**Parametric Study and Preliminary
Evaluation of Reverse Osmosis for
Seawater Desalination**

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MASTER

OAK RIDGE NATIONAL LABORATORY

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REVERSE OSMOSIS FOR SEAWATER DESALINATION

P. Glueckstern S. A. Reed
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CONTENTS

	<u>Page</u>
ACKNOWLEDGMENT	v
ABSTRACT	1
1. INTRODUCTION	1
2. SUMMARY AND CONCLUSIONS	3
3. SCOPE AND OBJECTIVES	7
4. EVALUATION METHOD	8
5. RO PLANT CONCEPTS EVALUATED	9
5.1 Process Description and Basic Assumptions	9
5.2 Basic Design Criteria	11
6. ECONOMIC GROUND RULES	12
6.1 Bases for Estimating the Direct Capital Cost	12
6.2 Bases for Estimating Indirect Capital and Total Annualized Capital Costs	13
6.3 Bases for Estimating the Operating Costs	13
7. RESULTS FOR OPTIMIZED RO SYSTEMS	16
7.1 Optimum Product Recovery and Other Design Data	16
7.2 Capital Cost of the Optimized RO Designs	16
7.3 Unit Water Costs from the Optimized RO Designs	16
7.4 Summary of Economic Ground Rules and Evaluation of Unit Water Cost	27
8. ANALYSIS OF RESULTS	28
8.1 Range of Capital and Unit Water Costs	28
8.2 Effect of Seawater Salinity and Temperature	28
8.3 Effect of System Pressure and Membrane Life	30
8.4 Effect of Energy Cost and Comparison with Multistage Flash Distillation Plants	30
8.5 Overall Sensitivity Analysis of Technological and Cost Parameters	34
REFERENCES	37
APPENDIX A. Computer Printout of a 2.5-Mgd RO Plant Based on Current Technology	39
APPENDIX B. Computer Printout of a 2.5-Mgd RO Plant Based on Projected Technology	43
APPENDIX C. Conversion Factors	47

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ABSTRACT

A parametric study of one- and two-stage reverse osmosis (RO) systems for seawater desalination was made with the aid of a recently developed computer program for optimization of RO systems. Membrane manufacturer's performance data and predicted cost data were applied. Based on these data and other equipment and operating costs considered, a preliminary evaluation of the economics of RO for seawater desalination was made. Plant capacities in the range of 1 to 12.5 Mgd were considered and compared economically with the current most common multistage flash distillation process at different energy price levels.

1. INTRODUCTION

Recently the reverse osmosis (RO) process, which is now most commonly applied for quality improvement of brackish water, has also become a potential contender for seawater desalting. The RO process may be attractive for small plant capacities, especially when the dual-purpose electric power-seawater desalting cannot be applied. For this situation, the less energy consuming RO process is becoming more and more attractive, as energy costs continue to increase.

Although RO membranes for seawater conversion have been demonstrated at several locations, in the last two years, this technology is not yet fully commercial and cannot be considered as "proven technology." This is due to several operational uncertainties which strongly affect the economics of the process, namely, membrane life and the required raw seawater pretreatment. These factors are affected primarily by the quality of the raw seawater and the developmental status of membrane technology.

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No detailed economic evaluation of the RO process for seawater desalination has been published. Most of the available data concerning economics was published by membrane manufacturers involved in developing membranes for seawater application. Usually, only overall figures predicting the cost for "projected technology" were stated. Using this type of information, a preliminary cost study employing thin film cellulose triacetate membranes, was published recently.¹

The main objective of the present investigation was to perform a parametric study to identify the interaction between the various operational and economic parameters, and to make a preliminary evaluation of the economics of RO for seawater desalination for a reasonable set of operational and economic assumptions. This investigation includes much more detailed cost information than other previously published cost estimates.

2. SUMMARY AND CONCLUSIONS

This investigation was basically a parametric study of "near-commercial" RO technology for seawater desalination. Based on performance data obtained from membrane manufacturers and various technological and economical assumptions, optimal systems were designed and evaluated economically for several plant sizes in the 1- to 12.5-Mgd range.

As detailed membrane performance data (as a function of the operating parameters) were obtained only for hollow-fibre aromatic polyamide membranes, the parametric study was based mainly on this membrane technology. Currently, the largest module size (8 in. diam) at reference operating conditions (800 psig, 25°C (77°F), 30% recovery and a feed salinity of 30,000 ppm NaCl) has a productivity of 5000 gpd and a salt rejection of 98.5%. The assumed costs of these modules (termed "current technology") were \$0.75/gpd (\$3750/module) for initial cost including housing, and \$0.60/gpd for replacement membranes.

In order to evaluate the effect of predicted near-term developments (termed "projected technology"), a 12-in. module with a productivity of 10,000 gpd (at reference operating conditions) has been assumed to cost \$0.50/gpd initial cost and \$0.40/gpd membrane replacement cost.

The bases for evaluating the capital costs of the various optimized RO systems were as follows:

1. Direct costs of basic RO equipment:
 - Initial membrane module costs: Unit costs (\$/gpd at reference operating conditions) as mentioned above.
 - Other standard RO equipment cost (pumps, piping, instrumentation etc.): Detailed cost functions based on conceptual designs of brackish water RO systems adapted to two-stage seawater RO systems and escalated to First Quarter 1976 prices.²
2. Site and pretreated seawater supply:
 - Site and supply of raw seawater: Cost functions applied to ORNL's computer programs for seawater distillation processes.
 - Raw seawater pretreatment system: Updated cost of equipment recommended by Ralph Stone and Co. Inc., based upon the findings of the Chula Vista, California, Pilot Plant Testing program to evaluate seawater pretreatment methods (OSW Report No. 890).³

3. Indirect capital cost and contingency:

— Engineering, design and supervision; interest during construction; escalation during construction and owner's cost for training and start up: Cost functions applied in ORNL's computer programs for seawater distillation processes assuming a 10% interest rate and a 8% annual escalation rate.

— Contingency: 15% of all direct and indirect capital costs.

For "open ocean" seawater with a salinity of 3.5% and an average feed temperature of 25°C (77°F), the resulting capital costs for optimized two-stage systems were as follows:

	(\$/gpd)	
	1.0 Mgd systems	12.5 Mgd systems
(a) For current technology:		
Basic RO equipment	2.16	1.94
Site, seawater intake & pretreatment	1.14	0.56
Indirect capital & contingency	1.91	1.53
Total without power recovery turbine	5.21	4.03
Power recovery turbine (including indirect costs)	0.22	0.05
Total with power recovery turbine	5.43	4.08
(b) For projected technology:		
Basic RO equipment	1.60	1.28
Site, seawater intake & pretreatment	0.99	0.50
Indirect capital and contingency	1.49	1.07
Total without power recovery turbine	4.08	2.85
Power recovery turbine (including indirect costs)	0.19	0.04
Total with power recovery turbine	4.27	2.89

It is noted that the resulting capital cost is approximately twice as much as the figures recently estimated by membrane manufactures. The major factors for the higher estimated costs in this study were the more conservative assumptions regarding long-term membrane performance (log flux decline), materials of construction and standby equipment, higher

cost for site and pretreated seawater supply and the much higher allowance for indirect costs and contingency (usually underestimated if applied at all by membrane manufacturers). Obviously, a meaningful comparison can only be made by applying identical ground rules concerning site development, seawater intake, pretreatment and indirect capital cost items. The assumptions made in the current study (excluding pretreatment) were identical to those made in the evaluation of current technology seawater distillation process, recently made at the Oak Ridge National Laboratory (ORNL), so that a comparative analysis can be made on an objective basis.

By comparing the capital cost of a 1-Mgd RO seawater desalination plant with that of a 1-Mgd multistage flash (MSF) plant, it was found that the capital cost of the RO plant based on current technology is approximately the same as that of an MSF plant with a performance ratio of 13.4 lb/1000 Btu (optimized for high energy cost). The cost of an RO plant based on projected technology would, however, be approximately 20% lower.

The resulting unit water costs, calculated for a 10% interest rate, 85% plant factor and a \$0.020/kWhr electricity cost, were as follows:

	(\$/kgal)	
	1.0-Mgd systems	12.5-Mgd systems
For current technology	4.70	3.90
For projected technology	3.30	2.50

The long-term membrane performance factors which must be realized in order to obtain the above costs are a membrane life of 2 years and log flux decline of 0.05 for current technology, and a membrane life of 3 years and a log flux decline of 0.03 for projected technology.

It should be emphasized that the unit water cost is very sensitive to those not yet fully proven (even the current technology assumptions) membrane performance factors. On the other hand, the assumed operating pressure (800 psig), as well as other operating and performance factors, are based on a rather conservative approach. After some further development, it is probable that capital and unit water costs will eventually be lower than those obtained for projected technology. The results of a

sensitivity analysis made for a 12.5 Mgd system indicated that a set of optimistic operational and costing assumptions would, if realized, eventually reduce unit water cost to approximately \$1.50/kgal, compared to approximately \$2.50/kgal obtained for projected technology based on a conservative approach. It is, however, questionable if optimistic assumptions such as higher operating pressures and longer membrane life can be realized simultaneously.

One of the most attractive features of the RO process, in comparison to distillation processes, is obviously its lower energy requirement. RO systems are therefore considerably less sensitive to energy price fluctuations. The comparison made with single- and dual-purpose MSF plants indicated that current technology RO systems can compete with single-purpose plants in the 1-Mgd size at locations with prevailing high energy costs, while projected technology RO systems are not realistically competitive with small MSF dual-purpose plants in the 1.0- to 2.5-Mgd range.

The unit water costs from the RO systems evaluated were found to be considerably less sensitive to energy price levels than other desalting processes. A comparison made with single- and dual-purpose multistage flash distillation (MSF) plants indicated that RO systems, at the present stage of technology, might eventually be competitive with single-purpose MSF plants in the 1-Mgd range at locations with prevailing high energy costs.

The development potential of RO technology for seawater desalination was analyzed by making several basic assumptions for an improved 12-in.-diameter module with extrapolated initial performance data based on the current-technology 8-in. module. The basic assumptions for this module termed "projected technology" were: (a) a unit membrane cost decrease from \$0.75 to \$0.50/gpd, (b) a projected membrane life of 3 instead of 2 years, and (c) a lower log flux decline factor of 0.03 rather than 0.05. If these assumptions could be realized, the unit water cost would be decreased by about 30 to 35%. In this case, RO technology is not realistically competitive with small MSF dual-purpose plants in the 1-25 Mgd range at locations even with prevailing high fuel costs.

3. SCOPE AND OBJECTIVES

The objectives of the present study were as follows:

1. To identify the interaction between operational and economical parameters of the RO process for seawater desalination.
2. To make a preliminary evaluation of the economics of the RO process for a reasonable set of operational and economical assumptions.
3. To compare the potential economics of the RO process with the most commonly used multistage flash distillation process, at various site and economic conditions.

The study was based on performance data obtained from membrane manufacturers. Most of the data were obtained for hollow-fibre polyamid membranes. The parametric study considered, mainly, this membrane type. The effects of most operational parameters, such as system pressure, feed temperature and product recovery on membrane performance, would probably not be very different for other types of seawater RO membranes having similar salt rejection properties. Most of the results of the parametric study can therefore be considered indicative for the RO process in general.

The economic evaluation considered plant capacities in the range of 1.0 to 12.5 Mgd.

The parametric study considered the following range of feed, operational, and economic variables:

Feed salinity	25,000-45,000 ppm
Feed temperature	18-35°C (65-95°F)
First stage pressure	750-1000 psi
First stage product recovery	10-50%
Second stage product recovery	50-90%
Membrane life	1-5 years
Log flux decline	0.02-0.05/m/
Electrical power cost	0.02-0.05 (\$/kWhr)
Independently operating train sizes	0.5-1.25 Mgd
Membrane module sizes*	5000 and 10,000 gpd
Unit membrane costs*	
First stage initial cost	0.75 and 0.50 (\$/gpd)
First stage replacement cost	0.50 and 0.40 (\$/gpd)
Second stage initial cost	0.20 and 0.15 (\$/gpd)
Second stage replacement cost	0.15 and 0.10 (\$/gpd)

*At reference operating conditions defined by the manufacturer.

4. EVALUATION METHOD

The study was performed with the aid of the computer program RO-75,² developed at ORNL. This program was aimed at optimizing RO systems applying defined membrane technology, by varying the operational parameters, such as system pressure, feed temperature and product recovery in a range defined by the membrane manufacturers. For each specific requirement of plant capacity and product quality, RO-75 defines a reference design based on the applied membrane performance data. This reference design will apply (optionally, upon request) to a two- or one-stage membrane system, provided the rejection rate of the applied membrane is high enough, and can include, or not include, a power recovery turbine generator set. The capital cost of the reference design was then evaluated by applying appropriate cost functions to all required site work (buildings, seawater intake and outfall, etc.) and all other process equipment, including pumps and drivers, valves, piping, instrumentation, electrical equipment, pre- and posttreatment systems and the power recovery turbines (if used). The cost of the membrane modules is computed by multiplying the number of required modules and the assumed cost per module. The unit water cost corresponding to this design is then evaluated on the basis of a defined set of operational and economic ground rules (see Sects. 5 and 6).

The reference designs were optimized (with the aid of a special subroutine included in RO-75), by varying the operational parameters in a defined range. On the basis of the cost data obtained, only the recovery ratios of the first stage and, if applicable, a second stage, could be optimized.

5. RO PLANT CONCEPTS EVALUATED

5.1 Process Description and Basic Assumptions

The schematic flow diagram of the RO plants which were evaluated is shown in Fig. 1. The inlet feed was assumed to come from seawater wells for small plant capacities (up to 5 Mgd) and via a submarine line for larger plant capacities. The raw seawater is pumped to a pretreatment system, which has to be designed in accordance with raw seawater quality and the membrane technology. In the present evaluation the pretreatment system, recommended in OSW Research and Development Progress Report No. 890,³ has been adopted. This pretreatment includes chlorination (10 ppm Cl_2), detention (40 min), sand filtration (2.36 gpm/ft^2), and activated carbon filtration (4.60 gpm/ft^2).

In addition to the equipment required to remove organic material and residual chlorine, and to reduce the turbidity level below 1 JTU,* an acid injection system for pH adjustment (to prevent calcium carbonate precipitation and hydrolysis in the case of cellulose acetate membranes) and micron filter cartridges (5 microns for hollow-fibre membranes) were included. From the pretreatment system the treated feed is pumped by centrifugal pumping units to a number of trains composed of parallel (and optionally tapered) RO modules. Reference train sizes of 0.5 to 1.25 Mgd were considered. The outcoming permeate, or product water, is transferred to a storage tank and the concentrated brine is rejected to the outfall system or optionally via a power recovery turbine.

For two-stage systems, a part or all of the permeate produced by the first stage is pumped by the second-stage high-pressure feed pumps through the second stage RO modules, arranged in one or more trains. The rejected brine from the second stage, which is usually at lower salinity than the incoming seawater, is recycled to the inlet of the first stage while the permeate is transferred to the posttreatment system. This system consists of a decarbonating tower and a neutralizing agent (caustic or soda-ash) feeder. Finally, the treated product is pumped by one, or more, pumping units to the water supply system.

*Jackson turbidity units.

Fig. 1. Schematic diagram of reverse osmosis plant.

5.2 Basic Design Criteria

As discussed in the previous sections the present evaluation was based on actual membrane performance data for specified modules which have been published by membrane manufacturers. These data as well as the basic criteria of the associated process equipment on which the evaluation has been based are summarized in Table 1.

Table 1. Summary of basic design criteria

	First stage	Second stage
1. <u>Membrane modules:</u>		
Membrane material	Aromatic polyamide	
Membrane configuration	Hollow fibre	Hollow fibre
Manufacturer	Du Pont	Du Pont
Type	B-10 ^a	B-9
Fibre bundle diameter, in.	8	8
Initial productivity, gpd	5000 ^b	14,000 ^c
Initial rejection, %	>98.5 ^b	>90.0 ^c
Rated operating pressure, psig	800	400
Temperature range, °F	0-30°C (32-86°F)	0-35°C (32-95°F)
Recovery range, %	10-50	25-90
3. <u>Train sizes:</u>		
0.5 Mgd (max) for 8-in. diam modules		
1.25 Mgd (max) for 12-in. diam modules		
3. <u>High-pressure pumping units:</u>		
Number of units:	One per train + one standby for five trains	
Overall efficiency:	76%	
4. <u>Seawater delivery pump:</u>		
Head requirement:	138 ft	
Overall efficiency:	82%	
5. <u>Product pump delivery pump:</u>		
Head requirement:	92 ft	
Overall efficiency:	82%	
6. <u>Overall efficiency of power recovery:</u>		
Turbine-generator:	65%	
7. <u>Pretreatment system:</u>		
Chlorination:	10 ppm Cl ₂	
Detention:	40 min	
Sand filtration:	2.36 gpm/ft ²	
Activated carbon filtration:	4.60 gpm/ft ²	
Cartridge filters:	5 microns	
Acid injection:	2 × 100% pumps sized to inject up to 100 ppm H ₂ SO ₄ (93%)	
Chemical cleaning systems:	Sized for cleaning one train	
8. <u>Posttreatment system:</u>		
Decarbonating tower + soda-ash or caustic feeder		

^aThe 8 in. B-10 is not yet commercial and the cited performance data was based on Ref. 4 and extrapolation of the commercial 4 in. B-10. For "projected technology" a 12-in. module having the same basic characteristics has been assumed.

^bBased on operation with a feed of 30,000 ppm NaCl at 800 psig, 77°F and 30% recovery.

^cBased on operation with a feed of 1500 ppm NaCl at 400 psig, 77°F and 75% recovery.

6. ECONOMIC GROUND RULES

6.1 Bases for Estimating the Direct Capital Cost

Two seawater membrane module costs, including housing, were assumed. A cost of \$0.75 per gpd was assumed for current technology 8-in.-diam modules and \$0.50 per gpd for projected technology 12-in.-diam modules. These unit costs correspond to the reference productivity rates defined in Table 1. Similarly, two different cost values were assumed for the second stage membrane modules; a unit cost of \$0.20 per gpd for current technology 8-in.-diam brackish water modules and a unit cost of \$0.15 for projected technology 12-in.-diam brackish water modules.

The costs of high-pressure pumping units were based on a cost function presented in Ref. 5, escalated to First Quarter 1976 prices. The costs of other pumping units were evaluated with cost functions applied in ORNL's computer programs MSF21⁶ and VTE21⁷ escalated to First Quarter 1976 prices.⁸

The costs of power recovery turbines were based on equipment cost functions published in OSW Research and Development Report No. 457⁹ escalated to First Quarter 1976 prices. A 100% additional cost was applied for mechanical and electrical installations.

The costs of seawater intake and outfall systems and that of the product water posttreatment systems were calculated on the basis of cost data applied in ORNL's computer programs VTE21 and MSF21. However, for small plant capacities, it was assumed that seawater wells will be cheaper than a long submarine intake by about 20%.

The capital costs of valves and piping, electrical equipment, and instrumentation were based on cost data from the Fluor study for the OSW,¹⁰ downscaled to smaller size trains and escalated to First Quarter 1976 prices.

The cost of site development was based on cost functions published in Ref. 10. For buildings, a unit cost of \$16.50/ft² was used.

The pretreatment system, as already discussed in Sect. 5, utilized equipment recommended in OSW Report No. 890.³ The capital cost for this equipment was upscaled to the capacities required by applying a scale factor of 0.7 and escalated to First Quarter 1976 prices.

6.2 Bases for Estimating Indirect Capital and Total Annualized Capital Costs

The indirect capital costs included the following components:

- (a) Engineering design and supervision.
- (b) Escalation during construction.
- (c) Interest during construction.
- (d) Owner cost for training and startup.

The first three components were calculated by the same cost function applied in the evaluation of distillation plants,¹¹ by applying an escalation rate of 8% and an interest rate of 10%. The owner costs were assumed to be 1.9% of the total direct and indirect capital costs. Finally, a contingency of 15% on all capital (direct + indirect) costs was applied to estimate the total investment.

The total annualized capital costs were calculated on the basis of the following economic ground rules:

- interest rate: 10%
- sinking fund depreciation rate based on a 30-year plant life
- insurances and taxes: 1%

6.3 Bases for Estimating the Operating Costs

The operating costs included the following components:

- (a) Membrane replacement
- (b) Electrical power
- (c) Operation and maintenance labor
- (d) Supplies and maintenance materials
- (e) Chemicals

The membrane replacement costs were calculated on the basis of unit membrane replacement cost per gpd, corresponding to the two assumed values of initial membrane cost (including housing), and the assumed membrane life. The latter was evaluated parametrically; however, for the reference designs, lives of 2 and 3 years were assumed for seawater membranes based on current and projected technology, respectively. For the second-stage membranes a 3-year life was assumed for both. The applied unit membrane replacement costs are listed in Table 2.

Table 2. Summary of economic ground rules and parameters

Interest rate	10%
Insurances and taxes	1%
Plant operating factor	85%
Equipment cost basis	First Quarter 1976
Plant life	30 years
Electric power cost (reference)	\$0.02/kWhr
Contingency	15%
Annual escalation of costs during construction	8%
O&M labor cost	\$16,000/man-year (av)
General and administration cost	40% of O&M labor
Annual costs for supplies and materials	0.7% of total capital cost

Unit prices of chemicals:

Sulfuric acid	\$31/ton (bulk, 100% basis)
Chlorine	\$176/ton

	<u>Current technology</u>	<u>Projected technology</u>
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Unit costs of membrane modules:

Initial costs, \$/gpd

First stage	0.75	0.50
Second stage	0.20	0.15

Replacement cost, \$/gpd

First stage	0.60	0.40
Second stage	0.15	0.10

Membrane life, years

First stage	2	3
Second stage	3	3

Log flux decline, /m/

0.05	0.03
------	------

For the electrical power cost, a range of \$0.02 to \$0.05/kWhr was used. The reference designs however were optimized for an assumed power cost of \$0.02/kWhr.

The annual operation and maintenance labor costs of a 1-Mgd plant were calculated on the basis of 5 men at an average cost of \$16,000/man-year, with an additional general and administration cost of 40% of direct labor cost. For larger plants a scale factor of 0.5 was applied.

The annual costs for supplies and maintenance materials were assumed to be 0.7% of the total capital costs.

The annual costs for chemicals were based on the specific consumptions listed in Table 1 and the unit costs of chemicals listed in Table 2. An allowance of \$0.05 per 1000 gallons of product water was made for chemical cleaning of the membranes.

7. RESULTS FOR OPTIMIZED RO SYSTEMS

7.1 Optimum Product Recovery and Other Design Data

The resulting optimum product recovery ratios versus electric power cost, plant capacity, feed salinity and membrane cost, for systems with and without power recovery, are summarized in Table 3.

Typical design characteristics of two-stage systems applying 8 in. (current) and 12 in. (projected) hollow-fibre, polyamid, membrane modules are presented in Table 4, for 1- and 12.5-Mgd plant capacities. More detailed design data for a 2.5-Mgd plant optimized for an electrical power cost of \$0.05/kWhr is shown in the computer printouts included in the Appendices A and B, for plants applying current and projected technology, respectively.

7.2 Capital Cost of the Optimized RO Designs

The resulting capital cost versus plant capacity and feed salinity of optimized current technology (8 in. membrane modules) and projected technology (12 in. membrane modules) RO systems applying power recovery are summarized in Table 5.

A detailed capital cost breakdown for 1- and 12.5-Mgd plant capacities, operating with a feed salinity of 35,000 ppm, is presented in Table 6.

The capital costs versus plant capacity and versus feed salinity are also illustrated in Figs. 2 and 3, respectively.

7.3 Unit Water Costs from the Optimized RO Designs

The resulting unit water costs versus electric power cost, plant capacity, feed salinity and unit membrane cost for systems with and without power recovery are summarized in Table 7.

A detailed unit water cost breakdown for 1- and 12.5-Mgd plant capacities (for which the capital cost breakdown is given in Table 6) is presented in Table 8.

The unit water costs versus plant capacity and versus seawater salinity are also illustrated in Figs. 4 and 5, respectively.

Table 3. Optimum fraction recovery vs electric power cost, plant capacity, feed salinity and membrane cost^a

Power cost, \$/kWhr		0.02				0.05			
Power recovery		Yes		No		Yes		No	
Plant capacity, m ³ /d (Mgd)	3785	47,318	3785	47,318	3785	47,318	3785	47,318	
	1.0	12.5	1.0	12.5	1.0	12.5	1.0	12.5	
<u>High membrane cost^c</u>									
<u>Feed salinity, ppm</u>									
25,000	0.45	0.40	0.44	0.42	0.46	0.44	0.50 ^b	0.48	
30,000	0.35	0.33	0.35	0.34	0.35	0.35	0.35	0.35	
35,000	0.27	0.25	0.29	0.27	0.30	0.29	0.32	0.31	
40,000	0.22	0.23	0.25	0.23	0.26	0.25	0.28	0.27	
45,000	0.19	0.18	0.20	0.19	0.21	0.21	0.23	0.22	
<u>Low membrane cost^c</u>									
<u>Feed salinity, ppm</u>									
25,000	0.48	0.47	0.48	0.50 ^b	0.50 ^b	0.50 ^b	0.50 ^b	0.50 ^b	
30,000	0.35	0.35	0.39	0.36	0.40	0.37	0.42	0.41	
35,000	0.35	0.30	0.34	0.33	0.34	0.34	0.39	0.36	
40,000	0.29	0.27	0.30	0.38	0.30	0.29	0.32	0.31	
45,000	0.23	0.22	0.25	0.23	0.27	0.25	0.27	0.27	

^aOptimized for conditions defined in Table 1.

^bOne-stage design.

^cSee Table 2.

Table 4. Major design characteristics of RO systems
for seawater desalination^a

Membrane type: hollow-fibre-aromatic polyamide

System pressure: 800 psig (first stage); 400 psig (second stage)

Feed parameters: temperature (25°C); salinity (35,000 ppm)

Membrane module diameter	8 in.		12 in.	
Plant capacity, Mgd	1.0	12.5	1.0	12.5
Number of trains				
First stage	2	25	1	10
Second stage	1	5	1	2
1. Number of modules				
First stage	328	3962	173	1951
Second stage	25	287	15	173
2. Product recovery, %				
First stage	28.1	25.8	36.2	31.2
Second stage	88.5	90.0	84.4	90.0
3. Total production, Mgd				
First stage	1.035	12.840	1.070	12.960
Second stage	0.275	3.070	0.375	4.150
4. Product salinity, ppm				
From first stage	645	620	735	680
From second stage	120	125	115	140
5. Electric power				
Total installed, kW ^b	1315	17,680	1085	15,070
Specific consumption, kWhr/1000 gal	22.80	24.20	19.75	21.25

^aOptimized for conditions defined in Table 1.

^bOperating with power recovery.

Table 5. Capital cost^a of RO desalting plants vs unit membrane cost, feed salinity and plant capacity, \$/gpd

Membrane module diameter	8 in.				12 in.			
Unit membrane cost \$/gpd	0.75				0.50			
Plant capacity, Mgd	1.0	2.5	5.0	12.5	1.0	2.5	5.0	12.5
Feed salinity, ppm								
25,000	4.00	3.47	3.21	2.97	3.19	2.60	2.33	2.10
30,000	4.53	3.91	3.63	3.36	3.67	2.98	2.65	2.38
35,000	5.43	4.71	4.39	4.08	4.27	3.54	3.19	2.89
40,000	6.41	5.61	5.26	4.94	4.94	4.16	3.77	3.46
45,000	7.61	6.76	6.40	6.06	5.81	4.94	4.53	4.20

^aSystems include power recovery turbine; optimized for an electric power cost of \$0.020/kWhr.

Table 6. Capital cost breakdown of RO plants vs unit membrane cost and plant capacity

Membrane type: hollow-fibre-aromatic polyamide

System pressure: 800 psig (first stage); 400 psig (second stage)

Feed parameters: temperature (25°C); salinity (35,000 ppm)

Membrane module diameter	8 in.				12 in.			
Unit membrane cost, \$/gpd	0.75				0.50			
Plant capacity, Mgd	1.0		12.5		1.0		12.5	
	\$ × 10 ³	%	\$ × 10 ³	%	\$ × 10 ³	%	\$ × 10 ³	%
I. Direct capital components								
Modules and housing	1,300	37.74	15,661	49.58	928	34.23	10,486	46.54
Pumps and motors	369	10.71	3,313	10.49	334	12.32	2,262	10.04
Valves and piping	281	8.16	2,833	8.97	179	6.60	1,611	7.15
Instrumentation	94	2.73	1,175	3.72	54	1.99	565	2.51
Electrical equipment	118	3.43	1,268	4.01	101	3.73	1,084	4.81
Pretreatment system	695	20.17	4,161	13.17	587	21.65	3,651	16.21
Posttreatment system	28	0.81	109	0.35	28	1.03	109	0.48
Seawater intake or wells	238	6.91	1,921	6.08	194	7.16	1,651	7.33
Site and buildings	183	5.31	769	2.43	183	6.75	769	3.41
Power recovery turbine	139	4.03	376	1.20	123	4.54	342	1.52
Total direct	3,445	100.00	31,586	100.00	2,711	100.00	22,530	100.00
II. Indirect capital components	1,273	36.94	12,722	40.28	999	36.82	8,876	39.40
Total before contingency	4,718		44,308		3,710		31,406	
III. Contingency	707	20.54	6,646	21.04	557	20.52	4,711	20.91
Total investment	5,425		50,954		4,267		36,117	

ORNL-DWG 76-2949

Fig. 2. Capital cost vs reverse osmosis plant capacity.

Fig. 3. Unit water cost vs reverse osmosis plant capacity (35,000 ppm seawater salinity - current and projected technology)

Table 7. Unit water cost from RO plants vs electric power cost, plant capacity, feed salinity and membrane cost (\$/1000 gal)^a

Power cost, \$/kWhr		0.02				0.05			
Power recovery		Yes		No		Yes		No	
Plant capacity, Mgd		1.0	12.5	1.0	12.5	1.0	12.5	1.0	12.5
<u>High membrane cost</u>									
<u>Feed salinity, ppm</u>									
25,000		3.50	2.80	3.50	2.90	3.95	3.25	4.05 ^b	3.45
30,000		3.90	3.20	3.95	3.30	4.45	3.75	4.70	4.05
35,000		4.70	3.90	4.80	4.05	5.35	4.60	5.65	4.95
40,000		5.60	4.75	5.70	4.95	6.35	5.55	6.75	6.00
45,000		6.75	5.90	6.90	6.75	7.70	6.85	8.75	7.40
<u>Low membrane cost</u>									
<u>Feed salinity, ppm</u>									
25,000		2.45	1.80	2.50	1.85 ^b	2.90 ^b	2.20 ^b	3.00 ^b	2.35 ^b
30,000		2.80	2.05	2.85	2.15	3.30	2.55	3.50	2.80
35,000		3.30	2.50	3.35	2.60	3.90	3.10	4.10	3.40
40,000		3.80	3.00	3.90	3.15	4.50	3.70	4.80	4.05
45,000		4.50	3.60	4.65	3.80	5.35	4.45	5.70	4.90

^aOptimized for conditions defined in Table 1.

^bOne-stage design.

Table 8. Unit water cost breakdown of RO plants vs
unit membrane cost and plant capacity

Membrane type: hollow-fibre-aromatic polyamide

System pressure: 800 psig (first stage); 400 psig (second stage)

Feed parameters: temperature (25°C); salinity (35,000 ppm); electrical power cost (\$0.02/kWhr)

Membrane module diameter		8 in.				12 in.			
Unit membrane cost, \$/gpd		0.75				0.60			
Plant capacity, Mgd		1.0		12.5		1.0		12.5	
Unit water costs		\$/kgal	%	\$/kgal	%	\$/kgal	%	\$/kgal	%
1. Capital		1.95	41.6	1.44	37.1	1.54	46.7	1.03	41.4
2. Membrane replacement		1.56	33.2	1.51	38.9	0.72	21.8	0.65	26.1
3. Power cost		0.46	9.8	0.48	12.4	0.39	11.8	0.42	16.9
4. Operating cost									
Staff		0.36	7.7	0.10	2.6	0.36	10.9	0.10	4.0
Maintenance + supplies		0.12	2.6	0.09	2.3	0.10	3.0	0.07	2.8
Chemicals		0.24	5.1	0.26	6.7	0.19	5.8	0.22	8.8
Total operating		0.72	15.4	0.45	11.6	0.65	19.7	0.39	15.6
Total		4.69	100.0	3.88	100.0	3.30	100.0	2.49	100.0

Fig. 4. Capital cost vs seawater salinity and plant capacity — current technology.

Fig. 5. Unit water cost vs seawater salinity and plant capacity (current technology).

7.4 Summary of Economic Ground Rules and Evaluation of Unit Water Cost

All annual operating costs and the corresponding unit water costs were based on a plant operating factor of 85%. A summary of the applied economic ground rules is presented in Table 2.

8. ANALYSIS OF RESULTS

8.1 Range of Capital and Unit Water Costs

For the assumed unit membrane cost of \$0.75/gpd, representing current membrane technology and the other cost data detailed in Sect. 6, the resulting capital costs ranged between approximately \$5.40 to \$4.10/gpd, for RO plant capacities between 1.0 and 12.5 Mgd, respectively. These costs correspond to plants operating with an "open ocean" seawater salinity of 35,000 ppm and a feed temperature of 25°C (77°F). The effect of these site parameters is analyzed in the next subsection.

For the assumed lower unit membrane cost of \$0.50/gpd, representing projected technology and the other potential cost savings made possible by application of larger size membrane modules, the resulting capital costs ranged between approximately \$4.30 and \$2.90/gpd, for RO plant capacities of 1.0 to 12.5 Mgd.

The corresponding unit water cost, based on the assumed economic ground rules listed in Table 2, ranged between approximately \$4.70 and \$3.90 per 1000 gallons for the current technology, and between \$3.30 and \$2.50 for projected technology.

8.2 Effect of Seawater Salinity and Temperature

From the resulting capital costs versus seawater salinity shown in Table 5 and Fig. 3, and the unit water costs in Table 7 and Fig. 5, it is quite evident that the prevailing seawater salinity is one of the key-factors in determining the competitiveness of the RO process for seawater desalting. The resulting unit water costs for a low seawater salinity of 25,000 ppm were about 25 to 20% lower, and those for a high seawater salinity (e.g., prevailing in the Red Sea) of 45,000 ppm, were about 40 to 50% higher than those obtained with reference salinity of 35,000 ppm water.

The effect of feed temperature on unit water cost is illustrated in Fig. 6. In the range of 18–30°C (65–86°F), the cost with feed at 30°C is about 19% lower than that obtained with low temperature feed of 18°C. The corresponding change per degree centigrade is in the order of 1.5%,

Fig. 6. Effect of feed temperature on capital and unit water costs.

compared to an effect of approximately 3% imposed by feed temperature on membrane flux, or module productivity. It should, however, be emphasized that temperature has also an adverse effect on membrane life, predicted to be considerably lower at high temperatures. The overall effect of feed temperature on the economics of the RO process is therefore still pending on more reliable data concerning the effect on membrane life.

8.3 Effect of System Pressure and Membrane Life

The effect of system pressure and membrane life on unit water costs are illustrated in Fig. 7. Obviously, for a given membrane life and a constant unit membrane cost, a higher system pressure which strongly affects membrane flux would result in a considerably lower unit water cost. Membrane module cost, however, would be higher and membrane life lower at higher system pressure. The optimum pressure therefore cannot be determined without applying reliable data of membrane cost and membrane life versus pressure.

8.4 Effect of Energy Cost and Comparison with Multistage Flash Distillation Plants

From the unit water cost breakdown shown in Table 8, it is evident that for an electrical power cost of \$0.020/kWhr, the energy cost in RO systems operating at pressure of 800 psig amounts to approximately 10-15% of the total unit water cost. This percentage is relatively low in comparison to other desalting processes and is even lower than in nuclear dual-purpose plants using current technology, high temperature distillation processes.²

An economic comparison was made between RO plants and small single- and dual-purpose MSF plants recently evaluated at ORNL² using identical economic ground rules. The comparison with single-purpose plants versus energy costs is shown in Fig. 8 and that with dual-purpose plants in Fig. 9. The combined figures of process heat cost applied in MSF, and electrical power cost applied in both processes were based on the corresponding evaluations made for single- and dual-purpose distillation plants.²

Fig. 7. Unit water cost vs system pressure and membrane life.

Fig. 8. Unit water costs from reverse osmosis and single-purpose multistage-flash plants vs energy costs.

ORNL-DWG 76-2956

Fig. 9. Unit water costs from reverse osmosis and dual-purpose multistage-flash plants vs energy costs.

From the results shown in Fig. 8, it is evident that current technology RO plants, based on the technological and cost assumption applied in the present study, might eventually be competitive in the 1-Mgd range with MSF single-purpose plants only at prevailing high energy cost. If, however, the projected RO technology could be realized, RO could also be competitive at lower energy price levels even in the 2.5-Mgd range. It must be emphasized however, that this comparison is made for plants operating with seawater having a salinity of 35,000 ppm. For higher salinities at which the distillation processes are much less affected, the competitiveness of the RO process would be reduced considerably.

Based on the results shown in Fig. 9, it is evident that RO seawater desalting plants cannot compete with dual-purpose MSF plants, even with the assumed projected technology and at high energy cost. In this case, the difference between the unit water costs obtained from both processes in the 1-Mgd range is quite small, especially when considering the accuracy of the results for both processes and the operational and cost uncertainties related to the RO process for seawater desalination.

8.5 Overall Sensitivity Analysis of Technological and Cost Parameters

The present evaluation was based on a set of technological and cost parameters which still have, at least in part, to be verified. As already discussed, the major unknown factors are membrane life, long-term performance, and the requirements for raw seawater pretreatment. However, several additional technological parameters, such as the largest applicable train size, the efficiencies of high pressure pumps and of the power recovery turbines and the cost of the seawater intake system, may also effect the economics of the RO process considerably. Therefore, a brief sensitivity analysis of these and several other factors was made in order to determine their effect on the unit water cost. As reference case for this investigation, a 12.5 Mgd plant based on current technology was selected. The effect of these parameters, each varied singularly while all others were kept at their reference design values, is summarized in Table 9. The combined effect of varying several parameters

Table 9. Effect of technological and cost parameters on the unit water cost from a 12.5-Mgd RO seawater desalting plant based on current technology

(Unit water cost from reference design = \$3.89/kgal)

Parameters varied	Cost increase		Cost savings	
	\$/kgal	%	\$/kgal	%
1. <u>Module and train size</u>				
a. Smaller: module-4 in.; train-0.25 Mgd	+0.34	+8.7		
b. Larger: module-12 in.; train-1.25 Mgd			-0.08	-2.1
2. <u>Unit membrane cost (first stage) - \$/gpd</u>				
a. Higher: initial-1.0; replacement to 0.80	+0.71	+18.3		
b. Lower: initial-0.50; replacement to 0.40			-0.73	-18.8
3. <u>Membrane life (first stage) - years</u>				
a. Lower - 1	+1.45	+37.3		
b. Higher - 3			-0.51	-13.1
c. Higher - 5			-0.91	-23.4
4. <u>Log flux decline /m/ (first stage)</u>				
a. Higher - 0.08	+0.41	+10.5		
b. Lower - 0.03			-0.24	-6.2
c. Lower - 0.02			-0.35	-10.0
5. <u>First stage pressure - psig</u>				
a. Lower - 750	+0.44	+11.3		
b. Higher - 900			-0.64	-16.5
c. Higher - 1000			-0.92	-23.7
6. <u>Feed temperature - °C</u>				
a. Lower - 18 (65°F)	+0.48	+12.3		
b. Higher - 30 (86°F)			-0.31	-8.0
7. <u>Overall efficiency of high pressure pumps, %</u>				
a. Lower - 68	+0.07	+1.8		
b. Higher - 84			-0.07	-1.8
8. <u>Overall efficiency of power recovery turbine, %</u>				
a. Lower - 60	+0.01	+0.3		
b. Higher - 75			-0.03	-0.8
9. <u>Capital cost of pretreatment system, %</u>				
a. Higher - +50	+0.10	+2.57		
b. Lower - -50			-0.10	-2.57
10. <u>Capital cost of seawater intake, %</u>				
a. Higher - +50	+0.05	+1.29		
b. Lower - -50			-0.05	-1.29
Combined effect of 3a, 9a and 9b	+1.62	+41.6		
11. Combined effect of all cost increasing factors	+6.40	+164.5		
12. Combined effect of 1b-4b and 7b to 10b			-1.12	-28.8
13. Combined effect of 1b to 10b			-2.08	-53.5
14. Combined effect of largest potential cost savings			-2.41	-62.0

simultaneously, representing overall sets of optimistic and pessimistic assumptions were also evaluated and are shown in Table 9.

The results of the sensitivity analysis emphasized again the very strong effect of membrane life on the unit water cost. However, the other factors characterizing the membrane, such as log flux decline and, of course, the unit membrane cost, will also have a major influence on the potential competitiveness of the RO process for seawater desalting. The cost of the pretreatment and the seawater systems, as well as the other "nonmembrane" factors, such as the efficiencies of the pumps and the power recovery turbine, have each alone only a minor effect.

The combination of several of these factors may however increase or decrease the unit water cost substantially.

REFERENCES

1. S. A. Reed, "The Impact on Increased Fuel Costs and Inflation on the Cost of Desalting Seawater and Brackish Waters," ORNL-TM-5070 Revised, January 1976.
2. P. Glueckstern, J. V. Wilson and S. A. Reed, "RO-75: A Fortran Code for Calculation and System Design Optimization of Reverse Osmosis Seawater Desalination Plants," ORNL-TM-5231, December 1976.
3. Ralph Stone and Co., Inc., "The Chula Vista, California, Pilot Plant Testing Program to Evaluate Seawater Pretreatment Methods for Reverse Osmosis," OSW Research and Development Progress Report No. 890, September 1973.
4. N. W. Rosenblatt, "Seawater Desalination by Reverse Osmosis," presented at the Special Symposium "Marine Chemistry in the Coastal Environment," 169th ACS National Meeting, April 8-10, 1975, Philadelphia, Pa.
5. D. G. Thomas, "Engineering Development of Hyperfiltration with Dynamic Membranes, IV, Economic Analysis Desalination," 15 (1974), pp. 343-369.
6. W. G. S. Fort, "Computer Codes for Office of Saline Water Program Analysis, Part V: MSF-21: A Fortran Code for the Rapid Calculation or Optimization of Multistage Flash (MSF) Desalination Plants," ORNL-TM-3535, February 1972.
7. W. G. S. Fort, J. V. Wilson, Ibid., Part VI: VTE-21: A Fortran Code for the Rapid Calculation or Optimization of Vertical Tube Evaporator (VTE) Desalination Plants," ORNL-TM-3535 (Part VI).
8. P. Glueckstern, J. V. Wilson and S. A. Reed, "Design and Cost Modifications of ORNL's Computer Programs MSF-21 and VTE-21 for Evaluation of Multistage Flash and Vertical Tube Evaporators," ORNL-TM-5230, December 1975.
9. K. E. Hickman et al., "Pumping and Energy Recovery Systems for Reverse Osmosis Desalination Plants," OSW Research and Development Report No. 457, August 1969.
10. Fluor Corporation, "Engineering and Economic Study of Reverse Osmosis Brackish Water Desalting Plants," Final Report to Office of Saline Water, October 1972, published in Membrane Digest, 1(4), Spring 1973.
11. P. Glueckstern and S. A. Reed, "A Comparative Investigation of the Economics of Seawater Desalting Based on Current and Advanced Distillation Concepts," ORNL-TM-5232, December 1975.

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Appendix A

COMPUTER PRINTOUT OF A 2.5-Mgd RO PLANT
BASED ON CURRENT TECHNOLOGY

REVERSE OSMOSIS SEAWATER DESALTING PLANT - ROUTINE RC75 PAGE 1

PLANT CAPACITY - MGD 2.500

UNIT COST OF DESALTED WATER - CENTS PER KILGALLON 460.518

CAPITAL COST OF PLANT(DIRECT+INDIRECT)- MILLIONS OF DOLLARS 9.431

POWER RECOVERY -YES

THE PLANT HAS BEEN OPTIMIZED

COST SUMMARY PAGE

CAPITAL COMPONENTS	COST-\$/YR	COST-C/KGAL	COST-\$DIRECT
MODULES AND FOLDS	509152.95	65.6442	3490350.00
PUMPS AND MOTORS	79319.03	10.2265	488441.53
VALVES AND PIPING	73821.47	9.5177	454587.88
INSTRUMENTS	18335.17	2.3639	112906.79
ELECTRICAL	45771.27	5.9012	281856.54
PRETREATMENT SYSTEM	155129.90	20.0006	955280.01
POSTTREATMENT SYSTEM	7273.10	0.9377	44787.27
SEA-WATER INTAKE	72004.61	9.2834	443399.77
SITE, BLDGS, CRANES	46763.10	6.0291	287964.17
POWER RECCV. TURBINE	29547.24	3.8095	181949.99
TOTAL CAPITAL	1037117.83	133.7138	6741523.96
TOTAL INVESTMENT, \$DIR+IND			9431229.91
SPECIFIC INVESTMENT, \$/GPD			3.77249
MEMBRANE REPLACEMENT	1194132.14	153.9574	
OPERATING COMPONENTS			
POWER	811194.66	104.5859	
CHEMICALS	284758.03	36.7134	
OPERATING	178668.69	23.0355	
MAINT. + SUPPLIES	66018.60	8.5117	
TOTAL OPERATING	1340639.98	172.8464	
TOTAL (CAP+RETUB+OP)	3571889.95	460.5177	

CCST FACTORS

COST OF POWER, C/KWH	5.000000
ANNUAL CHARGE RATE	0.116079
INTEREST RATE	0.100000
HIGHER CCST FACTOR	1.398576
PLANT LOAD FACTOR	0.850000
PLANT LIFE, YEARS	30.000000
PUMP AND MOTOR EFFICIENCY	
HIGH PRESSURE PUMP - 1ST STAGE	0.760
HIGH PRESSURE PUMP - 2ND STAGE	0.760
OTHER PUMPS	0.820
POWER RECOVERY TURBINE	0.650
MEMBRANE LIFETIME, YEARS	
FIRST STAGE	2.00
SECOND STAGE	3.00
LOG FLUX DECLINE RATE	
FIRST STAGE M FACTOR	-0.0500
SECOND STAGE M FACTOR	-0.0500
MEMBRANE REPLACEMENT CCST, \$/GPD	
FIRST STAGE	0.55
SECOND STAGE	0.15

I SUMMARY OF PLANT DATA

GENERAL

PLANT CAPACITY, MGD	2.50
SEA WATER CONCENTRATION, PPM	35000.00
PRODUCT CONCENTRATION, PPM	500.00
PRODUCT RECOVERY	0.32
NO OF STAGES	2.00
YEAR CONSTRUCTION STARTED	1976.00
OCEAN TEMPERATURE, F	77.00

II DESCRIPTION OF PHYSICAL PLANT

	TOTAL	STAGE 1	STAGE 2
TOTAL PLANT AREA, SQ FT	12500.		
NO OF TRAINS	3.	2.	1.
NO OF MODULES	949	877	72

III MEMBRANE PARAMETERS

MEMBRANE TYPE	POLYAMID - HOLLOW FIBER	
MODULE OUTSIDE DIAM., IN.	8.00	8.00
SHELL DIMENSIONS		
OUTSIDE DIAM., IN.	9.50	9.50
LENGTH, IN.	48.00	48.00
NO. OF MODULES IN SHELL	1	1
MEMBRANE SURFACE, SQ.FT.	5000.00	6000.00
NOMINAL OPERATING PRESSURE, PSI	800.0	400.0
NOMINAL PRODUCT RECOVERY	0.3000	0.7500
NOMINAL REJECTION RATE	0.9850	0.9000
INITIAL PRODUCTIVITY, GPD/MODULE	5000.00	14000.00
REFERENCE FEED CONCENTRATION, PPM	30000.00	15000.00
REFERENCE FEED TEMPERATURE, F	77.00	77.00
MEMBRANE LIFE, YEARS,	2.00	3.00

IV ACTUAL OPERATING PARAMETERS

MODULE INLET PRESSURE, PSI	800.00	400.00
PRODUCT RECOVERY	0.3225	0.8993
SALT PASSAGE	0.0192	0.2020
FIRST STAGE FEED CONCENTRATION, PPM	34679.	
MODULE PRODUCTIVITY, GPC		
INITIAL	3897.17	14379.37
END OF LIFETIME	2802.74	10133.73
WEIGHTED AVERAGE	2950.26	10667.08
TOTAL STAGE PRODUCTION, MGD	2.5869	0.7764
STAGE PRODUCT QUALITY, PPM	664.75	134.30
MODULE REJECTION RATE	0.9808	0.7980
BLENDING RATIO, 2ND STAGE/TOTAL		0.3106

V PUMPING PARAMETERS

	FLOW-GPM	HEAD-FT	POWER-KW
1ST STAGE HIGH PRESSURE	5570.2	1840.0	2551.1
2ND STAGE HIGH PRESSURE	599.5	920.0	123.5
SEA-WATER DELIVERY	5509.8	138.0	175.4
PRODUCT	1736.0	92.0	36.8
TOTAL			2924.1
POWER RECOVERY TURBINE	3773.8	1716.9	745.3
NUMBER OF PLANT CALLS 101	0		

Appendix B

COMPUTER PRINTOUT OF A 2.5-Mgd RO PLANT BASED ON PROJECTED TECHNOLOGY

REVERSE OSMOSIS SEAWATER DESALTING PLANT - ROUTINE R075

PLANT CAPACITY - MGD	2.500
UNIT COST OF DESALTED WATER - CENTS PER KILOGALLON	326.385
CAPITAL COST OF PLANT(DIRECT+INDIRECT)- MILLIONS OF DOLLARS	7.261
POWER RECOVERY -YES	

THE PLANT HAS BEEN OPTIMIZED

CLST SUMMARY PAGE

CAPITAL COMPONENTS -----	COST-\$/YR -----	COST-C/KGAL -----	COST-\$DIRECT -----
MODULES AND HOUSING	342284.51	44.1302	2328277.50
PUMPS AND MOTORS	74725.63	9.6342	460521.19
VALVES AND PIPING	47756.64	6.1572	294315.99
INSTRUMENTS	10522.51	1.3566	64848.44
ELECTRICAL	41620.32	5.3660	256498.88
PRETREATMENT SYSTEM	143201.72	18.4628	882527.58
POSTTREATMENT SYSTEM	7267.32	0.9370	44787.27
SEA-WATER INTAKE	65522.85	8.4477	403806.06
SITE, BLDGS, CRANES	46725.98	6.0243	287964.17
POWER RECOV. TURBINE	27677.62	3.5684	170572.44
TOTAL CAPITAL	807305.11	104.0845	5194119.52
TOTAL INVESTMENT, \$CIR+INC			7260680.05
SPECIFIC INVESTMENT, \$/GPD			2.90427
MEMBRANE REPLACEMENT	477288.94	61.5361	
OPERATING COMPONENTS -----			
POWER	763246.87	98.4041	
CHEMICALS	254192.32	32.7726	
OPERATING	178668.69	23.0355	
MAINT. + SUPPLIES	50824.76	6.5527	
TOTAL OPERATING	1246932.63	160.7649	
TOTAL (CAP+RETLB+OP)	2531526.68	326.3854	

COST FACTORS

COST OF POWER, C/KWH	5.000000
ANNUAL CHARGE RATE	0.116079
INTEREST RATE	0.100000
HIGHER COST FACTOR	1.397865
PLANT LOAD FACTOR	0.850000
PLANT LIFE, YEARS	30.000000
PUMP AND MOTOR EFFICIENCY	
HIGH PRESSURE PUMP - 1ST STAGE	0.760
HIGH PRESSURE PUMP - 2ND STAGE	0.760
OTHER PUMPS	0.820
POWER RECOVERY TURBINE	0.650
MEMBRANE LIFETIME, YEARS	
FIRST STAGE	3.00
SECOND STAGE	3.00
LOG FLUX DECLINE RATE	
FIRST STAGE M FACTOR	-0.0300
SECOND STAGE M FACTOR	-0.0300
MEMBRANE REPLACEMENT COST, \$/GPC	
FIRST STAGE	0.34
SECOND STAGE	0.10

I SUMMARY OF PLANT DATA

GENERAL

PLANT CAPACITY, MGD	2.50
SEA WATER CONCENTRATION, PPM	35000.00
PRODUCT CONCENTRATION, PPM	500.00
PRODUCT RECOVERY	0.35
NO OF STAGES	2.00
YEAR CONSTRUCTION STARTED	1976.00
OCEAN TEMPERATURE, F	77.00

II DESCRIPTION OF PHYSICAL PLANT

	TOTAL	STAGE 1	STAGE 2
TOTAL PLANT AREA, SQ FT	12500.		
NO OF TRAINS	2.	1.	1.
NO OF MODULES	472	431	41

III MEMBRANE PARAMETERS

MEMBRANE TYPE	POLYAMIC - HOLLOW FIBER	
MODULE OUTSIDE DIAM., IN.	12.00	12.00
SHELL DIMENSIONS		
OUTSIDE DIAM., IN.	13.50	13.50
LENGTH, IN.	48.00	48.00
NO. OF MODULES IN SHELL	1	1
MEMBRANE SURFACE, SQ. FT.	5000.00	6000.00
NOMINAL OPERATING PRESSURE, PSI	800.0	400.0
NOMINAL PRODUCT RECOVERY	0.3000	0.7500
NOMINAL REJECTION RATE	0.9850	0.9000
INITIAL PRODUCTIVITY, GPD/MODULE	10000.00	28000.00
REFERENCE FEED CONCENTRATION, PPM	30000.00	1500.00
REFERENCE FEED TEMPERATURE, F	77.00	77.00
MEMBRANE LIFE, YEARS	3.00	3.00

IV ACTUAL OPERATING PARAMETERS

MCCULE INLET PRESSURE, PSI	800.00	400.00
PRCCCT RECCVERY	0.2625	0.9000
SALT PASSAGE	0.0212	0.2037
FIRST STAGE FELD CONCENTRATION, PPM	34556.	
MCCULE PRCCCTIVITY, GPE		
INITIAL	7237.80	28577.51
END OF LIFETIME	5867.11	23165.52
WEIGHTED AVERAGE	6048.57	23881.98
TOTAL STAGE PRODUCTION, MGD	2.6102	0.9918
STAGE PRODUCT QUALITY, PPM	730.90	148.88
MCCULE REJECTION RATE	0.9788	0.7963
BLENDING RATIC, 2ND STAGE/TOTAL		0.3967

V PUMPING PARAMETERS

	FLOW-GPM	HEAD-FT	POWER-KW
1ST STAGE HIGH PRESSURE	5000.1	1840.0	2290.0
2ND STAGE HIGH PRESSURE	765.2	920.0	157.7
SEA-WATER DELIVERY	4923.5	138.0	156.7
PRCCUCT	1736.0	92.0	36.8
TOTAL			2678.6
POWER RECOVERY TURBINE	3187.5	1722.1	628.5
NUMBER OF PLANT CALLS 101	C		

Appendix C

CONVERSION FACTORS

<u>Multiply</u>	<u>By</u>	<u>To obtain</u>
Million of gallons per day (Mgd)	3.785×10^3	Cubic meters per day (m^3/d)
Dollars per gallon per day (\$/gpd)	264.2	Dollars per cubic meter per day ($\$/\text{m}^3 \cdot \text{d}$)
Pounds per square inch (psi)	703.1	Killograms per square meter (kg/m^2)
Inches (in.)	2.54	Centimeters (cm)
Square feet (ft^2)	0.0929	Square meters (m^2)
Gallons per minute per square foot (gpm/ft^2)	0.7451	Liters per second per square meter ($\text{liter}/\text{sec} \cdot \text{m}^2$)

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