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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.

Burns and Roe, Inc.

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A MANUAL ON WATER DESALINATION

SUPPLEMENT NO. 1
TO
VOLUME ONE - TECHNOLOGY

OCTOBER 1972

PREPARED FOR
UNITED STATES
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AGENCY FOR INTERNATIONAL DEVELOPMENT
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WASHINGTON, D.C. 20523

BURNS AND ROE, INCORPORATED
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Notice

"This manual was prepared for the Agency for International Development (AID), to guide its officers as well as to aid foreign administrators and technicians involved in water resources development in formulating and evaluating desalination programs and projects for which United States Government assistance may be furnished or sought. The information contained herein is not sufficient 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 assistance as required for development of technical and economic feasibility, plans and designs, construction, and the initiation of operation of such a project.

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I. INTRODUCTION

The preceding Volumes I and II of this Desalination Manual described the state-of-the-art through 1966. Subsequent advancements presented herein do not reveal any new technologies, but rather considerable refinement and commercialization of what has already been stated.

The objective of this Supplemental Manual of Water Desalination is to present updated technical and economic guidelines on water desalination, together with information and an outline of studies required to enable foreign administrators and technical personnel, as well as AID officers, to develop and appraise water development programs.

In the arrangement os this Manual, a condensed section on state-of-the-art is followed by sections on typical plants built since 1967 and on conceptual designs. These sections require reference to Volumes I and II for thorough detail. Finally there is a comprehensive section on Water Development Planning, feasibility analysis, and updated budget water plant pricing for commercially available types of desalting plants. Some environmental considerations and feed treatment and typical brine disposal methods have been added. Glossaries of Terms used in this Manual, to aid in its understanding, appear under III Saline Water Phenomena.

Repetition of material already presented in Volumes I and II is not a purpose of this Supplement. Such material is referenced where necessary and is available through AID as needed. However, sufficient nomographs and other aids for budget pricing of flash distillation plants for coastal sites have been reprinted from Volume II, and for pricing membrane plants at inland sites from OSW's R&D #555, so that a coherent costing procedure is contained in this supplement.

II. SUMMARY

1. Commercialization of Vertical Tube Evaporators (VTE) and Combinations with Multi-State Flash (VTE/MSF) Evaporators.

In distillation, 1972 has seen commercial, high performance Vertical Tube Evaporator plants come into prominence, incorporating preheat stages based upon multi-stage flashing followed by VTE effects. The economics of these units are favorable for seawater and where large volumes of product water (1-60 mgd) are produced. Capital costs at the 2.5 mgd level are in the range of $1.50 per daily gallon of capacity at a performance ratio of 14. These plants compare favorably with multi-stage flash evaporators both in capacity and performance. Total world-wide capacity as of January, 1971, was 45 mgd in 105 plants.

2. Reverse Osmosis Updated

Since 1967, three Reverse Osmosis designs have demonstrated technical and economic viability on a fully commercial scale. These are:

a. Nylon hollow fine-filament cartridges.
b. Spiral-wound cellulose acetate sheet modules.
c. Cast, tubular cellulose acetate modules.

Plants are presently in operation in capacities up to 0.8 mgd, serving a variety of municipal and industrial needs. All the designs have proven to be successful on brackish water; but because of pressure limitations, inadequate salt-rejection capabilities, and shortened membrane life at high salinities they are less applicable to the desalting of seawater. 12 mgd of total capacity was on order, or had been installed in late 1972.

3. Vapor Compression with Scale Control

Vapor compression has reached a total capacity of 3 mgd

in twenty-six installations as of January 1, 1971.

Vapor compression plus scale control by seeding is a new process for inland sites in small to moderate sizes:

Concentration to 25% solids is possible, which yields a reduced volume of brine for disposal. Aluminum condensers are used, which allow low capital cost. Constraints are the same as for other distillation processes. The module size is anticipated to be 0.2 mgd; however, the only present installation, on a cooling tower in El Paso, Texas, is of a 50,000 gallon per day size.

4. Vacuum Freeze - Vapor Compression

Theoretically, this should be a very economical process since the evaporation of one pound of water will freeze 7.5 pounds of ice. Scale, corrosion and pretreatment problems, except for deaeration, are minimized. The principal problem is scale-up to commercial sizes. Three small plants are in commercial operation on this process, totalling about 0.3 mgd.

5. Electrodialysis

With a world-wide installed capacity of 15.5 mgd in 57 plants as of mid 1972, electrodialysis has become a large volume water producer from brackish water sources. Siesta Key, Florida, and Gillette, Wyoming at 1.5 mgd each, are the largest operating plants.

6. Pretreatment

Pretreatment of brine is called for in most desalination processes. In all processes, except freezing, capacity and maximum operating temperature levels are limited by the formation of calcium scales. Alkaline scales are soluble in acidic solutions. Control is primarily effected by softening, using ion

exchange or lime-soda, and by pH adjustment with H_2SO_4. Coagulating and softening processes have the advantage of simultaneously reducing the other metallic ions such as iron and manganese which interfere with the membrane processes and which increase corrosiveness of the saline solutions. Because of their expense, minimum acceptable pretreatments are constantly being sought.

Organic constituents of brackish waters must be controlled for the membrane processes. This is also true of inorganic colloids which tend to accumulate on the membrane. Coagulation, generally followed by filtration, is the most feasible method for removal. The presence of bacteria and algae must be controlled to prevent their attack of membranes. Control of these is by chlorination, unless this, too, attacks the membranes

7. Brine Disposal

Brine disposal may be a particularly costly problem of inland desalting plants. Environmental factors sometimes preclude even controlled release into surface streams. The two processes most used are evaporation to dryness with possible regional stockpiling and by-product recovery, or concentration in ponds and pumping into subsurface aquifers where feasible, salt lake beds, or other non-leaching sumps. When concentration in a pond or by electrodialysis is practiced, the strong solution is usually disposed of elsewhere. Encapsulation of salt briquettes in heat shrunk films for disposal constitutes a recent advancement (43).

8. Materials of Construction

The allowable materials of construction of saline water plants are influenced by the pretreatment which the brine has received. Recent experience has shown that more expensive alloy materials are needed for surfaces coming into contact with raw or acid-treated saline waters. Corrosion has proven to be pitting

rather than generalized and is of particular importance when surfaces are in contact with stagnant warm saline solutions containing oxygen. Alkaline pretreatments and oxygen scavengers have been used successfully in conjunction with a corrosion allowance where carbon steel equipment was furnished, but a well maintained organic coating to protect the steel at temperatures below 200°F is to be preferred.

Concrete linings composed of mono-calcium aluminate have been found advantageous for protecting both steel and ordinary concrete against salt water environments. Concrete shells for evaporators seem to show some cost advantage over steel construction, but are still being tested.

Aluminum has shown promise because of its good heat conductivity in vapor compression equipment, (as already mentioned under 3) and in vacuum freezing equipment, where low temperatures reduce galvanic corrosion. In the Middle East, its use has been proposed in horizontal tube evaporators (HTE) using multiple effects for temperatures below 170°F. Preliminary test results are promising. With it, however, copper-bearing alloys, as well as iron, must be avoided throughout the system, and the concept is still viewed as highly experimental.

III. SALINE WATER PHENOMENA AND DEFINITIONS

A. General

Volume I of this Manual presents typical analyses of seawater, Tables III-1 and III-2, which may be used for design purposes. These include proximate analyses of the various ions and physical properties of the water and a complete analysis of the elements present in more than trace amounts. A chart is also included, Figure III-1, which illustrates variations in salinity of the waters of the North Atlantic. Table III-3 gives the composition of seawater from several locations which may support desalination plants. The tabulation compares these analyses with normal seawater.

Table III-4 and subsequent discussions present the proximate analyses of brackish water from several sources where experimental work has been carried out on desalination of these waters. The saline ground waters are typed by their processes of origin:

1. Brackish due to leached minerals.
2. Brackish due to seawater intrusion.
3. Connate waters.
4. Waters trapped in closed basins.

Tables III-3 and III-4 are condensed and reprinted in this Supplement as Table III.

B. Glossary of Saline Water Terms (Pages III-3 to III-6)

The "Salinity Terminology" presented in Volume I is somewhat incomplete by 1972 standards. Therefore, in this Supplement a glossary has been reproduced from the USDI Office of Saline Water's Distillation Plant Data Book for 1971 (Reference No. 44).

TABLE III

IONIC COMPOSITION OF SELECTED SALINE SOURCES
(mg/kg or ppm)

Constituent	Eastern Mediterranean(3)	Kwajalein Atoll Lagoon(8)	Arabian Gulf Shuaiba, Kuwait(4)	Seawater(6)(21)	Dead Sea(2) (100 m Below Surface)	Inland Bahrein Arabian Gulf Site	Caspian Sea(2)
Total Dissolved Solids	38,600	38,200	42,000	34,482	271,000***	2,196	12,858***
Chloride (Cl^-)	21,200	22,000	23,100	18,980	178,600	1,020	5,500
Sulfate (SO_4^{--})	2,950	2,300	3,100	2,649	400	192	2,970
Bicarbonate (HCO_3^-)	155	62*	-**	140	-	242	-
Bromide (Br^-)	72	-**	80	65	5,200	-	-
Fluoride (F^-)	2	-	-	1	-	-	-
Iodide (I^-)	-	-	-	-	-	-	-
Nitrate (NO_3^-)	1	1	-	-	-	-	-
Carbonate (CO_3^{--})	-	10	-	-	Trace	-	48
Sodium (Na^+)	11,800	11,200	-**	10,556	32,000	458	3,200
Magnesium (Mg^{++})	1,403	82	1,665	1,272	35,700	67	773
Calcium (Ca^{++})	423	742	500	400	12,700	217	297
Potassium (K^+)	463	-**	-**	380	6,400	-	70
Strontium (Sr^{++})	-	-	-	13	-	-	-
Silicon (Si^{++++})	-	-	-	-	-	-	-
Iron (Fe^{+++})	-	1	-	-	-	-	-
Manganese (Mn^{++})	-	1	-	-	-	-	-
Boric Acid (H_3BO_3)	-	-	-	26	-	-	-
Silica (SiO_2)	-	7	5	-	-	-	-
Metasilicic Acid (H_2SiO_3)	1	-	-	-	-	-	-
Others	131*	1,793*	13,550*				

* Computed Value
** No Value Listed in Reference
*** Total of Listed Values Only

GLOSSARY OF DESALTING TERMS - Source Ref. #44

Acid Treatment - A method for controlling scale by removing the alkaline scaling constituents through the addition of an acid to the makeup water.

Alkalinity - The sum of the carbonate, bicarbonate, and hydroxide ion concentrations, all expressed as parts per million of equivalent calcium carbonate ($CaCO_3$).

Approach to Equilibrium - The temperature difference between the average flashing brine temperature leaving a stage and the temperature of the vapor immediately above this flashing brine.

Backward Feed - The make up sea water is fed to the last effect (lowest temperature) and the concentrated brine blowdown is discharged from the first effect (highest temperature).

Blowdown Water - The water that leaves the evaporation section which has not been flashed to product. All of the salts entering the evaporators from the makeup water are concentrated into the blowdown water.

Boiling Point Elevation - The increase in boiling point temperature of a solution with respect to the boiling point of the pure liquid at the same pressure.

Brackish-Water - Water with total dissolved solids content greater than that recommended by the Public Health Service as suitable for drinking but less than sea water.

Brine - Any stream in the desalting plant with a dissolved solids content greater than the dissolved solids content of the makeup water.

Brine Heater - The heat exchanger through which external heat is supplied to a multistage flash desalting plant. The temperature of the brine here is the highest reached in this distillation process.

Chlorinity - The total amount of chloride (grams) contained in one kilogram of solution with all the bromide and iodide expressed as chloride.

Chlorosity - The total amount of chloride (grams) contained in one liter of solution at 20° C with all the bromide and iodide expressed as chloride.

chlorosity = chlorinity x density

Concentration Factor - The ratio of total dissolved solids to the total dissolved solids in standard sea water.

Concentration Ratio - The ratio of total dissolved solids in any stream in a desalting plant to that in the feed. The feed is considered as the value one.

Cross Tube Flash Evaporator - A flash evaporator in which the condenser tubes are oriented perpendicular to the flow of the flashing brine in the flash chamber.

Deaeration - Removal of dissolved air and other dissolved gases from water by stripping the water at its saturation temperature.

Decarbonation - Removal of only carbon dioxide from water.

Degasification - A general name for processes such as deaeration and decarbonation.

Demister* - See Entrainment Separator

Density - The mass of a unit volume of a substance. The density is a function of temperature, usually decreasing with increasing temperature.

Dry Air Equivalent - The quantity of dry air that would require the same amount of motive steam (for a steam jet-ejector) as the actual mixture of wet gas for a given set of operating conditions.

Effect - The evaporator body or chamber receiving heat from an external source or another effect, and producing vapor which may serve as the heat source for another effect or which may be condensed by other means.

Effluent - The portion of the brine leaving the desalting plant that is employed for cooling purposes only (cooling water). The term is used also to denote the mixed effluent and blowdown.

Enhanced Tube - A heat exchanger tube modified to increase the overall heat transfer coefficient. These modifications may include alteration of the actual surface configuration of the tube, insertion of devices which will induce turbulence in the evaporating side or application of materials which promote dropwise condensation.

Entrainment - The mechanical carrying of solute drops with a vapor during an evaporation, stripping, or distillation process.

* Trade Mark of Otto H. York Co., Inc.

Entrainment Separator - A device for removing liquid drops entrained with the generated vapor. This usually refers to a layer of knitted wire mesh oriented perpendicular to the vapor flow upon which the droplets impinge, coalesce, and run off.

Erosion - The wearing away of solids by continuous mechanical or physical action.

Falling Film Evaporator - A vertical shell and tube evaporator with the brine being fed to the top of the tubes. The liquid is distributed around the inner periphery of the tube and falls in an annular film down the inside of the tube. The center of the tube is filled with the generated vapor and some entrainment droplets. Condensing steam on the shellside is used for the heat source.

Flash Down - The temperature difference between the flashing brine entering and leaving a stage.

Flash Evaporator - An evaporator in which a preheated liquid is subjected to a pressure below its vapor pressure so that boiling or flashing will occur.

Flashing Brine Flow Rate - The mass flow rate in pounds per hour per foot of evaporator width of brine in the flash chamber.

Fluted Tubes - An enhanced tube with the tube wall being a series of axial corrugations or ridges and valleys, usually on both inner and outer surfaces. The tube is used in VTE plants for vapor generation and heat recovery.

Forward Feed - The makeup sea water is fed to the first effect (highest temperature) and concentrated brine blowdown is discharged from the last effect (lowest temperature).

Gain Ratio - See Performance Ratio

Hardness - See Total Hardness

Heat Capacity - The amount of heat required to raise a unit mass of substance one degree in temperature.

Heat Input Section - The heat exchange section in which external heat is added to the plant or effect.

Heat Recovery Section - The heat exchange section of the plant or effect in which available heat in produced vapors is recovered by preheating brine or generating a lower pressure vapor.

Heat Rejection Section - The heat exchange section of the plant or effect in which heat is removed from the unit (heat sink) by condensing the last effect's or stage's vapors, cooling product, and cooling brine.

Horizontal Spray Film Evaporator - A type of single or multieffect film evaporator with tube orientation being horizontal and brine being sprayed on the tubes. Steam condensation occurs on the tube inside walls which serves as the driving force.

Hybrid Plant - A desalting plant which combines the best features of two or more desalting methods to obtain higher efficiencies.

In-Stage Temperature Losses - See Stage Inefficiencies.

Intergrandular Corrosion - Corrosion taking place preferentially at boundaries between the crystal grains.

KBTU - Thousand BTU.

KGAL - 1000 Gallons.

KKBTU - Million BTU.

Lime Magnesium Carbonate (LMC) - A proposed method for controlling scale by a two stage coagulation process. Lime is added to precipitate magnesium hydroxide, and magnesium carbonate is added to precipitate calcium carbonate. The calcium carbonate is partially recalcined and slaked for lime recovery with the carbon dioxide generated during recalcining reacting with the magnesium hydroxide to form magnesium carbonate. The material balance is such that mainly calcium carbonate is removed from the water although separate magnesium hydroxide or carbonate byproduct recivery is possible.

LMTD - Logarithmic Mean Temperature Difference.

Load Factor - The ratio of actual production for a given period of time (usually one year) to the design production for the given time period.

Load Gas - The quantity of gas, noncondensables plus water vapor, which is fed to the suction of a steam jet ejector.

Long Tube Flash Evaporator - A multistage flash evaporator with a number of successive stages being spanned by a single tube bundle. The condenser tubes are oriented parallel to the flow of the flashing brine.

LTV - Long Tube Vertical. See Vertical Tube Evaporator.

Makeup Water - The salt or brackish water that enters the evaporation section of a desalting plant. The makeup equals the sum of the product and blowdown in an MSF evaporator.

MEMS - See Multieffect Multistage Flash Evaporators

Methyl Orange Alkalinity - The total alkalinity (See Alkalinity) indicated by methyl orange test.

MGD - Million Gallon per Day

Mist Eliminator - See Entrainment Separator

Motive Steam - Normally, the high pressure steam supplied to a steam jet ejector.

Multieffect Evaporator - An evaporator system with more than one effect with the vapors generated in one effect used to provide heat to the next and the brine from one effect serving as feed to the next.

Multieffect, Multistage Flash Evaporators (MEMS) - A series of multistage flash evaporators, with brine recirculation at different concentrations separating each effect. The heat input section to the first effect is from an external source, but the heat rejection section of the first effect serves as the heat input to the second effect, etc., with an external heat rejection section for the final effect. Essentially, it is several flash evaporator plants in series.

Multistage Flash Evaporator - An evaporator in which a preheated liquid passes through a series of stages or chambers with each successive stage at a low pressure so that some liquid will flash in each stage. Stages are numbered in order of decreasing boiling temperature.

Noncondensable Gases - Gases such as air and carbon dioxide which do not condense with the water vapor when heat is removed.

Non-Equilibrium Flashing - See Approach to Equilibrium

Once-Through Multistage Flash - A multistage flash evaporator in which the entering sea water recovers heat in the condenser tubes of the heat recovery section, is heated further in the heat input section, is flashed in each stage, and is then returned to the sea. Recirculation is not utilized.

Parasitic Losses - The in-stage temperature loss factors, excluding boiling point elevation, which must be taken into account in calculating the LMTD for the condenser. These factors are (1) approach to equilibrium, (2) vapor pressure loss through mist eliminator, and (3) vapor pressure loss through condenser bundle.

Performance Ratio - The ratio of pounds of product water (distillate) produced per 1000 BTU supplied.

Phosphate Treatment - A method which retards the formation of alkaline scale by the addition of controlled amounts of a polyphosphate. With this treatment, scale will still form, but at a much slower rate than without any treatment.

Pitting - An indentation or cavity on the surface of a solid indicating that localized corrosion has occurred. Pits in metals are usually caused by oxygen setting up a galvanic cell and in concrete by a chemical reaction.

Preheater - A heat exchanger in the heat recovery or rejection section of an effect in which brine or makeup water is heated by condensation of steam.

Pretreatment - The processes such as chlorination, clarification, coagulation, acidification and deaeration that may be employed on the feed water to a desalting unit to minimize algae growth, scaling and corrosion.

Product - The distillate or fresh water produced in a desalting plant.

Product Water - The addition of chemicals such as lime or soda ash to the product water to reduce its corrosiveness.

Rising Film Evaporator - See Upward Brine Flow Evaporator

Saline Water - Seawater, brackish water or other mineralized water.

Salinity - The total dissolved solids in the sea water in grams per kilogram with all halogens expressed as chloride, carbonates as oxide and all organic matter oxidized.

Salinity = 0.03 + 1.805 Chlorinity.

Scale - Salts deposited on heat transfer surfaces which retard the rate of heat transfer. These salts, precipitated from a brine that is locally supersaturated, are classified either as alkaline, such as calcium carbonate, or non-alkaline, such as calcium sulfate.

Sea Water Intake - The total quantity of water entering a desalting plant. This includes both the make up water to the evaporation unit and the coolant water.

Seeding - A method for controlling scale by the addition of crystals of scale to the brine in order to promote precipitation of scaling constituents on these particles rather than on the heat transfer surfaces.

SEMS - See Single Effect Multistage Flash Evaporator

Single Effect Multistage Flash Evaporator (SEMS) - A multistage evaporator with a single heat input section, a single heat rejection section, and either once-through operation or brine recirculation.

Stage - (Applies only to flash evaporation). The individual flash chamber in which vapors are produced from the heated brine.

Stage Inefficiencies - The factors which reduce the effective performance of a flash stage by reducing the available temperature driving force, the heat transfer rate or the amount of vapor produced in the stage. The major causes are: presence of non-condensable gases, interstage vapor leakage, boiling point elevation, and the parasitic losses of non-equilibrium flashing, mist eliminator pressure drop and condenser bundle pressure drop.

Standard Sea Water - Sea water with a total dissolved solids content of 34,483 ppm and chlorinity of 19.000. This sea water is used as the basis for calculating a concentration factor for the various sea waters throughout the world. A list of the major constituents is given in Table 6.1.

Stress Corrosion Cracking - A tendency for metals to crack at points of stress under nominal loading caused by the action of chloride ions and oxygen.

Submergence Losses - See Approach to Equilibrium

Sump-to-Sump Brine Feed - A process by which the brine from the sump of one VTE effect is fed directly into the sump of the next effect so that flashing does not occur in the water box.

Terminal Temperature Difference (TTD) The temperature difference between the saturated vapor entering a stage or effect and the make up recycle brine inside the tubes leaving the stage. Also known as the approach temperature.

Total Hardness - The concentration of calcium and magnesium ions in water expressed as ppm equivalent calcium carbonate ($CaCO_3$).

Transgranular Corrosion Cracking - A form of localized failure which is more severe under the combined action of stress and corrosion than would be expected from the sums of the individual effects of stress and corrosion acting alone. The surface direction of the crack is perpendicular to the direction of the stress and crosses the grain boundaries.

Traveling Screens - A moving screen that is used to remove debris from the lines that are used for supplying the sea water to the desalination plant.

TTD - See Terminal Temperature Difference.

Upward Brine Flow Evaporator - A vertical one-pass shell and tube evaporator with the brine being fed to the bottom of the tubes. The brine and the vapor generated rise or flow upward concurrently through the tubes and are separated in a chamber located directly above the tubes.

Vapor Release Rate - The volume of vapor generated in a flash stage per unit time and area.

Vertical Tube Evaporator (VTE) - A vertical one-pass shell and tube evaporator with the liquid flowing inside the tubes (usually falling film) and the vapor condensing on the outside of the tubes. (Previously referred to as Long Tube Vertical, LTV)

VTE - See Vertical Tube Evaporator

Water Box - The liquid collection chamber, just prior to the heat exchanger tubes in which the liquid is redistributed so that the brine flows evenly through all of the tubes.

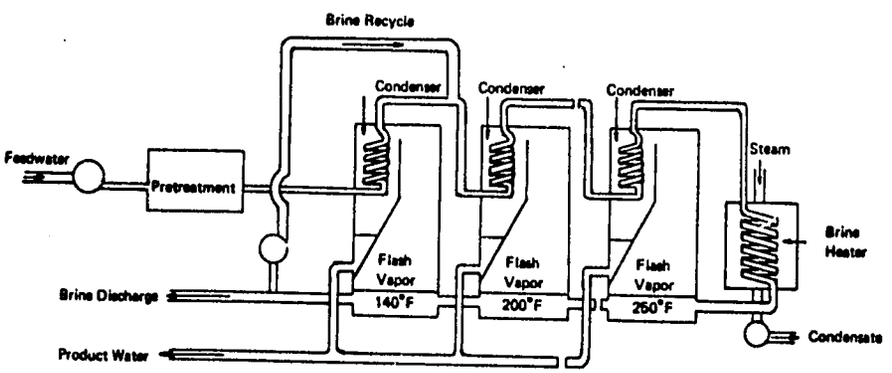
IV. DESALINATION PROCESS STATE-OF-THE-ART

A. Distillation

1. Multi-stage Flash Evaporators (MSF)

a. General: Under the present state-of-the-art, (1972), multistage flash (MSF) desalting plants can be designed for converting water containing up to 50,000 ppm total dissolved solids (TDS) to potable water containing 50 ppm or less.

The operation of the multi-stage flash distillation process is pictured schematically in Figure IV-1. Basically, the process involves heating saline water progressively up to 250° F, then flashing this water in a number of successive chambers or stages operating under progressively lower pressures. In each chamber, vapor is condensed by heat exchange with the incoming feedwater. Low-temperature low-pressure steam is used to heat the water, and the condensate from the brine heater is pumped back into the steam system. The feedwater is pumped through a pretreatment section to remove impurities which would damage the system. Part of the brine is recycled to improve system performance. The process requires large volumes of cooling water and hence is more suitable for coastal than inland sites.



Source: Office of Saline Water and Southwest Research Institute

FIGURE IV-1

SCHEMATIC FLOW DIAGRAM OF THE MULTISTAGE FLASH DISTILLATION PROCESS

b. Advancements: As of January, 1971, there were 216 multistage flash evaporator plants in operation, or under construction, throughout the world with a total capacity of 210 million gallons per day. The OSW demonstration plant, built in 1962, near San Diego, California, then dismantled and moved to Guantanamo Naval Base, Cuba, has a capacity of 1 mgd. Another 1-mgd desalting plant, based on a MEMS version of the MSF cycle, has been built near San Diego. The new plant is twice as efficient as the initial plant, mainly because its operating temperature is 300°F. A 7-mgd plant has been built and operated in Tiajuana, Mexico, and a 5-mgd plant of the OSW Universal MSF plant design at Jeddah, Saudi Arabia. Kuwait now has a 30 mgd plant, and a 48 mgd plant is scheduled for construction in Hong Kong.

c. Problem Areas: Problems are the high costs of the alloy construction materials such as cupro-nickel condenser tubes and liners for the evaporator bodies. Aluminum and concrete have been suggested, but are still highly experimental. Cooling water requirements pose problems with the environment, both as to pollution by heavy metal corrosion products and by heat. Other environmental considerations involve noises that must be muffled and plumes that must be dissipated. Another problem of this process is that corrosion, scale and fouling limit the top temperature to 250°F. Research is under way to develop control methods which will be effective at higher temperatures. When these methods are developed, (1) temperatures will be increased and higher plant efficiencies will be attained. Currently, only the MEMS experimental test plant can operate at a temperature of 300°F. The associated lime-magnesium carbonate softening plant has been built which shows some promise of allowing operation on seawater up to 350°F, but not at an attractive cost.

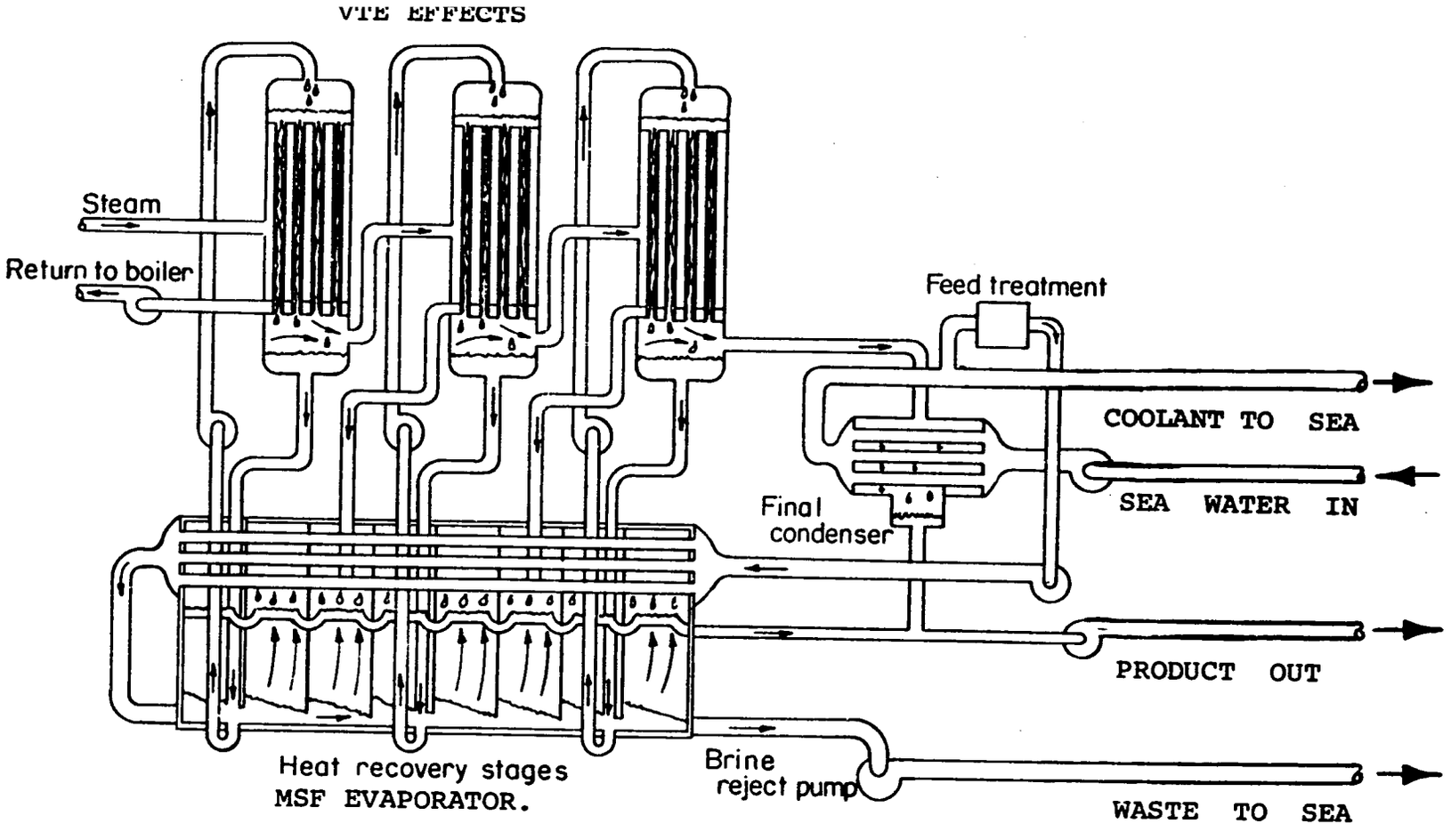
2. Vertical Tube Evaporators (VTE) and Combinations with MSF (VTE/MSF) Evaporators

a. General: 1.0 to 10 mgd Vertical Tube/Multi-Stage Flash Evaporator (VTE/MSF) units are capable of converting water containing not over 50,000 ppm total dissolved solids (TDS) to potable water containing 50 ppm or less with high steam economy.

A schematic flow diagram of a representative VTE/MSF distillation plant is shown in Figure IV-2. Warm feedwater from the final condenser is pre-treated and then pumped through the heat-recovery-stage tubes of an MSF evaporator. Here it is heated to within a few degrees of the operating temperature of the vertical tube evaporator (VTE) by condensing both the vapors of the brine and product, as they flash down and cool prior to being discharged. After being preheated, the feedwater is pumped sequentially into the feedwater chests of the VTE effects, where it boils as a falling film in the vertical tubes. Heat is supplied by steam at about 265° F. which enters the first effect steam chest. Vapors from the final effect are condensed in a final condenser, using raw seawater as the coolant. Steam condensate goes back to the boiler, and the concentrated brine from the last effect goes to waste. A steamjet ejector and a barometric condenser are used to maintain a vacuum on the last MSF stage. Seawater is concentrated up to three times normal salt content by the process. The process requires large volumes of cooling water and hence is more suitable for coastal than inland sites.

b. Advancements: As of January 1, 1971, there were 105 commercial VTE plants in operation, world-wide, with an installed capacity of 45 mgd. A 3.0 mgd module is being completed, expandable to 15 mgd, at Fountain Valley, California. Two 2.25 mgd units have been selected for the Virgin Islands.

Major advancements have been made since 1967 in attractively priced multi-mgd modular packages, in the enhancement of heat-transfer surfaces, in



COMBINATION VTE-MSF DISTILLATION PROCESS

FIGURE IV-2 (Source: Ref. #58)

scale control, and in the stability of VTE/MSF hybrids to startup and turndown conditions. By the use of special inlet distribution nozzles and doubly-fluted vertical tubes, once-through flow of the brine without starving the evaporator tubes has been made possible, with its consequent hotter allowable first-effect temperatures and freedom from scale. Improved feedwater treatment allows higher blowdown concentrations and lower corrosion of the parts exposed to brine. VTE heat-transfer coefficients, including normal fouling factors, now average about 1,400 btu/ft^2-hr-^oF.

c. Problems Areas: Indications are that the reliability, stability, and durability of VTE/MSF plants will prove to be on a par with those of conventional MSF plants, and therefore that solutions to the MSF operational problems already mentioned will also be solutions to the remaining VTE/MSF problems.

3. Vapor Compression Evaporators (VC)

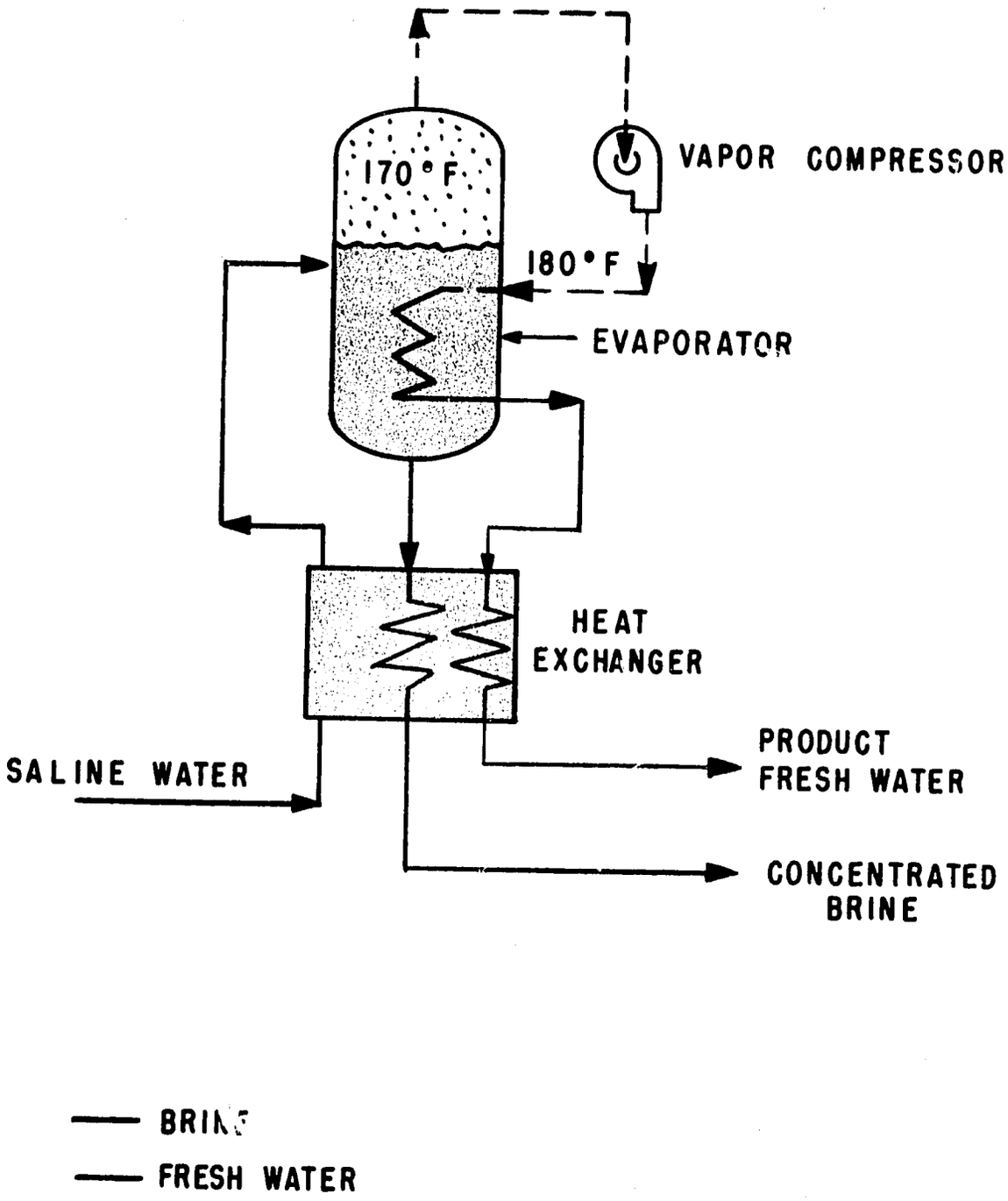
a. General: Up to 0.2 mgd vapor compression evaporator (VC) desalting plants are capable of converting water containing not over 50,000 ppm total dissolved solids (TDS) to potable water containing 50 ppm or less, while simultaneously concentrating to a heavy brine.

The primary difference between VC and other distillation processes is the manner in which heat is added to the system. Vapor compression distillation uses mechanical energy to desalt water, and uses heat only for startup.

The principle of forced brine circulation vapor compression is shown in Figure IV-3. As saline water is sprayed on the tube bundle in the evaporator, it is heated by the steam inside the tubes. A mixture of vapor and hot brine enters the entrainment separators. The dry vapor travels to a

FIGURE IV-3

VAPOR COMPRESSION DISTILLATION



centrifugal compressor. The compression of the vapor (steam) increases its energy content. The vapor is then returned to the evaporator tubes to serve as the heating medium for the incoming feedwater. As the vapor condenses, it gives up enough heat to boil more saline water. Feedwater is preheated by the brine discharge and product water streams.

Vapor compression distillation, in comparison with other distillation processes, has a very low energy requirement. Also, the vapor compression process does not require large amounts of cooling water, an advantage at inland sites. For plants rated 20 mgd or less, this concept appears to have the potential to reduce water costs below conventional MSF distillation costs.

b. Advancements: The normal single plant capacity at present is limited to about 60,000 gpd, and primarily shipboard installations. The largest land-based unit was the 1 mgd demonstration plant of OSW at Roswell, New Mexico. A new development is being tested by El Paso Natural Gas Company in the 50,000 gpd size on a cooling tower blowdown. This unit uses aluminum heat transfer surfaces, and a $CaSO_4$ seeding technique for scale control. Brine concentration can run as high as 25% for easy disposal at inland sites, and the modular size can be expanded to 0.2 mgd.

c. Problem Areas: VC stills operate on heat derived from the work of compression. Shaft work, however, is a very costly source of heat, and so VC units must operate at high thermal efficiency in order to be economical. Large condensers are required to allow use of small thermal driving forces, and also to minimize scaling potential at high salinities. Scale tends to form on the compressor blades, unbalancing the unit and shortening its life. Hydrojet compressors are being tested at the Roswell facility, as well as the aforementioned scale-seeding techniques, which removes the scale and

entrainment ahead of the compressor, in order to overcome this problem. A description of commercially operating units is given under Section VI Dual-Purpose Plants.

4. Miscellaneous

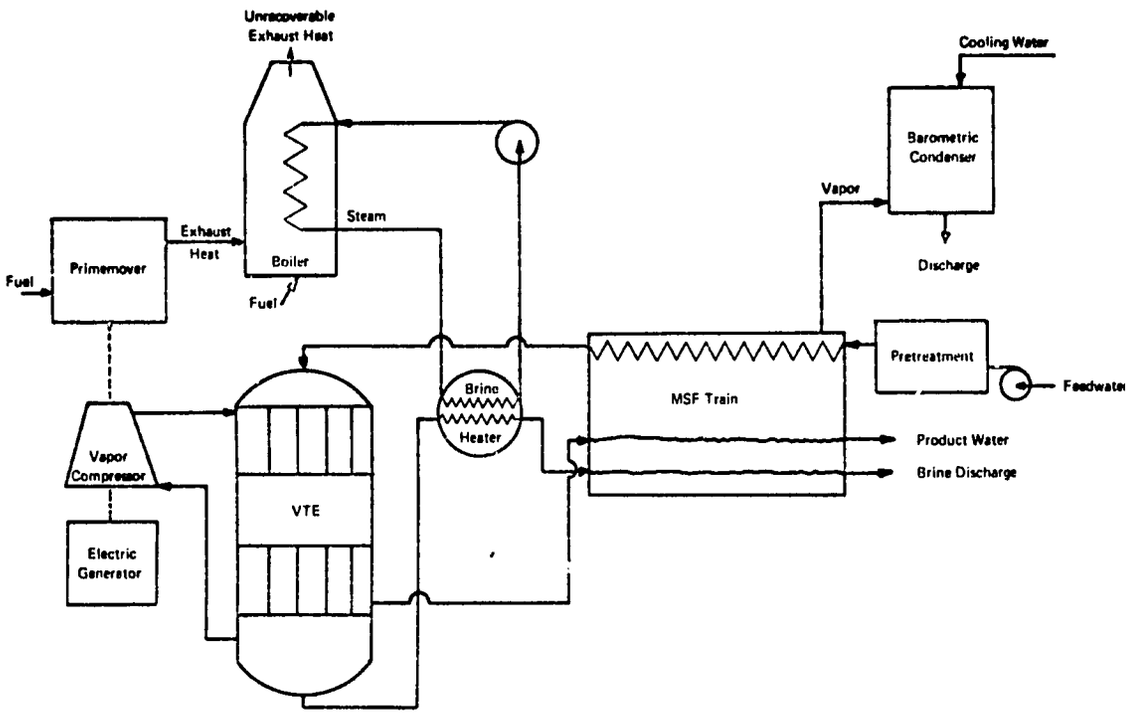
a. VC/VTE/MSF Hybrid: In the previous sections those distillation processes which are useful for production of potable water from the sea have been covered. Another large area of usefulness is for inland areas in utilizing brackish waters and renovating waste water and irrigation runoff. In these areas it is desirable to concentrate the waste brine as much as possible to facilitate disposal and also to avoid large volume usage of cooling water.

The large size VC unit, just mentioned, can be combined with the other distillation processes to do just this, and is expected, when a hybrid demonstration plant is completed, to operate more economically, while renovating waste water than any distillation process acting singly.

This hybrid VC/VTE/MSF distillation process was scheduled for testing at Brownsville, Texas, but has now been postponed. One million gpd vapor compression-vertical tube evaporator-multistage flash (VC/VTE/MSF) desalting plants are theoretically capable of converting water containing up to 50,000 ppm total dissolved solids (TDS) to potable water containing 50 ppm or less. None is yet being built although the special vapor compressor required has been purchased, but not yet tested, by OSW.

The primary difference between VC/VTE/MSF and other distillation processes is the manner in which heat is added to the system. The process uses mechanical energy as well as heat energy to desalt water.

A schematic flow diagram of a representative VC/VTE/MSF distillation plant is shown in Figure IV-4. Feedwater is pumped into the system through a pretreatment section and then through the heat recovery tubes of an MSF unit. Here it is heated to a few degrees below the operating temperature of the vertical tube evaporator (VTE).



Source: Oak Ridge National Laboratory and Southwest Research Institute

FIGURE IV-4

SCHEMATIC FLOW DIAGRAM OF THE VC/VTE/MSF PROCESS

After being preheated, the feedwater is pumped into the VTE where it is evaporated in one or more effects. The required heat is supplied by the vapor compressor which compresses the vapor produced in the last effect to the saturation pressure required for heating in the first effect.

An exhaust heat boiler is used to recover the exhaust heat from the primemover driving the vapor compressor and electric generator. The steam generated in the boiler is used to heat the brine entering the MSF train. The steam is condensed in the brine heater; the condensate is pumped back into the boiler.

A barometric condenser is used to maintain a vacuum on the last MSF stage. Cooling water for the condenser is furnished from plant intake or by cooling towers.

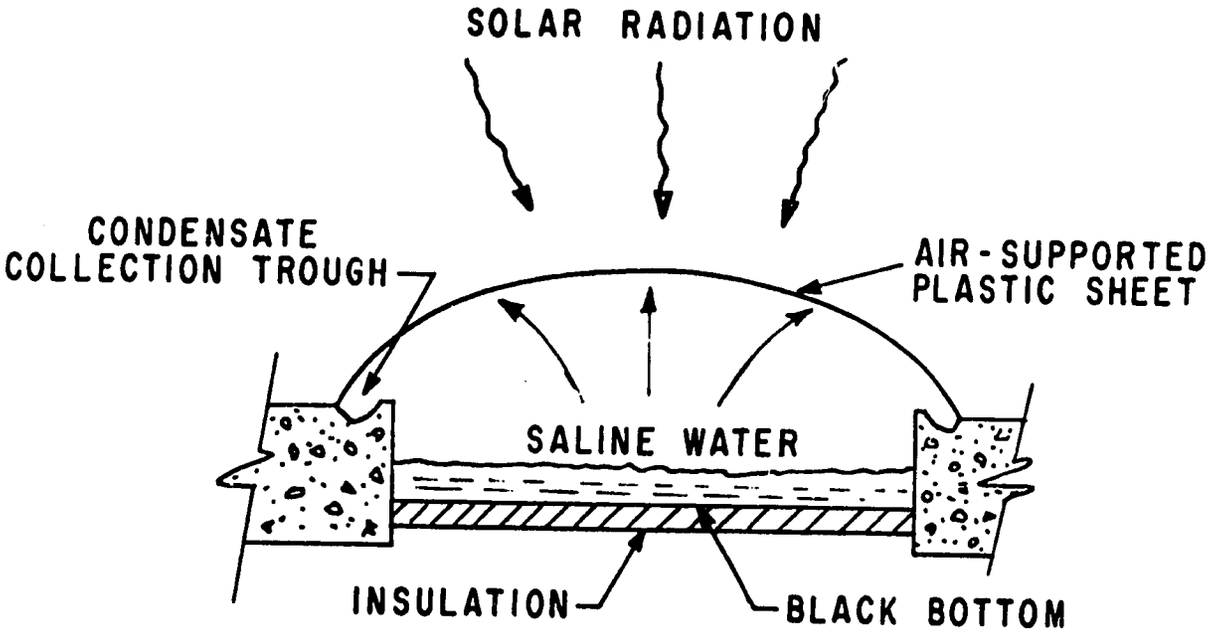
b. Simple Solar Still: To supply very small amounts of potable water in small isolated rural communities situated at low latitudes, the solar still is a simple and durable producer. The apparatus can be hand operated, easily maintained, and requires no feed preparation or fuel.

Figure IV-5 depicts the unit. Water vaporized under glass by solar heat, condenses on the inside glass surface and runs off into a collection gutter for use. Salt water is added to the supply basin as needed for production and for flushing the unit.

Construction details are given under Section VIII, Solution of Water Shortage Problems. Variations of this simple survival type solar still are also possible, based upon nocturnal radiation, soil moisture condensation, dew collection, etc.

FIGURE IV- 5

SIMPLE SOLAR STILL

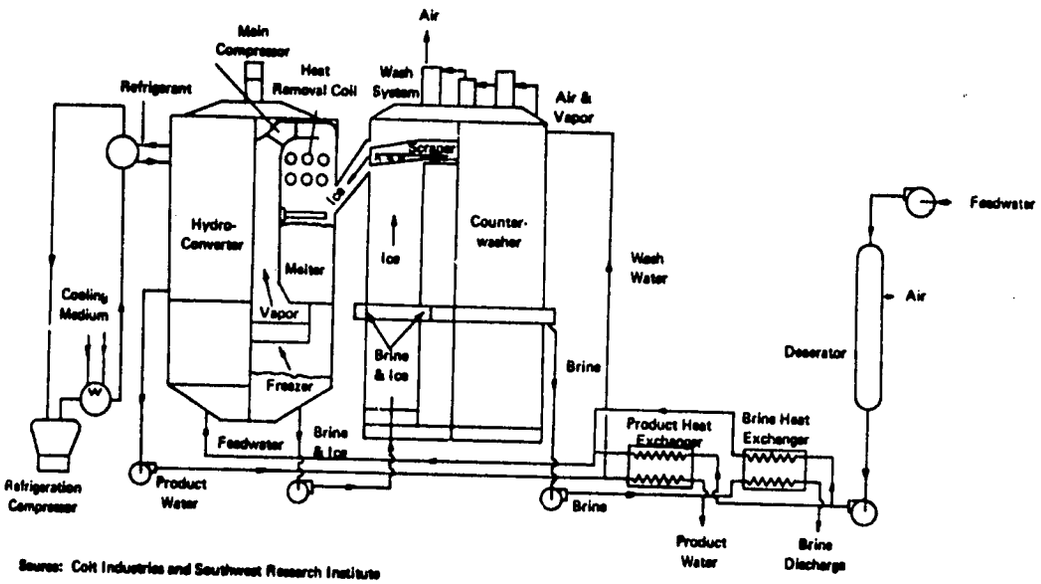


BRINE

FRESH WATER

B. Freezing and Crystallization Processes

1. General: Of the various freezing processes described in Volume I, the vacuum-freezing vapor-compression process has emerged as the most successful candidate. 0.1 to 1.0 mgd vacuum-freezing vapor-compression (VFVC) desalting plants are capable of converting water containing 1,000 ppm to 10,000 ppm total dissolved solids(TDS) to potable water containing 500 ppm or less at costs comparative with electrodialysis and reverse osmosis. Due to the high counter-washing costs at high salinities, the process is not competitive with MSF on seawater.



Source: Colt Industries and Southwest Research Institute

FIGURE IV-6

SCHEMATIC FLOW DIAGRAM OF THE VACUUM-FREEZING VAPOR-COMPRESSION PROCESS

In the vacuum-freezing vapor-compression process, latent heat of fusion is given up when precooled feedwater is introduced into a chamber under very low pressure. Figure IV-6 is a schematic flow diagram of the VFVC desalting system. Feedwater is pumped into the system through a deaerator and then through heat exchangers in which it is cooled almost to freezing by cold brine and cold product water flowing out of the system. The cooled and deaerated feedwater is then pumped into the freezer where a simultaneous boiling-

freezing process takes place. The low pressure causes the water to boil and part of it vaporizes. The vaporization extracts heat from the remaining feedwater, part of which then freezes. Approximately 7.5 lb of ice are formed for each pound of water vaporized, and about 30 to 40 percent of the feedwater is converted into ice or vapor. The concentration of total dissolved solids in the remaining water increases until the water contains 5 to 6 percent TDS. The ice-brine slurry is transferred continuously from the freezer to the counterwasher. Here the ice crystals are washed free of adhering brine and are scraped into a chute leading to the melter. The water vapor produced in the freezer is compressed and discharged into the melter. Here it condenses on the washed ice, melting it.

2. Advancements: Secondary refrigerant freezing, using a high capacity pressurized wash column and high capacity freezing, is under development. The vacuum freeze-ejector absorption process is also being developed as a means of by-passing one mechanical compressor and its inherent limitation on plant size.

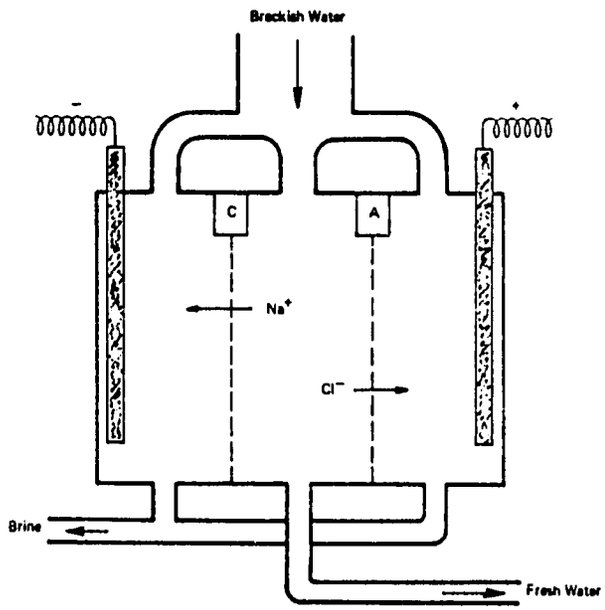
3. Problem Areas: The principal problem of freezing processes, in general, is scale-up to commercial sizes.

The above difficulty, plus those of deaeration at low temperatures and of removing the refrigerant from the product water in secondary refrigerant freezing have, so far, limited the commercial application of the process.

C. Electrodialysis Process

1. General: As already described in Volume I, Section IV-C, 0.144- to 10.0-mgd electrodialysis (ED) desalting plants are capable of converting brackish water containing 1,000- to 5,000-ppm total dissolved solids (TDS) to potable water containing 500 ppm or less. Plants have been in commercial operation since 1956.

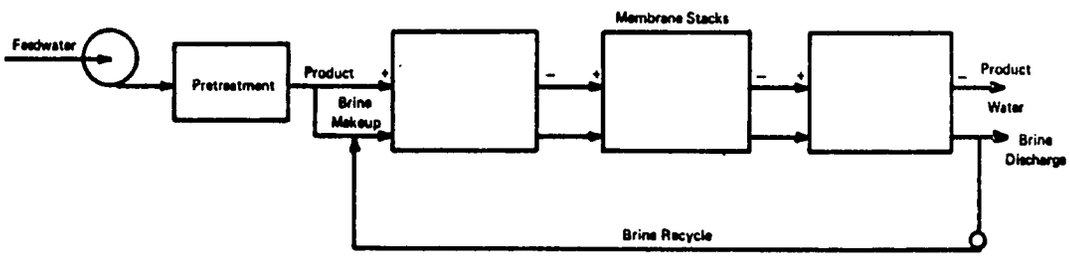
The principle of operation of electrodialysis is shown in Figure IV-7. Salts, when dissolved in water, are present in the form of negatively and positively charged ions. When an electric current is applied, positively charged ions in the brackish feedwater, such as sodium, are forced through the cation-permeable membrane toward the cathode, and negatively charged ions such as chloride are forced through the anion-permeable membrane toward the anode. The water in the compartment between membranes is depleted of salt while the water in the adjacent compartments increases in mineral content.



C – Cation-Permeable Membrane
A – Anion-Permeable Membrane

Source: Southwest Research Institute

PRINCIPLE OF ELECTRODIALYSIS



Source: Southwest Research Institute

Figure IV-7

SCHEMATIC FLOW DIAGRAM OF THE ELECTRODIALYSIS PROCESS

The membrane stack, or the electrodialysis process unit, consists of several sets of anion- and cation-permeable membranes. The quantity of salts removed by passage through one stack ranges from 30 to 50 percent of the entering minerals. Additional stacks are added in series to increase the degree of demineralization. Each added stack is a stage. Volume throughput is achieved by adding additional stacks in parallel.

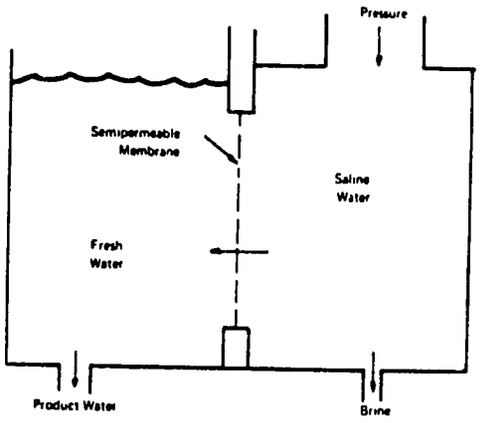
In Figure IV-7 is a schematic flow diagram of an electrodialysis system. Feedwater is pumped through a pretreatment section and then through successive membrane stacks, being processed into product water and a concentrate stream.

2. Advancements: With the construction of the Siesta Keys, Florida 1.2 mgd ED plant and on-schedule completion by June 1, 1972 of the 1.5 mgd Gillette, Wyoming plant, electrodialysis has become a large volume producer of water from brackish sources. Experience since 1960, when the Coalinga, California plant was put into operation, has shown that the plants have good on-stream capability and that corrosion, which plagues distillation plants, is not a problem at ambient temperatures prevailing in ED plants. Thermodynamic efficiency is higher than for any other water-making process.

3. Problem Areas: Major problems are scale potential due to $CaSO_4$ and $Mg(OH)_2$ fouling by iron and manganese, and membrane failure, which is a function of the operating and handling procedures. In the Siesta Keys and Gillette plants these problems have all been handled by minimal pretreatment techniques employing lime coagulation-softening and filtration. Hydrogen sulfide, an additional problem, is removed by aeration.

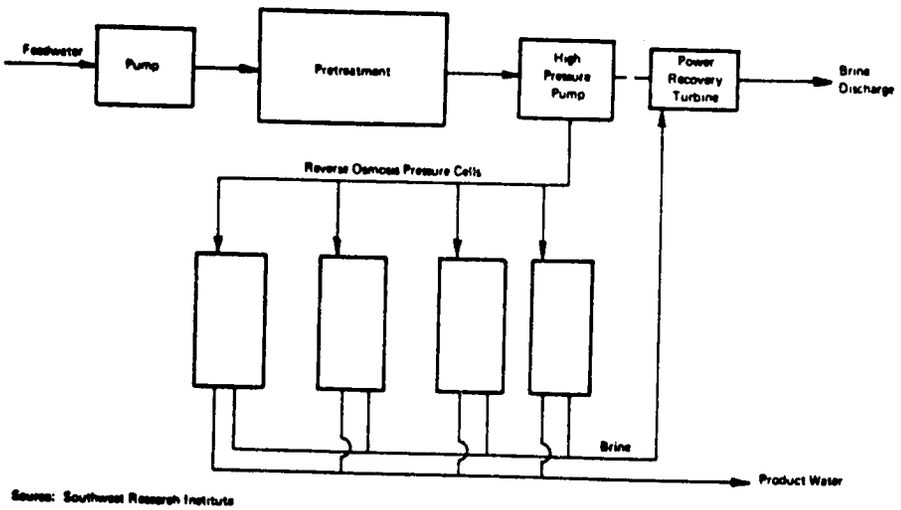
D. Reverse Osmosis Process

1. General: In 1967, when Volume I of this series was being written, reverse osmosis was still in the experimental stage. 0.1 to 3 mgd reverse osmosis (RO) desalting plants are now commercially capable of converting brackish water containing 1,000 to 10,000 ppm total dissolved solids (TDS) to potable water containing 500 ppm or less at a cost generally less than for distillation.



Source: Southwest Research Institute

Reverse Osmosis Cell



Source: Southwest Research Institute

Modular Arrangement of Reverse Osmosis Cells

REVERSE OSMOSIS SCHEMATIC FLOW CHART

FIGURE IV-8

Osmosis occurs when pure water and a salt solution are on opposite sides of a semipermeable membrane and pure water diffuses through the membrane, diluting the salt solution. The osmotic process can be reversed by applying pressure to the salt solution.

Under the reverse osmosis principle, saline feedwater is pumped into a pressure cell at high pressure. This cell is divided into two compartments by a thin semipermeable membrane that allows water to pass through more readily than salt. When the pressure in the saline water compartment exceeds the osmotic pressure, fresh water flows from the saline water compartment through the membrane into the fresh water compartment. The constant flow of water in the saline compartment is necessary to maintain the proper system dynamics.

Figure IV-8 is a schematic flow diagram of a reverse osmosis system. Feedwater is pumped through a pretreatment section which removes feedwater constituents capable of damaging other parts of the system. Feedwater is then pumped to a pressure level high enough to overcome the osmotic pressure and to produce the designed plant capacity. The fresh water which permeates the membrane is collected as product water. The brine stream is reduced in pressure by a power recovery turbine which helps drive the high pressure pump.

2. Advancements: Three systems are now in commercial use: spiral-wound sheet, cast tubular, and hollow fine-filament.

a. Spiral wound membrane processes use large areas of flat sheet cellulose diacetate. The capacity of semi-permeable membranes is a function of their efficiency in rejecting salt. Membranes which will reject 90% of the NaCl in the saline solution have a capacity of 22-30 gals/day per square foot of effective membrane surface. With membranes capable of 98% rejection, capacity drops to 9-14 gpd/ft^2. In order to attain the large volumes of deionized water desired in most installations, many square feet of surface are required. To accomplish this, long strips of flat sheet membrane material are edge-sealed to an impervious membrane with a highly porous material between them. To conserve space

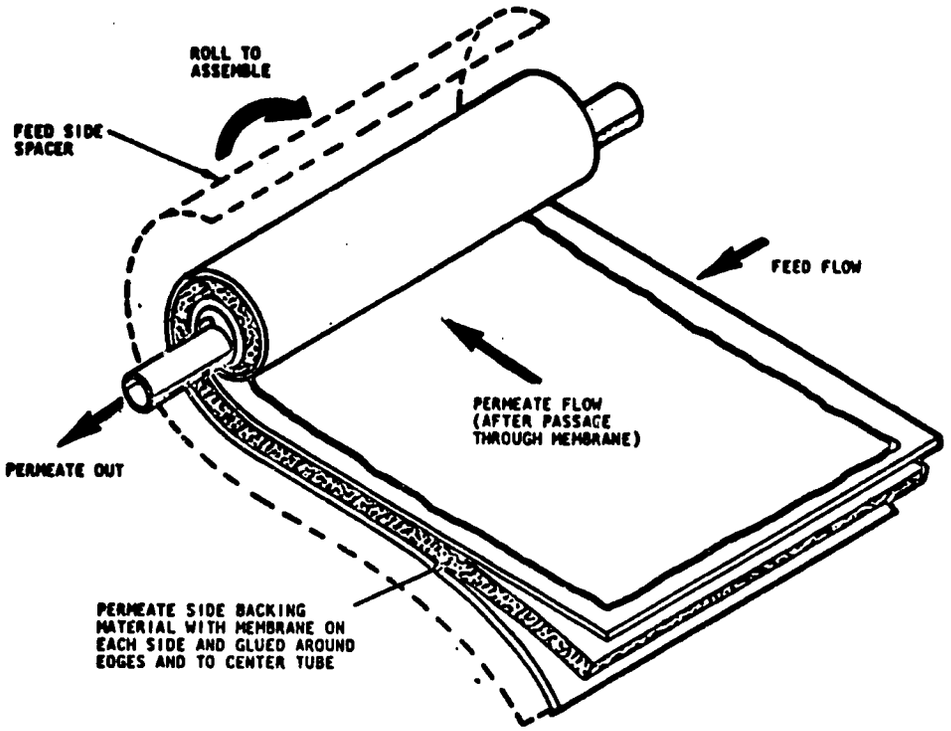


FIGURE IV - 9A SPIRAL MEMBRANE MODULE

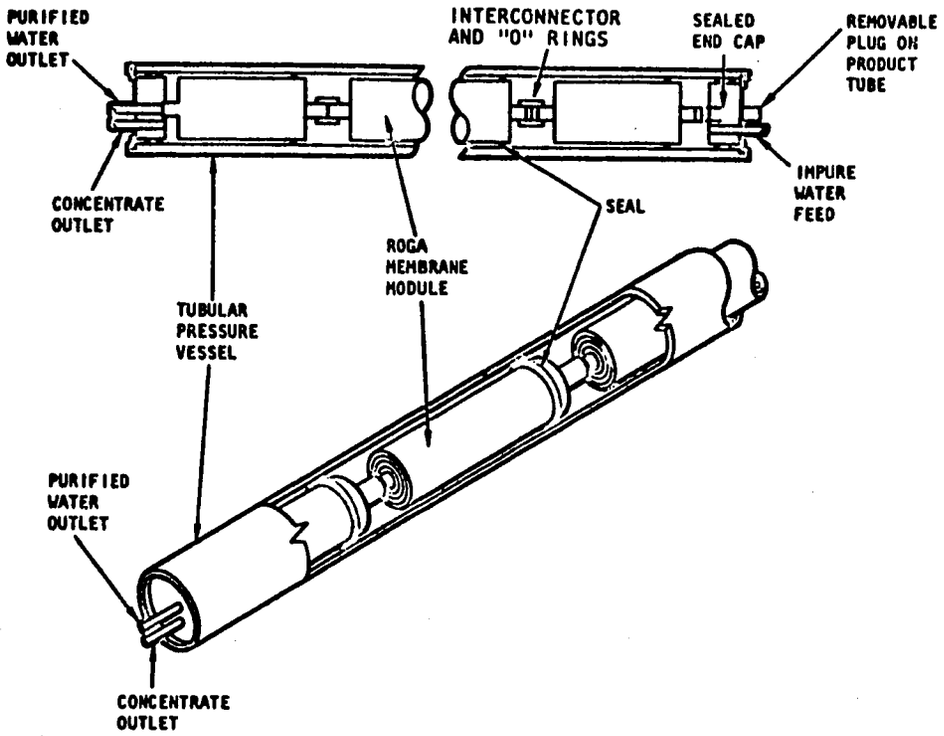


FIGURE IV - 9B A TUBULAR PRESSURE VESSEL

these are then spirally wound and inserted into a pressure shell. The desired water leaves the sandwich thru the porous material. This construction is shown in Figure IV-9A. Up to 6 spiral-wound cartridges are joined together within a single 4" diameter pipe shell as shown in Figure IV-9B, to form a module, designed for a 600 psig operating pressure.

b. Another variation of the cellulose diacetates membranes uses the semi permeable material cast in the form of tubes of approximately one inch diameter. These tubes are mounted in a configuration resembling a tubular heat exchanger with the shell side space filled with a porous material which will support the tubes while they are under pressure. Such a material may be sand in which the particles are bonded to prevent their shifting. In other configurations the tubes may be helically wound and mounted in a shell so as to resemble a helical flow heat exchanger. The tubular type units have the advantage that it is easy to flush the saline water side to prevent build-up of salt or other contaminants. Another advantage is the ease of replacing a faulty tube. Disadvantages are the larger building area required to house the same capacity unit, and difficulties with the end-seals. This construction is shown in Figure IV-10. Design pressures can exceed 1000 psig.

c. The hollow fine-filament units use a bundle of nylon capillary tubes, cemented into a cast epoxy tube sheet at one end and a cast epoxy plug at the other. The pure water diffuses thru the filaments and passes along them and out at the tube sheet end. Concentrated saline solution passes into a porous annulus surrounding the bundle and is led to the outside and to waste. This construction is shown in Figure IV-11. Though the flux is small by comparison to other constructions, the large surface area of the capillaries and the fact that they are self supporting makes them attractive.

d. <u>Three-way Combination Reverse Osmosis System for High Percent Recovery of Product Water</u>: The three basic types of reverse osmosis modules can possibly be combined into a system wherein each type is utilized most effectively when staged on the brine stream.

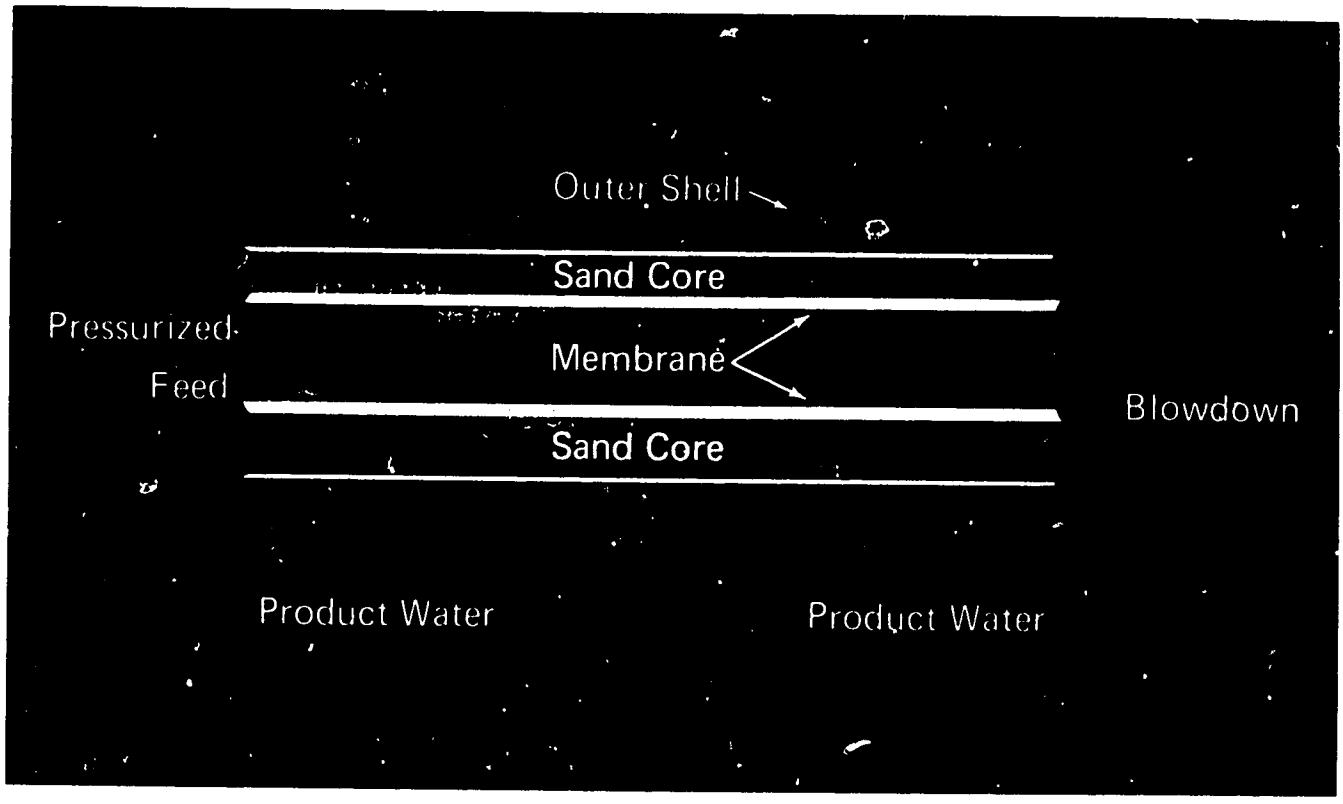
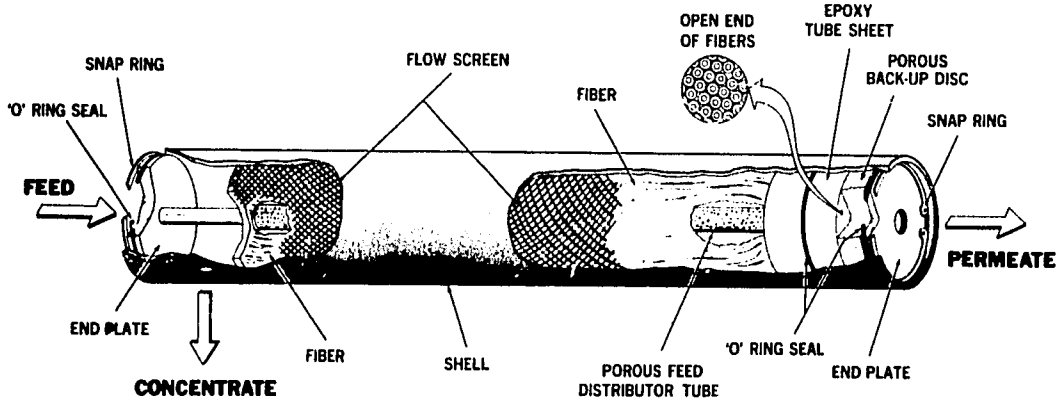


FIGURE 1V-10



CUTAWAY DRAWING OF PERMASEP® PERMEATOR

FIGURE 1V-11

HIGH RECOVERY REVERSE OSMOSIS

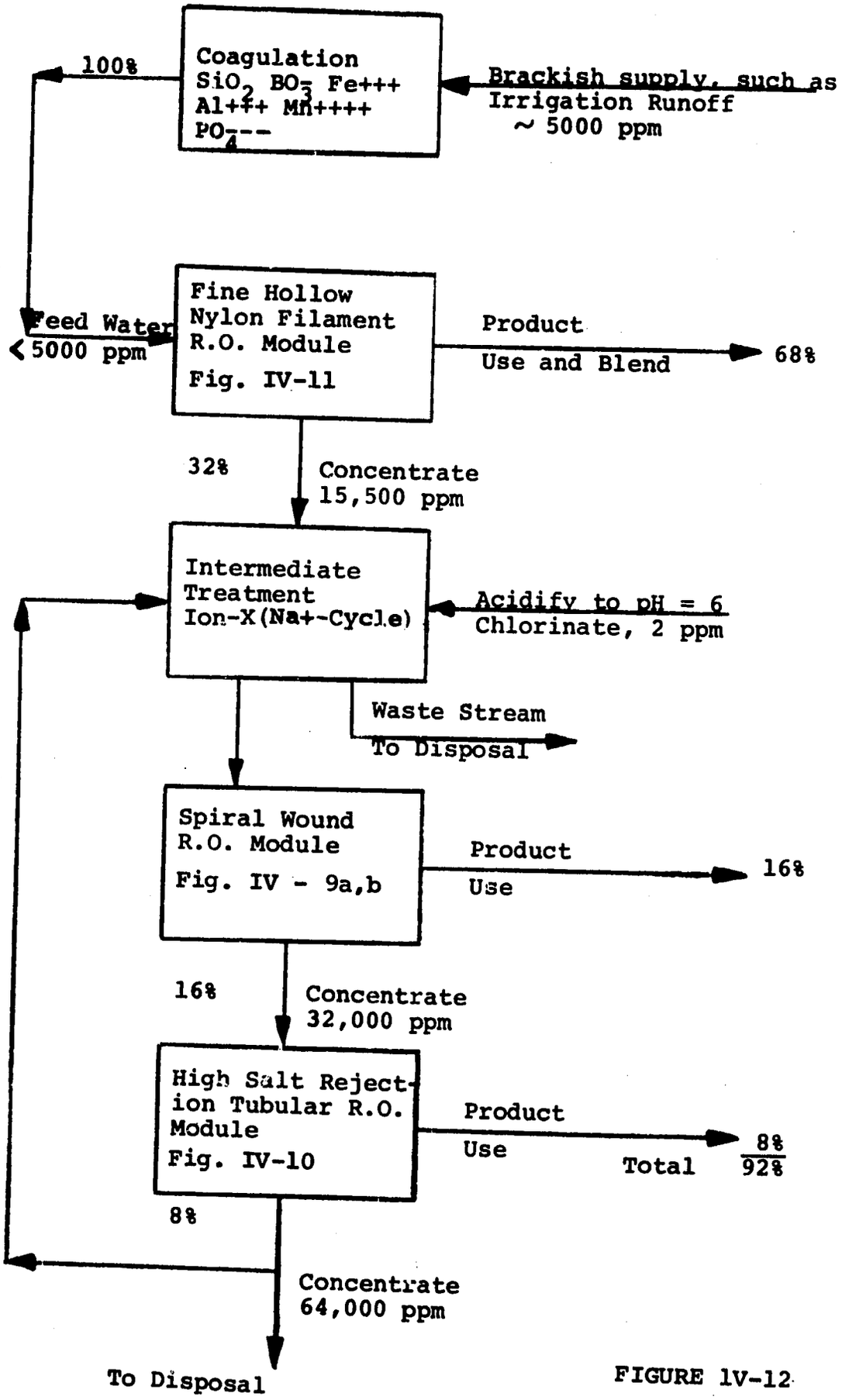


FIGURE IV-12

Starting with a 5,000 ppm brackish water supply and recovering 93% of the feed as product would concentrate the brine stream twelve-fold so that only one-twelfth of the flow work would be done on the brine, and no pressure recovery equipment is needed. At one-twelfth of the original volume, the brine disposal problem would be greatly reduced and, if sodium-cycle, ion exchange softening is used, most of this brine can be used to regenerate the resin. Notice from the flow sheet for the combination unit, Figure IV-12, that the feed stream becomes less for each reverse osmosis unit in sequence. If the flow rate entering the first unit is 100, the rate entering the third unit would be only 16, so that with equal size units there would be only one-seventh as many units in parallel for the third stage as for the first. For each particular case, based upon stated feedwater and brine compositions, the hydrodynamics would have to be worked out separately. Stagnant pockets and concentration potentials at membrane surfaces should be eliminated, preferably without using recirculation.

The operating pressures employed in the different stages will depend both upon the feed salinity to the stage and upon the membrane characteristics as mentioned in Volume 1, Section IV-D. Generally, for brackish water up to 6,000 ppm TDS it will amount to about 500 psi in the first stage and 1,500 psi for stages operating at or near seawater concentration. One effect of high salinity and the associated high operating pressure is to shorten the lifetime of the membranes used in that stage. Another is to require interstage ion exchange and repressurization equipment in order to prevent precipitation of scaling components at the higher salinity.

The advantage of the combination system is to provide for the first time an economical means for water supply together with economic brine disposal for small inland rural communities that would otherwise have to import drinking water.

(1) <u>Pretreatment</u>: An engineering study of pretreatment systems for membrane desalting processes[5] has been made and shows that the cost of pretreatment is extremely high for the benefits derived,

being, for Fresno, Calif. irrigation return flow 20¢/Kgal. (5) With the combination system a minimum feed treatment using coagulation to remove iron, manganese and phosphorus, followed by rapid sand filtration is envisioned, for which the cost is nominal. A typical pretreatment plant may also involve a carbon filter to remove organic loading and a manganese zeolite filter for complete iron and manganese removal. Hopefully, the last two units can be avoided by placing a nylon hollow fine-fiber reverse osmosis unit first in the series, as these units have wide latitude for the water composition.

(2) Nylon Hollow Fine-Fiber Reverse Osmosis Units:

With brackish water of about 5,000 ppm which has received a preliminary filtration, the nylon hollow fine-fiber unit can recover up to 70% as product, leaving 30% as a concentrated brine. Beyond 70% recovery there are internal hydrodynamic problems. The unit has advantages of relatively low capital cost, high through put, wide latitude as to feed pH and composition; also absolute clarity is not essential and no chlorination is required. As shown by the flow sheet, (Figure IV-10), the concentrate can be about 15,500 ppm of TDS.

(3) Intermediate Feed Treatment: The pH of the concentrate stream will have to be adjusted to approximately 6 before going to the next reverse osmosis unit. This has the dual function of preventing alkaline scale deposition, which would cause compaction of cellulose acetate membranes, and also of stabilizing them against hydrolysis. Chlorination to about 2 ppm residual is required at this point in order to destroy bacteria which would otherwise feed on the cellulose acetate. Softening by means of sodium zeolite will be used so as to stay below the scaling limits of calcium sulfate for the remaining processes. It may be possible to utilize waste brine from the last stage at 62,000 ppm to regenerate the sodium-cycle ion exchanger. This will drastically reduce, if not eliminate, need for purchasing regenerant salt.

Following intermediate feed treatment the feed pressure will have to be boosted before the feed enters the spiral wound reverse osmosis stage.

(4) Spiral Wound Module: The feed stream volume has been greatly reduced by the first stage so that to maintain its turbulence the spiral wound unit possesses outstanding advantages. It can operate at about 50% product water recovery, furnishing thereby 16% of the total product and reducing the brine stream to 16% of the original value. The membrane can be selected to produce product water of potable quality when the unit is operated on a feed of 15,500 ppm. The capital cost of these units is attractive, and absolute clarity of the feedwater is not essential. As with the fine hollow filament modules, the spiral wound modules are presently in mass production and have proven themselves on a variety of industrial and municipal applications.

(5) Tubular Reverse Osmosis Modules: From the spiral wound module 16% of the original feedwater enters a high salt rejection reverse osmosis tubular unit at 31,000 ppm. If this unit is purchased with special high salt rejection tubes the product water will be 8% of the original feed and, when blended with the production of the preceding two units will result in 92% overall recovery of potable water. About 8% of the original feedwater will be rejected as waste brine. However, as already suggested, this can possibly be used for regenerating the ion exchange unit, in which case the total reject stream will be that leaving the intermediate feed treatment system. The unit must be designed for 1000 psig.

(6) Alternatives: The main problems in reverse osmosis are encountered following the first unit, after the solids have been concentrated to over about 7,000 ppm. To avoid these in the subsequent units, it would be possible to substitute the vapor compression evaporator, using $CaSO_4$ seeding, which has been described in Section IV, A-3.

The reverse osmosis unit should concentrate up to the scaling limit, or up to about 70% water

recovery and the remaining brine then fed into the VC unit. 8000-12,000 ppm suspended solids should be used in the VC unit and temperature differences of not more than 6°F, to allow scale-free operation. The brine can then be discharged in a small volume at up to 250,000 ppm of TDS.

3. Problem Areas: The success of reverse osmosis processes depends upon careful pretreatment of the feed stock. Iron, manganese and hydrogen sulfide must be removed and the pH controlled on the acid side, pH6.0, to prevent alkaline scale and membrane hydrolysis. Colloidal solids, oils and precipitated solids will physically block the membranes and must be removed before they contact them. Lime-alum coagulation, if practical, is the least costly and most generally satisfactory pretreatment. (5)

The cellulose acetate materials are susceptible to the action of bacteria and algae. Consequently, the saline solution is chlorinated to a residual chlorine content sufficient to prevent their growth. The chlorine residual must be carefully controlled as excess chlorine will attack the membrane materials.

The nylon hollow fine-filament units have certain advantages over the other processes, being much more resistant to hydrolysis and to the action of bacteria and algae. This eliminates the necessity of chlorination and acidification, except to control scale. However, the hollow fibers are pressure limited to about 600 psig and will collapse if over-pressured.

The principal problem with all designs is to provide a permeable backing, or septum, that will support the membranes against the pressures of up to 1,500 psig that are required for sea water conversion. The membranes are particularly vulnerable to damage at the edge and end seals. Even where suitable backing has been provided, moreover, new membrane materials, possessing extra strength, toughness, and durability must be found if sea water is ever to be processed economically in a single pass.

E. Feed and Product Water Treatment

1. General: The subject of feed treatment to reduce corrosion and to effect scale control has been extensively reviewed in Vol. I of this manual. In distillation processes the principal problem is the formation of calcium carbonate, magnesium hydroxide and calcium sulfate scales. The first two are controlled by acidification, followed by atmospheric decarbonation and vacuum deaeration. If sulfuric acid is used for acidification, they are converted to $CaSO_4$ and $MgSO_4$. $MgSO_4$ is highly soluble and presents no further problems. $CaSO_4$ precipitation must then be controlled by careful attention to its concentration and temperature.

Calcium and magnesium may be reduced by ion exchange with sodium to yield a non-scale forming feed. Cost analysis of this process shows it to be much more expensive than acidification. However, it is anticipated that higher temperature operation with more stages may result in compensating savings. (5)

The membrane processes are not affected by the inorganic scales to the same extent as processes utilizing heat transfer surfaces. With them flocculent iron and manganese hydroxides and certain organic compounds fill the ionic passages in the membranes. Ion exchange processes will oxidize and remove the iron and manganese compounds. Flocculation followed by filtration can remove a number of organic materials.

Product water should be adjusted to a slightly alkaline pH and to a hardness of over 50 ppm to render it non-agressive and chlorinated to a positive residual before use in a municipal distribution system.

2. Advancements: Continuous ion exchange using the concentrated sodium brine effluent to regenerate the ion exchanger has been tested. 70-80% calcium removal has allowed scale-free operation at 275°F in the evaporator and blowdown concentration of 2.0. The method proves to be more expensive than acidification even without purchase of NaCl for regeneration.

Scale prevention by "seeding" the solution with calcium sulfate crystals and heating to temperatures above those used in the distillation process has been successful in scale prevention in a pilot plant. The crystalline $CaSO_4$ is removed from the saline water in a carbon filter and recycled. A concentration factor of 1.47 was reached. The solids can be carted away.

F. Brine Disposal

1. General: Waste brine disposal into surface waters is undesirable in inland areas. Two alternatives present themselves, evaporation to dryness, and concentration in ponds followed by pumping into underground formations having no value as water sources and which positively cannot contaminate other water sources. Evaporation to dryness may only be practiced in areas where the annual evaporation rate greatly exceeds the annual precipitation and there is no net annual runoff. Under other conditions the crystalline solids or concentrated brines must be disposed of so that they cannot contaminate surface or potable subsurface aquifers. National stockpiling is sometimes feasible. (43)

Evaporation in ponds to produce a concentrate that may then be pumped underground is a more common procedure. The holding of concentrated brines in ponds for temporary storage and/or atmospheric evaporation presents the problem of the brine leaching into the earth and contaminating fresh water supplies. The study of effective low-cost liners for such ponds is continuing.

Plant effluents may be pumped underground directly, without concentration, but require much more disposal equipment. Where liquids are to be disposed of underground every effort must be made not to contaminate possible sources of potable water. This entails use of several strings of casing and cementing of the injection well completely through any aquifer which is a potential source of usable water. (7)

This subject has been covered in Desalting Cost Calculating Procedures (13) and other sources listed therein. A further description appears in Appendix C, Estimating Aids and in Water-1970 (57).

2. Advancements: Studies are under way involving the discharge of concentrated brines into the sea. These denser fluids mix slowly with normal seawater resulting in concentrated zones at depths in the receiving body of water. In these zones, besides the effect of greater salinity, there is also an increased concentration of copper ions leached from the plant surfaces.

In the Diablo Canyon Prototype Desalting Plant, titanium tubes were employed to prevent pollution with copper and diffusional systems for outfalls were developed to remedy the other difficulties.

Studies of increasing water evaporation by incorporating electrodialytic concentrators ahead of solar ponds show promise of success. A study of costs of deep-well injection of brines (7) was completed which showed a wide range of these costs. The method was feasible in all but one of the locations studied.

3. Problem Areas: The problems which face the distillation processes in coastal areas are the effects of the outfall brines on the ecology of the receiving body of water. The same is true to a lesser extent of the membrane processes. They, however, do not introduce copper ions. The studies just mentioned, which are continuing, are directed toward these problems.

The problems facing deep-well disposal are cost problems based upon the depth at which suitable acceptor strata are found. In certain areas there are no such strata available and alternate disposal methods must be used. Similarly, evaporative disposal is only practical where meteorological conditions permit. Since these are beyond our control, the only problems are the cost factors involved in minimizing contamination by the effluents stored in evaporation ponds. Controlled release during flood seasons is one method of relief sometimes practical.

The problems associated with brine disposal have been discussed in greater detail in Section X of Appendix C.

G. Materials of Construction.

1. General: The state-of-the-art of combating corrosion, up to 1967, was reviewed in Volume I, Section IV-D. In combating saline water corrosion of metals, attention has historically been centered upon copper and its alloys. More recently nickel and its alloys, aluminum alloys and titanium have entered this picture. Of these, only aluminum is cheap on a per pound basis and cheaper still in service because it is much lighter than the other metals mentioned.

The destructive action of turbulent seawater, especially when oxygenated, has long been observed in its action on ships' propellors. Since turbulence and heat transfer efficiency are related, it is necessary to use materials in this service which resist erosion-corrosion. These have proven to be the copper-nickel alloys and, to a lesser extent, Ni-Al bronzes.

Iron and its alloys present high strength and low initial cost. However, other than the austenitic alloys, they all suffer corrosion in seawater, and the austenitic alloys are subject to stress corrosion cracking in salt water environments and under some other conditions. Their use can only be justified in areas where they can easily be replaced in case of failure.

Many non-metallic materials have excellent corrosion-resistant properties under selected conditions. Many lack physical strength but are usable as protective coatings. Portland cement concrete is an excellent low-cost structural material but lacks suitable corrosion resistance in seawater principally through attack of its reinforcing structure. It can be protected by linings. Other cements have been developed which have superior properties. A number of plastic materials developed in recent years have limited uses in desalination plants.

2. Advancements: Recent work on aluminum alloys has shown them to have adequate seawater corrosion resistance in the absence of copper and other heavy metal ions. The corrosion resistance of aluminum is due to

the formation of Al_2O_3. In deaerated waters where oxygen is not present to repair a break in the Al_2O_3 film, aluminum is readily attacked by water. This is also true where slimes mechanically block oxygen from the surface. Thus, the greatest potential uses for aluminum appear to be in the membrane processes which do not require deaeration.

The use of aluminum alloys drawn into heat exchanger tubing has shown promise of reducing the capital cost of such items. One complete 50,000 gpd distillation pilot plant is being used to test these under OSW supervision. Two years' experience has shown extreme promise. These tests are being continued in an effort to find the alloys most suitable in this service. Aluminum alloy plate has also proven resistant in parallel tests. Such plate is usable for shells and vessel internals. Aluminum structural materials give good service. In all these applications metallic couples must be carefully avoided to prevent electrolytic corrosion of the aluminum.

Aluminum brasses are proving to be competitive with the more expensive nickel-bronzes. Apparently, the aluminum contributes protective Al_2O_3 films on the surfaces. Alloys 687 (2% Al) and 688 (3.5% Al) are being studied further on tubular goods.

A number of experiments have been performed on ferritic alloys of iron. To date, no low-cost alloys have proven successful. Testing has been carried out at temperatures approaching 212°F in deaerated saline water. These alloys exhibit pitting corrosion when the alloying element is under approximately ten percent.

Stress corrosion cracking of austenitic stainless steels is influenced by oxygen at the surface of the metal. The susceptibility of these steels is a function of their carbon content. The so-called "L" grades of types 304 and 316 SS exhibit superior resistance. Welding of stainless steel is responsible for the deposition of iron carbides in the heat-affected zones. Where welding must be done, the "L" grades must be used.

Concrete vessels proposed in vacuum service for the large volume plants have presented many structural

problems. Concrete has proven to be susceptible to corrosion by sulfate ions. Linings of calcium aluminum silicate cement have proven able to protect mild steel surfaces exposed to seawater at temperatures up to 120°C.

The tabulation on the following page lists typical materials for the two 2.25 mgd VTE plants to be constructed in 1972 in the Virgin Islands.

3. Problems: Work must still be continued on the corrosion resistance of aluminum alloys in order to develop those which are most resistant in saline water applications. Aluminum and its alloys are weldable with shielded arc techniques. Insufficient information exists at present on the effect of such welding on the corrosion resistance of the parent metal.

The development of non-metallic saline water-resistant coatings has been continuing at a rapid pace. The methods of effective bonding of these materials to low-carbon steels to give long-term protection must be developed further in order that the use of low-cost steel may be extended.

TABLE IV-2

Materials of Construction for Typical Desalting Plant Services and Components

Equipment	Service Fluid	Fluid Conditions	Size or Type	Base Material	Coating or Lining
1. Evaporators					
Condensers and tubes	Seawater-steam	100°-125°F	3/4" minimum	Titanium	-
	Deaerated brine-steam	125°-250°F	3/4" minimum	90-10 CuNi	-
Evaporator tubes (Vertical Enhanced)	Deaerated brine-steam	125°-250°F	3" dbl fluted (31)	90-10 CuNi	-
Structural steel	Brine	100°-230°F	-	Carbon steel	1" Pre-krete
Structural steel	Brine	230°-250°F	-	Carbon steel	3/32"Tp316L SS.
Structural steel	Steam	100°-200°F	-	Carbon steel	15-mil epoxy
Structural steel	Steam	200°-250°F	-	Carbon steel	3/32"Tp316L SS.
Product Water Components	Distillate	100°-250°F	-	90-10 CuNi or Tp316 SS.	(solid or clad on c.s.)
Water boxes[2] and tube sheets	Deaerated brine & seawater	100°-250°F	-	90-10 CuNi	(solid or clad on c.s.)
Entrainment Separator & internals	Steam & brine	100°-250°F	-	Tp 316 SS.	
2. Deaerator	Seawater	120°F		Carbon steel	16mil coal-tar epoxy
3. Brine Heater					
Tubes	Deaerated brine-steam	250°F	3/4" minimum	90-10 CuNi	-
Shell	Steam	250°F	-	Carbon steel	-
Water boxes and tube sheets	Deaerated brine	250°F	-	90-10 CuNi	(solid or clad)
4. Piping					
Process	Deaerated brine	100°-250°F	-	90-10 CuNi or Bondstrand below 180°	-
Vents & drains	Seawater	125°F	-	Cast iron	Cement
	Misc.	100°F	-	Tp 316 S.S.	-

TABLE IV-2 (continued)

Equipment	Service Fluid	Fluid Conditions	Size or Type[1]	Material: Casing	Material: Trim	Coating or Lining
5. Pumps						
Product water	Distillate	100°F	H or V	Ni-Al-Bronze	Tp 316 S.S.	-
Recirculation	Brine	100°F	H or V	Ni-Al-Bronze	Tp 316 S.S. or Monel	-
Makeup	Brine	100°F	H or V	Ni-Al-Bronze	Tp 316 S.S. or Monel	-
Blowdown	Concentrated brine	100°F	H or V	Ni-Al-Bronze	Tp 316 S.S. or Monel	-

Notes

1. H or V means horizontal or vertical
2. Water boxes may alternately be constructed of fiberglass below 180°F.

Burns and Roe, Inc.

V. DESCRIPTION OF TYPICAL PLANTS

A. General

Section IV of Volume I of this Manual listed 8 typical examples of desalting units being developed in 1967. The present status of these units is shown below:

1. Distillation Processes	Status 1972
Vertical Tube Evaporator OSW#1, Freeport, Texas	Replaced by VTE/MSF after successful operation.
Multi-stage Flash Evaporator, OSW#2, San Diego, California	Moved to Guantanamo, Cuba; operating at 750,000 gpd.
Multi-effect Multi-stage Flash, OSW, San Diego, California	Clair Engle plant now operating jointly with LMC process.
Vapor Compression, 2-stage OSW#4, Roswell, N.Mexico	Troublesome. Now being tested with water-jet compressor.
Solar Stills, Dodecanese Islands, Greece.	Operating satisfactorily.
2. Freezing Process	
VC - Vacuum Freezing, Wrightsville Beach	Successful. A new modification now being tested by Colt Industries.
3. Membrane Processes	
Electrodialysis, OSW#3, Webster, S. Dakota	ACI Unit has successfully supplied Webster with water since 1962.
Reverse Osmosis, 20,000 gpd commercial unit	Now successful @ 0.2 mgd in 3 basic configurations

B. Advancements

The size, performance and quality of the distillation plants and of the membrane plants have been greatly improved since 1967, with reverse osmosis having made the greatest strides. The following plants are cited as typical for 1972 and are described in this section:

1. Distillation Processes

MSF Process, 5 mgd	Jeddah, Saudi Arabia
VTE/MSF Process, 3-15 mgd	Fountain Valley, Calif.
VTE/MSF Process, ½ mgd	Odessa, Texas

2. Membrane Processes

Membrane Processes	Reference Figures
Electrodialysis, 1.5mgd	Fig. V-5
Hollow Fiber Reverse Osmosis 0.112mgd	Fig. V-6
Spiral Wound Reverse Osmosis 0.1mgd	Fig. V-7
Tubular Reverse Osmosis 0.025 mgd	Fig. V-8
Tubular R.O. Seawater Conversion Module	Fig. V-9

C. Problems:

Fouling with organic and inorganic scale-forming compounds is the principal problem with membrane processes, while high temperature scale and corrosion still plague distillation plants. Better remedies are continually being found. For example, a 1 mgd horizontal-tube-evaporator (HTE) using aluminum tubes and temperatures below 170° to minimize scaling and corrosion is being built at Eilat, Israel, based upon tests with smaller unit which have indicated that this construction is very economical.

DESIGN OF A 2.5-MGD UNIVERSAL DESALTING PLANT

By

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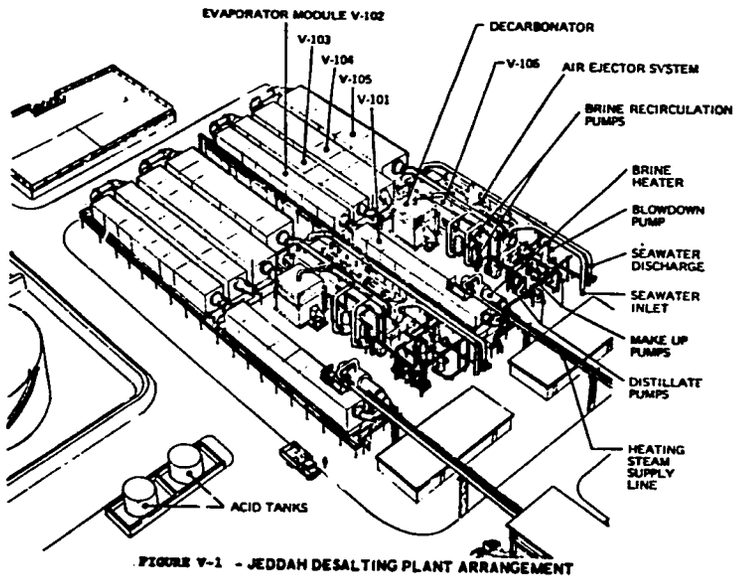


FIGURE V-1 - JEDDAH DESALTING PLANT ARRANGEMENT
(Courtesy of AquaChem, Inc.)

The United States Department of the Interior, Office of Saline Water, conducted studies prior to 1965 which showed that some standardization of multi-stage flash distillation plant designs would be desirable. The advantage of such standardization would be the provision of a "yardstick" with which to measure future progress in the world-wide development of desalting plants. The technology of the standardized design could be made available to emerging nations to permit the economic and industrial growth so vitally needed to provide for the needs of the soaring population. In the overall conceptual planning done by OSW, all desalting plants were divided into three groups sized by output capacity, small, intermediate, and large. The capacity in these ranges were to be approximately 1 to 10 MGD, 10-50, and the largest 50 MGD and above. The OSW determined that the most practical yardstick would be a standardized design in the smallest range. An output of 2.5 MGD of fresh water was selected because a plant of this size is about as big a plant as can be prefabricated, shipped and erected with comparative ease. Several of these plants may be grouped in trains to increase a total output. It was generally agreed that it would be practical to group together up to 4 trains, making a total of 10-MGD output, before it would be necessary to design larger capacity trains. The OSW required that the standard plant design represent not only the latest in desalting technology but also provide a design which would be close to the optimum design over a wide range of economic factors. The existence of the standard plant design will eliminate costly and time-consuming engineering studies which would otherwise be necessary to determine the optimum economic design for a particular locality.

The basic "yardstick" design will provide, in addition to its value as a practical contemporary design, a standard of comparison against which future improvements in plant and process design may be evaluated.

The standard design comprises the complete optimization of 11 different plants which vary in width and number of stages. Seven plants for use with acid-makeup treatment are provided having from 24 to 50 stages. Four plants for use with polyphosphate makeup treatment are provided, and these have wider stages than the acid treatment plants because of the higher brine flow required. The polyphosphate plants have from 24 to 32 stages. Under certain conditions, where fuel costs are exceptionally low, and fixed charges are high, a smaller number of stages will be feasible. However, these conditions fall outside the optimization limits of the standard plant computer program.

All of the design parameters, including steam cost, fixed charge rate, and power costs were included in the computer program which optimized the design for each combination of input parameters. Out of 60,000 cases examined, several hundred optimum plants were printed out in the computer output. From this large number of optimum plants, eleven (11) plants were selected, each satisfying the design requirements and each

providing fresh water at a cost no more than 1 percent above the optimum plant design cost. The standard design is based on the modular concept, each module having several stages, and being limited in length only by the maximum length of condenser tubes available. Although condenser tubes in excess of 110 feet in length are currently available in the United States, it was considered desirable to use module lengths of about 80 feet so that the tube drawing equipment available throughout the world might be used to manufacture the condenser tubes.

The final standard design consists of about 65 module design and process drawings, a volume of design specifications, a volume of bidding specifications, and a volume, in two parts, of instructions to the user on how to select one of the eleven plants for his particular site.

The 5-MGD desalting plant to be completed at Jidda, Saudi Arabia, in 1969, is being constructed for the Saudi Arabian Government, under the sponsorship of the United States Department of the Interior's Office of Saline Water. The use of standard design for this plant did, in fact, reduce the engineering costs for the Jidda project considerably, and provides an opportunity to apply the standard plant concept.

The natural conditions of the Port of Jidda are similar to those prevailing at the many arid coastal deserts throughout the world. Both the sea-water temperature and the salinity are relatively high compared to the open sea, or to areas where rainfall is greater. While the standard plant design was calculated for 85° sea-water temperature, the Red Sea at Jidda is closer to 87° on the average. The salinity at Jidda averages between 44,000 and 44,500 ppm TDS. For these reasons, it was determined that the standard plant would have to be operated at slightly different conditions than were provided in the standard plant design. The condensing water temperature was increased from 85° to 87° and the concentration ratio was reduced from 1.5 to 1.4. This necessitated some hand calculations to check the size of the makeup and blowdown pumps and to check the condenser surface.

The Jidda plant had to be designed as a dual-purpose plant because it was being built in conjunction with two 25,000-kw turbo-generator units. The steam for desalting plant brine heater is supplied from condensing-type extraction turbines, each turbine supplying steam to one of the 2.5-MGD trains. Bypass lines from the boilers to the desalting plant are provided in the event independent operation of the desalting plant becomes desirable.

The selection of the standard plant from the 11 designs available, began with the decision to use acid-makeup treatment instead of polyphosphate treatment. This permitted the use of a lower fixed capital cost plant but required that a supply of sulphuric acid be developed. From the seven acid plants, one plant had to be selected. The standard design sets forth a method for selection of the optimum plant based on the

prevailing economic conditions. This method of optimizing the dual-purpose plant resulted in a recommendation that a desalting plant having a specific performance ratio be used. Accordingly, the standard acid plant having a performance ratio closest to the dual-purpose plant optimized value was selected. The selection of the design and the procurement of the plant went forward on this basis. The plant selected has two trains, each with 6 module design, having four 8-stage modules, one 7-stage module, and one 3-stage heat rejection module. These modules were arranged, together with brine heaters and auxiliary pumps, so that each train could be operated completely independently from the other.

The standard design provides an option in the design of evaporator modules. The owner may select either the standard design of evaporator accompanying the specifications or he may select the specification which permits the use of a proprietary design which might be offered by the manufacturers of desalting apparatus. The Saudi Arabian Government was interested in the standard design because of their awareness of their own future desalting plant needs throughout Saudi Arabia. The presence of a plant constructed in accordance with the standard design, actually in operation in their system, will provide them with a yardstick to compare development of plants to satisfy their own water needs.

In that section of the technical specifications of the standard design which covers the standard designs modules, certain exceptions to the design are permitted in order that manufacturers may use the standard design drawings to construct the plant and yet guarantee the performance of the plant. Variations, within certain limits, from the standard design in condenser surface, stage width, tube pitch, stage elevation, and in certain other areas are allowed. In the Jidda project, for reasons already explained, proprietary design specification was not permitted.

Figure 1 shows the cycle diagram for a single-effect, multi-stage, flash evaporator of the type being built for the Jidda plant.

Seawater is heated and fed through a series of chambers at successively lower pressures, corresponding to the saturation temperature of the water in each chamber. In each stage, some seawater flashes into steam. The remaining seawater, cooled as a result, flows into the next stage through a control orifice. The vapor that flashes off proceeds through mist separators to a bank of condenser tubes, where it is collected in product water trays.

The seawater flows through the complete series of stages, cooling and giving up fresh water in each. It also becomes more saline in each stage. At the end of the flash-down, the cool brine is pumped back through the heat recovery stage condenser tubes where it recovers the heat given up by the condensing fresh water. Thus, the recycled brine is progressively heated until it leaves the first, or highest temperature stage. Finally, a brine heater brings it to the top temperature required by the process.

The recycle brine steam leaving the lowest pressure stage cannot be used in the condenser of that same stage because no temperature difference exists. Instead, it is piped directly to the fourth stage from the lowest, where it can efficiently condense to the fresh water vapor. Cooling water for bypassed stages is pumped into the condenser directly from the sea. After cooling these three stages, it is returned to the sea, the system heat sink. Thus, the entire plant is divided into a heat recovery section and a heat rejection section.

Recycling the brine increases its salinity with each cycle, raising its boiling point and reducing efficiency. This could lead to precipitation of scale-forming chemicals and scaling of condenser and brine heater tubes. Further, cycling reduces flow because fresh water is lost. By adding seawater, makeup flow volume is maintained and salinity kept below the scaling value. As seawater is fed into the evaporator, concentrated brine is blown down to the sea.

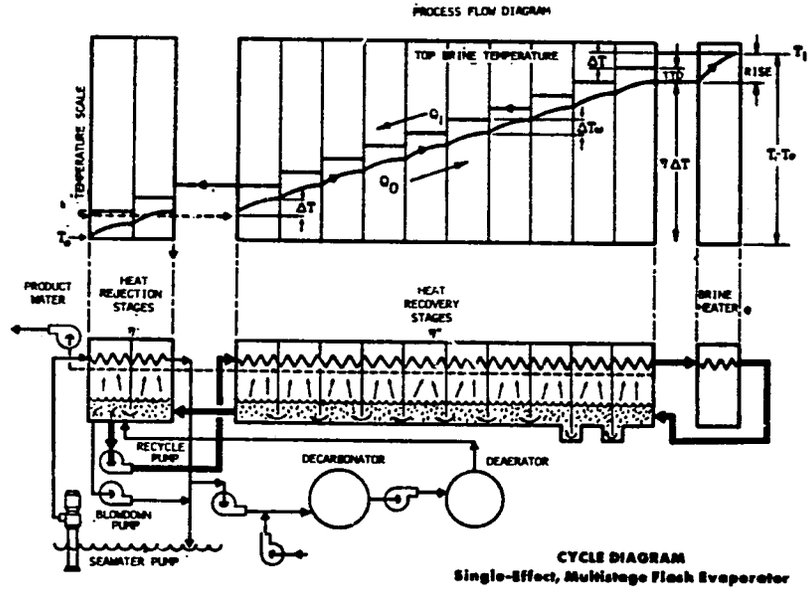
The governing thermodynamic laws dictate that the brine recycle flow is automatically fixed if: (1) the output is fixed by demand, and (2) the flashing range is fixed by the temperature differential between seawater (sink) and maximum allowable brine temperature. The higher the brine temperature, the greater the fresh water output. But brine temperature is restricted by: (1) the temperature at which brine heater tubes will scale, and (2) the process used to prevent scaling.

Makeup brine is treated before it enters the recycle stream. In a typical acid treatment plant, sulphuric acid is mixed into the makeup brine stream which then flows into an atmospheric decarbonator where most of the CO_2 generated by the reaction of the acid with the carbonates is removed. Makeup is next pumped into a low-temperature vacuum deaerator, where air and residual CO_2 are reduced to 50 parts per million. If necessary, acidity is adjusted with caustic, and foaming is inhibited by anti-foam agents. The makeup brine is then ready for recycling.

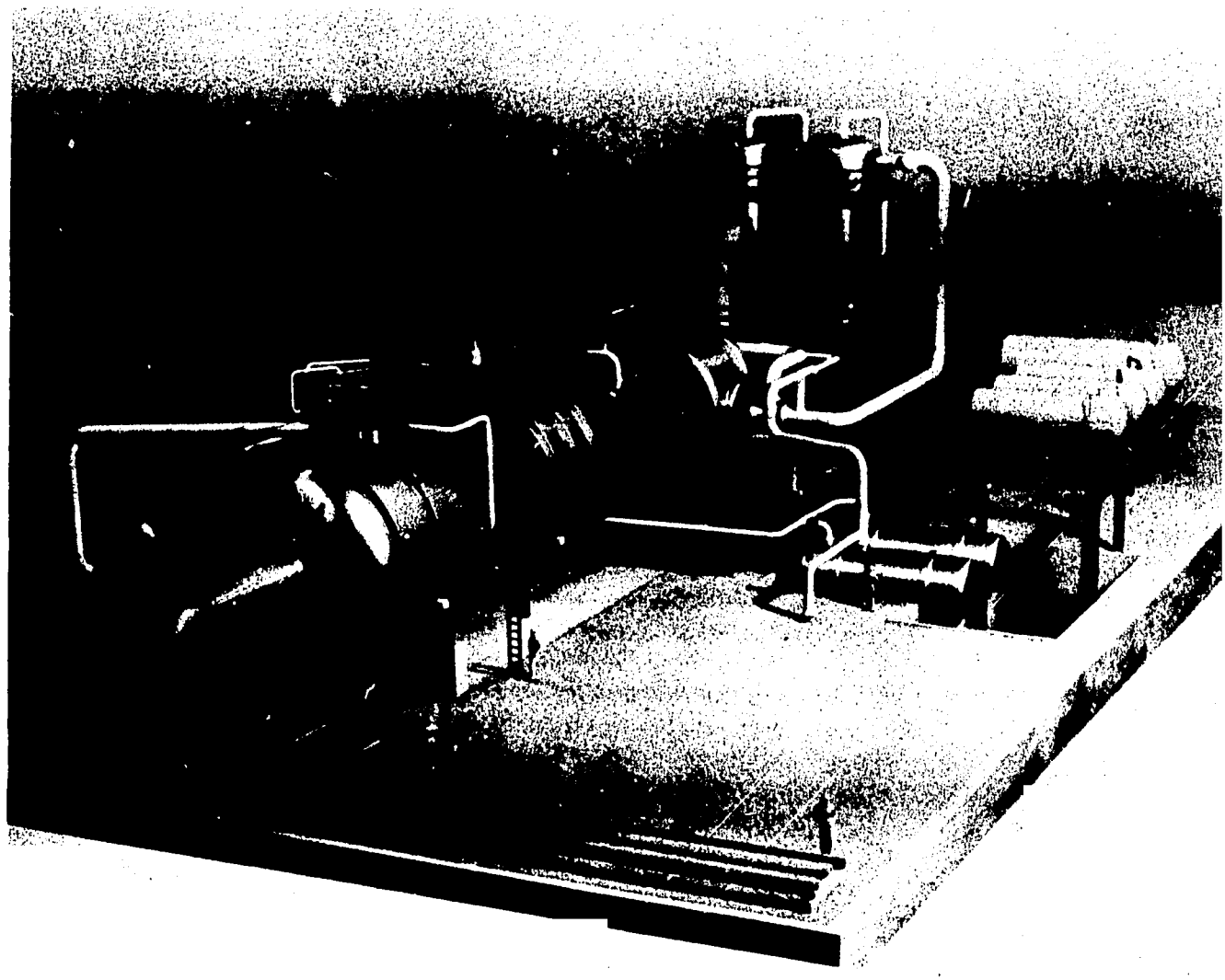
The manufacturers' bidding on the standard design made many suggestions for changes in addition to the allowances permitted by the specifications. Those suggestions which were obvious improvements were adopted; those which could not be substantiated in terms of improvement in thermodynamic efficiency or equal value at less cost were not adopted.

A good deal of time was devoted to discussions regarding the use of an integral deaerator contained in the last heat rejection stage, instead of the external deaerator specified in the technical specifications for the standard plant. After a thorough analysis, the internal deaerator was accepted, primarily because of the reduction in cost.

The accepted design is currently under construction and when it comes on stream in 1969, will increase the water supply for Jidda by about 20 percent. The plant will provide a start for the industrial and municipal growth which Jidda must accomplish in its role as one of the most important commercial centers on the Red Sea. Over the projected 30-year life of the plant, improvements will be made in desalting technology, some of which may be included into the Jidda installation but, in any event, the yardstick plant will be installed and running to provide the comparison for future plants.



CYCLE DIAGRAM
Single-Effect, Multistage Flash Evaporator
FIGURE V-2



THREE MILLION GALLONS OF FRESH WATER DAILY will be produced from seawater in this distillation desalting plant being constructed in Fountain Valley, California by Envirogenics Co. for the U.S. Department of the Interior's Office of Saline Water and the Orange County Water District. Plans call for the plant to be expanded soon to produce from 12.5 to 15 million gallons of water daily. The photograph shows a 1/4 scale model of the three mgd plant. Pure water is produced in the largest horizontal vessel. Seawater enters the plant in the white pipe, bottom right. Fresh water leaves in the dark pipe directly below the white pipe. Plant is scheduled to be in operation in early 1973 as a principal component of Waterfactory 21---the Orange County Water District's designation for the combined seawater distillation unit and the waste water reclamation plant.

FIGURE V-3

From Public Communications
ENVIROGENICS COMPANY
Aerojet-General Corporation
9200 East Flair Drive
El Monte, California 91734

NEW U.S. DESALTER

Technical Background

The Fountain Valley, California seawater desalting plant is a dual purpose facility representing a combined effort on the part of government at the national and county levels. It is a plant using a combination of old and new processes to:

- produce fresh water from seawater more efficiently and, therefore, more cheaply.
- produce operating data to gain confidence in using the system for massive desalting plants of the mid and late 1970's in the 150 million to 200 million gallon per day (gpd) capacities.

The advanced technique of desalting used in the plant is a new twist on the age old process of distillation. It is a system called VTE (vertical tube evaporation), and features the use of fluted metal tubes. The fluting increases the surface area available to heat transfer and therefore more water is produced at less cost. It is this VTE technique that has demonstrated itself to experts as being the most economical way for desalting the massive amounts of water needed in arid areas of the world. This facility will be operating in 1973.

It will be a VTE technique test bed for the U.S. Department of the Interior's Office of Saline Water, the agency having the responsibility for developing technology needed to economically desalt water.

The water will be used by the Orange County Water District to resupply its source of ground water. In so doing it will also retard the incursion of salinity from seawater.

Since the plant serves both the needs of the U.S. Office of Saline Water and the Orange County Water District, it is both a practical producer of water as well as a test bed for technical information needed to prepare for the future. While it is now sized to produce three million gallons of fresh water per day, plans call for expansion of the plant to production levels of from 12.5 to 15 million gallons per day.

DESALTING PLANT FACTS

1. PLANT PURPOSE

   Provide fresh water for Orange County Water District. Produce operating data on the use of VTE technique as a base for scale-up to VTE plants producing as much as 200 million gallons of water per day.

2. DESALTING METHOD

   Combined multi-stage flash (MSF) and doubly fluted vertical tube evaporation (VTE). Most advanced desalting process now available.

3. DESIGN CAPACITY

   Initial, with four effects, three million gpd. Expanded, with 16 effects, 12.5 to 15 million gpd. (gallons per day)

4. SCHEDULE

   On stream at three million gpd size, early 1973.

5. COST

   $7.6 million at three million gpd size including steam generator, seawater intake and outfall and cooling system sized for 12.5 - 15 million gpd unit.

6. PLANT SPONSORS

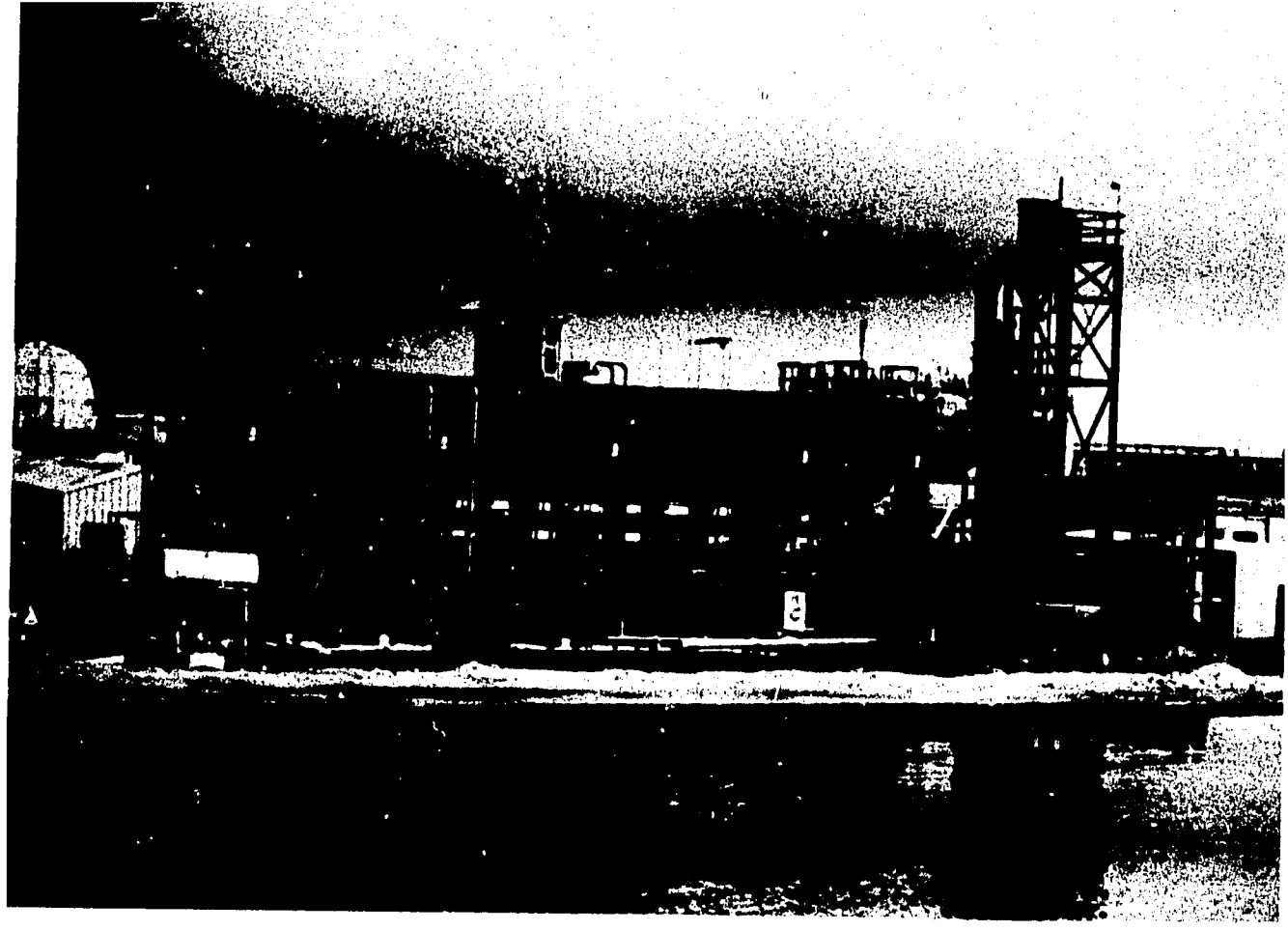
   U.S. Department of the Interior's Office of Saline Water and the Orange County Water District.

7. CONTRACTORS

   Prime--Envirogenics Co., El Monte, California. Major subcontractors, Ralph M. Parsons Co. architect and engineering services; Morrison Knudsen Co., facility erection; Capital Westward, Inc., vessel fabrication.

From Public Communications
9200 E. Flair Drive
El Monte, California 91734

Burns and Roe, Inc.



INDUSTRIAL WASTE WATER RECYCLING PLANT---500,000 gallons of clean water are returned each day to the General Tire & Rubber Company's rubber producing plant in Odessa, Texas by this 14 effect vertical tube evaporator. It is the first large scale use of distillation for fresh water reclamation. The plant was designed, produced and constructed by Envirogenics Company.

FIGURE V-4

From Public Communications
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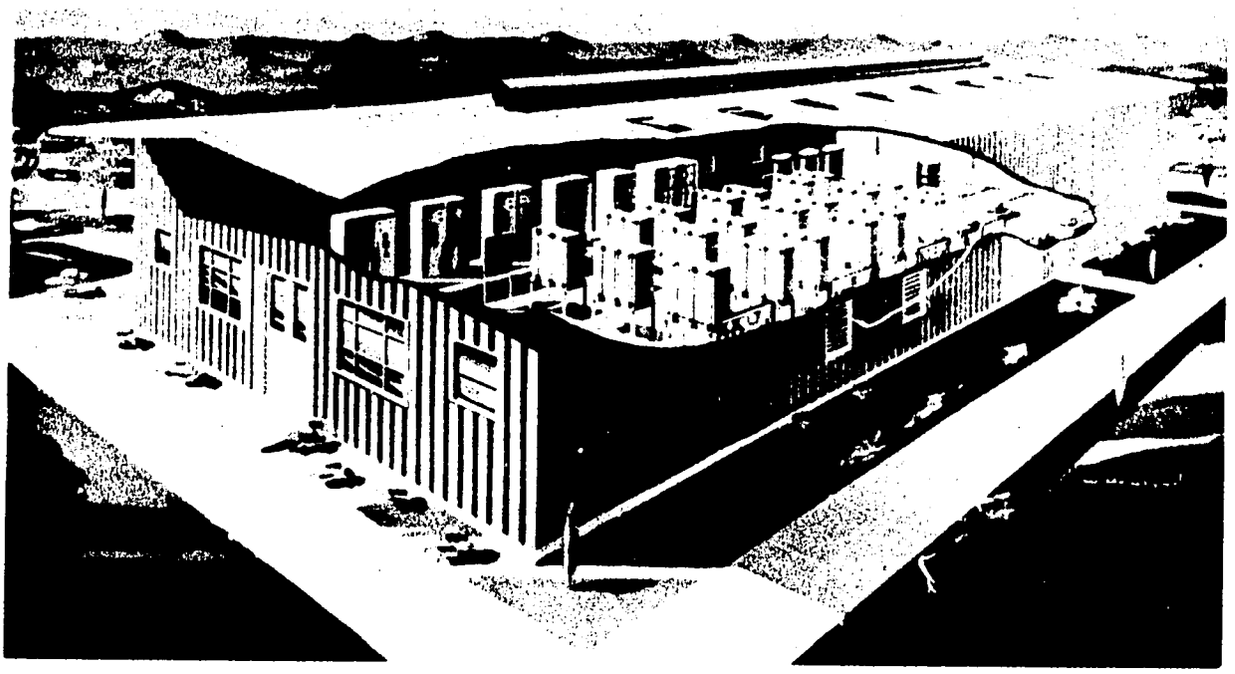
Aerojet-General Corporation
9200 East Flair Drive
El Monte, California 91734

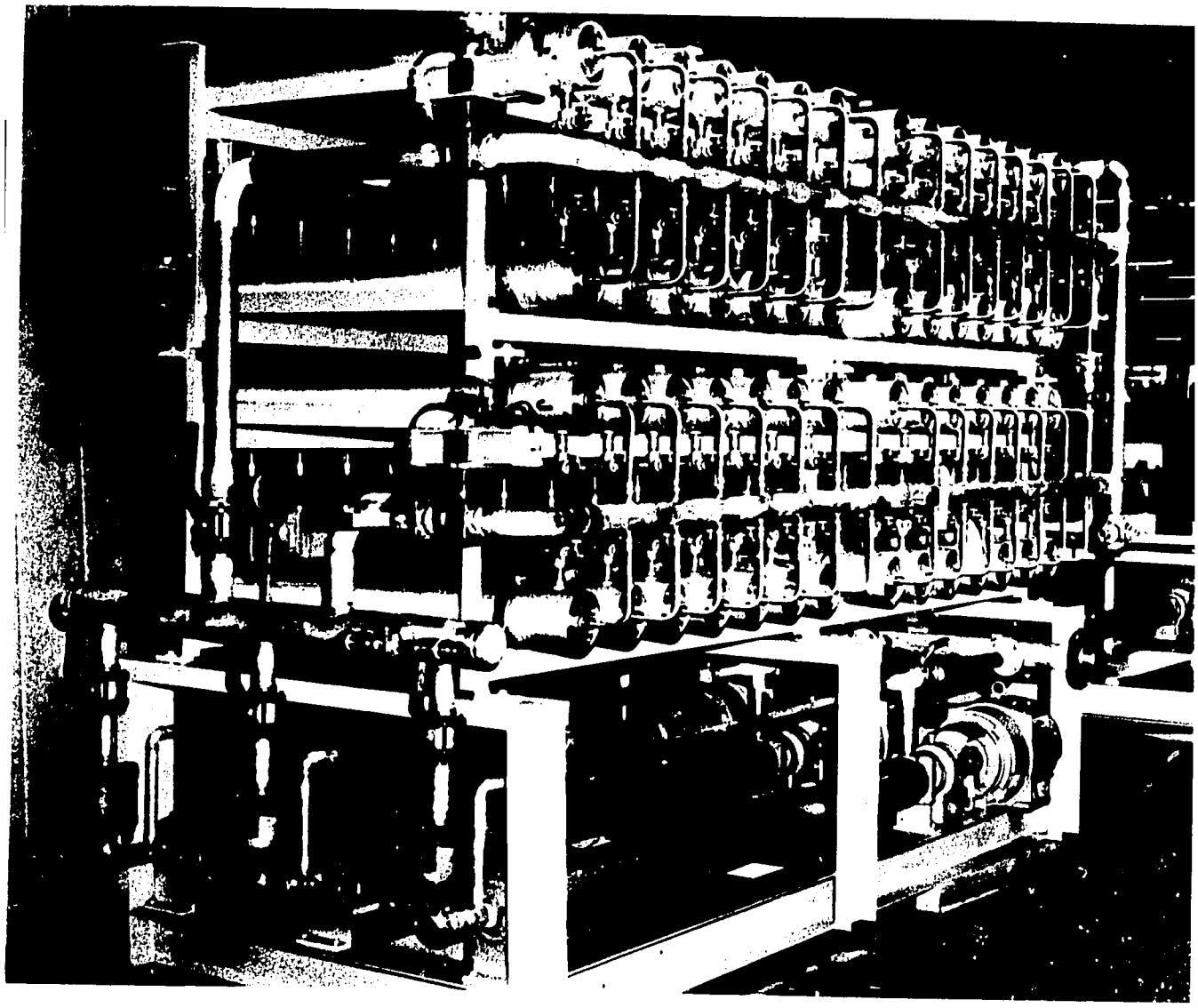


1.5 MGD ELECTRODIALYSIS DESALTING PLANT FOR GILLETTE, WYOMING---The 40 by 100 foot building in this artist's rendering is cut away to show the 1.5 mgd electrodialysis (ED) water desalting equipment inside. Cold (52°F), hard, gypsum-laden well water with total mineral content of 2,500 ppm is pretreated by lime and coagulated to remove iron, (after aeration) hydrogen sulfide, alkalinity and hardness. The softened water is pumped by pumps shown in the background through the 21 box-like structures called membrane stacks, emerging with about 750 ppm of tds. The excess minerals are removed by passing electricity through the water as it flows between hundreds of closely spaced special plastic membranes in each stack. The process requires less than one pound of chemicals per 1,000 gallons treated.

Scheduled for completion in June, 1972, the plant will equal in size another Ionics, Inc. plant serving Siesta Keys, Florida since 1969. The $465,500 desalting plant is part of a $1.8 million water improvement for Gillette, which also includes new wells, water mains, and pretreatment. Operating and maintenance cost for the ED plant is estimated at 27¢/kgal with 13¢ for electric power and consumables, and the balance for reserves for future replacement of membranes and other parts, plus a 5¢/kgal labor allocation.

Courtesy of
IONICS, INCORPORATED
65 Grove Street
Watertown, Mass. 02172

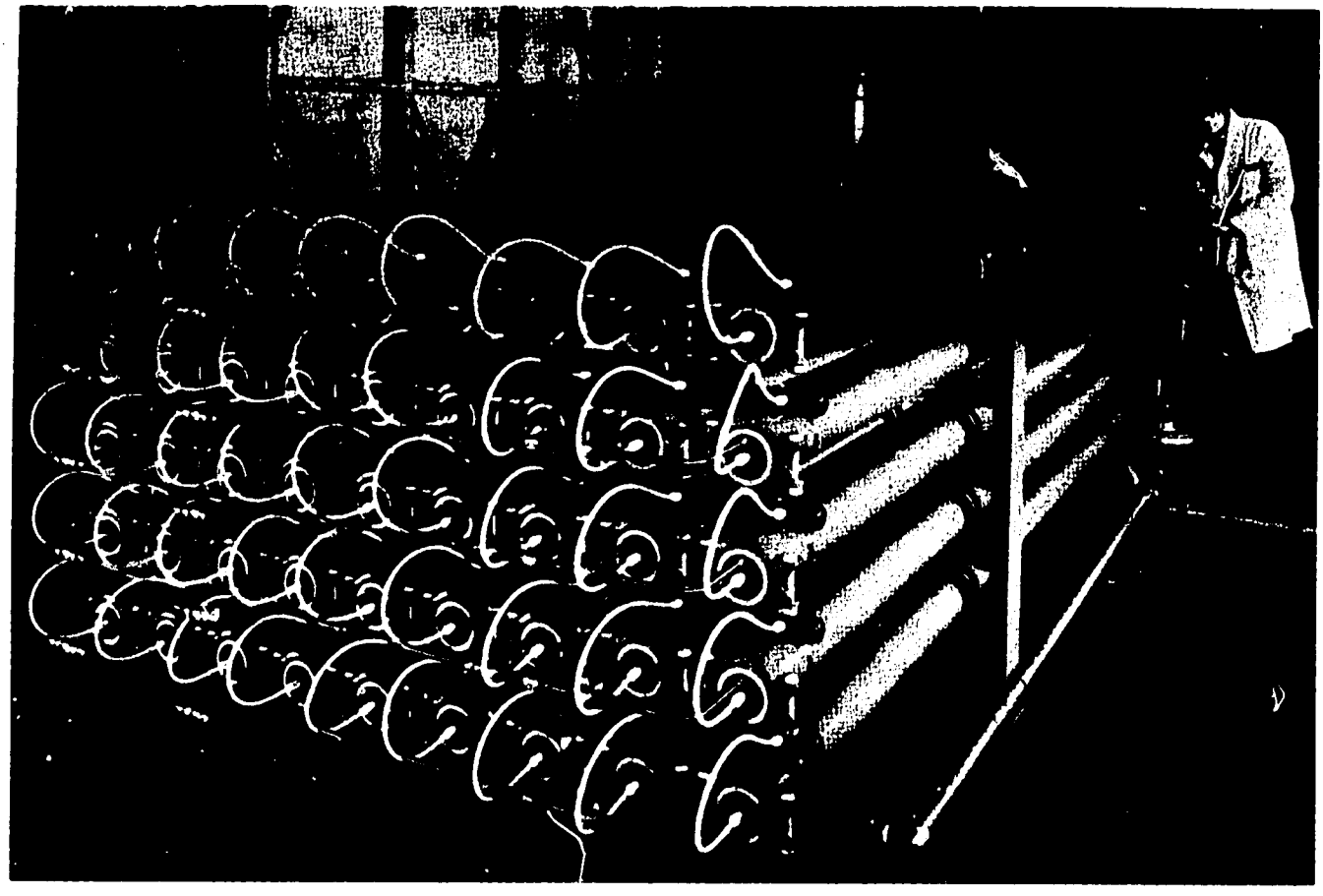
FIGURE V-5



HOLLOW-FIBER REVERSE OSMOSIS UNIT---A 112,000 GPD capacity skid-mounted module now operating in Ottawa, Ontario, Canada. A similar unit, employing 72 hollow fiber cartridges, processes 150,000 GPD of brackish water for municipal use at Greenfield, Iowa, reducing the total dissolved solids from an initial 2400 ppm to a final value of about 200 ppm, which is well within U. S. Public Health Service standards for potable water. Calculation sheets in Appendix C of this report show that this type operation can produce water at about 61¢/kgal, including pretreatment and brine disposal costs.

Photograph by courtesy of
THE PERMUTIT COMPANY, division of
Sybron Corporation
Paramus, New Jersey 07652

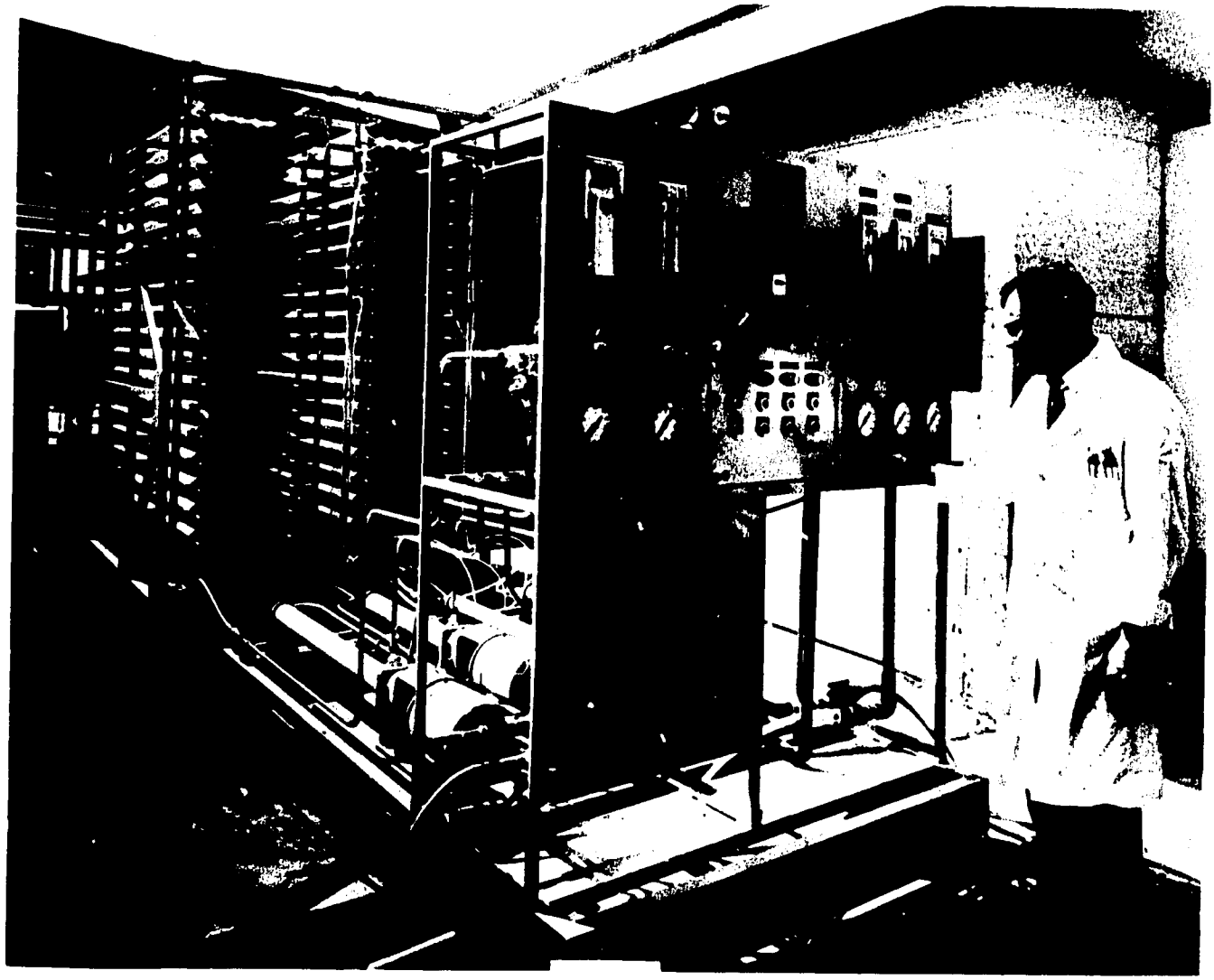
FIGURE V-6



100,000 GPD SPIRAL-WOUND REVERSE OSMOSIS UNIT---These Assemblies can be shipped on a flat-bed truck, and are presently being furnished to electronic component manufacturers as parts of ultra-pure water treatment plants throughout the country. Motorola has recently received 13 bids on their 320,000 GPD membrane desalting plant, proposed for their Mesa, Arizona electronics facility. Studies show that these units will prove economical also for inland municipalities that must rely on desalting brackish waters for their incremental supplies. A typical 1 MGD reverse osmosis system would consist of four 250,000 GPD modular units in parallel, each being 21 ft. long by 7 ft. high by 9 ft. wide. Suitable redundancy of pretreatment units, pumps, and membrane pressure vessel assemblies could be provided, and pump characteristics could be selected so that production could be maintained at about the same level if one of the four 250,000 GPD assemblies is temporarily removed from service. This assures high reliability for municipal supply systems.

Photograph by courtesy of
GULF GENERAL ATOMIC
P. O. Box 608
San Diego, California 92112

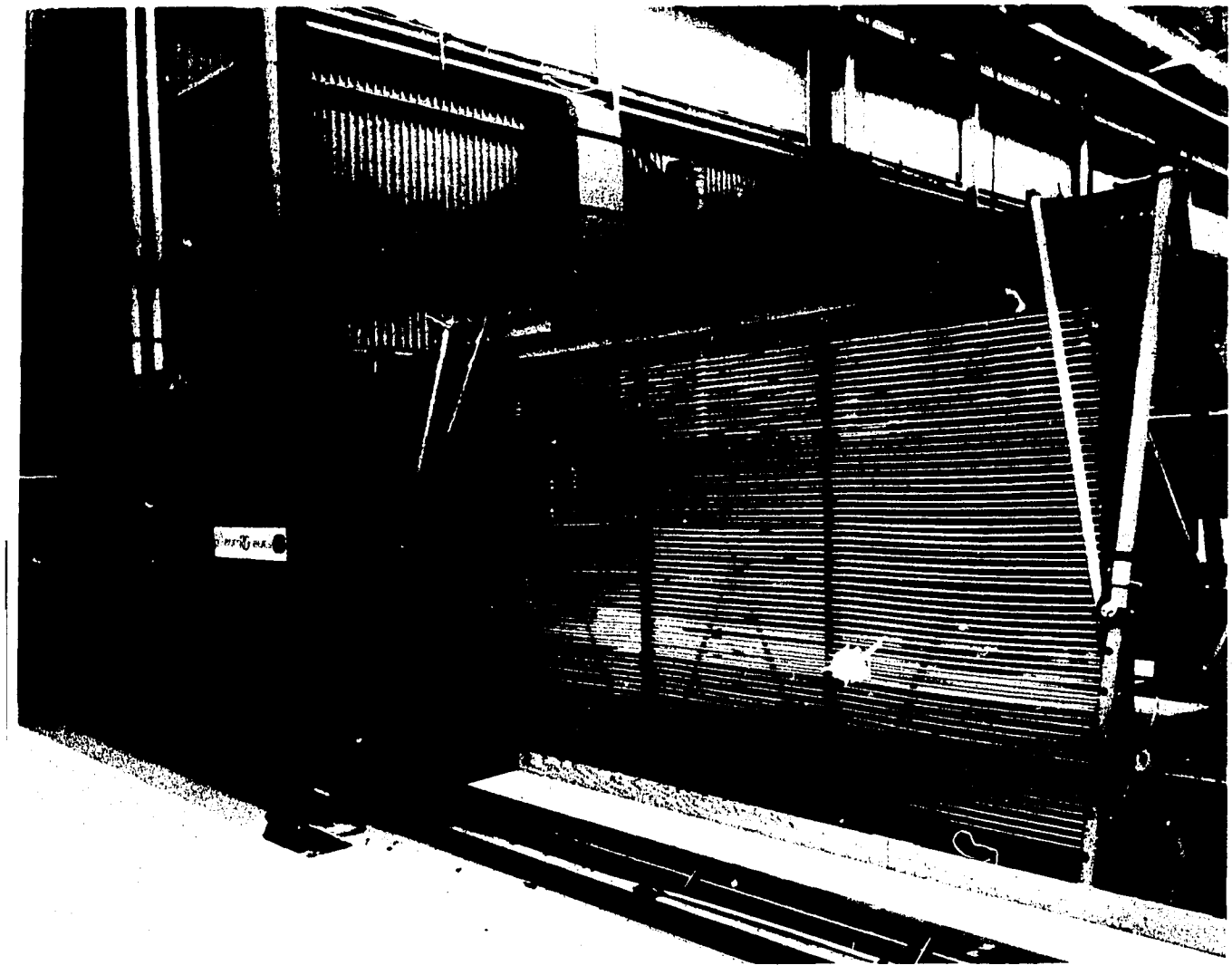
FIGURE V-7



CALGON-HAVENS TUBULAR REVERSE OSMOSIS MODULES---The City of San Diego has been experimenting with reverse osmosis units since 1964. A 20,000 GPD auto-pilot unit similar to the above, but containing 160 tubular modules, set in rows of 5 and 6 modules, and used in two stages, is now recovering usable water from raw sewage, which has received only a rough screening before entering the unit. The pilot plant produces 16,000 GPD of high purity water, which is used as makeup for two low-pressure boilers serving the main treatment plant digestors. To protect the cellolose-acetate membranes against erosion by the raw sewage, diatomaceous earth is applied as a precoat on a 7 hour operating cycle. Advantage is taken here of the open tube construction, which permits periodic flushing to remove accumulated solids.

Photograph by courtesy of
CALGON CORPORATION
Subsidiary of Merck & Co., Inc.
8133 Aero Drive
San Diego, California 92123

FIGURE V-8



LARGE REVERSE OSMOSIS SYSTEM turns 45,000 gallons of seawater into fresh water daily in a single pass. Envirogenics Co. unit operates at a pressure of 1,000 psi. Reverse Osmosis membrane is contained in metal tubing.

FIGURE V-9

From Public Communications
ENVIROGENICS COMPANY
Aerojet-General Corporation
9200 East Flair Drive
El Monte, California 91734
(213) 572-6583

1511 012

VI. DUAL PURPOSE PLANT DISCUSSIONS

A. General

The following items have been discussed in detail in Section VI of Volume I of this manual.

1. Dual-Purpose Electric Power Generation - Water Desalination Plant

All seawater and brackish water plants require energy for their operation. Electrical energy is required for electrodialysis to effect the ionic separation. All the processes require power for driving pumps and compressors. The distillation processes require heat energy, usually steam at temperatures above 250° F. This heat energy may be supplied by a self-contained boiler or it may be waste heat from a diesel generator or "extraction" steam from another operation. Frequently this steam is supplied by a turbine driving an electric generator. In this case the facility is known as a "dual-purpose electric power-water plant".

Dual-purpose plants, because of shared energy and facility uses, effect cost savings of considerable size. Since distillation processes are connected in series with the back pressure steam turbines supplying them, they recover a considerable portion of the heat energy that separate facilities would normally discharge to cooling water. The sum of the energy required by a single-purpose steam-electric power plant and by a single-purpose desalination plant is approximately 40% greater than that required by a dual-purpose plant of the same net output.

The question of whether to use a back pressure turbine or to extract steam from a condensing unit is based upon the relative sizes of the generation and desalination plants. In general, if the "Power/Water Ratio" (megawatts generated /mgd water produced) is less than 8, the use of a back pressure turbine is indicated. Where power generation is the prime consideration, power/water

ratio greater than 25, the use of an extraction turbine is indicated. Between these values a more detailed study must be made in order to determine the best combined mode of operation.

It must be pointed out that in meagerly industrialized areas, the projection of a large power generation or the juggling of power and water credits in order to lower the cost of the water is a fallacy.

2. Chemical Byproducts

The sea has long been a source of chemicals for mankind. The principal materials available are sodium chloride (NaCl, common salt), magnesium and bromine compounds. The concentration of the brine leaving the plant is a variable, dependent on the process used. The volume of this brine is extremely large. A single plant will be capable of supplying all of the sodium chloride normally demanded by any country many times over. The uses of magnesium and bromine are quite limited. The processes for recovery of these materials have been discussed at length in Volume I.

It must be remembered that any appreciable increase in the production of these chemicals will adversely affect the market values, resulting in limited financial return from this source. The value of sulfur as a consequence of natural gas sweetening has already shown the effects of oversupply.

B. Advancements

In Section VII which follows, there is a reprint of a paper detailing the desalting aspects considered during the conceptual design phase of a dual-purpose desalination plant planned for installation in Puerto Rico. It is to be noted that the authors considered the influence the complex would have on the overall economic planning and development of the island. The study examined the effects the power and water generated would have on industry and agriculture. The production of NaCl in this plant is justified by its use in the preparation of caustic soda for Puerto Rico's chemical industry. Similarly, a crystallized salt-and-water producing unit, utilizing a vapor-compression cycle, is described on the following pages.

RESOURCES CONSERVATION CO.
FLAT-PLATE, VAPOR-COMPRESSION EVAPORATOR

The RCC desalination system uses a vapor-compression evaporation process that retains nearly all latent heat being transferred by a falling-film heat exchanger. The basic components of the RCC evaporator are the flat heat-transfer plates, a steam compressor, an evaporator shell, and a brine recirculation pump. A schematic of the evaporative process is shown in Figure 1. A high heat-transfer coefficient is obtained in the RCC heat exchanger because of its low brine film resistance, low metallic membrane resistance, and a patented design for controlling the condensate film thickness. The flat-plate heat-exchanger design is adaptable to either vapor-compression or multiple-effect thermodynamic modes of operation. The additional incorporation of a seed slurry method, which eliminates scale deposition on evaporator surfaces by providing preferential sites within the concentrated brine for crystal growth, enables the RCC system to be used in pollution control and waste brine concentration applications also.

As a vapor-compression system, the RCC evaporator operates primarily from mechanical energy used to power the compressor. The compressor is adaptable to a variety of power sources, as diesel engines, steam turbines, electric motors, or gas turbines. These power options allow for direct conversion of the most economically available fuel to usable rotational energy. As a multiple-effect system the evaporator can be operated using the vapor compression process or by using waste steam, or steam generated from many common fuels.

a. Plant Description

The RCC design, Figure 2, stresses the use of modular construction, which allows factory assembly and checkout with attendant high equipment reliability, simplified transportation, and reduced site installation costs. Additionally, the RCC design allows coupling of multiple units, either in parallel to increase total capacity or in series to achieve higher brine concentration factors. The parallel or series coupling is done in a manner that permits units to be taken off line individually so that production loss is minimized during scheduled or unscheduled shutdown.

RCC desalination systems are fabricated from corrosion-resistant, high-strength, aluminum alloys. This offers substantial cost savings over more expensive materials while providing adequate corrosion resistance for most brines. For industrial water treatment applications where corrosion is a particular problem, other metal alloys can be used. In addition to lower fabrication costs, the high heat-transfer coefficient of aluminum reduces operating costs. Nonwetted structural components of the desalination system are of steel construction.

The design of the RCC evaporator provides an inherently reliable process, while a combination of advanced manufacturing techniques and system simplicity assure a low-maintenance operation. The system is operated from a centralized motor control, instrument, and alarm/monitor panel. The system can also be instrumented for fully automatic computer-controlled operation.

b. Process Description

Feed brine is introduced through a feed pump, heat exchanger and deaerator into the sump of the evaporator (Figure 1). The brine is then circulated to the top of the heat-transfer plates where it cascades as a thin film over the outer surface of the plates and back into the sump for recirculation. Steam that is generated as the brine film falls over the heat-transfer surfaces enters the compressor where its saturation temperature is increased to a few degrees above the boiling temperature of the recirculating brine. Then the hot, compressed, steam flows into the interior of the heat-transfer plates where condensation occurs on the inner wall surfaces, thus giving up energy to evaporate more water from the brine film. The condensate flows through a product pump to the heat exchanger, where it is cooled to nearly the incoming feed temperature. In the process, the feed is heated to approximately its boiling point. Hot concentrated waste brine from the plant flows through a brine pump to the heat exchanger also in order to give maximum heat recovery in the system.

c. Special Features

The major innovation of the RCC evaporator design is the use of a patented construction on the condensate side of the heat-transfer plates to control condensate film thickness, thereby improving condensate-side overall heat-transfer coefficients.

The patented heat-exchanger design has been adapted to mass production through the utilization of extruded aluminum sections, which make it competitive with enhanced vertical tubes on an installed heat-transfer-surface cost-per-unit production basis.

Efficient scale control in brine concentration applications has been obtained by a low heat flux across the heat-transfer surface, in combination with seed-slurry techniques.

d. Present Use

A modular system of 100,000 GPD capacity has been produced for installation at a resort site in the Caribbean in October, 1972. An illustration of this plant is shown in Figure 3.

A two-stage brine concentration evaporator system is on stream at a natural gas compressor station in El Paso, Texas. This system has been in service since July, 1971, and is processing approximately 50,000 GPD of blowdown from the station cooling towers having a nominal total dissolved solids concentration of 2400 mg/l. This plant returns 99% of the blowdown to the cooling tower system as pure water containing less than 10 mg/l TDS. The remaining 1% is discharged as a waste brine containing 250,000 mg/l dissolved solids, and 35,000 mg/l suspended solids.

e. Plant Staff

The RCC desalination system is designed for operation with minimal attention. The evaporator is operated from a centralized control panel which displays all necessary operating parameters. An alarm panel signals out-of-tolerance conditions with both audible and visual alarms, and under normal conditions only periodic checks by an operator are required. Utilization of off-the-shelf components insures easy maintainability by a user's normal maintenance staff.

f. Costs

Cost of a 200,000 GPD desalination plant is approximately $500,000. Typical energy costs per 1,000 gallons of fresh water produced are shown in Figure 4.

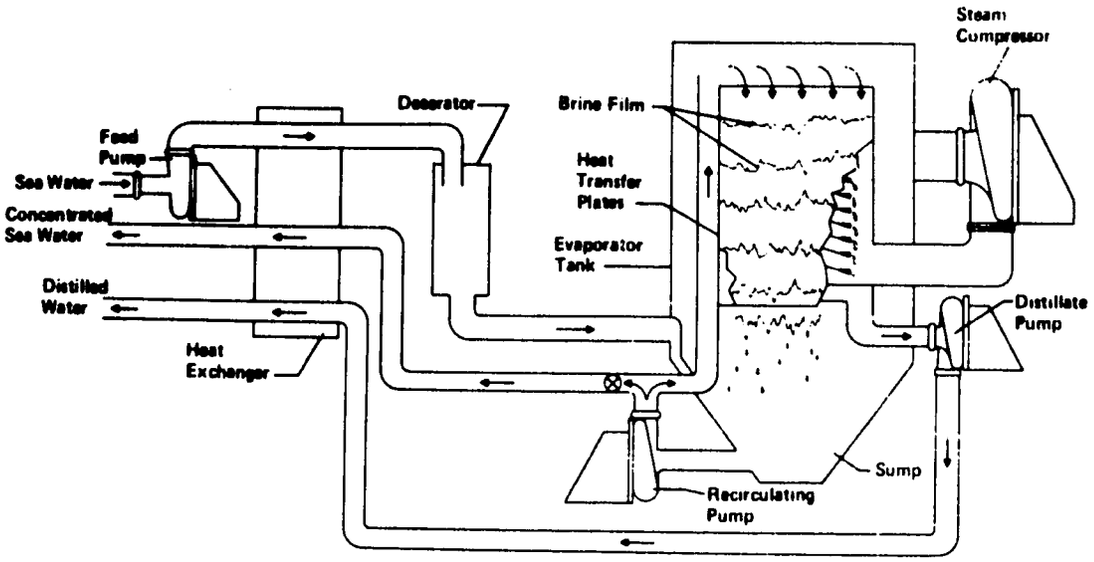


FIGURE 1 RESOURCES CONSERVATION CO. DESALTING PROCESS SCHEMATIC

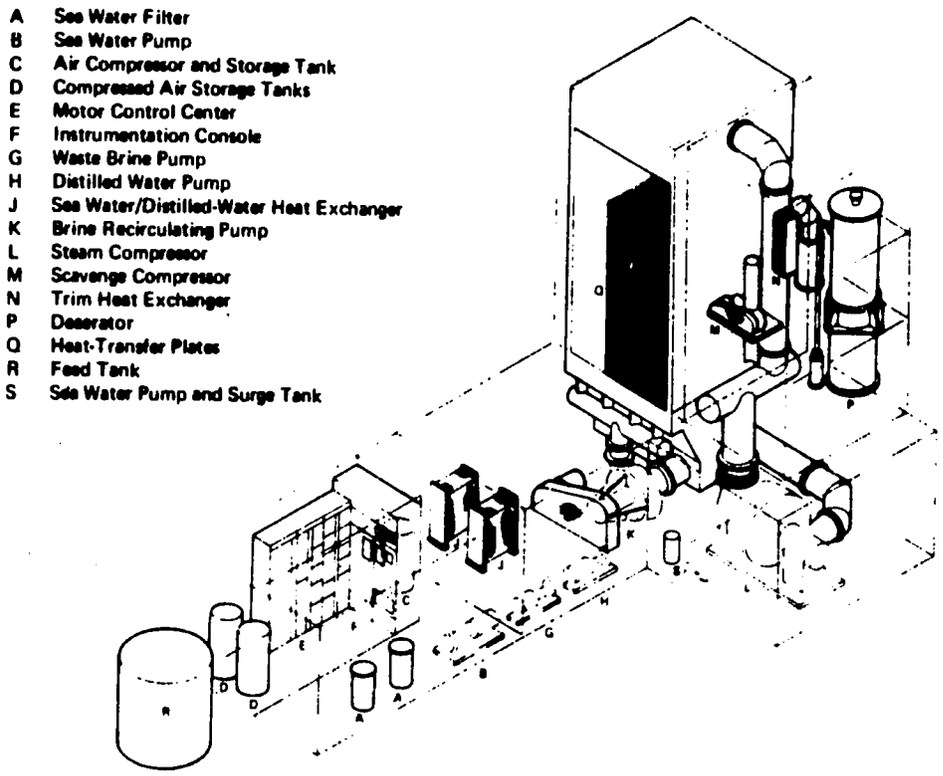


FIGURE 2 RESOURCES CONSERVATION CO. CANEEL BAY DESALTING PLANT

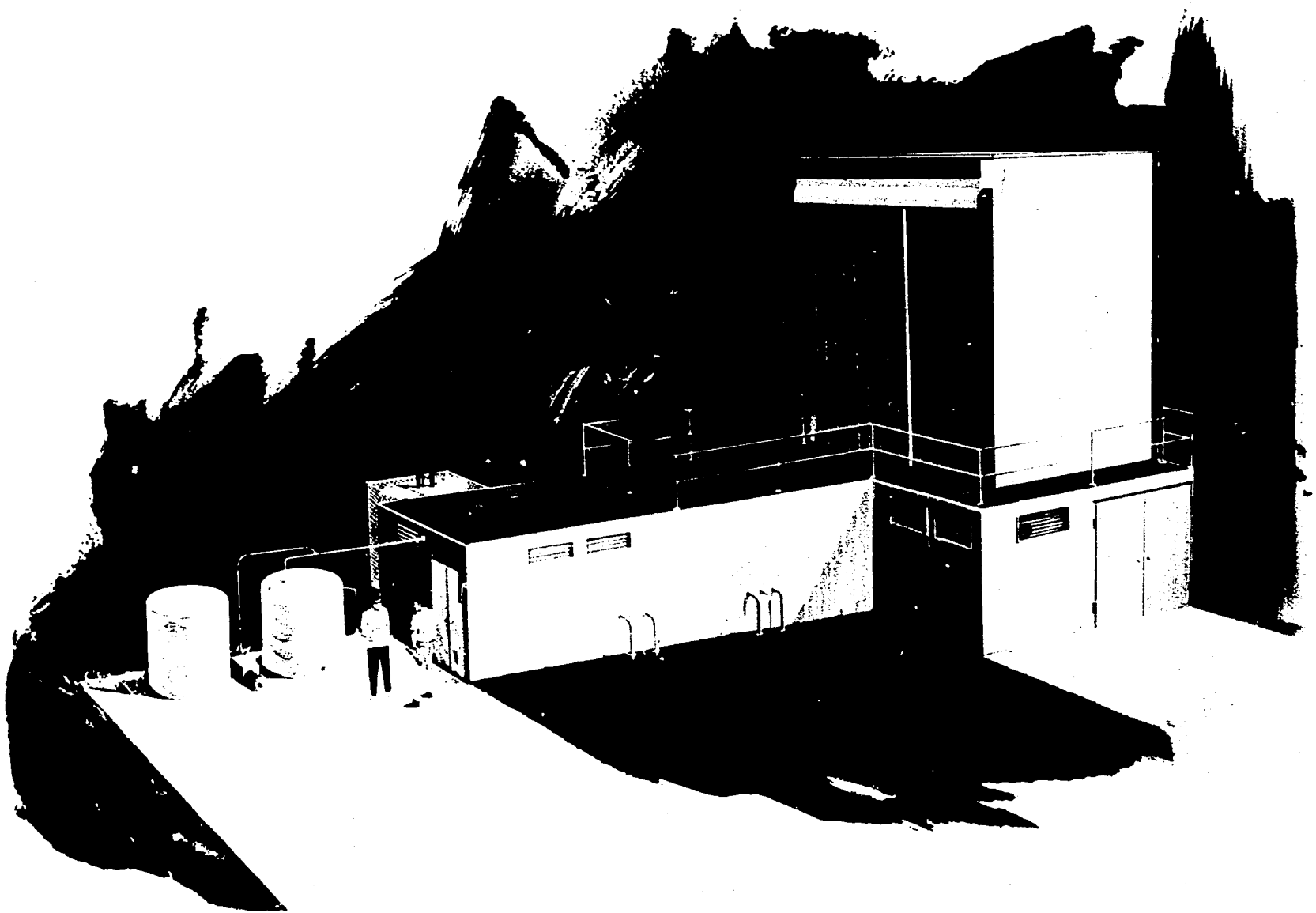


FIGURE 3 RESOURCES CONSERVATION CO.
DESALTING PLANT FOR CANEEL BAY
PLANTATION, ST. JOHN, VIRGIN ISLANDS

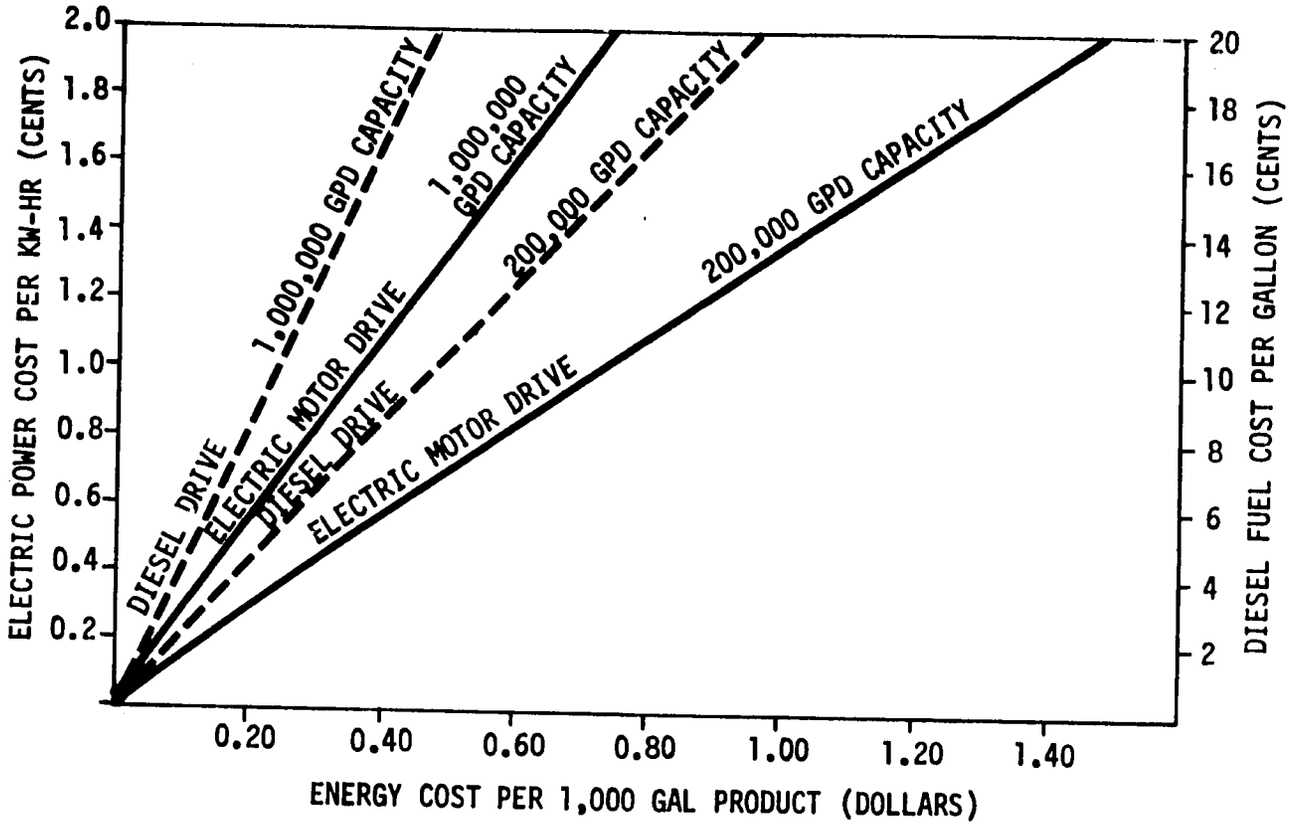


FIGURE 4 RESOURCES CONSERVATION CO. TYPICAL ENERGY COSTS FOR FRESH WATER PRODUCTION

VII. SUMMARY OF MAJOR CONCEPTUAL DESIGN STUDIES

A. General: Section VII, Volume I, presented summaries of 22 large conceptual designs for seawater distillation plants. The underlying philosophy in all of these designs was to develop a modular design having attractive costs when scaled up to 50 mgd capacity or more. The present status of these designs is as follows:

Description	1972 Status
200 Mw + 130 mgd Dual Purpose Plant Israel	Concrete MSF Section not yet proved. Project dormant.
1600 Mw + 150 mgd MWD Dual Purpose Bolsa Island, Cal.	MSF test module operating at San Diego. Project abandoned when water cost was escalated from 22¢ to 39¢/K gal.
20 Conceptual Designs Single Purpose 50 mgd Distillation Plants (OSW Study Grants)	Innovations still being tested. VTE/MSF has proved successful. LMC process to allow 350° F brine temperatures is under study. Economics of modular concepts to allow 50 mgd capacity is still under study using both steel and concrete vessel constructions.

B. Advancements:

Of the 20 conceptual designs listed, some, at least with modification, have proved successful.

With regard to MSF, the largest plant is the 7 mgd Tiajuana, Mexico, unit, built by Aquachem. A 50 mgd test module has been under test by OSW at San Diego, California, with good results. A 48 mgd plant, scheduled for Hong Kong, is presently in the detailed design stage.

VTE and VTE/MSF capabilities have been verified by OSW at San Diego and at Freeport, Texas. A 3 mgd VTE/MSF module is being built for Orange County at Fountain Valley, California, by Envirogenics, which later may be expanded to 15 mgd. The module provides an economical building block with an ultimate capacity of 200 mgd.

The emphasis in the Israeli conceptual design is for added economy and has now switched from MSF to a Horizontal Tube Evaporator (HTE), using aluminum tubes and temperatures below 170°F. A 40,000 gpd 10 effect pilot plant has operated successfully for 18 months, a 1 mgd plant is to be built in Eilat by 1974, and negotiations for building an 11 mgd prototype at a cost of $15 million are under way.

Vapor Compression, combined with the VTE/MSF concept, has been proposed in capacities of up to 20 mgd for inland sites where the cooling water supply is inadequate for ordinary distillation processes. A demonstration plant has been proposed for Brownsville, Texas. Construction has been postponed, but the large vapor compressor has been purchased and will be tested.

A conceptual design study involving a 20 mgd water plant associated with a nuplex has been conducted for Puerto Rico, which represents the latest state-of-the-art for multi-purpose plants with chemical byproducts. A summary of this study is presented in the reprinted AIChE paper at the end of this Section.

Within 10 years, it is anticipated that municipal and industrial usage, both of brackish water and seawater, will rapidly increase as desalting costs decline, in comparison to alternate means of supply, and as needs to improve water quality increase. The greatest U. S. potentials are in brackish water inland areas, the Southwest Pacific, the Gulf Coast and Caribbean areas. Within 30 years large applications can be anticipated for municipal or regional supply. A total U. S. desalted water need of 2.5 billion gpd is projected for the year 2000.

C. Problems

Technology developed to date provides only small scale evidence for the feasibility of future large-scale applications. The gap between current status and practical wide-spread applications must be bridged to reach desalting's potential of significantly

contributing to future water supply; that is, where desalting plants in sizes of the order of 100 to 500 mgd will become a reality. At the same time, advanced technology for a variety of practical desalting plants, in a size range of .5 to 20 mgd, is necessary to meet the increasing needs of smaller municipal and industrial users. These require new commercial methods such as the membrane processes, which involve no phase change, in order to reduce costs, cooling requirements and brine disposal volumes when providing water for inland villages and agriculture.

Concurrent with hardware applications is the need to continue research and development efforts to assure realization of projected cost levels of 25 to 35¢/Kgal of fresh water from large seawater plants and of even less cost from brackish water plants, particularly if agricultural needs are to be met.

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Chemical
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SYMPOSIUM
SERIES

PUERTO RICO ENERGY CENTER: DESALTING ASPECTS

J. Hernandez Fragoso

Puerto Rico Water Resources Authority, Santurce, Puerto Rico

and

James J. Byrnes

Burns and Roe, Incorporated, Hempstead, New York

Vol. 67, No. 107 1970 Pages 217-223

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and
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The Commonweaith of Puerto Rico with an area of 3,500 sq. mi., a population of 2.7 million, and a population density of 780 persons/sq. mi., is one of the most densely populated countries in the world. The Island has a high birth rate coupled with a decreasing mortality rate, and it is estimated that by 1985 the population of Puerto Rico will have increased to more than four million.

Prior to World War II, the economy of Puerto Rico was principally agrarian. Since that time, a dramatic transition towards industrialization has occurred, triggered by the economic incentives offered by the government. New facilities are established year after year, and this pace is expected to continue. The objective of the Puerto Rico Industrial Development Company is to triple present facilities by 1985. There is no doubt that industrialization has contributed considerably to the economic development of Puerto Rico. Present annual per capita income is $1,047.

The increased population accompanied by the industrial development and increased income have combined to produce an ever increasing need for a better utilization of the few natural resources available in Puerto Rico. This requires a higher degree of planning by the government. The interrelationships among available power and water with industries and agriculture coupled with the ever increasing demands of a growing population for a better standard of living in all its phases require deeper analysis than in previous years.

One of the steps taken by the Commonwealth was to initiate a study, with the assistance of the Federal Government, to determine the feasibility of an agro-industrial nuclear Energy Center for the coastal regions of Puerto Rico. This study is being conducted at the present time under a contract with the Puerto Rico Water Resources Authority and the Puerto Rico Aqueduct and Sewer Authority by Burns and Roe, Inc., which has as a principal subcontractor The Dow Chemical Company.

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PURPOSE OF THE STUDY

The study is intended to assist Puerto Rico in its plans for future development. The nuclear Energy Center concept involves a complex for the coastal regions of Puerto Rico consisting of a collection of industries clustered around a nuclear reactor and a desalting plant, in association with highly mechanized, scientific farming.

The guidelines for the study suggest that consideration be given to nuclear fueled electric power and desalting plants which would generate 500,000 to 1 million kw. of electricity and desalt 5 to 100 million gal. of water daily. The power and water produced would be for municipal, industrial, and agricultural use in connection with projects that could become operational between 1975 and 1990.

The 1 yr. study does consider possible sites on the southwestern, southeastern, and north-central regions of Puerto Rico, but will result in a recommendation and detailed analysis of a specific complex at a specific site. It will also consider the effect of a nuclear complex on the overall economic planning and development of the Commonwealth.

Four segments are thoroughly examined: power, water, industry, and agriculture. It is too early to derive any conclusion out of the study. Therefore, the discussion that follows does not necessarily represent the findings of the study, although the information and the data are pertinent to it.

This paper concentrates on the water segment of the

study and, most notably, the desalting aspects. The paper attempts to give the main concepts considered and the preliminary findings without going into details. Further details are given in the final project report issued in September, 1970.

The paper presents the requirements for water in the Energy Center and in the Guayama Sub-Region which was studied as a tentative location. The potential sources of water and their costs, including wells, dam projects, and desalting plants, are reviewed. Costs are compared for several different types of desalting plants, including several different methods of seawater pretreatment. A brief presentation is given on the economics of recovering salt and other marine chemicals from the brine waste of the desalting plant.

WATER REQUIREMENTS

The study assumed the Energy Center to be located at Aguirre in the Guayama Sub-Region which is shown in Figure 1. The Guayama Sub-Region consists of the municipalities of Salinas, Guayama, Arroyo, Patillas, and Maunabo.

The freshwater requirements of the Energy Center are listed in Table 1. This water is all used for process purposes; seawater is used for cooling. The industrial water requirements of the Guayama Sub-Region are shown in Figure 2 and are made up principally of the requirements of the Phillips Plant, the Energy Center, and an allowance for miscellaneous other developments.

The combined municipal and industrial water demands of the Guayama Sub-Region are shown in Figure 3.

The water requirements for the Guayama Sub-Region are largely satisfied by the existing South Coast Irrigation District and water pumped from wells. It is expected that the demand for irrigation water will remain steady or may decline in future years as food crops, other than sugar cane, are introduced which require less water and as sprinkler irrigation is introduced that uses water more efficiently than the present furrow type of irrigation.

Fig. 1. Location of Guayama subregion.

TABLE 1. PUERTO RICO ENERGY CENTER WATER REQUIREMENTS

Plant	Quantity, million gal./day
Power plant	1.4
Chemical refinery	.3
Ethylene	1.6
Ethylbenzene	.1
Cumene	.1
Propylene glycol	12.4
Ethylene dichloride	.1
Ethylene glycol	.2
Polyethylene	4.4
Chlorine-caustic	2.0
Aluminum	.3
Extra allowance	3.9
Total	26.8

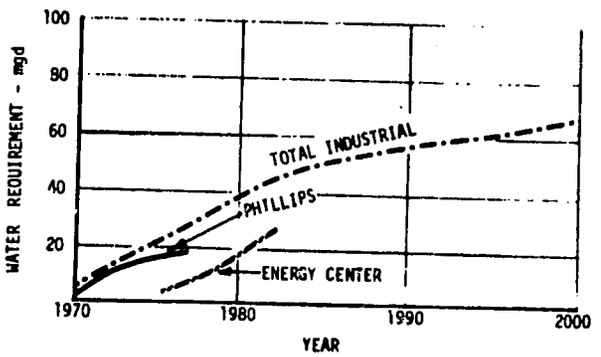

Fig. 2. Industrial water requirements of Guayama subregion.

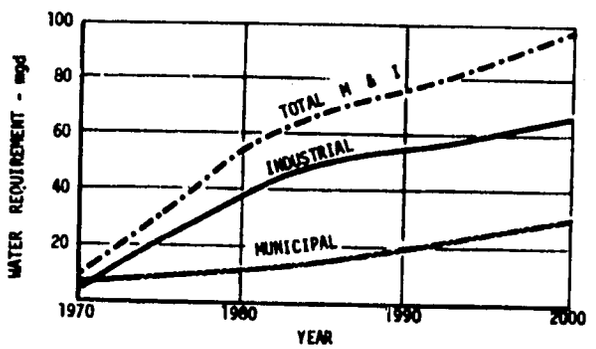

Fig. 3. Municipal and industrial water requirements of Guayama subregion.

WATER SUPPLIES

The water supplies available to meet the expected water requirements are shown in Figure 4 along with the total water requirement.

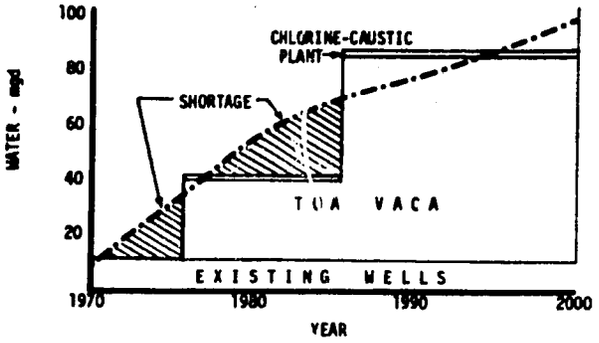

Fig. 4. Water supplies and requirements of Guayama subregion.

WELLS

Existing industrial wells have a capacity of 5 million gal./day. Existing municipal wells have a capacity of 2.85 million gal./day. In addition, 4 million gal./day from wells on the West Aguirre Energy Center site have been included. This is water that is presently used for irrigation that could be used for industry.

Well water supplies generally cost from 5¢/1,000 gal. or less at the wells.

TOA VACA

Figure 4 shows the amount and scheduled delivery dates for water from Toa Vaca Dam for the Guayama Sub-Region. The Toa Vaca dam project is planned in five stages. Stage 1, which is presently under construction, will dam up the Toa Vaca River. Stages 2, 3, 4, and 5 collect and divert water from north of the Cordillera Central to the Toa Vaca Dam for distribution on the south coast.

Table 2 shows the yield, costs, and schedule for Toa Vaca. Table 3 shows the unit water costs for Toa Vaca based on fixed charges of 6.33 and 100% use of water.

CHLORINE-CAUSTIC PLANT

The chlorine plant of the Energy Center produces about 1.4 million gal./day of water from the caustic evaporators.

TABLE 2. TOA VACA DAM PROJECT

Stage	Completion date	Yield, million gal./day	DAM cost $\times 10^{-6}$	¢/1,000 gal. Increment	¢/1,000 gal. Cumulative
1	10/71	23	$ 17.5	13.2	13.2
2	6/75	20	55.0	47.6	29.2
3	6/77	60	55.0	15.9	21.5
4	6/80	25	31.0	21.5	21.5
5	1/89	100	48.0	8.3	15.7
Total		228	$206.5		15.7

TABLE 3. DELIVERED UNIT COSTS OF WATER FROM TOA VACA

Items	Cost ¢/1,000 gal. Stages 1 and 2 (Ponce)	Stages 3, 4, and 5 (Guayama)	Total
Dams	29.2	12.6	15.7
Distribution	2.1	3.4	3.2
Total for raw water	31.3	16.0	18.9
Filtration	6.6	6.6	6.6
Total for filtered water	37.9	22.6	25.5

SHORTAGES

It can be seen from Figure 4 that two shortages of water occur, the first reaching a peak of 28 million gal./day in 1977 before it is relieved by Toa Vaca and the second reaching 33 million gal./day in 1989 before it is relieved by Toa Vaca.

These shortages must be supplied either from other wells or from desalting plants. It is estimated that the Guayama Sub-Region could probably yield an additional 10 million gal./day from wells on a reliable basis and over and above the present well water usage. The remainder must be obtained from desalting plants.

BASIS OF SIZING DESALT PLANT

There are two bases on which the desalting plant can be sized, the first being to recover enough salt from the waste brine of the desalting to feed the chlorine-caustic plant, that is, 3,000 tons of salt per day. The second basis is satisfying the water needs of the Guayama Sub-Region including the Energy Center and buying all the salt needed for the chlorine-caustic plant. The water supplied on the two bases is summarized in Table 4 and shows that satisfying the salt needs results in a surplus of 6.9 million gal./day of water which would be available to the Guayama Sub-Region outside of the Energy Center. Sizing to satisfy salt needs makes the Energy Center independent of the Toa Vaca project, and the surplus of water might make it desirable to delay the delivery of

TABLE 4. BASIS OF SIZING DESALT PLANT

Water source	Quantity, million gal./day Salt	Based on: Water
Desalt plant	20.0	15.0
Salt recovery plant	8.3	—
Chlorine-caustic plant	1.4	1.4
West Aguirre wells	4.0	4.0
Toa Vaca	—	6.4
Total	33.7	26.8

water from Toa Vaca until 1981. Based on satisfying the water needs, a plant of 15 million gal./day going into operation in 1974 will take care of matters until 1985. This solution makes the Energy Center dependent on receiving water from Toa Vaca.

It was decided to study the desalting plant on the basis of satisfying the salt needs, that is, a 20 million gal./day plant. This was arranged in two 10 million gal./day trains to give redundancy to reduce the chances of the total water supply being interrupted. The desalting plant is backed up by an underground reservoir with well capacity of 22 million gal./day to serve as an emergency supply. It is estimated that mining this underground reservoir could meet the water needs of the Energy Center for a period of 1 mo.

The separation of the water from the salt is a continuous evaporative process starting from seawater, so that the concentration ratio to assign to the desalting plant is somewhat arbitrary. It was decided that multieffect vertical tube evaporators (MEVTE) could produce a brine concentrate of three times normal seawater at no sacrifice to the cost of water, so that the desalting plants would use this as a design basis. The single-effect multistage flash distillation (SEMS) cannot go higher than a concentration factor of 2 at a peak temperature of 250°F. by using acid pretreatment, so that this was included primarily as a reference cost standard to which the multieffect plants could be compared.

Before we go into the results of the desalting plant studies, the results of the pretreatment of the seawater will be presented.

SEAWATER PRETREATMENT

The purpose of pretreating the seawater is to prevent scale formation on the heat transfer surfaces at elevated temperatures in the evaporators. Four processes were considered in detail, namely, acid, caustic, lime-magnesium carbonate (LMC), and ion exchange. The barium desulfating process was given preliminary consideration but was not considered a good choice for this particular application in Puerto Rico because its economics are related to the sale of by-product chemicals, an activity that would be inappropriate to the proposed desalting plant operator, the Puerto Rico Aqueduct and Sewer Authority.

The acid treatment process prevents the formation of alkaline scale deposits, calcium carbonate and magnesium carbonate, by removing temporary hardness, that is, by converting bicarbonates to sulfates and by removing the carbon dioxide.

Caustic pretreatment converts the bicarbonate to carbonate and results in precipitation of about 20% of the calcium carbonate in the seawater. This pretreatment was considered because the chlorine-caustic plant produces dilute caustic solution that could be used for the process.

TABLE 5. SEAWATER PRETREATMENT DATA

Item	Process			
	Acid	Caustic	LMC	Ion exch
Capital $\$ \times 10^{-3}$	440	680	2,750	2,630
Annual cost $\$ \times 10^{-3}$	264	328	555	577
Cost ¢/10^3 gal.	4.3	5.3	9.0	9.3

The lime-magnesium carbonate process adds magnesium carbonate and lime to the seawater to precipitate 70 to 80% of the calcium as calcium carbonate.

The ion exchange process uses a continuous countercurrent ion exchange column to adsorb and remove 65 to 70% of the calcium. The ion exchange resin is reactivated by using concentrated brine from the salt recovery plant.

The data developed for the four processes are compared in Table 5 and show that acid pretreatment is the lowest in capital cost as well as in the unit cost of water produced. This method of pretreatment was used for all the desalting plant design cases considered.

The data in Table 5 are based on an on-stream efficiency of 85%, producing 6.2×10^9 gal./yr. of desalted water from a desalting plant with a concentration factor of 3. The fixed charge rate on capital is assumed to be 7.835%. The costs of raw materials are assumed to be as follows:

Sulfuric acid	\$/ton	45
Caustic soda	\$/ton	40
Limestone	\$/ton	3.5
Coagulant	\$/ton	60
Fuel oil	\$/ton	18

DESALTING PLANTS

Five different desalting plant designs were considered as summarized in Table 6. Two multistage flash distillation plants were considered, one with a single-effect and the other with two effects. The single-effect multistage plant (SEMS) is the type of desalting plant most commonly built to date on which the greatest experience has been developed. It is, however, not capable of producing a brine at a concentration factor of 3 by using only acid pretreatment. It was primarily included to serve as a

TABLE 6. COMPARISON OF DESALT PLANTS

	Multistage flash		MEVTE		
	SEMS	MEMS	ORNL	WLB	A-G
Effects	1	2	11	16	12
Performance ratio	10.0	9.0	9.6	12.0	9.6
Capital $\$ \times 10^{-6}$	20.0	20.6	16.9	19.0	15.4
Annual cost $\$ \times 10^{-6}$	3.2	3.3	2.9	2.8	2.8
Cost of water ¢/10^3 gal.	52	53	47	44	44

point of reference on desalted water costs. The multieffect multistage flash distillation plant (MEMS) is similar to the SEMS except two effects are used and it can produce a brine of three times seawater concentration because the higher brine concentration occurs in the lower temperature effect.

Three multieffect evaporators which use vertical tubes with falling film evaporation were considered based on concepts developed by the Oak Ridge National Laboratory, Aerojet-General Corporation, and W. L. Badger Associates, Inc. The ORNL and Aerojet-General designs use multistage flash preheaters (MSF). Badger arranges the effects in two vertical towers. All the designs use enhanced heat transfer surface (fluted tubes) in the vertical tube evaporators.

The annual cost data are based on the fixed charge rate for capital of 7.835%, the use of exhaust steam from the dual-purpose nuclear power plant at 20¢/1,000 lb. and electrical power at 5 mills/kw.-hr.

The cost data given in Table 6 indicate that the multieffect vertical tube plants all produce water at lower costs than the multistage flash distillation plants. A waste brine stream at three times seawater concentration can be obtained from the VTE plants at no cost penalty to the production of water.

SALT RECOVERY PLANT

The salt recovery plant further evaporates the brine from the desalting plant to produce 1,035,000 short tons/yr. of salt and 8.3 million gal./day of desalted water. The design was developed by W. L. Badger Associates, Inc. A simplified flow sheet of salt recovery plants is shown in Figure 5. The first step in the salt recovery process is to heat the brine by using 40 lb./sq. in. gauge steam in a special contact chamber. This precipitates 75% of the calcium sulfate which collects in the bottom of the vessel. This prevents scale formation in the evaporators.

The further concentration of the brine is accomplished in an eight-effect, forward feed, vertical tube evaporator. The first effect is heated with 40 lb./sq. in. gauge steam, and the condenser of the last effect is maintained at 25 in. of vacuum with a steam jet. The salt concentration increases from 8.19 to 17.6% in this set of evaporators.

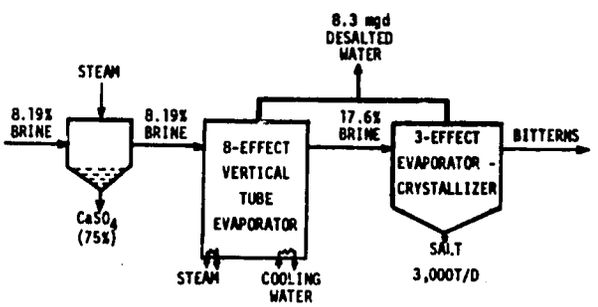

Fig. 5. Salt recovery flow diagram.

TABLE 7. SALT RECOVERY PLANT DATA

Salt production capacity	1,035,000 T/Y
Water production capacity	8.3 million gal./day
Direct fixed capital	$8,400,000
Allocated fixed capital	$2,600,000
Working capital	$ 300,000
Annual cost for sale	$2,100,000

Further concentration results in the beginning of crystallization, so salt recovery is done in a triple-effect backward feed externally heated forced circulation evaporator crystallizer. The concentrated brine is fed to the third effect of the evaporator. The salt concentration is increased to 22.6% without salt crystallization. However, calcium sulfate starts to precipitate. Scaling is prevented by keeping plenty of seed crystals in the brine slurry and by maintaining a high velocity through the heater. Excess calcium sulfate is withdrawn by means of a settler and centrifuge.

The concentrated brine from the top of the thickener is divided equally and fed to the first and second effects of the crystallizer evaporator. Salt and some calcium sulfate crystallize in these units and are separated in a centrifuge. The wet salt, which contains 0.7% calcium sulfate, is conveyed to a dissolving tank for solution transfer to the chlorine-caustic plant.

The capital cost and annual cost for sale are shown in Table 7. Figure 6 shows the discounted cash flow percentage for the salt recovery plant as a function of the selling price with the credit obtained for the by-product water as a parameter. Also shown in this figure is a dashed line drawn at $4.18/short ton which is the indicated price for the purchase of salt in 1975 based on delivery in 50,000 ton self-unloading ships.

BRINE PRODUCTS

The composition of the bitterns discharged from the salt production plant is shown in Table 8. Of the constituents listed in Table 8, the magnesium and potassium

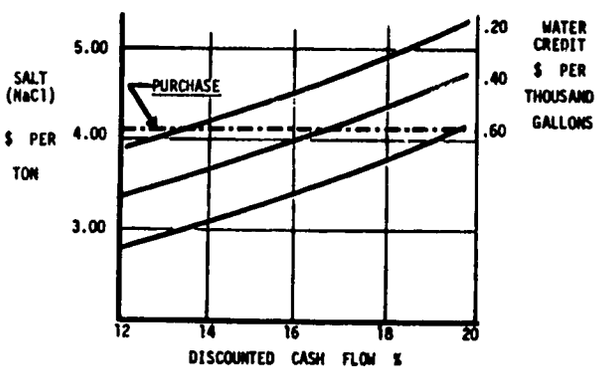

Fig. 6. Salt recovery plant returns.

TABLE 8. COMPOSITION OF BITTERN DISCHARGED BY SALT RECOVERY PROCESS

Ion	M/1,000 moles water
Chloride	134.22
Sodium	54.11
Magnesium	47.91
Sulfate	17.12
Potassium	8.90
Bromide	0.63
Calcium	0.55

appear in the bittern in amounts substantially undiminished from their respective quantitative occurrences in the original seawater. The quantity of bromide ion has suffered a slight but detectable loss resulting from its isomorphous replacement of the chloride ion in the salt removed in the salt recovery step. The latter process has removed approximately 90% of the total chlorine originally present in the seawater. The gypsum and anhydrite removed in the special equipment of the salt removal step have reduced the quantity of calcium to nearly 7% and the sulfate ion to about 70% of their respective values in normal seawater.

The bittern just described is saturated at 142°F. with respect to both sodium chloride and at least one other compound. Any further attempt to recover salt from it by continuing the process of evaporation will result in the coprecipitation of loeweite, a hydrated double sulfate of sodium and magnesium.

The ions present in the bittern will theoretically permit the recovery of the following materials in addition to sodium chloride: magnesium chloride, magnesium sulfate, sodium sulfate, potassium chloride, potassium sulfate, bromine (either as elemental bromine or as magnesium bromide), and various double salts, such as kainite or langbeinite. The calcium present may be eliminated from further economic consideration. From a practical standpoint, it is not present in amounts great enough to warrant either the recovery or the marketing of any of its compounds.

The marketing economics of the compounds listed above indicate the compounds with the greatest possibility of evolving into viable businesses: bromine (either elemental or as ethylene dibromide); potassium chloride, fertilizer grade; sodium sulfate, anhydrous; magnesium chloride, 36 to 40% solution, for magnesium metal. From a practical standpoint, these four products completely utilize all the dissolved constituents of the bittern. The only waste materials resulting from the recovery process will be a relatively small quantity of gypsum and the normally expected plumbing losses of brine.

The bromine is removed from the bitterns first. The process for recovery of bromine is well established. It consists of acidifying to $pH = 3.2$ to 3.5 and oxidizing the bromide to bromine by using chlorine. The bromine is then steamed out, condensed, and separated from the water phase. The bromine is then reacted with ethylene to form ethylene bromide.

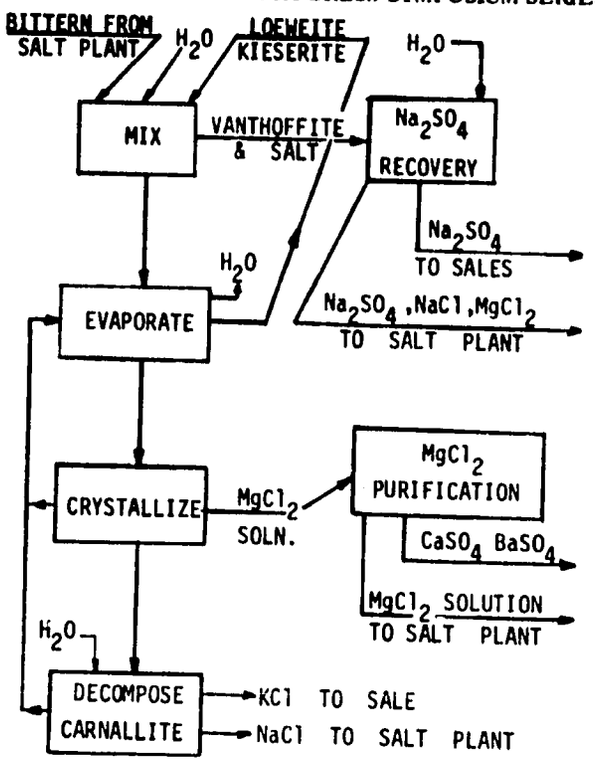

Fig. 7. Brine products block flow diagram.

After the bromine is removed, sodium sulfate, potassium chloride, and magnesium chloride are recovered by a process that consists basically of evaporation and fractional crystallization as shown in Figure 7. The data available seem to show that such a process to produce these products from seawater is feasible. However, most of the separation steps are undeveloped. It is expected that quite a bit of research and development work would be required to perfect this process.

TABLE 9. BRINE PRODUCTS DATA

Product	Quantity lb./yr. × 10^6	Selling price, $/lb.
Na_2SO_4	429.7	0.014
KCl	53.1	0.014
Br	5.5	0.150
Mg	100.0	0.305
Cl	220.0	—
H_2O	1.0 million gal./day	40¢/kgal.

(In Millions of dollars)	Brine products	Magnesium
Direct fixed capital	9.0	52.3
Allocated fixed capital	2.6	1.7
Working capital	1.1	5.6
Discounted cash flow %	13.9	

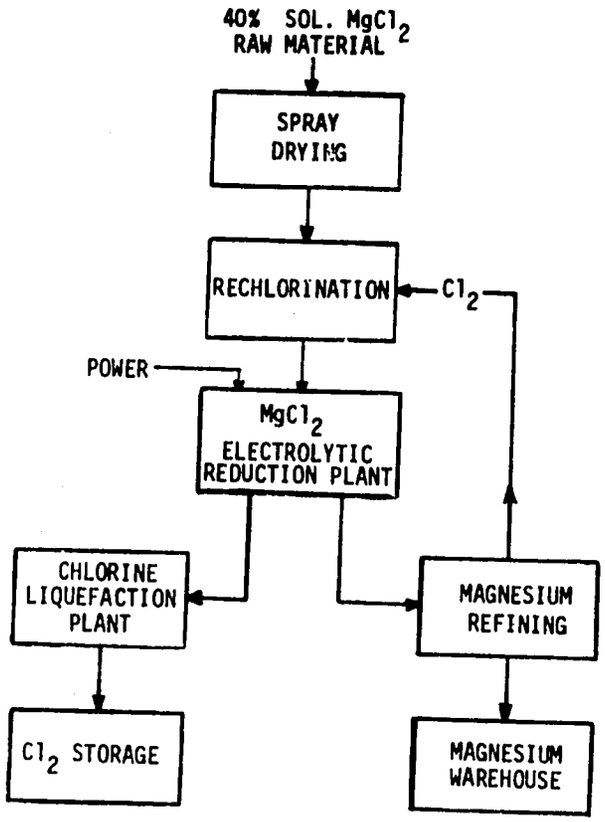

Fig. 8. Magnesium block flow diagram.

The magnesium chloride is converted to magnesium metal by the process shown in Figure 8. This process is used commercially.

The brine product recovery capacities and cost data are summarized in Table 9. Economic returns are on the lower edge of viability.

CONCLUSIONS

1. The proposed industrial development of the Guayama Sub-Region will result in a water shortage that will require development of all available sources including wells, the Toa Vaca Dam project, and desalting plants to satisfy the requirement most economically.

2. The approximate cost of the various water supplies is as follows:

Wells	5¢/1,000 gal.
Toa Vaca	20¢/1,000 gal.
Desalted	45¢/1,000 gal.

3. Multieffect vertical tube type of plants can produce desalted water as cheap or cheaper than multistage flash distillation plants in the 20 million gal./day size range. The VTE type of desalting plant produces a waste brine at three times seawater concentration at no cost penalty to water production cost.

4. Salt can be economically recovered by using multieffect evaporator plants in the 3,000 ton/day range when starting with free brine at three times seawater concentration, with steam at 18.5¢/1,000 lb. and with a market for the sale of the by-product water of 8.3 million gal./day. Some development work is needed.

5. Brine products theoretically can be efficiently recovered from the waste stream of the salt recovery plant. Economic return is on the lower edge of viability. Extensive research and development work would have to be done.