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9. ABSTRACT
(Engineering--Hydraulics R&D)
Volumes I & II; Supplements to Volumes I & II
A PROJECT: Intended to guide the officers of the Agency for International Development in evaluating and determining the technical and economic feasibility of desalination projects for which U.S. Government assistance may be sought. The original two-volume Manual -- one volume on technology and one on economics -- was prepared by Kaiser Engineers of Oakland, California in April, 1967. In October, 1972, Burns and Roe Inc., of Oradell, New Jersey produced a supplement to each volume.
Duration: 1967: 1972
DEVELOPMENTS: Technical improvements and cost reductions of desalination have added a new dimension to water supply planning: This method of supplementing freshwater supplies demands prominent consideration as a means of meeting future water needs. The Manual reviews various desalination processes and presents economic and technical guidelines for gathering and analyzing data on energy and water resources and requirements -- data that must be considered in deciding whether water desalination is a viable solution to an area's water shortage problem. Volume I discusses eight desalination methods and describes a typical plant for each: Distillation processes - 1. vertical falling and rising film, 2. multistage flash, 3. multieffect multistage flash, 4. vapor compressions, and 5. solar stills; Non-distillation processes - 6. vacuum-freezing vapor-compressions, 7. electrodialysis, and 8. reverse osmosis. It includes a feasibility study on combining desalination

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plants with electric power generating plants, and a discussion on combining desalination with production of chemical by-products. Volume II gives guidelines for studies to determine if desalination is the optimum solution to an area's water shortage problem. An Appendix lists existing plants, technical data on properties of seawater and brackish water, and sample survey questionnaires. The Supplements to Volumes I & II do not present new technologies but rather refinement and commercialization of the material in the two volumes. They update technical and economic guides - often pointing to the increased viability of desalination solutions. The Supplements describe typical plants built since 1967 and provide sections on conceptual designs, water development planning, and updated budget water plant pricing. They add as well some environmental considerations, and feed treatment and brine disposal methods. Whereas the Manual provides information for evaluation and determination of the feasibility of desalination programs, it does not provide the detailed technical data required by a desalination plant designer. Volumes I & II can be used independently of each other; the Supplements, however, do require reference to Volumes I & II.

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A MANUAL ON WATER DESALINATION

VOLUME ONE—TECHNOLOGY:

CHAPTERS 5 - 7

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PREPARED FOR

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VOLUME ONE

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Notice

"This manual was prepared for the Agency for International Development (AID), to guide its officers in evaluating and determining the technical and economic feasibilities of desalination programs and projects, for which United States Government assistance may be furnished or sought. The information contained herein is not sufficient for and is not intended to be the sole basis for design or construction of a project at any particular location, or as a substitute for professional judgment, advice, and design work by competent engineers in connection with such a project."

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V. DESCRIPTION OF TYPICAL PLANTS

A. General

This section contains descriptions of "typical" examples of eight of the more prominent processes or process variations discussed in Section IV. The examples selected are specific commercial installations or Office of Saline Water (OSW) demonstration plants. In the case of reverse osmosis no commercial or demonstration plant exists and a pilot plant has been selected. In the case of direct freezing, commercial plant data are not readily available, and a pilot plant has been selected. For each plant, nontypical features have been omitted from the flowsheet, heat balance and material balance.

The plants described in this section are as follows:

1. Distillation Processes

Vertical Falling and Rising Films	-	OSW Demonstration Plant No. 1 Freeport, Texas
Multistage Flash (MSF)	-	OSW Demonstration Plant No. 2 San Diego (Pt. Loma), California
Multieffect Multistage Flash (MEMS)	-	OSW Clair Engle Plant San Diego, California
Vapor Compression	-	OSW Demonstration Plant No. 4 Roswell, New Mexico
Solar Stills	-	Symi, Dodecanese Islands, Greece

2. Freezing Process

Direct Freezing	-	Pilot Plant at Wrightsville Beach, N.C.

3. Membrane Processes

Electrodialysis	-	OSW Demonstration Plant No. 3 Webster, South Dakota
Reverse Osmosis	-	20,000- to 40,000-gpd (gallons per day) commercial unit

B. Distillation Process

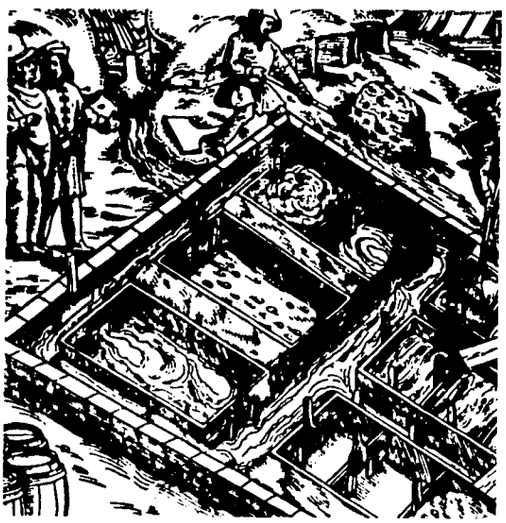
Distillation of potable water from seawater was known to the ancient Romans and for hundreds of years the distillation process has been used for obtaining salt from seawater, for drying oils and for concentrating chemicals. Until very recently, however, the distillation devices have been so crude and inefficient that drinking water could be produced only at great expense. Shipboard seawater distillation equipment was used as early as 1790 for producing potable water. Various early distillation devices are shown on Figure V-1.

More than 400 distillation untis, each with a capacity of 100,000 gpd or more, are installed throughout the world today. Many of these units supply water of very high purity for boilers and industrial processes, but the installations of largest capacity are for production of potable water for residential, municipal and industrial use. The most widely used distillation process is the submerged tube evaporator (described in Section IV). This process is now obsolete, having been replaced in popularity by the more efficient MSF evaporator process. Of the world's 20 desalination plants with capacities greater than 1 million gallons per day (mgd), 16 are of the MSF type.

With the exception of solar distillation, all of the modern distillation processes operate with a seawater or brine temperature in excess of 200 F. To achieve this operating temperature without the formation of scale, it is necessary to pretreat the seawater by decomposing the carbonates in the seawater with sulfuric acid and degasification, and by maintaining a pH low enough to prevent magnesium hydroxide scale, as described in Section IV. In each of the distillation plants described in this section (except solar distillation) the raw seawater is chlorinated to prevent marine growth, and passed through a traveling screen of approximately 1/2-inch mesh to exclude fish, other marine life and flotsam. Some of this seawater is then used for cooling but the remainder, which is to be used as makeup, is further pretreated with sulfuric acid using an amount slightly in excess of the stoichiometric requirements to decompose the carbonates. The acid-treated makeup seawater is then deaerated to remove the CO_2 resulting from the decomposition of the carbonates and to remove the noncondensables, particularly oxygen. The removal of the noncondensables prevents them from accumulating near the heat-transfer surfaces and thus degrading performance. Removal of the

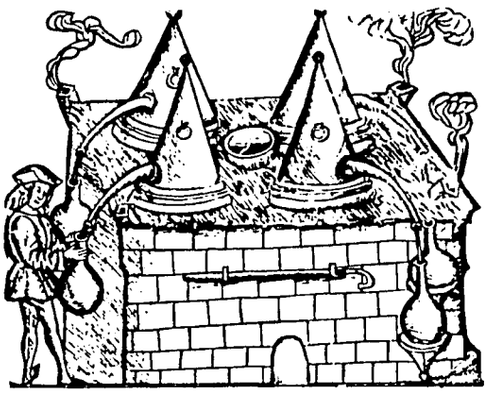
FIGURE V-1

EARLY DISTILLATION DEVICES



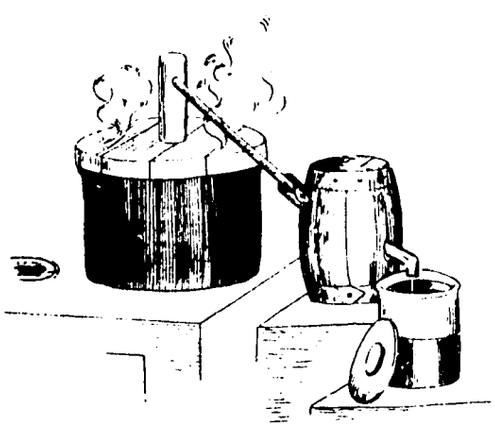
Solar Evaporation of Seawater (to obtain salt) as Practiced in the 16th Century

Courtesy: *See below



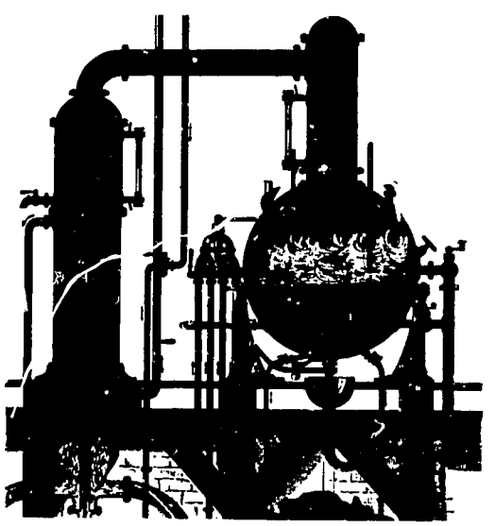
Mass distillation circa 1500

Courtesy: Cuno Engineering Corp



Marine distillation circa 1790

Courtesy: Cuno Engineering Corp



Process Vacuum Evaporation Late 19th Century

Courtesy: *See below

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oxygen is particularly important as its presence accelerates the corrosion of carbon steel - a material frequently used for evaporator shells and piping. Deaeration can reduce the oxygen content in the seawater makeup to as little as 5 parts per billion. The following typical distillation plants (except solar distillation) use seawater pretreated as described above. Brackish water feed, however, may require special pretreatment.

1. Vertical Falling and Rising Films

Relatively few falling and rising film distillation units have been built, though a few units have been installed in England and Germany and some small size units are installed on ships. Recently, several companies have offered the long-tube-vertical (LTV) design as a competitor to the MSF design. The plant chosen as representative of the application of the vertical falling and rising film distillation process is the OSW Demonstration Plant No. 1, located at Freeport, Texas. This plant employs a 12-effect falling film distillation process of LTV configuration, and has the following characteristics:

Capacity	1,000,000 gpd
Performance ratio	Approximately 10 lb product/ 1,000 Btu
Source of seawater	Gulf of Mexico via a seawater canal
Source of heat	Purchased steam from Dow Chemical Co

The startup of the Freeport Plant took place on July 1, 1961. By June 30, 1965 the Freeport Plant had produced over 1 billion gallons of fresh water. The plant produced 234,000,000 gallons of water from July 1, 1964 to June 30, 1965, and had an on-stream

factor of 85% during this period, disregarding shutdowns for strictly experimental, developmental and modification work. The construction cost of the Freeport Plant was $1,256,000, and the actual water production cost for a normal operating period in this plant in 1965 was approximately $1.05 per thousand gallons, as reported in the 1965 OSW Saline Water Conversion Report.

a. Plant Description

The OSW Demonstration Plant located at Freeport, Texas, is situated near the Dow Chemical plant. The demonstration plant itself is a 12-effect vertical falling film process. Twelve-effect evaporator vessels are arranged in two rows and are serviced by a steel support structure containing two platforms above the ground level. The administration building contains the office, control room and a small chemical laboratory, a separate small building adjacent to the plant contains the electrical switchgear. The plant receives its seawater from a seawater canal, which also serves the Dow Chemical plant, it receives its steam from the nearby Dow Chemical plant facility, it supplies its product water to the City of Freeport, and it returns the concentrated brine to the seawater canal downstream of the demonstration plant inlet. Photographs of the Freeport Demonstration Plant facility are shown in Figures V-2 and V-3.

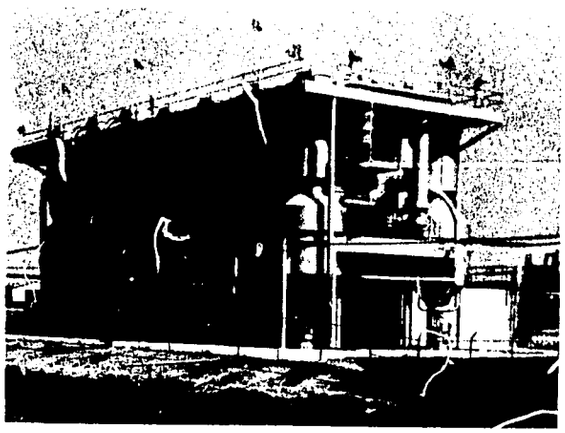
b. Process Description

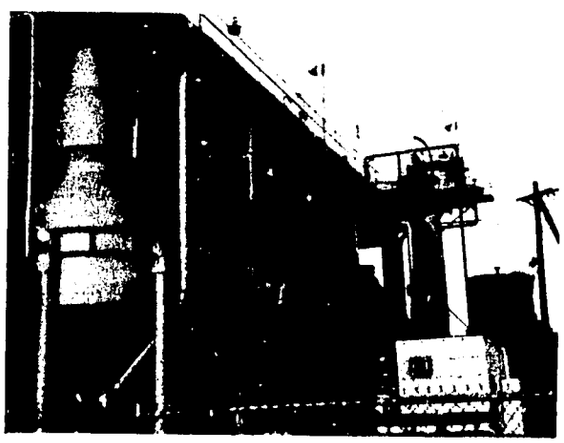
A simplified process flow diagram of the Freeport Demonstration Plant is shown on Figure V-4. The seawater feed is slightly preheated in the heat-recovery heat exchanger of Effect No. 12 (lowest - temperature effect), and then pretreated for scale and corrosion prevention, using the acidification-deaeration method. The treated seawater feed is heated successively in the heat-recovery heat exchangers of Effects No. 11 to No. 1, and enters Effect No. 1 on the tube side above the vertical tube-bundle.

Each of the 12 evaporator effects consists of a vessel housing a vertical tube-bundle with 2 inch OD, 24 foot long tubes, a brine feed and distribution chamber above the tube-bundle, and a vapor and liquid separation chamber below the tube-bundle. The brine or seawater feed enters the effect via the distribution

FIGURE V-2

OSW DEMONSTRATION PLANT NO. 1
FREEPORT, TEXAS, LTV PROCESS





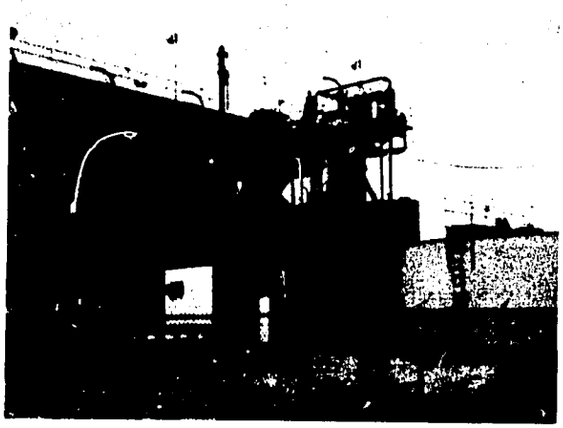
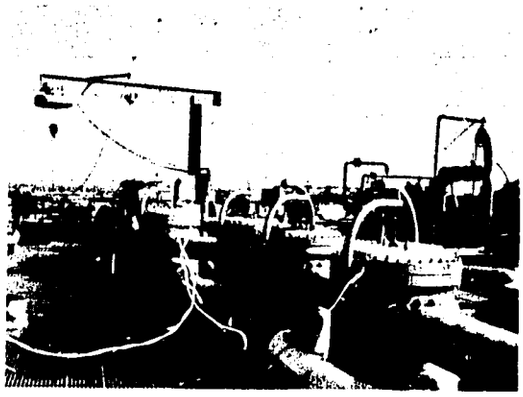
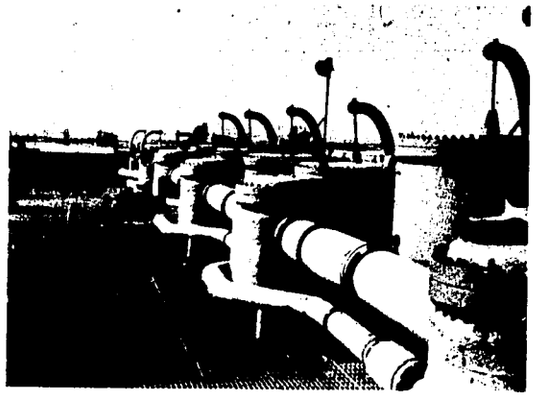
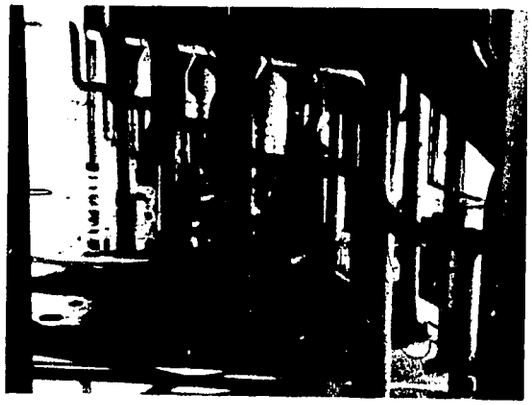


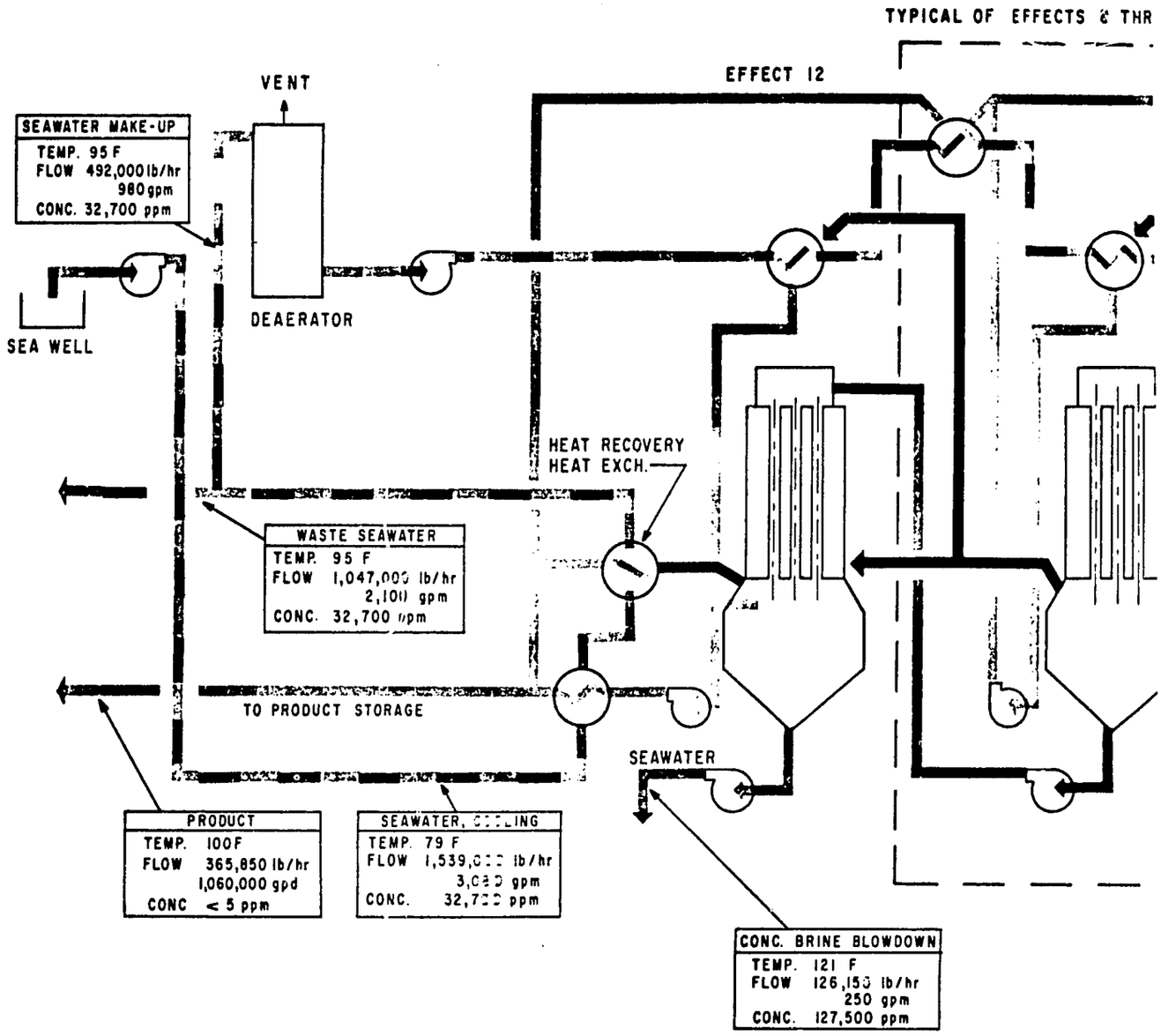
FIGURE V-3

OSW DEMONSTRATION PLANT NO. 1
VIEWS OF TOP DECK AND GROUND LEVEL








TYPICAL OF EFFECTS 8 THR
EFFECT 12
VENT
SEAWATER MAKE-UP
TEMP. 95 F
FLOW 492,000 lb/hr
980 gpm
CONC. 32,700 ppm
DEAERATOR
SEA WELL
HEAT RECOVERY
HEAT EXCH.
WASTE SEAWATER
TEMP. 95 F
FLOW 1,047,000 lb/hr
2,101 gpm
CONC. 32,700 ppm
TO PRODUCT STORAGE
SEAWATER
PRODUCT
TEMP. 100F
FLOW 365,850 lb/hr
1,060,000 gpd
CONC < 5 ppm
SEAWATER, COOLING
TEMP. 79 F
FLOW 1,539,000 lb/hr
3,080 gpm
CONC. 32,700 ppm
CONC. BRINE BLOWDOWN
TEMP. 121 F
FLOW 126,150 lb/hr
250 gpm
CONC. 127,500 ppm

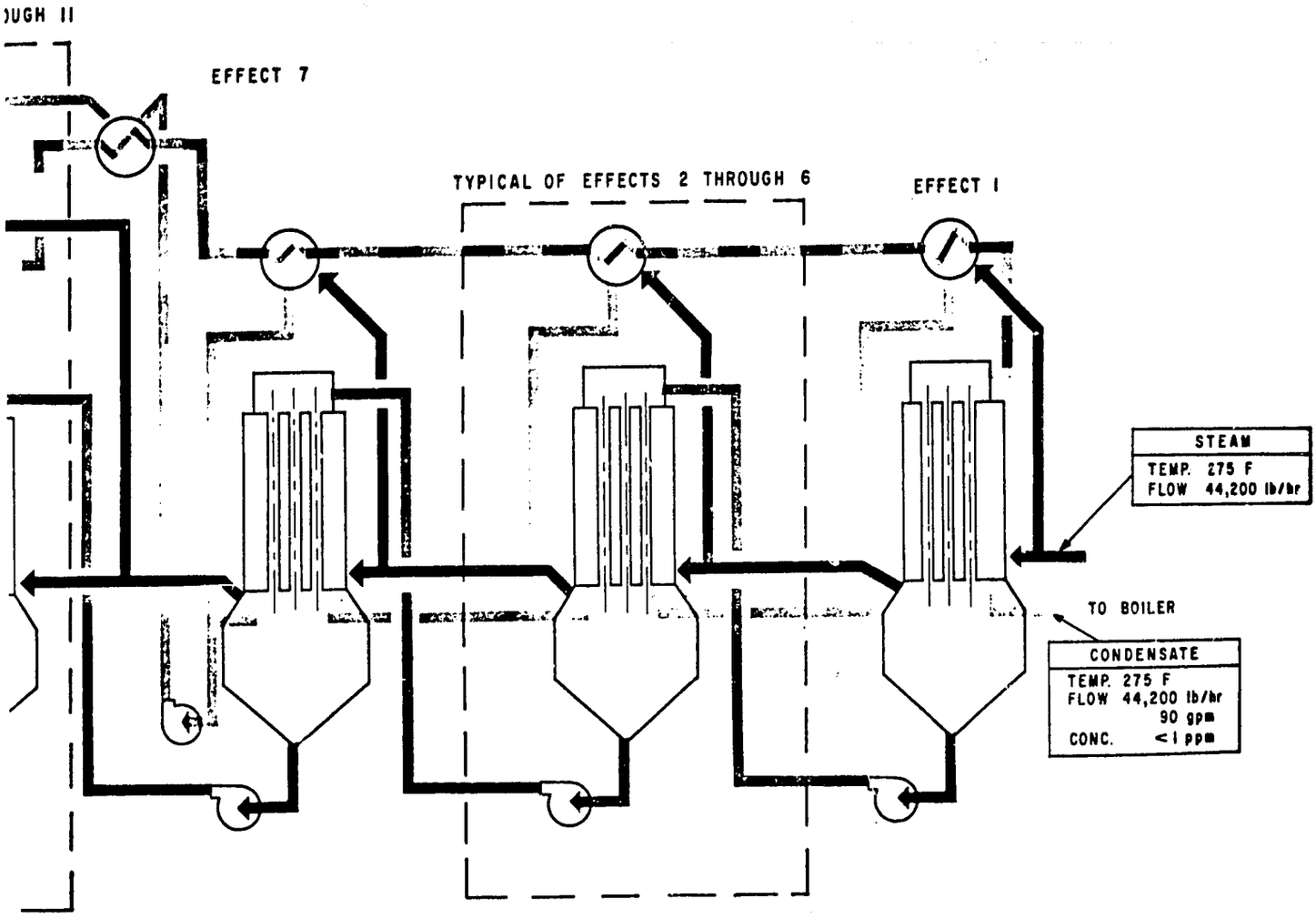


FIGURE V-4

VERTICAL FALLING FILM PROCESS

OSW DEMONSTRATION PLANT NO.1

FREEPORT, TEXAS

chamber above the vertical tube-bundle, and is distributed to the tubes. The liquid forms a thin-film layer on the inside of the tubes as it flows downward. Steam condensing on the shell side of these tubes evaporates a portion of the water in the falling film. The water and vapor pass into the bottom chamber and are separated. The liquid forms the brine feed to the next lower temperature effect; the major portion of the vapor flows to the shell side of the next lower temperature effect where it condenses as product water. A portion of this vapor also flows to a heat-recovery heat exchanger, where it heats the incoming seawater feed. The steam entering the shell side of Effect No. 1 is generated in a separate plant, it is the heat supply to the process. The temperature level in the first effect is approximately 265 F and it drops in each successive effect by approximately 12 F per effect.

A simplified heat and material balance for the process also is given on Figure V-4. The primary heat supply to the system is the steam supply to the shell side of Effect No. 1. The points of heat loss, other than those in the equipment and piping, are in the brine blowdown stream, in the product water stream and in the waste seawater used to condense the vapor from Effect No. 12. The process-supply streams are the seawater feed to the process, the seawater used for condensing vapor generated in Effect No. 12 and the steam supply to Effect No. 1. The discharge streams are the condensate return from Effect No. 1, the product water, the blowdown brine and the waste seawater used for condensing the vapor from Effect No. 12.

c. Special Features

As the first of the OSW demonstration plants, the Freeport facility incorporated a number of features for test and development purposes; e.g. the tubes in Effects No. 1, No. 5, No. 7, No. 8 and No. 9 were steel. Provisions were made for several types of pretreatment on incoming seawater; these included a seeding technique (Subsection IV.A.3.b), but this was abandoned after a very brief trial.

d. Problems Encountered

The corrosion of the steel tubes was excessive and they had to be replaced with aluminum-brass tubes. Also, modifications were made to increase the capacities of several pumps and to increase the heat-transfer capacity of Effect No. 12. The sulfuric acid seawater pretreatment system control was improved, thereby improving the stability of the pH of the makeup feed; the deaerator controls were also improved. Corrosion and scale, particularly magnesium hydroxide and calcium carbonates, were major problems initially, but the pretreatment system improvements have resulted in essentially scale-free operation. Furthermore, the capacity of the plant has been increased to about 1.2 mgd and can potentially be raised to 1.6 mgd without major equipment changes.

e. Present Use

The Freeport plant is presently used as a development and test facility to investigate improvements in the vertical falling and rising film process. Scale-free operation has been demonstrated at a first-effect brine outlet temperature of 265 F. A blowdown concentration ratio of as much as 3:1 can be achieved.

An additional small facility is in operation for conducting corrosion studies.

f. Plant Staff

The plant staff consists of approximately 13 personnel for continuous around-the-clock plant operation. These personnel are:

Plant Superintendent

Plant Engineer

Operations Supervisor and Chemist

Accountant

Plant Operators - 4 (one on each shift)

Day-shift operating and maintenance personnel - 5 - consisting of:

1 Instrumentation Operator
1 Welder
2 Maintenance Operators
1 Utility Man

The plant is sufficiently automated so that only one Plant Operator per shift is required.

For further discussion of the LTV process see Subsection IV.A.2.a.(2).

Additional information is contained in the following publications:

OSW Report No. 71 — First Annual Report Saline Water Demonstration Plant No. 1, Freeport, Texas, Order No. PB 181466

OSW Report No. 100 — Second Annual Report Saline Water Conversion Demonstration Plant, Freeport, Texas Stearns-Roger Manufacturing Co., Order No. PB 181681

OSW Report No. 123 — Third Annual Report Saline Water Conversion Demonstration Plant No. 1 Freeport, Texas Stearns-Roger Manufacturing Co

OSW Annual Saline Water Conversion Reports for 1959 through 1965

2. Multistage Flash (MSF)

The MSF process is the most highly developed of the distillation processes. Nearly 100 MSF evaporator units of 100,000 gpd or greater capacity have been installed throughout the world. Sixteen of these installations have a capacity of more than 1 mgd. The largest installation, located in Kuwait, has a capacity of approximately 6 mgd.

The most widely publicized and documented MSF evaporator plant is the OSW Demonstration Plant No. 2, initially located at Pt. Loma, California, and subsequently moved to Guantanamo, Cuba. This well-known plant has been selected as representative of MSF evaporator units. Its features are typical of more advanced units - the advances primarily being internal design improvements and cost reductions. This plant, commonly referred to as the Pt. Loma-I Plant, employs a 36-stage flash evaporator and has the following characteristics:

Capacity	1 mgd
Economy ratio	Approximately 11 lb of product/lb of steam
Source of seawater	Pacific Ocean water via inlet marine conduit
Source of heat	Oil-fired boiler

The San Diego Demonstration Plant was placed in service on March 5, 1962. It was shut down on February 26, 1964 for dismantling and subsequent shipment to Guantanamo, Cuba. During this slightly less than 2-year operation at Pt. Loma, the plant produced approximately 500 million gallons of distilled water and was in operation more than 70% of the total elapsed time. The construction cost of the San Diego plant was $1,663 000. and the normalized water production cost reported in the OSW 1964 Saline Water Conversion report was $1. 18 per thousand gallons.

a. Plant Description

The Pt. Loma-I Plant was located adjacent to the ocean, on the western side of the Pt. Loma peninsula. The 36-stage evaporator units were arranged in nine rectangular enclosures. In addition to the evaporator stages, brine heater and auxiliaries, the facility had an administration building containing an office, a control room, a small chemical laboratory and a work shop. The facility received its seawater through a 24 inch diameter marine conduit, extending 700 feet into the ocean. An oil-fired boiler was provided to generate steam for the brine heater and various steam-driven auxiliaries. The product water was pumped to the City of San Diego, and the concentrated brine returned to

the sea through a short outfall. Photographs of the San Diego facility are shown in Figure V-5. Additional photographs of the Guantanamo facility and other MSF facilities are shown in Figure V-6.

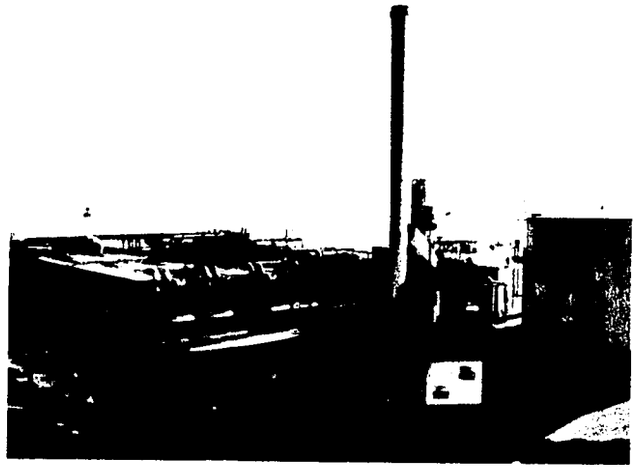
b. Process Description

A simplified process flow diagram of the Pt. Loma-I Plant is shown in Figure V-7. Raw seawater (which has been chlorinated for control of marine growth) passes through the heat-reject stages of the MSF evaporator. The major portion of this seawater is returned to the sea. That portion which is to become makeup is pretreated by acidification to decompose the carbonates and by degasification and deaeration to remove the CO_2 and noncondensables. This makeup stream enters the MSF evaporator in the lowest-temperature reject stage (36th stage). It joins the unflashed brine in the flashing-brine stream and enters the brine recycle pump, where it is pressurized and pumped through the tube bundles of the heat-recovery stages. In the heat-recovery stages the brine-recycle stream is successively heated to approximately 240 F. The brine-recycle stream, still under pressure to prevent boiling, is further heated in the brine heater to approximately 250 F, and is then permitted to flash in the flash chamber of the first stage. This flashing-brine stream flows successively to the flash chambers of each lower temperature (and lower pressure) stage, i.e., the heat-recovery stages and the heat-reject stages, flashing and cooling as it passes through these stages. The vapor generated when the brine stream flashes is condensed to product water on the tube-bundles through which the recycle stream is being pumped; this heats the brine recycle stream. A portion of the brine recycle stream is removed as concentrated brine blowdown; the brine concentration of the brine recycle stream is controlled by the quantity of brine blowdown removed. The quantity of treated seawater makeup to the flash evaporator must equal that removed from the flash evaporator as product and blowdown.

The primary heat supply to the flash evaporator is the steam supplied to the brine heater from a boiler or other steam supply source. The points of heat loss other than those in the equipment and piping, are in the brine blowdown stream, in the product water stream and in the waste

FIGURE V-5

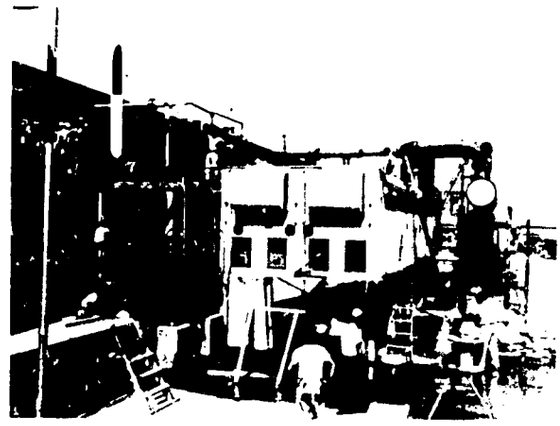
OSW DEMONSTRATION PLANT NO. 2
SAN DIEGO (PT LOMA) CALIFORNIA--MSF PROCESS



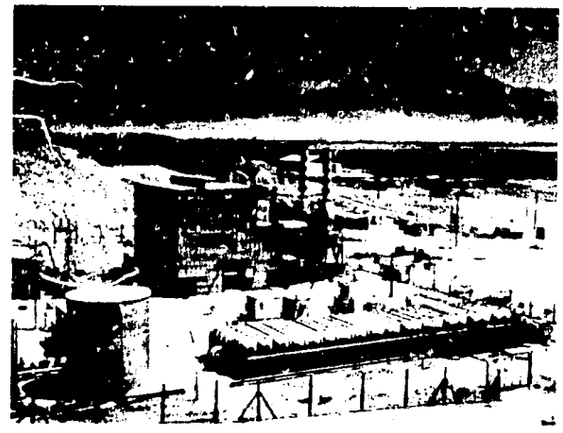
Source: U. S. Department of the Interior,
Office of Saline Water

FIGURE V-6

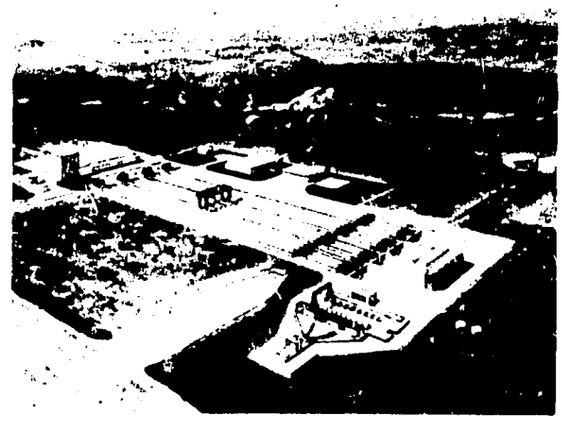
MULTISTAGE FLASH DISTILLATION PLANTS



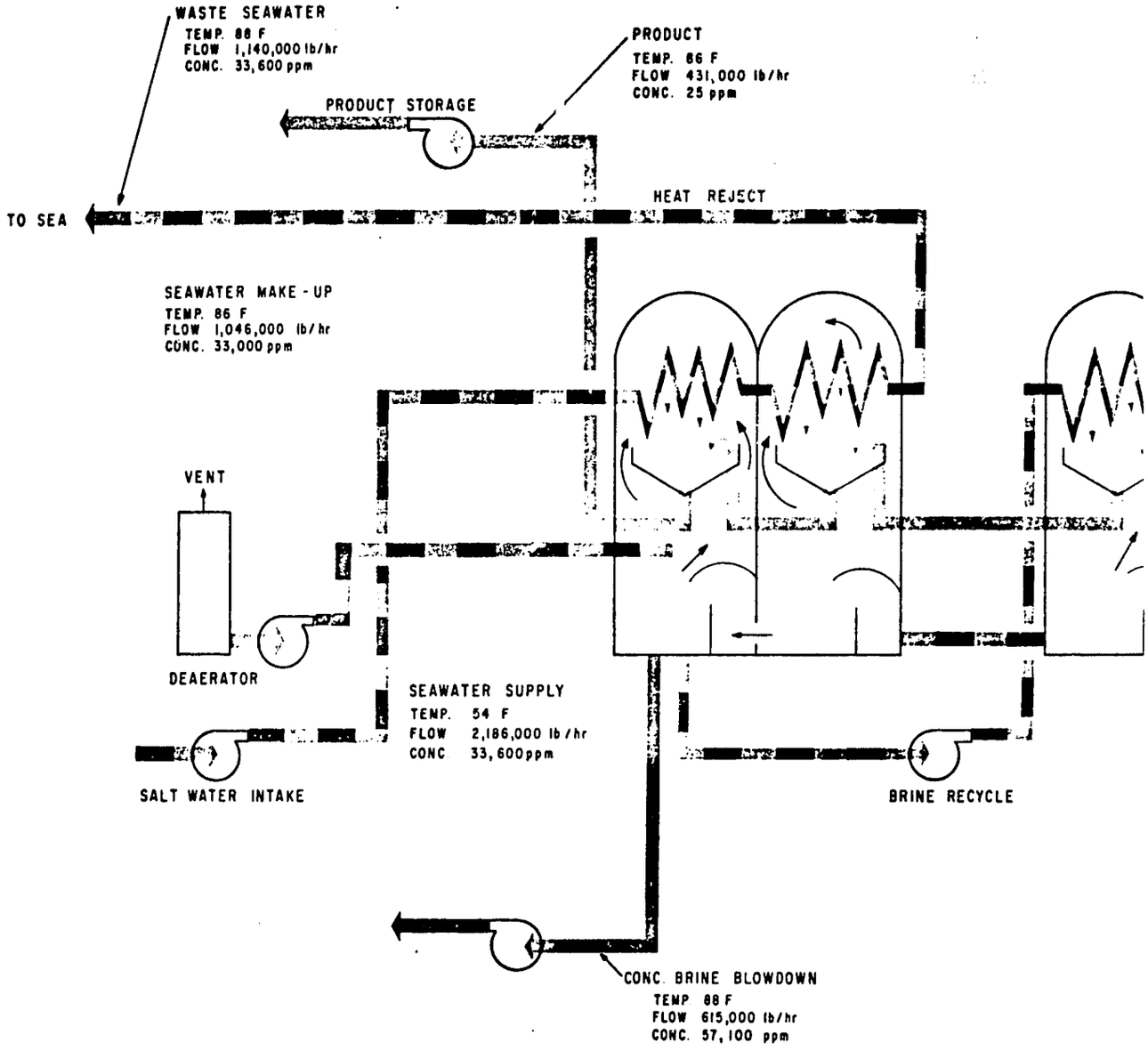
Pt Loma Plant Relocated at Guantanamo Bay, Cuba
Courtesy: Westinghouse Electric Corporation



1-mgd MSF Plant at Eilat, Israel
Courtesy: Baldwin-Lima-Hamilton Corporation



Proposed Dual-Purpose Power-Water Desalting Plant
for Israel--200 MWe/100 mgd


WASTE SEAWATER
TEMP. 88 F
FLOW 1,140,000 lb/hr
CONC. 33,600 ppm
PRODUCT
TEMP. 86 F
FLOW 431,000 lb/hr
CONC. 25 ppm
PRODUCT STORAGE
HEAT REJECT
TO SEA
SEAWATER MAKE-UP
TEMP. 86 F
FLOW 1,046,000 lb/hr
CONC. 33,000 ppm
VENT
DEAERATOR
SEAWATER SUPPLY
TEMP. 54 F
FLOW 2,186,000 lb/hr
CONC 33,600 ppm
SALT WATER INTAKE
BRINE RECYCLE
CONC. BRINE BLOWDOWN
TEMP 88 F
FLOW 615,000 lb/hr
CONC. 57,100 ppm

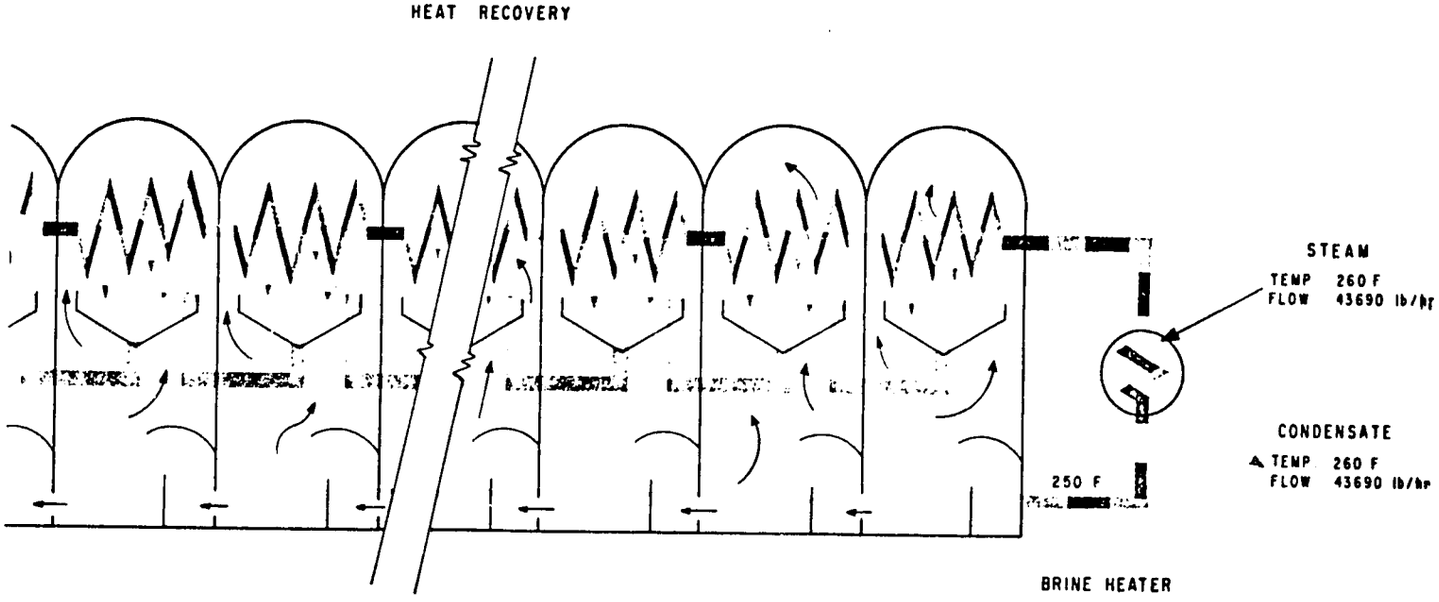


FIGURE V-7

MULTISTAGE FLASH PROCESS

OSW DEMONSTRATION PLANT NO.2

SAN DIEGO (PT. LOMA), CALIF

seawater passing through the heat reject stages. The process supply streams are the raw seawater to the heat-reject stages, the treated seawater makeup and the steam supply. The discharge streams are the product water, the blowdown brine, the waste seawater, and the brine heater condensate return to the point of steam supply.

c. Special Features

The Pt. Loma-I Plant was piped in a complex manner for experimental purposes. It was found, however, that the simplified piping shown in the typical process flow diagram was adequate. The plant was initially operated with a brine heater brine outlet temperature of 200 F and with Hagevap-LP pretreatment. The flash evaporator enclosures and the brine heater were designed to permit operation at a brine heater brine outlet temperature of 250 F. A sulfuric acid pretreatment system was added and the unit was then operated at this higher brine heater brine outlet temperature, thereby increasing the capacity of the plant to approximately 2 mgd with a maximum 24-hour capacity of 1.4 mgd.

d. Problems Encountered

The marine conduit seawater supply for the Pt. Loma-I Plant extended approximately 700 feet into the ocean. In spite of this distance, clogging of the intake screens and the evaporator tube sheets by marine growth caused frequent plant shutdowns during certain seasons. During periods of high wave, waterborne sand and silt was drawn in through the intake conduit. During one 24-hour period, 5 tons of sand and silt accumulated in the intake pump pit. Many other problems developed due to the clogging of tube sheets and the restricted seawater supply. Partially as a result of waterborne sand and silt in the recycle stream, recycle pump bearings and rotating parts failed. Plans for resolving these difficulties were being made at the time of the decision to move the plant to Guantanamo, Cuba.

e. Present Use

The Pt. Loma-I Plant was designed and operated as a demonstration facility for obtaining experimental data. With its reinstallation at Guantanamo, Cuba, this plant together with two plants of a similar design became the principal source of potable water for the U.S. Naval Base there.

f. Plant Staff

The plant staff consists of approximately 13 personnel for continuous around-the-clock operation. These personnel are:

Plant Superintendent

Chemical Engineer

Administrative Assistant

Clerk-Typist

Nine Plant Operators, including maintenance personnel

The plant is sufficiently automated so that only one Plant Operator is required per shift.

For further discussion of the Pt. Loma-I Plant see the following publications:

OSW Report No. 102 First Annual Report, Saline Water Conversion Demonstration Plant San Diego, California Burns & Roe Inc. Order No. PB 181684

OSW Report No. 114 Second Annual Report Saline Water Conversion Demonstration Plant Point Loma-I, San Diego, California Burns & Roe, Inc Order No. PB 166247

OSW Annual Saline Water Conversion Reports 1959 through 1965

3. Multieffect Multistage Flash (MEMS)

The MEMS process combines the advantages of the MSF process (boiling in chambers rather than in heat-transfer tubing) and the multieffect process (low brine concentration at the highest brine temperature). The MEMS process is potentially well suited to single-purpose desalination plants since it is possible to achieve higher brine temperatures with the MEMS process than with the MSF process. The higher brine temperature leads to a higher economy ratio which in turn, may result in lower unit water cost. For dual-purpose plants, the urgency of achieving higher brine temperatures is less since the lower cost of low-temperature steam from the turbine plant causes the economic optimum for maximum brine temperature to be lower than the limiting temperature of the MSF process. The first large MEMS evaporator is being constructed at San Diego, California as a replacement for the OSW Demonstration Plant No. 2 (Pt. Loma-I Plant) which has

been moved to Guantanamo, Cuba. The new MEMS unit, which will be called the Clair Engle Plant, has the following characteristics:

Capacity	1 mgd
Performance ratio	21.5
Source of seawater	Pacific Ocean water via San Diego Gas & Electric Plant inlet canal
Source of heat	Purchased from San Diego Gas & Electric

a. Plant Description

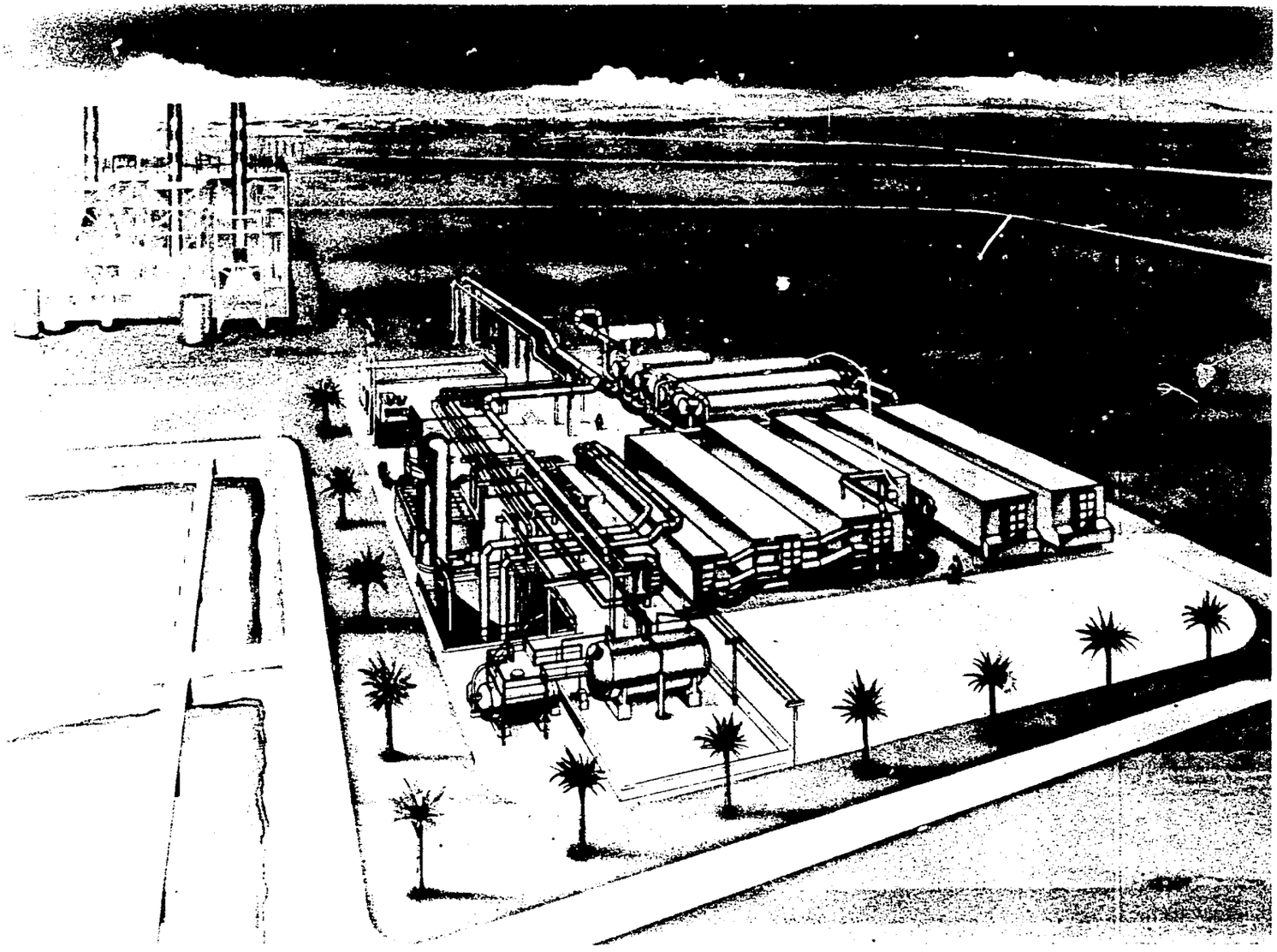
The Clair Engle Plant is situated adjacent to the San Diego Gas & Electric South Bay facility, and shares its site with the OSW West Coast Test Facility and the 17-mgd Module Test Facility. The facility has an administration building containing an office, a control room, a small chemical laboratory and a work shop, and office space for the test facility contractors. The product water from the Clair Engle Plant is pumped to the City of San Diego, and the concentrated brine is returned to the sea through a short outfall. An illustration of the San Diego facility is shown in Figure V-8.

b. Process Description

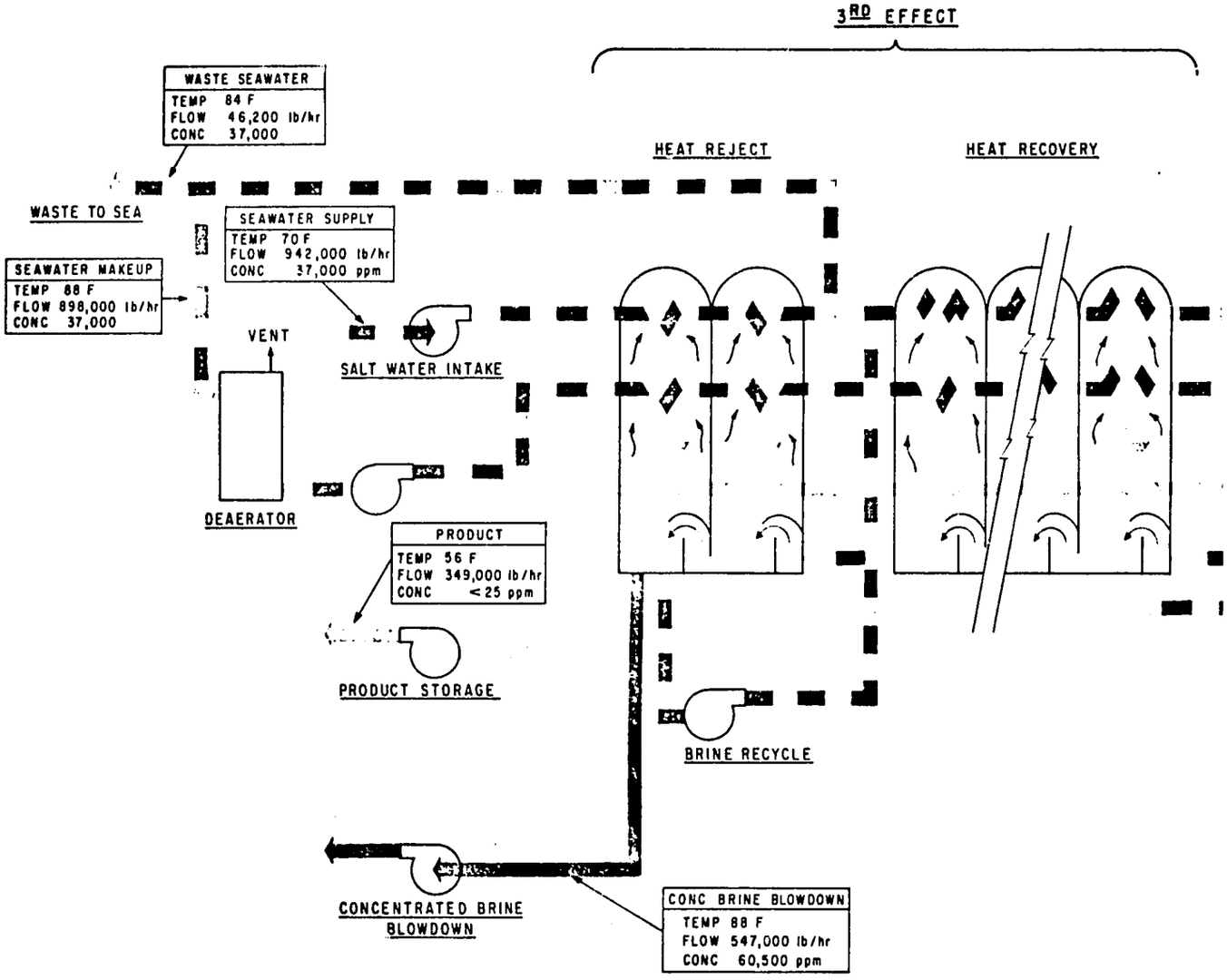
A simplified flow diagram of the Clair Engle Plant is shown in Figure V-9. Raw seawater (which has been chlorinated for control of marine growth) passes through the heat-reject stages of the third effect of the MEMS plant. A major portion of the seawater is returned to the sea. That portion which is to become makeup is pretreated by acidification to decompose the carbonates and by degasification and deaeration to remove the CO_2 and noncondensables. This makeup stream is pumped under pressure through tube-bundles in the heat-reject and heat-recovery stages of the third and second effects, and joins the brine recycle stream of the first effect recycle pump inlet. This brine recycle stream is pumped through heat-recovery stages of the first effect where it is successively heated to approximately 244 F, and through the brine heater where it is further heated to 250 F. It is then permitted to flash successively through the flash chambers of the first effect heat-recovery and heat-reject stages, being cooled as it flashes. The vapor generated when the brine stream flashes is condensed on the tube-bundles (through which the recycle stream was pumped) as product water, and the unflashed brine again joins the brine recycle

FIGURE V-8

OSW SAN DIEGO NO. 2 PLANT
SAN DIEGO, CALIFORNIA--MEMS PROCESS



Source: U. S. Department of the Interior,
Office of Saline Water


3RD EFFECT
WASTE SEAWATER
TEMP 84 F
FLOW 46,200 lb/hr
CONC 37,000
HEAT REJECT
HEAT RECOVERY
WASTE TO SEA
SEAWATER SUPPLY
TEMP 70 F
FLOW 942,000 lb/hr
CONC 37,000 ppm
SEAWATER MAKEUP
TEMP 88 F
FLOW 898,000 lb/hr
CONC 37,000
VENT
SALT WATER INTAKE
DEAERATOR
PRODUCT
TEMP 56 F
FLOW 349,000 lb/hr
CONC < 25 ppm
PRODUCT STORAGE
BRINE RECYCLE
CONCENTRATED BRINE BLOWDOWN
CONC BRINE BLOWDOWN
TEMP 88 F
FLOW 547,000 lb/hr
CONC 60,500 ppm

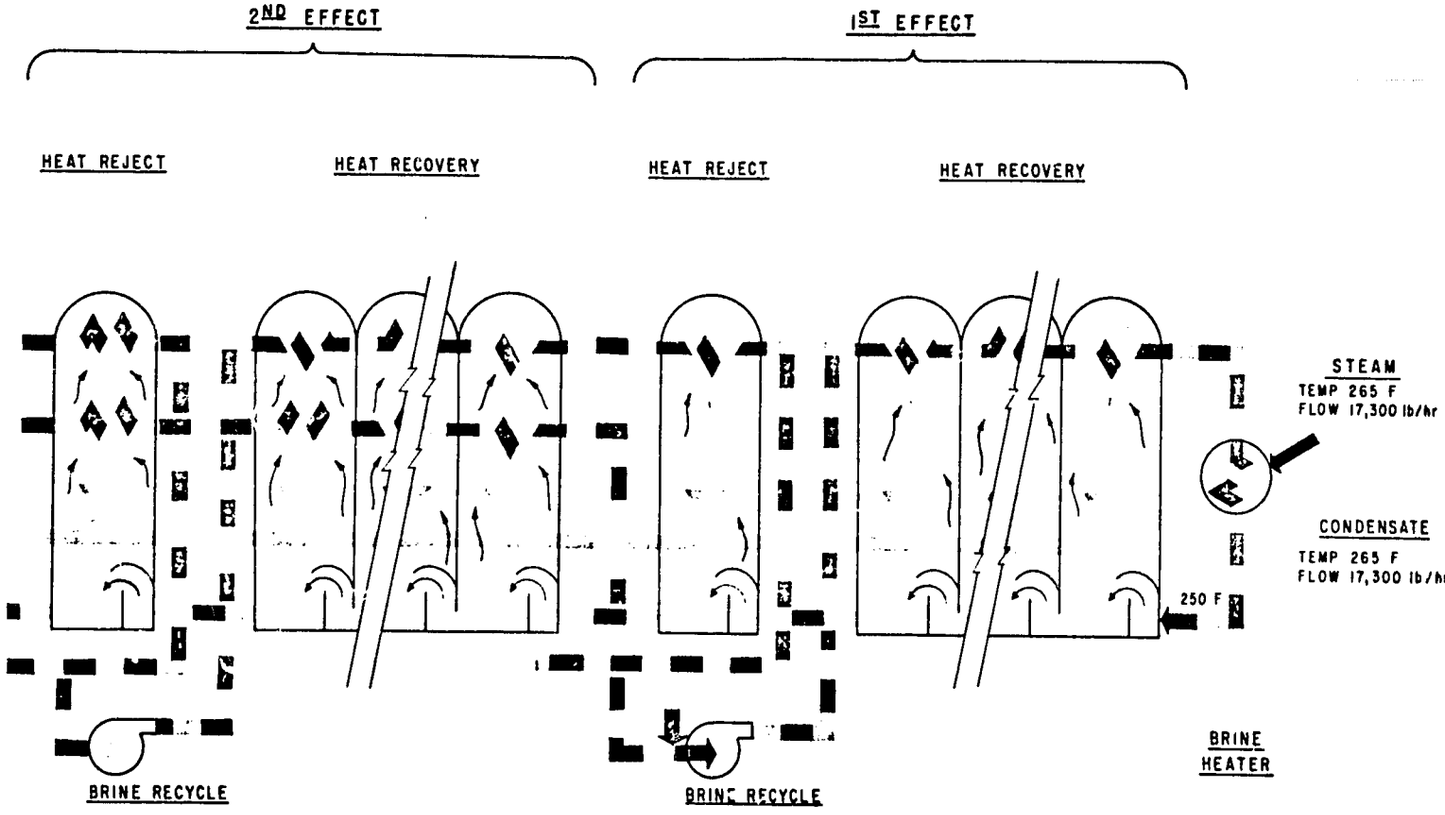


FIGURE V-9

MULTIEFFECT MULTISTAGE FLASH PROCESS

OSW DISTILLATION PLANT - SAN DIEGO NO. 2

SAN DIEGO, CALIF

stream. The concentration ratio in the recycle stream is held to about 1.2 by controlling the quantity of brine blowdown removed from the flashing brine as it leaves the heat-reject stage of the first effect. This blowdown becomes the makeup to the flashing brine stream of the second effect as this stream enters the flash chambers of the highest-pressure heat-recovery stage of the second effect.

The process flow in the second effect is similar to the process flow in the first effect. A brine recycle stream is pumped through tube-bundles in the heat-recovery stages, and is successively heated in each. As this brine recycle stream leaves the highest-temperature heat-recovery stage, it passes through the tube-bundle of the heat-reject stage of the first effect where it is further heated. This recycle stream then enters the flash chamber of the highest-temperature heat-recovery stage of the second effect, where it is joined by the brine blowdown from the first effect. This flashing brine stream then flashes successively through the flash chambers of the heat-recovery and heat-reject stages, and is successively cooled as it passes through these stages. The vapor generated when the brine stream flashes is condensed on the tube-bundles, as in the first effect. It should be noted that the heat-reject stage of the first effect supplies heat to the second effect in more or less the same manner as the brine heater supplies heat to the first effect.

The process in the third effect is the same as that in the second effect, except that the brine blowdown leaving the lowest-pressure heat-reject stage of the third effect has now reached a concentration ratio of approximately 1.6; this blowdown is then discharged to the sea.

Each of the three effects described above operates as if it were an MSF evaporator unit. It should be noted that the concentration of the brine recycle stream in each effect can be separately controlled, the concentration in the first effect is usually controlled to 1.2; that in the second effect is approximately 1.4 and that in the third effect is approximately 1.6.

Figure V-9 also includes a simplified heat and material balance for this process. The primary heat supply to the MEMS evaporator is the steam supplied to the brine heater. The points of heat loss, other than those in the equipment and piping, are in the brine blowdown stream, in the product water stream and in the waste seawater passing through the heat-reject stage in the third effect. The process supply streams are the raw seawater to the heat-reject stages, the treated seawater makeup (whose quantity must be equal to the product plus the blowdown) and the steam supply. The discharge streams are the product water, the blowdown brine, the waste seawater and the brine heater condensate return to the steam supply.

c. Special Features

The Clair Engle Demonstration Plant will be the first large MEMS unit to be constructed. It is designed to operate with a top brine-heater brine-outlet temperature of 350 F and a third-effect brine concentration ratio of 1.6. The design economy ratio of this unit is 20 lb of product per pound of steam. Because of the experimental nature of this plant, it has been equipped with a large number of instruments, and will be operated over a wide range, to obtain as much performance and design data as possible.

d. Problems Encountered

The Clair Engle Multieffect Multistage Flash Plant is at present in the construction stage.

e. Present Use

The Clair Engle Demonstration Plant is intended to be a developmental unit.

f. Plant Staff

The plant staff has not yet been established.

4. Vapor Compression

The vapor compression process is the only major distillation process which does not require large quantities of cooling water; this process is, therefore, particularly well suited for the conversion of brackish water at inland locations where cooling water may be in short supply. The OSW Brackish Water Conversion Demonstration Plant No. 4 is located at Roswell, New Mexico in a water-short area which has very brackish underground water, and this plant has been chosen as typical of plants using vapor compression processes. It is the fourth in a series of OSW demonstration plants designed, built and operated to demonstrate a particular process and to obtain operating, maintenance and cost data on an actual plant installation.

The Roswell plant utilizes the forced circulation-vapor compression distillation process to produce potable water from a brackish well-water: the plant has the following characteristics:

Capacity	1 mgd
Plant performance	55 kilowatt-hours per 1,000 gallons
Source of brackish water	Brackish well-water
Source of energy	Purchased power

The construction of the Roswell plant was completed in July 1963. Since that time, considerable difficulties have been experienced at the plant due to equipment failures and to scale and corrosion problems, therefore, the plant has not obtained a consistent level of operation. In the period July 1, 1964 to June 30, 1965 the plant produced water during only approximately 20% of the time. While the operation of the Roswell vapor compression process has not yet been fully demonstrated, many of the present problems can be attributed to the characteristics of the very brackish well-water

supply. This water contains approximately one-half as much total dissolved solids as seawater; however, it contains higher concentrations of calcium and silica than seawater and this makes scale-free processing difficult. At present, other vapor compression units are operating successfully on less difficult raw water and seawater.

The only major utility service required by the Roswell plant is electrical power for the vapor compressor; no cooling water is required for the process. The construction cost of the plant was $1,794,000. The actual water production costs are not typical of successfully operating vapor compression plants. Typical water production costs for a plant of this size would be between $1.00 and $1.50 per 1,000 gallons.

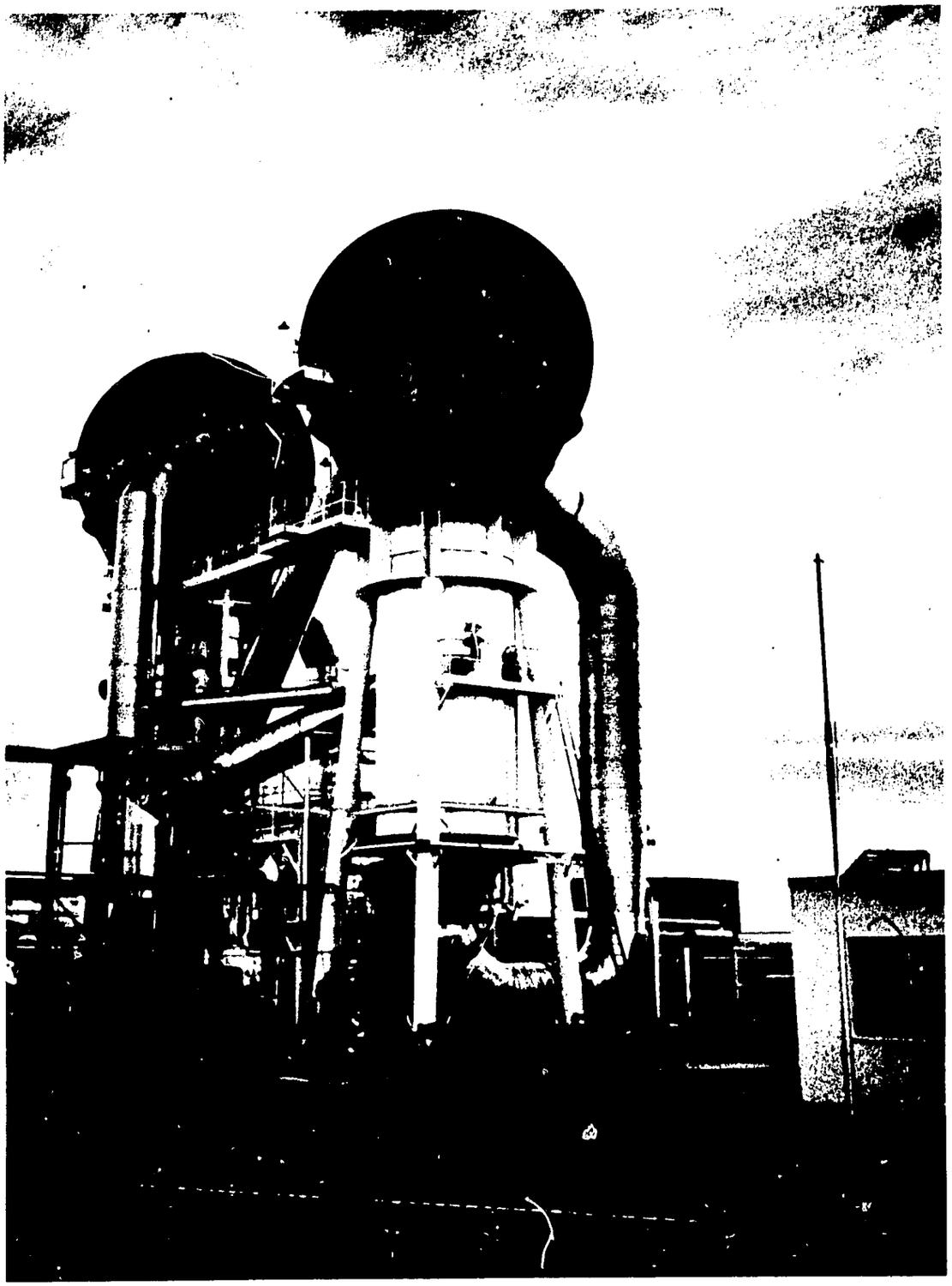
a. Plant Description

The OSW Demonstration Plant No. 4, located on the outskirts of Roswell, New Mexico, receives its water from a brackish water well. The demonstration plant itself is a two-effect vertical forced circulation vapor compression unit. Because of the high calcium and silica content of the brackish water supply, an ion-exchange feedwater pretreatment system has been installed. This is not typical of vapor compression plants; on seawater and most brackish water sources, sulfuric acid pretreatment with degasification and deaeration would be required instead of ion exchange.

The administration building at Roswell contains the office, a control room, a small chemical laboratory area and a maintenance facility. An oil-fired boiler provides steam for startup. Startup steam is required in all vapor compression plants to bring the plant to operating temperature. The plant supplies product water to the City of Roswell, and disposes of the concentrated brine in a plastic-lined brine evaporation pond located near the site. A photograph of the Roswell demonstration plant facility is shown in Figure V-10.

FIGURE V-10

VAPOR COMPRESSION PLANT

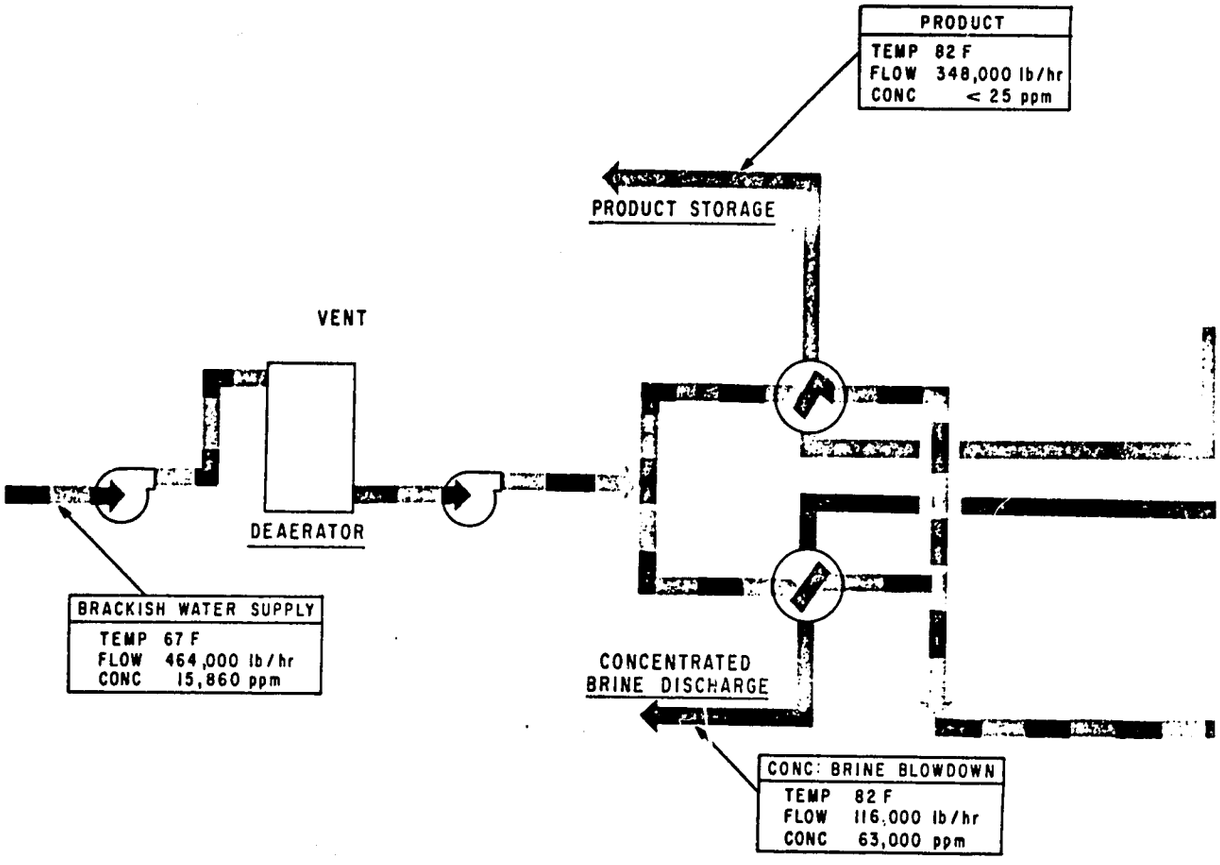


Source: U. S. Department of the Interior,
Office of Saline Water

b. Process Description

A simplified flowsheet for the Roswell plant is shown in Figure V-11. The incoming brackish water is treated as the water composition requires, typically by sulfuric acid pretreatment with degasification and deaeration. At Roswell, as previously noted, the incoming well-water is treated in an ion-exchange system in which a portion of the calcium and magnesium ions are replaced by sodium ions, and then acidified and deaerated to remove the CO_2 and oxygen. The treated feed stream is then preheated in heat-recovery heat exchangers, where heat is exchanged with the effluent product water and brine blowdown streams. The feed stream then joins the brine recycle stream of the first effect before that stream enters the recycle pump. The brine recycle pump forces the brine recycle stream upward through the tubes of this effect. Steam from the vapor compression discharge condenses on the shell side of the tubes and raises the temperature of the recycle brine in the tubes by approximately 4 F. The brine recycle stream flashes as it leaves the tube bundle and this mixture of steam and hot brine is separated in a large spherical flash chamber. The steam passes to the shell side of the second effect and is therein condensed as product water. The heat of condensation is utilized in the second effect to heat the brine recycle stream therein. The unvaporized brine in the first effect flows to the bottom of the flash chamber, from where it is circulated through an external downcomer to the recycle pump. A small part of the brine recycle stream is removed as blowdown from the first effect and flows into the downcomer of the second effect as the makeup to the brine recycle stream of the second effect.

As the brine recycle stream passes through the vertical tube bundle of the second effect it is heated by the condensation of steam from the first effect. This heated brine stream emerges into the flash chamber, where it flashes. The mixture of steam and hot brine is separated in the flash chamber, and the steam is pumped by the vapor compressor to the first effect shell, where it condenses as product water. This vapor compressor, an axial flow unit driven by a 2,500-hp electric motor, is one of the largest such compressors ever built; it compresses 175,000 pounds of steam per hour from 17 psia to 24 psia. The work of this compressor is the main energy input required for the process.


PRODUCT
TEMP 82 F
FLOW 348,000 lb/hr
CONC < 25 ppm
PRODUCT STORAGE
VENT
DEAERATOR
BRACKISH WATER SUPPLY
TEMP 67 F
FLOW 464,000 lb/hr
CONC 15,860 ppm
CONCENTRATED
BRINE DISCHARGE
CONC. BRINE BLOWDOWN
TEMP 82 F
FLOW 116,000 lb/hr
CONC 63,000 ppm

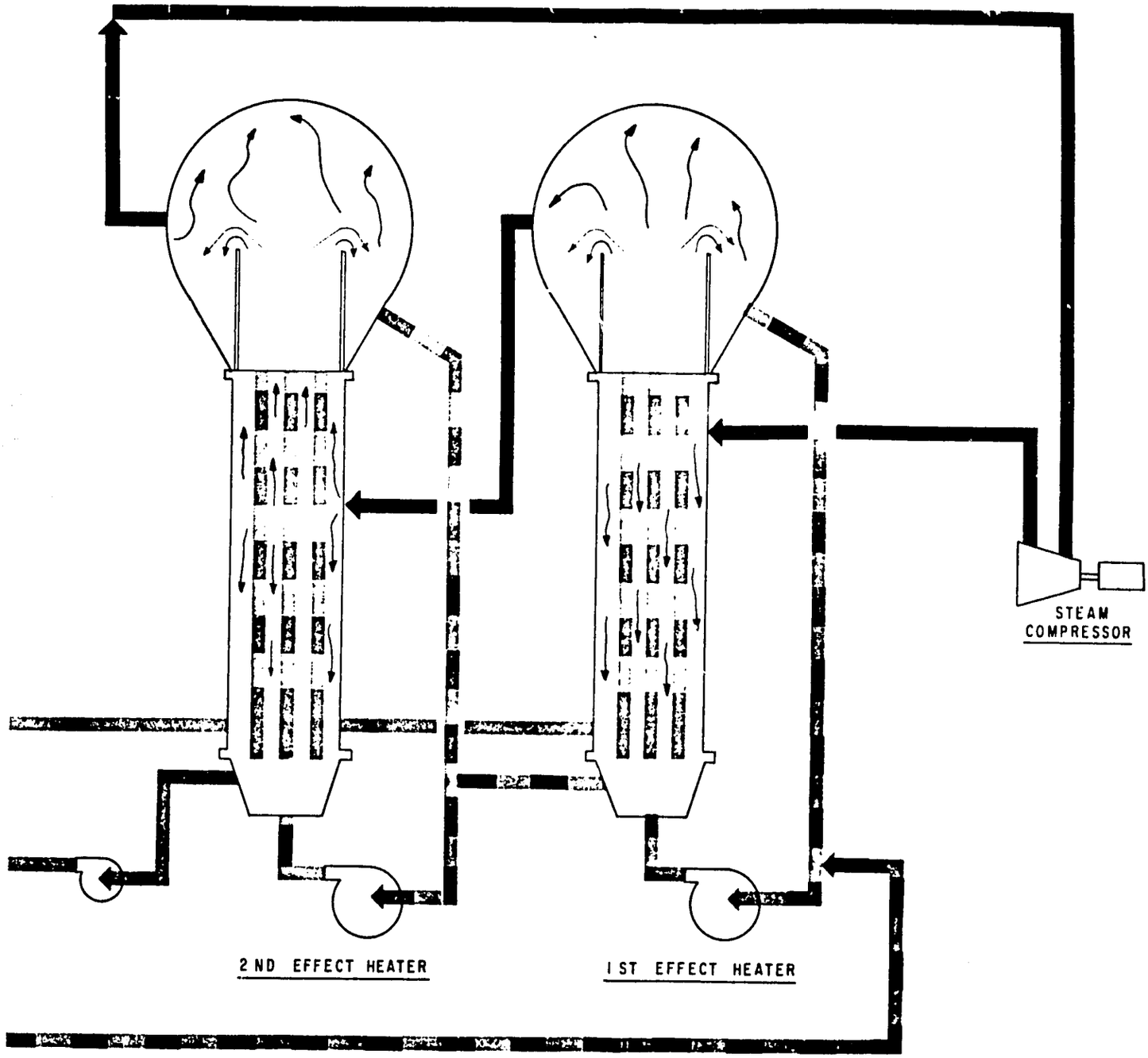


FIGURE V-11

VAPOR COMPRESSION PROCESS

OSW DEMONSTRATION PLANT NO. 4

ROSWELL, NEW MEXICO

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

A simplified heat and material balance for the process is also given in Figure V-11. The primary energy supplied to the system is the electrical energy supplied to the compressor. Heat losses from the cycle are counter-balanced by the energy supplied to the process by the compressor work. The only process supply stream is the brackish water feed to the process, and the discharge streams are the product water and the brine blowdown.

c. Special Features

Typically, sulfuric acid pretreatment with degasification and deaeration is required for brackish water processing, but the composition of brackish water varies greatly from location to location and, occasionally, special pretreatment systems will be necessary for inland brackish water desalination plants. The high calcium and silica content of the brackish water supply at Roswell required a number of special features in the pretreatment system.

The vertical forced circulation configuration utilized at Roswell is one configuration of the vapor compression process. The heat-transfer tube-bundles can equally well be situated in a horizontal manner, as in a flash evaporator. In fact, one or two stages of the flash evaporator can be utilized in the vapor compression process by removing the steam from the flash chamber of the lowest temperature stage, compressing it and utilizing this steam as the heat source to the brine heater.

d. Problems Encountered

The difficulties encountered at Roswell are primarily associated with the difficulty of adequately pretreating the brackish water to prevent formation of calcium and silica scales.

e. Present Use

The Roswell plant continues as a developmental facility for vapor compression and is being expanded as a test facility for other brackish water processes.

f. Plant Staff

The staff required to operate this demonstration plant facility is approximately 14 personnel:

Plant Manager

Plant Engineer

Plant Chemist

Accountant-Clerk

6 Plant Operators

4 Maintenance Men

For further discussion of the vapor compression process, see Subsection IV. A. 2. b. Additional information is contained in the following publications:

OSW Report No. 169 First Annual Report Brackish Water Conversion Demonstration Plant No. 4, Roswell, New Mexico (American Hydrotherm Corp.)

OSW Report No. 170 Second Annual Report, Brackish Water Conversion Demonstration Plant No. 4, Roswell, New Mexico (American Hydrotherm Corp.)

OSW Annual Saline Water Conversion Reports for 1959 through 1965

5. Solar Stills

Solar stills are the oldest and simplest of the distillation processes, and can be so constructed as to require no energy except the heat received from the sun. This solar energy, however, even in the most favorable areas in the world seldom exceeds 2,000 Btu per square foot per day, and most solar stills require approximately ten square feet for each daily gallon of capacity. Although the construction of the still is very simple, and inexpensive materials such as an asphalt lined basin covered by a thin plastic are utilized, the cost is of the order of 50 cents per square foot. This means that the capital cost is approximately $5 per daily gallon of capacity.

In smaller sizes, the costs of other desalting processes, such as the MSF process, are sufficiently high for solar stills to be competitive, but in larger sizes the capital cost of other processes is as low as 50 cents to $1 per daily gallon of capacity. Thus capital cost limits the use of solar stills to installations of less than 50,000 gpd and to date has limited their size to 20,000 gpd.

The first large solar still was built in Chile in 1875, had a capacity of about 6,000 gpd and operated for a number of years. Solar stills which are in operation or under construction today are shown in Table V-1.

TABLE V-1

TODAY'S LARGE SOLAR STILLS

Location			Capacity (gpd)	Type
Australia	-	Northam	330	Glass
	-	Coober Pedy	3, 500***	Glass
Greece	-	Aegina	2, 500	Plastic
	-	Kimolos	3, 000	Glass
	-	Nisyros	3, 000	Glass
	-	Patmos	10, 000	Glass
	-	Salamis	500	Plastic
	-	Santorini	*	--
	-	Symi	4, 000**	Plastic
Mexico	-	Puerto Penasco	4, 000	Plastic - Multiple Effect Humidification
Portugal	-	Cape Verde (Santa Maria do Sal)	--	Plastic
Spain	-	Las Marinas	1, 000	Glass
	-	Tabarca	2, 500	Glass
Tunisia	-	Gafia	750	Glass

*Planned
**Variously reported as 4, 000 to 7, 500 gpd
***Yearly average, but production expected to be 6, 500 gpd during summer

The plant chosen as representative of solar stills is the 4,000-gpd plant recently installed on the island of Symi, Dodecanese Islands, Greece, sponsored by the Church World Service and constructed in the summer of 1964. This plant has the following characteristics:

Capacity	4,000 gpd plus rain catchment (variously reported as 4,000 to 7,500 gpd)
Performance ratio	Not applicable
Source of seawater	Aegean Sea
Source of heat	Solar energy

a. Plant Description

The 4,000-gpd solar distillation plant at Symi is located in the village square, and provides water for a village of approximately 3,000 people. It consists of 14 bays approximately 150 feet long and covering a total area of approximately 29,000 square feet. The facility consists of a saltwater feed pump and sedimentation reservoir, the still itself and a freshwater sump. The still itself consists of a lined basin covered with an air-supported sheet of tough, stable plastic film. The air is provided by four blowers which consume a total of 160 watts of electrical energy. The air-supported wettable film is resistant to winds and punctures, and when a puncture occurs it is easily repaired. Photographs of the Symi Solar Distillation Plant are shown in Figure IV-12, and photographs of other solar distillation facilities are shown in Figure V-13.

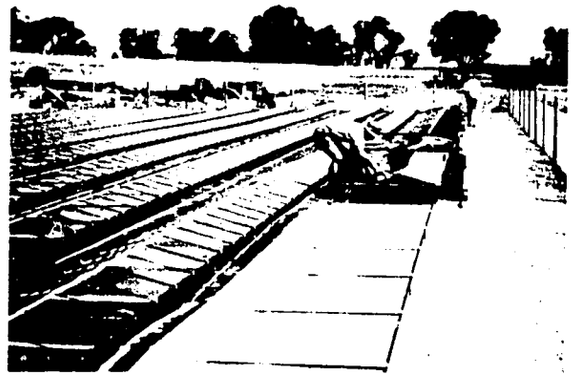
b. Process Description

The solar still is a very simple device. A flowsheet for a solar still is shown in Figure V-14. Each night, the brine resulting from the day's operation is flushed from the lined basins of the solar still by pumping fresh salt water through the still. This also recharges the basins with fresh salt water for the next day's operation. During the day, solar energy

FIGURE V-12

SOLAR STILLS

Solar Distillation Plant, Symi, Dodecanese Islands, Greece Courtesy: Aqua-Sol, Inc

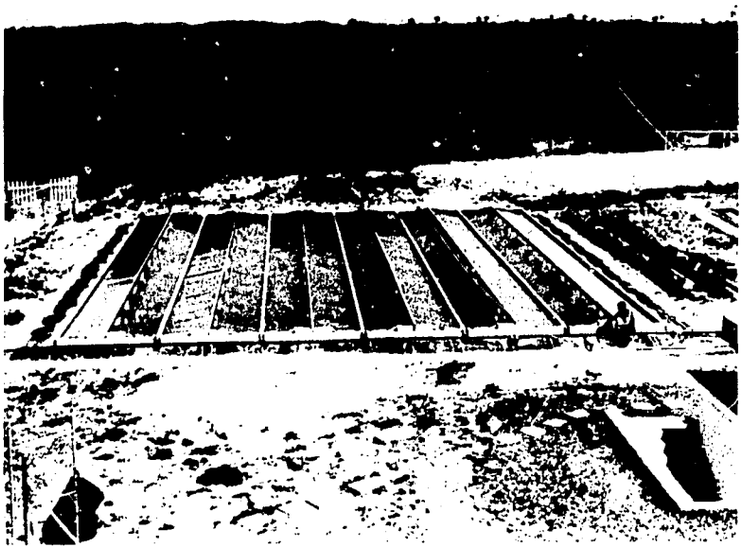


Coober Pedy, Australia

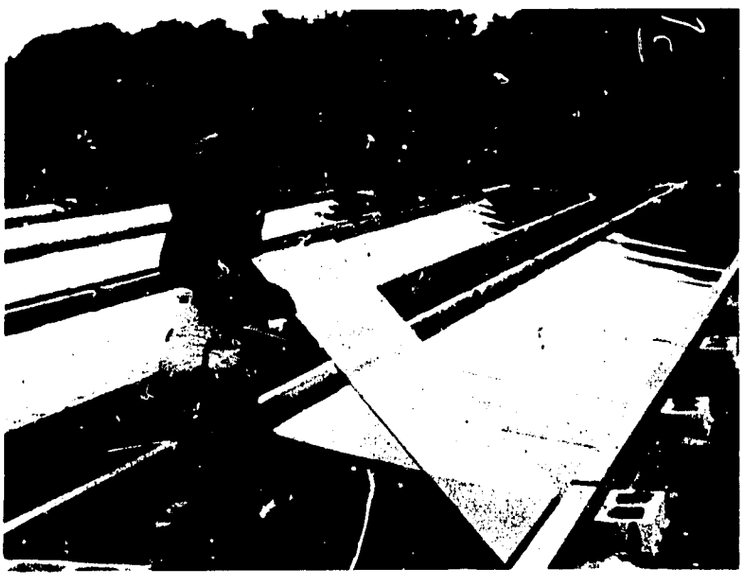
Solar Distillation Plant, Symi, Greece
Source: U. S. Department of the Interior, Office of Saline Water

FIGURE V-13

OTHER SOLAR DISTILLATION PLANTS

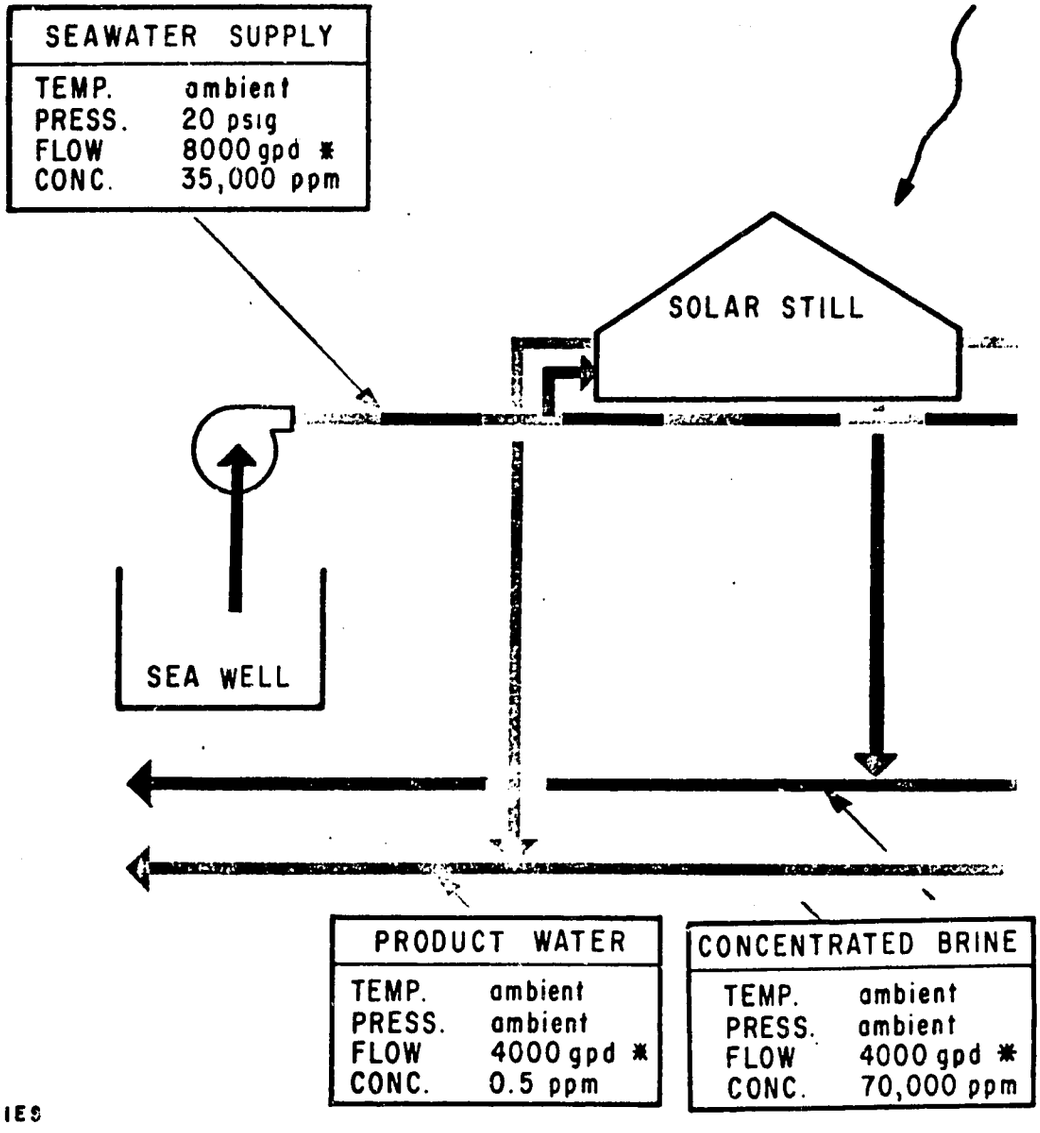


Deep Basin Still, Daytona Beach, Florida



Deep Basin Still, Under Construction

Source: U. S. Department of the Interior,
Office of Saline Water



* VARIES

NOTE: SEAWATER IS PUMPED THROUGH THE SOLAR STILL BASINS AT NIGHT REFILL THE BASINS FOR THE NEXT DAY'S RUN. IF THE RESIDUAL TO THE ADMITTING OF THE FRESH SEAWATER SUPPLY, THE QUANTITY THAT NECESSARY TO REFILL THE BASINS.

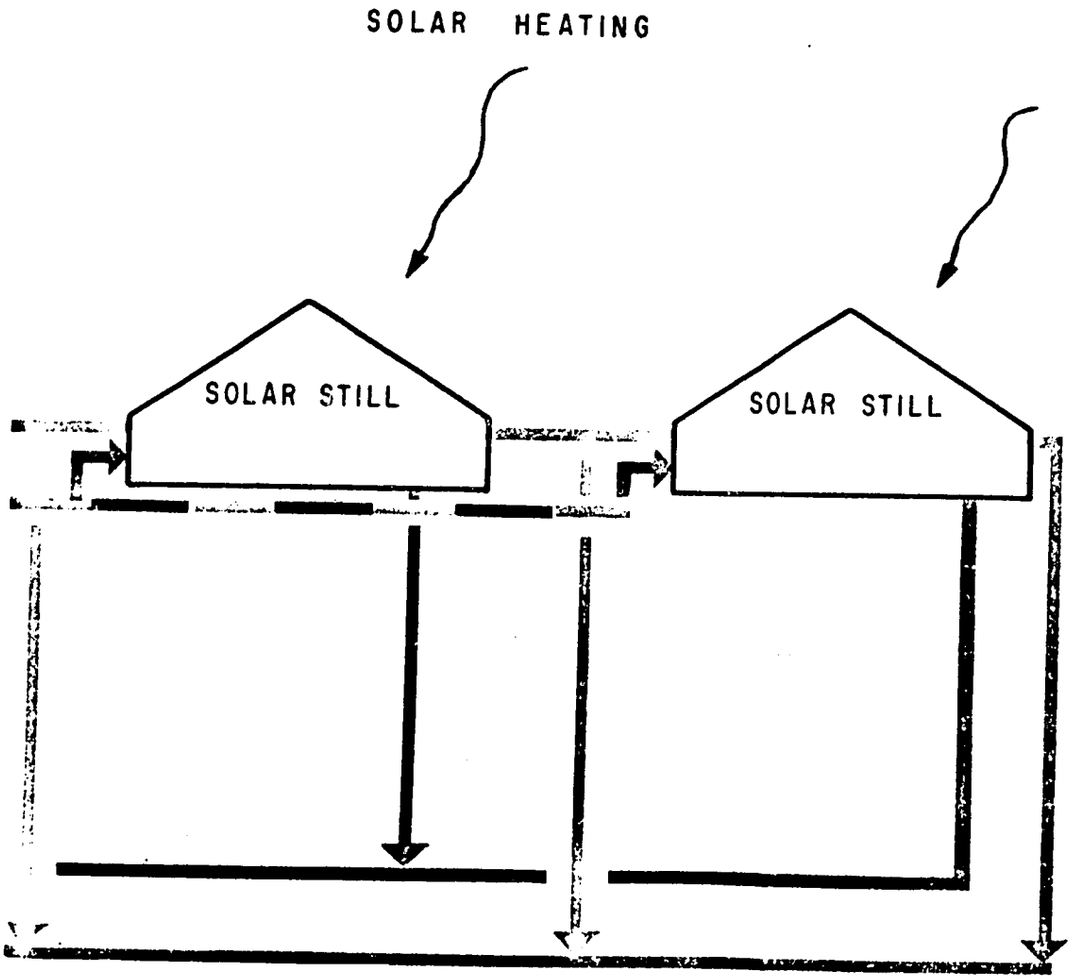


FIGURE V-14

SOLAR DISTILLATION

4,000-TO 5,000-GPD PLANT

SYMI, DODECANESE ISLANDS, GREECE

TO FLUSH THE RESIDUAL BRINE AWAY AND
BRINE IS DRAINED FROM THE BASINS PRIOR
OF SEAWATER SUPPLY REQUIRED IS ONLY

passing through the wettable film evaporates a portion of the seawater. This in turn condenses on the film, and drains to collection troughs and thence to the freshwater sump. The unit can be constructed with a self-supported rather than an air-supported film, in which case it requires no power for operation. The components of the solar distillation unit, with the exception of the plastic film, have an expected life of 20 years or more. A film life of at least 5 years can be reasonably expected. The still is highly reliable, and requires a minimum of maintenance and minimum of operating labor. One bay can be removed from service for repair without affecting operation of the remaining part of the still. The product water is very pure, and usually contains less than one-tenth of one ppm of dissolved salts or minerals.

c. Site Considerations

Because solar stills rely upon solar energy, they are generally limited to the area bounded by latitudes 35° north and 35° south. This boundary, however, may be considerably extended or reduced by local climatic conditions. The ideal site would be on the windward side, on a grade less than 5%, and clear of shadows from hills or large buildings. The orientation of the still on the site is not important. When considering the required capacity for solar stills, it is important to know that average yearly water production will vary as much as plus or minus 30% due to annual climatic changes. The production of the still is normally increased by the rain catchment in areas of significant rainfall, and this should be considered in the design.

d. Problems Encountered

The principal problem of solar stills continues to be their high initial cost. Many stills are constructed with a glass rather than a plastic top structure, which further increases the initial cost. If the plastic films are to operate satisfactorily in solar stills, they must be wettable and tough, and their transparency should not change significantly during their expected life. One problem frequently encountered in solar stills is the accumulation of dust and dirt on the outside surface of the plastic or glass covering. To prevent degradation of performance, the dust must be removed, but this is a time-consuming and sometimes water-consuming process.

e. Present Use

The solar still at Symi and other solar stills listed in Table V-1 are commercial installations; in the smaller capacity range, solar stills are competitive with other processes.

f. Plant Staff

Operating personnel are required for only a short time each day to pump fresh saltwater into the basins and to pump the day's accumulation of freshwater to the storage system. Maintenance, likewise, is on a part-time basis. The requirements for construction labor, and for operating and maintenance personnel can be met by using unskilled labor.

For further discussion of the solar distillation process, see Subsection IV. A. 2. e. Additional information is contained in the following publications:

OSW Report No. 50	First Two Years Progress on Study and Field Evaluation of Solar Seawater Stills PB 171934
OSW Report No. 60	Development of an Improved Solar Still, PB 181144
OSW Report No. 147	Second Two Years' Progress on Study and Field Evaluation of Solar Seawater Stills
OSW Report No. 190	Final Three Years' Progress on Study and Field Evaluation of Solar Seawater Stills

C. Freezing Process - Vacuum Freezing - Vapor Compression

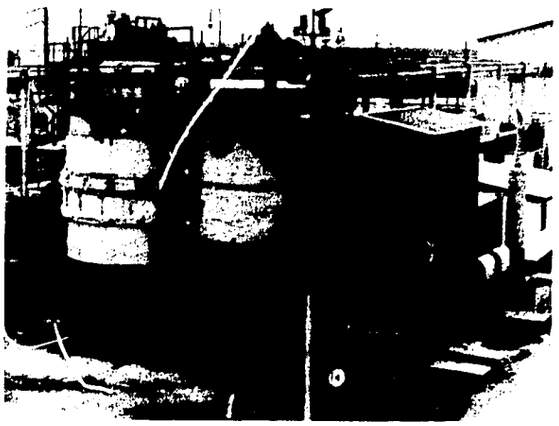
So far, the freezing processes have not been utilized in commercial installations--although the industry is prepared to quote on the vacuum freezing-vapor compression freezing process at the present time. The vacuum freezing-vapor compression process is under development in a four-unit plant in operation at Eilat, Israel; each unit has a capacity of 60,000 gpd.

The plant chosen as typical of this process is the pilot plant installation at Wrightsville Beach, North Carolina; a photograph of this installation is shown in Figure V-15. This plant has the following characteristics:

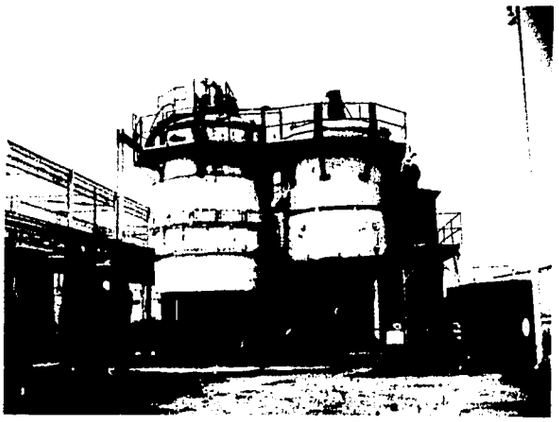
Capacity	Approximately 100,000 gpd
Performance	Approximately 44 kwh per 1,000 gallons
Source of seawater	Atlantic Ocean

FIGURE V-15

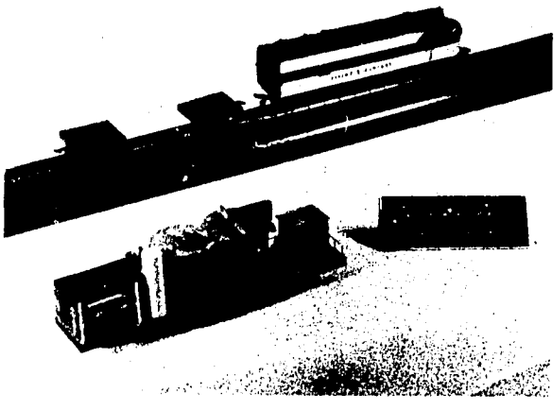
VACUUM FREEZING VAPOR COMPRESSION PILOT PLANT
WRIGHTSVILLE BEACH, N. C.



View of Pilot Plant



View of Pilot Plant



Model of 100,000-gpd Skid Mounted Unit

Courtesy: Colt Industries

The vacuum freezing-vapor compression plant at Wrightsville Beach has been operated on an experimental basis to test and develop various components. Improvements have resulted in increasing the unit capacity from 60,000 gpd to more than 100,000 gpd.

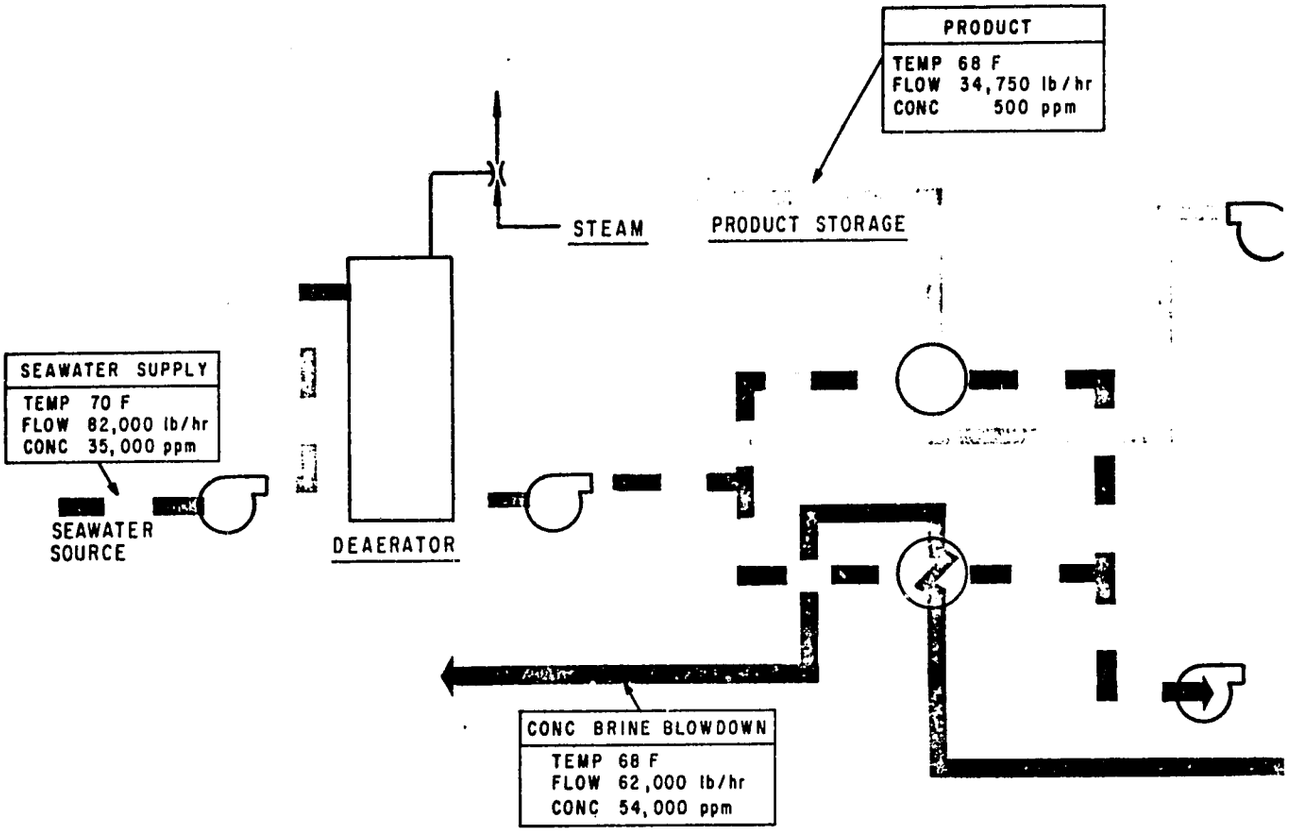
A model of a railroad-flatcar mounted, 100,000-gpd unit is shown in Figure V-15. This unit is self-contained and requires very little field erection. Its operating costs (depending upon the relative price of fuel and electrical power) may be less than those of distillation processes.

1. Plant Description

The vacuum freezing-vapor compression pilot plant is located at the OSW Wrightsville Beach test station. This test module receives its seawater from the test station seawater supply, and utilizes the maintenance and laboratory facilities and other services of the OSW test station. The pilot plant itself consists of two large vessels and the process auxiliaries. These vessels are the hydroconverter unit and the counterwasher. The hydroconverter unit contains a freezing zone, a melting zone, a vapor compressor and a heat removal coil. The counterwasher unit is used for washing the ice crystals which have been removed from the freezing zone in a brine slurry. The washed crystals from the counterwasher are returned to the melting zone of the hydroconverter. Auxiliaries include the refrigeration system, feed heat exchangers and air removal system.

2. Process Description

A simplified process flow diagram of the vacuum freezing-vapor compression plant is shown in Figure V-16. Seawater is supplied to the plant through a filter and deaerator; no chemical treatment of the seawater is required. Deaeration is particularly important because of the high vacuum maintained in the vessels. The deaerated seawater is cooled by exchanging heat with the freshwater product and the concentrated brine blowdown, and then is passed to the hydroconverter. The hydroconverter operates at a pressure of less than 0.1 psia. A portion of the seawater and brine flashes due to the low pressure in the vessel, thereby cooling to


PRODUCT
TEMP 68 F
FLOW 34,750 lb/hr
CONC 500 ppm
STEAM
PRODUCT STORAGE
SEAWATER SUPPLY
TEMP 70 F
FLOW 82,000 lb/hr
CONC 35,000 ppm
SEAWATER SOURCE
DEAERATOR
CONC BRINE BLOWDOWN
TEMP 68 F
FLOW 62,000 lb/hr
CONC 54,000 ppm

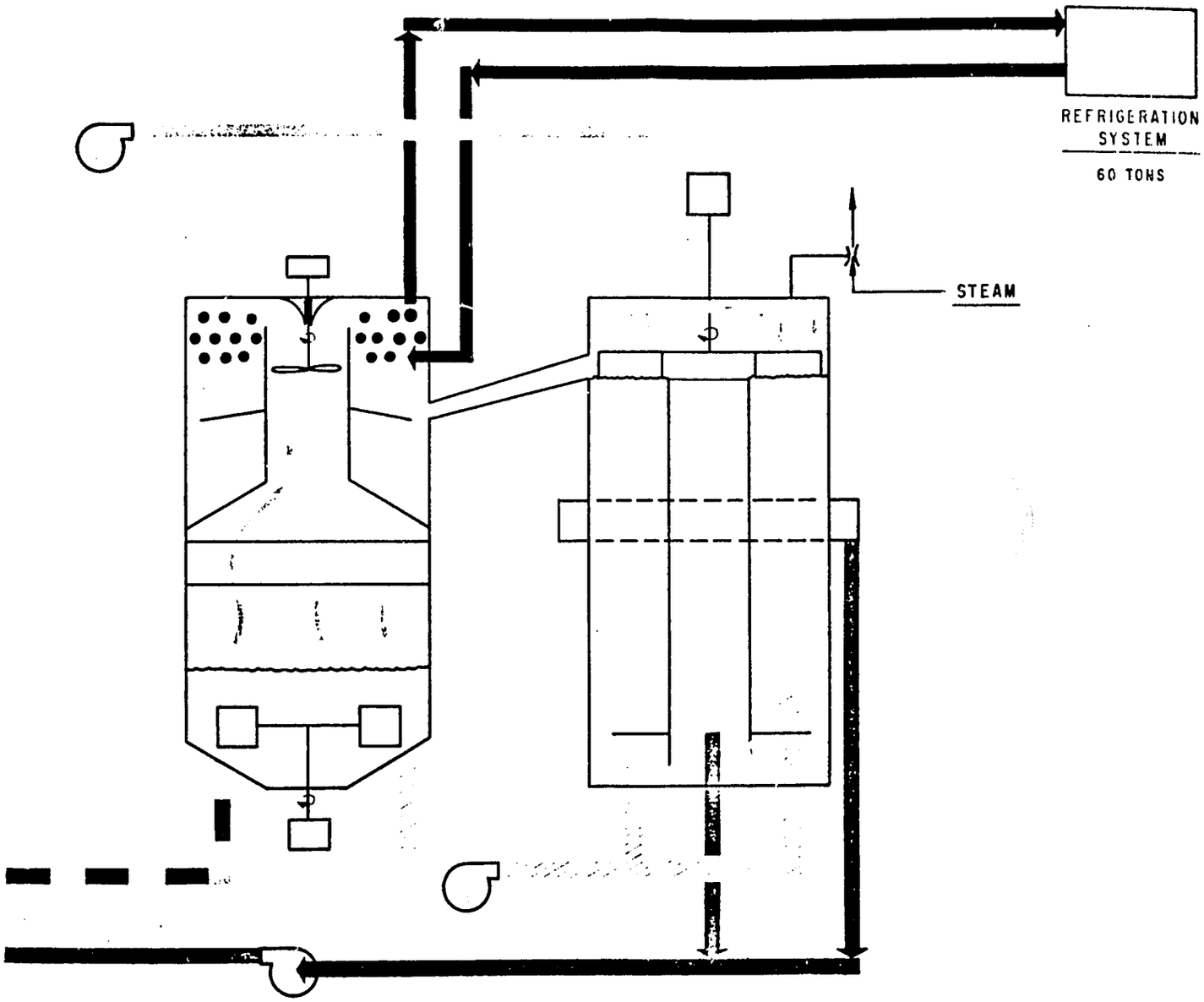


FIGURE V-16

VACUUM FREEZING - VAPOR COMPRESSION PLANT

OSW PILOT PLANT

WRIGHTSVILLE BEACH N.C.

freezing temperatures. To limit the size of the ice crystals which are forming, and to prevent the formation of an ice crust on the brine, a mechanical agitator is rotated in the freezing brine mixture. The vapor generated in the flashing process is compressed in the centrifugal compressor at the top of the hydroconverter and then directed to the melting zone of the hydroconverter. The brine and ice crystal slurry in the hydroconverter is pumped to the counterwasher where the ice crystals, which are lighter than the brine, rise to the top of the washing zone. A small amount of product water flows countercurrent to the rising ice crystals. Ice crystals which have risen through the washing zone are collected by a scraper and conveyed to the melting zone of the hydroconverter. Direct contact between the ice crystals and the compressed vapor in the melting zone results in the ice melting and the vapor condensing to form product water. The compressed vapor contains more latent heat than that required to melt the ice crystals, and this excess heat and in-leakage heat from the atmosphere is removed by a small refrigeration system.

3. Special Features

Because of the low-temperature operation of the freezing process, scaling is not encountered. Furthermore, corrosion is much less severe and the vessels and internal parts of the hydroconverter and counterwasher are constructed of carbon steel. The compressor used in the vacuum freezing-vapor compression process is of an unusual design using flexible blades, and is a high-volume, relatively high-performance unit. Heat in-leakage from the atmosphere is reduced by insulation of the exterior of the vessels.

Heat losses from the product streams are minimized by an exchange of heat between the incoming seawater and the product water and brine blowdown. These heat-recovery exchangers are of a special design utilizing stacked plates, rather than the conventional shell-and-tube design.

4. Problems Encountered

Many problems have been encountered in the development of the freezing process. These are control of ice crystal size, adequate washing of ice crystals without loss of a large amount of product water in the washing process, conveyance of ice slurry through the process and efficient compressor performance. All

of these difficulties have been overcome in the installation at Wrightsville Beach, and the unit capacity and performance have been good. While corrosion of the vessel interior is not a problem, condensation from the atmosphere tends to occur on the exterior of the vessels, piping and equipment and, in oceanside locations, this condensation and the salt sprays can cause corrosion and maintenance problems.

5. Present Use

The pilot plant at Wrightsville Beach is being used to test developments and modifications to the vacuum freezing-vapor compression process. In future studies it is planned to obtain long-term operating data to evaluate the reliability of the process.

6. Plant Staff

Since no freezing plants are in commercial operation, there is insufficient experience to assess staff requirements.

Additional discussion of the freezing process may be found in Subsection IV. B. 2., and in the OSW Annual Saline Water Conversion Reports for 1964 and 1965.

D. Membrane Process

It is possible to construct semipermeable membranes which have the characteristic of selectively passing certain types of ions while not passing other types. One family of semipermeable membranes which selectively permit either cation or anion passage forms the basis for the electrodialysis desalting process. In this process, ions carry electrical energy through the saline solution to electrodes situated at either end of the membrane assembly; this electrical energy transports the ions to the electrodes.

In a second family of membranes, the phenomenon of water migration through the crystal lattice of the membrane is utilized for desalting purposes; these membranes permit water to pass through, but exclude virus, bacteria and ionic impurities. In this process, a pressure of the solutions on either side of the membrane is applied across the membrane as a driving force; the process is therefore known as reverse osmosis.

1. Electrodialysis

The plant selected as representative of the application of the electrodialysis process is the OSW Demonstration Plant No. 3 located at Webster, South Dakota. This plant reduces the salinity of the brackish well-water supply from approximately 1,500 ppm to less than 500 ppm. The plant has the following characteristics:

Capacity	250,000 gpd
Performance	6.5 kwh/1,000 gallons
Source of seawater	Brackish water from 5 city wells
Source of heat	Purchased power

The startup of the Webster plant took place on March 5, 1962 and since that time it has supplied the water needs of the City of Webster. The construction cost of the plant was $486,000, and the actual water production cost for a period of normal operation in this plant today is approximately $1.20 per 1,000 gallons. A comparison of the Webster plant with a plant installed at Buckeye, Arizona is shown in Table V-2.

a. Plant Description

The OSW demonstration plant located at Webster, South Dakota is situated in a one-story building, and consists of four electrodialysis stages operating in series; it desalts water from a brackish well-water supply. There are 216 membrane pairs in each stage, and the membranes are 44 inches square. The membranes are separated by 0.030 inch, using monofilament separators to maintain the flow space between them. The membranes are assembled in a vertical position in a conventional plate and frame filter press. In addition to the electrodialysis unit, the demonstration plant building contains an office space, a chemical laboratory, a workshop and storage space.

TABLE V-2

COMPARISON OF WEBSTER, SOUTH DAKOTA AND BUCKEYE, ARIZONA ELECTRODIALYSIS PLANTS

Item	Buckeye Plant	Webster Plant
Designed Installed Capacity (gpd)	650,000	250,000
Capital Cost	$297,741	$433,470
Electrodialysis Stages (total)	6, in 3 Parallel Groups; 2 Stages in Series in Each Group	4, in Series
Total Effective Membrane Area--All Stages (sq ft)	11,500	18,000
Raw Water Composition	Essentially NaCl; No Iron or Manganese	See Table V-3
Raw Water Temperature (average)	86 F	48F
Pretreatment Required	None, Other Than Polishing Filters	Fe/Mn removal; Polishing Filters
Raw Water Total Dissolved Solids (ppm)	2,076-2,140	1,375-1,700
Product Water Total Dissolved Solids (ppm)	500	350-400; Lower Total Dissolved Solids Required to Reduce Hardness

TABLE V-2
(Cont'd)

Item	Buckeye Plant		Webster Plant	
Waste Water Total Dissolved Solids (ppm)	6,900		4,000	
Chemical Consumption (lb/1,000 gal.):				
(1) 66° Be Sulfuric Acid	0.6		2.8	
(2) Potassium Permanganate	None		0.04	
Normal Operating Current Densities (ma/cm^2)	ma/cm^2	amp/sq ft	ma/sq cm	amp/sq ft
Stage 1	19.0	17	3.9	3.5
Stage 2	10.0	9	2.7	2.4
Stage 3			1.8	1.6
Stage 4			1.1	1.0
Electrical Power Consumption (process only) kwh/1,000 gal.	12.0		6.5	

Source: 1963 Saline Water Conversion Report - OSW

TABLE V-3

ANALYSIS OF WATER (PARTS PER MILLION) FROM WELLS AT WEBSTER, SOUTH DAKOTA

	Well No.	
	3	4
Total dissolved solids	1,384	1,387
pH	7.55	7.62
Calculated hardness ($CaCO_3$)	787.4	788.2
Iron (Fe)	1.0	1.4
Potassium (K)	9.85	9.23
Sodium (Na)	104.0	102.1
Calcium (Ca)	187.4	193.0
Magnesium (Mg)	76.79	73.99
Manganese (Mn)	.90	.92
Fluoride (F)	.65	.68
Chloride (Cl)	9.50	7.00
Sulfate (SO_4)	724.4	765.00
Bicarbonate (HCO_3)	298.9	297.60
Specific Conductance, micromho/cm	1,627	1,626
Average temperature	48 F	48 F

Source: 1963 Saline Water Conversion Report - OSW

Recently a brackish water test facility has been added to the plant. A photograph of the Webster, South Dakota demonstration plant is shown in Figure V-17.

b. Process Description

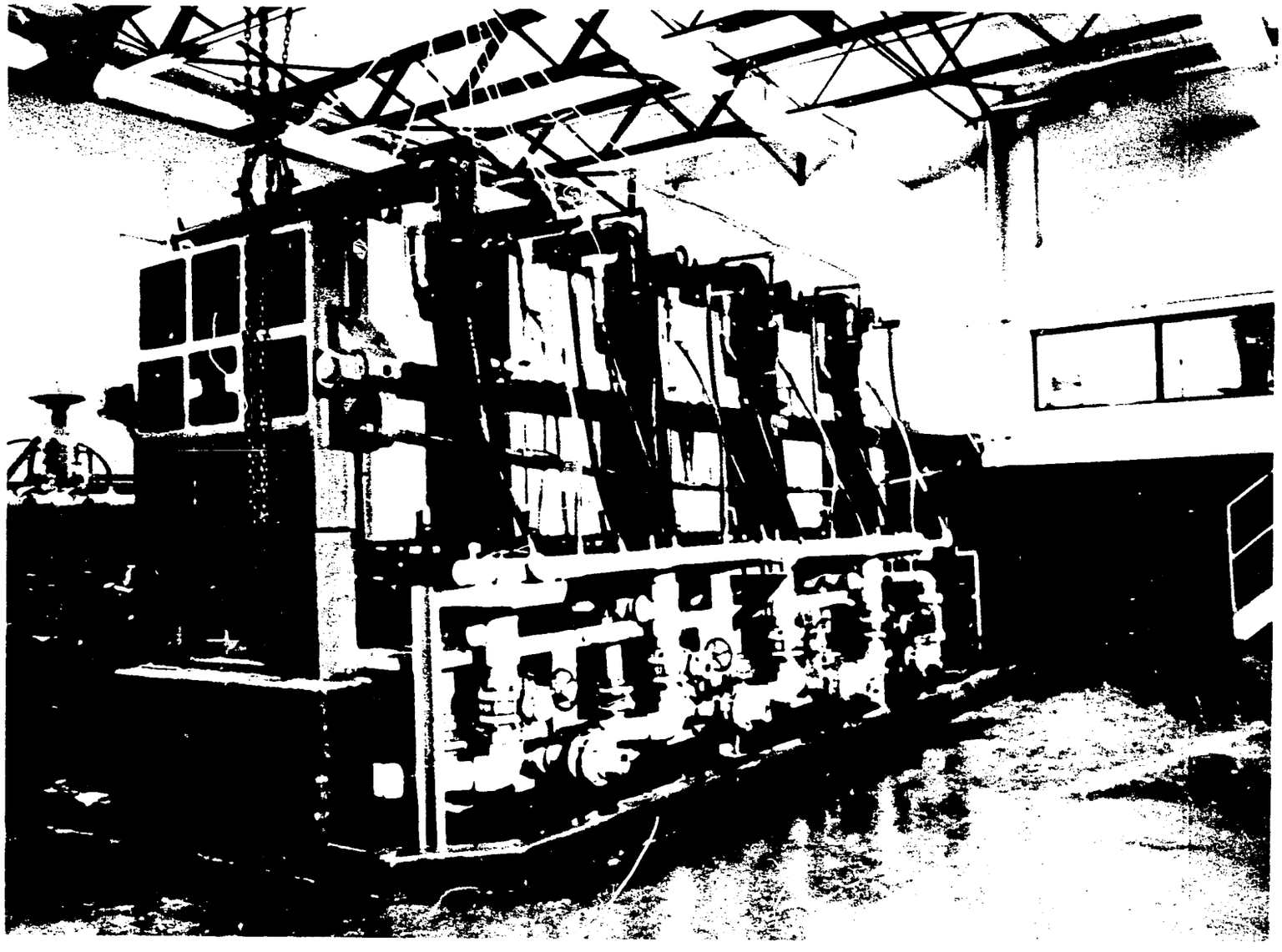
A simplified process flow diagram of the Webster plant is shown on Figure V-18. The brackish water feed is pretreated by ion exchange to remove constituents detrimental to electrodialysis. In this case, the constituents removed are iron and manganese; these occur in 1 to 2 ppm concentrations, and must be reduced to less than 0.1 ppm to avoid the formation of iron hydroxide scale on the membranes. After the water has been pretreated it is separated into two parts--one part is the dilute stream which passes successively through alternate passages of Stacks No. 1 through No. 4, and the other part becomes the concentrate stream makeup. In each stack, approximately 40% of the salt ions pass from the dilute stream through the membranes to the concentrated brine stream. The concentrate stream portion of the treated water joins the recirculating concentrate stream and passes in parallel through alternate spaces between the membranes where it receives the ions which have been transported through the membranes, and thereby becomes more concentrated. A portion of the concentrate stream is diverted to the waste tank as blowdown and thence to a wastewater well. The product water emerging from the last electrodialysis stage is pumped to a clear well and thence to the city water system.

c. Special Features

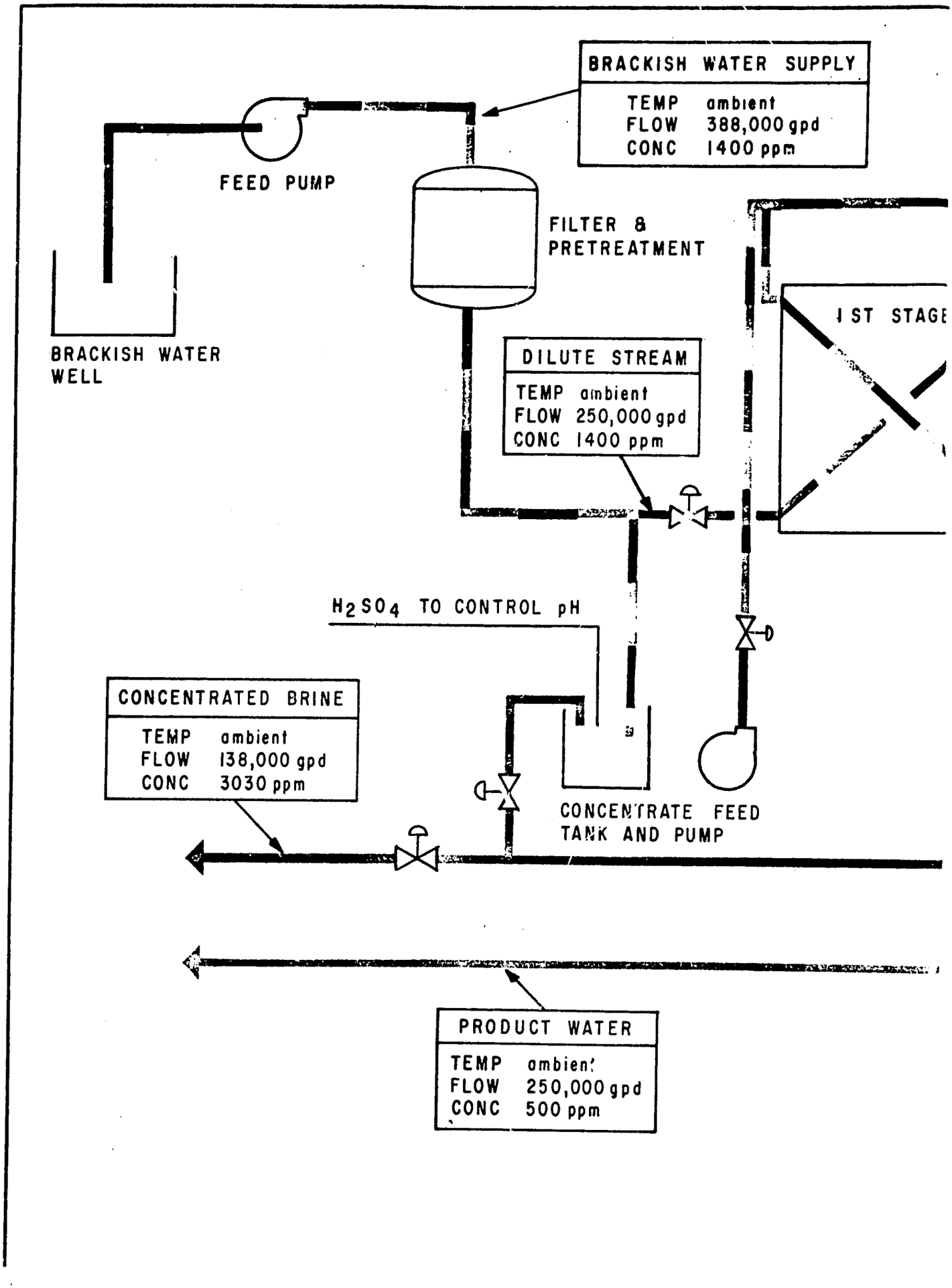
Occasionally brackish water facilities will require special pretreatment of the brackish water supply to permit satisfactory operation of electrodialysis. The brackish well-water at Webster, South Dakota contains an unusually high amount of iron and manganese, which is detrimental to performance of electrodialysis plants. This facility therefore contains an ion-exchange pretreatment system for removal of the iron and manganese.

FIGURE V-17

OSW DEMONSTRATION PLANT NO. 3--WEBSTER, SOUTH DAKOTA



Source: U. S. Department of the Interior,
Office of Saline Water


BRACKISH WATER SUPPLY
TEMP ambient
FLOW 388,000 gpd
CONC 1400 ppm
FEED PUMP
FILTER & PRETREATMENT
BRACKISH WATER WELL
1 ST STAGE
DILUTE STREAM
TEMP ambient
FLOW 250,000 gpd
CONC 1400 ppm
H_2SO_4 TO CONTROL pH
CONCENTRATED BRINE
TEMP ambient
FLOW 138,000 gpd
CONC 3030 ppm
CONCENTRATE FEED TANK AND PUMP
PRODUCT WATER
TEMP ambient
FLOW 250,000 gpd
CONC 500 ppm

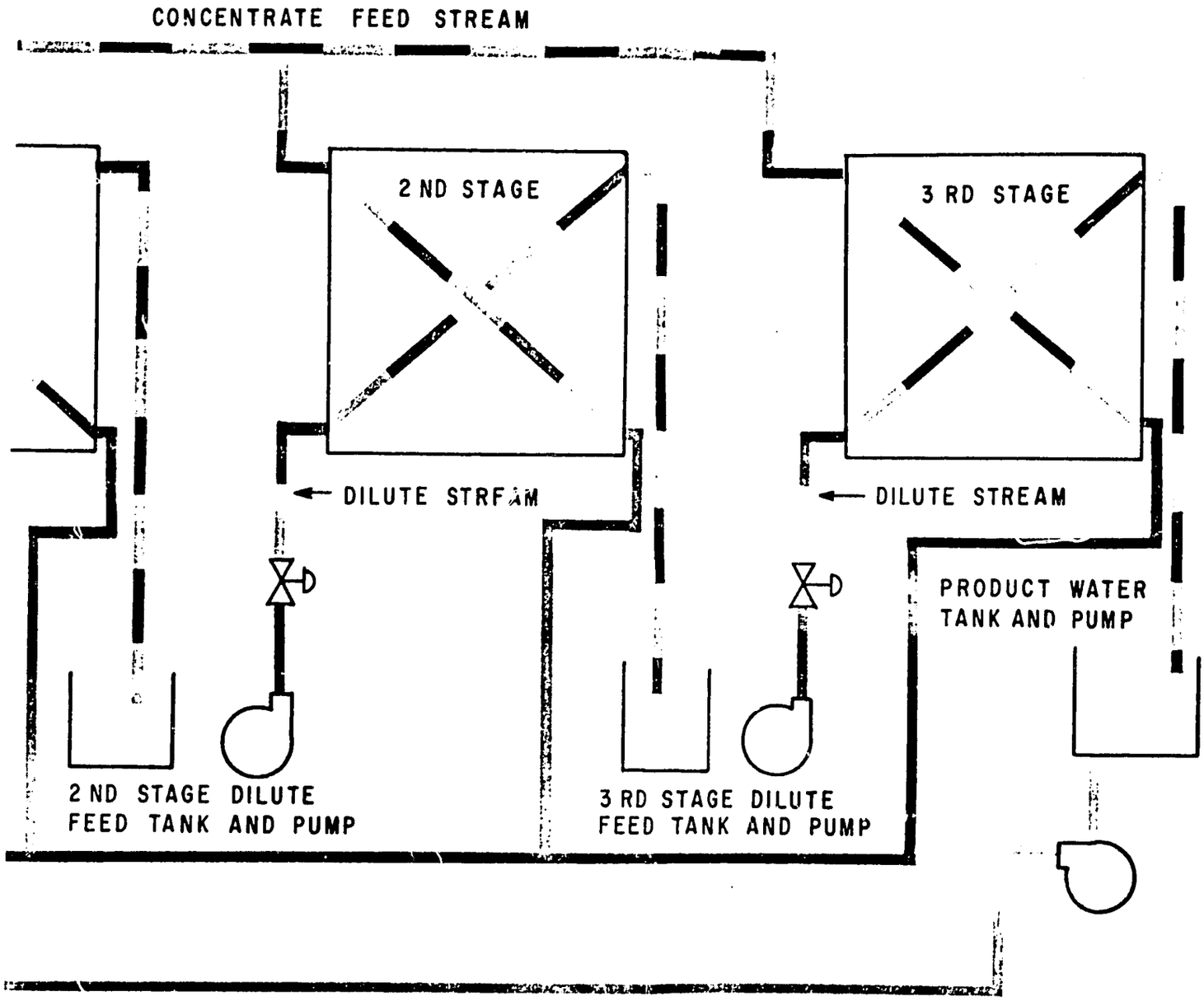


FIGURE V-18

ELECTRODIALYSIS PROCESS

OSW DEMONSTRATION PLANT NO. 3

WEBSTER, SOUTH DAKOTA

d. Problems Encountered

Five wells were drilled to obtain brackish water for the Webster, South Dakota plant. The chemical analyses of the waters obtained from these wells have large variations--Well No. 5, in particular, has a high content of calcium and a higher salinity than the other wells. Pretreatment of the feedwater was found to be inadequate for removing the iron and manganese and for controlling the calcium during the first several years of this plant's operation; therefore, iron hydroxide and calcium carbonate scales were deposited on the cell membranes and caused frequent outages. Cell membranes were damaged by uneven pressure distribution resulting from the scaling, and membrane replacement costs, therefore, have been higher than would normally be expected. When pretreatment is adequate, scale does not form on the membranes and high maintenance costs are not encountered. The electrodialysis plants at Buckeye, Arizona and Port Mansfield, Texas operate satisfactorily.

e. Present Use

Since its startup, the Webster, South Dakota plant has been the primary source of fresh water for the City of Webster. Occasionally, improvements and modifications have been made to reduce the outages due to inadequate pretreatment and to increase the capacity of the plant. During 1963, the plant operated approximately 90% of the time, and produced approximately 80,000,000 gallons of potable water. Currently, modifications are being made to increase the plant capacity from 250,000 to 325,000 gpd.

f. Plant Staff

The staff of the Webster plant is as follows:

Plant Manager

Staff Engineer

Clerk/Accountant

3 Operators/Maintenance

Additional information on electrodialysis may be found in Subsection IV. C, and in the following reports:

OSW Report No. 101	First Annual Report, Saline Water Conversion Demonstration Plant, Webster, South Dakota, Mason-Rust, Order No. PB 181680
OSW Report No. 132	Second Annual Report, Saline Water Conversion Demonstration Plant Webster, South Dakota. Mason-Rust
OSW Report No. 164	Third Annual Report, Brackish Water Conversion Demonstration Plant Webster, South Dakota. Mason-Rust

2. Reverse Osmosis

Although the reverse osmosis process has not been utilized in commercial applications to date, the industry is prepared to quote on reverse osmosis units of small-scale for processing brackish water. Pilot plants of 10,000-gpd and a 50,000-gpd capacity are being constructed for the OSW. A 40,000-gpd unit is being offered by one manufacturer, and has been chosen as representative of this process. This plant has the following characteristics:

Capacity	Up to 40,000 gpd (on brackish water)
Performance	Approximately 7.5 kwh/1,000 gallons
Source of water	Brackish
Source of heat	Purchased power

a. Plant Description

The reverse osmosis unit utilizes membranes and separator plates arranged horizontally in a vertical stack, and includes polishing filters, a high pressure pump and instrumentation

The membranes are cast from solvent solutions of cellulose acetate containing additions to give the membrane its desalting capabilities. This unit contains approximately 1,840 square feet of effective membrane surface area. Plastic separators are utilized to support the membranes against the pressure differential of approximately 750 psi between the saltwater side and the freshwater side. The stack is arranged so that the brackish water flows through several stack elements in series; the required velocity over the brackish water side of the membranes is thus attained, and the brackish water is exposed to sufficient surface to obtain the required brine blowdown concentration. Product water, from each of the stack stages, is collected in a central portion of the vertical stack. The stack consists of an assembly of plastic separator plates and membranes held in a vertical filter press; this press is constructed of steel which is coated with plastic on the saltwater side.

Figures V-19 and V-20 show photographs of various units and a conceptual design of a 1-mgd unit.

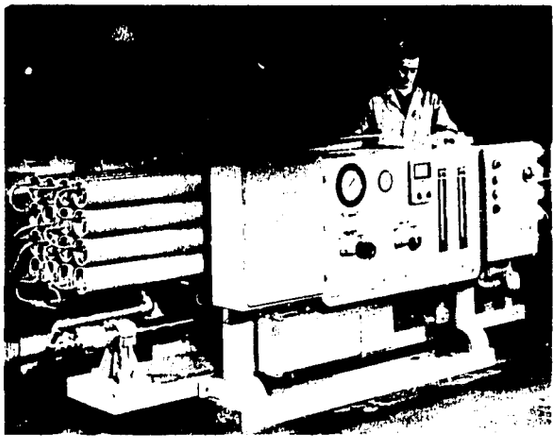
b. Process Description

A simplified process flow diagram of this reverse osmosis process unit is shown in Figure V-21. Brackish water is passed through a filter to remove suspended particles, acid treated for pH control, raised in pressure to approximately 750 psi and distributed to the high pressure side of the reverse osmosis membranes. No heating or cooling is required unless the feedwater temperature is below 60 F. As the brackish water passes through a unit, approximately 80% of the feed permeates the membranes as fresh water; this is typical for brackish water applications. The brine, now concentrated by a factor of 5, is rejected from the system. Typically, a gross energy input of about 7.5 kwh per 1,000 gallons is required for brackish water applications. In very large installations, it may be economical to recover energy from the high pressure brine stream.

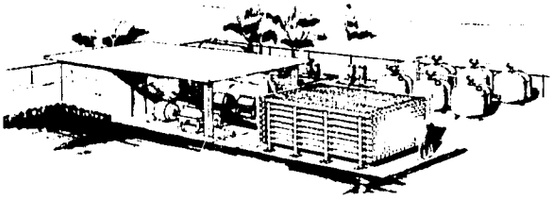
This unit will operate on seawater for a limited time only, but improvements in the membrane life and characteristics may permit reverse osmosis units to be used on seawater applications at a later date. Should this occur, the pressure

FIGURE V-19

REVERSE OSMOSIS PLANTS



10,000-gpd Reverse Osmosis Unit



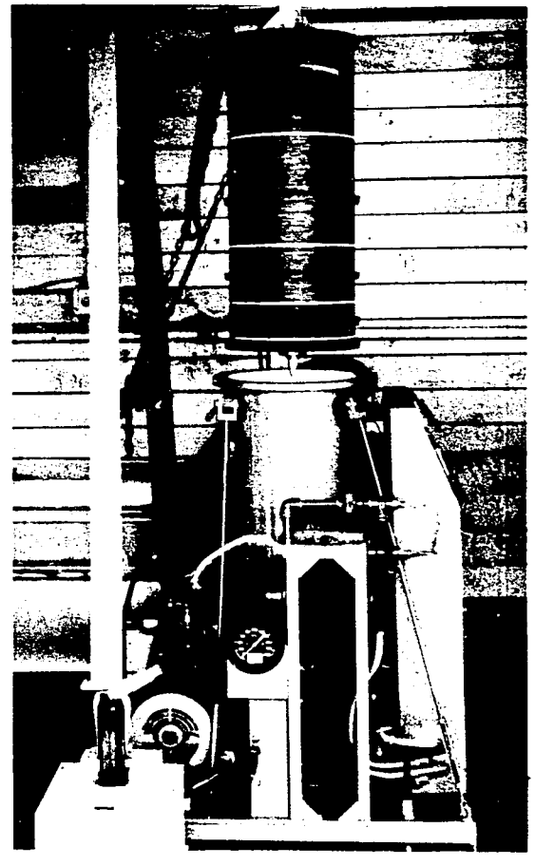
Perspective of 1-mgd Reverse Osmosis Plant

Courtesy: General Atomic Div of
General Dynamics Corp

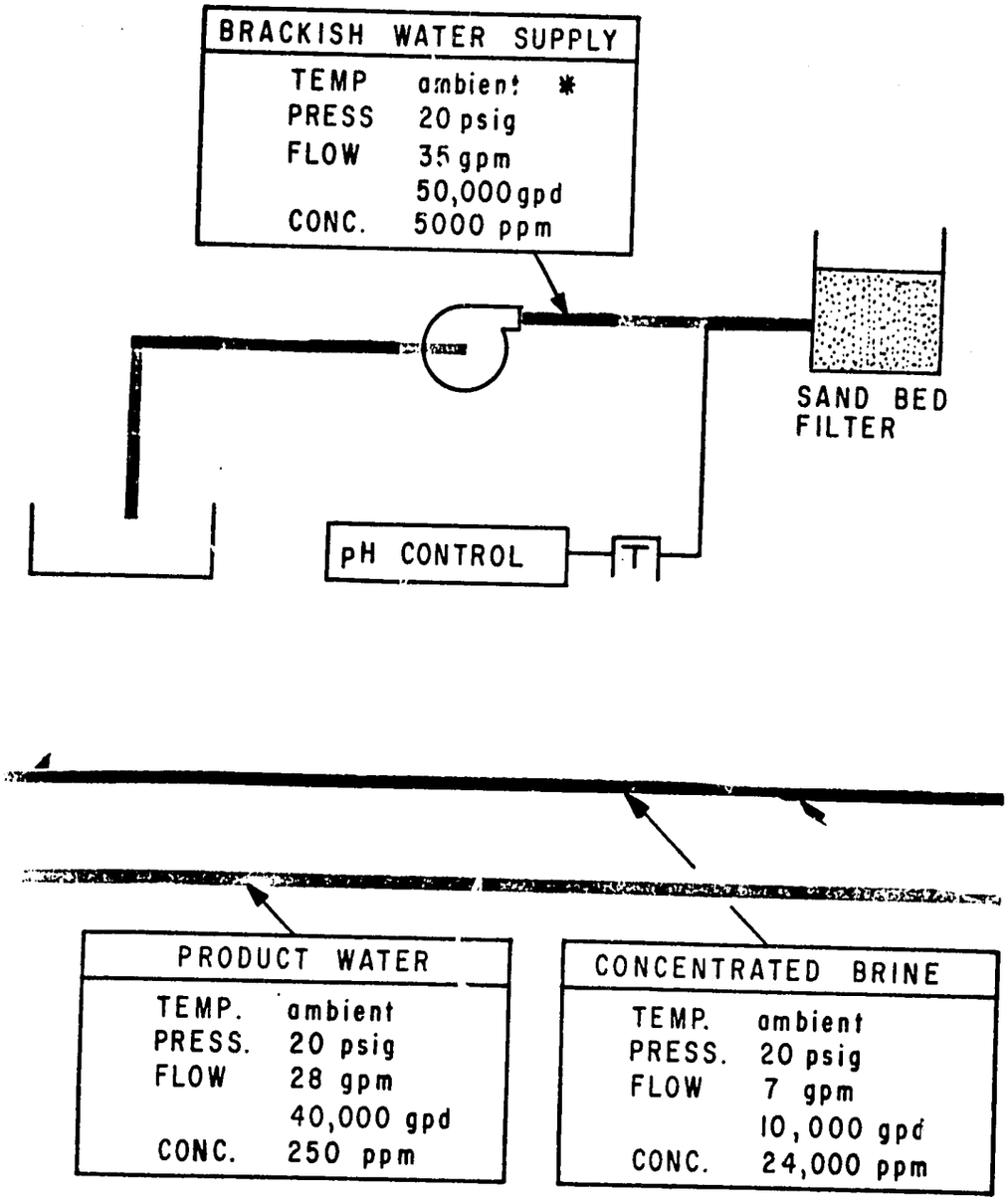
FIGURE V-20

REVERSE OSMOSIS UNITS

20,000-gpd Reverse Osmosis Unit



Membrane Plate Stack Being Inserted In Pressure Vessel



* NO HEATING OR COOLING REQUIRED IF WATER IS BETWEEN 60 F AND 100 F. OPTIMUM PERFORMANCE IS ACHIEVED BY OPERATING BETWEEN 70 F AND 80 F.

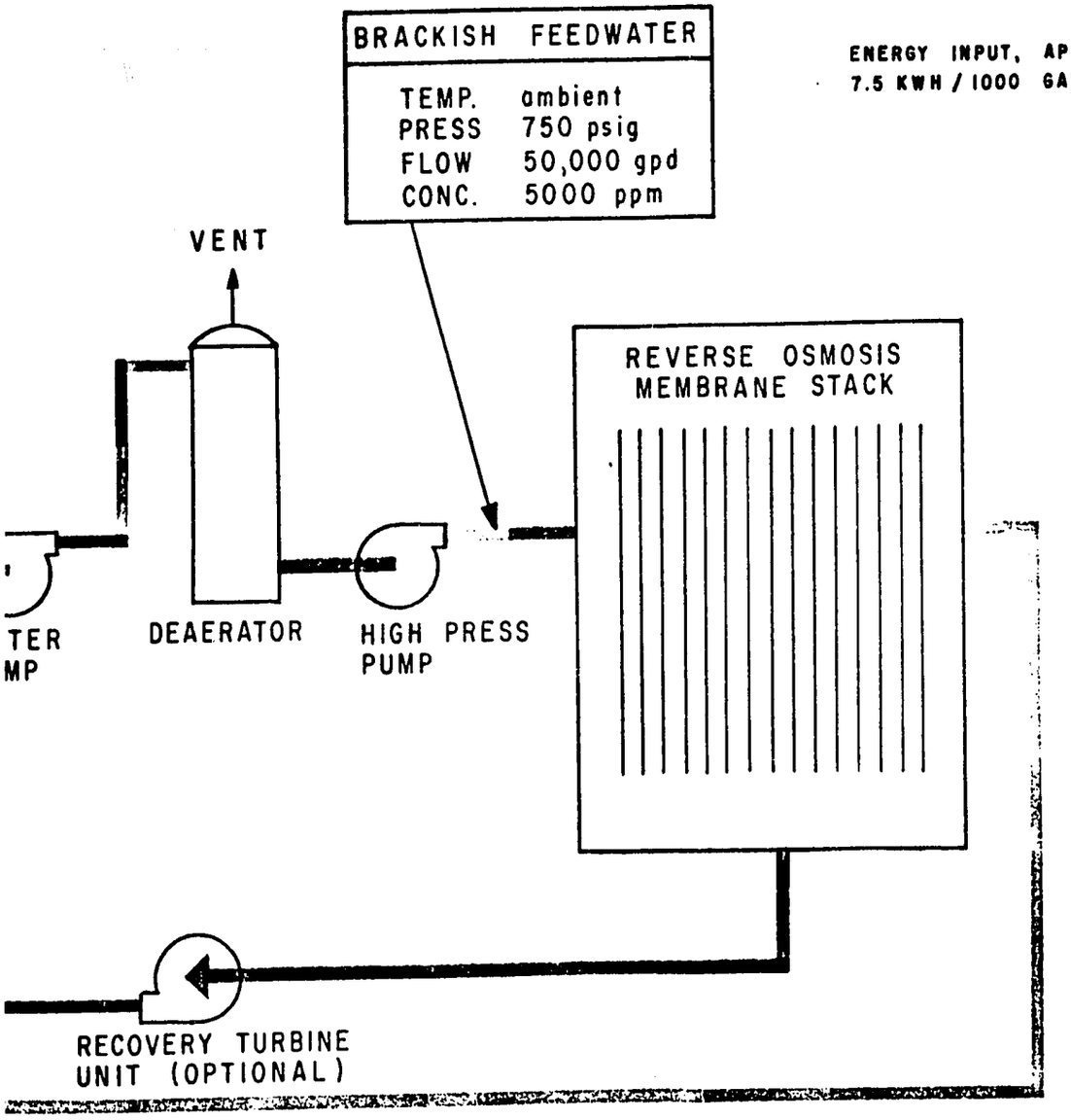


FIGURE V-21

REVERSE OSMOSIS

20,000- TO 40,000-GPD PILOT PLANT

required for operation will be approximately 1,500 psi, the product to feed ratio will be approximately 1:3 and the gross electrical input will be approximately 18.6 kwh per 1,000 gallons. In large seawater installations, energy recovery from the high-pressure brine stream may reduce the energy requirements by a factor of 2.

c. Problems Encountered

The most significant problem encountered in reverse osmosis units is membrane life. With seawater, this life is seldom more than a few months, partly due to the effect of high pressure on the membrane characteristics. At present, even the brackish water applications have not demonstrated a membrane life of more than 4 to 6 months. The cost of membrane replacement and the time required to make this replacement are the most serious problems associated with the present reverse osmosis units. In addition, to avoid plugging of the membrane with suspended material, a very good feedwater filtration system is required. Mechanical maintenance of these units has not been a problem except for membrane replacement.

d. Present Use

Until more data are available, the units now being offered should be considered for special applications or for testing purposes only. During the next one to two years, performance of these units may justify there use for commercial brackish and seawater desalination.

e. Plant Staff

Because no reverse osmosis units are in commercial operation, it is not possible to predict the staff requirements. It is reasonable to expect, however, that even large size units may operate efficiently unattended and that staff will be required for administrative and maintenance purposes only.

VI. DUAL-PURPOSE PLANT DISCUSSIONS

A. Dual-Purpose Electric Power Generation-Water Desalination Plants

1. General

Seawater and brackish water desalination plants require energy for operation. Electrical energy is required by all processes to operate pumps. Electrical energy is also required to operate compressors for the freezing and crystal hydrate processes and the vapor compression distillation process. Other distillation processes, however, require heat--usually in the form of steam at about 260 F to 280 F.

Where energy in the form of heat is required, it may be obtained by including a boiler in the desalination plant facility or by extracting heat from a nearby process plant. The electrical energy may be obtained from an existing electrical distribution system or may be generated at the desalination facility. This possibility of on-site generation of electricity suggests that there may be advantages in combining the desalination plant facility with an electric power plant, to provide low cost, low-pressure steam and electrical energy. This type of facility is designated a "dual-purpose electric power-water plant."

2. Processes Requiring Heat Energy

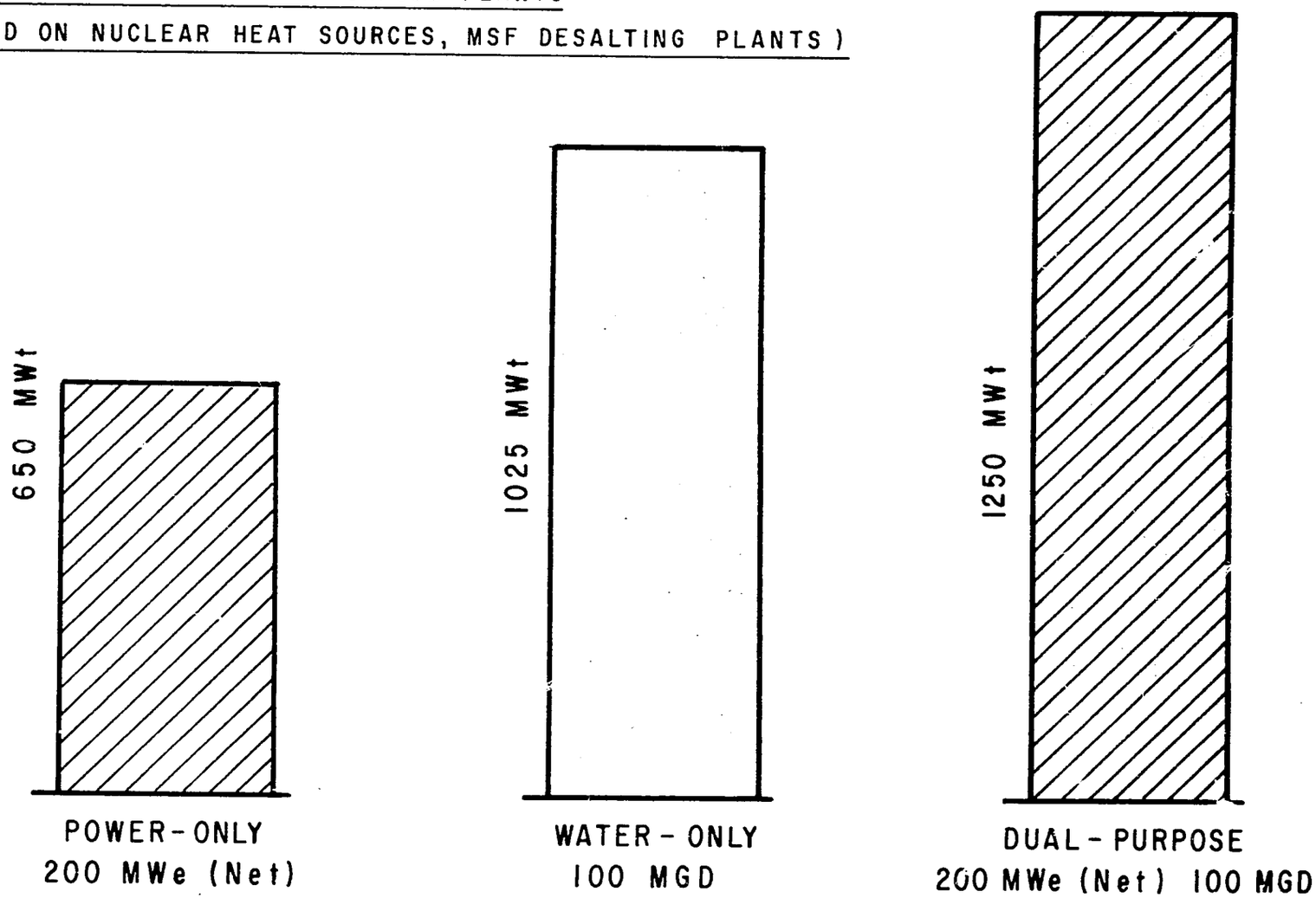
There are technical and economic advantages of using extraction or exhaust steam from a steam power plant as the heat source for a seawater or brackish water distillation facility (1)(2)(3)(4). These dual-purpose plants, which produce electric power and fresh water, can readily be shown to have lower capital costs and lower energy consumption than two separate plants for power generation and saltwater or brackish water distillation.. Figure VI-1 depicts the relative thermal energy needs for a large nuclear power-only, a large water-only and a large dual-purpose operation; it illustrates the savings in thermal energy that can be achieved by combining these two operations. For the case shown, the sum of the thermal energies required for a single-purpose power plant and a single-purpose desalination plant are approximately 40% greater than the thermal energy requirement for a combined operation of these two plants. For fossil-fueled plants, the thermal energy requirement for single-purpose plants would be approximately 30% greater than the thermal energy requirement for a combined operation.

FIGURE-VI-1

COMPARISON OF ENERGY REQUIREMENTS

SINGLE-PURPOSE VS DUAL-PURPOSE PLANTS

(BASED ON NUCLEAR HEAT SOURCES, MSF DESALTING PLANTS)



a. Power/Water Ratios

There are a large number of possible steam power plant cycles which are suitable for use in a dual-purpose plant (4). These are all variations of three basic cycles: (1) backpressure or noncondensing, (2) condensing-backpressure and (3) condensing-extraction. In discussing these three steam cycles it is useful to introduce the concept of the "Power/Water Ratio," i.e., the ratio of the net power production capability in megawatts electrical (MWe) to the water production capacity in millions of gallons per day (mgd). Thus for the dual-purpose plant shown in Figure VI-1, the Power/Water ratio is 200 MWe/100 mgd, or 2.0 watts per gallon per day. This is a typical Power/Water ratio for a nuclear-fueled dual-purpose plant using a simple backpressure or noncondensing turbine cycle.

The Power/Water ratio is one of the most important considerations in establishing the feasibility of a dual-purpose electric power-water plant. There must be a market for the power which is generated in the dual-purpose plant, just as there must be a market for the water which is produced in such a plant. Often the power customers and water customers are the same people and, when this is the case, the Power/Water ratio must be related to the Power/Water consumption of this community. In the United States the ratio of the consumptions of power and water is approximately 10 watts per gallon per day. In many less industrialized areas of the world, however, this Power/Water ratio is much lower.

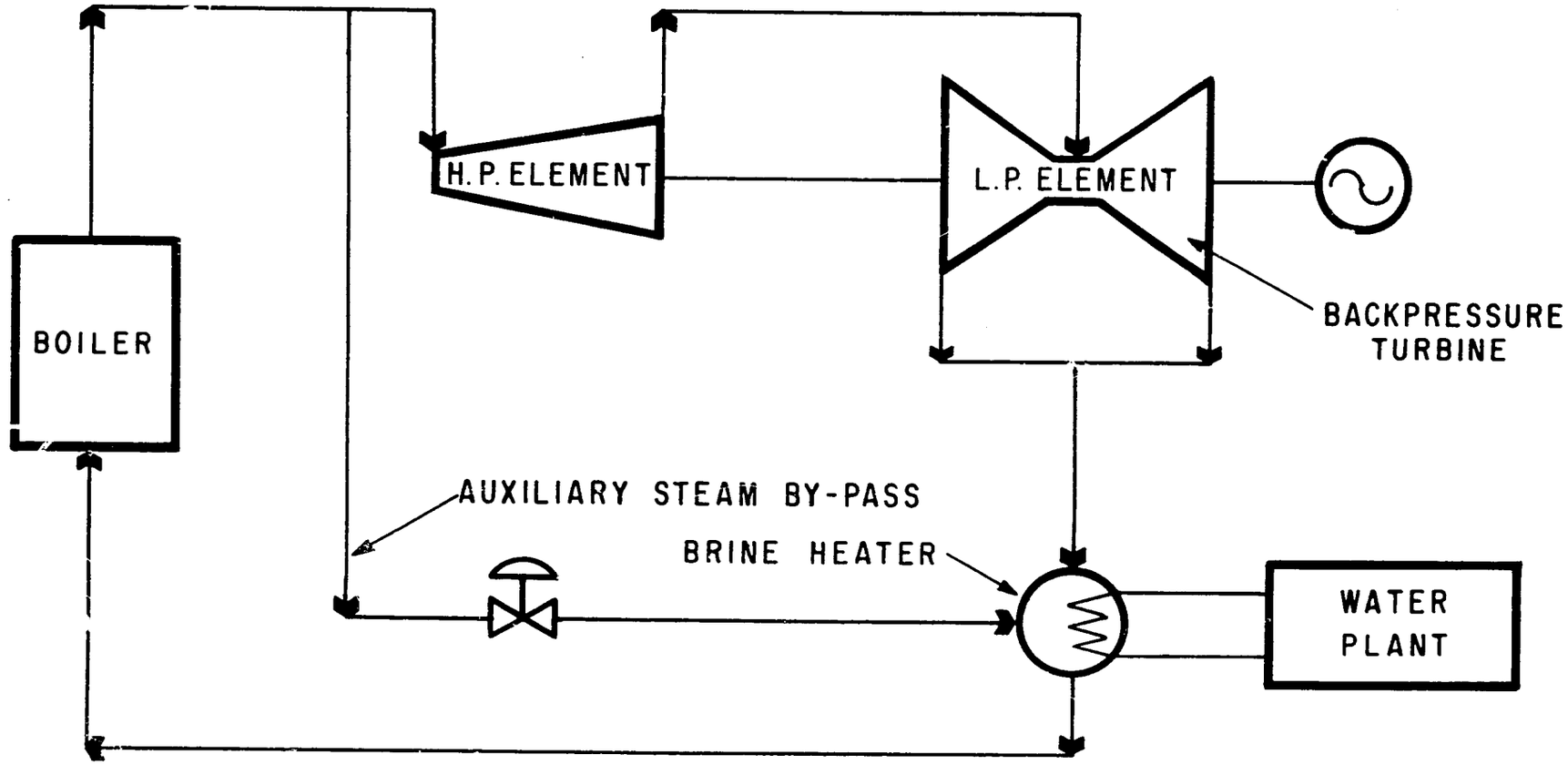
The type of power plant cycle which is appropriate for the dual-purpose plant operation is determined primarily by the required Power/Water ratio. The backpressure or noncondensing cycle is appropriate for the lowest Power/Water ratios and the condensing-extraction cycle is appropriate for the highest Power/Water ratios.

b. Power Plant Cycles

The backpressure cycle, as shown in Figure VI-2, is the simplest of the turbine cycles useful in dual-purpose plants. All of the power is generated in one backpressure or noncondensing turbine, and except for that used for feedwater heating, all of the steam leaving the turbine is used as the energy

FIGURE VI - 2

BACKPRESSURE CYCLE



HIGH BACKPRESSURE:

LOWER POWER PRODUCTION
HIGHER WATER PRODUCTION

LOW BACKPRESSURE:

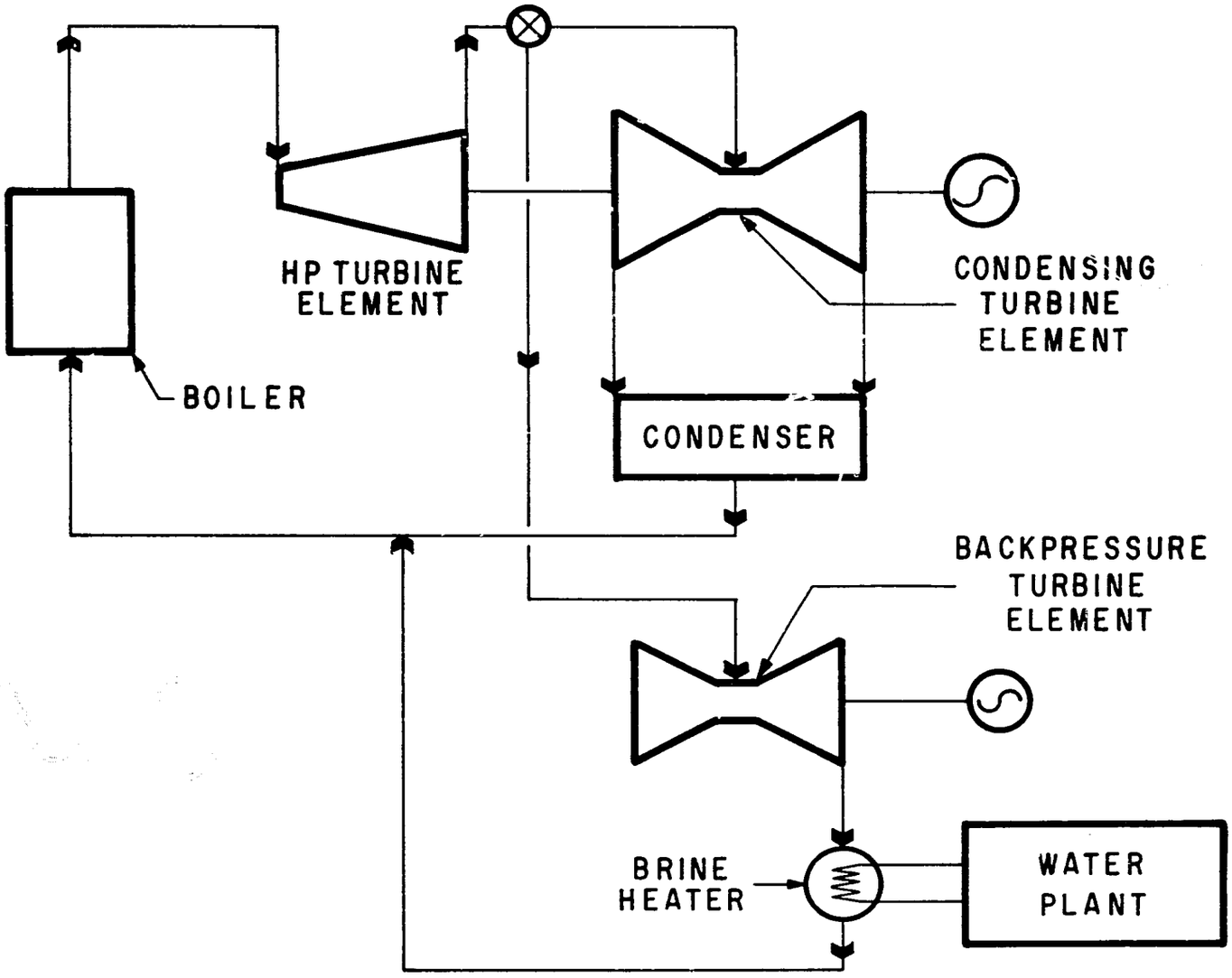
HIGHER POWER PRODUCTION
LOWER WATER PRODUCTION

source for the desalting plant. For the turbine cycles considered, the backpressure cycle also shows the greatest savings in energy consumption and capital cost compared to two separate single-purpose plants. The dual-purpose plants now operating at Guantanamo, Cuba and Eilat, Israel are examples of plants using this turbine cycle. The auxiliary steam bypass shown in Figure VI-2 serves three purposes: (1) it provides a means by which the turbine exhaust pressure can be controlled, (2) it provides a means by which the desalination plant can be operated at full capacity during periods of low power demand or turbine-generator outage and (3) it provides a means by which the heat input to the desalination plant can be varied for constant power generation--a feature which may be required for plants with a large seasonal variation in seawater temperature. For plants using a nuclear heat source, the simple backpressure cycle can be used for Power/Water ratios from about 1.5 to 4.0, as is planned for the Dual-Purpose Electric Power-Water Desalting Plant for Israel (6); the lower values provide high maximum brine temperatures and turbine exhaust pressures, while the higher values result in lower maximum brine temperatures. For plants using a fossil-fuel heat source, the backpressure cycle can be used for Power/Water ratios between about 2.0 and 8.0. When the required Power/Water ratio falls below the 1.5 for nuclear and 2.0 for fossil-fueled plants it is necessary to obtain at least some of the heat directly from the steam source; when the ratio exceeds 4.0 for nuclear and 8.0 for fossil-fueled plants, the condensing-backpressure cycle is required.

An example for the condensing-backpressure cycle is shown in Figure VI-3. In this cycle, all of the steam first passes through a single high-pressure turbine element, after which the steam flow splits. Part of the steam then passes through the condensing low-pressure turbine element to the condenser; the other part passes through the backpressure or noncondensing low-pressure turbine element from which it is exhausted to the brine heater of the desalination plant. The particular arrangement shown, using a separate generator for the backpressure turbine, was described in more detail in Reference 5; it permits water production to be maintained constant even when there are large electrical load swings, and can be designed to provide excellent power peaking and emergency power capability. Other arrangements of a condensing-backpressure cycle use separate

FIGURE VI-3

CONDENSING-BACKPRESSURE CYCLE



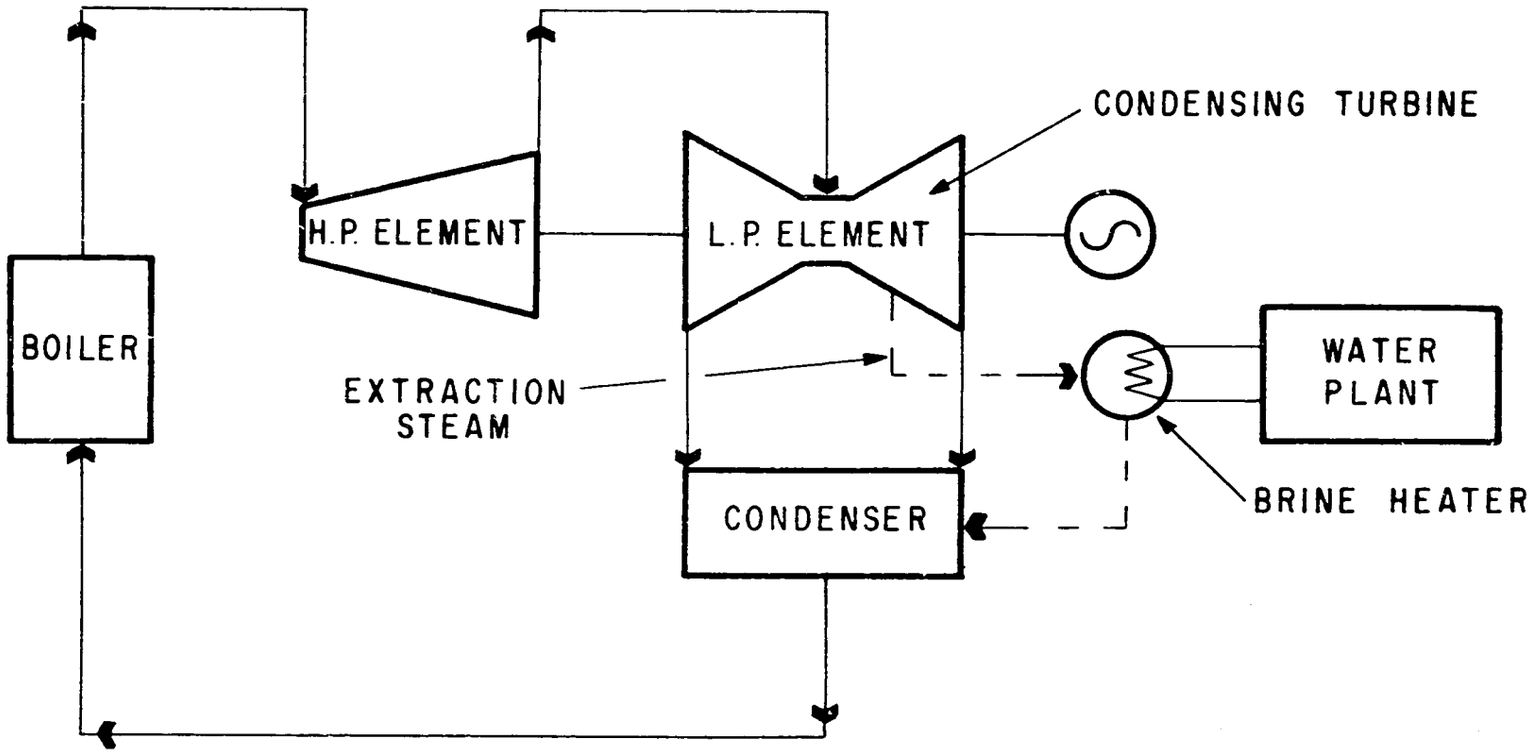
high-pressure turbines with the condensing and noncondensing turbines; the latter arrangement is the basis for the currently proposed dual-purpose plant for the Metropolitan Water District of Southern California(7). The backpressure-condensing cycle is most suitable for Power/Water ratios from lower limits of 4 for nuclear plants and 8 for fossil-fueled plants to an upper limit of about 25, at which point it becomes feasible to use a condensing-extraction cycle.

A condensing-extraction cycle, as shown in Figure VI-4, uses a single, condensing turbine for power generation, with steam extracted from an appropriate point in the turbine for use as the heat source in the desalination plant. Since turbine design limits the quantity of low-pressure steam which can be extracted from a single point, this cycle is limited to dual-purpose plants with low water production, i.e., a Power/Water ratio of 25 or greater.

Up to this point, the effect of the Power/Water ratio on turbine steam cycle selection has been discussed without reference to the manner in which the Power/Water ratio itself is established for a given dual-purpose plant. The lower Power/Water ratios, which are suited to the use of a backpressure cycle, are typical of plants which are designed to meet an incremental requirement for both power and fresh water; for example, the power generating capacity might represent the incremental load growth for two or three years and the desalting plant capacity represent the projected needs for fresh water for approximately the same period. The intermediate Power/Water ratios, for which the condensing-backpressure cycle is well suited, are usually utilized in plants for which the power capability is set by the incremental needs of a region for a few years, but the freshwater production capacity is set by supplemental or emergency needs. The plants with very high Power/Water ratios usually produce desalted water as a by-product for in-plant use or local needs, while meeting a regional need for power.

FIGURE VI - 4

CONDENSING - EXTRACTION CYCLE



c. Base Load Versus Peaking Operation

There is a substantial investment in the desalination plant, whether it is a single purpose plant or a portion of a dual-purpose electric power-water plant. The fixed costs, such as interest and depreciation associated with the investment in the water plant, is the largest cost item in determining the unit cost of water. The full-load continuous operation of the water plant spreads the very substantial fixed costs over the maximum production. While there will be scheduled shutdowns during the year for routine maintenance, inspection and repair of the desalination plant, this time should be kept at a minimum; during the time when the desalination plant is available for operation, it should be operated as near its maximum capacity as possible to reduce the unit water cost. Except for unusual situations, the water plant should operate as a base load plant. One unusual situation where other than base load is justified is on the Island of Guernsey. The damage to the local economy due to an extended drought is substantial, and a drought can be anticipated every 6 to 8 years. On Guernsey, it was found economical to install a desalination plant to supply water during this drought, in order to prevent the loss of the islands crops. The annual crop value (mostly greenhouse-grown tomatoes and flowers) is approximately $20 million. The desalination plant investment and occasional operation were found to be less expensive than the crop insurance and less expensive than the damage to the economy in the event of crop loss.

The operation of the electric power plant is related to the requirements of the desalination plant. If the desalination plant is base loaded, as will normally be the case, it is necessary to operate the electric power plant with a load sufficiently high to supply the steam requirements. If the full economy of the dual-purpose plant operations is to be realized in determining the feasibility of a dual-purpose power-water distillation plant, it is important to examine the economics of the total plant over its operating cycle, with particular regard to the capability of the electric power system served by the dual-purpose plant to absorb the required amount of electrical energy generated in the dual-purpose plant facility. This requires an operational analysis of the dual-purpose plant and of the electric power and water systems which it serves.

d. Allocation of Production Costs

The dual-purpose plant produces two commodities: electric power and product water. The determination of the capital costs and annual costs of the facility is not difficult, but allocation of the capital costs and annual costs to either product is arbitrary. This allocation is compounded by the differing accounting procedures of the electric power industry and the water industry. For example, power plants are usually amortized over a 30-year period, while water plant facilities are often amortized over a 40-year period and may be amortized over a period as long as 100 years. The interest on borrowed capital may also be different for the two types of facilities. These differences in accounting procedures make it very difficult to determine an appropriate allocation of costs between the two products of a dual-purpose plant.

One common method of allocating costs in a dual-purpose plant is to establish a value of electric power based on the cost of producing this power in the same quantities and load conditions in a single-purpose electric power plant. This value of electric power is then used in determining the credit for power sold during the year, and in applying this credit against the total annual costs of the dual-purpose electric power-water plant facility. The remaining annual costs of operating the dual-purpose plant facility are then designated "the annual cost of producing product water," and the unit water costs are determined by dividing this annual water cost by the annual quantity of water produced. The value for electric power used in this cost allocation has an important influence on the resulting unit cost of water. To avoid distortions in the unit water cost, it is very important to arrive at the value of power in a fair manner. Other methods of allocating production costs are described in Reference 13.

3. Processes Requiring Electrical Energy

The vapor compression distillation system, the freezing and hydrate processes, reverse osmosis and electrodialysis all primarily require electrical energy for their operation. The cooling water requirements of all of these processes is small and therefore, these plants are particularly suited to inland brackish water locations. Moreover, the vapor compression distillation and freezing processes are suitable for seawater as well as brackish water locations. The electric power requirements for the freezing, reverse osmosis and electrodialysis processes are directly related to the salinity of the saltwater source.

A dual-purpose plant facility consisting of an electric power plant and one of the desalination processes which require electrical energy does not have the advantages offered by the dual-purpose plants utilizing desalination processes which require heat. The thermal energy requirement of the dual-purpose plant is the same as the sum of the thermal energy requirements of the single-purpose electric power plant and single-purpose desalination plants of the same capacity. The advantages of a dual-purpose plant utilizing electrically operated desalination processes are restricted to the savings which may result from combining the operating staff, maintenance facilities and savings in electrical transmission cost between a single-purpose electric power plant and a single-purpose desalination plant. The allocation of costs between the production of electrical power and water is much simplified, because the desalination plant is essentially an electric power customer.

B. Recovery of Chemicals as By-Products from Desalination Plants

1. General

The sea has long been a source of chemicals for use by mankind, and various extraction methods have been employed. However, only three chemicals have been economically "mined" in any significant quantities: sodium chloride or common salt, magnesium compounds and bromine. Since a desalination plant processes a large volume of seawater and produces concentrated brine, it is logical to consider the combined installation of a desalination plant and a plant to recover the seawater chemicals as by-products. If an appropriate process to produce salable by-product chemicals from this brine is selected and developed, it could be used to reduce the cost of water and/or treat the seawater before it enters the desalination plant. This pretreatment would remove the calcium and magnesium salts and thus enable the desalination plant to operate more efficiently, with a higher concentration ratio and with less scaling problems.

Since appreciable benefits may be achieved by combining by-product recovery with a desalination plant, studies have been made and continuing development work is being performed to obtain feasible and economic processes to produce chemicals as by-products from desalination plants. One of the recent studies on this subject was performed for the Office of Saline Water (OSW) by American Cyanamid Co.(9) The resulting report provided information on the basic value of chemicals in standard seawater, the seawater elements that may be of commercial significance are shown in Table VI-1. Other elements found only in trace amounts in seawater would have less value.

TABLE VI-1

VALUE OF MINERALS IN SEAWATER

Element	Concentration in Seawater (ppm)	Value (¢/1,000 Gal. Seawater)	Mineral Reclaimed	Market Price (U.S.A., 1966 Basis)
Sodium	10,561	41	Crude Salt	$3.70/ton
		162	Refined Salt	$14.60/ton
Magnesium	1,272	53	95% MgO	$61/ton
Sulfur	884	47	Na_2SO_4	$28/ton
Bromine	65	12	Br_2	$0.22/lb
Potassium	380	10	KCl	$33/ton
Calcium	400	3.3	Quicklime	$14/ton
Boron	4.6	0.9	Borax	$47/ton
Strontium	13	0.6	$SrSO_4$	$56/ton

It should be noted that the chemicals that may be obtained from seawater have a relatively low value per unit weight. Therefore, the economics of recovering by-products are partly dependent on location, since transportation costs may be high in relation to the market value of the by-product. Furthermore, an economic analysis must include consideration of the potential market volume, since the potential production of some of the chemicals, particularly salt, may be large in relation to the demand in the area. The potential by-product production rates from a single 10-mgd desalination plant facility, are listed in Table VI-2(9).

TABLE VI-2

POTENTIAL BY-PRODUCT PRODUCTION

FROM 10-MGD SEAWATER DESALINATION PLANT

Material	Potential By-Products Production (tons/year)
Salt (NaCl) Refined	420,000 to 740,000
Magnesia (MgO)	32,000 to 58,000
Quicklime (CaO)	8,600 to 15,000
Potash (as KCl)	11,000 to 20,000
Bromine	1,000 to 1,800

In this table the higher production rate figure is based on a desalination plant concentration ratio of 2; the lower limit is based on a concentration ratio of 10, which is the maximum possible based on sodium chloride solubility.

2. Selection of Process

Various processes have been employed for the commercial recovery of chemicals from seawater; however, there is as yet no commercial installation combining a desalination plant and a chemical recovery plant. Research and development work towards this goal has been conducted for several years by the OSW, and combined installations have been operated on a pilot plant scale(10).

A by-product chemical recovery plant can be combined with a desalination plant in any one of three ways:

- The chemical recovery plant processes the seawater before it enters the desalination plant.
- The chemical recovery plant utilizes the brine blowdown from the desalination plant as its feed.
- The chemical recovery plant processes both the incoming seawater feed to the desalination plant and the blowdown from the desalination plant.

Studies have been conducted on processes for each of the above categories.

Several factors must be considered in selecting a chemical recovery process for a specific desalination plant installation and in determining if such a combined plant is technically and economically feasible at a particular location. These factors are discussed below.

a. Chemicals to be Produced

A variety of chemicals may be produced, depending upon the particular process employed. The chemicals produced commercially from seawater have been limited primarily to common salt, magnesium and bromine. However, variations of processes can be utilized to produce a salt solution instead of refined salt, potash, calcium sulphate, calcium carbonate, chlorine, sodium hydroxide and fertilizer components.

b. Market Price and Market Volume Potential at Plant Location

A market survey must be made to determine which, if any, of the chemicals should be considered for recovery at the particular proposed plant installation. This information would be used to select the most suitable process or process variation. Since transportation cost may be high in relation to the market value of the by-products, this factor should be considered in the original selection of plant location.

c. Cost and Availability of Raw Materials

Certain of the processes require relatively special and expensive raw materials, while others require only materials that are readily available. In certain processes some of the necessary raw materials can be produced from the by-products; for instance, soda ash, which is utilized in some of the processes, can be produced from a portion of product salt.

d. Required Process Development

While none of the proposed processes have been used in combination with full-scale desalination plants, portions of these processes are in common usage. On the other hand some processes contain steps that have been tried only in pilot plants or on a bench scale. A critical evaluation must be made, to determine if the proposed process is sufficiently proven to warrant the investment.

e. Patents and Licensing

A survey of possible patent and licensing problems should be made for each process seriously considered.

f. Process Flexibility

Some of the processes are sufficiently flexible to produce a variety of products, depending upon market conditions, while others are limited only to the production of a single product.

g. Compatibility with Desalination Plant

In the case where the output from a chemical plant is used as a desalination plant feed, the compatibility between the chemical composition of the feed and the desalination plant materials and design must be determined. As an example, one of the processes proposed to produce fertilizer involved the use of ammonia. The ammonium ions in the desalination plant feed would have been detrimental to the materials of construction in the desalination plant.

h. Potential Benefit to Desalination Plant Design and Economics

Certain of the processes remove magnesium and calcium from the seawater feed. This allows the chemical treatment costs for the desalination plant to be reduced, and also enables the desalination plant to be operated with a higher concentration ratio and at higher temperatures than would otherwise be possible. Cost reductions can therefore be made in both the capital and operating costs in the desalination plant.

i. Optimization of Combined Desalination By-Product Recovery Plant

The selected process should be sufficiently flexible in its operating conditions so that its design can be optimized with the design of the desalination plant. For example, the chemical process should not have an absolute requirement for a desalination plant concentration ratio in the order of 10 (even though technically feasible to do so without scaling) since this requirement might raise desalination plant costs appreciably because of boiling point elevation. The ideal arrangement would be to have both the chemical process and the desalination plant design sufficiently flexible so that overall optimum plant design may be obtained.

3. Process Descriptions

The processes described in this section are representative of the by-product recovery processes that may be utilized with a desalination plant. Their listing should not be considered complete, since a large number of combinations are available for adaptation to the particular requirements in the country which is planning the desalination plant. Examples of processes in each of the categories listed in the preceding section are provided.

The most versatile type of process, and the one that appears most promising in reducing overall water costs in areas that can utilize the several chemicals produced, is the integrated process that removes the scale-forming components from the desalination plant feed and removes additional by-product chemicals from the desalination plant brine blowdown. Two examples of this process type are described in the American Cyanamid report(9).

a. American Cyanamid Process I (Figure VI-5)

This process would recover magnesia, common salt and bromine, although other chemicals could be produced by process variations. Raw seawater is treated with soda ash and lime to precipitate calcium carbonate, and the clarified liquid is treated in ion-exchange units where magnesium and the remaining calcium are removed. The descaled water is then acidified and degasified, and then fed to the desalination plant. Product magnesia is obtained from ion-exchanger spent-regenerant solution in a series of steps. The product salt is obtained from the desalination plant blowdown after calcium is removed. The bromine is obtained from the brine effluent of the salt evaporation step by acidification and chlorination.

(1) Process Variations

In addition to magnesia, salt and bromine, process variations could be used to yield:

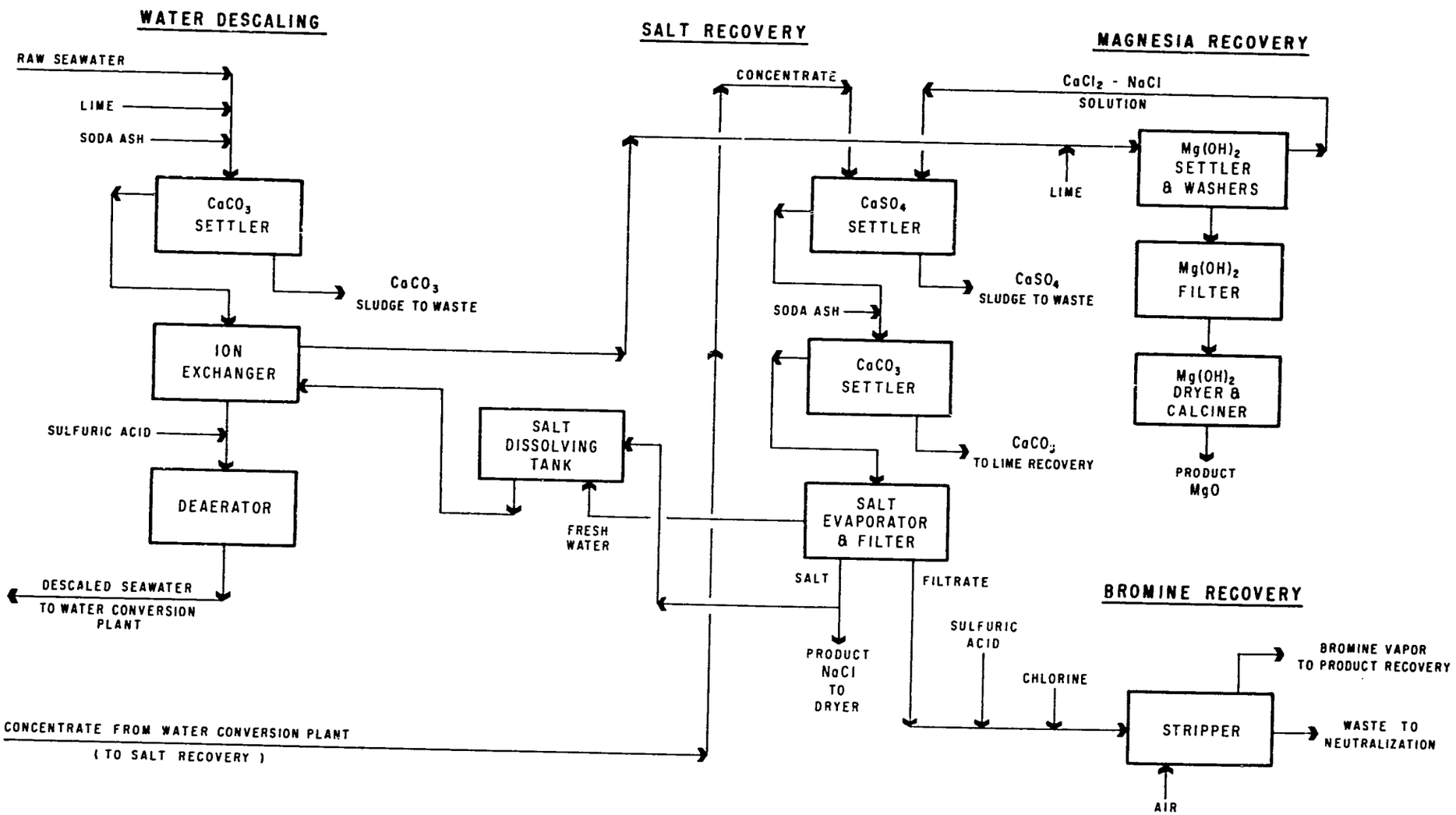
(a) Calcium sulfate from the thickened slurry, if sufficient demand existed in the immediate area.

(b) Calcium carbonate from either the salt recovery step or the raw-seawater treatment step.

FIGURE VI-5

FLOW SHEET PROCESS I

DESCALING AND RECOVERY OF MINERALS FROM SEAWATER



(c) Potash could be recovered from the concentrated effluent by dipicrylamine precipitation.

(d) Common-salt brine, if a market does not exist for all the salt production if this is refined salt only.

(2) Raw Materials

The raw materials needed would be:

(a) Lime (which could be obtained from calcium carbonate by-product).

(b) Soda ash (which would be manufactured from product salt by the Solvay process).

(c) Sulfuric acid.

(d) Chlorine for the bromine extraction step.

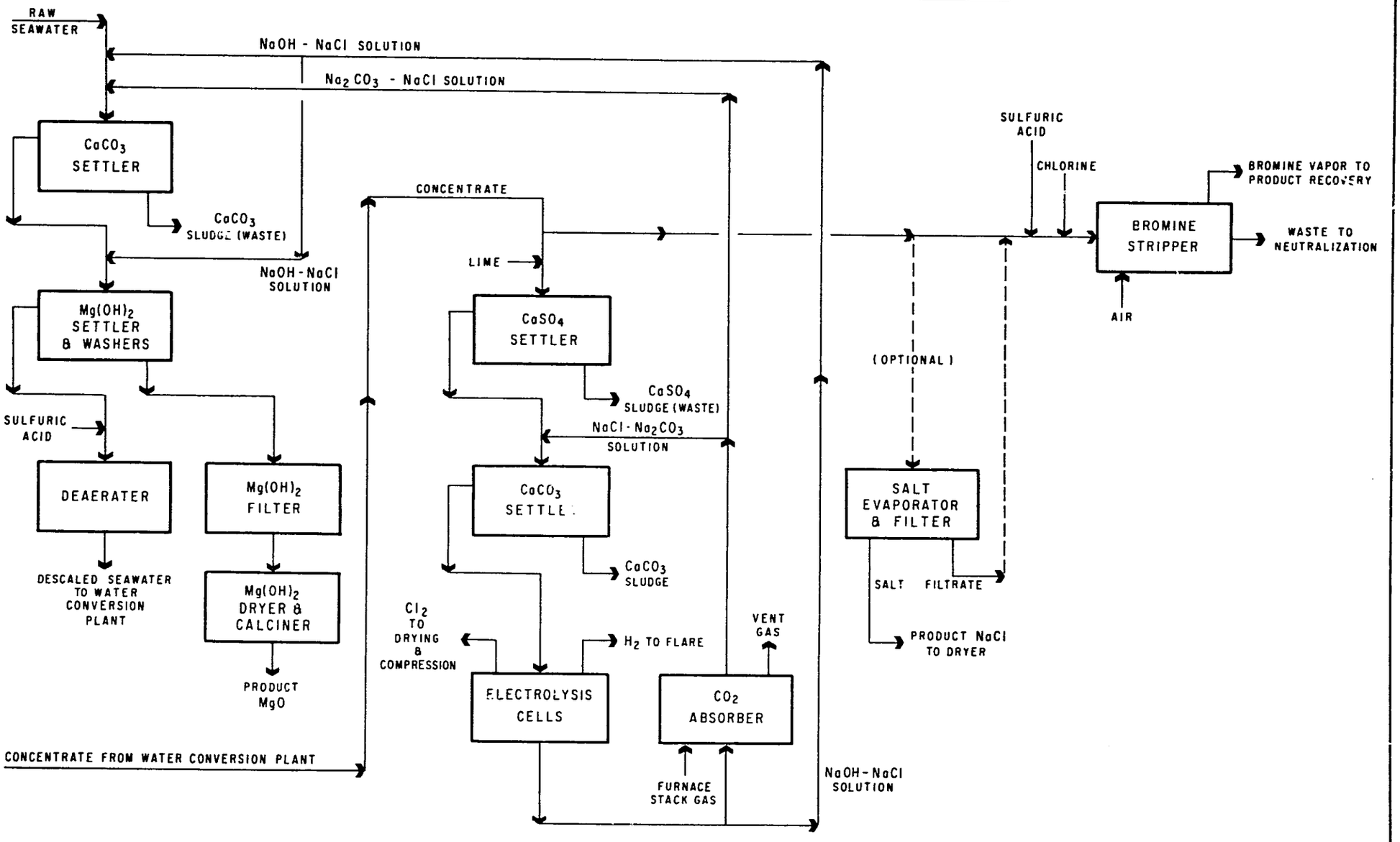
b. American Cyanamid Process II (Figure VI-6)

This process would produce magnesia and chlorine as well as refined salt, sodium hydroxide and bromine in the amounts required by local markets. Raw seawater is treated with sufficient sodium carbonate and sodium hydroxide solutions (both recycled from later stages in the process) to precipitate the calcium as carbonate. After separation, additional sodium hydroxide solution is added to the treated seawater to precipitate magnesium hydroxide, which is processed further to obtain product magnesia. As in Process I, the descaled water is acidified and degasified, and then fed to the desalination plant. After the sulfate is removed, the chlorine and sodium hydroxide are obtained from a portion of the blowdown brine from the desalination plant by electrolysis. Refined salt and bromine are recovered from the other part of the blowdown brine.

FIGURE VI-6

FLOW SHEET PROCESS II

DESCALING AND RECOVERY OF MINERALS FROM SEAWATER



This process has several advantages over Process I:

(1) Process II has greater flexibility and is capable of producing a larger variety of products, and the proportions of the amounts of these products can be adjusted to suit local markets.

(2) The only raw materials required for this process are lime, soda ash and sulfuric acid, and the first two of these could be made from plant products.

(3) Less laboratory development would be required since the individual sequences in the process have been verified.

(4) The economics of both Processes I and II are more favorable if the desalination plant concentration ratio approaches 10, the limit imposed by sodium chloride solubility; however, as high a concentration ratio as this is not necessary with Process II if the market conditions do not warrant the production of refined salt.

c. Blowdown Brine Processes

If calcium is not removed in a desalination plant feed pretreatment step, the maximum concentration ratio attainable in an MSF process without calcium sulfate scale formation will be approximately 2. Therefore, the blowdown brine from such a plant will be only twice as concentrated as seawater; only a limited advantage is offered by processing this instead of seawater. To obtain refined salt, additional evaporation steps would be necessary; solar evaporation could be utilized in a manner similar to the present commercial methods of obtaining salt from seawater.

Magnesium compounds could be obtained from the blowdown brine by precipitation of magnesium hydroxide through the addition of lime, and by the subsequent separation of the magnesium hydroxide from the calcium sulfate which would precipitate with it.

Bromine could be obtained from the blowdown brine in the same manner as that utilized in Processes I and II. Additional evaporation of the blowdown brine would not be necessary for the recovery of either bromine or magnesium.

d. Desalination Plant Feed Pretreatment

The processes proposed to recover by-product chemicals from the desalination plant feed stream are dual-purpose processes, since they produce a feed relatively free of scale-forming calcium and magnesium besides producing the chemical by-products. The head-end portions of integrated Processes I and II are two examples of processes in this category.

W. R. Grace & Co has performed considerable work for the OSW on feed pretreatment processes. One proposed process(11) would provide a product having value as a fertilizer, and would remove scale-forming chemicals. Phosphoric acid and soda ash are added to precipitate the calcium and magnesium as phosphates. The precipitate would have value as an ingredient for fertilizer manufacture, and the clarified liquid would be fed to the desalination plant. This process has been demonstrated satisfactorily on a pilot-plant scale, with ammonia used as a neutralizer instead of soda ash; however, the product water contained dissolved ammonia, and corrosion problems developed on the ferrous materials in the evaporators due to the presence of ammonium ions and the consequent decrease of pH. These problems are eliminated when soda ash is used; however, the process economics are not as favorable as with ammonia, and the process requires additional pilot plant work using soda ash.

The recovery of potassium from seawater by phosphate precipitation has also been investigated in laboratory-scale experiments by W. R. Grace & Co(12). Good recoveries of potassium, magnesium and calcium in a combined phosphate precipitate were obtained by treating seawater with phosphoric acid and sodium hydroxide. This combined precipitate will be a good source of both potash and phosphate in fertilizer manufacture.

The head-end processes that produce a fertilizer material and provide the desalination plant with a "descaled" water feed should be considered particularly for those countries where fertilizer costs are high and a large fertilizer market exists or is developing.

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VII. SUMMARY OF MAJOR CONCEPTUAL DESIGN STUDIES

A. Israel Dual-Purpose Electric Power-Water Desalting Plant

Israel anticipates a serious water problem by the early 1970's. In 1964, action taken toward solution of this problem involved the signing of an agreement for cooperation between the United States and Israel in the preparation of a study to determine the feasibility of a large desalination plant to be built in Israel. In October 1964, a Joint Team of experts from Israel and the U.S. completed a preliminary study of a combination seawater desalination and electric power plant. In the same month, an agreement was reached between the United States and Israel to initiate a feasibility study of a combined power generation and desalting plant.

In December 1964 the U.S. Department of the Interior, with the concurrences of the U.S. Atomic Energy Commission and of the State of Israel, contracted with Kaiser Engineers and its principal subcontractor, Catalytic Construction Company, to conduct an engineering feasibility and economic study of a dual-purpose power generation and water desalination plant to be located in Israel (1). The bases for the study were specified as follows:

- A plant capacity of 175 to 200 megawatts of salable electricity and 100 to 150 million cubic meters of water per year (i.e., 85 to 130 million gallons per day (mgd))

- A plant ready for initial operation in 1971 and commercial production in 1972

The study was conducted in two phases. In the first phase, completed in July 1965, the technical feasibility of the plant was determined, and comparisons were made of various dual-purpose nuclear power and desalination plant combinations using the multistage flash (MSF) evaporation process. A reference plant utilizing a nuclear reactor was selected for detailed evaluation as a result of these comparisons. A comparison was made between this reference plant and a comparable fossil-fueled dual-purpose plant.

The second phase was the preparation of a detailed evaluation of the reference plant and the development of information for use in preparing an application to funding agencies. The conceptual design was refined, and new estimates of capital cost, annual cost and unit water cost were prepared. In addition, several plant alternatives and factors important in the determination of the total annual cost were examined. The second phase of the study was completed in January 1966.

The study concluded that a dual-purpose power-water desalination plant of 200 megawatts electrical (MWe) and 100-mgd capacity for Israel is technically feasible for initial operation in 1971 and full commercial operation by late 1972. The demand for electric power and low-salinity water requires a solution such as a dual-purpose plant. The conceptual design on which the cost estimates were based employed certain desalination plant process and component design criteria requiring confirmation in a development program However, these design criteria were considered to be reasonable extrapolations from currently proven technology.

1. Reference Dual-Purpose Plant

The reference dual-purpose plant consists of a light-water nuclear reactor supplying steam to a turbine-generator producing electric power; the 25-psia turbine exhaust steam in turn supplies heat to the brine heaters of the desalination plant, wherein this exhaust steam is condensed before returning to the power plant. In the brine heater, the turbine exhaust steam furnishes the thermal energy required by the desalination plant for evaporating fresh water from seawater. The concentrated brine resulting from the partial evaporation of seawater is returned to the sea.

Two available sites on the Israel coast near the port of Ashdod were determined to be suitable locations for nuclear dual-purpose plants. Either of these locations would accommodate several plants of this size, thus ensuring room for expansion.

A survey was made of all contending U.S. reactor manufacturers to determine those which could supply reactors for commercial operation in 1971 for a plant of this size. Only the manufacturers of the light-water Pressurized Water Reactor (PWR) and Boiling Water Reactor (BWR) types indicated that their reactors would be available commercially and were able to supply operating characteristics and cost data for use in the feasibility studies.

The nuclear plant chosen for the reference plant design was rated at about 1,250 megawatts thermal (MWt) with a 250-MWe (200 MWe salable) turbine-generator supplying steam at 25 psia to the desalination plant. Either a PWR or a BWR was found to be suitable for this application. In addition to the normal back-pressure turbine exhaust system, a provision was made in each cycle to supply additional steam to the brine heater through a line with a pressure-reducing station, thus ensuring a continuous water plant operation when the turbine-generator is shut down, and providing for maximum use of the desalination plant during the winter when colder seawater temperatures would permit more water production. The PWR or BWR systems were housed in containment or pressure-suppression type reactor type buildings, respectively, which would contain the radioactive products released in the unlikely event of a rupture in the reactor coolant system.

The desalination plant conceptual design was based on the MSF evaporation process. This design was studied in sufficient detail to permit the determination of a development program to test those items of process and component design requiring validation in time to ensure commercial operation by the scheduled date. An optimization program was performed which yielded the following plant characteristics for the reference plant conceptual design:

Capacity	100 mgd
Trains	4
Stages	31
Maximum brine temperature	220 F
Performance ratio	10.3 lb water/1,000 Btu heat
Steam pressure at brine heater	25 psia
Tube-bundle material (recovery stages 3-28 and reject stages)	7/8 in. diameter 20 BWG 90-10 copper-nickle alloy
Tube-bundle material (brine heaters and recovery stages 1 and 2)	1 in. diameter 18 BWG 90-10 copper-nickel alloy

Tube-bundle material (reject-deaerator stage and air cooling sections in all stages)	7/8 in. diameter 20 BWG stainless steel/90-10 copper-nickel bimetallic
Chlorine consumption	140 metric tons/yr
Sulphuric acid consumption	33,000 metric tons/yr

Each of the four desalination plant trains has a capacity of 25 mgd and each is capable of operating independently of each other. Within each train there are four module streams operating in parallel, each having a capacity of 6.25 mgd and representing only a four-fold capacity scale-up of existing plants of this kind.

The reference nuclear dual-purpose plant was compared with a comparably sized fossil-fueled dual-purpose plant, with respect to characteristics and cost. This comparison showed that, for all fixed charge rates through 10%, the cost of water production in a nuclear dual-purpose plant of 100-mgd desalting capacity was lower than that of water produced in a fossil-fueled dual-purpose plant of the same capacity.

2. Cost Estimates

The capital cost estimates made for the reference plant design were based on 1965 costs, and reflect information received from Israeli construction contractors and Israel Electric Corporation as well as from several U.S. equipment manufacturers. Because of the uncertainty in the interest rate, the cost estimates were prepared for fixed charge rates of 5%, 7% and 10%. The estimates of total capital cost for these three fixed charge rates are shown in Table VII-1.

In determining the unit water cost estimates, the capital costs upon which the fixed charges were calculated and the cost for operating labor, materials and power credit were based on 1965 costs; these estimates of unit water cost are given in Table VII-2. A determination was then made of the cost of power from a single-purpose 200-MW fossil-fueled power plant. This value was credited against the total annual costs of the dual-purpose plant, and the differences divided by the quantity of water produced yields the unit cost of water. For this study the Joint Board specified a power credit of 5.3 mills per kilowatt-hour, which is the cost of electric power

TABLE VII-1

ESTIMATE OF TOTAL CAPITAL COST - 1965 COST BASIS
NUCLEAR DUAL-PURPOSE POWER-WATER DESALTING PLANT FOR ISRAEL
(Thousands of Dollars)

Nominal Fixed Charge Rate	5%	7%	10%
Dual-Purpose Plant			
Depreciable Items			
Power Plant	$ 52,850	$ 52,850	$ 52,850
Desalination Plant	74,540	74,540	74,540
Intake and Outfall	4,950	4,950	4,950
General Plant Facilities	2,440	2,440	2,440
Other Construction Costs	16,980	26,090	37,560
Subtotal	$151,760	$160,870	$172,340
Contingency @ 12%	18,240	19,330	20,660
Total Depreciable Items	$170,000	$180,200	$193,000
Nondepreciable Items			
Land Cost	$ 0	$ 0	$ 0
Working Capital	16,000	16,000	16,000
Subtotal	$ 16,000	$ 16,000	$ 16,000
Contingency @ 6%	1,000	1,000	1,000
Total Nondepreciable Items	$ 17,000	$ 17,000	$ 17,000
Total Dual Purpose Plant	$187,000	$197,200	$210,000

Note: It is estimated the additional investments required for electrical transmission facilities and water conveyance facilities are $5,000,000 and $25,000,000 respectively.

TABLE VII-2

ESTIMATES OF UNIT WATER COST AT THE PLANT BOUNDARY-1965 COST BASIS
NUCLEAR DUAL-PURPOSE POWER-WATER DESALTING PLANT FOR ISRAEL

Nominal Fixed Charge Rate	5%	7%	10%
Capital Cost*			
Depreciable Capital - Thousands of Dollars	$170,000	$180,200	$193,000
Nondepreciable Capital	17,000	17,000	17,000
Total Capital Cost	$187,000	$197,200	$210,000
Annual Cost - Thousands of Dollars			
Fixed Charges (Depreciable Capital)	$ 8,500	$ 12,610	$ 19,300
Fixed Charges (Nondepreciable Capital)	370	880	1,540
Operating and Maintenance Expenses	3,470	3,470	3,470
Fuel Expense	4,430	4,410	4,390
Total Annual Cost	$ 16,770	$ 21,370	$ 28,700
Power Credit	(7,900)	(7,900)	(7,900)
Net Annual Water Cost	$ 8,870	$ 13,470	$ 20,800
Water Production			
Million Gallons Per Day	100	100	100
Million Gallons Per Year	31,050	31,050	31,050
Million Cubic Meters Per Year	117.5	117.5	117.5
Unit Water Cost			
Cents Per Thousand Gallons	28.6	43.4	67.0
Cents Per Cubic Meter	7.6	11.5	17.7
Dollars Per Acre Foot	93.1	141.0	218.0

*Excluding electrical transmission and water conveyance facilities

produced in Israel at the prevailing fixed charge rates and 85% plant operating factor. It should be noted that fixed charges constitute the major portion of the total annual cost of the dual-purpose plant. The fixed charge rate, in turn, depends mainly on the interest rate. For instance, with a fixed charge rate of 7%, interest is 4.6% and depreciation (on a 30-year sinking fund basis) is 1.6%--the two together constituting 90% of the total fixed charge.

3. Recommended Development Program

One important basis for the study was that the reference dual-purpose plant would be an operating utility rather than a developmental-type installation. Therefore, either the process and component technology must have been proven or there must have been reasonable assurance that development work needed to confirm certain features would be concluded by the required date for freezing the desalting plant design, June 1968. The study indicated that the only items requiring development were a part of the desalination plant. These items include:

- Tube-bundle design
- Evaporator hydrodynamics
- Large pumps for recycle service
- Special mechanical hardware

It was anticipated that the MSF module work to be conducted by the Office of Saline Water in its program at the San Diego Saline Water Test Facility would be a source of information about the development items mentioned above. In addition, the acquisition and evaluation of oceanographic data and seawater chemical properties would be required in Israel, in the area of the proposed sites. Another area of work involves obtaining test information relative to the possible use of concrete for the evaporator shells.

B. Metropolitan Water District Dual-Purpose Electric Power-Water Plant

The Metropolitan Water District (MWD) of Southern California supplies water to six southern California counties - Los Angeles, Orange, Riverside, San Bernardino, San Diego and Ventura. It presently receives its water supply from the Colorado River and is to be supplied additional water from northern California through the California State Water Project aqueduct now under construction. MWD has been actively interested in determining the feasibility of desalting seawater to furnish additional water supply.

In August 1964, MWD entered into an agreement with the OSW and the Atomic Energy Commission to jointly sponsor an engineering and economic feasibility study for a combination power and desalination plant. In December 1964, MWD awarded Bechtel Corporation a subcontract to perform the engineering and economic study(2).

While the initial scope included studies of power and desalination plants ranging from 50 mgd to 150 mgd and 150 MWe to 750 MWe, the emphasis was placed on a plant with sufficient capacity to deliver 150 mgd to an MWD filtration plant located in Orange County.

The engineering and economic studies consisted of:

- Site studies including comparative costs, evaluations and preliminary safety analyses
- Exploratory drilling and oceanographic studies to establish the feasibility of constructing a man-made island for siting the dual-purpose plant
- The selection of an optimum plant configuration as a result of conceptual designs and analyses
- Preparation of construction cost estimates for the reference plant configurations at the selected sites

In April 1965, a group of electric utilities in southern California offered to participate in the proposed dual-purpose plant. Their proposal was to supply and operate a two-unit nuclear power plant which would have sufficient capacity to supply the power requirements of

the utilities as well as the steam and power needs of the desalination plant. This proposal was utilized in the subsequent studies and served as one of the bases for the final recommendations So that an equitable basis for the division of plant capital costs could be determined from these studies, cost estimates of single-purpose plants for producing only electric power were included in addition to those of the dual-purpose plants.

The engineering and economic feasibility studies concluded that a combination nuclear power and desalination plant to produce 150 mgd of desalted water was technically feasible for operation in 1971 Of the sites evaluated, a man-made island approximately 1/2 mile off the Orange County coast was concluded to be acceptable for the plant and to offer the lowest overall cost. The estimated cost of producing water at the desalination plant was 21.9¢ per 1,000 gallons based on acceptance of the proposal from the electric utilities The reference plant design developed in these studies utilized the MSF distillation process, which was concluded to be the process of demonstrated commercial reliability most suitable for extrapolation to a large-sized plant.

1. Site Investigations

The initial site investigations reviewed 17 sites located along the Southern California coast and evaluated these sites on considerations of nuclear safety, land cost and cost of delivering the water for use by MWD. As a result of the initial site studies two of the sites were selected for detailed evaluations and cost estimates These sites were:

- The Bolsa Island Site located approximately 3 500 feet offshore from Bolsa Chica State Beach in Orange County

- The Pelican Point Site located in Orange County on Irvine Company property, between Corona del Mar on the upcoast side and Crystal Cove on the downcoast side

Detailed engineering analyses were made for each site. To determine the feasibility of the island location, exploratory drilling and oceanographic investigations were made

The engineering analyses and site examinations established the technical feasibility of both sites for the location of a dual-purpose nuclear power and desalination plant. The cost estimates indicated that the Bolsa Island site would offer a lower overall project cost to MWD and the electric utilities than the Pelican Point site; the difference was approximately $30 million. Of this difference, $20 million is attributable to a difference in captial costs for power transmission facilities, a $2 million difference for product water conveyance and an $8 million difference for construction costs at the site itself.

2. Plant Description

The reference dual-purpose plant includes two commercially available light-water nuclear reactors, each rated at approximately 3,000 MWt, and a 150-mgd MSF evaporator desalination plant. The power plant uses a condensing back-pressure cycle; it includes two large condensing turbines each rated at approximately 700 MWe and a noncondensing back-pressure turbine rated at approximately 375 MWe. The back-pressure turbine would furnish steam to the desalination plant at approximately 35 psia. The electric utilities would obtain approximately 1,500 MWe, and the remainder of the electrical power produced would be utilized as auxiliary power for the desalination plant and by MWD for exchange power to pump Colorado River water. While the reference plant design used for developing cost estimates utilized a PWR nuclear steam supply system, the study concluded that both PWR and BWR systems were equally suitable from a technical standpoint. Since the two types of reactor systems were considered competitive, the cost estimates would be valid for the purposes of estimating water costs, regardless of the type of reactor finally selected.

The reference design for the desalination plant developed in the study consists of three identical MSF evaporation trains, each having a rating of 50 mgd. The steam from the back-pressure turbine unit is furnished to six brine heaters. The maximum brine temperature in the brine heaters is set at 250 F, based upon economic analyses.

Pretreated seawater is used both as a coolant and makeup feed for the desalination plant. About 60% of the seawater is returned to the sea after circulating through the heat-rejection sections; the remaining 40% is used as makeup feed for the plant. Sulfuric acid is injected into the feed stream to neutralize carbonate and bicarbonate alkalinity, releasing carbon dioxide. Vacuum deaerators are used to remove the carbon dioxide and other noncondensable gases from the makeup feed.

Each 50-mgd evaporation train has a total of 53 stages--49 heat recovery stages and 4 heat reject stages--enclosed in nine steel evaporator vessels. The eight heat recovery vessels are approximately 60 feet long by 150 feet wide, and the heat rejection vessel is 64 feet long and 150 feet wide. These vessels are designed for pressures ranging from a full vacuum to 15 psig.

Optimization studies were performed to establish the process and physical configurations of the reference desalination plant. These studies assumed a concentration ratio of the brine recycle stream of 2.0, a 30-year plant life, a plant capacity factor of 90%, an electric power cost of 2.5 mills per kilowatt hour and annual fixed charges of 5.44%. The characteristic features of the reference desalination plant design are as follows:

Capacity	150 mgd
Maximum brine temperature	250 F
Evaporator trains	3
Stages per train	53
Performance ratio	10.6 lb water/1,000 Btu Heat
Evaporator condensing area	10,500,000 sq ft
Brine heater area	680,000 sq ft
Seawater flow	532,000 gpm
Tube size and material	3/4 in. OD, 70-30 cupronickel

3. Cost Estimates

The capital cost estimates prepared for the study included those for both dual-purpose and power-only plants of comparable electric power output at the two sites considered most suitable the Bolsa Island and Pelican Point sites. Estimates of comparable power-only plants were necessary to provide a basis for the distribution of capital costs between MWD and the electric utilities since, according to the proposal from the electric utilities, the difference between capital costs for single- and dual-purpose plants was allocable to MWD.

The capital cost estimates were made on the basis of the conceptual designs developed in the studies and on 1965 prices, wages and technology; no provisions were made for escalation Design and price information for nuclear steam supply systems was received from four suppliers of light-water type reactors. Preliminary prices were obtained also on the turbine-generator units and other power plant auxiliary equipment. Desalination plant design and price information was obtained from five U.S. manufacturers of MSF evaporator plants. In addition, preliminary proposals were obtained for individual items of desalination plant equipment.

The summary of estimated capital costs, as given in the study report(2), are shown in Table VII-3. In each case, MWD receives 150 mgd of product water and 100 MWe net of electrical power while the electric utilities receive 1,529 MWe net of electrical power.

The capital cost estimates given in Table VII-4 do not include distributable costs for interest during construction, owner's engineering, nuclear insurance during construction and spare parts The estimated total plant investments, including distributable costs for the four cases, are shown in Table VII-4.

TABLE VII-3

CAPITAL COST ESTIMATE - 1965 BASIS
(Millions of Dollars)

	Pelican Point		Bolsa Island	
	Power-Only	Dual-Purpose	Power-Only	Dual-Purpose
Desalination Plant	-	$ 92.7	-	$ 94.0
Basic Power Plant	$149.6	178.6	$153.7	182.7
Other Facilities				
Site Preparation	2.8	7.0	-	-
Island	-	-	13.9	18.9
Causeway	-	-	1.8	1.8
Marine Lines	8.6	13.8	-	-
Switchyard, Including Main Transformers	6.7	8.2	10.2	13.4
Miscellaneous Facilities*	6.7	8.7	8.1	10.0
Land and Land Rights	18.7	18.7	0.4	0.4
Total	$193.2	$327.7	$188.1	$321.2
MWD Capital Costs	$134.5		$133.1	

*Includes seawater intake and discharge structures, yard, fire and service water systems, reactor unloading facilities, access and plant roads, yard lighting, fencing, landscaping and similar site improvements and office, shop, warehouse and associated equipment and furnishings

TABLE VII-4

TOTAL PLANT INVESTMENT - 1965 BASIS
(Millions of Dollars)

	Pelican Point		Bolsa Island	
	Power-Only	Dual-Purpose	Power-Only	Dual-Purpose
Water Production (mgd)	–	$ 150	-	$ 150
Power Production (MW)	$1,529	1,629	$1 629	1 629
MWD Investment	–	146 2	–	144 3
Electric Utilities Investment	220.3	218 9	213.7	213 1
Total Investment*	$220.3	$365.1	$213.7	$357 4

* Not including product water conveyance or electrical transmission costs

The study provided the estimates of water cost at the two plant sites under final consideration. These estimated costs were 21.9¢ per 1,000 gallons at the Bolsa Island site and 22 2¢ per 1,000 gallons at the Pelican Point site. These unit water cost estimates include the annual fixed charges on the MWD investment, operating and maintenance costs for the desalination plant and a portion of the power plant, acid costs for water treatment and a portion of the fuel and nuclear insurance costs The bases were:

Plant life	30 years
Fixed charge rate	5.44% (3.5% interest + 1 94% sinking fund depreciation)
Plant operating factor	90%
Sulfuric acid costs	1965 prices - $24/ton
Power credit for exchange power	4.0 mills/kwhr
Fuel cost (30 year average cost)	11.7¢/million Btu

4. Development Program

The study indicated that while the MSF process uses known evaporation and condensation principles, the size of the MWD plant equipment would be much larger than in existing plants. To permit the design and construction of a plant without incorporating large design margins, experimental development work was recommended in the following areas:

- Mechanics of flashing brine
- Heat and mass transfer of vapor
- Decarbonation
- Deaeration
- Overall plant performance

Work in these areas, particularly with large-sized equipment, would provide the required design confirmation. The OSW has included a large test module in their development program at the San Diego test facility. Results of testing with this module are expected to benefit both the MWD and Israel dual-purpose plant designs.

C. 50-mgd Conceptual Design Studies

In the latter part of 1964, the OSW initiated a series of desalination plant design studies under their 50-mgd Conceptual Design Study program. The purpose of this program was to make a technical and economic comparison of the several desalination processes that had been proposed, and by means of these engineering studies, to produce preliminary designs which would advance the technology of seawater conversions. These studies also served to identify specific areas which would benefit from additional research and development work.

The OSW contracted for a total of 15 of these conceptual design studies. The contracting firms included both equipment manufacturers and architect-engineering firms. Each firm selected the particular process or process variation it felt was most promising In all cases, the firms selected distillation or evaporation processes as being most suitable for the particular requirements established by the OSW. The list of these studies, and a summary of the significant results of each are given in Table VII-5. A brief summary of certain design features and conclusions from each study are provided in this section, to supplement the information given in the table.

1. OSW Ground Rules

To ensure that the studies would be performed on a comparable basis, the OSW established a detailed set of ground rules as presented in Table VII-6, page VII-18. The principal features of these ground rules are as follows:

"The study is to cover only the desalination section of a dual-purpose power-water plant. Design capacity of the water plant is 50 million gallons per day (50 mgd). It is assumed that the heat source is a fossil-fueled plant. Extraction steam from the power plant is delivered to the water plant and hot condensate from the water plant s brine heater is returned to the power plant. Electric power generated by the power plant is made available to the water plant."

	Allis Chalmers	Applied Research & Engr Ltd
Process	MSF	MSF
Water Cost (¢ per 1,000 gal.)	35.4	34.2
Maximum Brine Temperature (F)	250	233
Brine Blowdown Concentration Ratio	2.0	2.0
Performance Ratio (lb water per 1,000 Btu)	16.1	14.4
Brine Heater Concentration Ratio	1.71	1.74
Trains	1	1
Effects	1	1
Stages, Heat Recovery	38	69
Heat Rejection	2	4
Total	40	73
Feed Treatment Method	Acid.-Deaer.	Acid.-Deaer.
Flashing Range, (degrees F)	165	150
Recycle Ratio (lb brine per lb product water)	6.9	7.5
Evaporator Tube Material	Al-Brass & Cu-Ni	Al-Brass
Evaporator Shell Material	Carbon Stl	Carbon Stl
Steam Required at Brine Heater, Flow 10^6 lb/hr	1.15	1.26
Pressure (psia)	37	23
Power Requirements (mw)	14.4	13.7
Capital Cost ($\$10^6$)	27.0	28.7
Condensing Heat Transfer Surface (10^6 Sq Ft)	5.4	6.2

Legend

MSF	Multistage Flash Evaporation
COMEX	Combined Multiple Extraction
LTV-MEMS	Long-Tube-Vertical, Multieffect Multistage
LTV-MEFF	Long-Tube-Vertical, Multieffect Falling Film
MEMS	Multieffect Multistage
TFMED	Thin-Film Multieffect Distillation

TABLE VII-5

SUMMARY OF 50-MGD CONCEPTUAL DESIGN STUDIES

Aqua-Chem	Badger	Baldwin Lima Hamilton	C. F. Braun	Burns and Roe		Dow Chemical	H. K. Ferguson
MSF	MSF	COMEX	LTV-MEMS	MSF		LTV-MEFF	MEMS
29.1	32.3	36.8	31.3	34.7	40.5	42.7	47.4
245	290	290 & 255	229	250		308	300 & 292
1.87	1.43	1.85	3.0	2.0		4.8	3.0
18.0	21.0	18.2	16.7	13.1	13.4	16.4	20.4
1.6	1.15	1.0@290F 1.2@255F	1.0	1.7		1.0	1.02
1	2	1	1	1	3	1	2
1	1	4	18	1		20	21
80	73	-	51	26		-	100
4	-	-	0	2		-	0
84	73	65	51(MSF)	28		-	100(MSF)
Acid. - Deaer.	Acid. - Deaer.	Acid. - Deaer.	Acid. - Deaer.	Acid. -Deaer.		Acid. - Deaer. +IX	Acid. - Deaer.
171	212	205	137	170	172	204	200
6.9	6.1	-	1.5	6.3		1.25	1.5
Ti	Ti	90-10 Cu-Ni	Ti	Ti	90-10 Cu-Ni	Al-Brass & Cu-Ni	Al-Brass & Cu-Ni
Concrete	Steel-Concrete	Carbon Stl	Concrete & Steel	Con-crete	Steel	Carbon Stl	Carbon Stl
1.02	0.87	1.03	1.09	1.43	1.39	1.17	0.85
30	71	72 & 42	23	48		85	72'
14.6	0.16	17	8.9	16.2	43.4	16.0	13.2
21.6	28.2	38.4	34.0	25.5	28.2	39.7	61.0
6.1	5.9		5.77	4.31	4.04	5.0	5.7 + 2.4

Fluor Corp	Foster Wheeler	General Electric	Lockheed Missiles & Space Co	Ralph M. Parsons	Worthington Corp
MSF	MSF	TFMED	MSF	MSF	MSF
42.1	35.0	31.4	43	52.5	36.1
250	250	252	250	273	270
2.25	2.35	2.4 - 3.0	2.0	1.8	1.993
14.8	16.2	17.1	12.4	13.9	21.4
1.93	2.0	1	1.7	1.5	1.625
4	1	1	4	3	4
1	1	22	1	1	1
40	68	-	27	94	59
2	4	-	3	4	2
42	72	-	30	98	61
Acid. - Deaer.	Acid- Deaer.	Acid. - Deaer.	Acid. - Deaer.	Acid. - Deaer.	Acid. - Deaer.
165	175	180	172	194	196
7.0	6.7	1.5	6.67	5.9	5.4
90-10 Cu-Ni Concrete	70-30 Cu-Ni Carbon Stl	Al-Brass Concrete & Steel	90-10 Cu-Ni Carbon Stl	70-30 Cu-Ni Carbon Stl	90-10 Cu-Ni Carbon Stl
1.25	1.16	1.09	1.50	1.34	0.88
38.6	48	35	38	48	48
13.9	12.9	5.4	15.4	7.1	11.4
34.8	33.2	27.8	35.4	52.9	41.1
5.38	5.82	1.45	3.5	4.5	4.65

TABLE VII-6

OSW GROUND RULES FOR 50-mgd CONCEPTUAL DESIGN STUDIES

Site	Southern California Coast
Plot Elevation	20 feet
Cost of Land	$25,000 per acre
Seawater Temperature	65 F
Seawater Total Dissolved Solids	33,600 ppm
Product Water Temperature	90 F maximum
Capital Charges	5.185% amortization (3-1/8% for 30 years) + 0.25% insurance + replacement costs determined by contractor
Steam Cost	Varies with pressure from 21¢ per million Btu at 23 psia (sat.) to 28.3¢ per million Btu at 72 psia (sat.) for steam to brine heater
Electric Power Cost	6.2 mills per kwh
Plant Operating Factor	81% for combined dual-purpose plant; 90% for water plant only
Brine Concentration Ratio	2.1 maximum @ 250 F 1.15 maximum @ 290 F
Blowdown Concentration Ratio	3 maximum
Plant Life	30 years
Construction Cost Estimate Basis	December 1964 construction cost index for Southern California
Water Plant Portion of Shared Facilities Cost	$2,175,000

2. Summaries of Conceptual Design Studies

As indicated in Table VII-5, the conceptual design studies performed by the 15 firms produced a wide range of estimated water costs for a 50-mgd desalination plant. The figures varied from 29.1¢ per 1,000 gallons to 52.5¢ per 1,000 gallons, and these two limiting figures were obtained for the same basic process - the MSF evaporation process. This wide variation indicates that different factors of conservatism were employed by the various firms in the preparation of cost estimates and in the selection of materials and equipment utilized in their conceptual designs; it also indicates some differences in design ingenuity between the firms. The study results provide no indication of any particular process or process variation having clear advantage over the others.

Brief summaries of the individual conceptual design studies are provided below:

a. Allis-Chalmers Manufacturing Company

Allis-Chalmers chose the single-effect MSF process for their conceptual design study, since they concluded that it was the simplest and least expensive for dual-purpose operation and it lends itself to multiple-type construction modules. The standard modules used for evaporator shells are of welded carbon steel construction, and would be prefabricated to minimize field erection cost. The use of concrete shells was considered but was not recommended because of the unresolved questions concerning the ability of concrete to provide 30 years' service. Allis-Chalmers estimated that the lifetime of their aluminum-brass and 90-10 copper-nickel tubes would be 15 years, and their water cost calculations included the cost of retubing once during the 30-year design life. They considered titanium tubing too expensive to justify its apparently excellent corrosion-resistant qualitites.

Allis-Chalmers indicated their belief that additional research was justified in the areas of steam separation and demisters, particularly for the low-pressure stages of the plant; in the determination of fouling factors to provide additional information for the calculation of overall heat-transfer coefficients; and in the area of tube-bundle design and manufacture, because the heat exchange equipment represents such a large percentage of the total capital investment in an MSF plant.

b. Applied Research & Engineering, Ltd

Applied Research & Engineering indicated their belief that, for the immediate future, the MSF evaporator process is the only type that could be used in a 50-mgd desalination plant operating on a commercial basis. As a result of their consideration of several tube materials for their flash evaporator, they selected aluminum-brass tubing since they felt that this material would have a better resistance to pitting corrosion than the 90-10 copper-nickel tubes. In their cost analysis, they assumed one complete tube replacement during the 30-year plant life. In their conceptual design, they fitted tubes along the length of the evaporator and passed them through several stages. They indicated that a shallow tube-bundle design was advantageous from the standpoint of overall heat-transfer coefficient.

For future development work, Applied Research & Engineering suggested that better heat-transfer data be obtained so that design margins could be lowered; that heat-transfer coefficients be increased by the development of better venting techniques and optimized tube-bundle designs; that thinner gauge tube materials be used, and longer and cheaper tubes be perfected by additional development work on tube materials; and that development work be performed on the use of concrete for the evaporator shell structures.

c. Aqua-Chem Incorporated

The process utilized by Aqua-Chem in their conceptual design studies was the MSF type distillation process. They utilized concrete evaporator construction, although they recognized that only limited knowledge was available on its behavior at

elevated temperatures and in saltwater and brine environments. Titanium tubing was selected for their evaporator tube material because of its indicated 30-year life. Their conceptual design utilized single lengths of condenser tubes for the heat-recovery and heat-rejection sections of the plant. They stated that, although the development of equipment for producing tubing in 900-foot lengths was not complete at the time, such construction was selected in the belief that equipment development could be completed before 1970. In their cost studies, they assumed anticipated 1970 costs for their titanium tubing.

In their recommendations for their engineering studies and tests Aqua-Chem included the development of a satisfactory concrete evaporator shell; the improvement of tube welding techniques to produce the long lengths of titanium tubing and methods of joining the titanium tubing to the titanium-clad tube sheets; a reduction in the quantity of acid necessary to control scale in the plant; improvements in flashing chamber geometry; and investigation of lower cost materials, such as wood, for the construction of certain of the evaporator internals.

d. The Badger Company, Inc

The process chosen by the Badger Company Inc for their conceptual design study is the MSF distillation process with intermediate recycle. There are three major process sections: the high-temperature section with 60 stages, the intermediate-temperature section with 3 stages and the low-temperature section with 10 stages. The recycle brine is extracted downstream of the intermediate-temperature section, and is combined with the seawater feed before it enters the tubes of the high-temperature section The evaporator stages are housed in long tunnel-like structures. The first 35 stages of the high-temperature section (the evaporator stages operating above atmospheric pressure) are housed in a steel shell structure 322 feet long The remainder of the plant's 73 stages (operating at subatmospheric pressure) are housed in a concrete shell structure 632 feet long After consideration of other tubing materials, including aluminum titanium tube material was selected for its corrosion-free properties in brine service and its predicted 30-year life without maintenance or replacement.

The plant capital cost estimate is based on predicted 1967 titanium tubing costs. The feed treatment to prevent scaling is acidification with sulfuric acid followed by deaeration in Stage No. 1, downstream of the brine heater. The Badger Company, Inc indicates that this location for deaeration has several advantages.

Their recommendations for future work to reduce operating and/or capital costs include investigations on scale-control, acid treatment to reduce acid consumption; experimentation on brine flow hydraulics of small flashdowns with large mass flows, to provide definitive design data for flashing areas and flash chamber dimensions; and the determination of the feasibility of computerized monitoring and control of a multistage process, to achieve efficient operation at optimum conditions and to achieve labor savings.

e. Baldwin-Lima-Hamilton Corporation

The conceptual design presented in the Baldwin-Lima-Hamilton (BLH) report is based on their Combined Multiple Extraction (COMEX) process. This process consists of a multieffect multistage system with a top temperature of 255 F, combined with a once-through system with a top temperature of 290 F. BLH indicates that this process has the advantages that only comparatively dilute seawater is exposed to the top plant temperature, and that the time of such high-temperature exposure is appreciably shortened. In this process, the feed stream is preheated to 145 F prior to decarbonation and deaeration BLH indicates that this is preferable to low-temperature pretreatment; they indicate that the reaction rate differs by a factor of 30. The evaporator tube material selected was 90-10 copper-nickel, and on the basis of past experience, they estimated that approximately 50% of the tubing would have to be replaced during the anticipated 30-year life of the plant.

BLH's recommendations for required research and development work were in the area of combating corrosion in a desalination plant. They indicated that continued basic research must be devoted to the mechanism of corrosion as a function both of brine composition and of the metallurgy of components of the desalination plant; they also indicated that the present long-term exposure tests under actual operating conditions using metal samples covered with various types of protective coatings should be continued.

f. C.F. Braun & Company

After several preliminary studies of the various distillation processes, C.F. Braun & Company concluded that the most efficient process scheme is one that utilizes both the LTV multiple effect evaporation and the MSF concepts. The major portion of the process heat load is transferred at a constant temperature differential in the falling film evaporators. This part of the process has the best thermal efficiency. The remainder of the process heat load is required for heating the seawater. A new and unique design was developed for the multiple-effect evaporator stages. It consists of a series of annular contiguous effects, with common walls, and both concrete and steel were used in its construction The use of titanium for tubes was recommended by C.F. Braun & Company, although they indicated that little experience was available to prove its superiority. In their cost estimate, they used the anticipated 1968 price of titanium tubing.

In their recommendations for future testing work, C.F. Braun & Company recommended that a plant module of one-eighth full plant capacity be built in accordance with their design, and tested to provide a firm basis for an optimum full-scale design. They also indicated that it may be possible to obtain further improvements by combining the MSF and LTV structures, thus eliminating much of the transfer piping.

g Burns & Roe, Inc

The process selected by Burns & Roe, Inc for their conceptual design study is the MSF evaporation system. After considering several different concepts for this system, they selected two concepts for detailed study and optimization Both schemes utilize the basic MSF process, with 26 heat-recovery stages and 2 heat reject stages. The equipment used for the two schemes represents completely different approaches For one scheme the flash evaporator stages are housed in horizontal vaults, arranged with their longitudinal axis in parallel. In the second scheme the stages are enclosed in vertical columns.

If one column was used to contain all 26 stages it would have to be about 400 feet high; therefore, the total brine flow through the heat-recovery stages was divided into three parallel streams, with two columns containing 13 stages each connected in series. Each of the six flash-heat recovery columns are 26 feet OD by 168 feet high. Each of the two parallel heat-injection columns is 34 feet OD by 55 feet straight height. Because of the foundation problems, the scheme utilizing the tall columns would be feasible only in locations where good foundation conditions exist. Based on the ground rules provided by the OSW, the scheme utilizing the horizontal evaporator structures provided lower-cost water. It was concluded, however, that for certain conditions the two schemes may be competitive.

h. Dow Chemical Company

For their conceptual design study Dow Chemical Company chose the multiple-effect falling film distillation process, since they felt that this was one of the most economical and most technically advanced methods of producing potable water from saline water. In addition to the conventional acidification-deaeration treatment for the incoming seawater feed an ion-exchange pretreatment also was included so that the seawater could be heated to more than 325 F and the brine concentration ratio could be nearly 5 without the formation of calcium sulfate. As a result of design studies an unusual and compact arrangement of the evaporator effects was developed. The effects are arranged in a circular structure approximately 300 feet in diameter. The evaporator effects are of common-wall construction so that a savings in steel can be achieved by designing for only the pressure difference between adjacent effects. The common pressure-walls also lower the amount of surface that has to be insulated, and minimize the interconnecting piping.

Dow Chemical Company's recommendations for future studies included investigation of the use of steam-turbine drives for the brine circulating pumps, and investigations on the substitution of steel for copper alloys in the evaporator. The latter would include the removal of the oxidizing agents normally present in seawater, since these accelerate the corrosion of steel.

i. H.K. Ferguson Company

The combination multieffect multistage flash (MEMS) evaporation process was chosen by H.K. Ferguson Company as the basis of their conceptual design study. They used carbon steel for the enclosures, and aluminum-brass and 90-10 copper-nickel for the tube materials. Their selection was made on the basis of current availability, technical status and cost. They considered other materials, such as concrete for evaporator vessels and titanium for tubing and lining, as promising materials that might be used in the future when their performance and life expectancy are sufficiently established

The recommendations for a research and development program made by H.K. Ferguson Company included further investigations of the basic thermodynamic properties of saline solutions; obtaining additional basic data on suitable fouling factors; development work on the use of concrete for large vessels and on various metals for tube materials; and additional work on pump types and materials of construction.

j. The Fluor Corporation

The conventional MSF process was chosen by Fluor Corporation for their conceptual design studies. Four preliminary trains of evaporators were used, each sized for 12.5 mgd. All of the evaporator vessels are of concrete construction. prestressed by post-tensioning both longitudinally and circumferentially. These cylindrical vessels are 28 feet ID and have heads of carbon steel. The maximum length of the evaporator tube-bundles utilizing 90-10 copper-nickel tubing, is 93 feet.

The recommendations made by Fluor for development work include work on pumps for evaporator plant service; continued studies on the use of coatings or linings for concrete exposed to hot brine; and additional investigation of tube materials, particularly titanium, and fabrication methods.

k. Foster-Wheeler Corporation

The process utilized by Foster-Wheeler in their conceptual design study was the single-effect MSF distillation process with brine recirculation. The evaporator concept investigations

resulted in the selection of a concept employing a barrel-vault shell design built upon a system of structural members The modular concept of the plant permits shop fabrication of the 100 foot long modules containing the factory-tubed condenser bundles; internal water boxes are employed for field welding to join each 100-foot section to the next. The total length of the 68 heat recovery stages is 704 feet. Because the 70-30 copper-nickel material has a long history of operation, it was chosen over the 90-10 copper-nickel material. The maximum tube length was set at 100 feet, to allow shop fabrication of tube bundles and because of lack of experience in fabricating longer tube lengths.

Foster-Wheeler's recommendations for further research and development work included plant operating tests at low-temperature conditions; further work on the prevention of corrosion on carbon steel in a brine environment; and additional development work on tube materials and fabrication methods

1. General Electric Company

The conceptual design studies performed by General Electric were based on the application of their thin-film multiple-effect distillation process, using data obtained from their previous, OSW-sponsored pilot-plant studies on this process. This process utilizes vertical fluted tubes to generate thin-films on both sides of the metallic heat-transfer surface where evaporating and condensing takes place. Thin-film distillation increases the rate of heat transfer, thereby permitting a decrease in the required heat-transfer area. The falling film of feed is distributed on the inside wall of the tube by a spray nozzle located at the top of each tube The higher-temperature effects utilize 10 foot long double-fluted evaporator tubes, while the tube lengths in the lower-temperature effects are 5 feet. The 22 effects are arranged in a rectangular structure with an overall dimension of 260 feet by 400 feet While the thin-film process could have used a concentration factor of 3 at 110 F, the conceptual design used blended seawater in the last several effects to limit brine concentration to about 2.5 and increase the available temperature difference in the last effects.

The recommendations made by General Electric for future research and development work included improvements in spray decarbonation of the feed; work on reinforced concrete for desalination plant structures, and coatings and linings for steel and concrete structures; and experimental work to allow the proper sizing of orifices, loop seals and baffles between stages and effects.

m. Lockheed Missiles and Space Company

As a result of their preliminary evaluation of various seawater distillation processes, Lockheed chose the MSF system for their conceptual design study. Particular design features incorporated in their plant were an efficient condenser tube-bundle design using bent tubes and bulkhead insert pieces to eliminate conventional water boxes, expansion bellows and adaptors; a cathodic protection system to minimize corrosion of steel in contact with brine; and the use of a dropwise promoter to achieve more efficient dropwise condensation The 50-mgd plant is divided into four equal evaporator modules, each capable of producing 12.5 mgd, and each capable of independent operation and control. The 500 foot long carbon steel evaporators are slightly tapered, with a diameter varying from 28-1/3 feet on the hot brine end to 34 feet on the cold end.

Additional work recommended by Lockheed was investigations on materials protection and service life for steel and the tube materials; continued investigations on the flashing process and stage hydraulics, to allow higher flashing brine flows; studies on the use of a dropwise promoter and on the effect of variables on fouling factor; and investigations of less expensive methods for the fabrication and assembly of condenser tube-bundles.

n. The Ralph M. Parsons Company

The conceptual design studies performed by the Ralph M. Parsons Company are based on the MSF distillation process. For the heat-rejection section and most of the heat-recovery section, vessels made of corrugated metal pipe arch section, 50 feet wide and 25 feet high are utilized. The corrugations permit the longitudinal thermal expansion of the vessel to be absorbed

within the vessel walls. A single vessel structure approximately 830 feet in length is utilized for the complete heat-recovery and heat-rejection sections. Flat-plate construction is used for the high-pressure portion of the heat-recovery section. The 1-inch OD 70-30 copper-nickel tubes are assembled, in place, into lengths of 348 feet maximum; the tubes span a number of stages in order to minimize costs of tube sheets, water boxes, expansion joints and connecting pipe.

The suggestions made by the Ralph M. Parsons Company for further development work were to provide data on the mechanisms and limitations of flashing brine flow and vapor release with small interstage pressure and temperature differences, and to evaluate vapor and gas transfer between stages to determine optimum configurations for the gas removal system.

o. Worthington Corporation

The conceptual design study performed by Worthington Corporation covers the optimization of a 50-mgd seawater distillation plant of the MSF type. The plant design is conventional except for relatively small changes in the cycle. Worthington indicates that withdrawing the product from the heat-recovery evaporator rather than from the heat-reject evaporator results in a savings of 20% to 25% of the heat-transfer surface in the reject evaporators. Corrosion control is enhanced by dividing the air removal of a venting system into three zones: a low-pressure zone in the reject evaporator, an intermediate-pressure zone in the vacuum section of the heat-recovery evaporator, and a high-pressure zone in the positive pressure section of the heat-recovery evaporator.

Worthington's suggestions for future studies include the improvement of seawater deaeration to reduce plant corr sion, work on concrete vessels with linings for desalination plants; additional work on stage separation flashing devices; tests on the use of baffles to prevent liquid brine carry-over; and additional studies on evaporator tube materials and fouling resistance.

D. References

(1) Engineering Feasibility and Economic Study for Dual-Purpose Electric Power-Water Desalting Plant for Israel, Kaiser Engineers in association with Catalytic Construction Company, Report No. 66-1-RE, January 1966.

(2) Engineering and Economic Feasibility Study for a Combination Nuclear Power-Desalting Plant TID-22330, Vols I, II and III, Bechtel Corporation, January 1966.