient field experience with seawater reverse osmosis had not occurred by the time of procurement of the final Environgenics units. In addition, the availability of steam from the electric generating station makes a distillation plant theoretically the most inexpensive process to run due to the lower energy cost.

ii) <u>Operating Information</u>--Table 9-1 presents a summary of the production and on-line (operating) time for the six VIWAPA distillation plants. The table reveals that the overall performance of the facility is not particularly impressive. Although the demand for water exceeds the supply, the average annual plant factor during the fiscal years 1977-1979 has been about 0.3 versus the 0.85 to 0.9 often used by planners in making cost estimates for distillation facilities. During the last fiscal year the average production was about 2.3 mgd [8,710 m^3/d], versus an installed nameplate capacity of 9.25 mgd [35,000 m^3/d].

Problems have arisen with both design and operation. Many of the plants were prototypes--the first of that particular design to be built. Maintenance budgets and spare parts inventories have been generally low, thus adding to the downtime of many units. Moreover, several of the units have required major overhauls to restore them to operating condition.

Because of the high percentage of downtime, the actual cost of the water produced is extremely high since costs for items such as fixed charges, labor, and much of the steam continue whether or not the plants are operating. Thus in fiscal year 1979 the unit costs could be 3 to 4 times as high as expected.

This experience in the Virgin Islands attests to the validity of three important facts.

1. A primary consideration in the selection of any desalination plant should be its realistic plant factor.

2. Reliability of performance is extremely important.

Table 9-1
WATER PRODUCTION DATA FROM THE VIRGIN ISLANDS WATER AND POWER AUTHORITY'S (VIWAPA) DESALINATION FACILITIES

	F/Y 1977			F/Y 1978			F/Y 1979		
Plant Information	Time On-Line (hours)	Annual Production (MG)	Annual Plant Factor	Time On-Line (hours)	Annual Production (MG)	Annual Plant Factor	Time On-Line (hours)	Annual Production (MG)	Annual Plant Factor
ISLAND OF ST. THOMAS									
Plant No. 1 Design Capacity--0.25 mgd [946 m^3/d] Est. Actual Capacity--0.27 mgd [1,020 m^3/d]	7,763	71.7	0.78	2,731	21.7	0.24	1,459	11.9	0.13
Plant No. 2 Design Capacity--1 mgd [3,790 m^3/d] Est. Actual Capacity--0.7 mgd [2,650 m^3/d]	1,347	21.6	0.06	810	10.2	0.03	928	15.3	0.04
Plant No. 3 Design Capacity--2.5 mgd [9,460 m^3/d] Est. Actual Capacity--1.6 mgd [6,060 m^3/d]	4,480	223.3	0.24	1,868	73.2	0.08	5,965	287.4	0.31
Plant No. 4 Design Capacity--2.25 mgd [8,520 m^3/d] Est. Actual Capacity--1.8 mgd [6,810 m^3/d]	6,566	441.7	0.54	6,414	235.2	0.29	547[b]	41.1	0.05

Table 9-1 is continued on the next page.

Table 9-1--Continued

Plant Information	F/Y 1977			F/Y 1978			F/Y 1979		
	Time On-Line (hours)	Annual Production (MG)	Annual Plant Factor	Time On-Line (hours)	Annual Production (MG)	Annual Plant Factor	Time On-Line (hours)	Annual Production (MG)	Annual Plant Factor
ISLAND OF ST. CROIX									
Plant No. 5	116	3.4[b]	0.01[b]	3,506	41.7	0.11	4,500	91.4	0.25
Design Capacity--1 mgd [3,790 m^3/d] Est. Actual Capacity--0.7 mgd [2,650 m^3/d]									
Plant No. 6 Design Capacity--2.25 mgd [8,520 m^3/d] Est. Actual Capacity--1.9 mgd [7,190 m^3/d]	7,860	577.8	0.70	7,355	451.6	0.55	6,313	402.4	0.49
TOTAL	21,132	1,339.5	.40	22,684	833.6	0.24	19,712	849.5	0.25
TOTAL POSSIBLE[a]	52,704	3,385	1.00	52,560	3,376	1.00	52,560	3,376	1.00

Notes:
1. Information based on data furnished by the VIWAPA in a letter to Buros on 4/11/80.
2. Annual plant factor determined using design capacity.
3. Annual plant factor equals actual annual production divided by annual production at the design capacity.
4. Actual capacity estimated by VIWAPA.
5. F/Y = fiscal year July 1 through June 30 of the year specified.
6. Metric conversion factors are: mgd x 3,785 = m^3/d; MG x 3,785 = m^3.

[a] Total possible if all plants were run at design capacity 100% of the time. (Plant Factor = 1).

[b] Estimate by Buros of plant's production based on potential actual capacity.

3. The predicted efficiency (or performance) of a plant is meaningless if the plant does not run.

As a result of this poor performance, the islands have been forced to buy water from privately operated distillation plants and to barge water in from Puerto Rico. Both of these remedial solutions have proved costly.

An interesting sidelight to the difficult experiences has been the publication of a short book by one of the operators, G. R. Sheppard (1978), which describes, from an operator's point of view, some of the day-to-day problems and activities at a distillation facility. The book contains observations, comments, tips, and many miscellaneous bits of information that could be valuable to the operator and engineer.

Industrial Desalting--Egypt

Almost all of Egypt, except that portion along the banks of the Nile, is extremely arid. Although ground water is available in some of the desert areas, it is often extremely brackish. This lack of freshwater causes problems in obtaining potable water for various industrial and military installations in the country. Two examples of the use of desalination for the applications discussed in this section are for an oil campsite and a portable unit which furnishes water for the Egyptian Army.

El-Alamein (Vapor Compressor). This is an oil campsite run by the Western Desert Operating Petroleum Company. Brackish ground water is utilized to produce potable water for the camp personnel, who number around 50. Water is produced, stored in an elevated tank, and distributed by gravity to the various camp buildings. Brackish water is used for sanitary flushing purposes to minimize the use of potable water.

i) Process Description--The camp uses a vapor compression unit manufactured by MECO (New Orleans, USA). The unit is electrically driven with upflow vertical tubes and operates at a temperature of about 215°F [101.7°C]. The compressor is a centrifugal type and in the evaporator the major material is 90/10 copper-nickel; the tubes are made from arsenical aluminum brass.

Vapor compression was selected for this application for several reasons. One is the familiarity of the oil companies with the vapor compression process and its component parts. The other is the difficult nature of the ground water, which places it in the range where reverse osmosis can also have problems in efficient processing of the water. The vapor compression unit can be moved to other locations for use on a variety of water sources including seawater.

ii) Operating Information--The plant utilizes a brackish ground water with a high calcium content and a TDS of about 12,000. This difficult brackish water causes scaling on the heat transfer surfaces, thereby lowering the unit's overall performance. Various chemicals have been tried to alleviate the problem, but with only limited success. When the unit is not producing, water must be trucked in from nearby cities.

Sinai and Western Deserts (40,000 gpd [150 m^3/d], RO). The Egyptian Army employes two potable reverse osmosis units which are used to supply freshwater for drinking, cooking, and general kitchen activities at various locations in the deserts. The 40,000-gpd [150-m^3/d] units are installed in an enclosed 8-foot by 30-foot [2.4 by 9.1 m] trailer together with storage tanks for feedwater, brine, and product water. The units were made by Culligan Italiana in Italy, and the first one began operation in 1978.

i) Process Description--A flow diagram and photo of the interior of the unit are shown on Figures 9-10 and 9-11, respectively. The unit has extensive pretreatment processes to enable it to treat the variety of brackish waters that might be encountered. A portable diesel generator is towed behind the unit to provide electricity. The unit is designed for a 75 percent recovery with 95 percent salt rejection.

This unit is experimental in nature. The process was selected with its special pretreatment train enabling it to treat a variety of brackish ground waters.

ii) Operating Information--The units appear to be functioning, and it is reported that the troops greatly prefer the treated water to the untreated brackish supplies. The water from the reverse osmosis unit is blended with treated brackish water to produce a final product with a TDS of about 500 mg/l.

Village Desalting--General

This presents a special case for desalination applications. There is a great need in many areas for village desalination activities. In India, for example, a considerable number of villages in the States of Rajasthan and Gujarat have only brackish water available for drinking purposes. This same situation exists in many other parts of the world. A village, however, can be a very difficult place to successfully introduce a new technology such as desalination. Villages and villagers vary widely and each one must be approached within the context of its own particular culture.

One trait that is fairly common is that villagers tend to be conservative and hence, in general, shy away from risk. They have found over years of experience that certain habits, traditions, skills, etc., have endured, and they tend to be skeptical of changing these, especially at the urging of a stranger who may know little about their way of life. Risk-taking is difficult and dangerous when there is little with which to gamble and the penalties for failure are severe.

Water is a basic necessity of life and is recognized as such in villages. Changes made in this basic necessity can mean changes in the daily routine and/or ritual of life. In some villages, this may cause a disruption which interferes with the routine and leads to confusion and discord.

Introduction of Desalting. Thus, if efforts are made to introduce an "improved" water supply for the villagers they should be made with particular care if they are to have long-term success. The villages should want the improvement, approve of the idea of desalination, and participate in discussions of what capacity it will be, where it will be located, and what will happen to the brine. Additionally, they should participate in the construction and/or installation of the unit and its subsequent operation. They need to identify with the unit so that they are willing to keep it running.

It is crucial that adequate supervision be given during the installation and initial operating period. If the unit, for even the most insignificant reason, stops operating in the beginning, the villagers may become extremely skeptical of the whole concept.

After the initial operating period, additional visits should be made to monitor its performance and assist in its operation and maintenance.

Cost. Cost is an important factor. There is often very little money in a particular village, and the price to purchase fuel and/or parts may be difficult to pay. If the cost is paid by the central government, it is important that the parts and/or fuel come in a timely fashion so that the unit continues to function. If it ceases to function for a time and water is needed (as it is every day), the villagers may judge the unit undependable and are apt to return to their previous methods.

Technology and Sociology. The problems of village desalting are both technical and sociological. Technically, a unit must function extremely dependably, with minimal maintenance and operator attention. Although energy efficiency is a consideration, it is not as important as reliability. Sociologically, the ideal situation is for the desalination unit to complement--or at least only minimally disrupt--village life. Some examples of the disruptions that could occur include the following:

1. The introduction of strangers (possibly central government employees) into the village to install, operate, and/or service the unit.
2. The change in patterns of meeting, working, and talking that might occur if the source of water for the village is changed in type and/or location.
3. The question over which sex should service and operate the desalting unit. In many societies obtaining water is traditionally a woman's job, whereas operating

machinery is a man's. Based on this, the question of who should operate a solar still or similar equipment which may involve both tasks can cause problems.

4. The installation of a community unit where traditions are not oriented toward community organization.

5. The decision or mechanism involved in allotting water produced by the desalting device.

Sociological problems are very site- or region-specific and vary considerably. However, they do exist and must be addressed.

<u>Village Desalting Programs</u>. Programs for village desalting require considerable planning to achieve long-term success. If they are to be accomplished with the cooperation of the villagers, and not by fiat, they demand considerable time and substantial funds. The cost per person served is bound to be considerably higher than for desalination projects in urban areas, where many thousands of people can be serviced with one installation. The program being carried out by the Mexican Government has the elements of a successful program. It is not inexpensive but it serves a social and technical function that is desired by the central government--and it works.

<u>Village Desalting Equipment</u>. The type of equipment used will vary with the characteristics of the specific location and the amount and type of technical and financial support available. If technical support is generally good (as in Mexico) rugged standard units, of the best process for the application, can be used.

However, if the intention is to let the villagers "go it alone" with minimal assistance after installation, serious consideration should be given to the use of solar stills where feasible. Although the capital cost is rather high and the structure requires considerable space, it is technically rather straightforward in construction and operation. It also allows for maximum local participation, operation, and repairs. Various designs, construction techniques, pitfalls, etc., are discussed in the U.S. A.I.D. book, <u>Fresh Water From the Sun</u> (Dunham, 1978).

Another potential apparatus for village desalting could be a specially designed reverse osmosis unit. This would have a low efficiency and low recovery but would require minimum attention, chemicals, etc. A commercial unit of this type that has been thoroughly proven in the field does not exist at present, due probably to the lack of a significant market for this type of unit; hence manufacturers have little incentive to design or manufacture such a unit. If there were a substantial market it is very probable that a unit would be developed.

It should be remembered, though, that reverse osmosis units continue to require fuel, filters, and other expendables for the life of the facility. Although alternate energy sources such as wind, solar, wave, etc., can replace the fuel component, these technologies are not (as discussed in Chapter 7) yet applicable at the village level. However, the progress of the village applications discussed in Chapter 7 should be followed so as to ascertain their long-term results.

Village Desalting--India

The ground water in the arid and semi-arid areas of the states of Gujarat and Rajasthan is used heavily for irrigation, and there are significant salinity problems in much of the area. The salinity problems, in turn, affect the village water supplies, especially during periods of drought.

In the mid-1970's the Central Salt & Marine Chemicals Research Institute (CSMCRI) in the State of Gujarat constructed and installed several desalination units in various small villages in the region as a practical demonstration of their utility in providing good water to the villages. These units included electrodialysis, reverse osmosis, and solar stills.

Motagokharwala (3,100 gpd [12 m^3/d], ED). An electrodialysis unit was installed and operated for about 1 year in this village, with a population of about 1,000, in the State of Gujarat.

i) Process Description--A 3,100-gpd [12-m^3/d] electrodialysis was unit selected. It was made using interpolymer type membranes manufactured in India. Feedwater was obtained from the village well, and the product water was distributed with the use of large water pitchers within the village.

The level of TDS of this water is in the range where either electrodialysis or reverse osmosis could be used. CSMCRI had both types of units available for their research program and placed them in the selected villages as available.

Data on the membranes used are presented below:

	Membrane	
	Cation Exchange	Anion Exchange
Ionogenic Groups	Sulphonic	Quaternary Ammonia
Ionic Form	Na^{+1}	Cl^{-1}
Thickness (mm)	0.1 to 0.12	0.1 to 0.12
Water Content (%)	35 to 40	15 to 20
Exchange Capacity (meg/gm)	1.8 to 2.0	1.8 to 2.0
Electrical Resistance (ohm/cm^2)	1-2	2-3

Note: Data from Mehta et al., 1979.

A photo of this plant appears as Figure 6-1 in the chapter on electrodialysis.

ii) Operational Information--The unit reduced the TDS from 4,000 mg/l to 1,000 mg/l with an energy usage of about 11.5 kWh/kgal [3 kWh/m^3]. Intermittent dosage of the concentrate stream with acid was used to control scaling. The plant was operated by a villager who was trained by CSMCRI. There was reported to be no acceptance problem among the villagers, who approved of its use for drinking and cooking (CSMCRI, 1980).

Rajasthali (4,000 gpd [15 m^3/d], RO). In this village with a population of about 1,000 people, a tubular reverse osmosis unit, designed, manufactured, and operated by CSMCRI, was installed for about 6 months. As in Motagokharwala, the village well produced brackish water which had been used by the inhabitants.

i) Process Description--The reverse osmosis plant had a capacity of about 4,000 gpd [15 m^3/d] and it used cellulose acetate tubular membranes. The membranes in tubular supports were mounted in parallel holders and stacked above a central collection tray. The permeate moved through membranes and porous supports and dripped into the collecting tray, where it was transferred by gravity to a small product storage tank.

The level of TDS of this water is in the range where either electrodialysis or reverse osmosis could be used. CSMCRI had both types of units available for their research program and placed them in the selected villages as available.

A photo of a CSMCRI tubular test unit similar to that used at Rajasthali is shown as Figure 9-12.

ii) Operational Information--The unit was operated by a villager who was trained by the Institute. The energy usage was about 23 kWh/kgal [6.1 kWh/m^3] of product water produced. The TDS was reduced from 4,000 to 600 mg/l. Acid dosing was used intermittently to control scaling in the membranes (CSMCRI, 1980).

The villagers were reported to be enthusiastic about the quality of the water. They remarked on the reduced quantities of soap used for washing and sugar used with their tea (Mehta et al. 1976).

Village Desalting--Mexico

The Mexican Government, through its central desalination agency, DIGAASES, has installed and operated a number of small desalination units in villages in various parts of the country. The activities of DIGAASES were discussed earlier in this chapter. Table 9-2 lists some of these village installations where the population is less than 1,000 people. The size of the units depends on other water resources available, general need, etc. One of these typical village units is described in further detail below.

Table 9-2
SOME MEXICAN VILLAGES WITH POPULATIONS LESS THAN 1,000 SERVED BY DESALINATION FACILITIES

Location Village	State[a]	Capacity (gpd)	Capacity [m^3/d]	Type of Unit[b]	Population Served
Bahía Asuncion	BCS	26,400	100	RO	900
Bahía de Los Angeles	BC	10,500	40	VC	650
Ejido 4 de Marzo	Coah	2,100	8	RO	468
Ejido 28 de Agosto	Coah	2,100	8	RO	700
Ejido El Sol	Coah	2,100	8	RO	702
Ejido La Leona	NL	2,600	10	RO	500
Estanque Palomas	Coah	2,600	10	RO	707
Isla Margarita	BCS	16,000	60	VC	862[c]
Isla Natividad	BCS	10,500	40	VC	300
Puerto Chale	BCS	260	1	SS	103
San Cayetano de Vacas	NL	2,600	10	RO	490

Source: Data from DIGAASES, 1979.

[a] BC = Baja California; BCS = Baja California del Sur; Coah = Coahuila; NL = Nuevo Leon.

[b] RO = Reverse Osmosis; SS = Solar Still; VC = Vapor Compression.

[c] Two 16,000-gpd [60-m^3/d] vapor compression units are located on Isla Margarita.

Bahía Asuncion (26,400 gpd [100 m^3/d], RO). This village is located on the southern part of the Baja California peninsula. The village used brackish water for drinking before a reverse osmosis desalination unit was installed in 1973. Presently the 900 villagers use the desalted water for all of their drinking needs but brackish water for some other household activities. The information on this installation was furnished by DIGAASES (Rodriguez, 1980).

i) Process Description--The flow diagram for this unit is shown in Figure 9-13. The unit, manufactured by Permutit (USA), has a capacity of 26,400 gpd [100 m^3/d] and and utilizes three hollow fine fiber permeators (DuPont B-9). Power is furnished by a 35 kW generator. The feedwater for the plant has a TDS of about 2,350 mg/l.

Although electrodialysis could be used for an installation of this type, DIGAASES has had considerable experience and expertise in operating reverse osmosis units. Hence they select and utilize reverse osmosis in almost all of their brackish water applications.

The product water is sold at a wholesale rate to a village committee which takes care of the resale and distribution of the water to the villagers.

ii) Operational Information--A summary of data for this facility is shown in Table 9-3. Difficulty in operating the facility arose immediately following installation due to problems with the electrical generator and too little feedwater. These were corrected in 1974 and 1975, and the plant's production noticeably improved during 1976. The replacement of the permeators about halfway through their expected life has been a major problem.

The operator is furnished a house as part of the compensation. In the event of technical problems, the operator can contact DIGAASES by a radio; otherwise a DIGAASES technician visits the facility every month.

Table 9-3

OPERATING DATA FOR THE BAHÍA ASUNCION REVERSE OSMOSIS PLANT[a]

	1975	1976	1977	1978	1979
Operation During the Year					
Number of Days	331	356	356	352	331
Total Hours	3,902	5,240	7,187	7,416	6,838
Total Production (MG)	4.33	6.17	6.10	6.59	5.92
[m^3]	16,300	23,400	23,100	24,900	22,400
Plant Factor	45	64	64	69	62
Demand/Production					
Estimated Daily Demand (kgal)	--	--	17.2	17.2	17.1
[m^3]			65.1	65.1	64.7
Average Daily Production (kgal)	--	--	16.7	18.1	16.2
[m^3]			63.2	68.5	61.3
Consumption of Supplies					
Acid (gal/kgal [l/m^3] product)	--	--	0.136	0.135	0.121
Fuel (gal/kgal [l/m^3] product)	--	--	3.06	2.473	2.735

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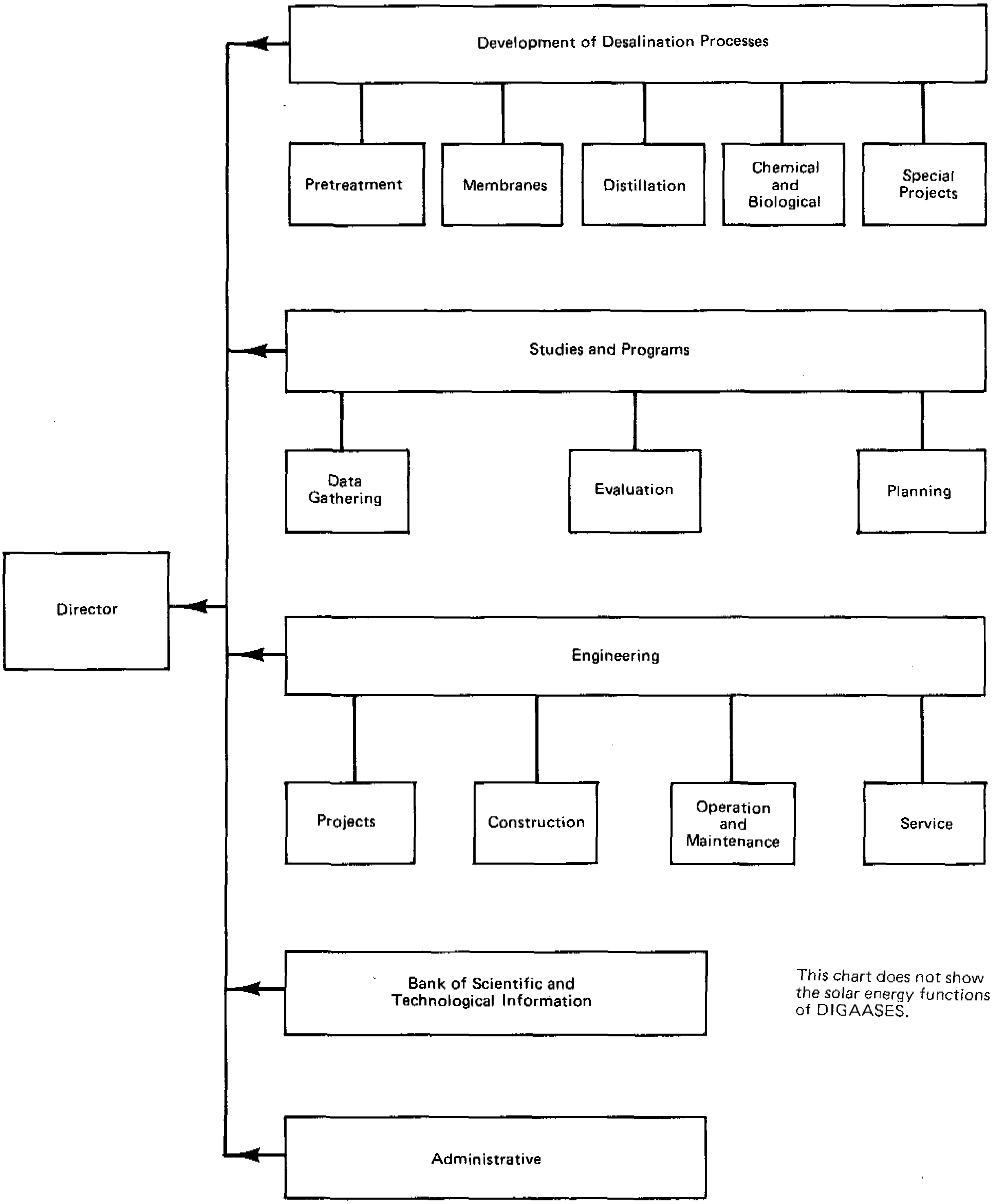
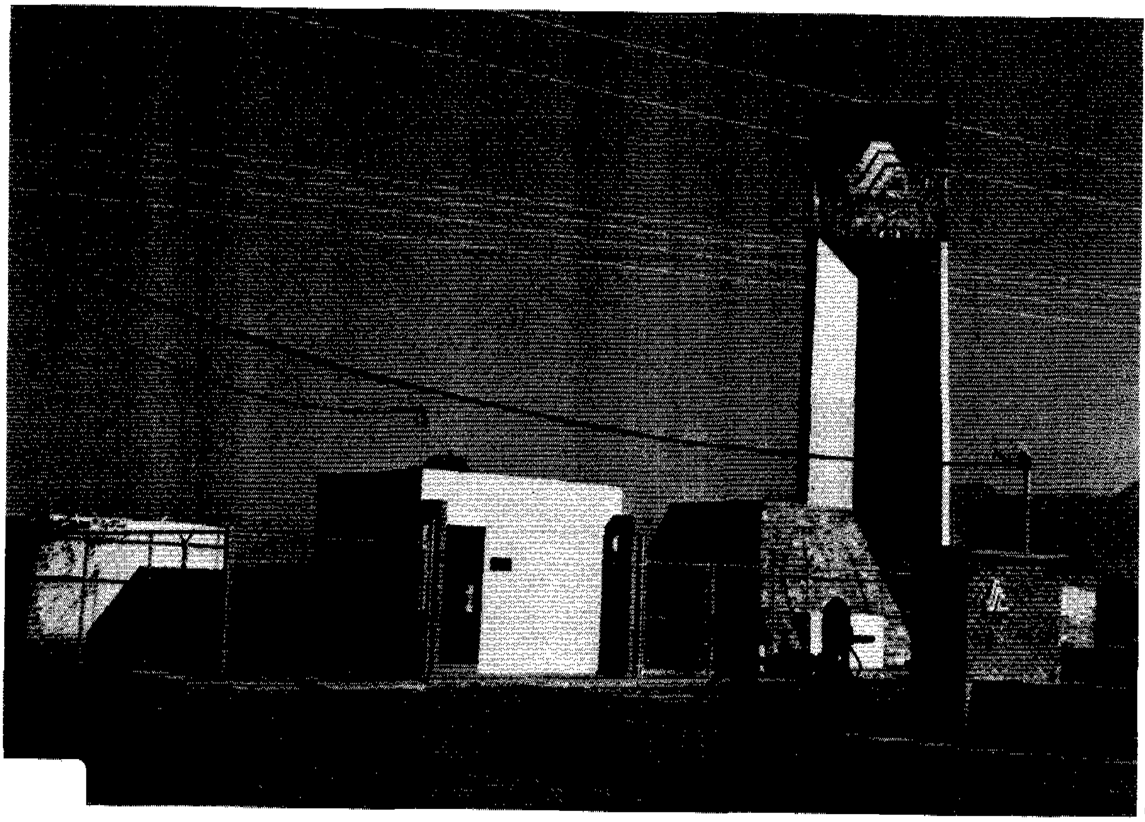
[a] Data from DIGAASES (Rodríguez, 1980).



This chart does not show the solar energy functions of DIGAASES.

DIGAASES (Direccion General de Aprovechamiento de Aguas Salinas y Energia Solar) is the Mexican Government's Desalination Agency.

FIGURE 9-1. Simplified organizational chart for DIGAASES.



Photos Courtesy of Direccion General de Aprovechamiento de Aguas Salinas y Energia Solar (DIGAASES).

FIGURE 9-2. Reverse osmosis facility at the Mexican Village, Ejido 4 de Marzo.

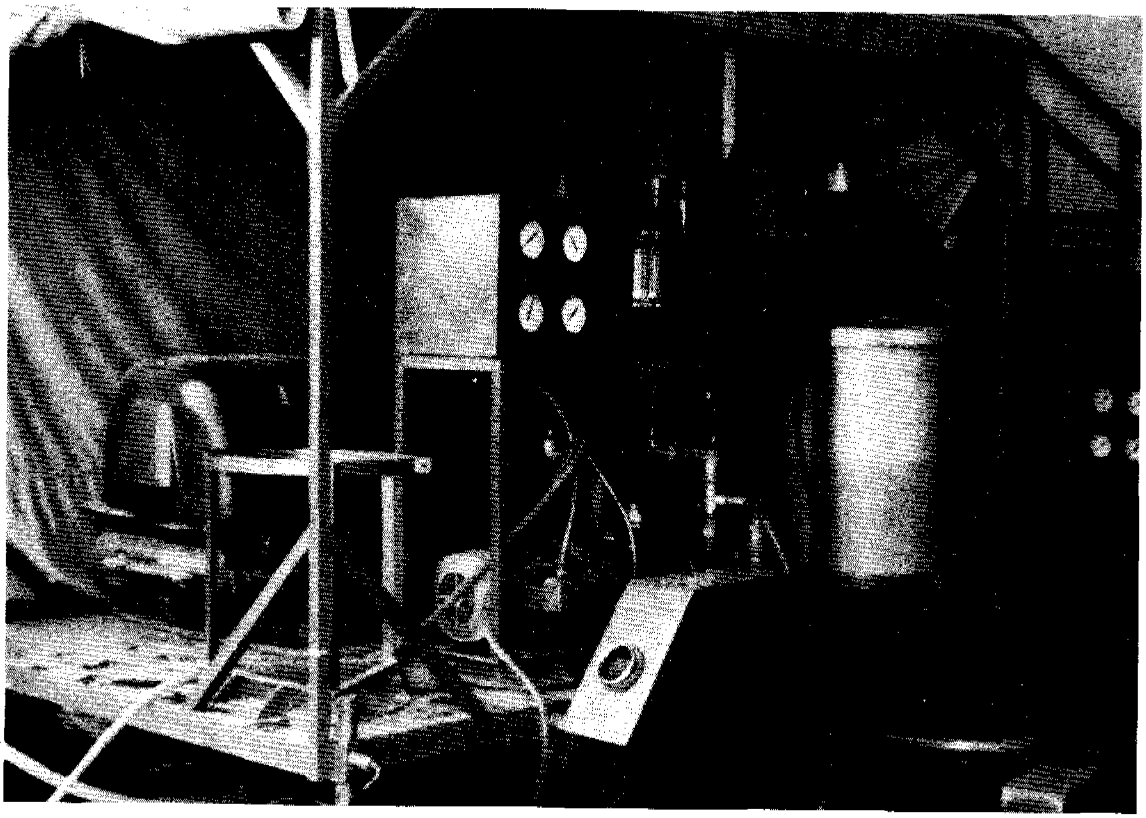


Photo Courtesy of O. K. Buros.

FIGURE 9-3. Portable reverse osmosis facility used to serve villages in Mexico.

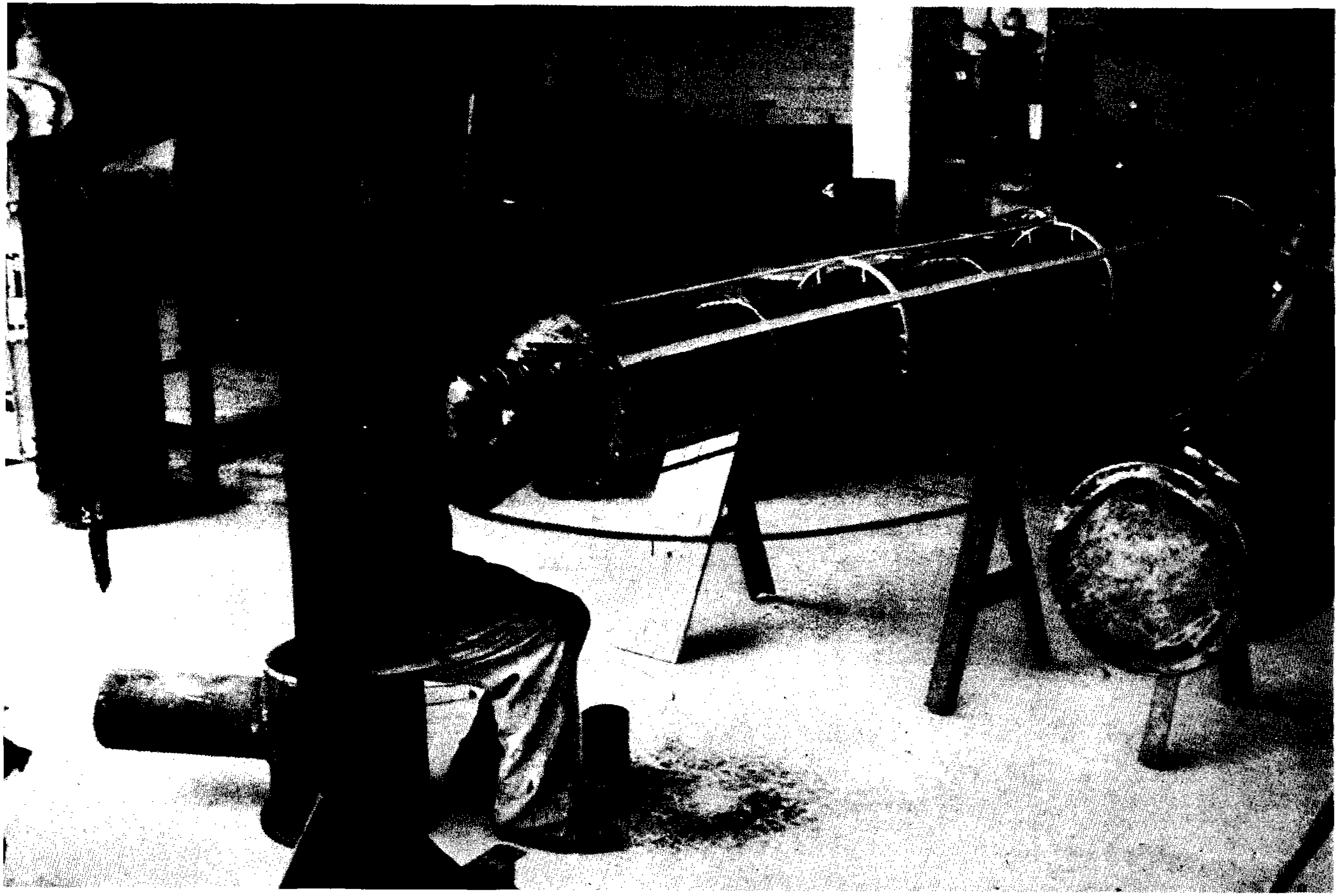


Photo Courtesy of Direccion General de Aprovechamiento de Aguas Salinas y Energia Solar (DIGAASES).

FIGURE 9-4. DIGAASES's desalination plant fabrication facility in Mexico City.

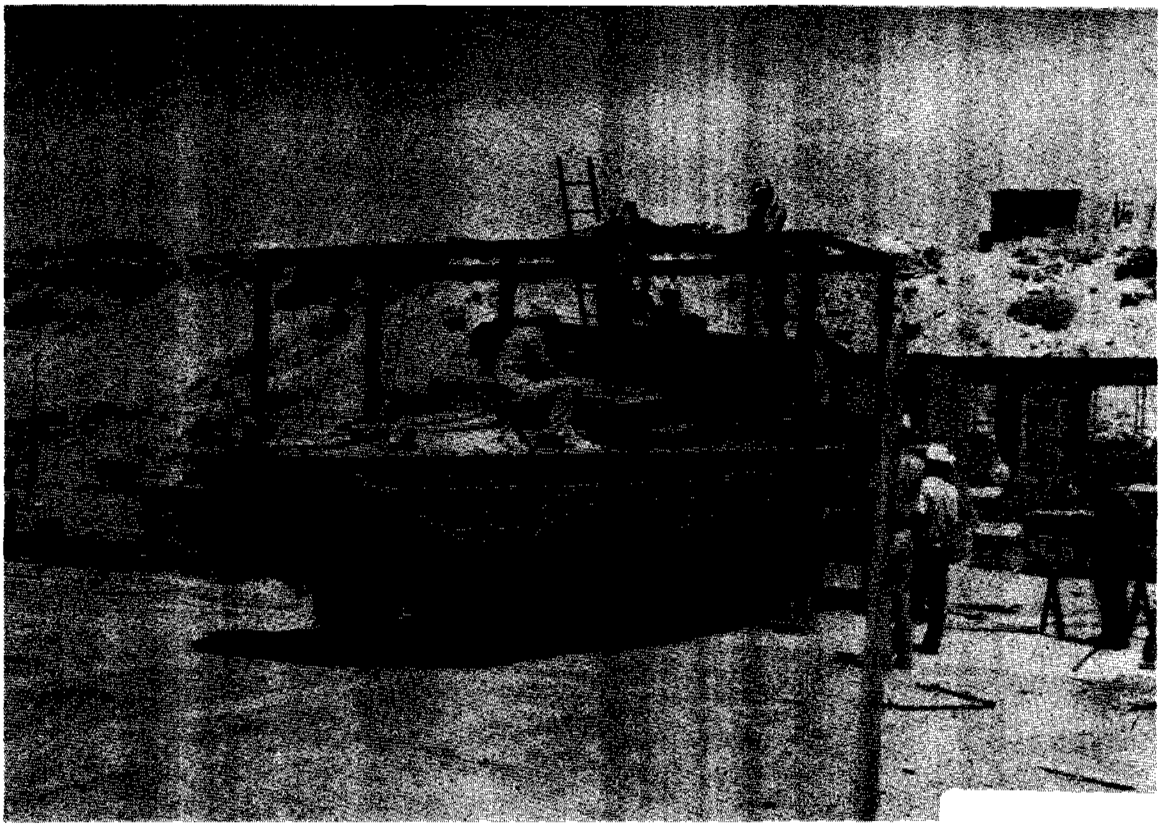


Photo Courtesy of Direccion General de Aprovechamiento de Aguas Salinas y Energia Solar (DIGAASES).

FIGURE 9-5. Field erection of an 0.08-mgd [300-m^3/d] distillation plant at Bahía Tortugas.

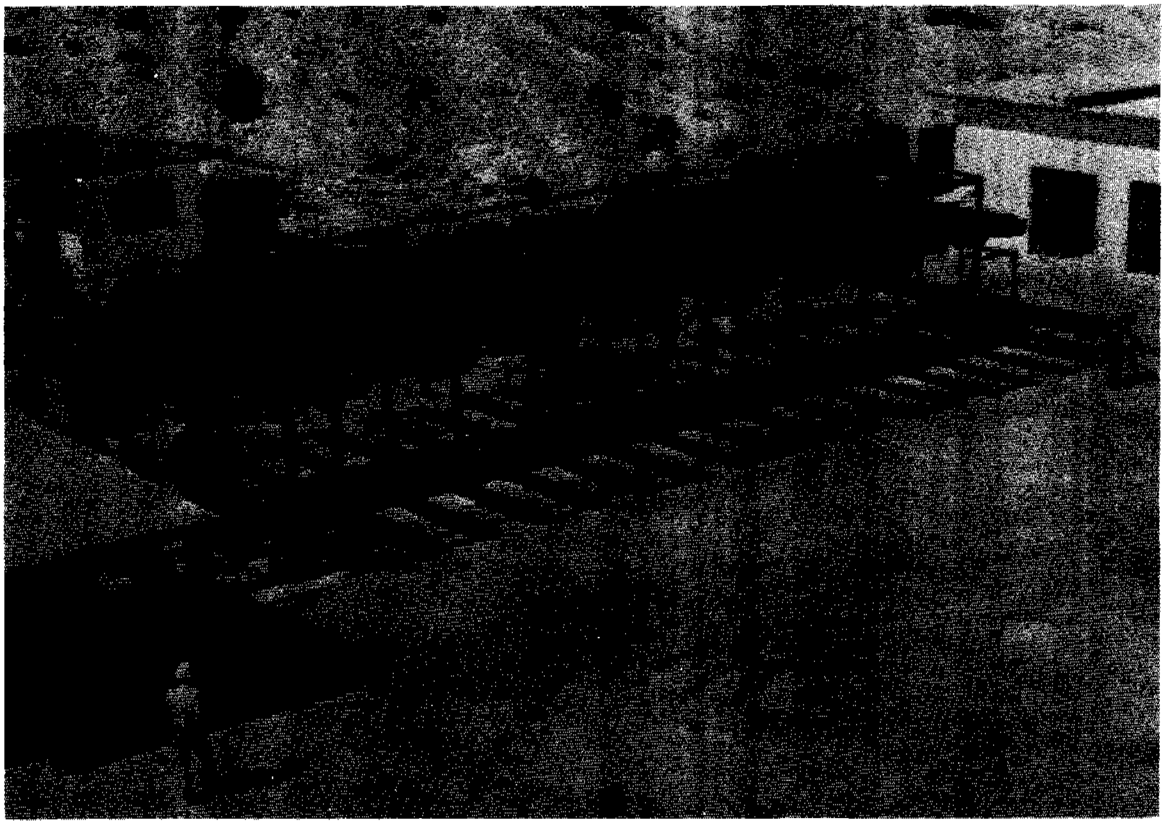
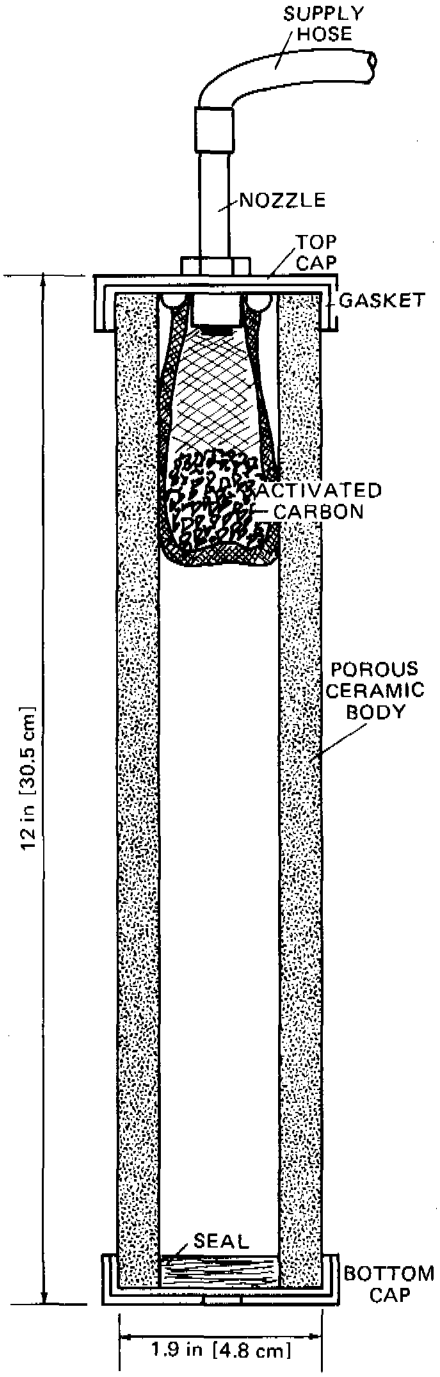


Photo Courtesy of Direccion General de Aprovechamiento de Aguas Salinas y Energia Solar (DIGAASES).

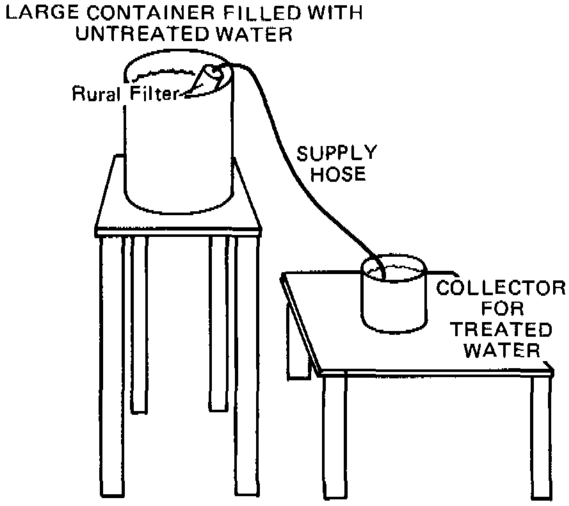
FIGURE 9-6. Completed 0.08-mgd [300-m^3/d] distillation plant at Bahía Tortugas.



CROSS SECTION

The major component of the rural filter is a ceramic cylinder 12 inches long x 1.9 inches [30.5 x 4.8 cm] in diameter.

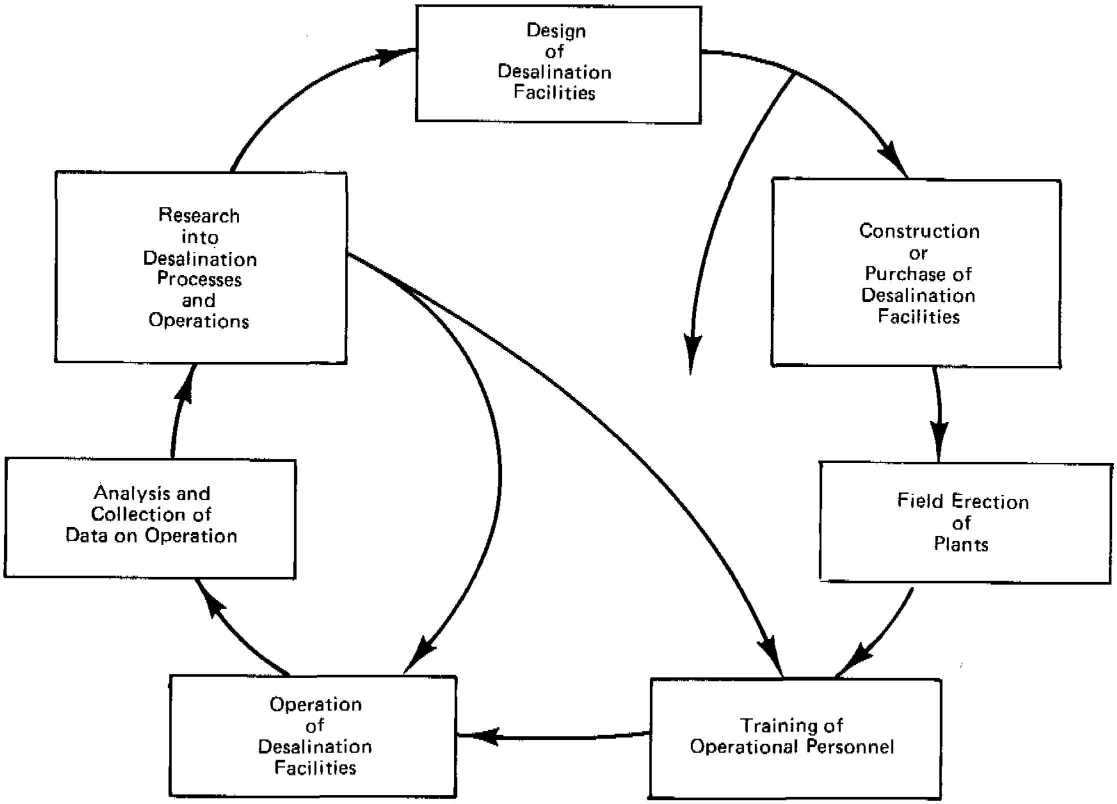
This figure was adapted from the bulletin "Como Usar el Filtro y Purificador Rural" by DIGAASES.



Rural Filter In Use

DIGAASES (Direccion General de Aprovechamiento de Aguas Salinas y Energia Solar) is the Mexican Government's Desalination Agency.

FIGURE 9-7. Details of the DIGAASES rural filter.



DIGAASES (Direccion General de Aprovechamiento de Aguas Salinas y Energia Solar) is the Mexican Government's Desalination Agency.

FIGURE 9-8. Technology interaction in the DIGAASES organization.

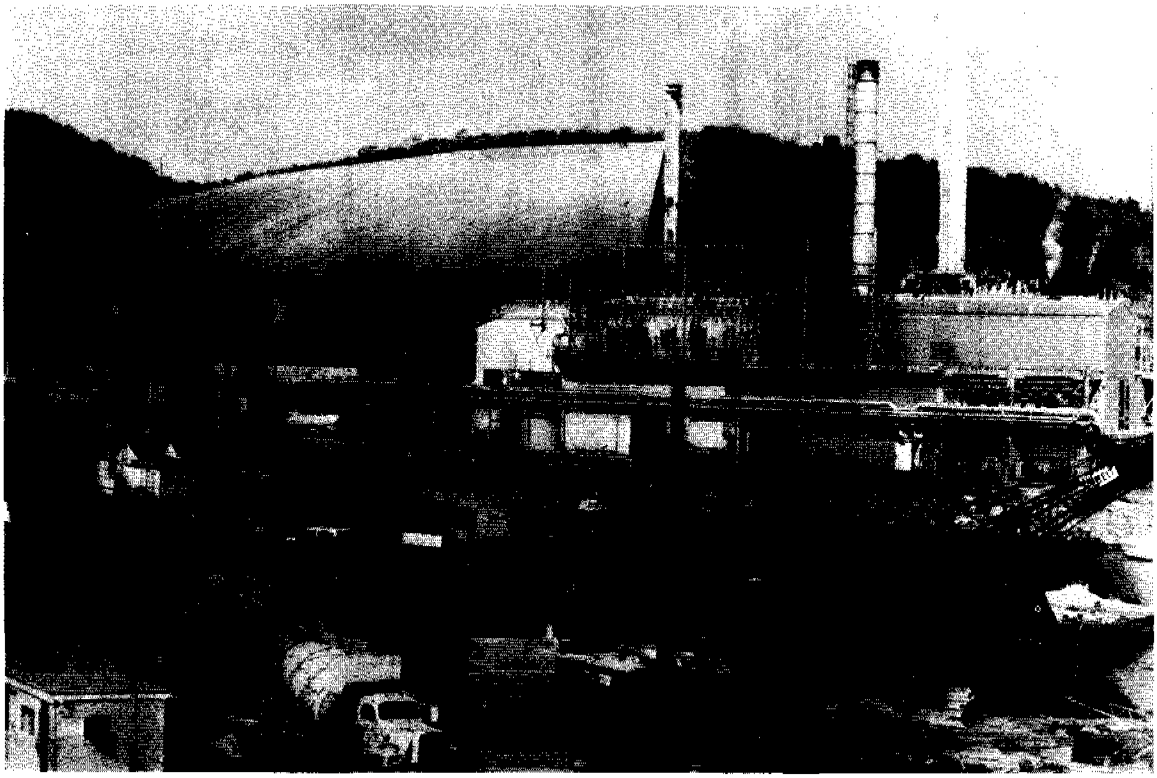


Photo Courtesy of O. K. Buros.

FIGURE 9-9. Distillation facilities at St. Thomas.

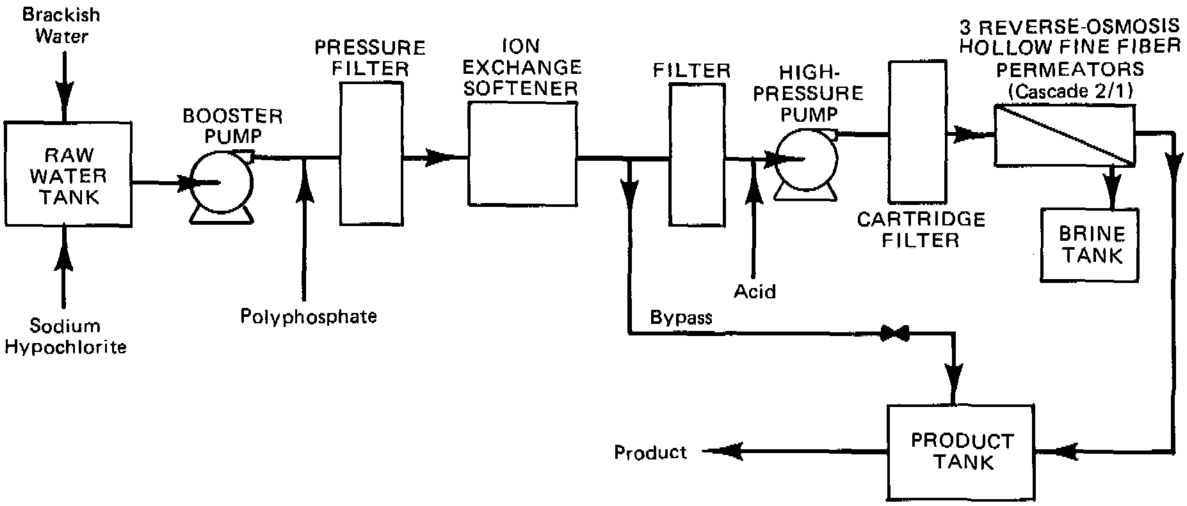


FIGURE 9-10. Flow diagram of a portable reverse osmosis unit in Egypt.

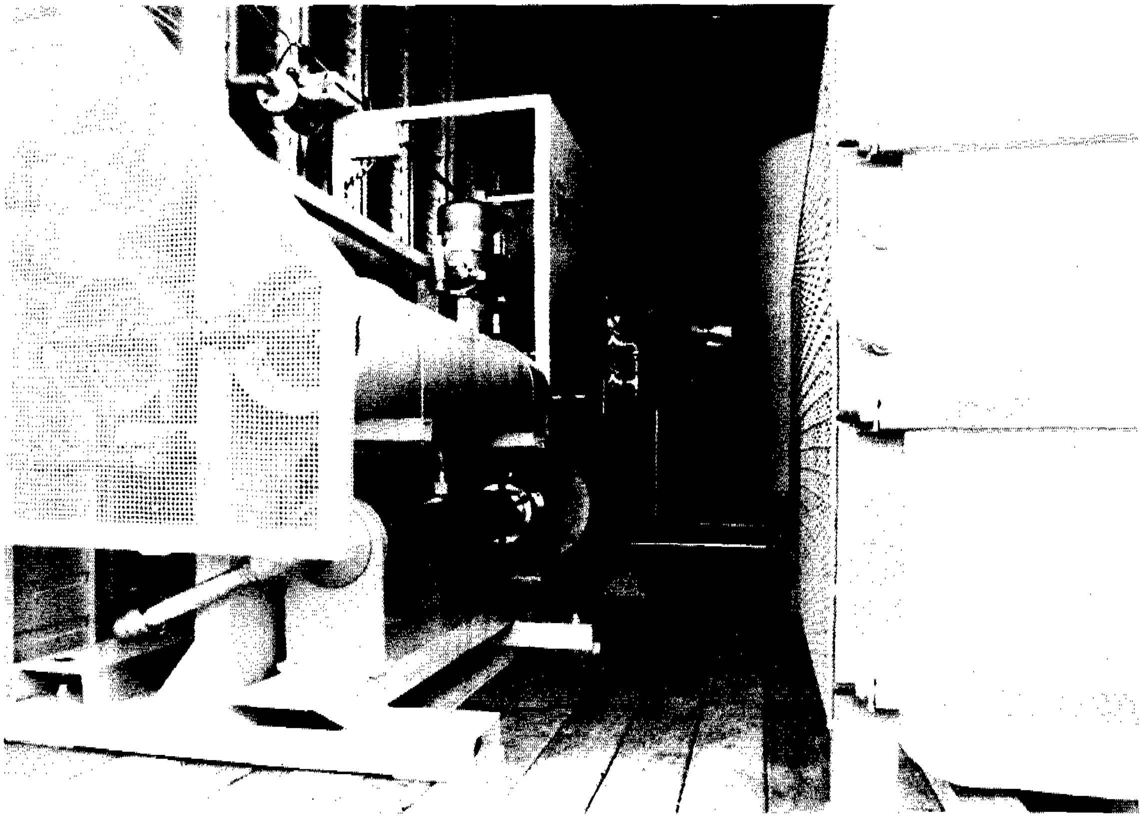


Photo Courtesy of Culligan International Co.

FIGURE 9-11. Interior view of a portable reverse osmosis unit in Egypt.

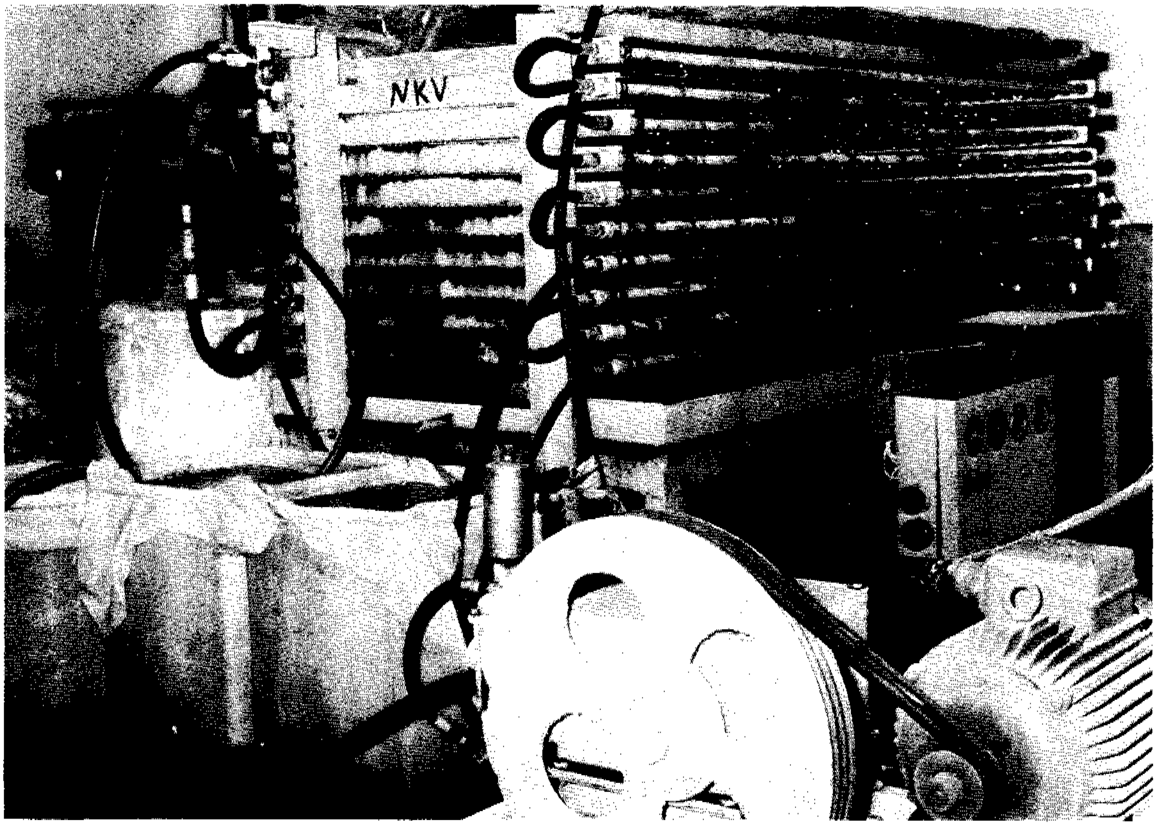


Photo Courtesy of Central Salt & Marine Chemical Research Institute, Bhavnagar, India.

FIGURE 9-12. A 4,000-gpd [15-m^3/d] tubular reverse osmosis unit developed in India.

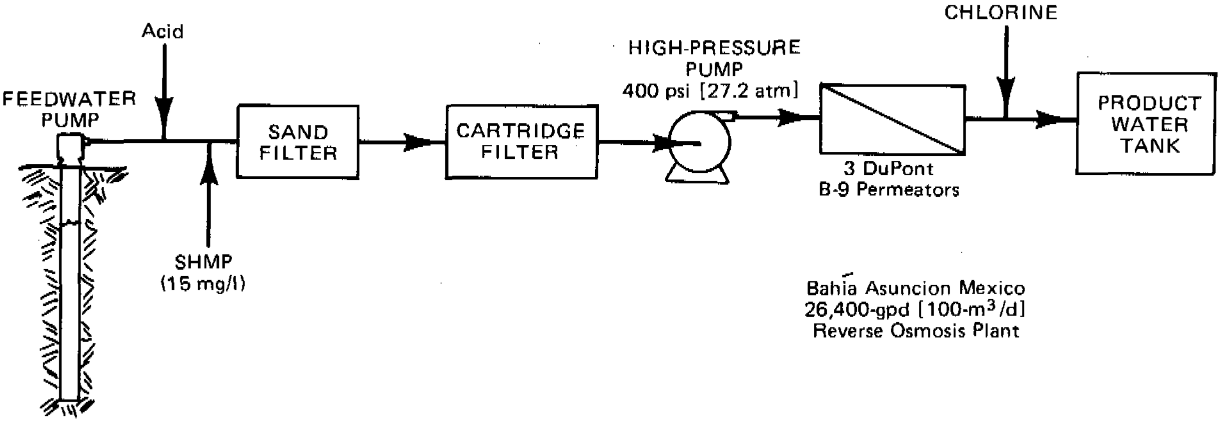


FIGURE 9-13. Flow diagram for the reverse osmosis plant at Bahía Asuncion.

APPENDIXES

Appendix A
LIST OF MANUFACTURERS

Manufacturers involved in desalination in the USA were surveyed by mail as part of this manual. The responses were used to complete the information on Table A-1.

The list is for general information only, since it does not represent all manufacturers. The inclusion or exclusion of any entity is not to be construed as approval or disapproval by U.S. A.I.D. nor the authors of the manual.

Table A-1

INFORMATION ON DESALINATION UNITS AVAILABLE FROM MANUFACTURERS IN THE USA

Manufacturer	Type of Desalination Units Produced	In the Past 3 Years			Capacity[a] Installed (mgd)	Comments
		Largest kgal/day	Smallest kgal/day	Typical kgal/day		
Ajax International Corporation P.O. Box 26607 San Diego, California USA 92131 Telephone: 714/578-3310	Reverse Osmosis	--	--	--	--	--
Allied Water Corporation 213 East Harris South San Francisco, California USA 91731 Telephone: 213/573-9220 Telex: 698706	Reverse Osmosis brackish and seawater	2	0.04	--	--	Standard compact units (40 to 1,750 gpd [0.15 to 6.6 m^3/d]) for land-based or shipboard applications. Can be furnished to operate on a variety of electrical power sources in a wide range of climatic conditions. The 40-gpd [0.15-m^3/d] unit could be used with a windmill furnishing DC power. Service manuals use photos extensively for minimal translation.
Aqua-Chem, Inc. P.O. Box 421 Milwaukee, Wisconsin USA 53201 Telephone: 414/962-0100 Telex: 26-679	Reverse Osmosis					Broad range of desalination equipment. Many self-contained units available. Capacity installed uncludes that under contract.
	brackish water	50	0.2	10	1	
	seawater	15	0.2	10		
	Vapor Compression	250	2.4	15	10	
	Flash Evaporators	3,750	12	--	20	
Aqua Media, Ltd. 1049 Kiel Court Sunnyvale, California USA 94086 Telephone: 408/734-8181 Telex: 34-6405 Cable: AQUAMEDIA	Reverse Osmosis					Standard package units for brackish water with a variety of pretreatments available. Used to supply hotels, condominiums, and camps. Seawater units are being tested and developed.
	brackish water	2,300	0.25	250	12	
	seawater	10	0.1	0.1	0.001	
Basic Technologies, Inc. 3879 West Industrial Way Riviera Beach, Florida USA 33404 Telephone: 305/848-1717 Telex: 80-3439	Reverse Osmosis				--	Line of membrane processing equipment. Some standard package units for brackish and seawater treatment. Custom-designed larger sizes. Low-pressure softening units also produced.
	brackish water	300	0.07	60	3+	
	seawater	12	1	1	3+	
	softening	360	10	90	3+	
Continental Water Systems Corp. 12400 Darrington P.O. Box 20018 El Paso, Texas USA 79998 Telephone: 915/852-9090 Telex: 749-491 Cable: CONTINENTAL ELP	Reverse Osmosis					Custom-design to fit each application. Research in the use of ultrafiltration membranes as a pretreatment for RO.
	brackish water	60	0.4	--	0.50	
	seawater	40	0.25	--	0.05	
Culligan International 1 Culligan Parkway Northbrook, Illinois USA 60062 Telephone: 312/498-2000 TWX: 910 686 0005 Cable: CULLINC	Reverse Osmosis					Builds both small self-contained units and larger units to order. Many overseas dealers for Culligan water treatment equipment.
	brackish water	160	0.005	1 to 10	12	
	seawater	20	0.25	10	0.08	

Table A-1--Continued

Manufacturer	Type of Desalination Units Produced	In the Past 3 Years			Capacity Installed (mgd)[a]	Comments
		Largest kgal/day	Smallest kgal/day	Typical kgal/day		
Envirogenics Systems Company 9255 Telstar Avenue El Monte, California USA 91731 Telephone: 213/573-9220 Telex: 698706	Reverse Osmosis seawater and brackish water Distillation Multistage Flash Vertical Tube Evaporators	--	--	--	3+	Large international corporation owned by Sogex International. Manufactures reverse osmosis membranes, assembles units, and designs, manufactures, and installs distillation plants.
Fluid Systems Division of UOP 2980 North Harbor Drive San Diego, California USA 92101 Telephone: 714/299-9920 Telex: 910/335-1193	Reverse Osmosis seawater brackish water	 3,200 5,000	 4 4	 -- --	 3.31 9.59	Produces membranes, assembles complete units, and installs in the field.
Hydranutics Water Systems 6338 Lindmar Drive Santa Barbara, California USA 93017 Telephone: 805/964-7747 Telex: 658445 HYDRA GLTA	Reverse Osmosis brackish water	3,000	--	--	7+	Produces membranes and assembles and installs units. Membranes are spiral-wound cellulose acetate. Produces units for both potable and industrial use, many located in harsh environments.
IONICS, Inc. 65 Grove Street Watertown, Massachusetts USA 02172 Telephone: 617/926-2500 Telex: 92 2473	Electrodialysis Electrodialysis--Reversal	4,000	0.5	100	20+	Electrodialysis-reversal plants use DC power (usually rectified from an AC source), low pressure, no heat, and no operating chemicals.
L*A/Water Treatment Division Chromalloy America Corporation 17400 East Chestnut Street City of Industry, California USA 91749 Telephone: 213/912-5411 Telex: 67-0482	Reverse Osmosis brackish and seawater	--	--	--	--	Prepackaged reverse osmosis systems for seawater and brackish water system. Have developed the concept of a desalination ship which would involve all reverse osmosis equipment on board ship to supply shore locations with freshwater.
Mechanical Equipment Co., Inc. 861 Carondelet Street New Orleans, Louisiana USA 70130 Telephone: 504/523-7271 Telex: 058-388 & 682161 Cable: MECO	Vapor Compression Waste Heat Reverse Osmosis	170 10 per demand	1.3 2.4 1	10 to 50 7 15	2 0.15 0.15	Diesel units, all electric and offshore. Operates off engine jacket water or low-pressure steam. Brackish water and seawater.
Neptune Microfloc 1965 Airport Road Corvallis, Oregon USA 97330 Telephone: 503/754-7654	Reverse Osmosis brackish and seawater	--	--	--	--	Assembles and installs brackish and seawater units.

Table A-1--Continued

Manufacturer	Type of Desalination Units Produced	In the Past 3 Years			Capacity[a] Installed (mgd)	Comments
		Largest kgal/day	Smallest kgal/day	Typical kgal/day		
NOAH Water Systems P.O. Box 1620 Santa Cruz, California USA 95061 Telephone: 408/425-5117	Reverse Osmosis seawater	0.35	0.19	0.25	--	Three standard units (192, 250, and 350 gpd) [0.7, 1, and 1.3 m^3/d] are built for shipboard use, requiring minimum operational skills and maintenance. Developing a wind-powered option for the units.
Metito International, Inc. 340 North Belt East Houston, Texas USA 77060 Telephone: 713/931-6115 Telex: 791339 METITO HOU	Reverse Osmosis brackish and seawater	1,000	0.5	--	--	Water treatment and pollution control equipment. Turnkey systems and/or contract operations available. Standard or special system designs available.
Permutit Company, Inc. E. 49 Midland Avenue Paramus, New Jersey USA 07652	Reverse Osmosis brackish and seawater	--	--	--	--	Assembles and installs brackish water and seawater units.
Polymetrics, Inc. 1005 Timothy Drive San Jose, California USA 95133 Telephone: 408/288-7850	Reverse Osmosis brackish and	--	--	--	--	Assembles and installs brackish water and seawater units.
Resources Conservation Co. P.O. Box 3766 Seattle, Washington USA 98124	Reverse Osmosis	4,500	1	--	4.5	Primarily industrial applications. The vapor compression unit is generally used to highly concentrate cooling tower blowdown. It is also used to treat difficult brackish water.
	Vapor Compression	1,000	225	400	5	
Riley-Beaird, Inc. Maxim Evaporator Division P.O. Box 31115 Shreveport, Louisiana USA 71130 Telephone: 318/865-6351 Telex: 50-7472	Single-Effect, submerged-tube (waste heat recovery) evaporator	15	0.19	--	--	The low-temperature submerged-tube units--can be used on waste heat. The horizontal MSF plants use a cross tube low-temperature design.
	Multi-Stage Flash	216	72	--	1	
Solar Energy Innovations Corp. 54-45 44th Street Maspeth, New York USA 11378 Telephone: 212/361-9038	Solar Still	--	--	0.1	--	Developing a solar distillation unit using a Fresnel-type concentrating lens. Water is heated in a stainless steel water chamber.
Solar Water Purifier Corp. 1525 West Vista Way P.O. Box 878 Vista, California USA 92083 Telephone: 714/724-2163	Solar Stills	3 gpd	0.25 gpd	--	--	Passive solar stills made of stainless steel and glass or galvanized steel and glass. Used mainly to produce small quantities of drinking water for home use.

Table A-1--Continued

Manufacturer	Type of Desalination Units Produced	In the Past 3 Years			Capacity[a] Installed (mgd)	Comments
		Largest kgal/day	Smallest kgal/day	Typical kgal/day		
Standard Communications Corp. HRO Systems Division P.O. Box 92151 Los Angeles, California USA Telephone: 213/532-5300 Telex: 67-7173 Cable: STACOM	Reverse Osmosis seawater	1.6	0.1	0.3	0.2	Standard compact units (100-1,600 gpd) [0.4-6 m^3/d] are built for shipboard use. The units can use a variety of energy sources including AC, DC, reciprocating engine, and hydraulic motor. The units can be supplied with frames for land-based usage.
Ultra Pure Systems, Inc. 1150-G Dell Avenue Campbell, California USA 95008 Telephone: 408/374-8370	Reverse Osmosis Ion Exchange (deionizers)	432	6	50	5	Units with a variety of pretreatment and post-treatment options to process a variety of water sources. Many industrial high-purity water applications.
Unitech Graver Water Division 2720 U.S. Highway 22 Union, New Jersey USA 07083	Distillation	--	--	--	--	--
Village Marine Corp. 8888 West Washington Blvd. Culver City, California USA 90230	Reverse Osmosis brackish and seawater	3	0.07	0.5 to 1.0	--	Standard compact units (70 to 3,000 gpd) [0.3 to 11.3 m^3/d] are built for shipboard use. The units can use 12 VDC or 110/220 VAC. Spiral-wound membranes are used.
Water Purification Systems 3451 E. 26th Street Los Angeles, California USA 53223 Telephone: 213/263-8998 Telex: 658437 XL Global SNC	Reverse Osmosis brackish and seawater	1.997	0.5	100	3	Self-contained units. Have produced a trailer-mounted self-contained unit for use in supplying isolated areas.
Water Services of America, Inc. 8165 West Tower Avenue Milwaukee, Wisconsin USA 53223 Telephone: 414/354-6470	Reverse Osmosis brackish and seawater	3,300	0.3	28	8	Systems can be manufactured with automatic controls and safety features which allow minimum maintenance and supervision.
Westinghouse Corporation Desalinations Operations P.O. Box 251 Concordville, Pennsylvania USA 19331 Telephone: 215/358-4902	Distillation	--	--	--	--	--

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Note: Metric conversion factors are: kgal/day x 3.785 = m^3/d; mgd x 3,785 = m^3/d.

[a] Capacity installed or sold in the past 3 years.

Appendix B
CHARACTERISTICS OF COMMERCIALLY AVAILABLE REVERSE OSMOSIS MEMBRANES

There are a number of manufacturers of reverse osmosis membranes in the USA. The major manufacturers are listed in Table B-1.

Table B-2 lists some of the many models of membranes available along with their characteristics. These data are based on information submitted by the manufacturers as part of a mail survey that was conducted in connection with the publication of this manual. It should be expected that the models and their characteristics may change in the future.

The inclusion or exclusion of a manufacturer and/or membranes in the listings does not constitute a recommendation nor a disapproval by U.S. A.I.D. or the authors of this manual. It is only offered as information to be evaluated by the individual reader.

Table B-1
MANUFACTURERS OF COMMERCIALLY AVAILABLE REVERSE OSMOSIS MEMBRANES IN THE U.S.A.

Culligan USA
1 Culligan Parkway
Northbrook, Illinois
USA 60062
Tel: 312/498-2000
TWX: 910-686-0005
Cable: CULLINC

Desalination Systems, Inc.
1107 West Mission Avenue
Escondido, California
USA 92025
Tel: 714/746-8141
Telex: 697890

Dow Chemical USA
Functional Products and Systems Dept.
2020 Dow Center
Midland, Michigan
USA 48640
Attention: Product Sales Manager
Reverse Osmosis Membranes
Tel: 517/636-1000
Telex: 2274-50

E.I. DuPont de Nemours & Company
Permasep Products
Wilmington, Delaware
Attention: Product Manager
USA 19898
TWX: 510-666-2217 IK

Envirogenics Systems Co.
9255 Telstar Avenue
El Monte, California
USA 91731
Tel: 213/573-9220
Telex: 698706

Fluid Systems Division of UOP, Inc.
2980 North Harbor Drive
San Diego, California
USA 92101
Tel: 714/299-9920
TWX: 910-335-1193

Table B-1--Continued

FilmTec Corporation
15305 Minnetonka Boulevard
Minnetonka, Minnesota
USA 55343

Hydranautics Water Systems
6338 Lindman Drive
P.O. Box 3690
Santa Barbara, California
USA 93105
Tel: 805/964/7747
Telex: 658445 HYDRA GLTA

Table B-2
CHARACTERISTICS OF COMMERCIALLY AVAILABLE REVERSE OSMOSIS MEMBRANES

Description				Performance		Recommended Operating Conditions					
Model	Use	Size (inches) OD x Length	Type	Permeate Production (gpd)	Salt Rejection (%)	Pressure (psi)	pH Range	Temperature Max. °F	Temperature Min. °F	Maximum Cl_2 Exposure (mg/l)	Footnotes
Desalination Systems, Inc.											
4014-98	BW	4 x 40	Spiral/CA/--	1,400	98	400-600	6-8	95	--	0.5	c
4018-97	BW	4 x 40	Spiral/CA/--	1,800	97	400-600	6-8	95	--	0.5	c
4022-97	BW	4 x 40	Spiral/CA/--	2,200	96	400-600	6-8	95	--	0.5	c
4050-98	BW	8 x 40	Spiral/CA/--	5,000	98	400-600	6-8	95	--	0.5	c
4065-97	BW	8 x 40	Spiral/CA/--	6,500	97	400-600	6-8	95	--	0.5	c
4080-96	BW	8 x 40	Spiral/CA/--	8,000	96	400-600	6-8	95	--	0.5	c
B451	BW	4 x 40	Spiral/Non CA/Tfc	2,000	98.5	400-600	1-8	122	--	0.0	c, f
B452	BW	4 x 40	Spiral/Non CA/Tfc	1,400	98.5	400-600	1-8	122	--	0.0	c, f
B881	BW	8 x 40	Spiral/Non CA/Tfc	8,000	98.5	400-600	1-8	122	--	0.0	c, f
B424LP	BW	4 x 40	Spiral/Non CA/Tfc	1,400	96	250-300	1-8	122	--	0.0	c, f
B425LP	BW	4 x 40	Spiral/Non CA/Tfc	2,000	96	250-300	1-8	122	--	0.0	c, f
B824LP	BW	8 x 40	Spiral/Non CA/Tfc	8,000	96	250-300	1-8	122	--	0.0	c, f
B441	SW	4 x 40	Spiral/Non CA/Tfc	1,600	99	800-900	1-8	122	--	0.0	a, c, f
B442	SW	4 x 40	Spiral/Non CA/Tfc	2,000	99	800-900	1-8	122	--	0.0	a, c, f
Dow Chemical USA											
DOWEX^{R} RO-20K	BW	10.75 OD x 48.75	HFF/CTA/ASY	20,000	> 90	400	4-7.5	86	32	1.0	b, e, i
DOWEX^{R} RO-4K	BW	6.25 OD x 48.0	HFF/CTA/ASY	4,000	> 90	400	4-7.5	86	32	1.0	b, e, i
DOWEX^{R} RO-.25K	BW	3.75 OD x 14.75	HFF/CTA/ASY	250	> 90	200	4-7.5	86	32	1.0	b, e, g

Table B-2--Continued

Description				Performance		Recommended Operating Conditions					
								Temperature		Maximum Cl_2	
Model	Use	Size (inches) OD x Length	Type	Permeate Production (gpd)	Salt Rejection (%)	Pressure (psi)	pH Range	Max. °F	Min. °F	Exposure (mg/l)	Footnotes
Dow Chemical USA (Continued)											
XFS-4167.15L	SW	8.6 OD x 48.75	HFF/CTA/Dense	3,000	> 98.8	800	4-7.5	86	32	1.0	b, e, h
F-4167.16L	BW	10.75 OD x 48.75	HFF/CTA/ASY	20,000	> 90	250	4-7.5	86	32	1.0	b, e, q
E. I. DuPont de Nemours & Co.--Permasep® Products											
B-9/0410-010	BW	5-¼ OD x 17-1/8	HFF/Aramid/ASY	830	> 90	400	4-11	95	32	0.0	o, b
B-9/0420-021	BW	5-¼ OD x 25	HFF/Aramid/ASY	1,740	> 90	400	4-11	95	32	0.0	o, b
B-9/0440-042	BW	5-¼ OD x 47	HFF/Aramid/ASY	3,500	> 90	400	4-11	95	32	0.0	o, b
B-9/0840-140	BW	9-½ OD x 48	HFF/Aramid/ASY	11,600	> 90	400	4-11	95	32	0.0	o, b
B-9/0840-155	BW	9-½ OD x 48	HFF/Aramid/ASY	12,800	> 90	400	4-11	95	32	0.0	o, b
B-9/0840-170	BW	9-½ OD x 48	HFF/Aramid/ASY	14,000	> 90	400	4-11	95	32	0.0	o, b
B-9/0040-250N	BW	12-¼ OD x 53	HFF/Aramid/ASY	20,700	> 90	400	4-11	95	32	0.0	o, b
B-10/X6410-003N	SW	5-½ OD x 22	HFF/Aramid/ASY	200	> 98.5	800	5-9	95	32	0.0	j, b
B-10/6440-015	SW	5-½ OD x 48-½	HFF/Aramid/ASY	1,240	> 98.5	800	5-9	95	32	0.0	j, b
B-10/6440-016	SW	5-½ OD x 48-½	HFF/Aramid/ASY	1,370	> 98.5	800	5-9	95	32	0.0	j, b
B-10/6840-045	SW	10-3/8 OD x 59	HFF/Aramid/ASY	3,700	> 98.5	800	5-9	95	32	0.0	j, b
B-10/6840-050	SW	10-3/8 OD x 59	HFF/Aramid/ASY	4,100	> 98.5	800	5-9	95	32	0.0	j, b
B-10/6840-055	SW	10-3/8 OD x 59	HFF/Aramid/ASY	4,600	> 98.5	800	5-9	95	32	0.0	j, b
Envirogenics Systems Co.											
490/40	BW	3.9 OD x 40	Spiral/CA/Blend	1,800	90	400-500	4-8	95	33	1.0	c, l
495/40	BW	3.9 OD x 40	Spiral/CA/Blend	1,500	95	400-500	4-8	95	33	1.0	c, l

Table B-2--Continued

Description				Performance		Recommended Operating Conditions					
Model	Use	Size (inches) OD x Length	Type	Permeate Production (gpd)	Salt Rejection (%)	Pressure (psi)	pH Range	Temperature Max. °F	Temperature Min. °F	Maximum Cl_2 Exposure (mg/l)	Footnotes
Envirogenics Systems Co. (Continued)											
497/40	BW	3.9 OD x 40	Spiral/CA/Blend	1,350	97	400-600	4-8	95	33	1.0	c, l
498/40	SW	3.9 OD x 40	Spiral/CA/Blend	1,800	98	600-800	4-8	95	33	1.0	c, k, a
890/40	BW	7.9 OD x 40	Spiral/CA/Blend	6,200	90	400-500	4-8	95	33	1.0	c, l
895/40	BW	7.9 OD x 40	Spiral/CA/Blend	5,500	95	400-500	4-8	95	33	1.0	c, l
897/40	BW	7.9 OD x 40	Spiral/CA/Blend	4,800	92	400-600	4-8	95	33	1.0	c, l
898/40	SW	7.9 OD x 40	Spiral/CA/Blend	5,500	98	600-800	4-8	95	33	1.0	c, k, a
Fluid Systems Division of UOP											
8600PA	BW	7.88 OD x 40	Spiral/PA/Tfc	5,000	98	400-600	4-6	113	40	0.0	c, d, n,
8150S	BW	7.88 OD x 40	Spiral/CA/ASY	4,800	95	400-600	4-6	105	40	1.0	c, d, n,
8150HR	BW	7.88 OD x 40	Spiral/CA/ASY	4,000	97	400-600	4-6	86	40	1.0	c, d, n
8800HR	BW	7.88 OD x 40	Spiral/CA/ASY	8,000	97	400-600	4-6	86	40	1.0	c, d, n
1501PA	SW	5.97 OD x 40	Spiral/PA/Tfc	1,800	98.9	800-1000	4-6	113	40	0.0	c, m, n
1001PA	SW	3.93 OD x 40	Spiral/PA/Tfc	800	98.9	800-1000	4-6	113	40	0.0	c, m, n
4600PA	BW	3.93 OD x 40	Spiral/PA/Tfc	1,250	98	400-600	4-6	113	40	0.0	c, d, n
4160S	BW	3.91 OD x 40	Spiral/CA/ASY	1,200	95	400-600	4-6	86	40	1.0	c, d, n
4160HR	BW	3.91 OD x 40	Spiral/CA/ASY	1,000	97	400-600	4-6	86	40	1.0	c, d, n
4200HR	BW	3.93 OD x 40	Spiral/CA/ASY	2,000	97	400-600	4-6	86	40	1.0	c, d, n
4101	BW	3.91 OD x 38	Spiral/CA/ASY	700	96	400-600	4-6	86	40	1.0	c, d, n
4400PA	BW	3.93 OD x 38	Spiral/PA/Tfc	1,050	98	400-600	4-6	113	40	0.0	c, d, n

Table B-2--Continued

Description				Performance		Recommended Operating Conditions					
Model	Use	Size (inches) OD x Length	Type	Permeate Production (gpd)	Salt Rejection (%)	Pressure (psi)	pH Range	Temperature Max. °F	Temperature Min. °F	Maximum Cl_2 Exposure (mg/l)	Footnotes
Fluid Systems Division of UOP (Continued)											
7005PA	SW	2.41 OD x 21	Spiral/PA/Tfc	100	98.9	800-1000	4-6	113	40	0.0	c, n, p
701PA	SW	2.41 OD x 40	Spiral/PA/Tfc	220	98.9	800-1000	4-6	113	40	0.0	c, m, n
Hydranautics Water Systems											
8040XA	BW	8 x 40	Spiral/CA/ASY	7,000	≥ 95	400	4-6	104	32	1.0	l
8040XA	BW	8 x 40	Spiral/CA/ASY	5,700	≥ 98	400	4-6	104	32	1.0	l
8540XA	BW	8.5 x 40	Spiral/CA/ASY	8,100	≥ 95	400	4-6	104	32	1.0	l
8540XB	BW	8.5 x 40	Spiral/CA/ASY	6,600	≥ 98	400	4-6	104	32	1.0	l

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Note: Metric conversion factors are: inches x 2.54 = cm, gpd x 0.003785 = m^3/d; psi ÷ 14.7 = atm; (°F-32) 0.556 = °C.
ASY = Asymmetric; BW = Brackish Water; CA = Cellulose Acetate; CE = Cellulose Ester; CTA = Cellulose triacetate;
NON CA = Non-Cellulose Acetate; PA = Polyamide; SW = Seawater; Tfc = Thin film composite; HFF = Hollow Fine Fiber;

$$\text{Salt Rejection} = \left(1 - \frac{\text{Product Concentration}}{\text{Feedwater Concentration}}\right) \times 100\%$$

a. A first stage seawater membrane--two stages are recommended.
b. Dimensions are outside diameter by length of the outer shell of the permeator.
c. Dimensions are outside diameter by length measured from end-to-end of the product water tube.
d. Performance initially--test conditions 2,000 ppm NaCl, 420 psi [28.6 atm], 10 percent recovery, 77°F [25°C].
e. The permeator bundle is replaceable in the pressure case.
f. Performance after one year--test conditions 1,000 ppm NaCl, 410 psi [27.9 atm], 10 percent recovery, 77°F [25°C].
g. Performance initially--test conditions 500 ppm NaCl, 200 psi [13.6 atm], 10 percent recovery, 77°F [25°C].
h. Performance initially--test conditions 30,000 NaCl, 800 psi [54.4 atm], 30 percent recovery, 77°F [25°C].
i. Nominal performance after 3 years--test conditions 1,500 ppm NaCl, 400 psi [27.2 atm], 75 percent recovery, 77°F [25°C].
j. Performance after one year--test conditions 30,000 ppm NaCl, 800 psi [54.4 atm], 30 percent recovery, 77°F [25°C].
k. Performance initially--test conditions 2,000 ppm NaCl, 600 psi [40.8 atm], 10 percent recovery, 77°F [25°C].
l. Performance initially--test conditions 2,000 ppm NaCl, 420 psi [28.6 atm], 10 percent recovery, 77°F [25°C].
m. Performance initially--test conditions 35,000 ppm NaCl, 800 psi [54.4 atm], 7 percent recovery, 77°F [25°C].
n. For continuous exposure a pH range of 3-7 for cellulose acetate and 2-12 for polyamide is possible.
o. Performance after one year--test conditions 1,500 ppm NaCl, 400 psi [27.2 atm], 75 percent recovery, 77°F [25°C].
p. Performance initially--test conditions 35,000 ppm NaCl, 800 psi [54.4 atm], 3 percent recovery, 77°F [25°C].
q. Performance initially--test conditions 1,500 ppm NaCl, 250 psi [17 atm], 75 percent recovery, 77°F [25°C].

Appendix C
COST ESTIMATING OF DESALINATION PROCESSES

This appendix contains some cost data and worksheets to aid in making an order-of-magnitude cost estimate of a project and/or an evaluation of processes or proposals.

Costs for desalination facilities are rather site-specific and hence local costs and information are critical. It is important to define what is or may be included in an estimate. Does it include everything necessary for a complete operating facility, or just the basic desalination equipment? The difference could easily double or triple the overall project costs.

Cost estimates tend to become more realistic (and often higher) as more is known about particular projects. The worksheets shown on Figures C-1 and C-2 are included in this appendix to assist in organizing information on a project for estimating and/or cost evaluation purposes. Figure C-1 is a Desalination Cost Summary Sheet and Figure C-2 is the Desalination Project Information Form. Both should be used when making or evaluating estimates. The following subsections contain comments to assist in filling out both of these forms.

The important aspects of a cost analysis of a desalination facility are the capital and operating costs. The worksheets will help develop both of these costs.

The Desalination Project Information Form (Figure C-2) should be begun first. This will assist in determining (or estimating) the components for both the capital and operating costs. The following section offers suggestions and guidance for each line item in the form. There are also tables and figures (Table C-1 and Figures C-3 through C-16B) which will provide additional cost information for the form. These sources provide information which correspond to the numbered lines or items in the forms or sheets as follows:

Source	Desalination Project Information Form	Desalination Cost Survey Sheet
Table C-1	29	20
Figure C-3	18	1
Figure C-4	18	1
Figure C-5	18	1, 7
Figure C-6	18	2
Figure C-7	18	3
Figure C-8	18	3
Figure C-9	18	3
Figure C-10	18	4, 8
Figure C-11	18	6
Figure C-12	18	7
Figure C-13	13	20, 21, 23
Figure C-14	24	15
Figure C-15	35	24
Figure C-16A, B	39 - 43	26, 27

After the Desalination Project Information Form is completed as much as possible, the Desalination Cost Summary Sheet should be filled in.

The three numbers of primary interest are: (1) the overall capital cost; (2) the capital cost per installed gpd or m^3/d, and (3) the overall unit cost of production in \$/kgal or \$/m^3.

Obviously when a process is complicated as a cost estimate for a desalination facility is reduced to filling out a five-page form and using a series of graphs for determining the numbers, the cost estimate is just that--an estimate. It will assist at the planning stages to provide an order-of-magnitude estimate of the costs involved, but the estimate should be used with caution and judgment. The only thing certain about the cost determined is that it is bound to change as the project progresses.

DESALINATION PROJECT INFORMATION FORM

The numbers to the left of the following comments correspond to the numbered lines on the Desalination Project Information Form, Figure C-2.

1. Whether the capacity does or does not include blending with untreated feedwater or water from other sources should be clear.

2. The type of plant means distillation, electrodialysis, etc. A realistic plant life over 20 years is unusual.

3. This could be wells, a shore intake, a lake intake, etc.

4. A reliable analysis of the feedwater is extremely important for design. There can be considerable variation in temperature and constituents of a source over time (which can dramatically affect the selection and operation of a process).

5. The desalination facility should be described in greater detail.

6. The recovery factor is the amount of product water divided by the amount of feedwater used per unit of time expressed as a percent.

7. The performance factor in a distillation plant is the pounds of water produced by 1,000 Btu's [or kg/MJ]. It is not customary to use it for plants which do not employ steam for power.

8-9 This will aid in assessing the extent and results of the commercial experience of this type and capacity unit.

10. Will this plant be operated by a local agency, a contractor, or under some other arrangement? Table C-1 presents a guide to staffing.

11-12 The technical ability and experience of the operators can affect the selection and/or performance of the facility.

13. If the operating agency does not envision a realistic budget for the long-term operation of the project this could be an indication of future problems or the need to select a plant that will operate under budgetary contraints.

14. A designer or manufacturer will set a design plant factor (actual annual production divided by rated production), which is often based on past experience, materials, and/or design. However, factors such as lack of demand, experimental aspects, complexity, local conditions, operating staff, future budgets, etc., may change the actual plant factor. If this plant factor changes, it will alter the calculated unit cost of the water produced, potentially changing the economics of the project. A realistic prediction of a plant factor is one of the most important items in estimating costs and assessing the viability of desalination projects. See discussion in Chapter 8.

15. The disposal of brine can be critical at inland locations. Brine disposal in the wrong area could adversely affect existing surface- or ground-water resources.

16. Regulatory or financial aspects of a project may affect its feasibility, timing, and hence its actual cost.

17. Realistic scheduling of the project steps will assist in arriving at times for bidding and initial operation. These time periods may have a bearing on financing.

An estimate of design and construction time for various desalting processes has been given by Catalytic (1979):

Process	Capacity (mgd)	Capacity [m^3/d]	Design Time (months)	Construction Time (months)
Distillation	0.5	1,890	6	8
	2.5	9,460	8	20
	25	94,600	8	36
RO/ED	0.25	946	6	6
	0.5	1,890	6	6
	5	18,900	9	12

It is assumed that this is based on construction in the USA. To this must be added time for plans, specifications, bidding, bid award, startup, etc., plus additional time for shipping and delays associated with work in any particular country.

18-19 This table should be used to list and briefly describe the major components of a desalination project. If better information is not available, an order-of-magnitude cost estimate for these facilities can be obtained from cost curves in this appendix. Many of these graphs have time-related cost indexes which can assist in their updating. The assumptions and references for these cost curves are listed in Table C-2. The following is a list of cost curves in the appendix which relate to the various major components shown in the table.

Feedwater Supply Development

Figure C-3--Construction costs of feedwater intakes and brine outfalls.

Figure C-4--Construction costs for wells.

Figure C-5--Construction costs for long runs of pipelines.

Feedwater Treatment

Figure C-6--Construction costs for special pretreatment equipment. Standard pretreatment is included in equipment costs.

Desalting Equipment

Figure C-7--Construction costs for desalination plants.

Figure C-8--Capital cost of vapor compression and heat recovery evaporators.

Figure C-9--Capital cost of small reverse osmosis and electrodialysis units.

Site Development

Figure C-10--Construction costs for general site development.

Special Energy Sources

Figure C-16B--Cost of electricity based on fuel cost.

Electrical Equipment and Switchgear

Figure C-11--Construction costs for electrical utilities and switchgear.

Brine Disposal

Figure C-12--Construction costs for evaporation ponds.

Product Storage and Treatment

Figure C-10--Construction Costs for Storage Tanks.

Indexes which relate prices in the USA (those in the figures in this appendix) to those in the locality under study must be ascertained.

20-21 The construction period noted in Item 17 will affect this.

These costs will vary depending on the extent to which outside consultants are used and the scheduling necessities of the job. These owner's expenses can be estimated by using the following guide provided by Catalytic (1979):

Direct Costs (in 18 above)	Percentage for Owner's Expenses (Items 20 + 21)
Up to $100,000	15
$100,000 - $1,000,000	12
$1,000,000 - $10,000,000	9

22. This will vary with the complexity, schedule, locality, and unknowns of a job. It is supposed to cover "surprises," so it should be adequate. Five to ten percent of direct capital costs could be used if no better estimate exists.

23. Startup costs are usually based on O&M costs for the number of months required for plant startup. This would generally be 1 to 2 months.

24. An estimate of land area is shown on Figure C-14. A lesser area could probably be used depending on requirements for storage, buildings, parking, etc. An idea of the floor area for the basic desalination equipment itself is given in the "Engineering Considerations" section of each process chapter. Land is a non-depreciating capital cost, since it is recoverable upon completion of the project.

25. Working capital is the monetary value of stored material, spare parts, fuels, chemicals, and expense money needed for operating during a specific period. This can vary depending on many factors, including the predicted problems in procurement and transportation of items and the plant factor needed. Where a high plant factor is necessary in an isolated location, the working capital could be considerably higher than the same unit with either a low required plant factor and/or readily available chemicals, spares, etc. An estimate of 2 to 3 months of O&M costs could be used if other data are not available.

26. The rate of interest and plant life (Item 2) will affect the annual fixed charge.

The annual fixed charge is computed and used in the economic evaluation of a project to account for the use of capital funds on the facility. This is often called a capital recovery or fixed annual charge (for the use of money). It can be computed on the depreciable capital costs (direct and indirect) or both the depreciable and non-depreciable costs (such as land and working capital).

The capital recovery factor is based on the interest rate selected and the planned life of the plant. These factors are available in tables or can be computed by the formula

$$\text{Capital Recovery Factor} = \frac{i}{1 - (1 + i)^{-n}}$$

where

i = interest expressed as a decimal.
n = number of years (plant life).

For convenience an abbreviated table follows:

Interest (%)	n = 10	n = 15	n = 10
6	0.1359	0.1030	0.0872
8	0.1490	0.1168	0.1019
10	0.1628	0.1315	0.1175
12	0.1770	0.1614	0.1339
15	0.1993	0.1845	0.1598
20	0.2385	0.2253	0.2054

27-29 The labor involved in running a desalination facility in developing areas can easily be underestimated. Although the operation of a desalination plant does not seem to take a great deal of physical work, the plant does need to be checked and tended to one degree or another during its operating periods. In addition, continual maintenance work should be taking place. Chemicals, saline water, heat, etc., create a harsh environment for materials and machinery, requiring continual care, repairs, cleaning, painting, etc., which, if postponed, will eventually cause major problems.

Where personnel can be shared between facilities such as water- and power-generating stations, costs can be reduced. However, it should be remembered that there is more to keeping a facility going than simply operators. This includes administrators, secretaries, clerks, guards, janitors, drivers, chemists, mechanics, etc. If these must be hired, they will add to the operating costs.

A good, competent operating staff is an excellent investment in the long-term operation of a facility. If the future operating agency is unable or unwilling to supply this, the potentials for future problems, low plant factors, and excessive unit water production costs are high.

Table C-1 is a rough guideline for estimating staffing for desalination facilities. This is for a typical staff employed at a plant located in the USA and must be adjusted to local conditions to reflect the local culture, technology, conditions, and cost structure.

Figure C-14 can be used for a rapid estimate of labor-related O&M costs, including general and administrative overhead (G&A). This is for USA costs and should be adjusted for local staffing requirements and conditions. The costs in the graph also include the cost of supplies and maintenance materials.

30. G&A overhead covers expenses which are not chargeable directly to a particular operation or maintenance function (Parsons, 1967). It is usually based on a percentage of direct labor cost. This is included in the costs listed on Figure C-14.

31-32 The local cost and availability of chemicals can vary considerably. Some of the chemicals require special transportation, handling, and storage techniques, which can radically alter their costs and delivery schedule if they must be transported to certain areas. The type and amount of chemicals can be estimated by using Appendix D.

The following is a list of chemical costs in the USA used for the ORNL report (Larson, 1979) on desalination costs.

Chemical	Unit Cost [$/kg]	Unit Cost ($/lb)
Antifoam	2.13	0.97
Sulfuric acid (100%)	0.07	0.03
Polyphosphate	3.41	1.55
Sodium hexametaphosphate	0.64	0.29
Potassium permanganate	1.28	0.58
Caustic (NaOH)	0.40	0.18
Sodium sulfite	0.11	0.05
Chlorine (gas)	0.24	0.11

33. This includes cartridge filters for ED and RO plants. If the cost for labor and supplies is not used from Figure C-13 an estimate of 1 percent of the desalting plant capital cost could be used.

34-35 This will help in clarifying the cost data on membrane processes. The assumed membrane life for the process can affect the ultimate operating life. An estimate of membrane costs is given on Figure C-15.

36. If there are major items within the process such as tubes, vapor compressors, pumps, motors, boilers, etc., which are not anticipated to last the design life of the plant, they should be noted and the cost for replacement determined. An initial estimate of this could be 1 to 2 percent of desalting plant capital cost.

37-43 The energy costs should be estimated based on local conditions. The energy costs used in estimating proposals should be examined very closely to determine how they were determined and if they are valid for the particular location.

The cost of steam should be ascertained. It is dependent on whether the facility is a dual-purpose plant or not and the method of accounting utilized.

Figures C-16A and C-16B can be used as a guide to estimating electrical and steam costs if other information is not available. The sizes of the steam and electrical generation facilities have an impact on unit costs--this is not fully reflected in curves shown on the guide. As facilities become smaller, the unit costs rise very rapidly. The curves which show the relationship between the fuel cost and electrical or steam costs need to be updated periodically. These curves were taken from publications (Larson, 1979, and Catalytic, 1979) which are being used for desalination estimating but with the realization that costing energy is considerably more complex than they indicate.

DESALINATION COST SUMMARY SHEET

The data for the costs listed in Items 1-16 and 20-27 should be obtained from the Desalination Project Information Form and the accompanying explanation pages contained in this appendix.

Recurring costs of taxes, insurance, etc., should be determined for Items 17-19 if they are significant and relevant.

The method of calculating the capital, annual, and unit production costs are shown on the sheet.

Some capital items such as startup costs and working capital cannot be computed until the operation and maintenance costs are completed.

In order to complete the cost summary sheet it will be necessary to make assumptions and estimates--especially in the early stages of a project. As information on a particular project develops the estimates can be refined. However, it should be kept in mind that the plant factor in itself can have sufficient subjective variation (due to local conditions, etc.) that the actual overall unit cost, i.e., \$/kgal or \$/m^3, may be subject to considerable variation.

These potential variations in actual cost should be remembered when comparing the unit costs between various processes in an effort to make a final selection. Depending on the validity of the data which were used, minor variations between processes are often meaningless.

COST ESTIMATE--EXAMPLE

An example of a cost estimate using the two forms (Figures C-1 and C-2) plus the pricing and process data available is worked out in this appendix. A hypothetical, although realistic, situation is given below.

The Ministry of Water and Power of the water-short Caribbean island of San Manne (an imaginary country) has requested assistance in determining the planning level cost of a reverse osmosis unit to desalt brackish water for use as potable water on the island. This requested evaluation is shown in the five sheets which make up Figure C-17. The narrative that helps to clarify these data follows.

By a previous study the Ministry of Water and Power has determined that the island requires about 1.2 mgd [4,450 m^3/d] of potable water from the plant throughout the year.

San Manne has had some limited experience with reverse osmosis. It has only a small unit and it had problems with maintenance and repairs. San Manne's population of 75,000 has only limited commercial facilities and hence specialized mechanical parts and tools are not readily available. Parts and materials are delivered by plane or on the weekly cargo vessel.

To provide the water required (1.2 mgd [4,540 m^3/d]) it has been decided to install a desalination facility with a total nameplate (or design) capacity of 1.6 mgd [6,060 m^3/d] and count on a plant factor of only 0.75 versus the 0.9 or 0.95 that is possible in more industrialized countries.

A recovery factor was restricted to 50 percent to lessen the potential for scaling and simplify the plant, although several membrane manufacturers calculated that a recovery of about 65 to 70 percent could have been obtained by the use of reject staging.

The 1.6 mgd [6,060 m^3/d] plant would be constructed and installed in four independent modules of 0.4 mgd [1,510 m^3/d] each. These could be operated separately both to meet the water demand and to allow for repairs and/or utilization of parts from other units while awaiting replacements if a spare part is not available.

The ground water will be supplied by six 200-foot-deep [61-m] wells, each rated at 0.8 mgd [3,030 m^3/d] and hence each well will be able to supply one of the modules. This will allow four working wells and two spares. A 1,000-foot [305-m] pipeline is necessary to bring the feedwater (1.6 ÷ 50 percent recovery = 3.2 mgd [12,100 m^3/d]) to the plant site.

The facility will use electric power from the island's grid, which is available at the site.

Brine will be disposed of through a 2,000-foot [610-m] pipeline which will discharge it into the sea in a good mixing area on the south shore of the island.

Water storage tanks equivalent in capacity to 10 day's output will be constructed at the site. This will allow for shutdowns for repairs and maintenance as well as for prolonged operation at less than expected capacity while awaiting critical parts, etc. This will involve two storage tanks, each with a capacity of 6 MG [22,710 m^3/d].

In proceeding through the Desalination Project Information Form, certain assumptions and estimates had to be made based on the limited data available. A general cost index of 1.14 was assumed to update the 1978-1979 cost to 1980, and a local index of 1.2 was used to account for project cost differences due to shipping, productivity, labor costs, etc., between San Manne and the USA, on which the costs are based.

The interest during construction, A/E and project management fees, startup costs, working capital, etc., were based on suggestions in the manual. (As the project proceeds, these could be modified as required.)

The land area requirements were about 2 acres [8,100 m^2] for the plant and about 8 acres [32,400 m^2] for the brackish water well field.

The cost of money (capital recovery) and the general and administrative overhead (G&A) were set by consultation with the Ministry. The staffing reflects the recommendations in Table C-2.

The amount of chemicals (chlorine, acid, and sodium hexametaphosphate) needed is based on the data in Appendix D under brackish water reverse osmosis. Chemical costs, especially for acid and chlorine, will be high compared to those in the USA if extensive shipping (as hazardous cargo) is involved. The markup in cost from the table of chemical cost was arbitrarily (but with judgment) determined.

Mechanical and electrical problems (such as power surges at the central power plant in the capital) cause occasional burnouts of motor windings so that an allowance for special repairs was included.

The electrical rate of $0.10/kWh is high but reflects the power costs in many of the Caribbean islands.

In computing the annual costs no fixed cost for non-depreciable capital costs was utilized, by agreement with the Ministry.

The result of the computations was an estimate of production costs (including capital recovery) of about \$3.80/kgal [\$1.00/m^3] including the large storage facility, well field, and brine disposal.

The next steps, not included here, would be to perform the same analysis using electrodialyis as a method of desalting.

Table C-1
GENERAL GUIDE TO STAFFING AT DESALINATION FACILITIES

Staff	Distillation				Electrodialysis and Reverse Osmosis				Well Field	
	Plant Capacity				Plant Capacity				Capacity	
mgd	To 0.05	0.05-0.5	0.5-2.5[a]	2.5-10[a]	To 0.05	0.05-0.5	0.5-2.5	2.5-10	To 3	3-12
1,000 m^3/d	To 0.19	0.19-1.9	1.9-9.5	9.5-38	To 0.19	0.19-1.9	1.9-9.5	9.5-38	To 11	11-45
Supervisory										
Plant superintendent/administrator	1	1	1	1	1	1	1	1	--	1
Process or plant engineer	--	--	1	2	--	--	--	1	--	--
Maintenance and Day Staff										
Mechanic--general	1	1	2	3	1	1	1	1	1	2
Mechanic--electrical	--	--	1	2	--	--	1	2	1	1
Mechanic--instruments	--	--	1	2	--	--	--	1	--	--
Chemist	--	--	1	2	--	--	1	2	--	--
Laborer	--	--	2	4	--	--	2	4	--	--
Secretary/Clerk	--	--	1	1	--	--	1	1	--	--
Shift Staff										
Supervisor	--	2	4[b]	4[b]	--	--	1	4[b]	1	1
Operator--3 shifts/day	6-8	6-8	6-8	8-12	6-8	3-8	6-8	8-12	-	-
--2 shifts/day	3-6	3-6	5-6	-	3-6	3-6	5-6	-	-	3
--1 shift/day	1-2	2-3	-	-	1-2	2-3	-	-	1	2
TOTAL (24-hour operation)	1-10	3-16	6-22	12-31	1-8	3-16	5-21	10-29	1-4	2-9

Notes:

1. Caution--The staffing of any facility is extremely site-specific. In some cases, the staff can be reduced considerably from these figures and/or used on more than one facility such as at a dual-purpose (electricity-water) plant or other facility located in the vicinity. In other cases, specialists could be hired temporarily for specific repair work to eliminate some of these positions. The most important item in staffing is employee competence.
2. The table notes the number of full-time staff in each category. This is the equivalent of a full-time, well-trained person in the USA. Local conditions, culture, training, labor laws, etc., could affect this considerably in other countries.
3. This is for plant operation 24 hours per day. For safety reasons, it is best to have a minimum of two people on any shift as shown in this guide. Some facilities can run unattended for long periods at a time.
4. This labor is exclusive of power source staffing and requirements.

[a] These are recommended by Watson (1976) for a single purpose plant except that he specifies four operators (18 total staff) for the 0.5- to 2.5-mgd [0.19- to 1.9-m^3/d] plant and eight operators (29 total staff) for the 2.5- to 10-mgd [9.5- to 38-m^3/d] plant. The distillation portion of a dual-purpose (electricity-water) plant could be staffed with much fewer people, as work could be shared with the electric plant.

[b] One supervisor per shift in the large plants.

Table C-2
DATA ON COST CURVES USED IN APPENDIX C

Figure No.	Description	Includes	Excludes	Index	Reference
C-3	Construction costs for feedwater intake and brine outfall	Installed equipment costs	Indirect capital costs Non-depreciating capital costs	USBR Pumping Plant Cost Index--2.46 for January 1979 Useful life 20 yrs	Larson, 1979 OWRT, 1976
C-4A	Cost of constructing a well	Conventional water-supply well granular material Well depth of 1,000 ft [305 m] Well diameter 2 or 4 inches [5 to 10 cm] larger than pump Full-depth carbon steel casing, 0.375-inch [9.5 mm] wall Cage-type, wire-wound, stainless steel screen, 100 ft [30.5 m] Grouting, 2 inches [5 cm] thick, 100 ft [30.5 m] Designs and specifications Construction engineering Construction supervision	Interest during construction Owner's general expense Land costs Pumps, valves, and other surface facilities Collection and transport pipelines	ENR Building Cost Index--1739 for January 1979 Useful life 40 yrs	Catalytic, 1979 OWRT, 1976
C4-C	Cost of well pumping facilities	Installed cost of conventional deep-well turbine pumps including electric motor, column and shaft, discharge head, valve and connection	Interest during construction Owner's general expense Land costs Transport pipelines Electrical powerline and substation	USBR Pumps and Prime Movers Cost Index--2.71 for January 1979 Useful life 40 yrs	OWRT, 1976 Catalytic, 1979
C-5	Construction costs for pipelines	Pipeline Designs and specifications Construction engineering	Interest during construction Owner's general expense Removal or replacement of roadway Pumping facilities Land costs	USBR Concrete Pipeline Cost Index--2.26 for January 1979 Useful life 100 yrs	Catalytic, 1979 OWRT, 1976
C-6	Construction costs for pretreatment equipment	Designs and specifications Construction engineering Construction supervision	Indirect capital costs Non-depreciating capital costs	ENR Building Cost Index--1739 for January 1979	Morin and Latour, 1980 Dupont, 1976 Catalytic, 1979 Burns & Roe, 1979

Table C-2--Continued

Figure No.	Description	Includes	Excludes	Index	Reference
C-7	Construction costs for desalination plants	Installed equipment costs, electric utilities Distillation VTE MSF Performance ratio 12 12 Acid feed X X Concentration ratio 2 2 Single purpose X X Reverse Osmosis and Electrodialysis Membranes Minimum pretreatment Seawater RO 30% recovery Single pass Standard seawater Brackish RO & ED 5,000 mg/l TDS Brackish ED 78% recovery 70°F [21°C] feedwater	Site development Intake and outfall systems Indirect capital costs Non-depreciating capital costs	ENR Building Cost Index--1739 for January 1979 Useful life 20 yrs	Larson, 1979 Catalytic, 1979
C-8	Capital cost of vapor compression and heat recovery evaporators	Assembled skid unit at factory in the USA	Transportation, engineering, site preparation, source of feedwater, power, etc.	1st quarter 1980 prices for 2 USA manufacturers	--
C-9	Capital costs of small reverse osmosis and electrodialysis skid units	Assembled skin unit at factory in the USA	Transportation, engineering, site preparation, source of feedwater, power, etc.	1st quarter 1980 prices for 3 USA manufacfurers	--
C-10	Construction costs for general site development	Site development and common facilities, grading, roads, fencing, gates, basic service buildings	Indirect capital costs Non-depreciating capital costs	ENR Building Cost Index--1739 for January 1979 Useful life 20 years	Larson, 1979 Catalytic, 1979

Table C-2--Continued

Figure No.	Description	Includes	Excludes	Index	Reference
C-10	Construction costs for water storage tanks	Cone roof, carbon steel tank with foundation and painter	Indirect capital costs Non-depreciating capital costs	ENR Building Cost Index--1739 for January 1979 Useful life 20 yrs	Catalytic, 1979 Richardson, 1978
C-11	Construction costs for electrical utilities and switchgear	Installed costs--site development	Indirect capital costs Non-depreciating capital costs	1st quarter 1979	Larson, 1979
C-12	Construction costs for evaporation ponds	Pond embankments Buried membrane lining Interpond conduits Gravel protective cover on slopes Construction engineering Construction supervision	Interest during construction Owner's general expense Land costs Brine transport pipeline	USBR Canal Earth-Work Cost Index--2.32 for January 1979 Useful life 30 yrs	OWRT, 1976 Catalytic, 1979
C-13	Annual costs for O&M labor	O&M labor, G&A, miscellaneous materials and supplies	Chemicals and filter replacement Electric power Membrane replacement O&M chemicals	BLS Labor Costs Index for SIC--494-7 6.83 for December 1978	Larson, 1979 Catalytic, 1979
	MSF, well field, and intake	O&M labor, G&A, miscellaneous materials and supplies	Chemicals and filter replacement Electric power Membrane replacement O&M chemicals	BLS Labor Cost Index for SIC--494-7 6.83 for December 1978	Larson, 1979 Catalytic, 1979
	RO & ED information	O&M labor, G&A, miscellaneous materials and supplies	Chemicals and filter replacement Electric power Membrane replacement O&M chemicals	BLS Labor Cost Index for SIC--494-7 1st Quarter 1979	Larson, 1979 Catalytic, 1979 Larson, 1979
C-14	Land requirements for desalination facilities	Land suggested--less may be needed	--	--	OWRT, 1976 Catalytic, 1979

Table C-2--Continued

Figure No.	Description	Includes	Excludes	Index	Reference
C-15	Membrane replacement costs	RO membranes based on a 3-yr life and approximately 4,000 gpd [15 m3/d] per membrane element ED membranes based on a 5-yr life at 70° F [21°C]	--	--	Catalytic, 1979 Permutit, 1978 Burns & Roe, 1979 Morin & Latour, 1980 DuPont, 1976 Shannon, 1979
C-16A and C-16B	Estimating the cost of electricity and steam based on fuel cost	--	--	--	Cox, 1979 Larson, 1979 Catalytic, 1979 Watson, 1976 DOE/EIA, 1976 NATCO, 1980

Notes: ENR = *Engineering News Record*; USBR = U.S. Bureau of Reclamation (now called the Water and Power Resources Service);
BLS = Bureau of Labor Statistics.

DESALINATION COST SUMMARY SHEET

Location ______________________________ Capacity ____________ mgd [m³/d]
Type of Plant ______________________ Plant Life ________ Years Plant Factor ____________
Source of Feedwater ______________________________ TDS Level ____________
Recovery or Performance Factor ______________________________
Energy Source(s): Fuel (________________ type)/Electricity/Steam (Circle Sources)

CAPITAL COSTS

DIRECT CAPITAL COSTS

1. Feedwater Supply Development $________
2. Feedwater Treatment $________
3. Desalination Equipment $________
4. Site Development $________
5. Energy Source Development $________
6. Electrical Equipment (switchgear) $________
7. Brine Disposal $________
8. Product Water Storage and Treatment $________
9. Other $________

SUBTOTAL DIRECT CAPITAL COSTS (1-9) $ Ⓐ

INDIRECT CAPITAL COSTS

10. Interest During Construction $________
11. A/E, Project Management Fees $________
12. Contingencies $________
13. Startup Costs $________
14. Other $________

SUBTOTAL INDIRECT CAPITAL COSTS (10-14) $ Ⓑ

TOTAL DEPRECIABLE CAPITAL COSTS (A + B) → $ Ⓒ

OTHER CAPITAL COSTS (NON-DEPRECIABLE)

15. Land ____________ acres [m²] $________
16. Working Capital $________

TOTAL OF OTHER CAPITAL COSTS (15-16) → $ Ⓓ

TOTAL OF ALL CAPITAL COSTS (C + D) → $

UNIT CAPITAL COST ($/GPD [m³/d] INSTALLED CAPACITY) → $ /gpd [m³/d]

ANNUAL COSTS

Recurring Costs

17. Taxes $________
18. Insurance $________
19. Other $________

TOTAL ANNUAL RECURRING COSTS (17-19) → $ Ⓔ

OPERATION & MAINTENANCE (O&M) COSTS

20. Labor—Salaries $________
21. Labor—General & Administrative Overhead (_____%) $________
22. Chemicals $________
23. Supplies and Maintenance Materials $________
24. Membrane Replacement (_____ yr Life) $________
25. Special Repairs or Overhauls $________
26. Energy—Fuel/Steam (Circle one) cost/unit ______ $________
27. Energy—Electricity __________ $/kWhr $________
28. Other $________

TOTAL ANNUAL O&M COSTS (20-28) → $ Ⓕ

ANNUAL FIXED CHARGE

____% Interest & ____ yr Plant Life Capital Recovery Factor (CRF) ________

($______________ Ⓒ × ________) + (______________ Ⓓ × ________)* $ Ⓖ
(Depreciable Capital) (CRF) (Nondepreciable) (CRF)

TOTAL ANNUAL COSTS (E + F + G) → $ Ⓗ

*In many analysis the capital recovery is not computed for nondepreciable capital (D).

UNIT PRODUCTION COSTS

Annual Production ______________ kgal [m³] ☐ Design ☐ Actual

Operating Units

UNIT COST = Total Annual Cost Ⓗ / Actual Annual Production = ______________________ = $ /kgal [m³]

For Estimating Costs

UNIT COST = Total Annual Cost Ⓗ / Annual Design Capacity x Plant Factor = ______________________ = $ /kgal [m³]

Cost Estimator ______________________________ Date ____________ Project ____________

Attach the Desalination Project Information Sheets

FIGURE C-1. Desalination cost summary sheet.

DESALINATION PROJECT INFORMATION FORM Page 1

1. Location ____ Design Capacity ____ mgd [m^3/d]

2. Type of Plant ____ Est. Plant Life ____ yrs

3. Source of Feedwater ____

4. Feedwater Characteristics—Is a Complete Analysis Available? ☐ Yes ☐ No
   TDS ____ mg/l max ____ mg/l min
   Temperature ____ max ____ min
   Source of Feedwater Data ____
   How many of the feedwater analyses have been performed? ____ When? ____
   Where? ____
   How reliable is this feedwater data? ____
   What feed and/or cooling water temperature was used for plant design? ____

5. Description of Plant ____

6. Recovery Factor ____ (Membrane Plants)

7. Performance Factor ____ (Distillation Plants)

8. How many plants of this design and equivalent capacity have been built previously? ____
   Where? ____

9. Are any plants of this design and capacity operating under similar conditions? ☐ Yes ☐ No
   Where? ____ Any in the Same Country or Region? ____
   With what results? ____

10. Who will operate this plant? ____

11. What is the expected level of technical expertise of those who will operate the plant? ____

12. What experience have they in operating desalination or similar equipment? ____

13. Will there be an adequate budget available for operation and maintenance of the plant throughout the project? ____
    ____ Is a training program planned? ____

14. What is the design plant factor? ____
    Based on local conditions of funding, technology, climate, support facilities, etc., is this a realistic plant factor? ☐ Yes ☐ No Explain ____
    If not, what would be? ____

15. Will there be a problem disposing of the concentrated brine discharge from the plant? ☐ Yes ☐ No
    How will it be disposed? ____

16. What permits, licenses, ordinances, financing, etc., need to be obtained or passed before the desalination facility can be procured and operated? ____

FIGURE C-2. Desalination Project Information Form.

PROJECT DESCRIPTION ______________________ Page 2

17. What is the estimated construction period? ______________________ months
When will design begin? ______________ Bids be requested? ______________
Contract be awarded? ______________ Construction Begin? ______________
______________ Plant begin operation? ______________

CAPITAL COSTS

DIRECT CAPITAL COSTS

18.

Type	Description	Source of Cost Estimate	Cost	Cost Index		Est. Indexed Cost
				Time	Locality	
Feedwater Supply Development						
Special Feedwater Treatment						
Desalting Equipment						
Site Development						
Energy Source						
Electrical Equipment (Switchgear)						
Brine Disposal						
Product Storage and Treatment						
Other						

19. The cost index(es) used: ______________________
A local cost index factor for the area under study must be determined and must include transportation factors. This local index is ______________________

INDIRECT CAPITAL COSTS

20. Interest during construction: ________%. Basis for calculation is: ______________
______________________ Interest is: $______________

21. A/E and project management fees: ________% and/or basis for calculation is: ______________
______________________ Fee is: $______________

22. Contingencies: ________% and/or basis for calculation is: ______________
______________________ Contingency is: $______________

23. Startup costs: ______________ Basis for calculation is: ______________________

FIGURE C-2. Desalination Project Information Form (continued).

PROJECT DESCRIPTION ______________________________

OTHER CAPITAL COSTS

24. Land area needed: ____________________ Unit cost of land: __________ Total cost: ___________
Basis for unit cost of land: ____

25. Working capital: _____________ Basis for cost is: ____

ANNUAL FIXED CHARGE

26. Has this been included? ☐ Yes ☐ No Rate of interest: ___________% Time Period ____________yrs
Capital recovery factor: ________________________ If no charge is to be made, why? ___________
____

O&M COSTS

LABOR

27. How many hours per day will the unit run? ____________________________

28. How many hours per day will the unit be attended? ____________________________

29. The anticipated staff is:

Job Description	No. of Positions	Annual Salary	Annual Cost
Administrator/Superintendent			
Process or Plant Engineer			
Mechanic—General			
Mechanic—Electrical			
Mechanic—Instruments			
Chemists			
Laborers			
Driver/Guard			
Secretary/Clerk			
Shift Supervisor			
Shift Operator			
TOTAL ANNUAL COST			

If there is no idea of the staff required nor the salary levels, then use a general estimated of overall labor costs. This estimate is $___________ and ☐ Does ☐ Does Not include G&A.

30. What level of general and administrative (G&A) overhead is to be used? ____________________%
The basis for this percentage is: ____
The annual G&A cost is: $______________

FIGURE C-2. Desalination Project Information Form (continued).

CHEMICALS, SUPPLIES, AND MAINTENANCE MATERIALS

31.

Chemicals and Filters Description	Annual Quantity Needed	Local Unit Cost	Total Cost

32. Are there anticipated difficulties in purchasing, delivering, storing, and/or handling these materials during the life of the project? ☐ Yes ☐ No Details: ______________________

33. Cost of supplies and maintenance materials $__________ Basis ______________________

MEMBRANE REPLACEMENT

34. What type or configuration of membranes ______________________

35. Will spares be kept on hand? ☐ Yes ☐ No If yes, how many? __________
What membrane life is being used? __________ years. For RO membranes, what is the guaranteed productivity at the end of that period versus at startup? __________
Annual replacement cost $__________ Basis ______________________

SPECIAL REPAIRS

36. Will any major repairs or equipment replacement going to be required during the life of the project?
☐ Yes ☐ No Describe: ______________________
Estimated annual cost $__________ Basis ______________________

ENERGY

37. Source(s) ______________________

38. Dual- or single-purpose plant ______________________

39. Cost of Electricity __________/kWh; Electric usage __________kWh/kgal [kWh/m^3]

40. Does the electricity consumption include all pumping into and out of the plant? ☐ Yes ☐ No
If not, how much additional pumping energy will be required? __________kWh/kgal [kWh/m^3]

41. If fuel is required what type? __________ Heating value __________

42. Fuel usage (for diesel- or gas-operated plants) __________ gal fuel/kgal [l/m^3] water

43. Steam (for distillation plants)
Plant performance factor __________ Operating temperature ______°F [°C]
Steam requirement __________ MBtu/kgal [GJ/m^3] or ______/______
Single-purpose steam cost ______/MBtu [GJ] for prime steam @ ______ psig [atm] ______°F [°C]
Dual-purpose steam cost ______/MBtu [GJ] at operating conditions ______ psig [atm] ______°F [°C]

FIGURE C-2. Desalination Project Information Form (continued).

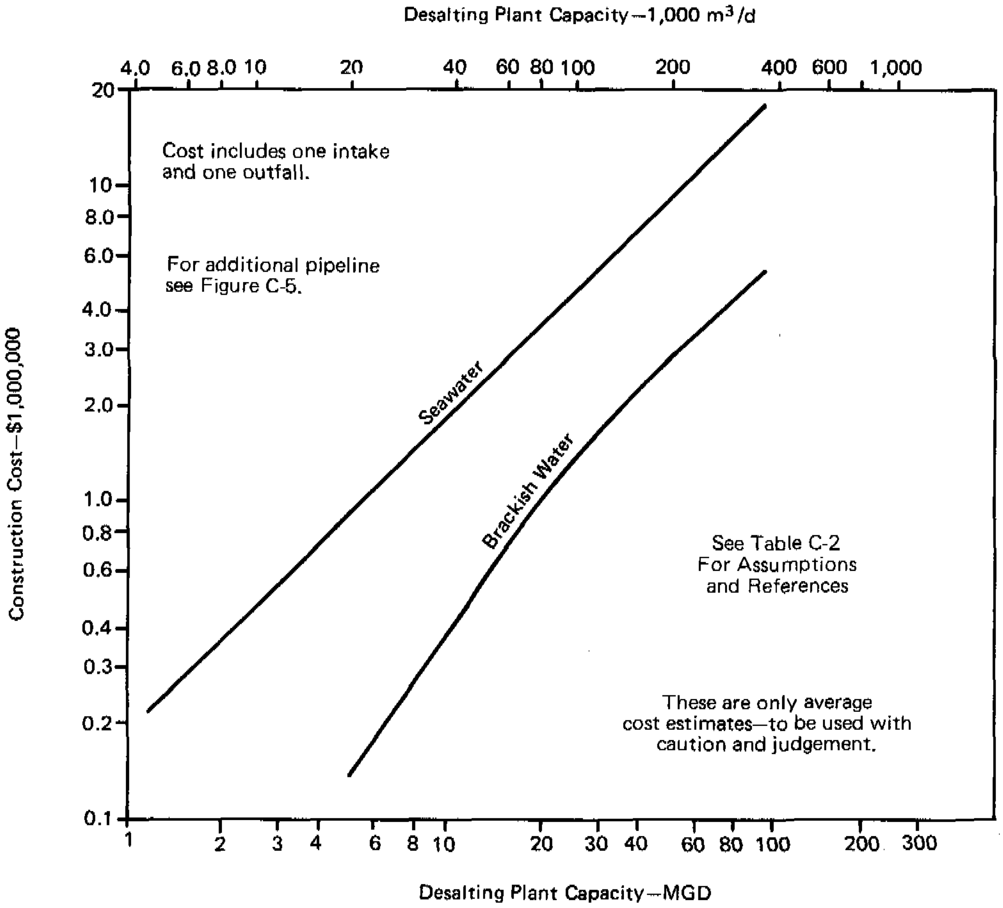
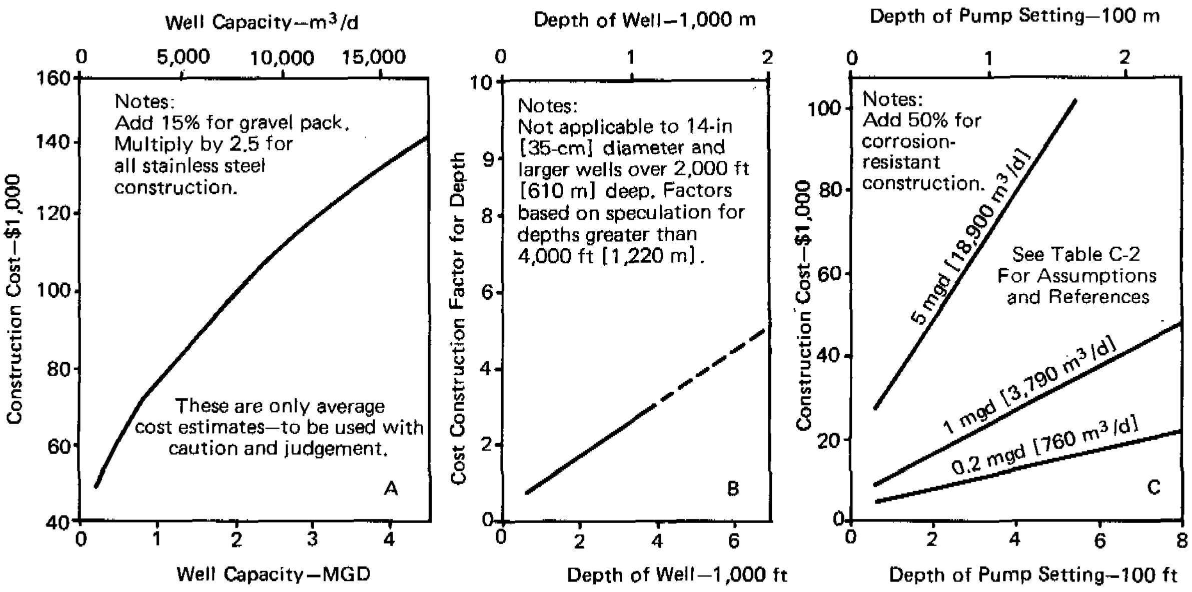


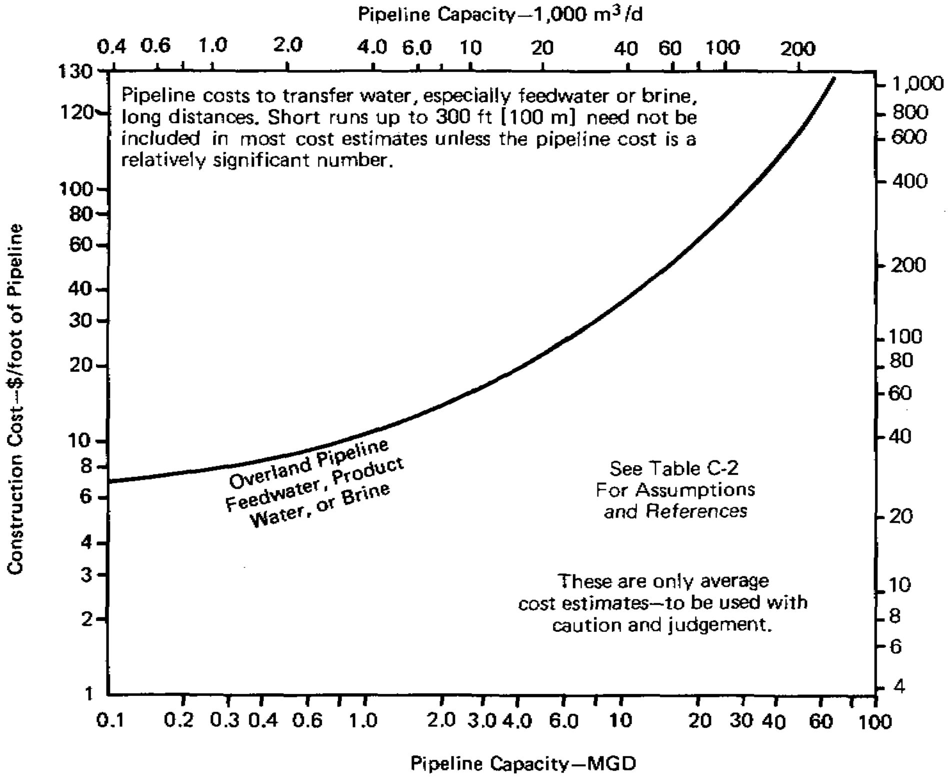
FIGURE C-3. Construction costs for feedwater intake and brine outfall.



Cost of Constructing a Well
(Use Curve A then Correct with B)

Cost of Well Pumping Facilities

FIGURE C-4. Construction costs for wells.

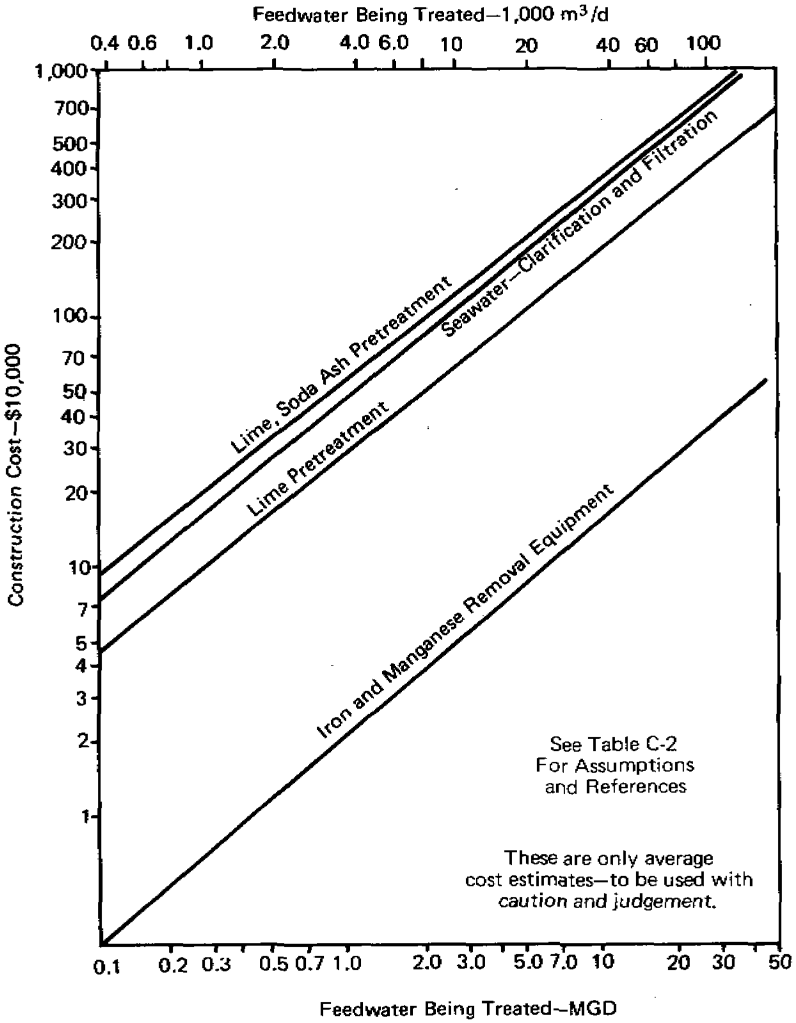


Pipeline costs to transfer water, especially feedwater or brine, long distances. Short runs up to 300 ft [100 m] need not be included in most cost estimates unless the pipeline cost is a relatively significant number.

See Table C-2 For Assumptions and References

These are only average cost estimates—to be used with caution and judgement.

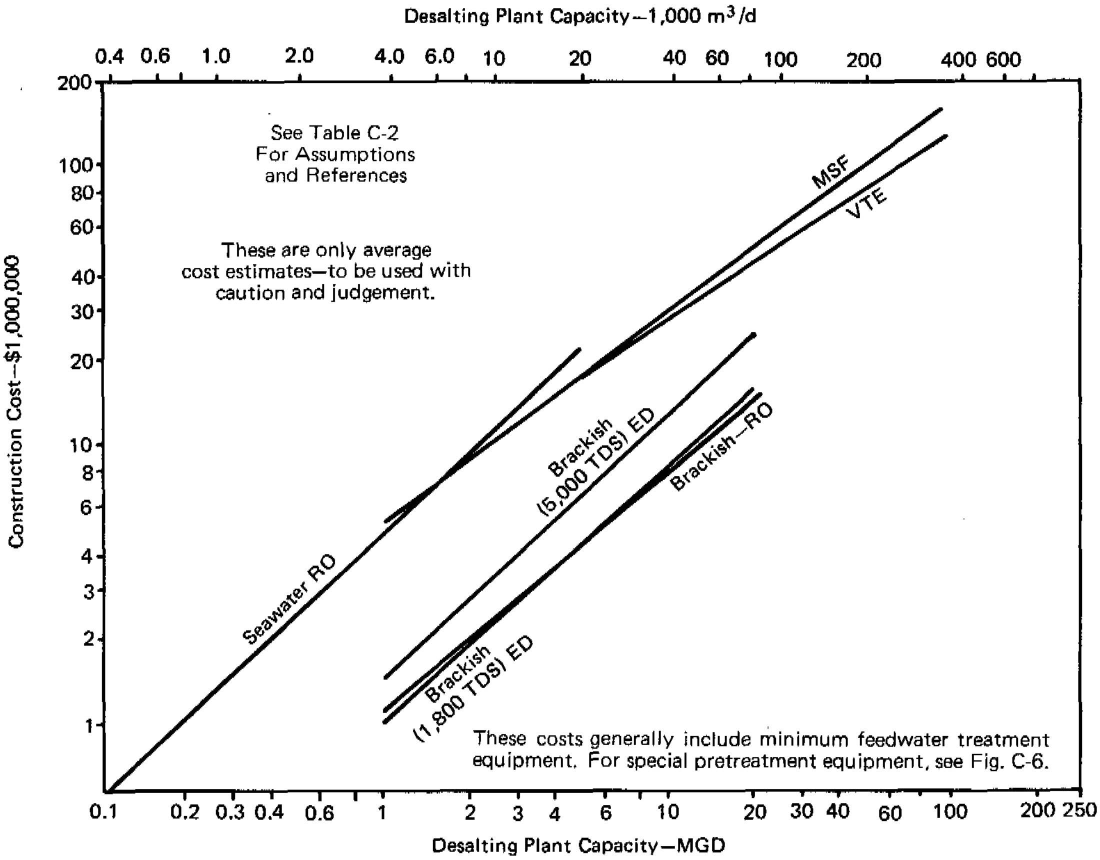
FIGURE C-5. Construction costs for pipelines.



See Table C-2 For Assumptions and References

These are only average cost estimates—to be used with caution and judgement.

FIGURE C-6. Construction costs for special pretreatment equipment.

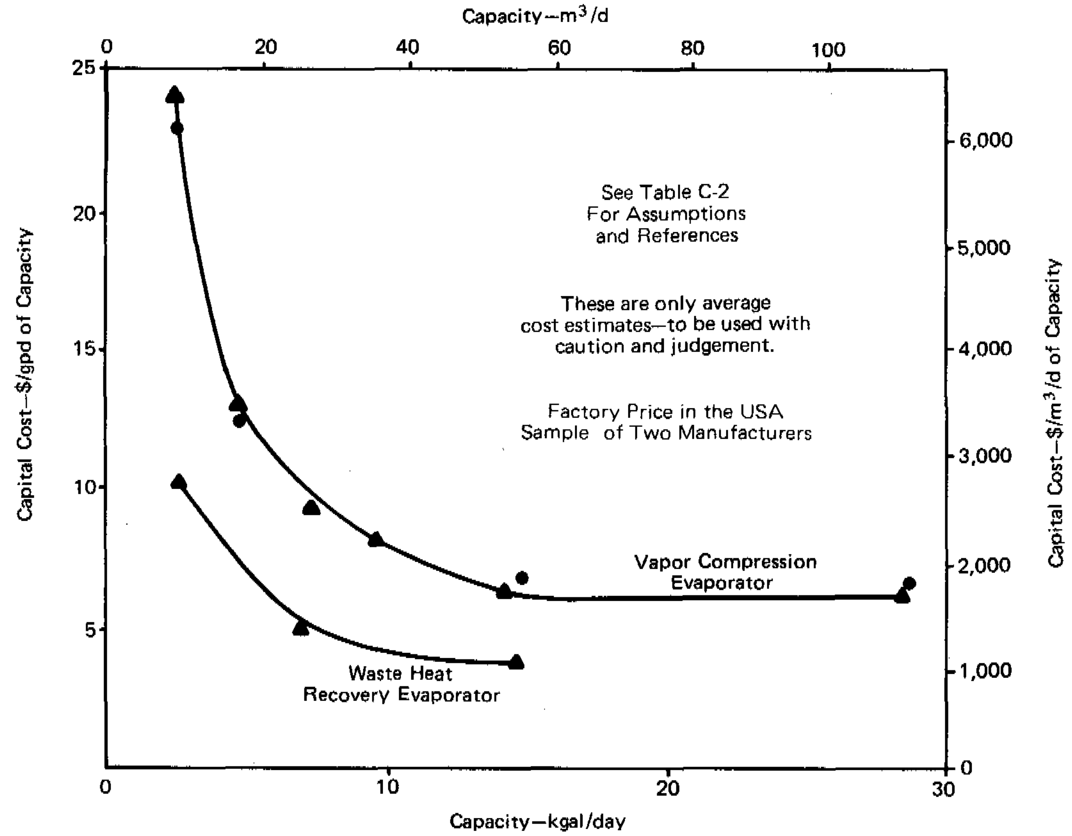


See Table C-2
For Assumptions
and References

These are only average cost estimates—to be used with caution and judgement.

These costs generally include minimum feedwater treatment equipment. For special pretreatment equipment, see Fig. C-6.

FIGURE C-7. Construction costs for desalination plants.

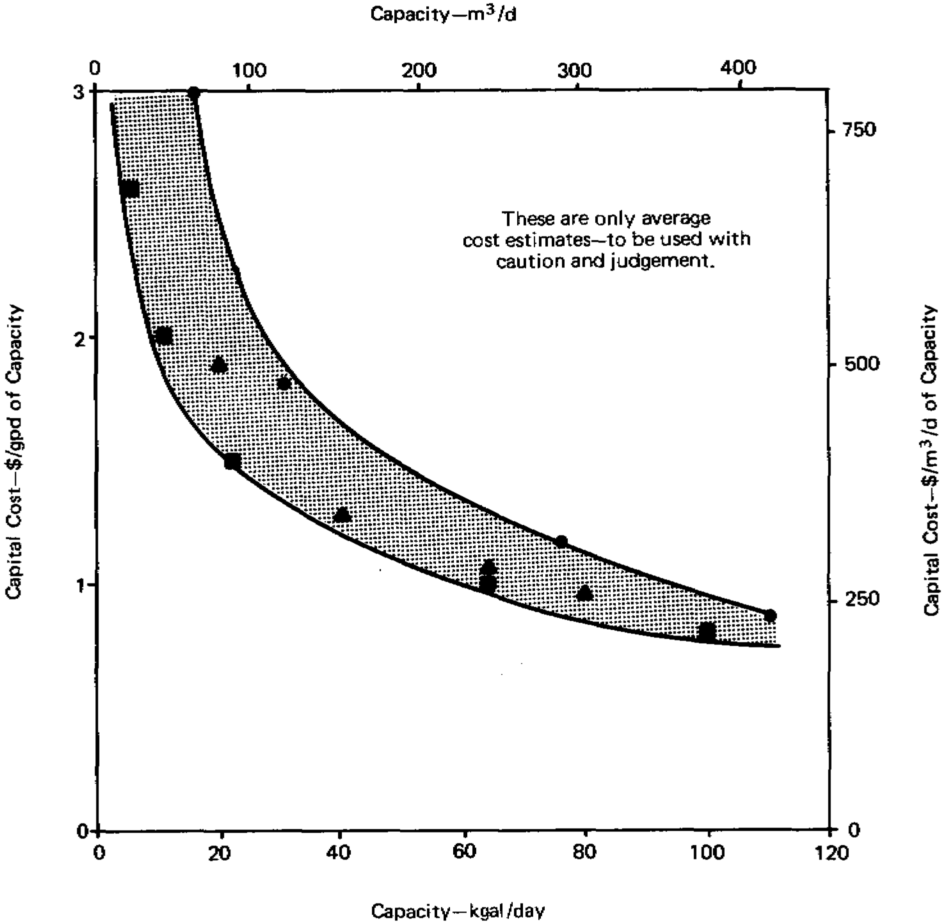


See Table C-2
For Assumptions
and References

These are only average cost estimates—to be used with caution and judgement.

Factory Price in the USA
Sample of Two Manufacturers

FIGURE C-8. Capital cost of vapor compression and heat-recovery evaporators.



1. Skid Units, FOB Factory in USA; prices do not include site work, feed water, power connection, engineering, start-up, etc.
2. These prices are for basic skid units which are self-contained. Depending on the unit, the extent of field-assembly work varies. Some added accessories may be needed and/or desired at added cost.
3. Prices will vary under bid conditions.
4. Data based on 1980 2nd Quarter price information from three manufacturers (2-RO and 1-ED).

FIGURE C-9. Capital cost of small reverse osmosis and electrodialysis skid units.

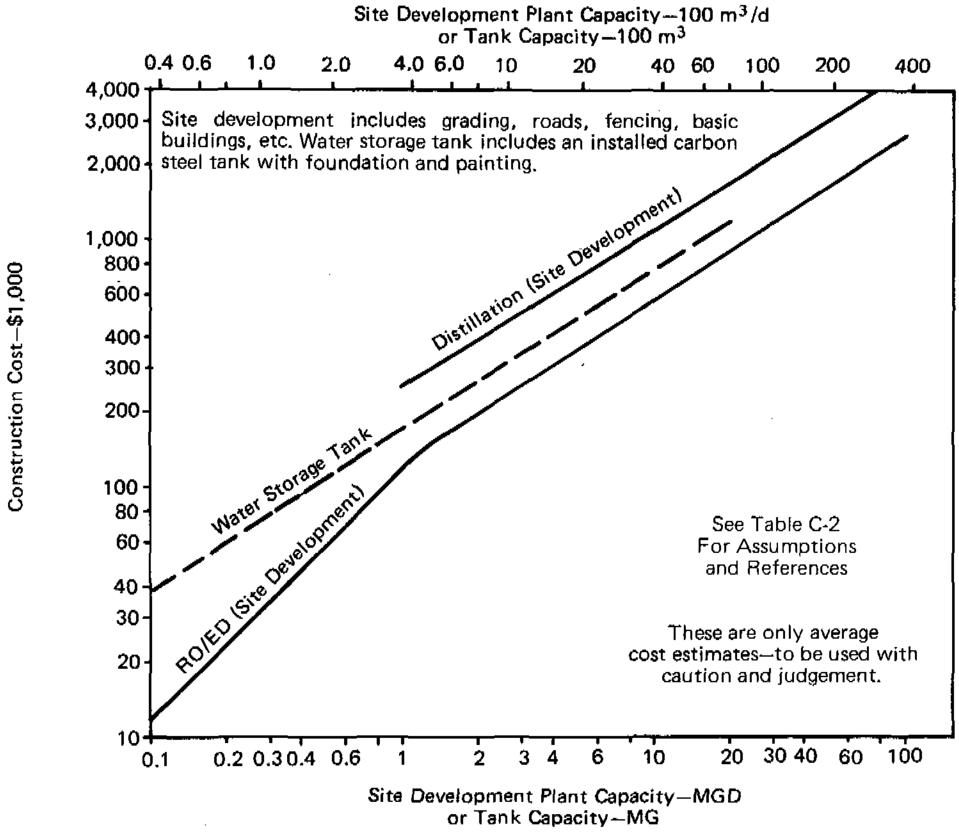


FIGURE C-10. Construction costs for general site development and water storage tanks.

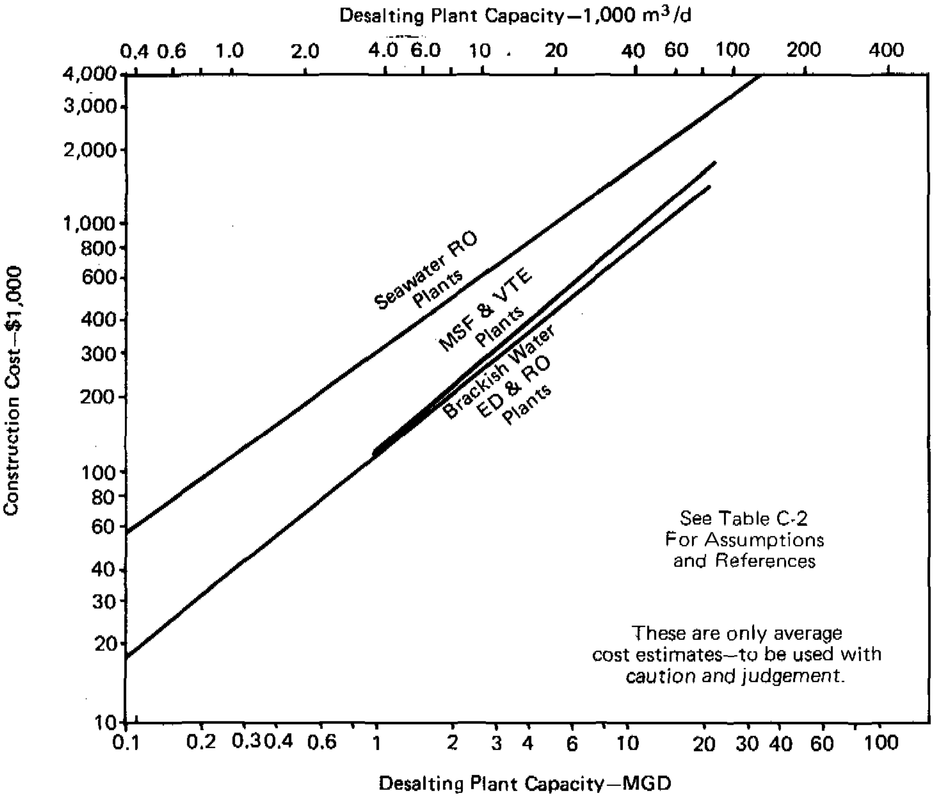


FIGURE C-11. Construction costs for electrical utilities and switch gear.

Example—If it is known that the brine discharge is 5 mgd [18,900 m^3/d] *and the average net evaporation* (E) is 4 ft [122 cm], average pond area would be about 1,400 acres [5,660,000 m^2] and the construction cost would be about $16.7 million. See ············ line.

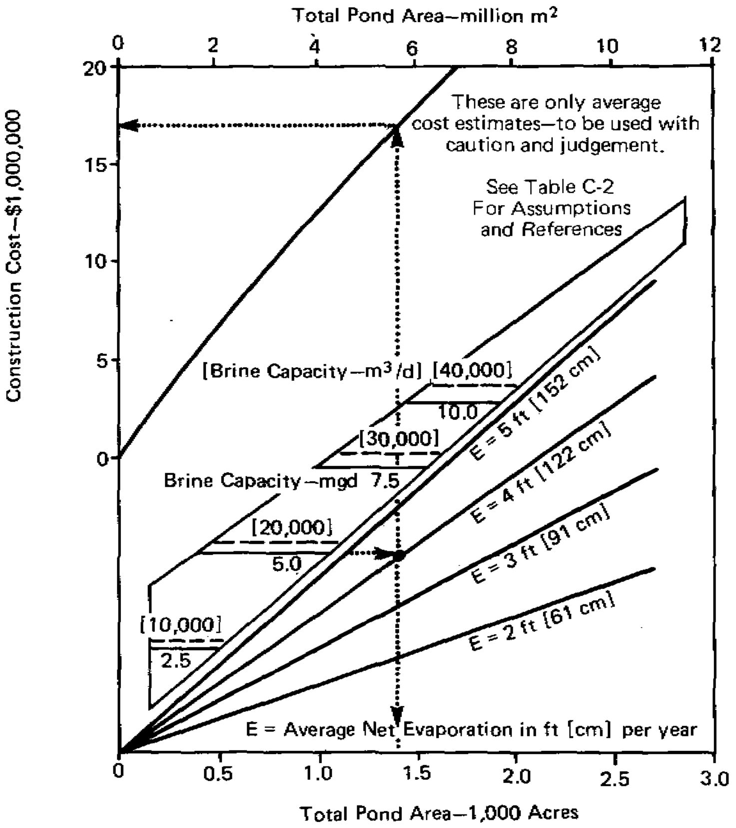


FIGURE C-12. Construction costs for evaporation ponds.

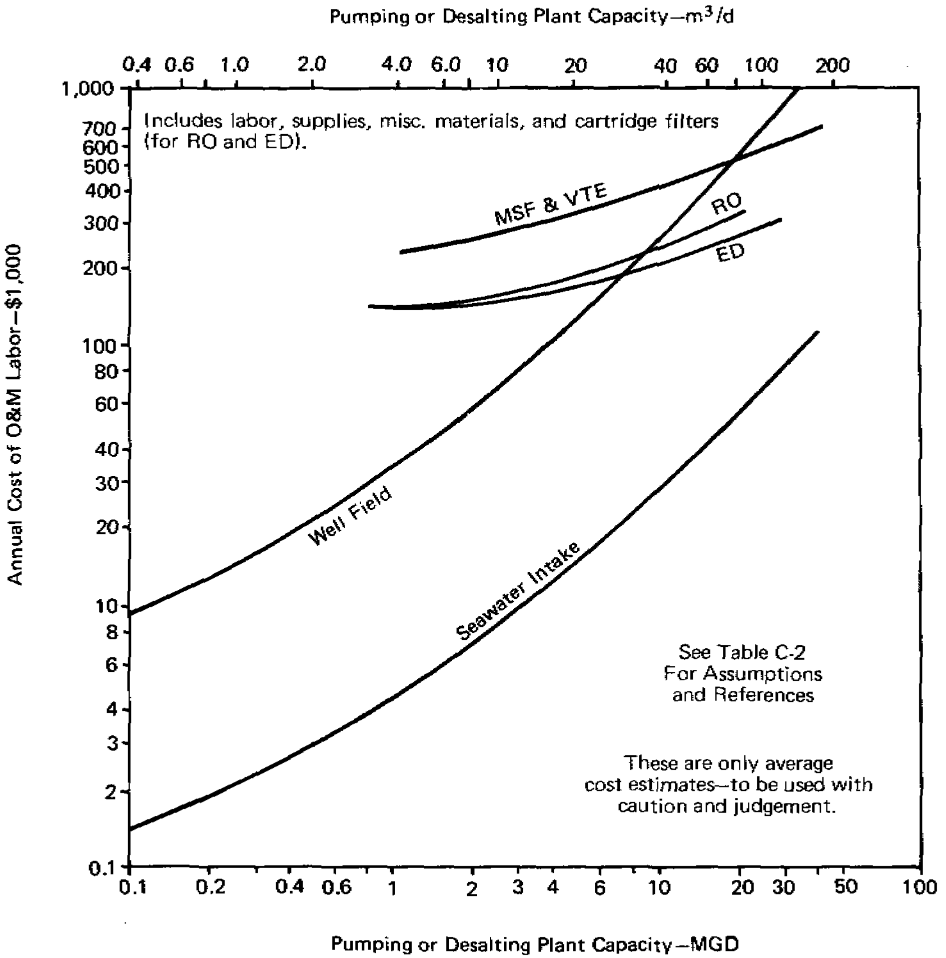
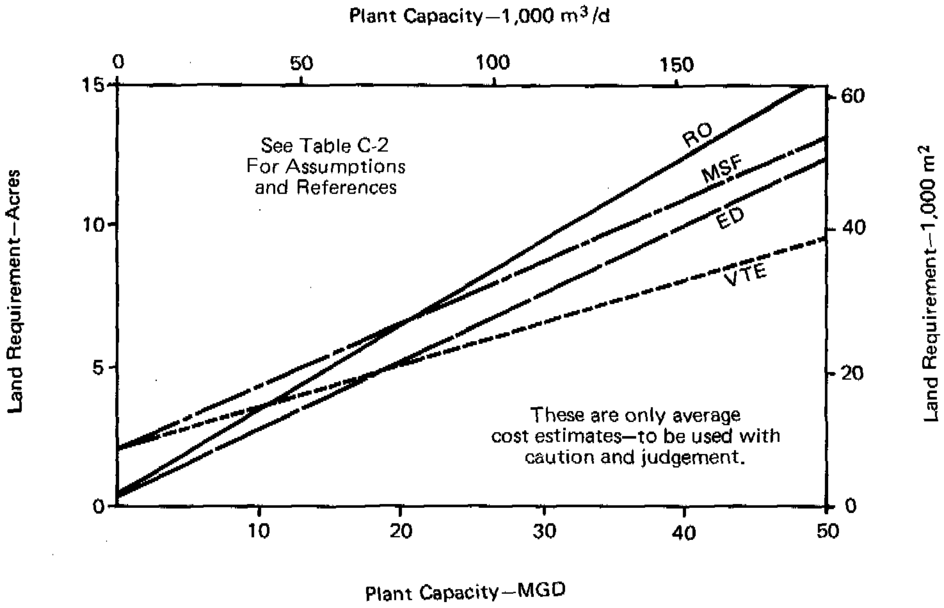


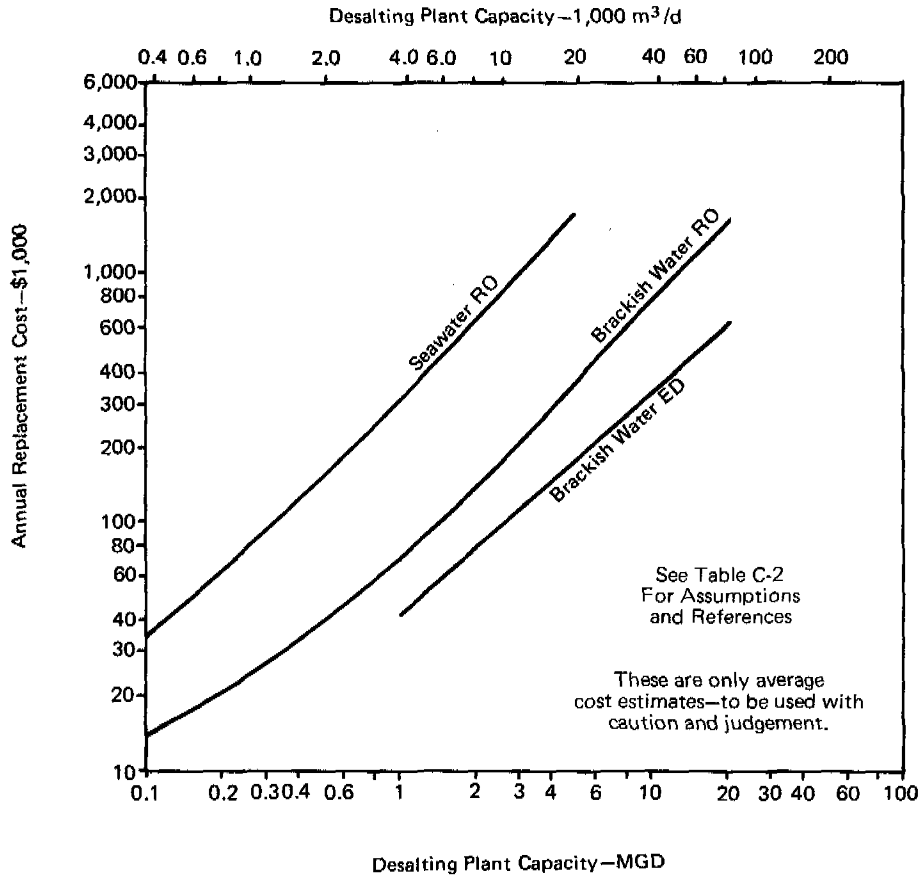
FIGURE C-13. *Annual costs for O&M labor, supplies, and miscellaneous materials.*



See Table C-2 For Assumptions and References

These are only average cost estimates—to be used with caution and judgement.

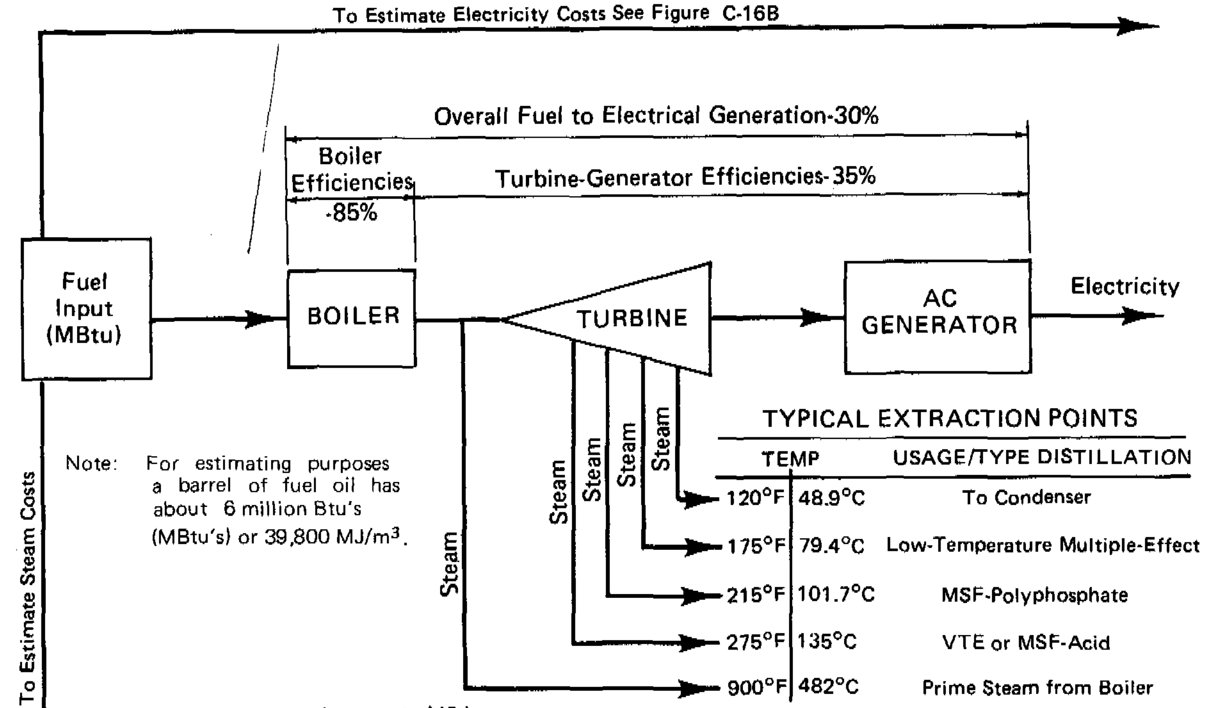
FIGURE C-14. Land requirements for desalination facilities.



See Table C-2 For Assumptions and References

These are only average cost estimates—to be used with caution and judgement.

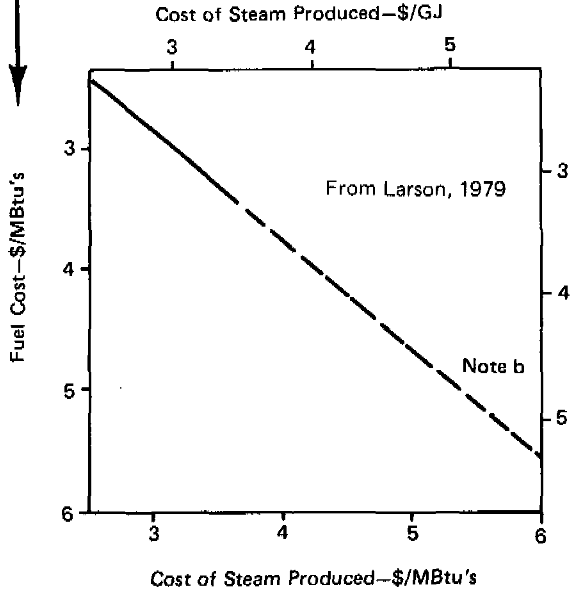
FIGURE C-15. Membrane replacement cost.



Note: For estimating purposes a barrel of fuel oil has about 6 million Btu's (MBtu's) or 39,800 MJ/m³.

TYPICAL EXTRACTION POINTS

TEMP		USAGE/TYPE DISTILLATION
120°F	48.9°C	To Condenser
175°F	79.4°C	Low-Temperature Multiple-Effect
215°F	101.7°C	MSF-Polyphosphate
275°F	135°C	VTE or MSF-Acid
900°F	482°C	Prime Steam from Boiler



For SINGLE PURPOSE plants use this steam cost. Single purpose plants produce steam for desalination purposes only.

For DUAL PURPOSE plants this steam cost can be used to compute the cost assigned to steam extracted at lower temperatures. See Column 4 in the adjacent table for the factor to use in determining the reduced cost of low-temperature steam.

These are only average cost estimates—to be used with caution and judgement.

DUAL-PURPOSE STEAM COSTS

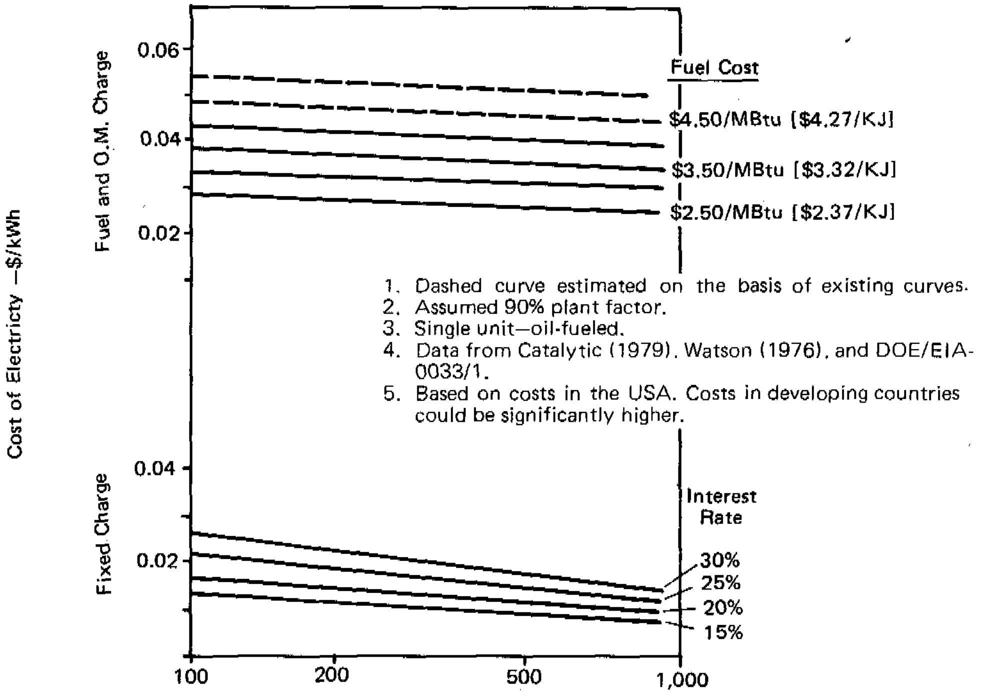
A typical calculation of extraction steam cost based on the usable work (enthalpy) available. Note a.

1		2	3		4
Extraction Steam Temperature (°F)	(°C)	Steam Enthalpy Btu/lb	Usable Work Above Condensing Temperature Amount Btu/lb	Percent of Total	Extracted Steam Value Percent of Steam Cost
900	482	1,448	516	100.0	100.0
275	135	1,133	201	39.0	39.0
215	101.7	1,057	125	24.2	24.2
175	79.4	1,020	88	17.1	17.1
120	48.9	932	0	0	0

Notes:
a. Other methods can and are being used to assign costs of extracted or backpressure steam in a dual-purpose plant.
b. Dashed curve estimated on basis of existing curves.
c. This method from Cox, 1979.

FIGURE C-16A. Estimating the cost of electricity and steam based on fuel cost.

BUS BAR ELECTRIC PRODUCTION COST ESTIMATES FOR A LARGER POWER PLANT



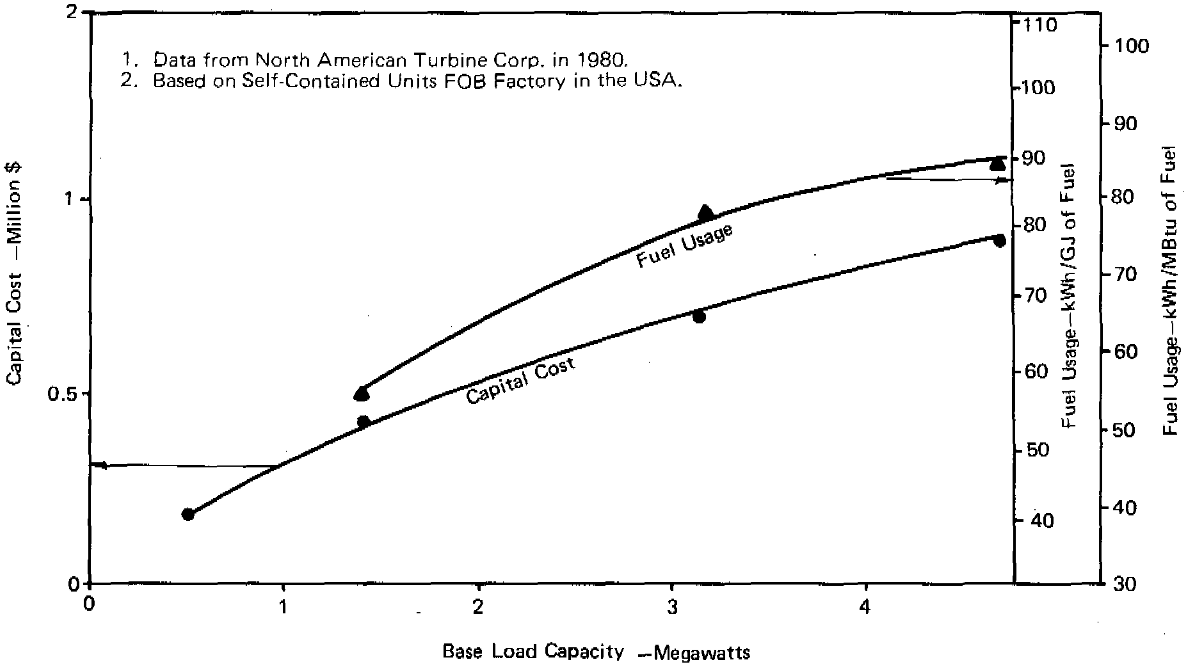
1. Dashed curve estimated on the basis of existing curves.
2. Assumed 90% plant factor.
3. Single unit—oil-fueled.
4. Data from Catalytic (1979), Watson (1976), and DOE/EIA-0033/1.
5. Based on costs in the USA. Costs in developing countries could be significantly higher.

POWER PLANT CAPACITY —Megawatts

COST/kWh = Fuel and O&M cost/kWh + Fixed Charge/kWh

These are only average cost estimates—to be used with caution and judgement.

ELEMENTS OF ELECTRIC PRODUCTION COSTS ESTIMATES FOR SMALL GAS TURBINE GENERATORS



1. Data from North American Turbine Corp. in 1980.
2. Based on Self-Contained Units FOB Factory in the USA.

Base Load Capacity —Megawatts

FIGURE C-16B. Estimating the cost of electricity and steam based on fuel cost.

DESALINATION PROJECT INFORMATION FORM

Page 1

1. Location San Manne Design Capacity 1.6 mgd ~~[m³/d]~~

2. Type of Plant Reverse Osmosis Est. Plant Life 15 yrs

3. Source of Feedwater Brackish Ground Water at a depth of ~ 200 Ft.

4. Feedwater Characteristics—Is a Complete Analysis Available? ☒ Yes ☐ No
   TDS 5,500 mg/l max 4,800 mg/l min
   Temperature 81°F max 80°F min
   Source of Feedwater Data Existing well serving 0.01 mgd RO Plant
   How many of the feedwater analyses have been performed? 15 When? 1979-80
   Where? Government Laboratory on San Manne
   How reliable is this feedwater data? Good
   What feed and/or cooling water temperature was used for plant design? 80°F

5. Description of Plant A 1.6 mgd Single-Stage RO Plant to be built with 4 - 0.4 mgd modules which can be operated independently.

6. Recovery Factor 50 % (Membrane Plants)

7. Performance Factor (Distillation Plants)

8. How many plants of this design and equivalent capacity have been built previously? Many
   Where? USA and other locations

9. Are any plants of this design and capacity operating under similar conditions? ☒ Yes ☐ No
   Where? Florida and the Caribbean Any in the Same Country or Region? yes
   With what results? There are 2 - 0.01 mgd RO Plants on San Manne which have operated but with some maintenance difficulties

10. Who will operate this plant? Local personnel from the Ministry of Water & Power

11. What is the expected level of technical expertise of those who will operate the plant? Fair - not much experience with instrumentation

12. What experience have they in operating desalination or similar equipment? 3 years operation (with some problems) of the 2 existing 0.01 mgd RO Plants

13. Will there be an adequate budget available for operation and maintenance of the plant throughout the project? The operating budget will be low and dependent on revenue from the system Is a training program planned? yes

14. What is the design plant factor? 0.75 has been selected
    Based on local conditions of funding, technology, climate, support facilities, etc., is this a realistic plant factor? ☒ Yes ☐ No Explain Plants of this design have had plant factors of 0.9 and 0.95 in the USA. This lower plant factor allows for longer downtime due to problems with the local power and delays in obtaining parts and materials

15. Will there be a problem disposing of the concentrated brine discharge from the plant? ☐ Yes ☒ No
    How will it be disposed? It will be discharged to the sea

16. What permits, licenses, ordinances, financing, etc., need to be obtained or passed before the desalination facility can be procured and operated? The ministry of Water & Power will obtain all the permits, etc.

FIGURE C-17. San Manne example.

PROJECT DESCRIPTION San Manne RO Plant Page 2

17. What is the estimated construction period? 12 months
When will design begin? June 1980 Bids be requested? March 1981
Contract be awarded? May 1981 Construction Begin? Dec. 1981
Plant begin operation? December 1982

CAPITAL COSTS

DIRECT CAPITAL COSTS

18.

Type	Description	Source of Cost Estimate	Cost	Cost Index		Est. Indexed Cost
				Time	Locality	
Feedwater Supply Development	6 - 0.8 mgd wells 6 - 0.8 mgd pumps 1000 ft. pipeline	C-4A C-4C C-5	6x$70,000 6x$10,000 1000 x $17	1.14	1.2	$574,560 82,080 23,256
Special Feedwater Treatment	NONE	REQUIRED				
Desalting Equipment	1.6 mgd RO Plant	C-7	$1,400,000	1.14	1.2	$1,915,200
Site Development	1.6 mgd RO Plant	C-10	$170,000	1.14	1.2	$232,560
Energy Source	NONE	REQUIRED				
Electrical Equipment (Switchgear)	1.6 mgd RO Plant	C-11	$165,000	1.14	1.2	$225,720
Brine Disposal	2000 Ft. Pipeline	C-5	2000 x $12	1.14	1.2	$32,832
Product Storage and Treatment	2 - 6 MG Tanks	C-10	2x$520,000	1.14	1.2	$1,422,720
Other						

19. The cost index(es) used: A General 14% adjustment for inflation
A local cost index factor for the area under study must be determined and must include transportation factors. This local index is 1.2 x USA Price - Based on other projects

INDIRECT CAPITAL COSTS

20. Interest during construction: — %. Basis for calculation is: using the method in the manual (page C-5) for combined Interest is: $ 405,900
items 20 & 21 9% of $4,510,000 = $405,900

21. A/E and project management fees: ____% and/or basis for calculation is: ____
see 20 above Fee is: $____

22. Contingencies: 5 % and/or basis for calculation is: manual suggestion (on page C-5) 5% x $4510,000 Contingency is: $ 225,700

23. Startup costs: $ 160,300 Basis for calculation is: 2 months O&M costs
($961,950 x 2/12)

FIGURE C-17. San Manne example (continued).

PROJECT DESCRIPTION San Manne RO Plant Page 3

OTHER CAPITAL COSTS

24. Land area needed: 10 acres incl. wells Unit cost of land: $15,000/ac Total cost: $150,000
Basis for unit cost of land: Information from the Ministry of Water & Power

25. Working capital: $240,500 Basis for cost is: 3 months O&M cost ($961,950 x 3/12)

ANNUAL FIXED CHARGE

26. Has this been included? ☒ Yes ☐ No Rate of interest: 10 % Time Period 15 yrs
Capital recovery factor: 0.1315 If no charge is to be made, why? —

O&M COSTS

LABOR

27. How many hours per day will the unit run? 24

28. How many hours per day will the unit be attended? 16 (2 shifts)

29. The anticipated staff is: Based on Table C-1

Job Description	No. of Positions	Annual Salary	Annual Cost
Administrator/Superintendent	1	$25,000	$25,000
Process or Plant Engineer			
Mechanic—General	1	12,000	12,000
Mechanic—Electrical	1	12,500	12,500
Mechanic—Instruments			
Chemists	1	13,000	13,000
Laborers	2	4,500	9,000
Driver/Guard	1	5,000	5,000
Secretary/Clerk	1	6,000	6,000
Shift Supervisor	1	12,000	12,000
Shift Operator	5	7,500	37,500
TOTAL ANNUAL COST			$132,000

If there is no idea of the staff required nor the salary levels, then use a general estimated of overall labor costs. This estimate is $________ and ☐ Does ☐ Does Not include G&A.

30. What level of general and administrative (G&A) overhead is to be used? 25 %
The basis for this percentage is: Ministry of Water & Power
The annual G&A cost is: $ 33,000

FIGURE C-17. San Manne example (continued).

PROJECT DESCRIPTION San Manne RO Plant

CHEMICALS, SUPPLIES, AND MAINTENANCE MATERIALS

31.

Chemicals and Filters Description	Annual Quantity Needed	Local Unit Cost	Total Cost
Acid	2 lb/Kgal X 438 MG = 876,000 lb	0.06/lb.	$52,560
Sodium Hexametaphosphate	0.2 lb/Kgal X 438 MG = 87,600 lb	0.06/lb.	52,560
Chlorine	1 lb/Kgal X 438 MG = 438,000 lb	0.22/lb.	96,360
TOTAL			$201,480

* The cost of ocean freight & handling was estimated at doubling the cost shown in the manual.

32. Are there anticipated difficulties in purchasing, delivering, storing, and/or handling these materials during the life of the project? ☒ Yes ☐ No Details: There is often a 2 to 3 month delay in shipping chemicals to the island.

33. Cost of supplies and maintenance materials $19,150 Basis 1% of the cost of the desalting portion of the facility.

MEMBRANE REPLACEMENT

34. What type or configuration of membranes Either spiral or hollow fine fiber

35. Will spares be kept on hand? ☐ Yes ☒ No If yes, how many? ______
What membrane life is being used? 3 years years. For RO membranes, what is the guaranteed productivity at the end of that period versus at startup? 1.6 mgd will be specified in the contract
Annual replacement cost $100,000 Basis Figure C-15

SPECIAL REPAIRS

36. Will any major repairs or equipment replacement going to be required during the life of the project?
☒ Yes ☐ No Describe: Fluctuating voltage in the island power system will probably damage motors during the life of the plant.
Estimated annual cost $38,300 Basis 2% of plant cost (see #33)

ENERGY

37. Source(s) Electricity from the central power station

38. Dual- or single-purpose plant single

39. Cost of Electricity $0.10 /kWh; Electric usage 10 kWh/kgal ~~[kWh/m³]~~

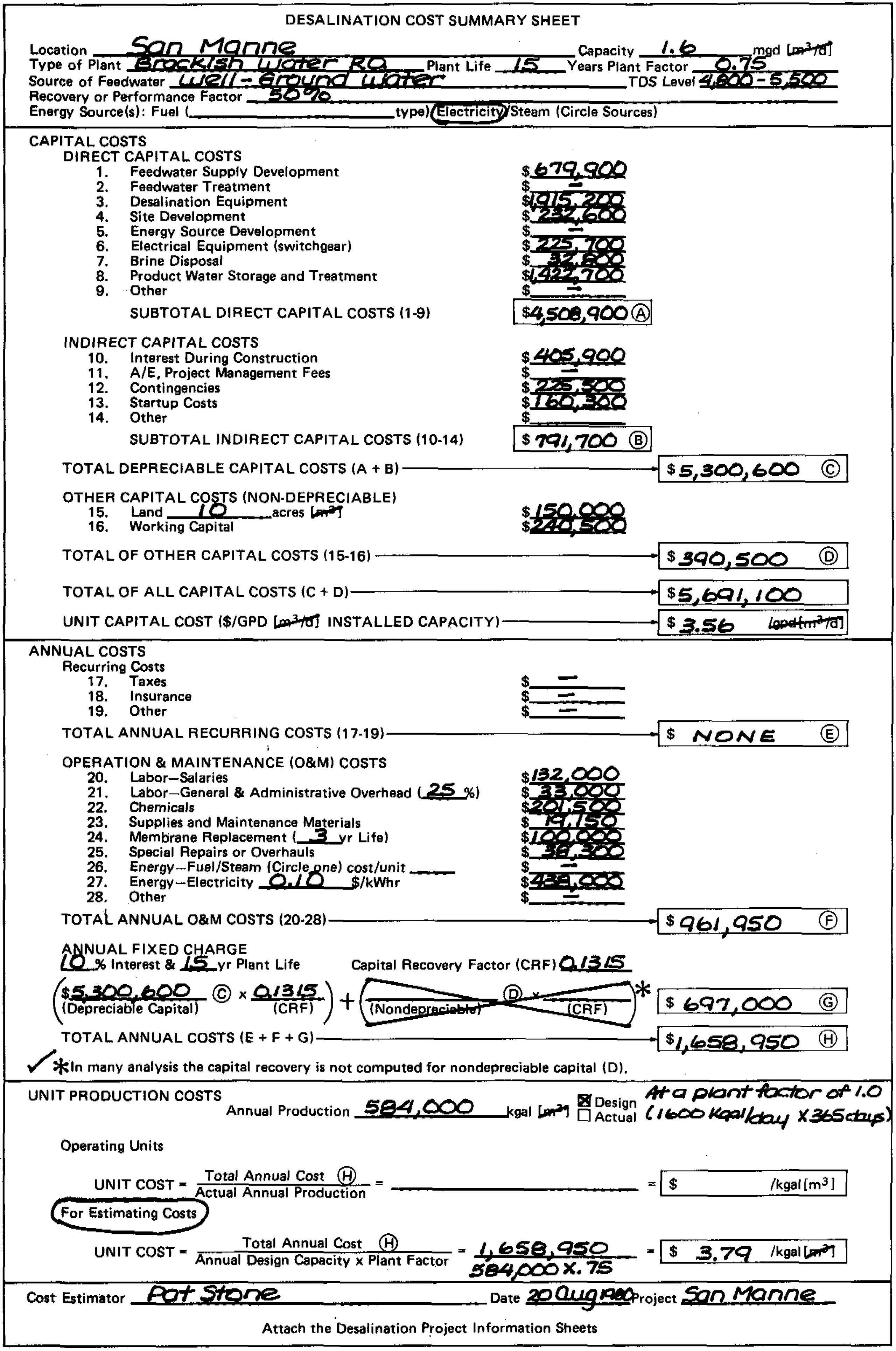
40. Does the electricity consumption include all pumping into and out of the plant? ☒ Yes ☐ No
If not, how much additional pumping energy will be required? —— kWh/kgal [kWh/m³]

41. If fuel is required what type? —— Heating value ______

42. Fuel usage (for diesel- or gas-operated plants) —— gal fuel/kgal [l/m³] water

43. Steam (for distillation plants)
Plant performance factor ______ Operating temperature ______ °F [°C]
Steam requirement ______ MBtu/kgal [GJ/m³] or ______/______
Single-purpose steam cost ______/MBtu [GJ] for prime steam @ ______ psig [atm] ______ °F [°C]
Dual-purpose steam cost ______/MBtu [GJ] at operating conditions ______ psig [atm] ______ °F [°C]

FIGURE C-17. San Manne example (continued).



DESALINATION COST SUMMARY SHEET

Location San Manne Capacity 1.6 mgd ~~[m³/d]~~
Type of Plant Brackish Water RO Plant Life 15 Years Plant Factor 0.75
Source of Feedwater Well - Ground Water TDS Level 4,800 - 5,500
Recovery or Performance Factor 50%
Energy Source(s): Fuel (_____ type) [✓ Electricity] / Steam (Circle Sources)

CAPITAL COSTS

DIRECT CAPITAL COSTS

1.	Feedwater Supply Development	$679,900	
2.	Feedwater Treatment	$ –	
3.	Desalination Equipment	$1,915,200	
4.	Site Development	$232,600	
5.	Energy Source Development	$ –	
6.	Electrical Equipment (switchgear)	$225,700	
7.	Brine Disposal	$32,800	
8.	Product Water Storage and Treatment	$1,422,700	
9.	Other	$ –	
	SUBTOTAL DIRECT CAPITAL COSTS (1-9)	$4,508,900 Ⓐ	

INDIRECT CAPITAL COSTS

10.	Interest During Construction	$405,900	
11.	A/E, Project Management Fees	$ –	
12.	Contingencies	$225,500	
13.	Startup Costs	$160,300	
14.	Other	$	
	SUBTOTAL INDIRECT CAPITAL COSTS (10-14)	$791,700 Ⓑ	
	TOTAL DEPRECIABLE CAPITAL COSTS (A + B)		$5,300,600 Ⓒ

OTHER CAPITAL COSTS (NON-DEPRECIABLE)

15.	Land 10 acres ~~[m²]~~	$150,000	
16.	Working Capital	$240,500	
	TOTAL OF OTHER CAPITAL COSTS (15-16)		$390,500 Ⓓ
	TOTAL OF ALL CAPITAL COSTS (C + D)		$5,691,100
	UNIT CAPITAL COST ($/GPD ~~[m³/d]~~ INSTALLED CAPACITY)		$3.56 ~~gpd [m³/d]~~

ANNUAL COSTS

Recurring Costs

17.	Taxes	$ –	
18.	Insurance	$ –	
19.	Other	$ –	
	TOTAL ANNUAL RECURRING COSTS (17-19)		$ NONE Ⓔ

OPERATION & MAINTENANCE (O&M) COSTS

20.	Labor—Salaries	$132,000	
21.	Labor—General & Administrative Overhead (25 %)	$33,000	
22.	Chemicals	$201,500	
23.	Supplies and Maintenance Materials	$19,150	
24.	Membrane Replacement (3 yr Life)	$100,000	
25.	Special Repairs or Overhauls	$38,300	
26.	Energy—Fuel/Steam (Circle one) cost/unit _____	$ –	
27.	Energy—Electricity 0.10 $/kWhr	$438,000	
28.	Other	$ –	
	TOTAL ANNUAL O&M COSTS (20-28)		$961,950 Ⓕ

ANNUAL FIXED CHARGE
10 % Interest & 15 yr Plant Life Capital Recovery Factor (CRF) 0.1315

($\frac{\$5{,}300{,}600}{\text{(Depreciable Capital)}}$ Ⓒ × $\frac{0.1315}{\text{(CRF)}}$) + ~~($\frac{}{\text{(Nondepreciable)}}$ Ⓓ × $\frac{}{\text{(CRF)}}$)~~* → $ 697,000 Ⓖ

TOTAL ANNUAL COSTS (E + F + G) → $1,658,950 Ⓗ

✓ *In many analysis the capital recovery is not computed for nondepreciable capital (D).

UNIT PRODUCTION COSTS

Annual Production 584,000 kgal ~~[m³]~~ ☒ Design ☐ Actual At a plant factor of 1.0 (1600 Kgal/day X 365 days)

Operating Units

UNIT COST = $\frac{\text{Total Annual Cost Ⓗ}}{\text{Actual Annual Production}}$ = __________ = $ _____ /kgal [m³]

[✓ For Estimating Costs]

UNIT COST = $\frac{\text{Total Annual Cost Ⓗ}}{\text{Annual Design Capacity × Plant Factor}}$ = $\frac{1{,}658{,}950}{584{,}000 \times .75}$ = $ 3.79 /kgal ~~[m³]~~

Cost Estimator Pat Stone Date 20 Aug 1980 Project San Manne

Attach the Desalination Project Information Sheets

FIGURE C-17. San Manne example (continued).

Appendix D
PROCESS SELECTION GUIDELINES

Appendix D consists of Table D-1, Summary of Guidelines for Desalting Process Selection. This table contains information on the characteristics, advantages, disadvantages, etc., of 18 different types of desalination equipment. The table provides selection guidelines on matching of equipment characteristics for the specific requirements and/or limitations of the application under study.

The table contains objective information such as the number and size of plants constructed and subjective evaluations regarding ease of operation and maintenance, etc., which can be used for guidance in the selection process.

The table is intended for overall assistance in process selection. By necessity it contains many generalities on the various types of equipment. Specific equipment and/or models produced by different manufacturers may have characteristics which vary from those shown in the table.

The table has been constructed so that the ten sheets comprising it can be reproduced and the copies joined together to form a large chart, five sheets (long side) wide and two sheets (short side) high. This large chart will show the 18 different types of desalination equipment across the top and the various characteristics, attributes, etc., for the processes in vertical columns. Thus, the various processes and their characteristics, etc., can be readily reviewed to aid in process selection and/or comparison.

The use of the table will aid in process selection but should not take the place of experienced, professional assistance in this task.

Table D-1
SUMMARY OF GUIDELINES FOR DESALTING PROCESS SELECTION

	Multistage (MSF)			
	Once-Through		Recycle	
	Low Temperature	High Temperature	Low Temperature	High Temperature
1980 Status				
Largest plant installed	1.25 mgd [4,730 m^3/d]	0.85 mgd [3,220 m^3/d]	6 mgd [22,700 m^3/d]	9 mgd [34,000 m^3/d]
Number of units installed[e]	Many	Few	Many	Many
Feedwater (FW)				
Source and TDS limit (mg/l)	Seawater, up to ~65,000	Seawater, up to ~65,000	Seawater, up to ~50,000	Seawater, up to ~50,000
Chemicals used in treatment	Polyphosphate or Polyelectrolyte	Polyelectrolyte	Polyphosphate or Polyelectrolyte	Acid or Polyelectrolyte
Typical chemical dosage (mg/l)	~0.5 to 4	~1 to 2	~4	Acid ~ 140; Poly ~ 5 to 10
Effect of higher FW temperature	Decreases output[c]	Decreases output[c]	Decreases output[c]	Decreases output[c]
Product Water (PW)				
TDS level (mg/l)	0.5 to ~25	0.5 to ~25	0.5 to ~25	0.5 to ~25
Reduction of bacterial level	High	High	High	High
Percent recovery	~9	10 to 15	~9	10 to 15
Typical Energy Consumption				
High-pressure steam (Btu/kgal)	~15,000 @ 100 to 150 psia	~15,000 @ 100 to 150 psia	~15,000 @ 100 to 150 psia	~15,000 @ 100 to 150 psia
Low-pressure steam (Btu/kgal)	~1,000,000 @ 18 psia	~700,000 @ 45 psia	~1,000,000 @ 18 psia	~700,000 @ 45 psia
Electricity (kWh/kgal) [kWh/m^3]	10 to 18 [2.6 to 4.8]	6 to 10 [1.6 to 2.6]	10 to 18 [2.6 to 4.8]	6 to 10 [1.6 to 2.6]
Performance				
Maximum operating temperature	195°F [90.6°C]	250°F [121.1°C]	195°F [90.6°C]	250°F [121.1°C]
Performance factor (lb/1,000 Btu) [kg/MJ]	Up to 8.5 [3.65]	8 to 13.5 [3.43 to 5.79]	Up to 8.5 [3.65]	8 to 13.5 [3.43 to 5.79]
Operating pressure (RO and ED only)	--	--	--	--
Relative O&M Characteristics				
Operational attention required	Minimal--plant can be automated	Minimal--plant can be automated	Continual--operation easily upset	Close, continual
Maintenance attention required	Medium[b]	Medium high[b]	High[b]	Highest[b] (acid); High[b] (poly)

	Multistage (MSF)			
	Once-Through		Recycle	
	Low Temperature	High Temperature	Low Temperature	High Temperature
Potential for scale formation	Low to medium[b]	Medium high[b]	Medium high[b]	Acid[b]--highest; Poly[d]--probably comparable[b]
Potential for corrosion	Medium high	Medium high	Medium high	Acid--highest; Poly--medium high
Membrane life (years)	--	--	--	--
Relative Advantages	1. High-quality (low-TDS) product. 2. Easy to start up, control, and shut down. 3. Can use low-cost turbine extraction steam. 4. Available from many manufacturers. 5. Can be built in large capacities. 6. Insensitive to sudden process changes.	1. High-quality (low-TDS) product. 2. Easy to start up, control, and shut down. 3. Can use low-cost turbine extraction steam. 4. Available from many manufacturers. 5. Can be built in large capacities. 6. Insensitive to sudden process changes.	1. High-quality (low-TDS) product. 2. Long history of operation. 3. Can use low-cost turbine extraction steam. 4. Available from many manufacturers. 5. Can be built in large capacities.	1. High-quality (low-TDS) product. 2. Long history of operation for acid plants. 3. Can use low-cost turbine extraction steam. 4. Available from many manufacturers. 5. Can be built in the largest capacities of any type plant.
Relative Disadvantages	1. Fairly high maintenance unless made of high-quality materials. 2. Relative to the high-temperature once-through MSF plant, it has a higher capital cost and uses more steam and power for pumping. 3. Limited performance factor.	1. Fairly high maintenance unless made of high-quality materials. 2. High-temperature excursion can produce hard scale.	1. Fairly high maintenance unless made of high-quality materials. 2. Relative to the high-temperature recycle MSF plant, it has a higher capital cost and uses more steam and power for pumping. 3. Requires constant operator attention; easily upset.	1. Acid plant has very high maintenance unless made of high-quality materials. The poly plants will probably have fairly high maintenance unless made of high-quality materials. 2. High-temperature excursion can produce hard scale. 3. Requires constant operator attention; easily upset. 4. Control of the acid addition is very critical to operation.
Principal Applications	Produces potable, process, or boiler feedwater from seawater.	Produces potable, process, or boiler feedwater from seawater.	Produces potable, process, or boiler feedwater from seawater.	Produces potable, process, or boiler feedwater from seawater.

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[a] Output and recovery depend on the difference in temperature between the waste heat and the cooling water.
[b] Depending on materials and construction.
[c] Output decreases since feedwater is also used for cooling. Heated feedwater would reduce the overall temperature drop in the plant.
[d] Insufficient field experience to fully evaluate.
[e] Few--0-20 plants; Some = 20-50 plants; Many--50 or more plants.
[f] Includes some units under construction.
[g] Projected--not in commercial operation.
[h] Upper temperature limited by membranes.
[i] Chlorine may or may not be needed. With chlorination, some units will then require dechlorination to protect the membrane.
[j] TDS--2,500 mg/l.
[k] TDS--4,500 mg/l.
[l] In the freezer.
[m] Metric equivalents: Btu/kgal x 0.279 = kJ/m^3; psi ÷ 14.7 = atm.

Table D-1--Continued

Multiple-Effect (ME)

	Horizontal Tube		Vertical Tube	
	Low Temperature	High Temperature	Upflow	Downflow
1980 Status				
Largest plant installed	1 mgd [3,790 m^3/d]	0.8 mgd [3,030 m^3/d]	0.025 mgd [95 m^3/d]	4.5 mgd [17,000 m^3/d]
Number of units installed[e]	Few	Some	Few--Pilot only	Many
Feedwater (FW)				
Source and TDS limit (mg/l)	Seawater, to ~70,000	Seawater, to ~70,000	Seawater, to ~70,000	Seawater, to ~70,000
Chemicals used in treatment	Polyphosphate	Acid	(Acid/Polyelectrolyte) + Surfactant	Acid
Typical chemical dosage (mg/l)	~5	125 to 140	(~135/~8) + 50	125 to 140
Effect of higher FW temperature	Reduces output[c]	Reduces output[c]	Reduces output[c]	Reduces output[c]
Product Water (PW)				
TDS level (mg/l)	0.5 to ~25	0.5 to ~25	0.5 to ~25	0.5 to ~25
Reduction of bacterial level	High	High	High	High
Percent recovery	8 to 15	10 to 25	10 to 25	10 to 25
Typical Energy Consumption				
High-pressure steam (Btu/kgal)	~15,000 @ 100 to 150 psia	~15,000 @ 100 to 150 psia	~15,000 @ 100 to 150 psia	~15,000 @ 100 to 150 psia
Low-pressure steam (Btu/kgal)	~1,000,000 @ 6 to 18 psia	400,000 to 2,000,000 @ 45 psia	400,000 to 2,000,000 @ 45 psia	400,000 to 2,000,000 @ 45 psia
Electricity (kWh/kgal) [kWh/m^3]	4.5 to 6 [1.2 to 1.6]	3.5 to 5 [0.9 to 1.3]	4.5 to 8 [1.2 to 2.1]	4.5 to 8 [1.2 to 2.1]
Performance				
Maximum operating temperature	160°F [71.1°C]	~250 [121.1°C]	~250 [121.1°C]	~250°F [121.1°C]
Performance factor (lb/1,000 Btu) [kg/MJ]	Up to 8.5 [3.65]	4 to 20 [1.72 to 8.6]	4 to 20 [1.72 to 8.6]	4 to 20 [1.72 to 8.6]
Operating pressure (RO and ED only)	--	--	--	--
Relative O&M Characteristics				
Operational attention required	Minimal	Minimal	--	Close--continual
Maintenance attention required	Average to low[b]	Average to low[b]	--	High

	Multiple-Effect (ME)			
	Horizontal Tube		Vertical Tube	
	Low Temperature	High Temperature	Upflow	Downflow
Potential for scale formation	Low	Low	Less than for downflow mode	Very high
Potential for corrosion	Low	Low	--	High
Membrane life (years)	--	--	--	--
<u>Relative Advantages</u>	1. High-quality (low-TDS) product. 2. Easy and fast to start up, operate, and shut down 3. Can use low-cost turbine extraction steam. 4. Uses less power for pumping than MSF or VTE. 5. Lower capital and operating costs than MSF or VTE. 6. Can operate on waste heat. 7. History of good availability.	1. High-quality (low-TDS) product. 2. Easy and fast to start up, operate, and shut down. 3. Can use low-cost turbine extraction steam. 4. Uses less power for pumping than MSF or VTE. 5. Lower capital and operating costs than MSF or VTE. 6. Can be designed for a high performance factor. 7. History of good availability.	1. High-quality (low-TDS) product. 2. Fast startup and shutdown. 3. Can use low-cost turbine extraction steam. 4. Uses less power for pumping than MSF. 5. Somewhat lower capital cost than MSF. 6. Can be designed for a high performance factor. 7. Can operate at low temperature on waste heat.	1. High-quality (low-TDS) product. 2. Fast startup and shutdown. 3. Can use low-cost turbine extraction steam. 4. Uses less power for pumping than MSF. 5. Somewhat lower capital cost than MSF. 6. Can be designed for a high performance factor.
<u>Relative Disadvantages</u>	1. Published operating data are limited. 2. Available from only 4 to 6 manufacturers. 3. Units larger than 1 mgd have not yet been operated. 4. Much less design and operating experience than with MSF plants.	1. Published operating data are limited. 2. Available from only 4 to 6 manufacturers. 3. Units larger than 0.8 mgd have not yet been operated. 4. Much less design and operating experience than with MSF plants.	1. High-temperature excursions during operation can produce hard scale. 2. Surfactant must be recovered from brine; it cannot be discharged. 3. Pilot plant only--no commercial experience.	1. High-temperature excursion or maldistribution of feed can produce hard scale. 2. Requires constant operator attention. 3. Fairly high maintenance unless made of high-quality materials. 4. Much less design and operating experience than with MSF plants. 5. High mechanical complexity because of many pumps.
<u>Principal Applications</u>	Produces potable, process, or boiler feedwater from seawater.	Produces potable, process, or boiler feedwater from seawater.	Proposed for desalting seawater or concentrating waste streams.	Produces potable, process, or boiler feedwater from seawater.

---

[a] Output and recovery depend on the difference in temperature between the waste heat and the cooling water.

[b] Depending on materials and construction.

[c] Output decreases since feedwater is also used for cooling. Heated feedwater would reduce the overall temperature drop in the plant.

[d] Insufficient field experience to fully evaluate.

[e] Few--0-20 plants; Some = 20-50 plants; Many--50 or more plants.

[f] Includes some units under construction.

[g] Projected--not in commercial operation.

[h] Upper temperature limited by membranes.

[i] Chlorine may or may not be needed. With chlorination, some units will then require dechlorination to protect the membrane.

[j] TDS--2,500 mg/l.

[k] TDS--4,500 mg/l.

[l] In the freezer.

[m] Metric equivalents: Btu/kgal x 0.279 = kJ/m^3; psi ÷ 14.7 = atm.

Table D-1--Continued

	Waste Heat	Vapor Compression (VC)		
		Mechanical Drive		
	Submerged Tube or Flash and/or Multiple-Effect	Low Temperature	High Temperature	Steam Jet
1980 Status				
Largest plant installed	0.07 mgd [265 m^3/d]	0.5 mgd [1.890 m^3/d]	1.0 mgd [3,785 m^3/d]	0.6 mgd [2,270 m^3/d]
Number of units installed[e]	Many	Many	Many	Some
Feedwater (FW)				
Source and TDS limit (mg/l)	Heated waste stream, any TDS	Seawater up to ~70,000	Seawater, up to ~50,000	Seawater, up to ~60,000
Chemicals used in treatment	None	Polyphosphate	Acid or Polyelectrolyte	Polyphosphate
Typical chemical dosage (mg/l)	None	0.5	~140 or 4 to 10	~5
Effect of higher FW temperature	Not applicable[a]	Capacity increases	Slight improvement in economy	Slight improvement in economy
Product Water (PW)				
TDS level (mg/l)	~25	0.5 to ~25	0.5 to ~25	0.5 to ~25
Reduction of bacterial level	No data	No data	High	No data
Percent recovery	About 1 to 2[a]	~40	~50	~50
Typical Energy Consumption				
High-pressure steam (Btu/kgal)	None	None	None	1,700,000 @ 120 psia
Low-pressure steam (Btu/kgal)	None	None	None	None
Electricity (kWh/kgal) [kWh/m^3]	15 to 35 [4 to 9.2]	64 [16.9]	65 to 100 [17.2 to 26.4]	15 [4]
Performance				
Maximum operating temperature	Usually below 150°F [65.6°C]	113°F [45°C]	215°F [102°C]	125°F [51.7°C]
Performance factor (lb/1,000 Btu) [kg/MJ]	0.95 [0.41]	--	--	--
Operating pressure (RO and ED only)	--	--	--	--
Relative O&M Characteristics				
Operational attention required	Minimal, plant can be automated	Minimal, plant can be automated	Minimal, plant can be automated	Very low
Maintenance attention required	Minimal	Minimal, but skilled	Minimal, but skilled	Minimal, comparatively unskilled

	Waste Heat	Vapor Compression (VC)		
	Submerged Tube or Flash and/or Multiple-Effect	Mechanical Drive		
		Low Temperature	High Temperature	Steam Jet
Potential for scale formation	Very low	Very low	Low	Very low
Potential for corrosion	Very low, even on steel	Very low	Low to high[b]	--
Membrane life (years)	--	--	--	--
Relative Advantages	1. High-quality (low-TDS) product. 2. Operates largely unattended. Easy to start up, operate, and shut down. 3. Standard package units are available for rapid delivery and quick installation and startup. 4. Can be fully automated. 5. Requires power only for process pumps. 6. Operating economics are attractive where waste heat is available. 7. Danger of operating error damaging the unit is small.	1. High-quality (low-TDS) product. 2. Operates largely unattended. Easy to start up, operate, and shut down. 3. Standard package units are available for rapid delivery and quick installation and startup. 4. Can be fully automated. 5. Can be installed anyplace where electricity or diesel fuel are available. 6. Economics are favorable for small plants. 7. Danger of scaling or damaging plant by operating error is small.	1. High-quality (low-TDS) product. 2. Operates largely unattended. Easy to start up, operate, and shut down. 3. Standard package units are available for rapid delivery and quick installation and startup. 4. Can be fully automated. 5. Can be installed anyplace where electricity or diesel fuel are available. 6. Economics are favorable for small plants. 7. Has neither a deaerator nor a decarbonator. Requires no vacuum system.	1. High-quality (low-TDS) product. 2. Operates largely unattended. Easy to start up, operate, and shut down. 3. Standard package units are available for rapid delivery and quick installation and startup. 4. Can be fully automated. 5. Practically foolproof operation.
Relative Disadvantages	1. Unit sizes are comparatively small (vs. MSF plants). 2. Published operating history is limited. 3. High capital cost.	1. Unit sizes are comparatively small (vs. MSF plants). 2. Published operating history is limited. 3. High-speed compressor requires skilled maintenance. 4. Operates under vacuum and therefore requires a vacuum system. 5. Only one supplier (Israel Desalination Engineering, Ltd.).	1. Unit sizes are comparatively small (vs. MSF plants). 2. Published operating history is limited. 3. High-speed compressor requires skilled maintenance.	1. Unit sizes are comparatively small (vs. MSF plants). 2. Published operating history is limited. 3. Requires a boiler or other steam source. 4. Not economical in terms of energy.
Principal Applications	Provides potable, process, or boiler feedwater from warm waste streams.	Provides potable, process, or boiler feedwater from brackish or sewater supplies.	Provides potable, process, or boiler feedwater from brackish or seawater supplies.	Provides potable, process, or boiler feedwater from brackish or seawater, especially at construction sites.

---

[a] Output and recovery depend on the difference in temperature between the waste heat and the cooling water.

[b] Depending on materials and construction.

[c] Output decreases since feedwater is also used for cooling. Heated feedwater would reduce the overall temperature drop in the plant.

[d] Insufficient field experience to fully evaluate.

[e] Few--0-20 plants; Some = 20-50 plants; Many--50 or more plants.

[f] Includes some units under construction.

[g] Projected--not in commercial operation.

[h] Upper temperature limited by membranes.

[i] Chlorine may or may not be needed. With chlorination, some units will then require dechlorination to protect the membrane.

[j] TDS--2,500 mg/l.

[k] TDS--4,500 mg/l.

[l] In the freezer.

[m] Metric equivalents: Btu/kgal x 0.279 = kJ/m^3; psi ÷ 14.7 = atm.

Table D-1--Continued

Seawater

	Freezing	Reverse Osmosis	Electrodialysis
	Vacuum-Freezing Refrigeration Compression (VFRC)	HFF or Spiral Membranes (SWRO)	Standard (SWED)
1980 Status			
Largest plant installed	0.025 mgd [95 m^3/d]	3.2 mgd [12,100 m^3/d]	0.03 mgd [114 m^3/d]
Number of units installed[e]	Few--pilot only	Some	Few
Feedwater (FW)			
Source and TDS limit (mg/l)	Brackish water or seawater	Seawater, 35,000-45,000	Seawater, 35,000-45,000
Chemicals used in treatment	Uusally none	Acid/SHMP	Acid
Typical chemical dosage (mg/l)	--	240-360/12-36	300
Effect of higher FW temperature	Increased power consumption	Increased efficiency[h]	Increased efficiency[h]
Product Water (PW)			
TDS level (mg/l)	As required, usually ~100	As required, usually ~500	As required, usually ~500
Reduction of bacterial level	Some	High	Negligible
Percent recovery	25	20-35	50-60
Typical Energy Consumption			
High-pressure steam (Btu/kgal)	None	None	None
Low-pressure steam (Btu/kgal)	None	None	None
Electricity (kWh/kgal) [kWh/m^3]	35 to 50 [9.2 to 13.2]	30 to 40 [7.9 to 10.6]	60 to 65 [15.9 to 17.2]
Performance			
Maximum operating temperature	20 to 30 of [-6.7 to -1.1°C]	~95°F [35°C]	~110°F [43.3°C]
Performance factor (lb/1,000 Btu) [kg/MJ]	--	--	--
Operating pressure (RO and ED only)	--	800 to 1,000 psi [54.4 to 68 atm]	60 to 100 psi [4.1 to 6.8 atm]
Relative O&M Characteristics			
Operational attention required	Minimal--can be automated[g]	Minimal--can be automated	Minimal--can be automated
Maintenance attention required	Minimal[g]	Minimal	Minimal

	Seawater		
	Freezing	Reverse Osmosis	Electrodialysis
	Vacuum-Freezing Refrigeration Compression (VFRC)	HFF or Spiral Membranes (SWRO)	Standard (SWED)
Potential for scale formation	Very low	Average	Average
Potential for corrosion	Very low	Very low	Very low
Membrane life (years)	--	3	Not known
Relative Advantages	1. Can be used to treat virtually any type of feed. 2. No chemicals or pretreatment necessary. 3. Corrosion is minimal. 4. No scale formation. 5. Can be operated anywhere power is available. 6. Moderate energy consumption for seawater desalination or for highly concentrating difficult brackish water. 7. Minimal O&M manpower projected.[g]	1. Fairly simple to operate compared to most distillation units. 2. Comes in a variety of sizes to match demand. 3. Energy cost is reasonable compared to single-purpose distillation plants. 4. Easy to quickly install (vs. distillation units). 5. An energy recovery system can be used to reduce overall energy consumption.	1. Fairly simple to operate compared to most distillation units. 2. Potential for significant energy savings by using hot feedwater. 3. Not as sensitive to suspended materials in the feedwater as RO. 4. Operates at low pressure. 5. Corrosion is not a major problem. 6. More forgiving of operational malfunctions than RO as the stacks can be disassembled and scales or other foulants can be cleaned by hand.
Relative Disadvantages	1. Not yet commercial--still in pilot plant stage. 2. Not yet demonstrated in large capacities. 3. Few suppliers--would be custom-made. 4. High energy usage for brackish water desalting.	1. Only a few years of full commercial experience. The precise service life of components, such as the membranes, is not certain. 2. Needs good pretreatment to remove suspended material in the water. 3. Operates at high pressures. 4. There is very little operating experience using energy recovery. 5. There are not significant economies-of-scale in large units (such as with distillation plants).	1. Only one land-based plant operating. It is on Noshima Island, Japan, made by Asahi Chemical Ind. Co., Ltd. 2. Little general operating experience. 3. Few suppliers--would be custom-made. 4. Has a higher energy consumption than most SWRO units if operated at ambient temperature. 5. No reduction in bacterial levels.
Principal Applications	Produces potable water from brackish water or seawater. Concentration of waste streams and recovery of byproducts.	Produces potable or process feedwater from seawater.	Produces potable or process feedwater from seawater.

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[a] Output and recovery depend on the difference in temperature between the waste heat and the cooling water.

[b] Depending on materials and construction.

[c] Output decreases since feedwater is also used for cooling. Heated feedwater would reduce the overall temperature drop in the plant.

[d] Insufficient field experience to fully evaluate.

[e] Few--0-20 plants; Some = 20-50 plants; Many--50 or more plants.

[f] Includes some units under construction.

[g] Projected--not in commercial operation.

[h] Upper temperature limited by membranes.

[i] Chlorine may or may not be needed. With chlorination, some units will then require dechlorination to protect the membrane.

[j] TDS--2,500 mg/l.

[k] TDS--4,500 mg/l.

[l] In the freezer.

[m] Metric equivalents: Btu/kgal x 0.279 = kJ/m^3; psi ÷ 14.7 = atm.

Table D-1--Continued

Brackish Water

	Reverse Osmosis	Electrodialysis	
	HFF or Spiral Membranes	Standard (ED)	Reversal (EDR)
1980 Status			
Largest plant installed	12 mgd [45,400 m^3/d]	5 mgd [18,900 m^3/d]	4 mgd [15,140 m^3/d]
Number of units installed[e]	Many	Many	Many
Feedwater (FW)			
Source and TDS limit (mg/l)	Brackish water up to 10,000	Brackish water up to 7,000	Brackish water up to 7,000
Chemicals used in treatment	Chlorine[i]/SHMP/Acid	Chlorine[i]/Acid/Base	Generally none
Typical chemical dosage (mg/l)	120/12-36/180-300	120/180/36	--
Effect of higher FW temperature	Increased production[h]	Increased efficiency[h]	Increased efficiency[h]
Product Water (PW)			
TDS level (mg/l)	As required, generally ~500	As required, generally ~500	As required, generally ~500
Reduction of bacterial level	High	Negligible	Negligible
Percent recovery	50 to 90	50 to 95	50 to 95
Typical Energy Consumption			
High-pressure steam (Btu/kgal)	None	None	None
Low-pressure steam (Btu/kgal)	None	None	None
Electricity (kWh/kgal) [kWh/m^3]	6-12 [1.6-3.2]	7.5[j] to 12.5[k] [2[j] to 3.3[k]]	7.5[j] to 12.5[k] [2[j] to 3.3[k]]
Performance			
Maximum operating temperature	85-120°F [29.3-48.9°C]	~110°F [43.3°C]	~110°F [43.3°C]
Performance factor (lb/1,000 Btu) [kg/MJ]	NA	NA	NA
Operating pressure (RO and ED only)	250 to 400 psi [17 to 27.2 atm]	50 to 60 psi [3.4 to 4.1 atm]	50 to 60 psi [3.4 to 4.1 atm]
Relative O&M Characteristics			
Operational attention required	Minimal but careful--usually automated	Minimal--usually automated	Minimal--almost always automated
Maintenance attention required	Minimal	Average--stack cleaning	Minimal

	Brackish Water		
	Reverse Osmosis	Electrodialysis	
	HFF or Spiral Membranes	Standard (ED)	Reversal (EDR)
Potential for scale formation	Possible in the membranes	Possible in the stacks	Minimal
Potential for corrosion	Exists for pumps, fitting, etc.	Minimal--much plastic used	Minimal--much plastic used
Membrane life (years)	3 to 5	5 to 10	5 to 10
Relative Advantages	1. 10-12 years of experience in design and operation. Many units operating. 2. Fairly simple to operate compared to most distillation units. 3. Comes in a variety of sizes. 4. Easy to quickly install small units.	1. 25 years of experience in design and operation. Many units operating. 2. Fairly simple to operate compared to most distillation units. 3. Comes in a variety of sizes. 4. Operates at a low hydraulic pressure--quiet operation. 5. Forgiving of operational malfunctions--stacks can be disassembled and cleaned. 6. Can operate at high recovery. 7. Membranes can be cleaned two ways--disassembly and clean-in-place.	1. 8 years of experience in design and operation. Many units operating. 2. Fairly simple to operate compared to most distillation units. 3. Comes in a variety of sizes. 4. Operates at a low hydraulic pressure--quiet operation. 5. Forgiving of operational malfunctions--stacks can be disassembled and cleaned. 6. Can operate at high recovery. 7. Membranes can be cleaned in three ways--disassembly, clean-in-place, and reversal. 8. No chemicals usually required.
Relative Disadvantages	1. Compared to distillation, the product is high in TDS. 2. Very large plants are multiples of smaller standard modules so that the economies-of-scale for large plants are limited. 3. Operates at a high hydraulic pressure. 4. Not as forgiving of operational malfunction as electrodialysis. 5. Vibration and noise from high-pressure pump can cause problems. 6. Needs chemicals for operation.	1. Compared to distillation, the product is high in TDS. 2. Very large plants are multiples of smaller standard modules so that the economies-of-scale for large plants are limited. 3. Depending on the water source and operation, the membrane stacks may require frequent disassembly and cleaning. Involves considerable hand labor. 4. Does not remove silica, bacteria, and suspended material. 5. Needs chemicals for operation.	1. Compared to distillation, the product is high in TDS. 2. Very large plants are multiples of smaller standard modules so that the economies-of-scale for large plants are limited. 3. Does not remove silica, bacteria, or suspended material. 4. Complicated plumbing required with several motor-operated valves and automation.
Principal Applications	Provides potable or process water from brackish supplies.	Provides potable or process water from brackish supplies.	Provides potable or process water from brackish supplies.

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[a] Output and recovery depend on the difference in temperature between the waste heat and the cooling water.

[b] Depending on materials and construction.

[c] Output decreases since feedwater is also used for cooling. Heated feedwater would reduce the overall temperature drop in the plant.

[d] Insufficient field experience to fully evaluate.

[e] Few--0-20 plants; Some = 20-50 plants; Many--50 or more plants.

[f] Includes some units under construction.

[g] Projected--not in commercial operation.

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[j] TDS--2,500 mg/l.

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[l] In the freezer.

[m] Metric equivalents: Btu/kgal x 0.279 = kJ/m^3; psi ÷ 14.7 = atm.

Appendix E
OTHER SOURCES OF INFORMATION ABOUT DESALINATION

The field of desalination is developing very rapidly. Although some textbooks and research texts have been written, the main thrust of commercial desalination has been in the form of published technical papers and reports. These include discussions of theoretical and commerical developments, performance of installations and processes, and special studies and surveys.

There are three conferences held on an annual or bi-annual basis whose proceedings form a valuable contribution to the industry's literature:

European Federation of Chemical Engineering's International Symposium on Freshwater From the Sea. This conference is held in various locations in Europe, generally in even-numbered years. For information on the conferences and past proceedings write:

Dr. Delyannis
Plastira St. 3
Amarousion, Pefki Greece

International Desalination and Environmental Association (IDEA). This conference is held in various locations around the world, generally in odd-numbered years. This is a membership organization which issues a bi-monthly newsletter containing news and technical articles. For information write:

IDEA Headquarters
1000 River Road
Teaneck, New Jersey USA 07666

National Water Supply Improvement Association (NWSIA). This conference is held annually in various locations in the USA. This is a membership organization which issues a bi-monthly general newsletter with membership news and a semiannual technical journal. For information write:

NWSIA Headquarters
26 Newbury Road
Ipswich, Massachusetts USA 01938

The U.S. Government's Office of Water Research and Technology (previously the Office of Saline Water) has issued over 1,000 reports on various aspects of desalination. A list of these reports can be obtained by writing:

The Office of Water Research and Technology
U.S. Department of Interior
Washington, D.C. USA 20240

Almost all of these reports are available from the National Technical Information Service (NTIS). The appropriate NTIS order number is noted in the list mentioned above. The address of NTIS is:

National Technical Information Service
U.S. Department of Commerce
Springfield, Virginia USA 22161

There is a charge for these publications.

Considerable helpful technical and commercial information can be obtained free of charge by writing the various desalination equipment and membrane manufacturers listed in Appendices A and B.

A weekly 4-page newsletter called "Water Desalination Report" covers much of the industry's activity, especially in the USA. The annual cost of this newsletter is about $250 in the USA and $275 overseas. For information write:

"Water Desalination Report"
P.O. Box 35-K
Tracey's Landing, Maryland USA 20869

A journal, Desalination, is issued several times a year containing technical articles on desalination. Desalination often publishes, as an issue, the proceedings from various international conferences. For information write:

Elsevier Scientific Publishing Company
P.O. Box 330
Amsterdam, The Netherlands

A number of textbooks exist which would be of value as reference material and as supplements to this manual. They would be more technical in nature and would be of major assistance to engineers interested in desalination. Two textbooks that might be obtained are:

Fundamentals of Water Desalination by E. D. Howe, published by Marcel Dekker, Inc., New York, 1974.

Principles of Desalination ed. by K. S. Spiegler and A. D. K. Laird. The 2nd edition is a 2-volume set which will be published by Academic Press, N.Y., in the latter part of 1980.

GLOSSARY

Agglomeration--The state of being gathered into a mass.

Approach Temperature--The difference in temperature between two streams entering and exiting a heat exchanger.

Boiling Point Rise--The elevation of the boiling point of a liquid due to the amount of salts in solution.

Downcomer--A pipe directed downward.

Draft (iceberg)--The depth below the surface of the water to which an iceberg extends.

Economy-of-Scale--The reduction in unit capital cost as the plant size increases.

Electrolyte--Any substance which, when placed in solution, dissociates into ions and is then capable of passing an electrical current.

Eutectic--Easily melted.

Eutrophication--The process by which an area, generally a body of water, becomes increasingly rich in nutrients. In bodies of water these excess nutrients can cause the growth of large quantities of planktonic algae. The algae in turn can reduce clarity and the dissolved oxygen level of the water at various times.

Galvanic Couple--A cell in which chemical change is the source of electrical energy. It can be formed from two different conductors in an electrolyte, a conductor in two different electrolytes, or one electrolyte at two different concentrations.

Ice Nucleation--The formation of new ice nuclei, either spontaneously or by bits breaking off of larger pieces of ice.

Immiscible Substance--A substance which cannot be mixed with another.

Motive Steam--High-pressure steam which is used to operate a steam-jet ejector.

Order-of-Magnitude Estimate--An approximate estimate made without detailed engineering data. It is normally expected that an estimate of this type would be accurate within +50 percent or -30 percent.

Surfactant--A surface-active agent such as a detergent which, when mixed with water, generally increases its cleaning ability, solubility, and penetration, while reducing its surface tension.

Thermosetting--The property of some pliable substances by which they become rigid when heated but incapable of being readily remelted or softened.

Transient--A momentary and instantaneous change in voltage level.

Triple Point--The condition of temperature, pressure, and salinity at which the solid, liquid, and gaseous states of a substance (such as water) coexist simultaneously in equilibrium.

Water Hammer--The phenomenon of oscillations in the pressure of water above its normal pressure in a closed conduit, flowing full, which results from too rapid acceleration or retardation of flow. Momentary pressures greatly in excess of the normal static pressure may be produced in a closed conduit by this phenomenon. (APHA, 1969.)

REFERENCES

ALCOA. Alcoa 387 Process. ALCOA, Alcoa, Pennsylvania.

Al Faisal, H.R.H. Prince M., 1976. "Water Supply and Weather Modifications through the Use of Transported Icebergs from the Antarctic." In Proceedings of the First Desalination Congress of the American Continent. October 24-29, 1976, Mexico City.

Al Faisal, H.R.H. Prince M., and Ismail, S. 1978. "Feasibility of Using Paddle-Wheels for the Propulsion of Icebergs." In Iceberg Utilization. A. A. Husseiny, ed. Pergamon Press, New York.

Al Gholaikah, A., El-Ramly, N., Jamjoom, I., and Seaton, R., 1979. "The World's First Large Seawater Reverse Osmosis Desalination Plant at Jeddah, Kingdom of Saudi Arabia." NWSIA Jour., 6, 1.

Al Taweel, A. M., Bushnak, A., and Bawarith, W., 1980. "A Strategy for the Use of Agricultural Drainage Waters in Saudi Arabia." NWSIA Journal, 7, 2.

Andreadis, G., and Arnold, J. W., 1978. Development of a Quality Water Supply for the Island of Corfu, Greece. In Proceedings of the Sixth Annual Conference of NWSIA. July 16-20, 1978. Sarasota, Florida.

APHA, New York. 1969. Glossary--Water and Wastewater Control Engineering.

Aqua-Chem, Inc., 1979. Material presented at Fairleigh Dickinson University Seminar "Desalination Operations--1979." December 9-14, 1979, St. Croix, U.S. Virgin Islands.

Arazzini, S., Tomei, G., Ciccone, E., and Fiorito, G., 1979. "Technical and Economical Evaluation of MSF Solar Energy Powered Desalination Plant." In Proceedings of the International Congress on Desalination and Water Reuse. October 21-27, 1979, Nice, France.

Arnold, J. W., 1979. "Operating Experience of a 15,000 Cubic Meter Per Day Municipal Desalting Plant at Corfu, Greece." In Proceedings of the International Congress on Desalination and Water Reuse. October 21-27, 1979, Nice, France.

Babcock & Wilcox Company, 1963. Steam--Its Generation and Use. Babcock & Wilcox Company, New York.

Bader, H., 1978. "A Critical Look at the Iceberg Project." In Iceberg Utilization. A. A. Husseiny, ed. Pergamon Press, New York.

Bailie, R., 1977. "Feasibility of Expansion or Modification of the VTE/MSF Desalting Module at Water Factory 21." In The Technical Proceedings of the Fifth Annual Conference of the National Water Supply Improvement Association. July 17-21, 1977. San Diego, California.

Baker, M. N., 1948. The Quest for Pure Water. AWWA, New York.

Bakish, R., 1979. "On the 'E' in IDEA." An editorial in Pure Water, 8, 2, p.1.

Barnett, K. M., 1978. A Wind Energy Summary for New Mexico. Report under auspices of New Mexico Energy Institute.

Benedict, C. P., 1978. "A Towing Concept for Small Icebergs." In Iceberg Utilization. A. A. Husseiny, ed. Pergamon Press, New York.

Bloemer, J. W., Eibling, J. A., and Irwin, J. R., 1965. "A Practical Basin-Type Solar Still." Solar Energy, 9, 4.

Böddecker, W., Hilgendorff, J., and Kaschemekat, J., 1976. "An Alternative Design Concept in Reverse Osmosis Desalination." In Proceedings of the First Desalination Congress of the American Continent. October 24-29, 1976, Mexico City.

Boegli, W. J., Suemoto, S. H., and Trumpeter, K. M., 1977. "Geothermal Desalting of the East Mesa Site." In Proceedings of the International Congress on Desalination and Water Reuse. November 27-December 3, 1977, Tokyo.

Bosich, J. F., 1970. Corrosion Prevention for Practicing Engineers. Barnes & Noble, Inc., New York.

Brace Research Institute, 1979. Commercially Available Windmill Systems and Components. BRI Report No. F.16. Quebec, Canada.

Brandon, C. A., 1980. "Closed-Cycle Textile Dyeing: Full Scale Hyperfiltration Demonstration." NWSIA Journal, 7, 2.

Breton, E. J., 1957. Water and Ion Flow Through Imperfect Osmotic Membranes, OSW Report No. 16.

Burns and Roe, Inc., 1979. Parametric Study on Brackish Water Membrane Desalting Plants. OWRT Report. NTIS No. 294 080.

Buros, O. K., 1976a. Wastewater Reclamation Project, St. Croix, U.S. Virgin Islands. EPA--600/2-76-134. EPA, Cincinnati, Ohio.

Buros, O. K., 1976b. "Conjunctive Use of Desalination and Wastewater Reclamation in Water Resource Planning." Desalination, 19, 587-594.

Buros, O. K., 1978. "Economics of Desalting." In AWWA 1978 Annual Conference Proceedings. June 25-30, 1978, Atlantic City.

Buros, O. K., 1979a. "Cowpet Bay Goes SWRO." Pure Water, 8, 2.

Buros, O. K., 1979b. "Economic Aspects of Membrane Processes." Desalination, 30, 595-603.

Buros, O. K., 1980. "The Role of Desalination in Water Reuse." Desalination, 32.

Burt, J. C., 1956. "Iceberg Water for California." Science Digest, 39, 1.

Burton, S. J., 1978. "Legal/Political Aspects of Antarctic Iceberg Utilization." In Iceberg Utilization. A. A. Husseiny, ed. Pergamon Press, New York.

Cadwallader, E. A., Westberg, J. E., and Williamson, W. R., 1977. The Application of Wind Energy Systems to Desalination. OWRT Report. NTIS No. PB-276 174.

Cameron, R. L., 1978. "Glaciers and Icebergs." In Iceberg Utilization. A. A. Husseiny, ed. Pergamon Press, New York.

Canadian Hunger Foundation and Brace Research Institute, 1979. A Handbook on Appropriate Technology. CHF & BRI, Ottawa.

Carrier Corporation, July 1964. Development of the Direct Freeze Separation Process. OSW R&D Progress Report No. 113.

Catalytic, Inc., 1979. Desalting Handbook for Planners. Prepared for OWRT.

Chalchal, S., 1979. "Operating Experience of a Large Multistage Flash Evaporator in Sultanate of Oman." In Proceedings of the International Congress on Desalination and Water Reuse. October 21-27, 1979, Nice, France.

Chan, Y. K., and Leung, K. K., 1979. "Chemical Removal of Calcium Sulphate Scale--Plant Experience of Lok On Pai Desalter." In Proceedings of the International Congress on Desalination and Water Reuse. October 21-27, 1979, Nice, France.

Christensen, V., 1976. "Concerning the Possibility of Procuring a Water Supply for St. Thomas, V.I." Translated from Danish by A. F. Johnson. In Report on Water Supply for St. Thomas. Bureau of Reclamation, Santa Fe, New Mexico.

Colt Industries, 1969. Vacuum-Freezing Vapor--Compression Process. OSW Report No. 451.

Colt Industries, 1972. "An Economic Study of the Vacuum-Freezing Ejector-Absorption Desalting Process." OSW R&D Report No. 833.

Colt Industries. *Vacuum-Freezing Vapor-Compression Process: 60,000-gpd Pilot Plant Evaluation*. OSW R&D Report No. 573.

Colt Industries. *Vacuum-Freezing Vapor-Compression Desalting Process*. OSW R&D Progress Report No 296.

Concentration Specialists, Inc., 1977. Personal communication to R. B. Cox.

Cox, R. B., 1969. "Sea Desalination." *Science and Technology*, April 1969.

Cox, R. B., 1974. "The Economics of Desalting Plants." Presented at Fairleigh Dickinson University seminar "Desalination Operations--1974." December 9-13, 1974, St. Croix U.S. Virgin Islands.

Cox, R. B., 1979a. "Energy Costs of Various Desalting Processes," *Pure Water*, *8*, 4.

Cox, R. B., 1979b. "Freezing Techology--An Update." Presented at Fairleigh Dickinson University Seminar "Desalination Operations--1979." December 9-14, 1979, St. Croix, U.S. Virgin Islands, 1979.

CSMCRI, 1980. Series of letters from the Central Salt and Marine Chemicals Research Institute (Bhavnagar, India) to O. K. Buros from February 2 through March 19, 1980, concerning activities of the Institute.

Curran, H. M., Dykstra, D., Kenkeremath, D., 1976. *State-of-the-Art of Membrane and Ion Exchange Desalting Processes*. Prepared for OWRT by Hittman Associates, Inc., Columbia, Maryland.

Daniels, F. 1964. *Direct Use of the Sun's Energy*. Ballantine, New York.

Davis, R. B., Burchesky, R. D., and Coplan, M. J., 1977. "Hollow Fiber Composite Membranes for Reverse Osmosis." *Proceedings of the International Congress on Desalination and Water Reuse*. November 27-December 3, 1977, Tokyo.

Davis, T. A., 1978. "Osmotic Propulsion of Icebergs." In *Iceberg Utilization*. A. A. Husseiny, ed. Pergamon Press, New York.

DeMaio, A. et al., 1979. "An Advanced Method for Sea Water Chemical Treatment in MSFD Plants." In *Proceedings of the International Congress on Desalination and Water Reuse*. October 21-27, 1979, Nice, France.

DIGAASES, 1979. La Desalacion en Mexico.

DOE, 1980. "Cheaper Solar Cells Aim of $9 Million Development Project." Energy Insider, 3 14.

DOE/EIA, 1976. Energy Data Reports: Steam-Electric Plant Construction Cost and Annual Production Expenses, 1975. 28th Annual Supplement, DOE/EIA-0033/1.

Dravo Corporation, 1973. "10,000,000-Gallon-Per-Day Secondary Refrigerant Freeze Desalting Plant." Prepared for OWRT. NTIS Publication PB-251 906.

Drost-Hansen, W., 1966. "The Puzzle of Water." International Science and Technology, October 1966.

Dunham, D. C., 1978. Fresh Water from the Sun. U.S. A.I.D., Washington, D.C.

DuPont Corporation, 1976. Permasep Permeators for Seawater Desalting.

Eibling, J. A., Talbert, S. G., and Löf, G.O.G., 1969. "Solar Stills for Community Use-Digest of Technology," Solar Energy, 13, pp 263-276.

Eibling, J. A., 1980. Personal Communication to O. K. Buros, July 2, 1980.

El-Nashar, A. M., 1976. "Desalting of Agricultural Drainage Wastewater Using Reverse Osmosis." Presented at the Conference "Egypt in the Year 2000." December 1976, Cairo.

El-Ramly, N. A., and Congdon, C.F., 1977. Desalting Plants Inventory--Report No. 6. U.S. Department of Interior, Washington, D.C.

Fluor Engineers, Inc. 1976. Economics of a Freezing Desalting Process Using Cold Seawater Effluent of a Liquid Natural Gas Plant. OWRT Report No. 52. NTIS No. PB-259 272.

Fluor Engineers, Inc., 1978. Desalting Plans and Progress, An Evaluation of the State-of-the-Art and Future Research and Development Requirements (2nd Edition). NTIS No. PB290 75. U.S. Department of Interior, Washington, D.C.

Friedlander, H. Z., and Rickles, R. N., 1966. "Membrane Separation Processes." International Chemical Engineering, February 28, 1966.

Gadelius K. K., 1977. "New Desalination Technology Utilizing Low-Temperature Energy." Translation of an article in Bulletin of the Society of Seawater Science (Japan), 30, No. 5 and 6.

Gasparini, Giuffrida, Zanelli, Migliorini, and Franzosi, 1979. "Technical and Economical Valuation on the Coupling of a Low Temperature Multiple Effect (LTME) Plant to Solar Collector Systems." In *Proceedings of the International Congress on Desalination and Water Reuse*. October 21-27, 1979, Nice, France.

Geishecker, E. P., 1977. "A Packaged Desalting Plant." In *The Technical Proceedings of the Fifth Annual Conference of the National Water Supply Improvement Association*. July 17-21, 1977. San Diego, California.

Ghulaigah, A., and Ericsson, B., 1979. "Riyadh's Reverse Osmosis Water Treatment Plants--The Largest Demineralization Complex in the World." In *Proceedings of the International Congress on Desalination and Water Reuse*. October 21-27, 1979, Nice, France.

Gilbert, P. T., 1975. "Selection of Materials for Heat Exchangers." Presented at the Sixth International Congress on Metallic Corrosion. December 1975, Sydney, Australia.

Glueckstern, P., 1979. "Comparative Economics of Current and Advanced Desalting Systems." In *Proceedings of the International Congress on Desalination and Water Reuse*. November 27-December 3, 1977, Tokyo.

Golding, E. W., 1955. *The Generation of Electricity by Wind Power*. (Reprinted in 1977) E&F.N. Spon Ltd., London.

Goldman, C. R., 1978. "Ecological Aspects of Iceberg Transport from Antarctic Waters." In *Iceberg Utilization*. A. A. Husseiny, ed. Pergamon Press, New York.

Goldstein, A. L., 1979. "Electrodialysis on the American Continent. In *Proceedings of the International Congress on Desalination and Water Reuse*, October 21-27, 1979, Nice, France.

Gomez, E., "Ten Years Operating Experience at 7.5 mgd Commercial Desalination Plant--Rosarito, B. CFA, Mexico." In *Proceedings of the International Congress on Desalination and Water Reuse*. November 27-December 3, 1977, Tokyo.

Green, J. A. S., Gamson, B. W., and Westerbaan, W.F., 1977. "Experience with Desalination Plants Containing Titanium Heat Exchange Tubing." In *Proceedings of the International Congress on Desalination and Water Reuse*. November 27-December 3, 1977, Tokyo.

Hammond, R. P., 1978. "The Application of Advanced Technology to Iceberg Utilization." In *Iceberg Utilization*. A. A. Husseiny, ed. Pergamon Press, New York.

Harashina, H., 1977. "5,000,000 Imp. Gal/Day Sea Water Desalination Plant for the Ministry of Electricity and Water, Government of Kuwait." In *Proceedings of the International Congress on Desalination and Water Reuse*. November 27-December 3, 1977, Tokyo.

Hickman, C. E., Jamjoom, I., Riedinger, A. B., and Seaton, R., 1979. "Jeddah Seawater Reverse Osmosis Installation." In *Proceedings of the International Congress on Desalination and Water Reuse*. October 21-27, 1979, Nice, France.

Hoffman, D., 1977. "Second Generation Low Temperature Vapor Compression Plants." In *Proceedings of the International Congress on Desalination and Water Reuse*. November 27-December 3, 1977, Tokyo.

Holden, G., 1980. Personal communication to O. K. Buros.

Howe, E. D., 1974. *Fundamentals of Water Desalination*. Dekker, N.Y.

Howe, E. D., 1980. Personal communication to H. A. LeSieur (U.S. A.I.D.) July 23, 1980.

Howe, E. D. and Tleimat, B. W., 1974. Twenty Years of Work on Solar Distillation at the University of California. *Solar Energy*, *16*.

Johnson W., 1976. *State-of-the-Art Survey and Economic Comparison of Freezing Processes*. Prepared for OWRT.

Johnson, W., 1979. "Indirect Freezing Process." *Desalination*, *31*, p. 417.

Juda, W., and MacRae, W. A., 1953. U.S. Patents No. 2,636,851 and 2,636,852.

Justus, C. G., 1978. *Winds and Wind System Performance*. Franklin Inst. Press, Philadelphia, Pennsylvania.

Kaiser Engineers, 1967. *A Manual on Water Desalination*. Prepared for U.S. A.I.D., Washington, D.C.

Kaiser Engineers, 1966. *Engineering Feasibility and Economic Study for Dual-Purpose Electric-Power Water Desalting Plant for Israel*.

Kanal, I., and Hughes, C. H., 1979. "The VTE Boiler: An Advanced Thermal System for the Desalination of Sea Water." Presented at the International Congress on Desalination and Water Reuse. October 21-27, 1979, Nice, France.

Katz, W. E., 1977. "Electrodialysis--the First Twenty-Five Years." In *The Technical Proceedings of the Fifth Annual Conference of the National Water Supply Improvement Association*. July 17-21, 1977, San Diego, California.

Katz, W. E., 1971. "Electrodialysis for Low TDS Waters" *Industrial Water Engineering*, June/July 1971.

Kawahara, T., Katsutoshi, A., and Tomoyasu, I., 1977. "The Development of a New Electrodialyzer for Desalination." In Proceedings of the International Congress on Desalination and Water Reuse. November 27-December 3, 1977, Tokyo.

Kays, D. D., 1965. "Desalting Seawater in the Multiple-Effect Long-Tube Vertical Evaporator." Paper WD/39, First International Symposium on Water Desalination. October 3-9, 1965, Washington, D.C.

Kellogg Company, 1975. Saline Water Conversion Engineering Data Book--1975. Prepared for OWRT, NTIS PB-250 907.

Kirchner, O. E., 1977. "Vapor Compression Evaporation for Inland Desalting." In Proceedings of the International Congress on Desalination and Water Reuse. November 27-December 3, 1977, Tokyo.

Kishi, M., Serizawa, S., and Nakano, M., 1977. "New Electrodialyzer System with Automatic Chemical Cleaning." In Proceedings of the International Congress and Water Reuse. November 27-December 3, 1977, Tokyo.

Klaxen, D. G., and Pieper, G. A., 1978. "The Multi-Stage-Flash/ Fluidized Bed Evaporator in Combination With Solar Energy as a Heating Source." In Proceedings of the Sixth International Symposium on Freshwater from the Sea, Las Palmas, Spain, 1978.

Korte, F., 1979. "Ecotoxicology of Belgard EV." In Proceedings of the International Congress on Desalination and Water Reuse. October 21-17, 1979, Nice, France.

Larson & Associates, Ltd., 1979. Desalting Seawater and Brackish Water: Cost Update, 1979. Report No. ORNL/TM-6912. Prepared for U.S. Department of Energy, Oak Ridge, Tennessee.

Lawand, T. A., 1967. The Economics of Wind Powered Desalination Systems. Brace Research Inst. Tech. Report No. T-36, Quebec, Canada.

Lawand, T. A., 1968. Historical Review of Solar Distillation. Brace Research Inst. Tech. Report No. T-48, Quebec, Canada.

Lawand, T. A., 1970. Experiences With Solar Stills Used By Developed and Primitive Societies in the West Indies. Brace Research Inst. Pub. No. R-53, Quebec, Canada.

Lawand, T. A., 1975. Systems for Solar Distillation. Brace Research Inst. Tech. Report No. R-115, Quebec, Canada.

Leimgruber, R., 1979. "Toxicological Aspects of Scale-Preventative Agents Used in Production of Drinking Water by Desalination." In Proceedings of the International Congress on Desalination and Water Reuse. October 21-27, 1979, Nice, France.

Libert, J. J., Maurel, A., and Lucas, R., 1979. "Dessalement et Environnement". In *Proceedings of the International Congress on Desalination and Water Reuse*. October 21-27, Nice, France.

Loeb, S., and Sourirajan, S., 1961. *Report No. 60-60*. University of California at Los Angeles.

Löf, G.O.G., Duffie, J., Smith, C., 1966. *World Distribution of Solar Energy*. University of Wisconsin, Madison.

Lonsdale H. K., and Podall, H. E., (eds.), 1972. *Reverse Osmosis Membrane Research*, Plenum Press, New York.

Lundstrom, J. E., 1979. "Water Desalting by Solar Powered Electrodialysis."

Manjarrez, R., and Gonzalez de Cosio, I., 1976. "Desalacion de Aguas de Drenes Agricolas en Mexicali, B.C." In *Proceedings of the First Desalination Congress of the American Continent*. October 24-29, 1976, Mexico City.

Manjarrez, R. and Galvan, M., 1979. "Solar Multistage Flash Evaporation (SMSF) as a Solar Energy Application on Desalination Processes--Description of One Demonstration Plant." In *Proceedings of the International Congress on Desalination and Water Reuse*. October 21-27, Nice, France.

Mason, E. A, and Kirkham, T. A., 1959. "Design of Electrodialysis Equipment." In Chemical Engineering Progress Symposium Series: *Adsorption*, *Dialysis*, *and Ion Exchange*, *55*, 24, 173-189.

Mattson, M. E., 1979. "Significant Developments in Membrane Desalination--1979." *Desalination*, *28*, 207-223.

Mattson, M. E., and Lundstrom, J. E., 1979. "New Developments in Brackish Water Desalting by Electrodialysis." In *Technical Proceedings of the 7th Annual Conference of NWSIA*. September 16-20, 1979, New Orleans, Louisiana.

Maurel, A., 1979a. "Dessalement et Energies Nouvelles." In *Proceedings of the International Congress on Desalination and Water Reuse*, October 21-27, 1979, Nice, France.

Maurel, A., 1979b. Communication to O. K. Buros, December 6, 1979.

McCarthy, J. V., and Leigh, J. M., 1979. *The Prospects for Desalination Using Solar Energy*. BHP Melbourne Research Laboratories, Australia.

McCue, D. M., 1975. "Titanium Tubes Cut Costs Even in Average Water." *Electrical World*, February 15, 1975.

McCutchan, J. W., Brice, D. B., and Yamamoto, K., 1976. "Reclamation of Irrigation Field Drainage Water." In Proceedings of the First Desalination Congress of the American Continent. October 24-29, 1976, Mexico City.

McRae, W., Parsi, E., and Ganzi, G., 1977. "Fresh Water From the Sea; A 380 Cubic Meter Per Day High Temperature ED Plant." In Proceedings of the International Congress on Desalination and Water Reuse. November 27-December 3, 1977, Tokyo.

Mehta, D. J., Pandya, V.P., and Rao, A. V., 1976. Reverse Osmosis Relevant to India. In Proceedings of the First Desalination Congress of the American Continent. October 24-29, 1976, Mexico City.

Mehta, D. J., Rao, A. V. and Govindan, K. P., 1979. "Water Desalination--A Perspective for Membrane Processes in India." In Proceedings of the International Congress on Desalination and Water Reuse. October 21-27, 1979, Nice, France.

Merten, U. (ed.), 1966. Desalination by Reserve Osmosis MIT Press, Cambridge, Massachusetts.

Miwa, T., Jordan, D. R., and McIlhenny, W. F., 1976. "Seawater Desalting by Electrodialysis." In Proceedings of the First Desalination Congress of the American Continent. October 24-29, 1976, Mexico City.

Morin O. J., and Johnson, R. A., 1979. "MSF Evaporator Fabricated from Reinforced Thermosetting Materials." In Proceedings of the International Congress on Desalination and Water Reuse. October 21-27, Nice, France.

Morin, O. J., and Latour, S. R., 1980. "Cost Comparison on Reverse Osmosis and Multi-Stage Flash Seawater Desalination." NWSIA Journal, 6, 1.

Mulford, S., Glater, J., and McCutchan, J. W., 1965. "An Appraisal of Scale Control Methods at the OSW Flash Evaporator Plant in San Deigo, California." Paper SWD/11, First International Symposium on Water Desalination. October 3-9, 1965, Washington, D.C.

Mustacchi, C., and Cena, V., 1978. "Solar Water Distillation." In Technology for Solar Energy Utilization. United Nations.

NAS, 1976. Energy for Rural Development. National Academy of Sciences, Washington, D.C.

Natu, G. L., Goghari, H. D., and Gomkale, S. D., 1979a. "Solar Distillation Plant at Awania Gujurat, India." In Proceedings of the International Congress on Desalination and Water Reuse. October 21-27, 1979, Nice, France.

Natu, G. L., Goghari, H. D., and Gomkale, S. D., 1979b. "Problems In Construction and Maintenance of Large Solar Stills Plants." In Proceedings of the National Solar Energy Convention, 1979.

Nebbia, G. and Menozzi, G., 1966. "A Short History of Water Desalination." In Acque Dolce Dal Mare, II^{a} Inchiesta Internazionale. Milan, Italy, April, 1966

Newton, E. H., Birkett, J. D., and Ketteringham, J. M., 1972. Survey of Material Behavior in Large Desalting Plants Around the World. OSW R&D Report No. 512 by A. D. Little, Inc., Cambridge, Massachusetts.

Nishimoto, W., 1977. "RF-FF VTE Evaporator." In Proceedings of the International Congress on Desalination and Water Reuse. November 27-December 3, 1977, Tokyo.

Nojiri, Y. and Fujii, K., 1976. "Utilization of Cement as an Evaporator Shell Material." In Proceedings of the First Desalination Congress of the American Continent. October 24-29, 1979, Mexico City.

OWRT, 1976. Desalting Handbook for Planners (1st Edition). May 1972. NTIS No. 253 755 OWRT Report 76/47.

OWRT, 1977. The A-B-C of Desalting. U.S. Government Printing Office. Washington, D.C.

Parsi, E. J., 1978. "Electrodialysis for Water Desalting--A Progress Report." Pure Water, 7, 3.

Parsons Company, 1967. Guidelines for Uniform Presentation of Desalting Cost Estimates. OSW R&D Report No. 264. U.S. Department of Interior, Washington, D.C.

Patel, K. G., Harris, F. L., and Channagassappa, K. C., 1976. "Operational Experience of Reverse Osmosis Plants of Wrightsville Beach, North Carolina, USA." In Proceedings of the First Desalination Congress of the American Continent. October 24-29, 1976, Mexico City.

Permutit Company, 1978. Increased Product Water Recovery by Reverse Osmosis Using Interstage Ion Exchange Softening with a Spiractor. OWRT Report No. 78/10.

Petersen G., Fries, S., Mohn, J., and Müller, A., 1979. "Wind and Solar-Powered Reverse Osmosis Units--Description of Two Demonstration Projects." In Proceedings of the International Conference on Desalination and Water Reuse. October 21-27, 1979, Nice, France.

Pleass, C. M., 1978. "The Use of Wave Powered Systems for Desalination: A New Opportunity." In Proceedings of the International Symposium on Wave and Tidal Energy, September 27-29, 1978, Canterbury, England.

Pleass, C. M., 1979. "Construction and Deployment of Delbouy, a Prototype 1,500-Gallon-Per-Day Sea-Wave-Powered Desalination Unit." A Sea Grant Project Summary for Project No. R/T-50, University of Delaware.

Rajvanshi, A. K., 1979. *Analytical and Experimental Investigation of the Effect of Dyes on Solar Distillation*. Ph.D. Dissertation, University of Florida, Gainesville.

Rao, A. V., 1980. Letter from Rao (of the Central Salt and Marine Chemicals Research Institute, Bhavnagar, India) to O. K. Buros, dated February 2, 1980.

Reed, S. A., and Wilson, J. V., 1977. *Desalting Sea Water and Brackish Waters: A Cost Update*. Report No. ORNL/TM-5926. Oak Ridge, Tennessee.

Reed, S. A., 1975. *The Impact of Increased Fuel Costs and Inflation on the Cost of Desalting Sea Water and Brackish Waters*. Report No. ORNL/TM-5070. Oak Ridge, Tennessee.

Richardson 1978. *The Richardson System--Process Plant Construction Estimating Standards*. 1978-1979 Edition.

Robertson R. S., 1978. *Developmental Testing of a Secondary Refrigerant Freezing Desalination Pilot Plant at the Wrightsville Beach Test Facility*. Report prepared for OWRT.

Rodriguez, H., 1980. Letter to O. K. Buros dated March 13, 1980 (Mexico City).

Rohm and Haas Company, 1979. "Industrial Waste Treatment with Butane Secondary Refrigerant Freezing," by R. D. Myers and R. W. Carter.

Romeijn, A. A., and Eimer, K., 1978. "Foam Rubber Ball Cleaning in MSF Evaporators in Terneuzen." In *Proceedings of the Sixth International Symposium on Freshwater from the Sea*. September 17-28, 1978, Las Palmas, Spain.

Ross, D. 1979. *Energy From the Waves*. Pergamon Press, Oxford.

Sackinger C. T., 1980. "The Energy Rquirements of Seawater Desalination: Reverse Osmosis vs. Multi-Stage Flash." *MEW&S*, April/May, 1980.

Satone, H., Funahashi, S., and Inque, S., 1977. "Experiences on VTE Distillation Plants in Japan." In *Proceedings of the International Congress on Desalination and Water Reuse*. November 27-December 3, 1977, Tokyo.

Saunders, R. J., and Warford, J. J., 1976. *Village Water Supply*. Johns Hopkins Univ. Press, Baltimore, Maryland.

Seto, T., Kawate, H., and Komori, R., 1977. "Development of Ion Exchange Membrane Technology." In Proceedings of the International Congress on Desalination and Water Reuse. November 27-December 3, 1977, Tokyo.

Shannon, E. 1979. "Design and Start-Up of the 3.0-mgd Reverse Osmosis Plant at Cape Coral, Florida." Presented at the OWRT Technology Transfer Workshop. July 19-20, 1979, Virginia Beach, Virginia.

Sheppard, G. R., 1978. Fresh Water from the Sea and What it Means to Me. GRS, Box 905, St. Croix, Virgin Islands 00820.

Silver, R. S., 1971. "Multi-Stage Flash Distillation--The First 10 Years." Desalination 9, 3-17.

Simmons, D. M., 1975. Wind Power. Noyes Data Corporation, Park Ridge, New Jersey.

Smith, M. C., 1980. Water Desalination Report, XVI, 8.

Sourirajan, S., 1980. "Reverse Osmosis--A New Field of Applied Chemistry and Chemical Engineering." Presented at ACS Symposium on Synthetic Membranes, and Their Applications. August 25, 1980, San Francisco, California.

Sourirajan, S. (ed.), 1977. Reverse Osmosis and Synthetic Membranes. National Research Council, Canada.

Standiford, F. C. and Sinek, J. R., 1961. "Stop Scale in Sea-water Evaporators." Chemical Engineering Progress, 57, 58.

Stewart, J., Bom, R., and Finan, M. A., 1979. "The Use of Acid and Additive--Practical Experiences in a Multistage Flash Plant." In Proceedings of the International Conference on Desalination and Water Reuse. October 21-27, 1979, Nice, France.

Sverdrup, H. U., Johnson, M. W., and Fleming, R. H., 1946. The Oceans. Prentice Hall, New York.

Talbert, S. G., Eibling, J. A., and Löf, G. O., 1970. Manual on Solar Distillation of Saline Water. OSW (NTIS PB 201 029), Washington, D.C.

Tazawa, Y., Nojiri, Y., and Fujii, K., 1977. "Behavior of Concrete Evaporator Shells During Operation of the Module Plant at the Chigasaki Test Facilities." In Proceedings of the International Congress on Desalination and Water Reuse. November 27-December 3, 1977, Tokyo.

Tidball, R. A. and Woodbury, R. E., 1965. "Methods of Scale Control in Flash Systems." In Proceedings of the First International Symposium on Water Desalination (Paper SWD/24). October 3-9, 1965, Washington, D.C.

Tleimat, B. W., 1969. "Novel Approach to Distillation by Vapor Compression Distillation." Paper 69-WA/PID-1, ASME, November 1969.

Tleimat B. W., 1978. "Solar Distillation: The State of the Art." In *Technology for Solar Energy Utilization*. United Nations.

Tleimat, B. W., and Howe, E. D., 1977. "Solar Assisted Distillation of Sea Water." *NWSIA Journal*, *4*, 1.

Trade Times (Tokyo) August 25, 1978.

United Nations, 1970. *Solar Distillation as a Means of Meeting Small-Scale Water Demands*. New York.

Van Hoek, C., and Mavis, J. D., 1977. "Test Results at the Yuma Desalting Test Facility." In *Proceedings of the Fifth Annual Conference of the National Water Supply Improvement Association*. July 17-21, 1977. San Diego, California.

Veenman, A. W., 1976. "The MSF/FBE: An Improved Multi-Stage Flash Distillation Process." In *Proceedings of the First Desalination Congress of the American Continent*. October 24-29, 1976, Mexico City.

Veenman, A. W., 1978. "A Review of New Developments in Desalination by Distillation Processes." *Desalination*, *27*, 21-39.

Veenman, A. W. et al., 1978. "The Fluidized Bed Heat Exchanger in an MSF/FBE--Its Self-Cleaning and De-Scaling Capabilities." Presented at the Sixth International Symposium on Freshwater from the Sea, September 17-28, 1978, Las Palmas, Spain.

Wade, N. M., 1979. "A Review of Scale Control Methods." In *Proceedings of the International Conference on Desalination and Water Reuse*. October 21-27, 1979, Nice, France.

Watson, B. M., 1976. *History, Status, and Future of the Distillation Process*. OWRT, Washington, D.C.

Watson, I., and Derowitsch, R., 1979. "Five Years of ED Experience on a Barrier Island, or '1.5 Billion Gallons Later'." In *Technical Proceedings of the 7th Annual Conference of NWSIA*. September 16-20, 1979, New Orleans, Louisiana.

Weeks, W. F., and Mellor, M., 1978. "Some Elements of Iceberg Technology." In *Iceberg Utilization*, A. A. Husseiny, ed. Pergamon Press, New York.

White, G. F., Bradley, D. J., and White, A. U., 1972. *Drawers of Water*. The University of Chicago Press, Chicago, Illinois.

Winters, H., Isquith, I., and Bakish, R., 1979. "Influence of Desalination Effluents on Marine Ecosystems." In Proceedings of the International Congress on Desalination and Water Reuse. October 21-27, 1979, Nice, France.

World Bank, 1976. "Village Water Supply." A World Bank Paper. Washington, D.C.

Yambe, T., 1977. "Present Status of Electrodialysis in Japan." In Proceedings of the International Congress on Desalination and Water Reuse. November 27-December 3, 1977, Tokyo.

Yorkshire Imperial Metals, Ltd., 1976. Technical Memorandum 2, YIM Heat Exchanger Tubes: Matching the Product to the Duty. Leeds, England.

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