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ENGINEERING STUDY OF DESALINATION USING SALT GRADIENT SOLAR PONDS

FINAL REPORT

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CONTENTS

I. SUMMARY

In this engineering study, technical and economic analyses have been made to investigate suitability of utilizing solar energy through salt gradient solar ponds for the purpose of desalination of seawater. A 550 m^3/day plant has been considered.

Suitably sized (in surface area and depth) solar ponds with hot brine bottom layer providing all or part of the energy required for desalination have been considered. Detailed considerations have been given to multistage flash distillation plants using hot pond brine for seawater heating, as well as reverse osmosis desalination plants using low temperature rankine-cycle power.

Specifically, the following candidate cases were studied to determine relative merits:

- Solar pond providing heating for MSF plant, purchased pumping power.

- Solar pond providing heating for MSF plant, pumping power provided by low-temperature rankine-cycle turbo-generator coupled with solar ponds.

- Solar pond providing energy for a low temperature rankine-cycle power generator driving high pressure pumps of an R.O. plant, balance purchased power.

- Solar pond providing energy for a low temperature rankine-cycle power generator providing full power for an R.O. plant.

MSF plants were considered for performance ratios of 6 and 8.

For convenience of comparison, dollars per m^3 of product water for all cases were determined. R.O. systems considered showed a cost range of $3.29 to $3.49, while the MSF systems considered showed a cost range of $4.15 to $4.74. It may be noted that MSF plant produces distilled water, while R.O. system as considered is adequate for potable water production. Cost per million Btu of solar pond heat energy ranges between $3.80 and $4.00 approximately.

The most promising system was found to be R.O. desalination system, fully or partly powered by solar pond. These two cases were considered for evaluating rate of return on investment. The internal rate of return* (IRR) values were determined to be 5.8% and 6.2% for these cases. For comparison, IRR values were determined for diesel-powered and wind-power supplemented R.O. plants. These latter cases showed returns of 14.0% and 12.2%.

In conclusion, this study showed that although technically suitable, present economic conditions and considerations make solar pond desalination less attractive to diesel-powered or wind-power supplemented systems. It is also recognized that due to the absence of long-term operating experience of solar ponds on a commercial scale, life-cycle cost projections for ponds are only approximate.

However, as shown in Section VII, Chapter G, a possibility exists for reducing costs for solar pond application, thereby increasing its commercial attractiveness. On the other hand, ever-increasing costs of imported fuel could render the conventional diesel power less attractive. These changes in the cost picture are time-dependent. It is, therefore, concluded that

*IRR calculations are based on constant purchasing power for the dollar. Real IRR may be approximately determined by simply adding the estimated inflation rate to the IRR values.

the economic merits of a solar pond project specifically for seawater desalination should be further reviewed at a later date, probably within a period of three years.

OBSERVATIONS

The economic attractiveness of a solar pond application for seawater desalination can be increased by the influence of the following significant cost components:

- Reduction of pond area - this study is conservative as regards assumed pond efficiency. Higher efficiency would yield less surface area requirement.
- Utilizing concentrate from a salt evaporation pond - this will reduce initial salt costs.
- Faster increase in fuel oil costs.
- Use of a natural depression on ground as pond location.
- Reduction in capital cost of the Rankine cycle power apparatus.

SECTION II - INTRODUCTION

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II. INTRODUCTION

(A portion of this section is based on AID work order document.)

A. Background

Fresh water is scarce in the Republic of Cape Verde. Many of the islands in the archipelago do not receive any precipitation for many years. Moreover, groundwater has become brackish to the point that it is becoming unsuitable for human consumption. To alleviate the scarcity of water, in fact, the natural sources are being supplemented by seawater desalination. The cost of desalination of seawater is very high. It varies between $7 and $11 per cubic meter (depending on the island). Most of the present installation use fuel oil and diesel oil and, consequently, the already high costs will probably increase as petroleum prices rise. It is, therefore, appropriate to look for energy sources for desalination which are less subject to price escalation and do not have to be imported. Wind and solar pond energies are two that are particularly attractive for investigating for possible application in Cape Verde.

A past study titled "Cape Verde Sal Island Desalination and Power Project", prepared in August, 1979, concluded that, for electrically powered desalination, the reverse osmosis (R)) process is the best among various processes, because of lowest energy costs. Subsequently, the wind potential on Sal Island was also investigated and a study "Wind Power Generation for Sal Island" was prepared in August, 1980. It was concluded in the latter study that when used for generating power for fuel saving purposes, wind electrical energy is competitive with diesel electric energy. The report considerd utilizing a 300 KW generator and with 1% differential inflation for diesel oil a 6.5% rate of return on investment was obtained. If one actually considered the same fuel rate escalation as averaged during the years 1970 to 1980, namely, about 17&; one would obtain a rate of return on investment approximating 30%. Thus, it may be assumed that wind energy could replace diesel energy (to an extent practical for design) in an RO plant.

Another local source of energy that could do the job is solar heat collected by salt gradient solar ponds. Cape Verde is located in a relatively good insolation area. The solar pond could collect, store and furnish heat in the temperature range 90°C to 100°C, directly to a flash evaporation desalination plant, or it could furnish electricity to drive an RO plant. A separate study* had been cited to show that estimated figures indicated heat and electricity could be furnished at about $1.29 per million Btu and 5.3 to 8.1 cents per Kwh, respectively.

This is competitive with petroleum in both cases. (Diesel fuel sells for 16 escudos/liter, or $460/ton, in Cape Verde. It produces heat at $14.40 per million Btu when an 80% boiler efficiency is used. Electricity sells for 16.5 cents (average) per Kwh on Sal, and for much higher in Praia. It is generated by diesel engines. The fuel cost predominates, being about 12 to 14 cents/kwh. Since low-temperature pond heat is theoretically less valuable than high-temperature petroleum heat (if petroleum heat is not used for low-temperature applications), one must devalue pond heat somewhat with respect to petroleum heat. However, in both cases, heat or electricity, pond energy is expected to be competitive with petroleum energy at present Cape Verdian prices. If, during the next decade, petroleum prices will rise faster than pond prices and pond energy will then be significantly cheaper

* (Tabor, H., Solar Ponds. Review Article Solar Energy, vol. 27, No. 3, pp. 181-194, 1981). Note: results of this study report, however, indicate a substantially higher figure as related to Sal Island conditions.

than petroleum energy in both heat and electrical forms. Cheap land and locations where salt leaks cause no environmental damage are available. Maintenance of a solar pond is less difficult than maintenance of diesel power equipment. The social impact should be the same as that for the present desalination plants.

Based on the above considerations, the current study has been performed to determine the technical and economical feasibility of pond desalination under the conditions prevailing on Sal Island. It may also be noted, as one negative aspect, that there is minimal operating experience with solar ponds, largely limited to Israel.

B. Objective

To determine the technical and economic feasibility of solar ponds for desalination and compare them to diesel and wind energy for desalination in Cape Verde at a reasonable location such as Sal Island.

C. Statement of Work

Appropriate portions of this study have been referenced on the righthand side of the work-statement.

The following tasks are performed:

a. Determine type and configuration of desalination equipment most appropriate to solar pond use. For example, this could be flash evaporation equipment using heat at $80^{\circ}C$ - $100^{\circ}C$ from the pond or RO using electricity generated from the pond. It could be a technique other than these. For example, it could be mixed system	Sec. IV and Sec. V

using diesel or wind to drive pumps, and ponds to supply heat in a flash-evaporation installation. However, solar pond heat should be the predominate energy input.

b. Make technical analysis of the two most promising systems: analyze particular advantages (such as long-term energy storage) and disadvantages of ponds in comparison to diesel and wind in the Cape Verdian environment. — Sec. VI

c. Provide sketches of the two most promising systems with representative numerical parameters, dimensions, temperatures and other pertinent data. These will be included in the final report. — Fig. IV-3; Fig. V-1 & Fig. V-2

d. Make an economic and financial analysis of the most promising system to be powered by a solar pond. This would include a calculation of the net present value and internal rate of return over the life of the system using reasonable price escalation and discount rates for all inputs and shadow pricing of the possible uses of the fresh water produced. The results of the economic and financial analysis for the solar pond powered system shall be compared with the results of an economic and financial analysis for diesel and wind powered — Sec. VII systems. A reasonable price escalation rate for diesel fuel shall be used. The analysis shall also comment on the feasibility of collecting a use charge for the water used.

D. Criteria

The desalination plant capacity is considered as 550 m^3/day installed, suitable for a township of about 6000 people with commercial activities. A utilization factor of 90 percent is considered. This would mean 50 liters per capita per day (lpdc) consumption, with an allowance of 32.5 lpdc for commercial establishments.

The chosen plant location is in the vicinity of Palmeiras, Island of Sal (see Figure II-2). Sal is one of the windward islands of the Cape Verde Archipelago (see Figure II-1). Salt is available locally in the towns of Pedra Lume and Santa Maria.

Sal Island is approximately located 17°N 23°W. Due to the absence of reliable data on insolation on Sal Island, insolation values for Dakar, Senegal have been used as close approximation. Dakar is located at approximately 15°N.

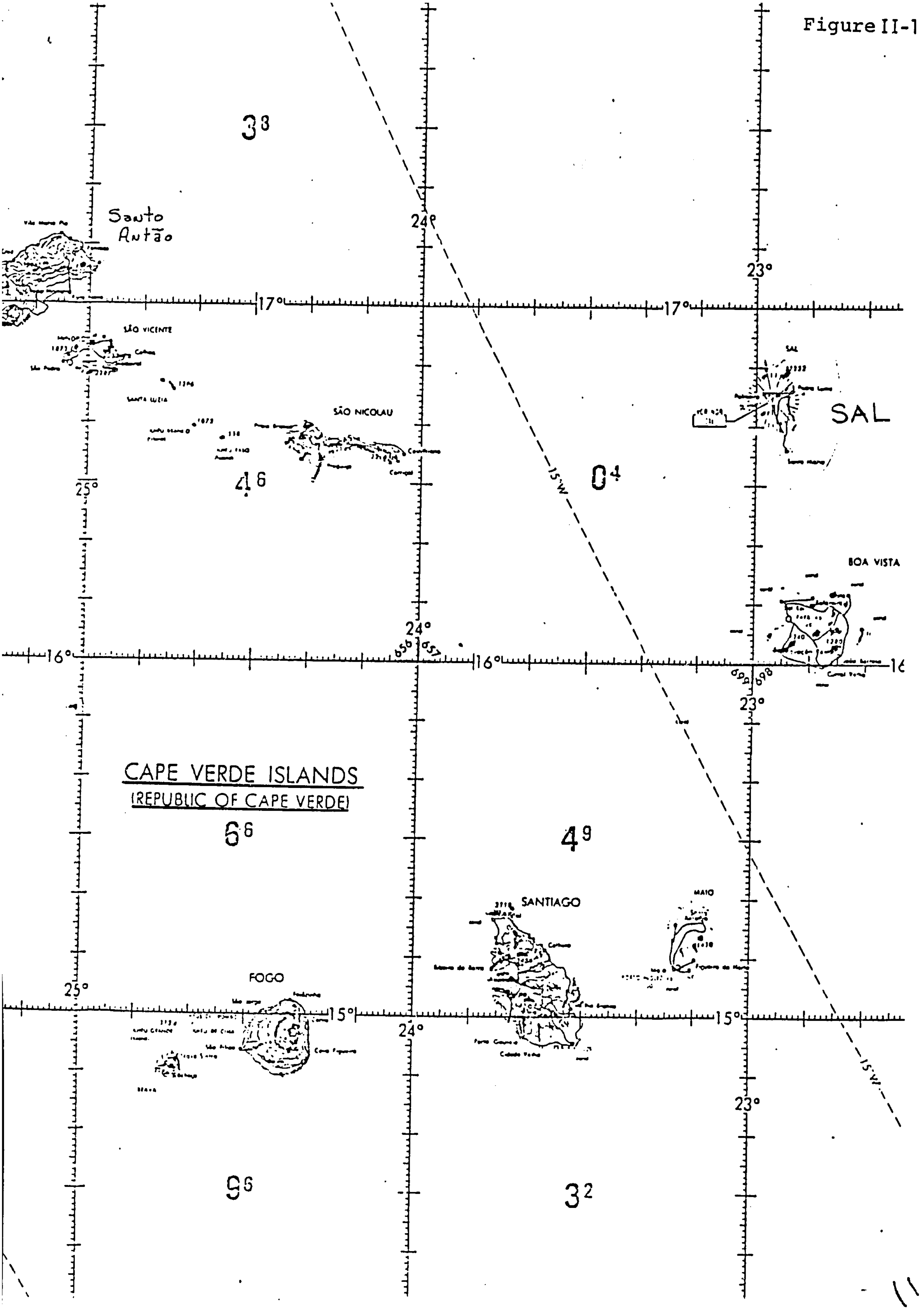
Salt gradient solar collecting pond, providing and storing hot brine at the bottom layer of the pond, has been used as the principal energy source.

All available and applicable data on Sal Island has been considered.

E. Economic Comparison

An economic comparison has been made in Section VII by evaluating the internal rate of return (IRR) for the cases considered for final analysis. For convenience of comparison in all other sections, unit production cost has been compared. Capital costs have been translated to annual fixed charges at a rate of 10 percent and operation and maintenance costs have been added thereto. It may be noted that 10% FCR is the amortization rate for a 20-year life project at a discount rate of slightly lower than 8 percent p.a.

Figure II-1


Santo Antão
SÃO VICENTE
SANTA LUZIA
SÃO NICOLAU
SAL
BOA VISTA
CAPE VERDE ISLANDS
(REPUBLIC OF CAPE VERDE)
SANTIAGO
MAIO
FOGO
15°W
17°
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3⁸
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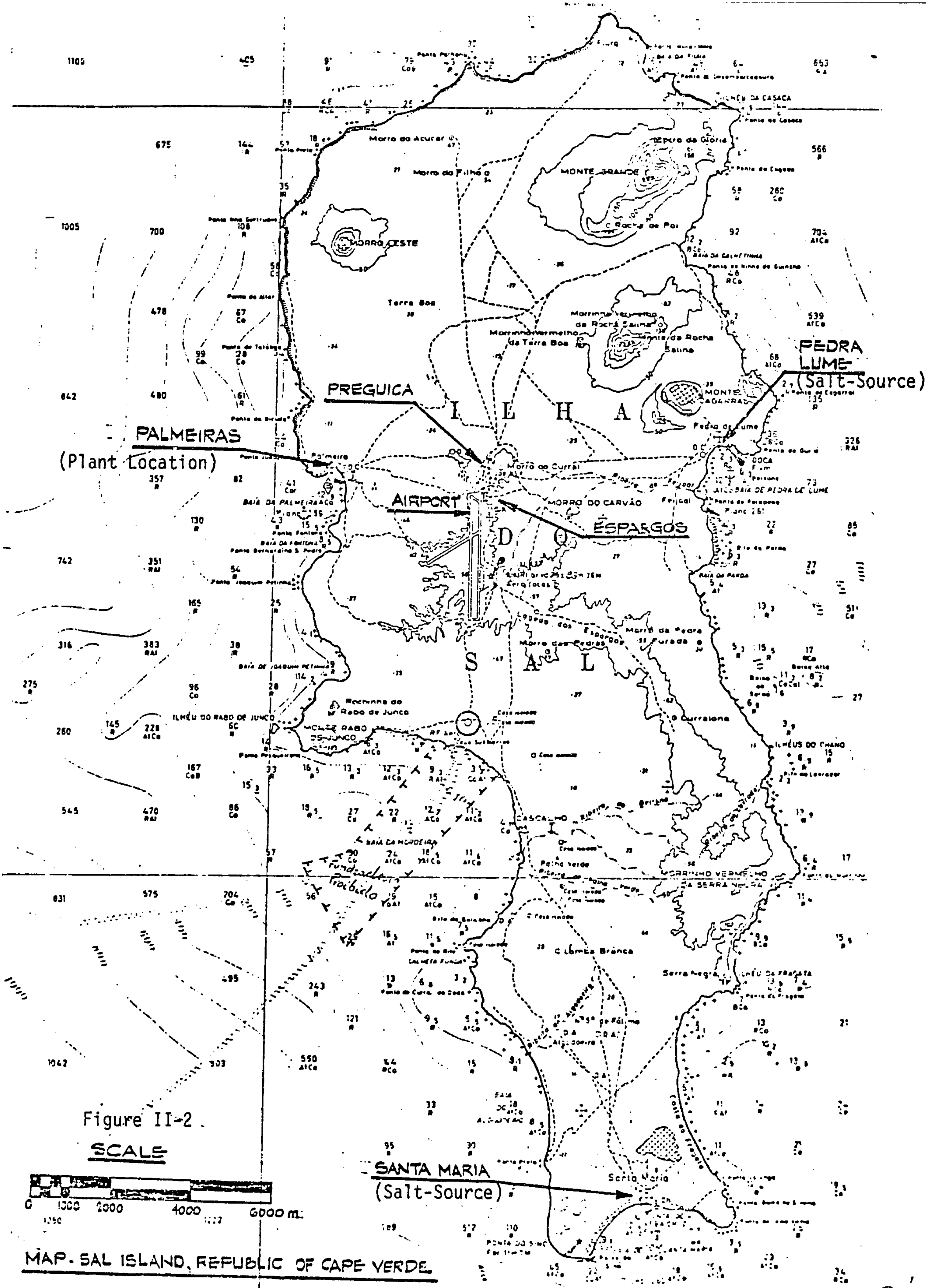

Figure II-2

MAP - SAL ISLAND, REPUBLIC OF CAPE VERDE

SECTION III

SALT GRADIENT POND

Prepared by: Albino Ko

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SECTION III - SALT GRADIENT POND

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III. SALT GRADIENT POND

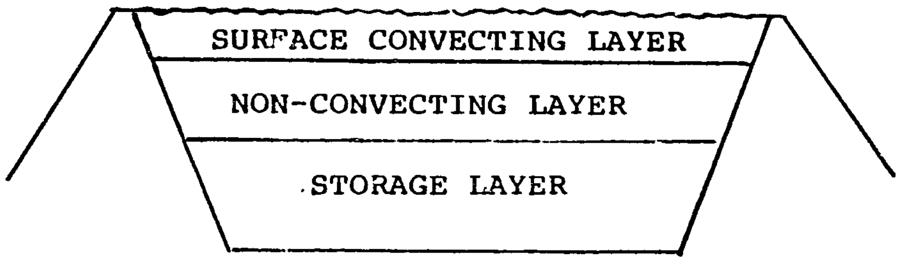
A. Introduction

The solar pond collects heat in the form of insolation absorbed into the pond body and heats the lower layers of water. This heat, due to the higher water density, reamins at the bottom layer of the pond.

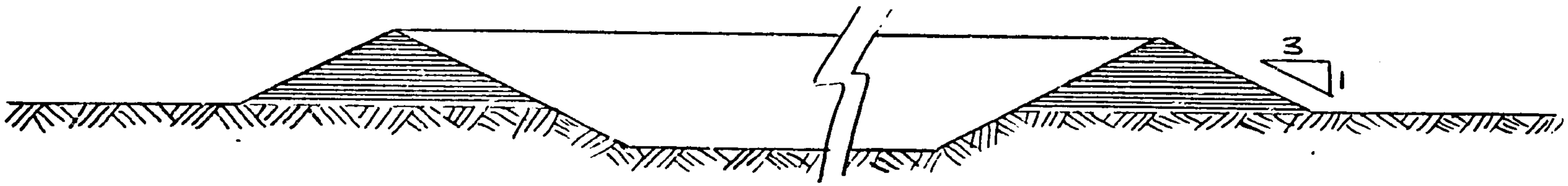
The pond consists of three layers: the surface layer - a thin layer of nearly saltless water in which there are vertical convection currents due to wind and evaporation; the nonconnecting layer - a layer in which a salt concentration gradient (positive downward) prevents vertical convection; and the lower convecting layer - a storage layer in which the salt concentration is constant. The nonconvecting layer serves to insulate the storage layer, preventing most of the heat loss. The convection of the surface layer is unavoidable, due to the effects of wind and evaporation. The bottom of the pond is usually lined with a blackened plastic film to prevent leakage and to absorb the insolation that reaches the bottom. Heat removal to supply the heat load in a desalting plant takes place in the storage layer by running the concentrated salt water through a heat exchanger. In order to maintain the density or salt gradient of the three layers, a concentrated solution is injected into the bottom of the pond, and the top of the pond is washed with fresh water from time to time. Figure III-1 shows the cross section of a salt gradient pond.

B. Design

The sizing of salt gradient pond to supply heat to a multistage flash desalting plant at performance ratio between 6 and 8 are based on the following:



a) Diagram showing water layers



b) Schematic cross section

FIGURE III-1 SALT GRADIENT POND CROSS SECTION

Average insolation, 246.3 w/m^2 (using Dakar data with max. insolation at 300 w/m^2 and min. insolation at 200 w/m^2) See Table III-1.

Ambient temperature, 20°C (using average wet bulk temperature)

Latitude, 17°-18°

Design Pond Temperature, 90°C (194°F)

Surface convecting layer thickness, 0.3 meter (m)

Nonconvecting layer thickness, 1.2 meters

Average optical transmission through top 2 layers, 0.31

Heat loss from pond surface through non-convecting layer - 0.4 w/m^2 °C

Edge losses 2.2 w/°C per meter of perimeter

Losses from pond bottom to ground, 0.1 w/m^2 °C

In the multi-stage flash plant, performance ratio is defined as pounds of product per 1000 Btu's. The higher the performance ratio, the more numbers of stages are required in the flash plant. Since more heat can be recovered in a multi-stage flash plant design at higher performance ratio, the lesser heat would be supplied from the salt gradient pond.

In a 145,310 gallons per day (550 m^3/day) multi-stage flash unit, the heat required in a multi-stage flash plant is 8,409,167 Btu's/hr at performance ratio of 6 and 6,306,875 Btu's/hr at performance ratio of 8.

	Total Sunshine Hours	Insolation w/m^2
January	220.8	219
February	232.1	253
March	196.3	291
April	212.7	292
May	185.0	293
June	172.8	272
July	173.9	237
August	181 8	223
September	185.3	223
October	180.8	235
November	222.0	217
December	218.6	200
		Avg. 246.3

TABLE III-1 MONTHLY TOTAL HOURS OF SUNSHINE AND INSOLATION IN DAKAR

Based on the above parameters, the bottom surface area of the salt gradient pond is sized at 65,361 m^2 (16.14 acres) and 49,441 m^2 (12.2 acres) at performance ratio 6 and 8, respectively. The salt gradient pond is designed in trapezoid configuration, with a side slope of 3:1 to eliminate concrete on the side wall. This type of salt gradient pond is essentially the same design as in any evaporation pond in the United States. Table III-2 shows the dimension of the pond at various pond depths (3.0 m to 6.5 m) between performance ratio 6 and 8. Note that the top surface area increases as pond depth increases.

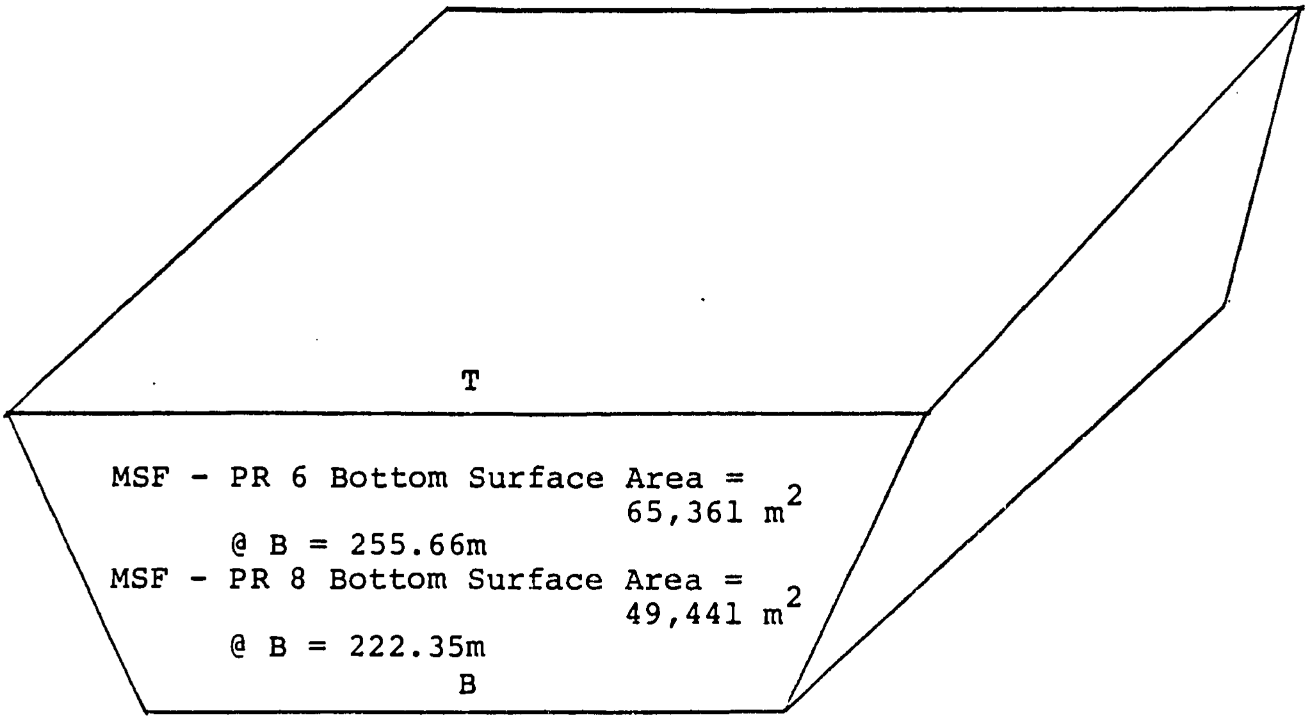
During the month of December, the insolation is at its lowest of 200 w/m^2, the fluctuation of insolation between summer and winter will affect the pond temperature. However, the greater the bottom layer depth of the pond, the greater its thermal mass and; therefore, the smaller its seasonal temperature fluctuations.

Table III-3 shows various pond depth at specified minimum pond temperatures. The designed pond maximum temperature is 90^oC at an average insolation of 246.3 w/m^2. However, during the winter season when the insolation is at 200 w/m^2, the pond temperature would be 76.8^oC at 3.0 m pond depth and 85.8^oC at 6.5 m depth.

Tables III-4 and III-5 show the salt gradient pond volume of various layers of the pond and total liner areas of the pond to supply heat to the multi-stage flash plant at performance ratio 6 and 8. The liner to be used will be reinforced Hypalon liner with polyester thread and 36 mils thick.

C. Salt Requirement

In order to have a high density of salt water at the bottom layer in the pond, it is essential to have a concentration of salt water as high as 30 percent. Thus, a salt with high solubility limit in water is required.



Pond Depth m	Pond Slope m	MSF - PR 6 Top Layer T m	MSF - PR 6 Top Surface Area m^2	MSF - PR 8 Top Layer T m	MSF - PR 8 Top Surface Area m^2
3.0	9.487	273.66	74,890	240.35	57,768
3.5	11.068	276.66	76,541	243.35	59,219
4.0	12.65	279.66	78,210	246.35	60,688
4.5	14.23	282.66	79,897	249.35	62,175
5.0	15.81	285.66	81,602	252.35	63,681
5.5	17.29	288.66	83,325	255.35	65,204
6.0	18.97	291.66	85,066	258.35	66,745
6.5	20.55	294.66	86,825	261.35	68,304

TABLE III-2 SALT GRADIENT POND DIMENSIONS AT VARIOUS POND DEPTH

Bottom Layer m	Total Pond Layer m	Min. Pond Temp. @246.3 w/m^2 Avg. 200 w/m^2 Min.
1.5	3.0	76.8°C
2.0	3.5	79.8°C
2.5	4.0	81.7°C
3.0	4.5	83.0°C
3.5	5.0	84.0°C
4.0	5.5	84.7°C
4.5	6.0	85.3°C
5.0	6.5	85.8°C
10.0	11.5	87.9°C

TABLE III-3 MINIMUM POND TEMPERATURE

Bottom Layer m	Mid Layer m	Top Layer m	Total Pond Depth m	m^3 Bottom Layer Volume	m^3 Mid Layer Volume	m^3 Top Layer Volume	m^3 Total Pond Volume	m^3 Total Liner Area
1.5	1.2	0.3	3.0	101,555	86,459	22,364	210,378	75,404
2.0	1.2	0.3	3.5	137,004	88,452	22,874	248,330	77,144
2.5	1.2	0.3	4.0	173,274	90,477	23,393	287,144	78,905
3.0	1.2	0.3	4.5	210,378	92,535	23,919	326,832	80,682
3.5	1.2	0.3	5.0	248,330	94,626	24,453	367,409	82,478
4.0	1.2	0.3	5.5	287,144	96,748	24,996	408,888	84,292
4.5	1.2	0.3	6.0	326,832	98,903	25,548	451,283	86,126
5.0	1.2	0.3	6.5	367,409	101,091	26,106	494,606	87,979

TABLE III-4 SOLAR POND VOLUMES AT VARIOUS LAYERS AND TOTAL LINER AREA AT PR6

Bottom Layer m	Mid Layer m	Top Layer m	Total Pond Depth m	m^3 Bottom Layer Volume	m^3 Mid Layer Volume	m^3 Top Layer Volume	m^3 Total Pond Volume	m^3 Total Liner Area
1.5	1.2	0.3	3.0	77,222	66,345	17,244	160,811	58,220
2.0	1.2	0.3	3.5	104,359	68,099	17,695	190,153	59,750
2.5	1.2	0.3	4.0	132,218	69,884	18,154	220,256	61,299
3.0	1.2	0.3	4.5	160,811	71,702	18,621	251,134	62,866
3.5	1.2	0.3	5.0	190,153	73,552	19,095	282,800	64,451
4.0	1.2	0.3	5.5	220,256	75,435	19,578	315,269	66,055
4.5	1.2	0.3	6.0	251,134	77,350	20,069	348,553	67,679
5.0	1.2	0.3	6.5	282,800	79,298	20,568	382,666	69,321

TABLE III-5 SOLAR POND VOLUMES AT VARIOUS LAYERS AND TOTAL LINER AREA AT PR8

The salt to be considered is based upon the solubility limit and cost. The salt that is produced in the Cape Verde Island is sodium chloride at about 95% to 98% purity and a solubility limit is about 39 grams per 100 cc of pure water. The cost of this salt is $20.15 per metric ton (or 0.9¢ per lb). The composition of this salt is as follows:

NaCl (Humid)	-	96.55%
(Dry)	-	97.94%
CaS_4	-	0.99%
$MgSO_4$	-	0.30%
$MgCl_2$	-	0.72%

Table III-6 shows the initial requirement of salt to maintain a 30 percent concentration at various bottom layer depths. At 1.5m bottom layer depth, the salt requirement is 82,433,052 lb. and 62,516,330 lb. in the salt gradient pond to supply heat to the multi-stage flash plant at performance ratios of 6 and 8, respectively.

In order to maintain a concentration gradient due to the effects of diffusion of salt from the bottom layer toward the surface, additional amounts of salt are required to inject into the bottom layer. Table III-7 shows the additional salt per day in the pond for multi-stage flash plant at performance ratio 6 and 8.

D. Pond Selection

The salt gradient pond is designed to supply heat to the multi-stage flash brine heater. The maximum designed temperature of the pond is 90°C. However, the pond temperature would change due to the fluctuation of seasonal insolation. Since the greater the pond depth, the more thermal mass can be stored

Total Pond Depth m	Total Salt Kg @MSF - PR6	Total Salt Kg @MSF - PR8
3.0	38,083,125	28,958,250
3.5	51,376,500	39,134,625
4.0	64,977,750	49,581,750
4.5	78,891,750	60,304,125
5.0	93,123,750	71,307,375
5.5	107,679,000	82,596,000
6.0	122,562,000	94,175,250
6.5	137,778,375	106,050,000

TABLE III-6 INITIAL SALT REQUIREMENT

ADDITIONAL SALT, Kg/DAY

Bottom Layer Depth m	MSF Plant @ PR 6	MSF Plant @ PR 8
1.5	1693	1287
2.0	1713	1305
2.5	1733	1322
3.0	1753	1340
3.5	1774	1359
4.0	1795	1377
4.5	1816	1395
5.0	1837	1414

(Bottom surface area maintained constant for each case)

TABLE III-7 ADDITIONAL SALT REQUIREMENT

(Based on top surface area of the lower-convecting zone) and a smaller change of pond temperature, the product output would be affected by the pond depth. Table III-8 shows the product output and percent of turndown capacity of the multi-stage flash plant at various pond depths and minimum temperature. During the month of December, when the insolation is at 200 w/m^2, the percent turndown product output capacity is 73.5% at 3.0 m pond depth and 95.8% at 6.5 m pond depth.

In order to establish a comparative costs figure at various pond depths, the capital and operating costs for the pond were developed. Table III-9 shows the costs of the pond at various pond depth. The costs of the pond for multi-stage flash plants is estimated to be approximately \$5.00 per m^3.

Table III-10 shows the initial salt costs for the pond at various pond depth between multi-stage flash plant performance ratio 6 and 8. The salt cost is based on the Cape Verde Island salt price at \$20.15 per metric ton and transportation costs at \$2.00 per metric ton.

Table III-11 shows the additional salt costs for the pond at various pond depth between multi-stage flash plant performance ratio 6 and 8. The differential additional salt costs are not significant at various pond depth. However, the differential salt costs between multi-stage flash plant performance ratio 6 and 8 is about \$3,000 per year.

Since the greater the pond depth, the more product water produced in the winter month of December, the product water sales are also greater at greater pond depth. Table III-12 shows the product water sales in December and differential costs at various

TOTAL POND DEPTH, m	MINIMUM POND TEMPERATURE, °C	PRODUCT OUTPUT m^3/day	TURNDOWN CAPACITY %
3.0	76.8	404.5	73.5
3.5	79.8	437.5	79.5
4.0	81.7	458.7	83.4
4.5	83.0	473.0	86.0
5.0	84.0	484.0	88.0
5.5	84.7	491.9	89.4
6.0	85.3	498.0	90.5
6.5	85.8	526.7	95.8

TABLE III-8 PRODUCT OUTPUT AT VARIOUS TOTAL POND DEPTH AND MINIMUM POND TEMPERATURE

TOTAL POND DEPTH	POND COSTS, $	
	MSF PLANT @ PR 6	MSF PLANT @ PR 8
3.0	$1,069,000	$ 823,000
3.5	1,247,000	962,000
4.0	1,436,000	1,101,000
4.5	1,635,000	1,257,000
5.0	1,837,000	1,414,000
5.5	2,050,000	1,575,000
6.0	2,280,000	1,735,000
6.5	2,515,000	1,906,000

TABLE III-9 SALT GRADIENT POND COSTS AT VARIOUS POND DEPTH

TOTAL POND DEPTH m	INITIAL SALT COSTS, $ @ $22.15/MT	
	@MSF PLANT PR 6	@MSF PLANT PR 8
3.0	$ 843,541	$ 641,425
3.5	1,137,990	866,832
4.0	1,439,257	1,098,236
4.5	1,747,452	1,335,736
5.0	2,062,691	1,579,458
6.0	2,714,748	2,085,982
6.5	3,051,791	2,349,008

TABLE III-10 INITIAL SALT COSTS

TOTAL POND DEPTH m	ADDITIONAL SALT COSTS, $/YR @$22.15/MT @MSF PLANT PR 6	@MSF PLANT PR 8
3.0	$12,319	$ 9,365
3.5	12,464	9,496
4.0	12,610	9,619
4.5	12,755	9,750
5.0	12,908	9,888
5.5	13,061	10,019
6.0	13,214	10,150
6.5	13,367	10,289

TABLE III-11 ADDITIONAL SALT COSTS

TOTAL POND DEPTH, m	PRODUCT WATER SALES IN DECEMBER @$11.0/$m^3$	DIFFERENTIAL COSTS, $
3.0	$137,935	Base
3.5	149,188	$11,253
4.0	155,417	17,482
4.5	161,293	23,358
5.0	165,044	27,109
5.5	167,738	29,803
6.0	169,818	31,883
6.5	179,605	41,670

TABLE III-12 PRODUCT WATER SALES IN DECEMBER AT VARIOUS POND DEPTH pond depth. The cost of product water is assumed as $11.00 per cubic meter.* In order to be comparative, the differential costs in product water sales in December would have to be subtracted from the total annual costs of the pond.

Table III-13 and III-14 shows the annual costs of the pond at various pond depth for multi-stage flash plant between performance ratio 6 and 8. The annual capital costs are based on a fixed charge rate of 10 percent. The result of the annual differential costs showed that a pond with 3m depth is the most economical. The total annual costs of the pond at 3m depth are $203,573 or multi-stage flash plant at performance ratio of 6 and $155,807 at performance ratio of 8.

E. Pond Sizing for the Reverse Osmosis Plant

The pond sizing for the reverse osmosis plant is continued in Section V-D.

F. Pond Process Description

As shown in Figure III-2, salt gradient pond brine is first transported into a duplex strainer located on the suction side of the pump. Large debris will be removed from the pond brine by a basket-type 3/8" screen. The pond brine is then pumped to a poro-edge automatic backflush-type strainer to remove solids of 0.02 inches and larger from the brine, with a pressure drop no greater than 2 psi.

The strainer does not require a separate water supply for backwashing or disruption of service to backwash and clean the strainer element. The strainer element is made of stainless steel and consists of a trapezoidal shaped wire wrapped helically on, and resistance welded to, a suitable core to provide 0.20 inch shaped openings or Delrin perforated discs with 1/64" tapered holes.

After the removal of solids, the pond brine then flows into the multi-stage flash unit brine heater to heat the recycle brine from the multi-stage flash unit first stage.

*current production cost (unsubsidized) on Sal Island

COSTS	SALT GRADIENT POND DEPTH, m 3.0	3.5	4.0	4.5	5.0	5.5	6.0	6.5
@MSF Plant-PR6								
Capital Costs								
Pond	$1,069,000	1,247,000	1,436,000	1,635,000	1,837,000	2,050,000	2,280,000	2,515,000
Salt	843,541	1,137,990	1,439,257	1,747,452	2,062,691	2,385,090	2,714,748	3,051,791
Total Capital Costs	$1,912,541	2,384,990	2,875,257	3,382,452	3,899,691	4,435,090	4,994,748	5,566,791
FCR @10%, $/yr.	191,254	238,499	287,525	338,245	389,969	443,509	499,474	556,679
Operating Costs								
Salt, $/yr.	$ 12,319	12,464	12,610	12,755	12,908	13,061	13,214	13,367
Total Annual Costs	$ 203,573	250,963	300,135	351,000	402,877	456,570	512,688	570,046
Diff. Water Sales in Dec.	Base	11,253	17,482	23,358	27,109	29,803	31,883	41,670
Relative Costs	$ 203,573	239,710	282,653	327,642	375,768	426,767	480,805	528,376
Differential Costs	Base	36,137	79,080	124,069	172,195	223,194	277,232	324,803

TABLE III-13 ANNUAL POND COSTS FOR MULTI-STAGE FLASH PLANT AT PERFORMANCE RATIO 6 (COMPARISON FOR VARIOUS DEPTHS)

COSTS	SALT GRADIENT POND DEPTH, m							
	3.0	3.5	4.0	4.5	5.0	5.5	6.0	6.5
@MSF Plant-PR8								
Capital Costs								
Pond	$ 823,000	962,000	1,101,000	1,257,000	1,414,000	1,575,000	1,735,000	1,906,000
Salt	641,425	866,832	1,098,236	1,335,736	1,579,458	1,829,501	2,085,982	2,349,008
Total Capital Costs	$1,464,425	1,828,832	2,199,236	2,592,736	2,993,458	3,404,501	3,820,982	4,255,008
FCR @10%, $/yr.	146,942	182,883	219,923	259,273	299,345	340,450	382,098	425,500
Operating Costs								
Salt	9,365	9,496	9,619	9,750	9,888	10,019	10,150	10,289
Total Annual Costs	$ 155,807	192,379	229,542	269,023	309,233	350,469	392,248	435,789
Diff. Water Sales in Dec.	Base	11,253	17,482	23,358	27,109	29,803	31,883	41,670
Relative Costs	$ 155,807	181,126	212,060	245,665	282,124	320,666	360,365	394,119
Differential Costs	Base	25,319	56,253	89,858	126,317	164,859	204,558	238,312

TABLE III-14 ANNUAL POND COSTS FOR MULTI-STAGE FLASH PLANT AT PERFORMANCE RATIO 8
(COMPARISON FOR VARIOUS DEPTHS)

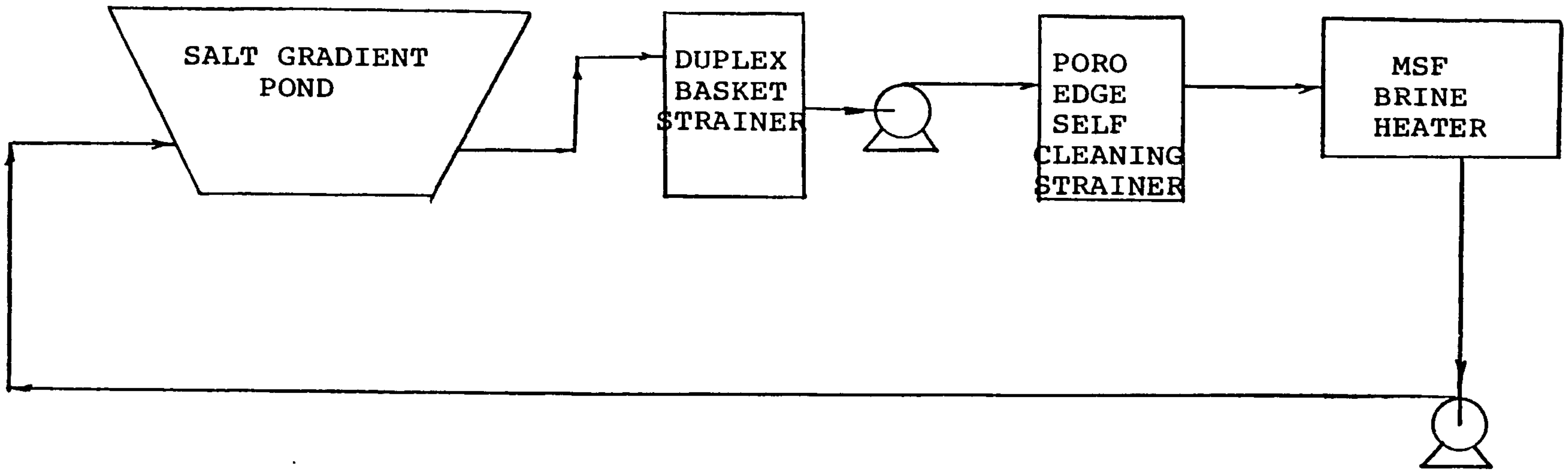


FIGURE III-2 SALT GRADIENT POND BRINE PROCESS FLOW DIAGRAM

In order to keep the recycle brine temperature as high as possible up to 180^{o}F, a liquid-to-liquid heat exchanger, operating at above atmospheric pressure, was selected against a steam-to-liquid heat exchanger. Although a steam-to-liquid heat exchanger requires much less surface area than a liquid-to-liquid heat exchanger, the added cost for the flash chamber operating at sub-atmospheric conditions, the lower top brine temperature to the first MSF stage, and the possibility of salt precipitation, will present many operational problems.

To prevent the growth of algae at the bottom of the salt gradient pond, copper sulfate will be added intermittently. In addition, hydrochloric acid is added to maintain a pH below 7 to keep the copper in solution.

Sodium chloride salt will be added to the recirculating pond brine to maintain the density gradient of the solar pond due to salt diffusion toward the surface layer. Seawater make-up will be added regularly to replace the pond top surface seawater due to evaporation and increases of salt concentration from diffusion.

Figures III-3 and III-4 show the piping and equipment of the salt gradient pond to supply heat to the multi-stage flash unit at performance ratio 6 and 8. The pond brine flow to the multi-stage flash brine heater is 1377 gpm at performance 6 and 1134 gpm at performance ratio 8. The piping material will be FRP.

Figures III-5 and III-6 show the configuration of the salt gradient pond for multi-stage flash plant at performance ratio 6 and 8.

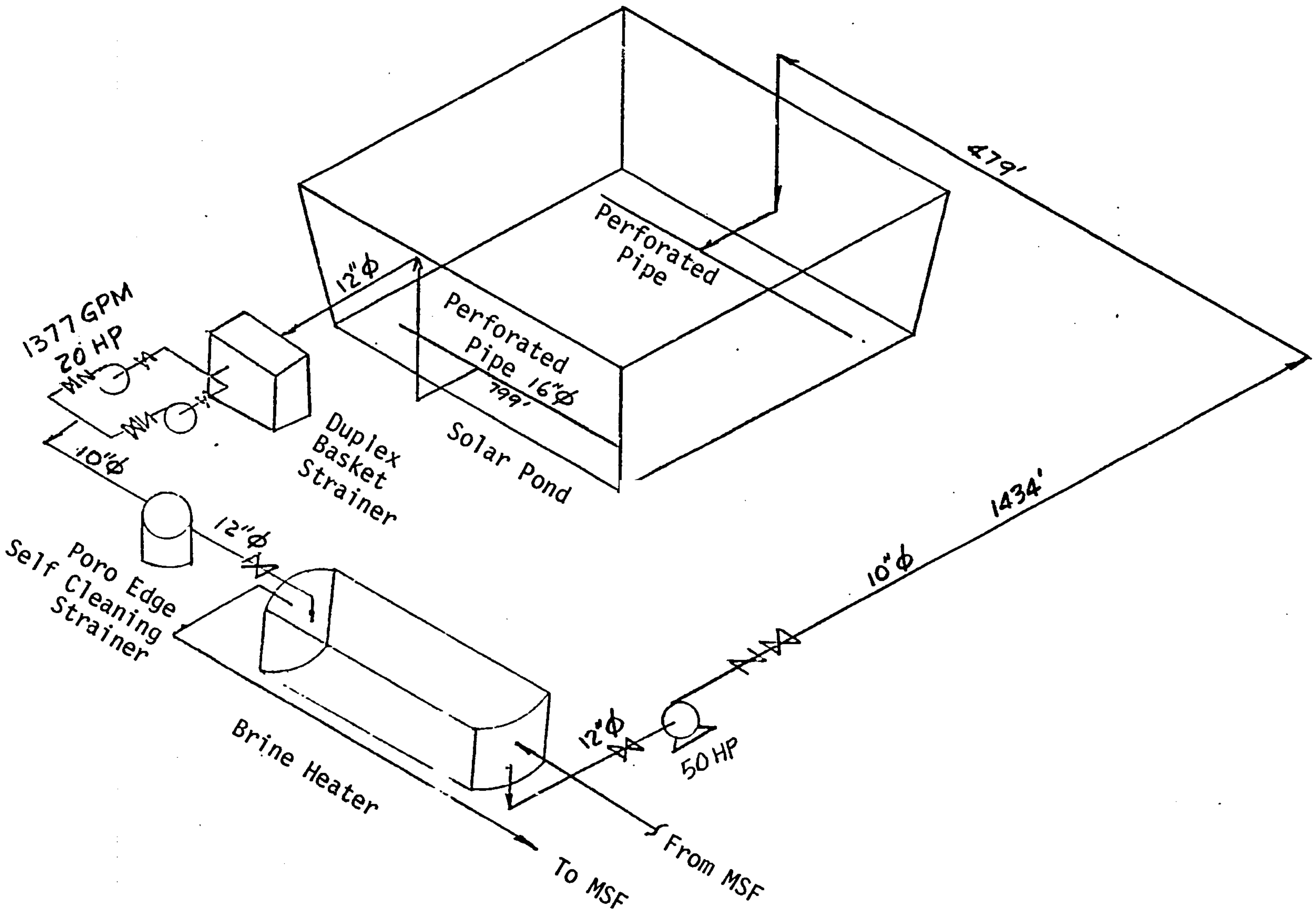


FIG. III-3 SALT GRADIENT POND PIPING AND EQUIPMENT AT MSF - PR6

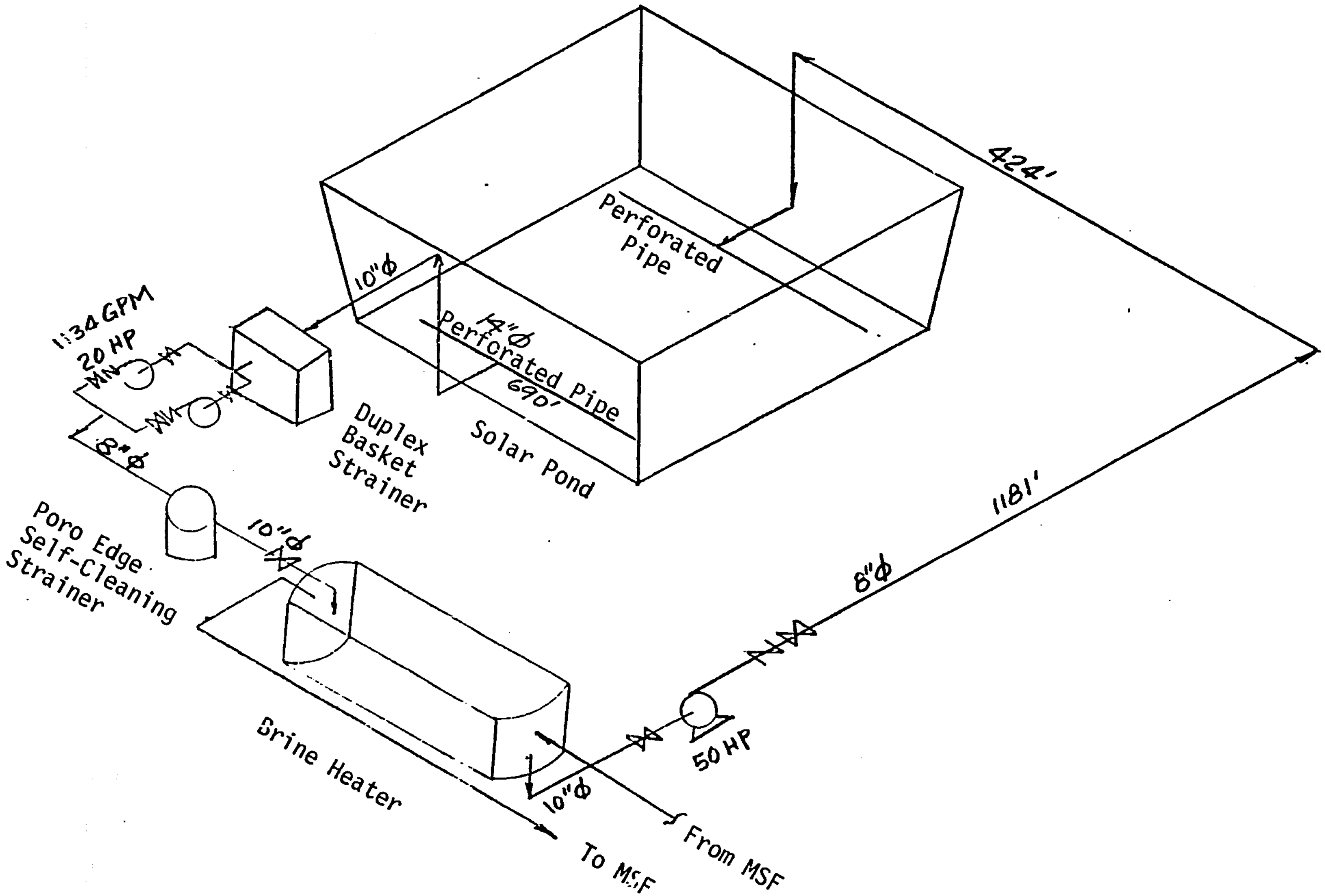


FIG. III-4 SALT GRADIENT POND PIPING AND EQUIPMENT AT MSF - PR8

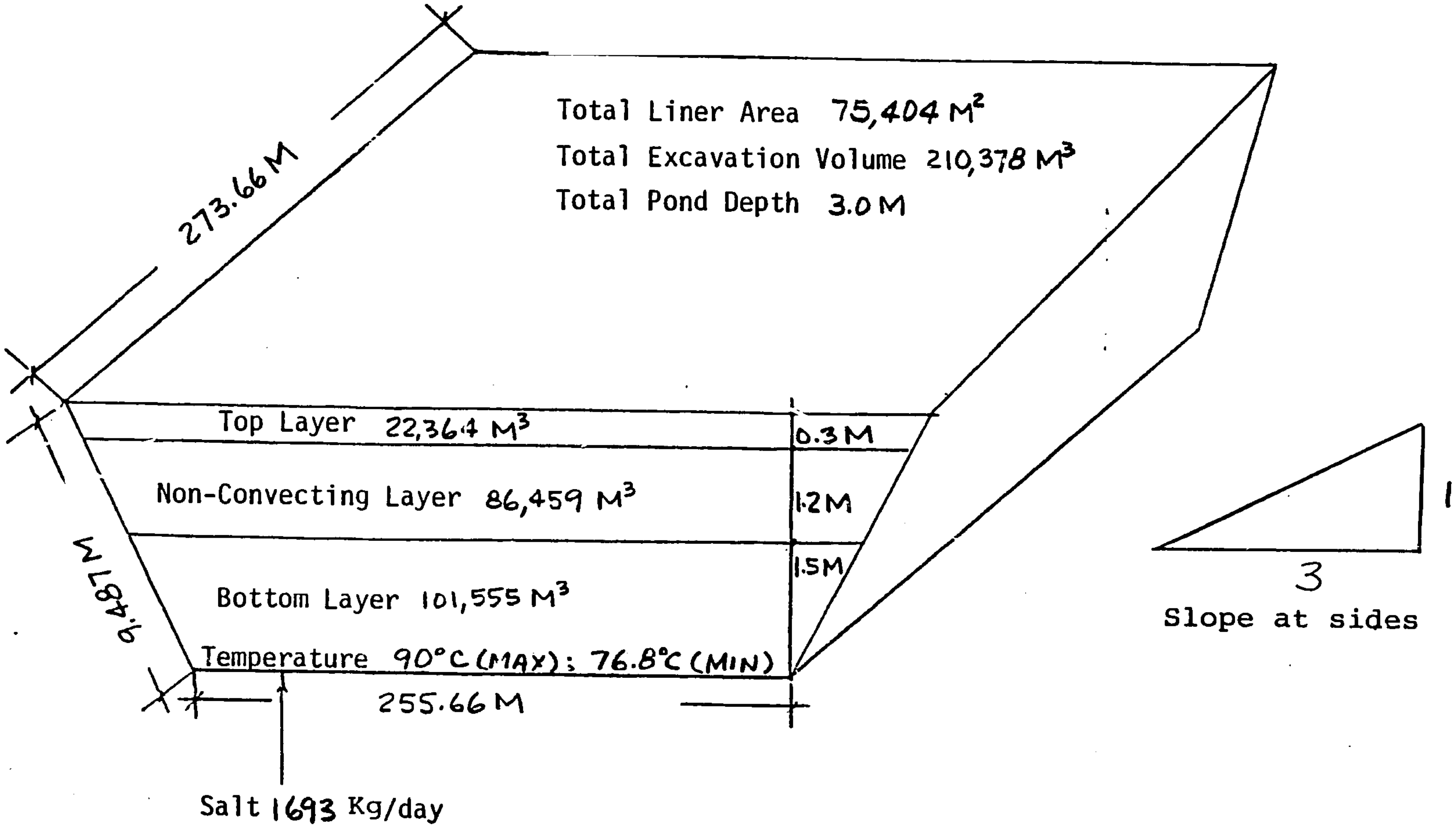


FIG. III-5 SALT GRADIENT POND AT PERFORMANCE RATIO 6

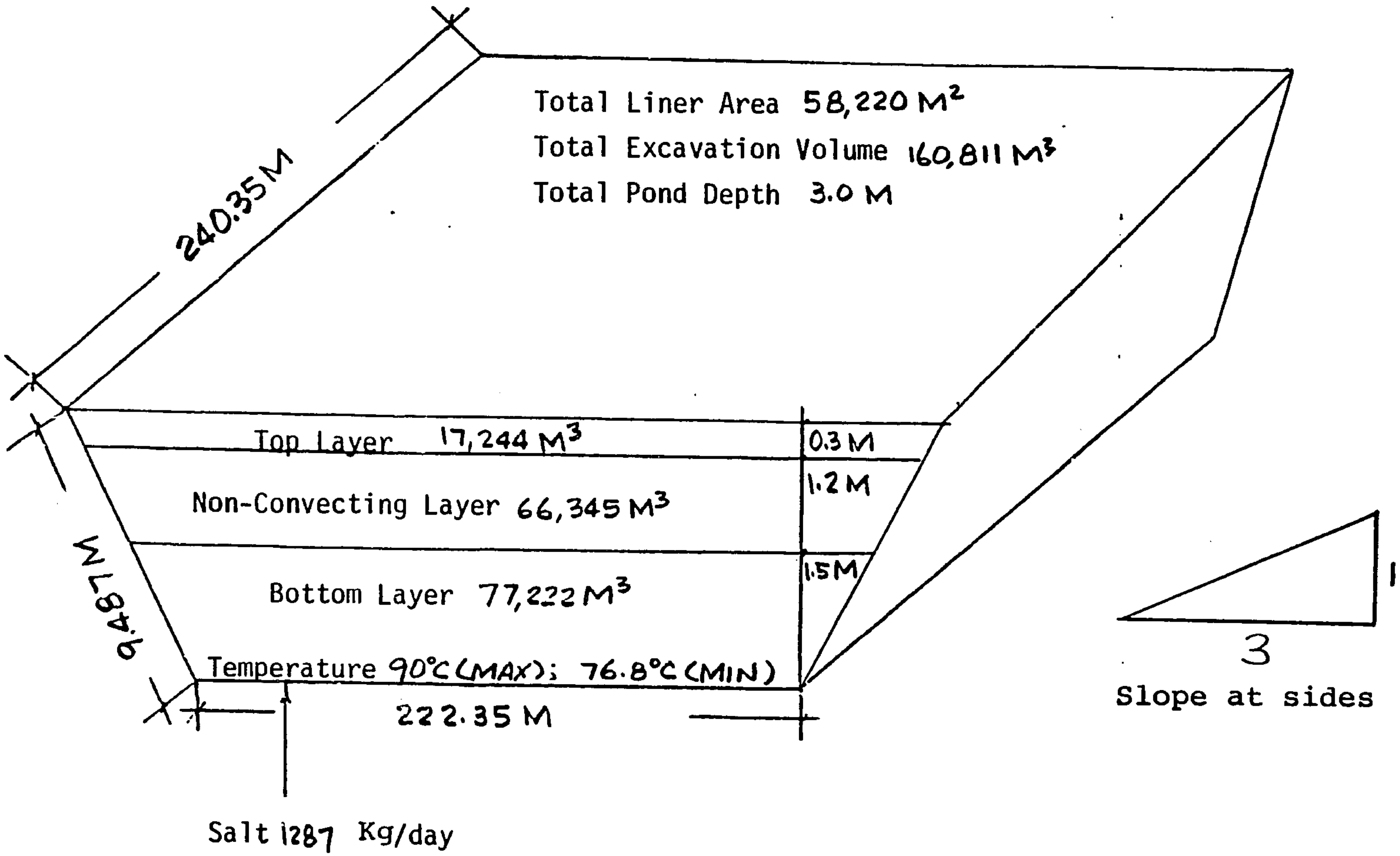


FIG. III-6 SALT GRADIENT POND AT PERFORMANCE RATIO 8

Pond Costs

The salt gradient pond costs can be summarized as follows:

	Salt Gradient Pond Costs @3 meter depth	
	MSF - PR6	MSF - PR8
Bottom Surface Area, m^2	65,361	49,441
Heat to MSF Brine Heater, Btu's/hr	8,409,167	6,306,875
Capital Costs		
Pond Costs (with liner)	1,069,000	823,000
Initial Salt Costs	843,541	642,425
Costs of Pond with Salt	1,912,541	1,465,425
Costs of Piping & Equipment (excluding brine heater)	475,000	362,000
Total Pond Costs (Incl. salt, piping and equipment)	2,387,541	1,827,425
Annual Capital Costs @FCR 10%	238,754	182,742
Operating Costs (at 90% utilization)		
Electricity @16.5¢/kwh	60,490	58,148
Salt @ $22.15/MT	12,319	9,365
Total Operating Costs	72,809	67,513
Total Annual Costs	311,563	250,255
Unit Costs		
Pond, $/$m^2$ (with liner)	16.35	16.64
Pond with Salt, $/$m^2$	29.26	29.64
Pond with Salt, piping & equipment, $/$m^2$	36.53	36.96
Energy, $/million Btu's	4.70	5.03
Pond heat $/million Btu's	3.79	3.86

SECTION IV

MULTISTAGE FLASH PLANT

(Direct Heat Use)

Prepared by: Albino Ko

SECTION IV - MULTISTAGE FLASH PLANT
(Direct Heat Use)

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IV. MULTI-STAGE FLASH PLANT (Direct Heat Use)

Of the desalting plant processes desalting seawater, over 90% of the installed world capacity is made up of the MSF plants. The MSF process is the most proven and the technology of this process is better defined, understood and developed.

In MSF plants, the latent heat of evaporation must be reused many times for distillation processes to be economically practical. The multi-stage flash distillation process serves this purpose by using seawater/brine flow-through heat-recovery and heat-rejection tubes to condense vapors successfully flashed at progressively lower pressures in individual stages.

This section is to cover the selection of a MSF plant between once-through and recycle system and between a performance ratio of 6 and 8.

A. MSF Alternative Designs

Basically, there are two multi-stage flash processes that can be used for desalting. They are the once-through and recycle process design.

In the once-through process, raw seawater enters the cold end of the evaporator heat transfer tubing and is progressively heated in each stage. The seawater exits from the hottest stage and passes through the brine heater, where it is heated to a temperature of 88^{o}C (190^{o}F). The hot brine is then admitted into the hottest stage flashing brine chamber. The flashed steam released becomes the product water as it is condensed by the seawater in heat transfer tubing. The flashing brine is progressively exposed to lower pressures in each stage.

At the lowest temperature stage, the brine chamber is at very low pressure (sub-atmospheric). A pump is used to draw off the remaining brine from the last stage for discharge back to the sea. In this process, the brine concentration in the last stage is about 10% greater than the seawater supply.

The recycle process operates in a similar manner except that the raw seawater is withdrawn after passing the first 3 (or 4) stages of the evaporator. These are termed "heat reject stages" and operate with seawater directly within the heat transfer tubing. The seawater, after passing through the heat reject stages, is split with the major portion of the flow discharged back to the sea. After the addition of make-up seawater, the brine is recycled through the heat exchanger tubing of the recovery section, the brine heater and returns as flashing brine to the last stage where the cycle is repeated. A portion of the brine in the last stage is discharged to the sea as blowdown. The blowdown brine concentration can be varied to maintain a preset maximum concentration, in this case 1.7 times the seawater feed.

The advantages of each process is as follows:

Once-Through

1. More economical system to fabricate.

2. Lower brine concentrations.

Recycle

1. Lower chemical costs for scale control additives.

2. Only in the heat reject section, 3 to 4 stages, is the heat transfer tubing exposed to the more aggressive raw seawater, the remaining stages are cooled by

recirculated brine. On the other hand, all stages, approximately 16 to 21 of the once-through system, are exposed to raw seawater.

3. Not sensitive to variations in seawater temperature.

4. Seawater make-up to the low temperature stage of the recycle system represents only one-third of the total seawater feed. Therefore, the amount of dissolved gas which is highly corrosive to the shell and tubes, entering the system is considerably less than the once-through system where practically all the dissolved gas in the seawater is released in the highest temperature stage.

5. The recycle arrangement lends itself to the application of a deaerator to reduce the amount of non-con densable gas (primarily oxygen and carbon dioxide) to less than 50 parts per billion. This feature can be expected to double the service life of the tubes and evaporator intervals.

Based on the above evaluation, we recommend the recycle system because it is more economical, has longer tube and evaporator shell life and provides more stability in plant operation.

B. Scale Control

The multi-stage flash process is generally used with either polyphosphate, polymer or acid treatment for evaporator scale control.

Acid treatment permits operation with brine temperature as high as 250°F. The high temperature results in a more efficient operation and/or a lower initial cost plant. However, acid feed rate is critical and automatic control of this rate requires a high level of maintenance to insure satisfactory operation. Improper operation of the acid feed system can result in rapid scaling of the heat transfer surfaces or excessive corrosion, depending upon the amount of acid fed. Either condition can have a major impact upon the plant function. Acid treatment also requires the handling of sulfuric acid, which is hazardous, generally utilizes skilled operators, and requires more maintenance.

Polyphosphate treatment limits operation to a 195°F maximum brine temperature. The lower thermal gradient results in an incremental increase in initial capital cost. However, the polyphosphate feed process has the distinct advantage of control simplicity and low impact upon plant operation when errors in feed rate take place. There is also polymer-type scale control which permits operation up to a maximum of 250°F. However, the costs of polymer are higher at this time and comprehensive technical proof from operating facilities is not conclusive enough to recommend at this time.

Since the solar pond is designed at 194°F (90°C) we, therefore, recommend polyphosphate treatment for scale control.

C. Recycle Plant Selection

The evaporator stage is the heart of the multi-stage flash plant and is where the production of product vapor takes place. The number of stages in a plant is determined by analyses of the total water costs. The object is to design a plant which has the lowest water costs. The total water costs are made up of the annual capital charges and annual operating costs which cover energy, chemicals, maintenance, etc.

The number of stages do not determine the plant production, but only the plant efficiency and water costs. The plant production is based solely on the total plant flashdown and flows. Therefore, a one-stage plant with a flashdown of 90^{O}F (180^{O}F to 90^{O}F) will produce as much water as a 20-stage plant with the same flashdown. The big difference, however, is in the energy (heat input to brine heater) requirements for the two plants, and the size of the equipment. The higher efficiency, the design of the plant, the more stages it will require. Therefore, a plant with a performance ration of 8 will have more stages than a plant design with a performance ratio of 6.

Figures IV-1 and IV-2 show the heat and material balances of the multi-stage flash plant at performance ratio 6 and 8. The number of stages required at performance ratio of 6 is about 16 stages, and at performance ratio of 8 is 22 stages. Table IV-1 summarized the major parameters between performance ratio 6 and 8.

In order to establish a comparative cost figure between performance ratio 6 and 8 for a multi-stage flash system desalting plant, cost data was obtained from the manufacturers for a single train with polyphosphate treatment. Tables IV-2 and IV-3 are chemical and electrical consumption estimates and chemical cost for the multi-stage flash plant based on 90% utilization at performance ratio 6 and 8.

Table IV-4 shows a breakdown of the quantities and crafts of workers, and their salaries, needed for operation of the salt gradient pond and multi-stage flash plant. The salaries are based on Cape Verde Island labor costs. The operating labor required is the same at performance ratio 6 and 8.

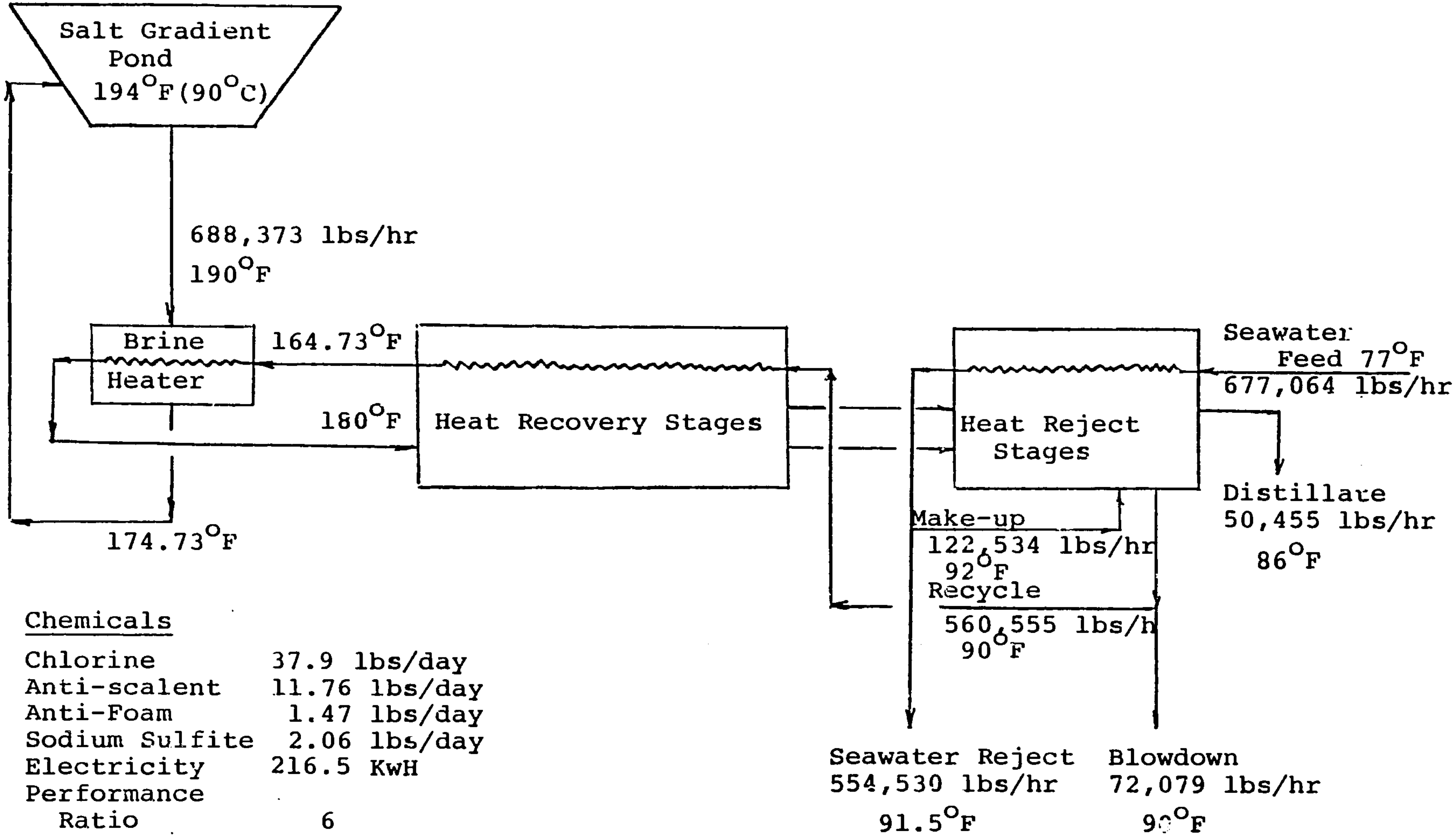
Table IV-5 presents the annual capital and operating costs for multi-stage flash plant at performance ratio 6 and 8. The

	Performance Ratio	
	6	8
Solar Pond, Acres	16.14	12.2
Solar Pond Salt Requirement, Kg	38,083,125	28,958,250
Brine from Solar Pond to MSF, Kg/hr	688,373	567,165
Heat Input to MSF Btu/hr	8,409,167	6,306,875
MSF - No. of Stages	16	22
Brine Heater, SF	3364	2523

TABLE IV-1 MULTI-STAGE FLASH PLANT MAJOR PARAMETERS

	Performance Ratio	
	6	8
Electricity, KW	216.5	200.7
Chemicals		
Chlorine, Kg/day	17.19	13.52
Polyphosphate, Kg/day	5.33	5.33
Anti-Foam, Kg/day	0.67	0.67
Sodium Sulfite, Kg/day	0.93	0.93
Salt Required in Solar Pond, Kg/day	1693	1287

TABLE IV-2 CHEMICAL AND ELECTRICAL CONSUMPTION OF MULTI-STAGE FLASH PLANT AT PERFORMANCE RATIO 6 AND 8



<u>Chemicals</u>

Chlorine	37.9 lbs/day
Anti-scalent	11.76 lbs/day
Anti-Foam	1.47 lbs/day
Sodium Sulfite	2.06 lbs/day
Electricity	216.5 KwH
Performance Ratio	6
Total Number of Stages	16

FIGURE IV-1 SALT GRADIENT POND - DESALINATION PLANT AT PERFORMANCE RATIO 6

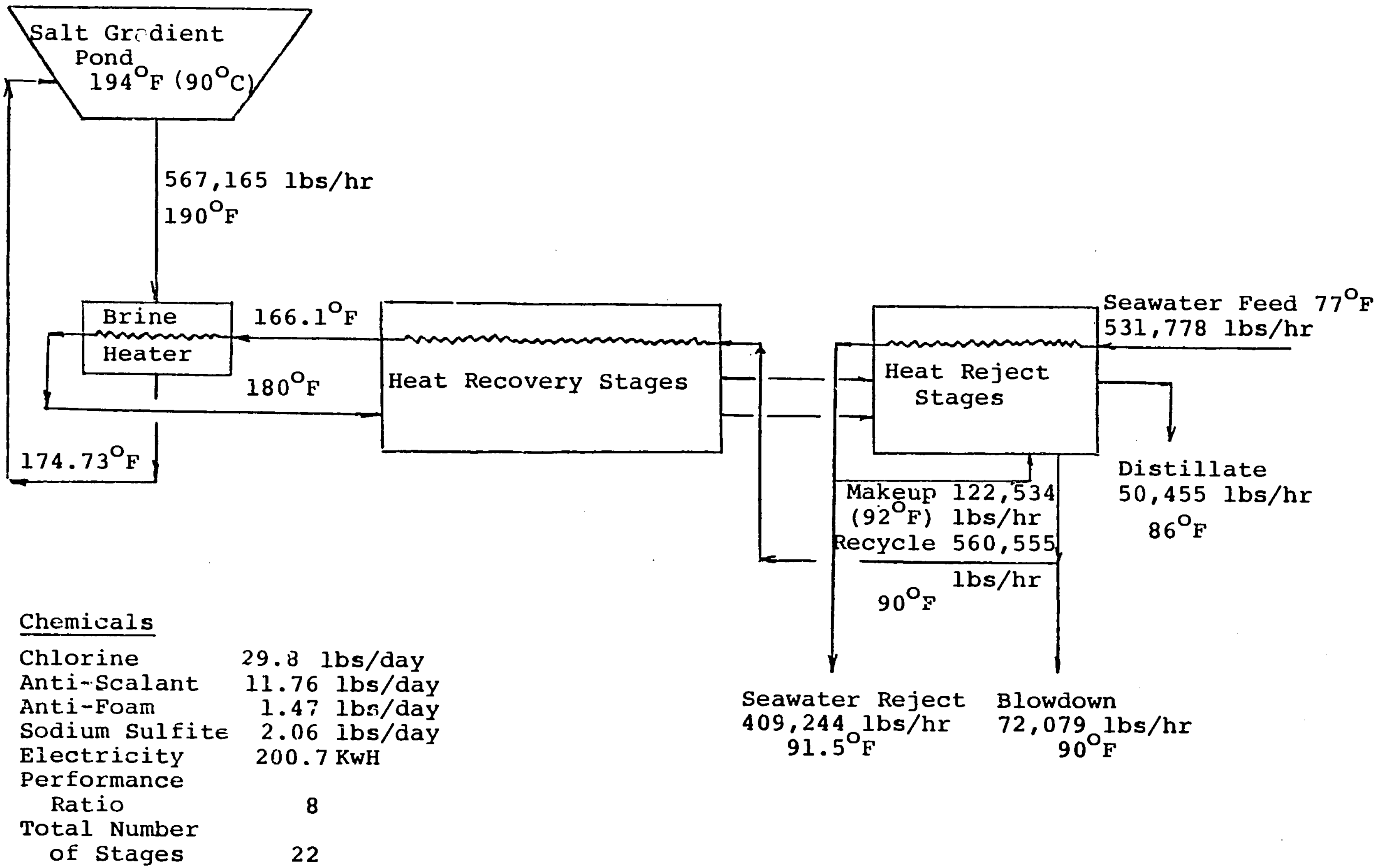


FIGURE IV-2 SALT GRADIENT POND - DESALINATION PLANT AT PERFORMANCE RATIO 8

	Performance Ratio	
	6	8
Chemical Costs $/yr		
*Chlorine @ $0.73/Kg	$ 4,122	$ 3,242
Polyphosphate @ $3.31/Kg	6,230	6,230
Anti-Foam @ $4.41/Kg	1,039	1,039
Sodium Sulfite @ $0.55/Kg	183	183
Total Chemical Costs	$ 11,574	$ 10,694

*Chlorine is based on 4.41 Kwh/Kg and $0.165/Kwh

Chemicals transportation cost from U.S.A. to Cape Verde Island @ $150/T

TABLE IV-3 CHEMICAL COSTS FOR MULTI-STAGE FLASH PLANT AT PERFORMANCE RATIO 6 AND 8

OPERATING LABOR	NO. REQ'D.	ANNUAL RATE	LABOR COSTS
Plant Superintendent	1	$ 8,000	$ 8,000
Maintenance Supervisor	1	5,500	5,500
Chief Operator	1	5,500	5,500
Chemist	1	4,500	4,500
Foreman	4	3,744	14,976
Field Operator	4	2,300	9,200
Field Helper	4	1,610	6,440
Mechanics	2	4,118	8,236
Electrician	1	4,118	4,118
Janitor	1	1,373	1,373
Secretary	1	2,300	2,300
TOTAL			$70,143

TABLE IV-4 OPERATING LABOR COSTS FOR MULTI-STAGE FLASH PLANT AT PERFORMANCE RATIO 6 AND 8 seawater intake costs are based on the Cape Verde desalting and power plant project escalated from September, 1980 to the fourth quarter of 1981 dollars. The total capital costs for multi-stage flash plant at performance ratio of 6 is $4,377,000 and at performance ratio of 8 is $3,993,000. The total annual cost based on the operating costs and capital costs at fixed charge rate of 10% is $813,372 at performance ratio of 6 and $750,585 at performance ratio of 8. The water cost at 90% utilization is $17.01 per 1000 gallons, or $4.50 per cubic meter at performance ratio of 6 and $15.72 per 1000 gallons, or $4.16 per cubic meter at performance ratio 8. The cost figures show that multi-stage flash plant at performance ratio 8 is more economical.

D. Plant Description

The 550 m^3/day unit is of conventional single effect multi-stage flash design similar to those which have been supplied by several commercial desalting plant manufacturers.

Figure IV-3 shows the multi-stage flash plant process flow diagram. The seawater flows enter the heat reject stage at 77^{o}F and leaves at 91.5^{o}F. In each stage, the seawater is heated by condensing vapor produced by brine which is flashing in the bottom of the stage and distillate which is flashing in the product trays. The flashing brine and product flow counter to the direction of the seawater. After passing through the heat reject module, most of the seawater is returned to the sea, but about 122,500 pounds of it is used per hour as makeup for the plant. The makeup stream is then sprayed into a vacuum deaerator which removes dissolved oxygen and reduces carbon dioxide. Oxygen and carbon dioxide levels of the makeup leaving the deaerator should be less than 50 ppb and 4 ppm respectively. Downstream of the deaerator, connections are provided for sodium sulfite addition to scavenge the remaining oxygen.

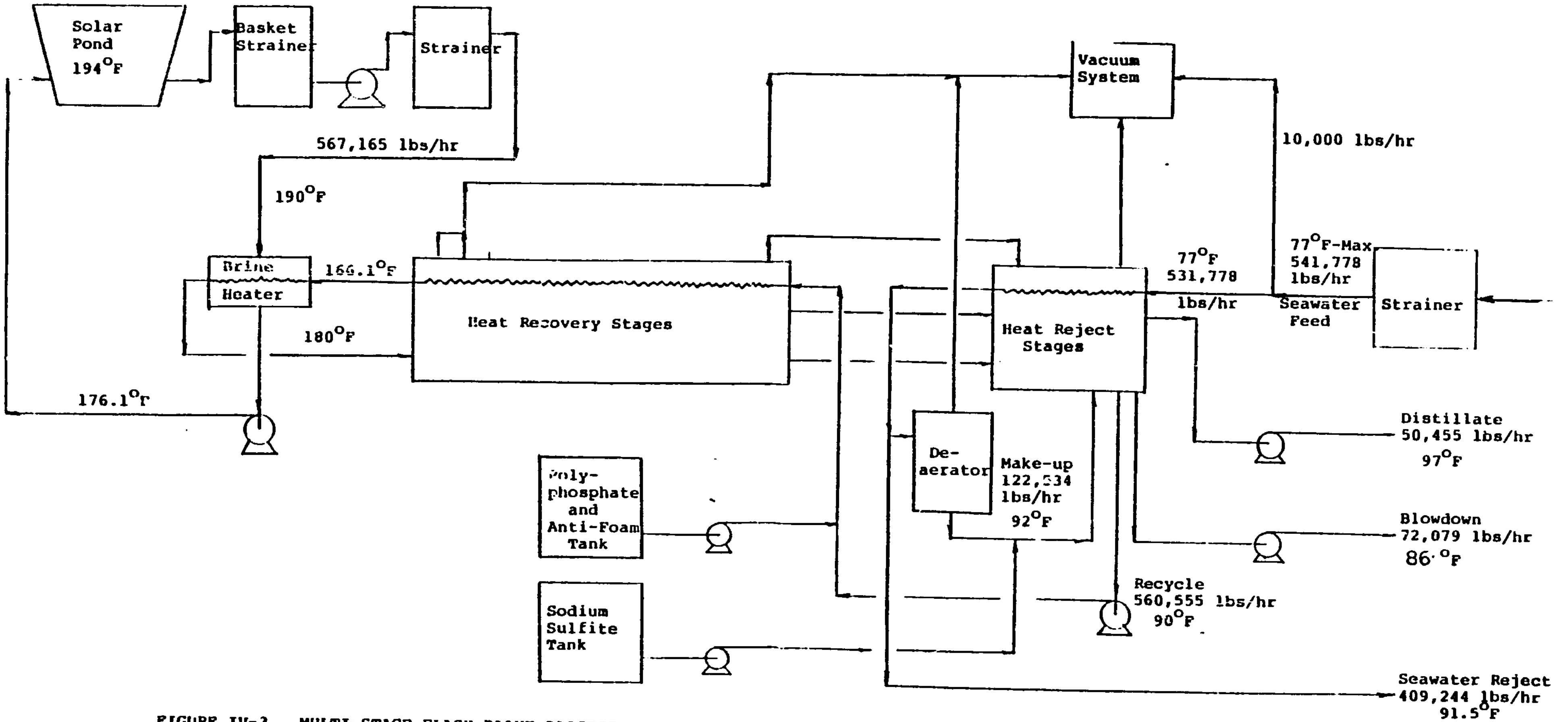


FIGURE IV-3 MULTI-STAGE FLASH PLANT PROCESS FLOW DIAGRAM

The pretreatment makeup then flows to the recycle side of the sump in the last stage, where it mixes with brine to form the recycle brine stream.

The recycle brine flows from the sump at 90^{o}F and a rate of 560,555 pounds per hour to the water box on the lowest temperature stage of the heat recovery module. Polyphosphate is added at the recycle brine to inhibit calcium carbonate scale formation. The recycle brine flows through the tube bundles in the heat recovery module in the direction of increasing temperature. In each stage, the recycle brine gains heat from vapor produced by flashing brine and distillate, which condenses on the outer surface of the tubes. Upon leaving the first stage water box at a temperature of 166.1^{o}F, the recycle brine enters the brine heater where solar pond brine at 190^{o}F raises the recycle brine temperature to the maximum of 180^{o}F.

A control valve is located downstream of the brine heater to control the recycle brine flows, and thus the plant output, as well as to help maintain enough back pressure in the brine heater to prevent flashing in the tubes and the result of scaling. The valve may be followed by an orifice which will reduce flashing at the valve. After passing through the valve (or orifice), the brine enters Stage No. 1 and flashes off enough vapor to bring it nearly to equilibrium with the pressure within the stage allowing for boiling point elevation due to salts and pressure drops due to demisters. The vapor passes through demister pads to the recycle brine tube bundle, where it is condensed. The condensed distillate falls to the product trays.

A vacuum and vent system is provided to evacuate air from the plant at startup and removes off-gases and air inleakage during operation. The proper tube bundle configuration and vapor baffle positioning is required to provide adequate sweep velocities and subcooling to prevent gas binding of the tube bundle without excessive vapor flow through the venting without excessive vapor flow through the venting system.

A water jet ejector is provided to evacuate the non-condensable gases. Motive water for ejector operation is circulated by a pump in a closed circuit. This circuit incorporates a motive water tank where any gases dissolved in the water are liberated and exhausted to atmosphere. The non-condensable gases, having a higher temperature than the motive water, would gradually cause an undesirable heating of the circulating motive water. A portion of the motive water is, therefore, continuously drained by an overflow device arranged inside the motive water tank and is substituted by an equal amount of cold seawater.

For start-up operation of the plant, a second water jet ejector (hogging ejector) is provided to evacuate the evaporator vessel to the required initial starting vacuum.

The flashing brine flows through an adjustable opening in the stage divider to Stage No. 2, which is at slightly lower pressure than Stage No. 1. Again, flashing and consequent cooling of the brine takes place. The vapor is again condensed on the recycle brine tube bundle and collected in the product tray. The distillate collected in Stage No. 1 flows in the product tray through a loop seal to Stage No. 2, where it flashes down to equilibrium with the product condensed in Stage No. 2. The vapor produced by the flashing distillate from Stage No. 1 is also condensed on the recycle brine tube bundle and collected in the product tray.

This process continues throughout the heat recovery stages with flashing brine transferred between modules by ductwork, while the product is transferred using piping. Loop seals are provided where necessary to prevent blow-through of vapor from one stage to subsequent lower pressure stages. The width of the modules and ductwork is increased for the lower temperature modules to provide adequate disengagement area and lower pressure drops between stages.

After leaving the heat rejection, stages, the flashing brine and product enter the heat reject module. Here, the flashing-down process continues in the same manner as in the heat recovery modules, with the exception that the cooling medium in the tubes is raw seawater rather than recycle brine. At the end of the last stage, the flashing brine, now at a temperature of 90^{o}F, enters a divided sump. On one side, make-up is added to form the recycle stream; on the other side, brine is withdrawn by a blowdown pump to maintain the solids content within the allowable limits.

Product is withdrawn from the sump at the end of the last stage product tray and pumped to the product storage facilities.

During the winter months, when the insolation is 200 w/m^2, and the minimum salt gradient pond temperature is 76.8^{o}C, the product water capacity will be reduced from 550 m^3/day to 405 m^3/day, due to the lower total flashdown of the plant. The turndown capacity at 73.6% is still within the acceptable operating range in multi-stage plant without fine tuning.

E. Multi-Stage Flash Plant Costs

As shown in Table IV-5, the total capital cost of the multi-stage flash plant at performance ratio of 8, with solar pond to supply heat to the MSF brine heater, is $3,993,000, and the operating cost is $351,285 per year. If the annual fixed charges are based on 10 percent of capital cost, the total annual cost is $750,585. The water cost at 90% utilization is $15.72

	Performance Ratio	
	6	8
Capital Costs		
Solar Pond*	$ 2,387,000	$ 1,827,000
Seawater Intake & Piping	277,000	241,000
Strainer	20,000	18,000
Deaerator	30,000	30,000
MSF	1,663,000	1,877,000
Total Capital Costs	$ 4,377,000	$ 3,993,000
FCR @10%, $/Yr	$ 437,700	$ 399,300
Operating Costs $/Yr @90% Utilization		
Salt	$ 12,319	$ 9,365
Chemicals	11,574	10,694
Electricity @ $0.165/Kwh	281,636	261,083
Labor	70,143	70,143
Total Operating Costs	$ 375,672	$ 351,285
Total Annual Costs	$ 813,372	$ 750,585
$/1000 gal @90% Utilization	$ 17.01	$ 15.72
$/m^3	4.50	4.16

*Complete with liner, salt, pumping equipment and piping

TABLE IV-5 CAPITAL AND OPERATING COSTS FOR MULTI-STAGE FLASH PLANT AT PERFORMANCE RATIO 6 AND 8 per 1000 gallons, or $4.16 per cubic meter. This case is also shown on Table IV-6 as Case I.

Due to the high electricity costs in Case I, the addition of a solar pond to supply electric power to the MSF plant is being considered in Case II. In order to supply 156 Kwh to the MSF plant and 53.6 Kwh to pump the solar brine to the heat exchanger and back to the pond, the pond will require a total bottom surface area at 74,659 m^2. The large pond area is due to the low overall efficiency at 7.5% to generate electric power. As shown in Table IV-6, the capital and yearly operating cost in Case II is $7,516,111 and $104,238. The water cost is $17.93/1000 gal, or $4.74 per cubic meter, which is higher than Case I.

As shown in Table IV-6, Case III is a typical conventional single-purpose MSF plant with a package boiler to supply heat to the brine heater, no solar pond is considered in this case. The total amount of steam required for the MSF plant is 2,976 Kg/hr and the total consumption of fuel oil for the boiler is 201.4 Kg/hr. Due to the high fuel oil costs in the Cape Verde Island at $460/MT, the annual cost for fuel oil at $731,922 made this case unattractive. The capital and annual operating costs in this case are $2,100,250 and $1,073,842. The water cost is $26.90/1000 gal, or $7.11 per cubic meter.

F. Alternative Process

Other processes that were reviewed and found to be substantially less attractive are:

1. Multiple-Effect Distillation Process

The multiple-effect distillation process may be considered as a number of single effects connected in series. Each of the

Capital Cost	Case I Solar Pond/MSF	Case II Solar Pond/Power Solar Pond/MSF	Case III Boiler/MSF
MSF	$ 1,877,000	$ 1,877,000	$ 1,811,250
Seawater Intake	241,000	241,000	241,000
Strainer	18,000	18,000	18,000
Deaerator	30,000	30,000	30,000
Solar Pond for Brine Heater	1,827,000	1,827,000	-
Solar Pond for Power Generation	-	3,523,111	-
Boiler	-	-	50,000
Total Capital Cost	$ 3,993,000	$ 7,516,111	$ 2,100,250
FCR 10%	399,300	751,611	210,025
Operating Cost $/yr			
Salt	$ 9,365	$ 23,401	$ -
Chemicals	10,694	10,694	10,694
Electricity @ $0.165/Kwh	261,083	-	261,083
Fuel Oil @ $460/MT	-	-	731,922
Labor	70,143	70,143	70,143
Total Operating Cost $/yr	351,285	104,238	1,073,842
Annual Capital Cost $/yr	399,300	751,611	210,025
Total Annual Cost $/yr	$ 750,585	$ 855,849	$ 1,283,867
@90% Utilization, $/1000 gals.	15.72	17.93	26.90
$\$/m^3$	4.15	4.74	7.11

TABLE IV-6 CAPITAL AND OPERATING COSTS FOR MULTI-STAGE FLASH PLANT AT PERFORMANCE RATIO 8 effects has a pump to recycle the brine. Steam heats the brine in the first effect. A portion of the brine is vaporized and the remainder is recycled. The vapor generated is directed to the heat transfer surfaces of the next effect, is condensed and becomes the product water. Each effect blows down brine to the next effect. This blowdown becomes the makeup to the next effect. The vapor from the last effect is condensed in a final condenser which uses seawater to reject the heat.

Multiple-effect plants employ either horizontal or vertical tube arrangements. In horizontal tube evaporators, the vapor is generated on the outside of the tube. In the vertical tube evaporators, the vapor is generated inside the tube. The recycle flow to each effect is by spray film through nozzles in either the inside of vertical tubes or the outside of horizontal tubes, depending upon the tube arrangement being employed. There are several variations of multi-effect plants which may differ somewhat. from the above.

The advantage of a multiple-effect distillation unit is that the brine concentration and flow can be varied for each effect. The high temperature effects are exposed to the lowest brine concentrations while blowdown concentrations can be higher. The multiple effect plant permits tailoring of the flashdown and flow rates in each effect to give the highest economy ratio. Multiple-effect plants usually are more costly due to the larger number of recycle pumps required; one for each effect, plus extra piping, controls and instrumentation. Commercial experience is limited to a few plants.

2. <u>Vapor Compression Process</u>

In vapor compression plants, distillation is similar to the multi-effect plant and usually employs two or three effects. the difference is that the steam from the lowest temperature effect is pumped by a vapor compressor to the first effect.

The compressed steam vapor is directed to the brine tubes containing the incoming brine.

The primary difference between this process and other distillation processes is in the method by which heat is added to the system. In other processes, brine is heated with steam in order to cause some of it to boil. In this process, however, heat is added to the vapor by converting mechanical energy into heat energy via compression.

Commercially, most of the vapor compression units are package units producing up to 125,000 gpd with high power consumption -- about 85 to 100 KWH per 1,000 gallons. Plant size is limited to the size of the vapor compressor available. The main disadvantage of this type of plant is the high degree of maintenance required for the vapor compressor.

3. Solar Humidification Process

The solar humidification process makes use of the fact that water will evaporate from a free surface even though the water is at a temperature below its boiling point. Under ideal solar transmission conditions approximately one pint of fresh water can be obtained daily for each square foot of surface that absorbs the solar energy. It is obvious that energy costs are minimal, but capital costs are in the order of 8 times those of the other processes because of the large solar collecting areas that are required. Maintenance costs have not been clearly established, but indications are that they are considerable. Experience is limited to a few small units (no systems in operation produce over a few 100 gallons per day), and long term economics have not been proven.

4. Electrodialysis Process

An electrodialysis conversion assembly is essentially a cell containing two different types of ion-selective membranes. One of the membrane types allow passage of positive ions or cations; the other allows passage of negative ions or anions. The cation-permeable membrane allows passage of the positive sodium ions, and the anion-permeable membrane allows passage of the negative chloride ions, leaving fresh water between the membranes. The electric current imposed on the electrodialysis cell provides the driving force for the ions. The amount of electric current required in a unit, which contains many sets of membranes between the electrodes, depends on the amount of dissolved salt to be removed. Therefore, the cost of the energy consumed in the process depends on the concentration of dissolved minerals in the feed water. As a result, the economics of electrodialysis are such that it is rarely considered for any water containing more than 3500 ppm TDS. Normal seawater contains salts in excess of ten times that amount. Research is currently being conducted to investigate the feasibility of operating the electrodialysis process at an elevated temperature of about $160^{\circ}F$ in order to reduce the high electrical energy required. High tempratures reduce the electrical resistance of the electrolyte and lower the electric power requirements. Seawater electrodialysis is under development and is not commercially available.

5. Freezing Process

It is well known that a salt solution cooled to its freezing temperature will deposit ice crystals of pure water. This principle forms the basis of desalting water by the freezing process. The application of freezing to the desalting of sea water involves partial freezing of a feed stream to an ice-brine slurry, separation of fine ice crystals from the brine, and melting of the ice as product water. Freezing is accomplished by flash evaporation of seawater at very low pressures (3mmHg) or by vaporizing a refrigerant such as butane in direct contact with the seawater. The theoretical advantage of the freezing process is it requires the lowest energy of any desalting process, which requires a phase-change; 144 BTU/lb compared to 970 BTU/lb for distillation. In addition, there is a minimum of corrosion and scaling because of the low temperature involved.

Although the freezing process has been under development for over 30 years, no commercial units are available. More development work is required in the areas of vapor compression and washing of the ice crystals before the process can be considered for commercial applications.

SECTION V

REVERSE OSMOSIS PLANT

Prepared by: Donald B. Guy

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SECTION V - THE REVERSE OSMOSIS PLANT

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FIGURES

SECTION V REVERSE OSMOSIS PLANT

A. Alternative Membrane and Power Production Processes

For the desalting plant, two membrane processes were considered. The first was reverse osmosis, and the second was high temperature electrodialysis process. In the standard electrodialysis process, the seawater feed is at ambient temperature, and the power consumption of the stack is from 2.2 to 2.7 kWh/1000 gallons/1000 mg/l TDS, or 75 to 93 kWh/1000 gallons of product in reducing the salinity from 35,000 mg/l to 500 milligrams per liter. The electrodialysis process does not require extensive pretreatment of the feedwater, since calcium precipitation is partially controlled by polarity reversing in the stack. Polarity reversal occurs automatically, three or four times per hour, reversing the electrical polarity of the stack and the product water, concentrate and electrode streams. The reversal reportedly purges the membrane and electrode surfaces of any scaling and fouling materials that may have been deposited.

The electrodialysis ion exchange membranes are chemically resistant, except to chlorine, and the membrane life is reported to be five to seven years. The electrode life is reported to be one to two years. The stacks can be manually disassembled for cleaning and repair.

Preheating the seawater with the solar gradient pond brine to 140-160^{o}F can drastically reduce the electrical requirements for the stack to 0.9 - 1.1 kWh/1000 gallons/1000 TDS, or about 35 kWh/1000 gallons. Seawater pumping would add several more kilowatt hours to the total. High temperature electrodialysis is not yet completely tested for commercial operation, although standard plants are operating at 110^{o}F.

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The reverse osmosis process option with power recovery has a base power consumption of 16.5 kWh/1000 gallons of product and, with all pumping included, a total consumption of about 22 kWh/1000 gallons, Table V-1.

The two module configurations commercially available for seawater desalting are the spiral wound element and the hollow fiber elements. Both are contained in pressure vessels, usually of fiberglass reinforced plastic material. The characteristics of each follow:*

Spiral Wound Module Characteristics

a) The pressure vessel can contain up to six membrane elements in series, resulting in lower pressure vessel to product water cost.

b) Have a productivity of 15 to 30 gallons per day of product per square foot of membrane area and 5100 gallons per day per cubic foot of element for brackish water.

c) Are the least prone to fouling.

d) Have dead-flow areas in the annulus between the element and vessel. These areas are prone to biological growths.

e) The spiral membrane leaves can be unrolled and examined for failure analysis.

f. O-ring seals in the center product tube and end caps can be nicked or roughened on assembly, causing concentrate leaks into the product water. The product tubes are

*D.B. Guy, S.V. Cabibbo, T.A. Ammerlaan, A. Ko and R. Singh, "Reverse Osmosis Technical Manual," Office of Water Research and Technology, U.S. Department of the Interior/Burns and Roe Industrial Services Corporation, NTIS Publication No. PB 80-186950 (July, 1979).

visual inspection. Membrane performance is uniform (±10 percent) if control is maintained on the various coating parameters.

p) Typical pressure drops from feed to reject are in the order of 10 psi per element. Product side parasitic pressure drops are said to be in the order of 30-40 psi, which lowers the effective driving force. Pressure drops are uniform from element to element, depending on the thickness of the spacer material, and element mass flow rate.

Hollow Fiber Module Characteristics

Characteristics of hollow fiber membrane modules are as follows:

a) The pressure vessel usually contains a single element, although development modules have contained two elements.

b) Have a productivity of 3 to 6 gallons per day of product per square foot of membrane area and 10,400 to 13,200 gallons per day per cubic foot of element.

c) The elements consequently are densely packed and are more prone to fouling. Hollow fibers may be 80 microns in diameter.

d) Have dead-end areas in annulus between product tube sheet and vessel which may encourage biological growth.

usually PVC or other plastic materials, and care must be taken in manufacture to assure good O-ring grooves.

g) The product water tube can be probed to locate leaks along the full length of the vessel.

h) The leading edge of the first element in series is the most prone location for fouling, biological degradation and washout.

i) Large membrane areas are required for the adhesive attaching membranes back-to-back in each leaf in some modules. These areas are non-productive.

j) Elements are prone to telescoping at high flow rates unless devices are installed for prevention.

k) Membrane defects can be patched before rolling if they can be located visually or with dye.

l) Flow appears to be well distributed throughout the membrane area. However, blockage by foreign material such as filter fibers can result in low flow areas and subsequent scaling.

m) Product recovery for each element is approximately 5 to 15 percent of the feed flow rate.

n) Element assembly at present is essentially semi-skilled hand labor and is subject to quality variation.

o) The membrane is machine-manufactured continuously in flat sheets with a width in excess of 40 inches. Membrane manufacturing defects can be located by

e) Membrane cannot be visually examined easily except by destruction.

f) The large O-ring on the outside diameter of the product tube sheet is a suspected source of internal leakage. It can become nicked on assembly with the pressure vessel when it is slid past the end plat retaining groove. The case can also be slightly out-of-round through storage and handling.

g) Large numbers of hollow fibers (10-20 percent) may be blocked by manufacturing operations on the face of the product tube sheet.

h) Large membrane areas are unproductive due to the extended fiber/epoxy laminate product tube sheet necessary to take the thrust from the 400 psi pressure drop between feed pressure and product pressure.

i) Broken fibers are said to be self-healing by collapsing. This may not be true adjacent to the tube sheets.

j) Flow distribution through the fiber bundle is not uniform and the areas adjacent to the tube sheets may have limited flow velocities.

k) Typical product recovery for each module is 50 to 60 percent of the feed flow rate.

l) Pressure drop appears to vary between modules of the hollow fiber PA type; hence, orifices or capillary tubes are inserted in the brine stream from parallel modules to equalize flow rates.

m) The pressure drop from feed to reject is in the order of 5 psi. The product water parasitic pressure drop is reported to be in excess of 200 psi for the hollow fiber element.

The polyamide (nylon) polymer hollow fiber membranes are relatively insensitive to pH extremes, permitting cleaning with strong acid or alkali solutions. They are very sensitive to oxydizing agents such as chlorine.

The DuPont B-10 permeator is the most reliable reverse osmosis module commercially available for single pass seawater desalting, at least at this point in membrane module development. It is recommended for this plant. Membrane life warranties have been extended to five years on a prorated basis by DuPont on acid using plants and by systems manufacturers on non-acid using plants. Spiral wound modules presently do not have such an extensive warranty.

For the power plant, three options were considered for driving the generator.

a) An organic Rankine cycle turbine system

b) A low pressure differential steam turbine system

c) An osmotic pump driving a Pelton wheel turbine.

d) A reverse electrodialysis system.

The first system is in commercial production and has been used with solar gradient ponds. The second is presently in the development stage. The third, the osmotic pump, would not use the solar gradient pond, but a source of concentrated brine such as is available on Sal Island. This option is presently a laboratory curiosity, though considered technically feasible. Preliminary calculations show that it would require an excessive quantity of evaporation to produce feed brine, even with the reject brine (4.4 percent solids) recirculated to the evaporation pond. The reverse electrodialysis system is also a laboratory curiosity. In this unit, alternate channels on each side of the ion exchange membranes will contain concentrated brine and seawater. Direct current power will be collected at the electrodes as the two streams slowly mixed.

Therefore, reverse osmosis was selected for the membrane plant and an organic Rankine cycle system for the power plant.

B. Reverse Osmosis Plant Description

The reverse osmosis plant delivering about 22.9 m^3/hr. of 440 mg/l total dissolved solids product water will be similar to that noted in Figure V-1. The product rate is somewhat less than that of the initial Sal Island plant. Twenty-five percent of the feed to the plant will be recovered as product water.

Seawater pretreatment will consist of chlorination at the seawell; coagulant injection, if required; roughing and polishing sand/anthracite filtration; antiscalant injection; one micron cartridge filtration; and the bisulfite dechlorination required for nylon membranes.

This pretreatment will be reviewed, based on the knowledge obtained about local seawater quality, from operation of the initial 27.5 m^3/hr plant on Sal Island. It is possible that the pretreatment process can be simplified.

The membrane desalting section will consist of three parallel trains to enable partial operation during periods when water requirements are reduced or when power production is reduced due to lower pond temperatures. Parallel train design also improves plant reliability.

The three high pressure, 850 psi, pumping sections will have positive displacement plunger pumps belted to TEFC, 1.15 SF motors having dual shaft extensions. The motor is belted to a pelton wheel power recovery turbine. The plunger pumps are extremely energy efficient; however, need careful pulsation dampening. Pelton wheel power recovery turbines are now available from the Hayward-Tyler Pump Company.

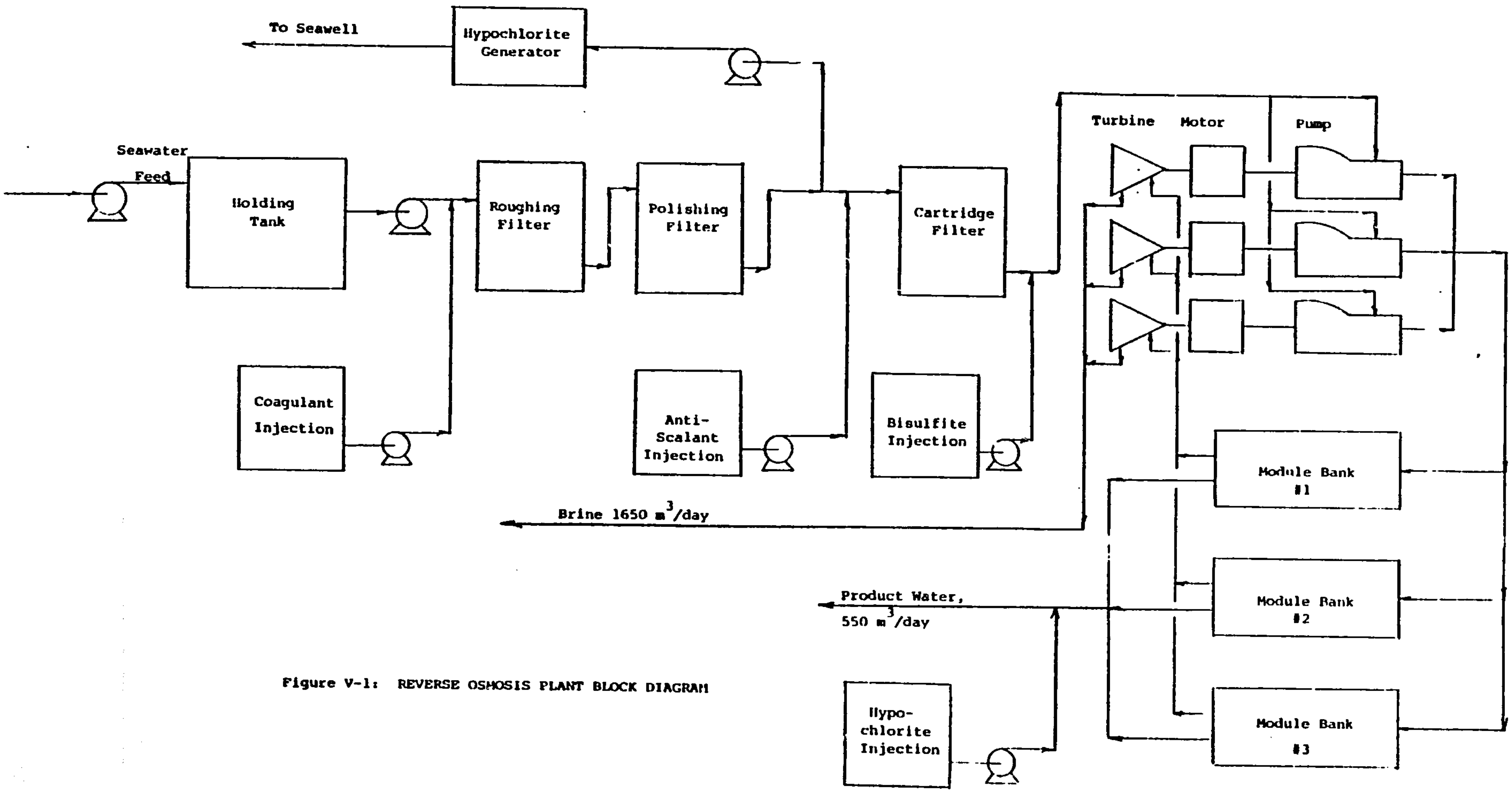

Figure V-1: REVERSE OSMOSIS PLANT BLOCK DIAGRAM

Membrane flushing during plant shutdown is accomplished by means of a product water tank atop each module bank. A cleaning and regeneration system for the module banks will be provided for use as required. The estimated use is every three months.

A hypochlorite system that generates sodium hypochlorite from seawater will be provided for injection into the seawater intake to control organic growths. A separate calcium hypochlorite injection system will be provided for control of bacteria in the product water.

The plant will not use sulfuric acid for calcium carbonate and sulfate control, but will use an antiscalant additive, such as Flocon-100 from Pfizer, Inc. This chemical will alleviate the handling problems of acid, particularly for a small, remote desalting plant.

The reverse osmosis modules will be 8-inch diameter Dupont B-10 Permaseps. At the present time, these are the most reliable modules available for single pass seawater desalting. Some systems manufacturers are extending the DuPont one-year warranty to five years, on a prorated basis. Seven-year lifetimes are reported.

Feedwater and product water diverting systems will be provided in event of pretreatment failure in the former and excessive product conductivity in the product water line.

Fail-safe instrumentation includes low feedwater pressure, excessive high pressure in the module banks and excessive feedwater temperature. The estimated electrical power requirements are shown in Table V-1.

TABLE V-1

REVERSE OSMOSIS DESALTING PLANT
ELECTRICAL POWER REQUIREMENTS

Description	kWh/1000 gallons of product
Seawater intake pumps	1.45
Boost pumps from storage tank	3.13
Filter surface wash pumps, intermittent use	Negligible
Filter backwash pumps, intermittent use	0.16
Filter air scour blower, intermittent use	0.02
Hypochlorite generator	0.33
High pressure pumps with energy recovery turbine	16.50
Cleaning pump (If used, at least one of the high pressure pumps will be off-line.)	Negligible
Chemical pumps and mixers	0.06
Desalting Plant Total	21.65 kWh/1000 gal

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A total of 21.65 kWh/1000 gallons is required from the power plant, or 130.5 kW, for the desalting plant. This assumes reliabile operation of the power recovery turbines. The pelton wheel is a classic turbine and, as such, is efficient (84 percent) and reliable.

However, currently there is little accumulated history of pelton wheel operation on seawater that has been published. A number of units are now being installed on reverse osmosis units, and reliability information will be forthcoming in the future.

The estimated consumables on an annual basis, based on 90 percent utilization, are shown in Table V-2.

The polyelectrolyte, used as a roughing filter aid, may not be required in the short term, if the seawater feed has low turbidity and silting index. Without the polyelectrolite, the roughing filter run time between backwash cycles may be extended. If the water quality is good, the roughing filters can be bypassed. However, they may be required at a later date, if the feedwater quality deteriorates or ferric iron is present in the feed.

The hypochlorite is used to aid in disinfection of the potable product water for health purposes.

The sodium bisulfite is needed for dechlorination, since the polyamide DuPont B-10 hollow fibers deteriorate in the presence of oxydizing agents. The formaldehyde is used to control organic growths in the modules during shutdown periods.

The tannic acid is a "regenerating" agent, used on polyamide membranes to recover monovalent ion rejection following cleaning.

TABLE V-2

ANNUAL CONSUMABLES FOR
REVERSE OSMOSIS DESALTING PLANT

Operating Items	Dosage	Quantity/year	Price	U.S. $/year*
Polyelectrolyte (if required)	3.0 ppm	2287 kg	$0.57/kg	1647
Sequestering agent (Pfizer Flowcon-100 or equivalent)	4.0 ppm	2320 liters	2.64/liter	6480
Sodium bisulfite	2.0 ppm	2168 kg	0.20/kg	782
Calcium hypochlorite (HTH or equivalent)	0.5 ppm	137 kg	0.90/kg	144
Cartridge filters (76.2 cm length)	twelve change per year	540	8.12 ea	4440
Pump hydraulic fluid	two changes/yr	379 liters	1.59/liter	672
Cleaning Chemicals				
Formaldehyde (37% active)	2 uses per year	161 liters	0.58/liter	118
Citric acid (commercial)	3 uses per year	233 kg	0.48/kg	147
Ammonium hydroxide (25% active)	3 uses per year	640 grams	0.54/kg	-
Tannic acid (commercial or DuPont PT-B)	3 uses per year	640 grams	3.40/kg	4
Enzyme detergent (BIZ or equivalent)	2 uses per year	26 kg	0.27/kg	11
			TOTAL	$14,445

*90% Utilization, $150/metric ton added (to U.S. costs) for shipping charge.

The approximate feedwater and product water ionic strength and the percent rejection of each ion is shown in Table V-3 for 25 percent product recovery.

It can be noted that the total dissolved solids (TDS) of the product is approximately 445 mg/l, a reasonable quality potable water, for a single stage system. The product water has a hardness of 32 mg/l as calcium carbonate, and therefore is extremely soft for household, commercial and industrial uses. The percent rejection based on the feed ions is calculated by the feed ion concentration minus the product ion concentration, divided by the feed ion concentration. For this membrane system it can be noted that the divalent ions have 99.5 percent rejection, while the monovalent ions are rejected by 98.6 to 98.8 percent. This difference is more vividly displayed in the last column, Reduction Ratio. This is the ratio of the feed ion concentration to the product ion concentration. For example, calcium is reduced in the product by a factor of 171 to 1. Using this method, the monovalent ions are reduced from 72-82 ions to 1 in the product, while the divalent ions are reduced from 171-199 to 1, over a factor of two times as great.

The facilities required are:

a. A concrete pad of 8000 mm x 20,650 mm in size for the pretreatment equipment.
b. A shelter of 6000 x 4000 mm in size for the chemical feed tanks and pumps and the hypochlorite generator.
c. A shelter of 10,500 x 13,500 mm for the module banks, pump skids, control panel and shops.
d. A raw water holding tank of 15,000 gallons to provide a total of 40 minutes of time for chlorine reaction.

TABLE V-3

APPROXIMATE FEEDWATER AND PRODUCT WATER
COMPOSITION AT 25 PERCENT PRODUCT RECOVERY

Ion and Physical Characteristic	Seawater	Product Water	Percent Rejection	Reduction Ratio
Calcium (as mg/l Ca)	341	2	99.4	171:1
Magnesium (as mg)	1,323	7	99.5	189:1
Sodium (as Na)	11,190	156	98.6	72:1
Potassium (as K)	489	6	98.8	82:1
Bicarbonate (as HCO_3)	137	4	97.1	34:1
Sulfate (as SO_4)	2,786	14	99.5	199:1
Chloride (as Cl)	20,026	256	98.7	78:1
Fluoride (as F)	0.4	0.0	-	-
Silica (as SiO_2)	8.0	0.2	97.5	40:1
Iron (as Fe^3)	0.1	0.0	-	-
Total Dissolved Solids (as mg/l)	36,300	445	98.8	82:1
Chlorine (residual Cl_2)	1.0	0.5	-	-
Carbon Dioxide as CO_2)	7.1	7.1	-	-
Hardness (as $CaCO_3$)	6,277	32	99.5	196:1
pH (units)	7.5	6.0	-	-

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C. Power Plant Description

Two active manufacturers of packaged power plant utilizing organic Rankine cycle turbines were located. One of the companies is Ormat Turbines, Ltd. of Yavne, Israel, and the other is AFI Energy Systems, Livingston, NJ, a partnership formed by Foster Wheeler Energy Systems and Harima Heavy Industries Company, Ltd. of Tokyo, Japan.

The current prices being quoted are as follows:

Capacity, kW	US$/kW	Manufacturer
150	4,000	Ormat Turbines, Ltd.
300	2,800	Ormat Turbines, Ltd.
500	2,000-2,500	AFI Energy Systems

AFI Energy Systems is presently not offering 150 kW systems, leaving the field to Ormat, Ltd. According to the literature provided, Ormat has over 2,500 Rankine Cycle Turbogenerators in operation in over forty countries on waste heat power generation applications.

A schematic of the cycle major components is contained in Figure V-2. The hot 190°F brine is circulated through the evaporator (vaporizer) and recirculated back to the bottom of the pond for reheating. The expanding vaporized organic fluid drives the turbine. Some of the fluids used are listed in Table V-4, together with their Figure of Merit in MMBtu/hr/ft^2 *. These are for 200°F evaporation and 104°F condensing temperatures. These figures cannot be used for equipment design, but they do allow a valid comparison

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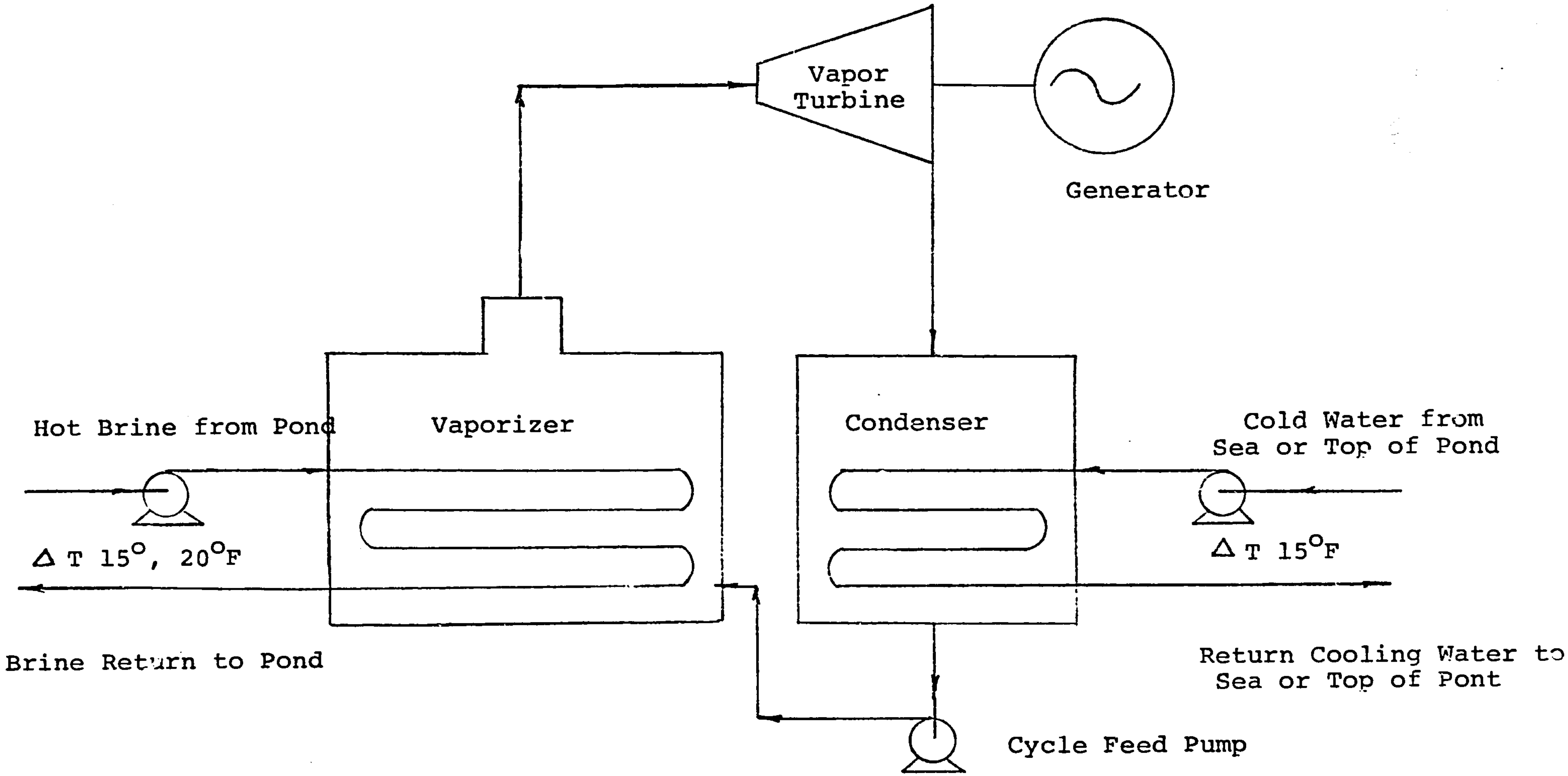

Figure V-2: Organic Rankine Cycle Power Production Schematic

TABLE V-4

HEAT CAPACITY FIGURES OF MERIT FOR VARIOUS FLUIDS[1]

Description	Fluid	Figure of Merit MMBtu/hr/ft^2 Inlet	Figure of Merit MMBtu/hr/ft^2 Outlet
Fluorocarbon	R-11	301	72
Fluorocarbon	R-21	621	148
Fluorocarbon	R-114	513	123
Fluorocarbon	R-133	613	138
Propane	C_3H_8	2237	438
n Butane	C_4H_{10} (n)	714	177
iso Butane	C_4H_{10} (iso)	909	231
Water	H_2O	153	14.7

1: Evaporation temperature, 200°F.
Condensing temperature, 104°F.

*R.H. Sawyer, and S. Ichikawa, "The Organic Rankine Cycle System, Its Application to Extract Energy from Low Temperature Waste Heat," AFI Energy Systems, a paper presented at the Second Annual Conference on Energy Conservation and Technology, Houston, TX, 1980.

For a given combination of evaporator and condensing temperatures, each working fluid will result in different rankine-cycle efficiencies. It is possible to determine the fluid (among the candidates) providing maximum efficiency. On the other hand, fluid characteristics, such as specific volume and heat capacity will determine the size (and, thus, cost) of equipment and piping. To compare the working fluids for their ability to carry heat per unit of flow area, a relative figure of merit may be established for the operating temperature levels. The values shown on the above table allow a comparison of sizes; larger values predict smaller expanders and transport line sizes and costs.

of sizes. Larger values predict smaller expanders, line sizes, and cost.

The cooling water can either be from the top of the pond or from the seawater intake system. Some advantage can be obtained by using seawater and then injecting a portion of the heated seawater into the desalting plant feedwater to maintain a year-round seawater temperature of 25° Celsius. It may also help in preventing instabilities in the solar gradient pond. The condensed fluid is then recirculated to the evaporator by pumping.

The units are skid mounted with the generator to ease installation problems at the site. Ormat reports that the overall efficiency is 7.5 percent, which should include the small feed pump.

Figure V-3 is an end view of a 300 kW Ormat unit. The condenser would appear to be on the upper left, with the vaporizer on the right side. The installation does not require protection, except for the generator.

In size, the plant would appear to be 20 feet in width by 30 feet in length and 25 feet in height.

Figure V-4 shows a 300 kW unit installed at a solar gradient pond.

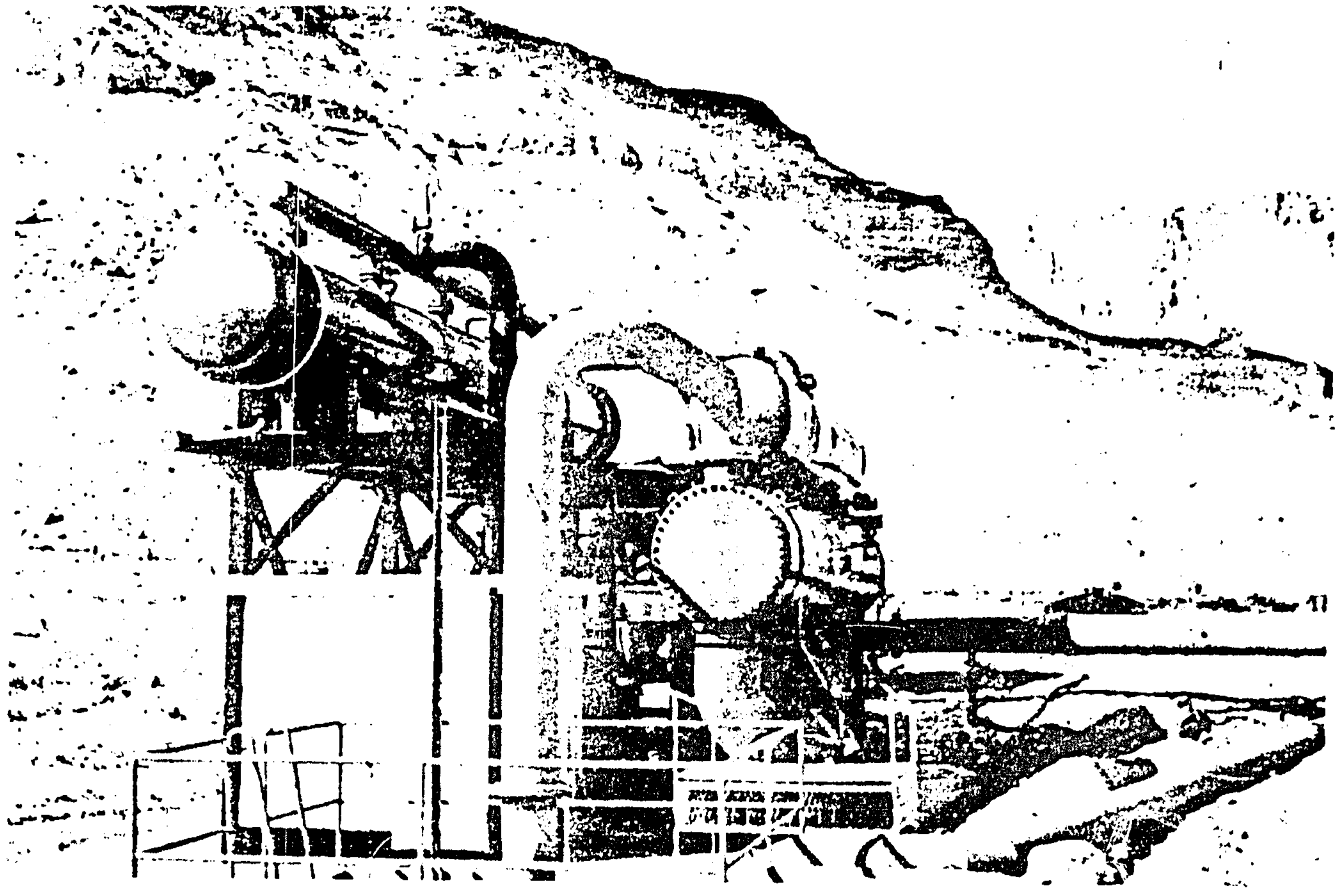

Figure V-3: View of 300 kW Power Station

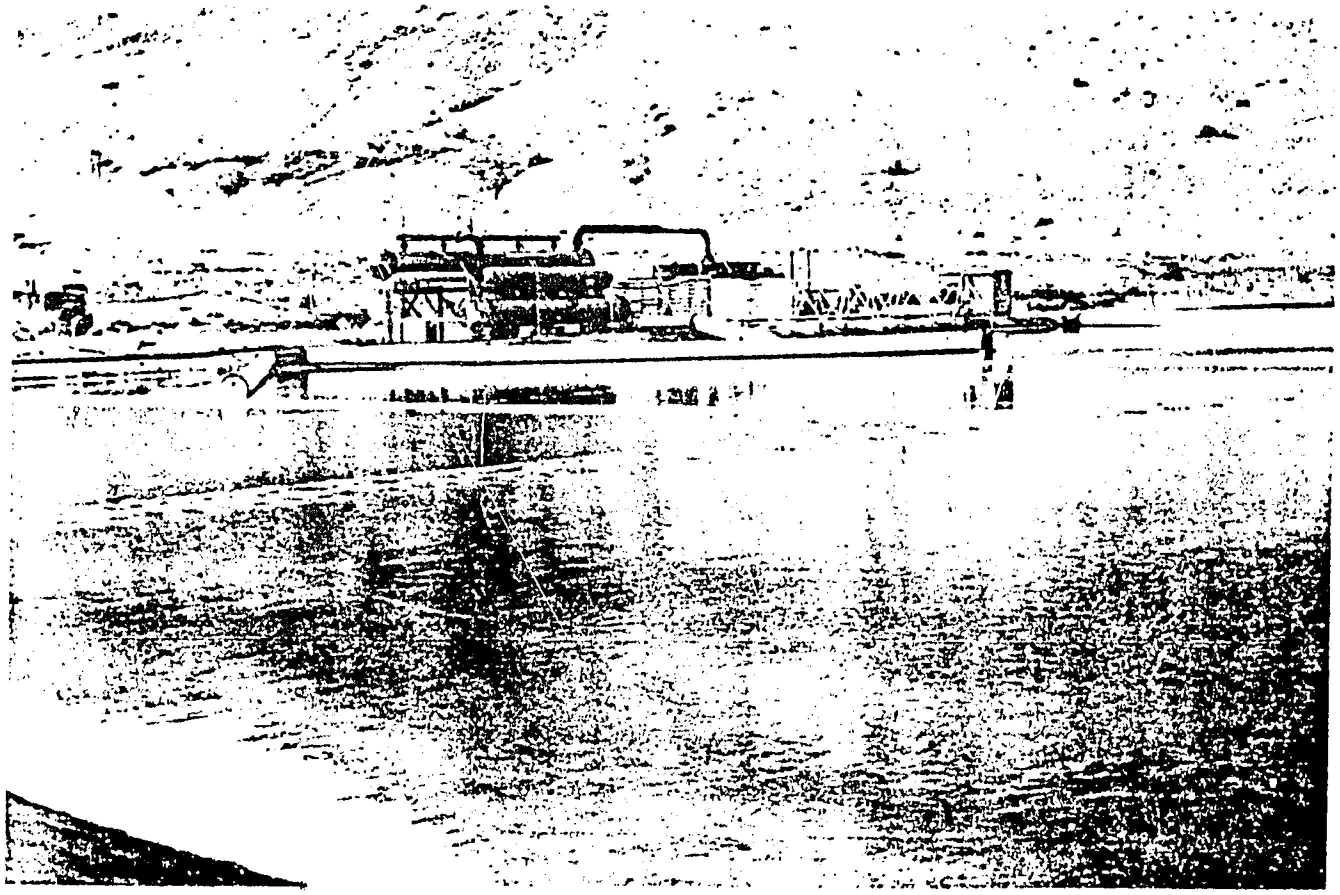

Figure V-4: 300 kW Solar Pond Power Plant

Source: Ormat Turbines, Ltd. Brochure

D. Salt Gradient Solar Pond Sizing and Power Requirements

The size of the solar gradient pond and the hot brine and cooling water pumping requirements, pump sizes and the heat exchanger will be different for the MSF and reverse osmosis plant options due to the variation in power requirements; however, design details will be similar, as noted in Section III. In all cases, the plant will be tied into the local power network for startup of the pond pumps.

Six cases were examined in some detail, involving differences in power generation capacity and in temperature difference between the incoming hot brine to the vaporizer from the pond and the brine return from the vaporizer.

The cases are:

Case 1, Base Case, 133.5 kW generated for the desalting plant only and a $15^{o}F$ delta T for the turbogenerator vaporizer.

Case 2, 150 KW generated for the desalting plant and partial power for the pond pumps, and a $15^{o}F$ delta T for the vaporizer.

Case 3, Full power generated for the desalting plant and for the pond pumps, and a $15^{o}F$ delta T for the vaporizer.

Case 4, Base case, 133.5 kW generated for the desalting plant only and an optimistic $20^{o}F$ delta T for the vaporizer.

Case 5, 150 kW generated for the desalting plant and partial power for the pond pumps, and 20^{o} F delta T for the vaporizer.

Case 6, Full power generated for the desalting plant and for the pond pumps, and a 20^{o}F delta T for the vaporizer.

Other factors were held constant:

a) An overall efficiency of 7.5 percent for the organic Rankine cycle turbogenerator.

b) Bottom of pond sizing at 128 Btu/h/m^2.

c) The desalting plant energy requirement of 4.5461 X 10^5 Btu/h; base case pond capacity requirement of 6.0615 x 10^{6}Btu/h (133.5 kW); base case pond bottom surface area of 47,335 m^2.

d) For pond brine, C_p = 0.8 and C_d 1.27 in flow rate and power calculations.

e) For cooling water (delta T 15^{o}F), C_p = 0.975 and C_d = 1.05 in flow rate and power calculations.

f) A pond pump head of 70 feet, and pump efficiency of 60 percent.

g) The miscellaneous Btu losses in the turbogenerator system at 10 percent of the net Btu's (Btuin - Btu of work).

The results of these calculations are shown in Table V-5, which lists the case number, the pond pumping power in kw

TABLE V-5

SOLAR POND SIZING AND POWER REQUIREMENTS

Case	Pumping Power, kW			Generator, kW	Outside* Power, kW	Pond Size, m^2	Acres
	Brine	Cooling Water	Total				
1	27.8	15.7	43.5	133.5	43.5	47,355	11.7
2	31.4	17.7	49.1	150	32.6	53,363	13.2
3	38.2	21.6	59.8	193.3	0	64,960	16.1
4	20.9	15.7	36.6	133.5	36.6	47,355	11.7
5	23.5	17.7	41.2	150	24.7	53,363	13.2
6	27.4	20.6	48	174.7	0	62,149	15.4

* After startup.

for the hot brine and the cooling water, the generator output in kW, the outside power required in kW following startup of the turbogenerator, and the pond size is in m^2 and acres.

It can be calculated that the pond pumping power varies from 21.5 percent to 32.7 percent of the total power, or from 6 kWh/1000 gallons to 9.9 kWh/1000 gallons of product water. This increases the total power required from 22 to 27:7 to 31.5 kWh/1000 gallons of product. The inefficiency of the turbogenerator (7.5 percent) is the major contributor.

The power required from the network varies from nothing to 43.5 kW.

The pond size varies from 47,355 m^2 to 64,960 m^2, depending on the cost and how much work the pond will do.

The first three cases are the most probable for selection of the prime candidate process.

E. REVERSE OSMOSIS POWER AND DESALTING PLANT COSTS

For comparative purposes, Cases 1, 2 and 3 will be costed, as noted in Table V-5. These have outside network power assist of 43.5, 33.8 and zero kilowatts, respectively. Ormat Rankine Cycle turbogenerators of 133.5, 150 (standard) and 193.3 kW at $4,000 and $3,600 per kW will be installed, respectively. 30% of equipment cost has been added to achieve the installed costs. The pond sizes will be 47,355, 53,363 and 64,960 square meters.

The desalting plant of 550 m^3 capacity is estimated at U.S. $884,500 based on vendor quotations. With a 15 percent contingency and escalation of U.S. $132,675, the cost is U.S. $1,017,175. This price includes packaging for overseas shipment. Shipping charges to a port warehouse and spare parts would add about U.S. $50,000. Installation costs with pads, building and raw water storage tanks are estimated at $90,000 plus 30% of the basic plant cost, or a total of $304,925. The total capital cost of the desalting plant is $1,462,100.

The tabulated pond costs and the capital and operating costs are in Tables V-6 and V-7. The solar gradient pond cost is shown in Table V-6. Increasing the generator capacity increases the pond size, the pond flow rates and power and also pond costs. Electricity costs would have to escalate from current 16.5¢/kWh to about 22.5¢/kWh in order to make the full power production pond competitive in annual costs.

Table V-7 contains capital and annual costs for the pond, power production and desalting plant.

The total capital costs range from $4,450,274 for the smaller pond to $5,317,460 for the full power pond. Annual

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TABLE V-6

SOLAR GRADIENT POND COSTS FOR THE REVERSE OSMOSIS PLANT*

Item	Case 1	Case 2	Case 3
Generator Capacity, kW	133.5	150	193.3
Outside Network Power, kW	43.5	33.8	0
Bottom Surface Area, m^2	47,355	53,363	64,950
Heat to Turbine Vaporizer, Btu's/hour x 10^6	6.0615	6.8304	8.3149
Total Pond Flow, gpm	1701	1917	2333
Capital Costs, U.S.$			
Pond Construction (incl.liner)	786,093	880,489	1,061,932
Salt Cost, Initial	611,158	688,696	838,237
Piping and Pump Cost	539,207	579,300	651,752
Total Pond Cost	$1,936,458	$2,143,485	$2,551,921
FCR @ 10 percent	193,646	214,848	255,192
Annual Operating Cost, U.S.$**			
Electricity, 16.5¢/kWh	56,587	43,969	0
Replacement Salt, $22.15/MT	8,925	10,058	12,242
Annual Operating Cost	$65,512	$54,027	$12,242
Total Annual Costs, U.S.$	$259,158	$268,875	$267,434
Cost of Pond Heat $/$10^6$ Btu	4.24	4.18	4.08

* 3-meter depth

** 90 percent utilization

TABLE V-7

REVERSE OSMOSIS PLANT CAPITAL AND OPERATING COSTS

Capital Costs, U.S.$*	Case 1	Case 2	Case 3
Solar Pond	1,936,458	2,148,485	2,551,921
Seawater Intake	249,616	261,382	290,895
Ormat Turbogenerator & electricals	802,100	887,900	1,012,544
Reverse Osmosis Plant	1,462,100	1,462,100	1,462,100
Total Capital Costs	$4,450,274	4,759,867	5,317,460
FCR @ 10 percent	445,027	475,987	531,746
Annual Operating Costs, U.S.$**			
Replacement Salt	8,925	10,058	12,242
Chemicals and Supplies	14,445	14,445	14,445
Electricity, 16.5¢/kWh	56,587	43,969	0
Labor	70,143	70,143	70,143
Annual Operating Costs	$150,100	$138,615	$ 96,830
Total Annual Costs, U.S.$	$595,127	614,602	628,576
U.S. Dollars/1000 gallons	12.47	12.87	13.17
U.S. Dollars/m^3	$ 3.29	3.40	3.49

*Installed costs

**90 percent utilization

costs vary from $595,127 to $628,526, or from $12.47/1000 gallons of product to $13.17/1000 gallons.

SECTION VI

COMPARISON OF THE MSF AND REVERSE OSMOSIS PROCESSES

Prepared by: Donald B. Guy

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SECTION VI - COMPARISON OF THE MSF AND REVERSE OSMOSIS PROCESSES

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VI. COMPARISON OF THE MSF AND REVERSE OSMOSIS PROCESSES

The selection of the prime candidate process is based upon the comparative annual cost, process advantages and disadvantages, suitability for the location, and miscellaneous factors such as intangibles, space requirements, etc.

A. Comparative Annual Costs

The comparative capital and annual costs are given for the MSF plant with a performance ratio of 8 and the reverse osmosis plant, Case 2, having a turbogenerator of 150 kW capability in Table VI-1. In both plants power is used from an outside network to supplement the available energy from the pond.

The MSF plant capital costs are $3,993,000, about 16 percent less than those of the reverse osmosis plant with the 150 kW turbogenerator.

In annual operating costs the reverse osmosis plant at $138,615 is 60 percent less than the MSF plant. The major cost difference is in the use of external power for all the plant pumps on the MSF plant.

In total annual costs, including fixed charges at 10 percent of total capital costs, the reverse osmosis plant at $614,602 is 18.2 percent less than that of the MSF plant.

Table VI-2 shows the relatively high cost of producing power by means of the solar gradient pond/turbogenerator combination. A comparison is given of the capital and annual costs for the MSF and reverse osmosis plants with full power generated by the solar ponds. For startup of the ponds, network power will still be required.

TABLE VI-1

COMPARATIVE COSTS OF MSF AND REVERSE OSMOSIS DESALTING PLANTS WITH SOLAR GRADIENT POND AND PARTIAL EXTERNAL POWER

CAPITAL COSTS, U.S.$	MSF PLANT PR of 8	REVERSE OSMOSIS PLANT, 150 kW
Solar Salt Gradient Pond	1,827,000	2,148,485
Seawater Intake	241,000	261,382
Desalting Plant	1,877,000	1,462,100
Turbogenerator & Electricals	-	887,900
Strainer, Deaerator	48,000	-
Total Capital Costs	3,993,000	4,759,867
CR, 10%	399,300	475,987
ANNUAL OPERATING COSTS, U.S.$		
Replacement Salt	9,365	10,058
Chemicals and Supplies	10,694	14,445
Electricity, $0.165 kW	261,083	43,969
Labor	70,143	70,143
Annual Operating Costs	351,285	138,615
Total Annual Costs, U.S.$	750,585	614,602
Cost in $/1000 gallons of water	$15.72	$12.87
Cost in $/m^3 of water	$ 4.15	$ 3.40

TABLE VI-2

COMPARATIVE COSTS OF MSF AND REVERSE OSMOSIS DESALTING PLANTS WITH THE SOLAR GRADIENT PONDS CAPABLE OF FULL POWER GENERATION

CAPITAL COSTS, U.S.$	MSF PLANT PR of 8	REVERSE OSMOSIS PLANT, 193.3 kW
Solar Salt Gradient Ponds and Turbogenerator	5,350,111	3,564,465
Seawater Intake	241,000	290,895
Desalting Plant	1,877,000	1,462,100
Strainer, Deaerator	48,000	-
Total Capital Costs	7,516,000	5,317,460
FCR, 10%	751,611	531,746
ANNUAL OPERATING COSTS, U.S.$		
Replacement Salt	23,401	12,242
Chemicals and Supplies	10,694	14,445
Electricity, $0.165/kWh	-	-
Labor	70,143	70,143
Annual Operating Costs	104,238	96,830
Total Annual Costs, U.S.$	855,849	628,576
Cost in $/1000 gallons of water	$17.93	$13.17
Cost in $/m^3 of water	$4.74	$ 3.49

It can be seen that the capital costs of the reverse osmosis plant at $5,317,460 are 30 percent less than the MSF plant due to the cost of the additional pond area and turbo-generator for the MSF plant. The annual operating cost of the reverse osmosis plant at about $100,000 is quite nominal. The cost in $/1000 gallons of potable water is $17.93 and $13.17, respectively, for the MSF and reverse osmosis plant.

The reverse osmosis plant total annual costs are $135,983 less than the MSF plant for the plants in Table VI-1 having less than full power production, but are $227,273 less for the plants in Table VI-2 having full power production from the salt gradient ponds.

B. Comparative Process Advantages

The advantages of the reverse osmosis plant include the following:

- Faster startup and shutdown
- Similarity to the initial Sal Island Plant
- Lower occurence of corrosion with desalting being at ambient temperatures
- Ambient temperature product water
- Product water is less corrosive than distilling plant water, without treatment.
- Does not operate under vacuum conditions, so leaks are external fluid, rather than internal air leaks, which may be difficult to locate.
- Somewhat more simple to operate, since inter-stage adjustments are not necessary.
- Has theoretical and actual advantages in power consumption.

- Has started to make inroads on thermo plant sales, particularly in small sizes
- Is not vulnerable to air in the seawater feed.

The advantages of the MSF process include the following:

- Is the most proven and established process
- The product water is of high quality (distilled)
- It may be more reliable.
- It is not vulnerable to oxidation by chlorine, or by bacterial sliming.
- Plants can be descaled by strong chemicals, or by ball cleaning of evaporator tubes.
- Plants with adequate operating procedures have run more than twenty years.
- All the large desalting plants use the multistage flash (MSF) process.
- Are widely used in cogeneration plants producing power and desalted seawater.

The last two advantages listed for the MSF process are not applicable to this plant.

C. Suitability for this Installation

On the basis of these characteristics, there is no clear advantage to either the reverse osmosis or MSF plant. The small size of the plant and the reduced power requirements tend to favor the reverse osmosis plant selection for this installation.

A third advantage is that the initial reverse osmosis plant could provide interchangeable operators and spare parts.

D. Other Comparative Factors

The MSF plant does not require protective shelter, while the module banks do, to protect from heat during shutdown. The MSF plant does not require the variety of chemicals or pretreatment equipment that the reverse osmosis plant does. on the other hand, the startup time for the MSF plant would be extensive, due to the low temperature of the brine, compared to boiler type MSF plants. The reverse osmosis plant shipping weight and cost would be less than the MSF plant. Environmentally, there would be limited differences between the two plants. The MSF product water can be mixed with the available brackish water to provide a potable blend with reduced corrosive quality.

E. Summary and Conclusions

The comparative annual cost analysis strongly favors the Reverse Osmosis Plant, whether partial or full power production is selected.

The comparative process advantages and the suitability for installation on Sal Island favors the reverse osmosis plant to a lesser extent.

Analysis of other minor comparative factors tends to favor the MSF plant.

In conclusion, the reverse osmosis plant would be the prime candidate choice, although either plant configuration is acceptable.

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F. Technical Comparison of Solar Gradient Ponds as An Energy Source to Diesel and Wind-Driven Generators

This comparison is provided as required by the Statement of Work. Among other considerations, the long term effects of the arid Sahel-belt environment, semi-skilled labor, and limited availability of materials and supplies, the comparison items can be summarized as follows:

Solar Gradient Pond

1. Utilizes a local salt source.
2. Has a renewable energy source.
3. Is relatively insensitive to settling of windblown sand.
4. Does not produce noise or emissions to pollute the atmosphere.
5. Operating labor skills are low.
6. Has relatively low efficiency, increasing capital costs.
7. Has low demand for imported materials and supplies.
8. Has energy storage capacity.
9. Maintenance requirements are low due to few moving parts.
10. The excavation has an extended life, liners, where used, have a reported life of fifteen years.
11. Does not corrode.
12. Does not require shutdown and startup for demand changes.
13. Requires larger land areas for installation.

Diesel Generators

1. Utilizes imported fuel.
2. The energy source is not renewable.
3. As an air breather, it can be sensitive to windblown sand and silt.
4. It produces noise, odors and smoke as pollutants.

5. The local labor force is familiar with the diesel units.
6. It has a relative high efficiency.
7. It is a classic, well-developed machine.
8. It has high demand for imported parts.
9. It has no energy storage capacity.
10. Maintenance requirements may be high, due to numerous moving parts.
11. It requires small space, but protection from the weather.

Wind Generators

1. Has a renewable energy source.
2. Must be located in advantageous areas.
3. Wind power is greatly susceptible to wind velocity changes.
4. Cannot store energy, except in external batteries.
5. Is vulnerable to storms, hurricanes, and sea gulls.
6. Is subject to corrosion and blade pitting by airborne particles.
7. It requires relatively less space for installation.
8. It can produce either mechanical or electrical power.
9. Operation and maintenance labor is low.
10. It utilizes imported parts.

Conclusions

It would appear that, from a technical viewpoint, each site would have to be evaluated for determination of the most suitable energy source or sources. In the case of Sal Island, both the solar and wind potentials exist.

SECTION VII

ECONOMIC AND FINANCIAL ANALYSIS OF THE MOST PROMISING SYSTEM

Prepared by: Amit Chattopadhay

SECTION VII - ECONOMIC & FINANCIAL ANALYSIS OF THE MOST PROMISING SYSTEM

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VII. ECONOMIC & FINANCIAL ANALYSIS OF THE MOST PROMISING SYSTEM

A. Introduction

In this section an economic and financial analysis of the most promising system to be powered by a solar pond has been performed. The results of the economic and financial analysis for the solar powered system has been compared with the results of an economic and financial analysis for diesel and wind powered systems.

B. Cases Considered in the Analysis

There are four cases considered in this analysis. Based on the study presented in the prior sections, there are two candidates for the solar powered system. The third and fourth ones are diesel-powered and wind-power supplemental cases. These cases are identified as follows:

Case i) Reverse osmosis desalination with part power generated by solar pond and part purchased power.

Case ii) Reverse osmosis desalination with full power generated by solar pond.

Case iii) Reverse osmosis desalination with diesel power generation.
(Relevant data extracted from the project documents titled "Cape Verde Desalination and Power (Sal) Project," AID Project No. 655-0005.)

Case iv) Reverse osmosis desalination with diesel power supplemented by wind-power to a maximum practical extent.
(Relevant data extracted from the study "Wind Power Generation for Sal Island", dated August, 1980)

C. Assumptions and Considerations

i) The analysis for each case results in determining the net present value over the life of the system and calculating internal rate of return (IRR).

ii) For each case, net present values are calculated at varying discount rates and then IRR is determined by interpolation. See Figure VII-1.

iii) Costs and incomes are presented in terms of constant purchasing power, rather than real dollar values over the life of the project. Price escalation to take into effect of differential cost growth for applicable items have been considered. Escalation has been accounted for:

a) Diesel fuel - an annual cost growth of 3.5 percent has been considered. (NOTE: Within an expected range between 2% and 5%, an average value of 3.5% is adopted.) Current price is taken as $460 per metric ton.

b) Purchased power - Current average cost of electricity on Sal Island is 16.5 cents per Kwh. It is assumed that 75% of this cost is attributable to fuel cost and thus adjusted accordingly over the years. It is calculated that at the end of 20 years cost of purchased power would rise to approximately 29 cents per Kwh. (See Note at the bottom of this page.)

c) Subsidy removal - (also refer to sub-section D "Shadow pricing"). It is considered that 70% of the current subsidy of $234,000* per year is attributable to fuel cost and thus adjusted accordingly over the years of operation.

NOTE: Many countries in Africa sell electricity at much higher price. Current prices have been reported to exceed 25 cents/Kwh in some places.

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*Source: AID Project Paper "Cape Verde - Desalination and Power (SAL)" Vol. I, pg. 78.

iv) A 20-year operating project life has been considered for all cases. A salvage value* has been considered after end of project life. A two-year period has been allocated for engineering, procurement and construction.

v) A 10% cost has been allowed for engineering and administration to determine final capital cost for each case.

vi) Recent labor cost data for Sal Island has been utilized for estimating operating labor costs as well as construction cost for the pond.

vii) The following categories of costs have been included in the operation and maintenance costs:

a) Operating labor
b) Maintenance parts
c) Operating chemicals
d) Fuel cost (when applicable)
e) Major overhauls and replacements:

- replacement membranes for R.O. system
- replacement of pond liner material
- overhaul of diesel engines

viii) A utilization factor of 90 percent has been considered.

ix) Costs of land and rights of way have been neglected.

D. Shadow Pricing

Shadow pricing of the possible uses of the fresh water produced has been considered in the following manner.

Since the sales prices of fresh water are subsidized, the use of sales price alone to determine income for economic analysis will be erroneous. "Shadow pricing" or true cost should be used for comparison purposes. Due to the fact

*Salvage value considers the whole of pond construction and salt costs, part of liner life, and remaining life of project components.

that the present installed water plants are significantly smaller than the one proposed in this study, care should also be taken not to overestimate the shadow price. Thus, the current cost of production per m^3 on Sal Island alone has not been used. Instead, the following logic is taken into account:

i) Sales price information:

Current average sales price in Mindelo : $0.70/$m^3$ (subsidized)

Current domestic sales price on Sal Island : $2.50/$m^3$ (subsidized)

Current non-domestic sales price on Sal Island : $3.00/$m^3$ (subsidized)

Proposed average sales price on Sal Island : *$2.00/$m^3$ (unsubsidized)

(Ref: "Cost-benefit study for the Initial Phase approach", dated March, 1980)

ii) Production cost information:

Current production cost at Mindelo old plant : $7.00/$m^3$
Expected production cost at Mindelo new plant: $3.50/$m^3$
Current production cost on Sal Island : $11.00/$m^3$

iii) Adopted data:

Sales price : $2.00/$m^3$
Removal of current subsidy of $254,000/year, regarded as income. Which signifies a current shadow price of <u>$3.29/$m^3$</u>

E. <u>Financial and Economic Analysis</u>

i) Case i) R.O. plant with 150 KW solar-pond powered turbo-generator, with balance purchased power.

*This is based on an average water charge of $2.50/$m^3$ less 50 cents/m^3 to compensate for trucking and/or distribution piping amortization costs.

a) Total capital cost for this system is estimated as 5,166,927 and has been used for IRR calculation. Breakdown of this cost is shown on Table VII-1.

b) Year by year net annual income has been calculated by taking the difference of revenues (shadow price) and costs (investment and O&M) as shown on Table VII-2.

c) Present values of year by year net annual income has been calculated for 2%, 5% and 7% discount rates as shown shown on Table VII-3. Interpolation shows rate of return of 6.2% as shown on curve A) on Figure VII-1.

ii) Case ii) R.O. plant with full-power turbine generator using solar pond.

a) Total capital cost considered in the analysis is $5,765,382. Breakdown of this cost is shown on Table VII-4.

b) Year by year net annual income has been calculated by taking the difference of revenues (shadow price) and costs (investment and O&M) as shown on Table VII-5.

c) Present values of year by year annual income has been calculated for 2%, 5%, and 7% discount rates as shown on Table VII-6. Upon interpolation, IRR is found as 5.8%, as shown on curve B, on Figure VII-1.

iii) Case iii) R.O. plant with diesel-generated power.

In this case, no solar pond is used. Conventional diesel-generator produces the required electrical power to run the R.O. plant.

TABLE VII-1

CAPITAL COSTS

(R.O. Plant with 150 KW Turbine-Generator)

I.	Solar Pond:	
	Pond Construction with liner	$ 880,489
	Initial salt cost	688,696
	Pumps, piping & appurtenances	579,300
	Total, Solar Pond	$ 2,148,485
II.	Seawater intake, with pumps & piping, installed	261,382
III.	Power Generator (Ormat turbine-generator)	600,000
IV.	Switchgear and electricals	83,000
V.	Reverse Osmosis Plant, including pretreatment and initial spares	1,067,000
VI.	Plant building; storage tanks, equipment pads	90,000
VII.	Erection & installation @30% of equipment cost	510,000
VIII.	Engineering & Administration @10%	407,060
	Total Capital Cost	$ 5,166,927

TABLE VII-2

CALCULATION OF NET "SHADOW" INCOME

REVERSE OSMOSIS PLANT, 150 KW, TURBO-GENERATOR

(IN THOUSANDS $)

		Costs		Revenues		
Elapsed Year	Operating Year	Investment	Operation & Maintenance	Water Sales	Subsidy Removal	Net Income
(1)	(2)	(3)	(4)	(5)	(6)	(7)
0						
1		2,583.4				(2,583.4)
2		2,583.5				(2,583.5)
3	1		148.9	361.4	250.0	462.5
4	2		188.9		256.3	428.8
5	3		196.8		262.8	427.4
6	4		225.3		269.6	405.7
7	5		249.0		276.6	389.0
8	6		227.8		283.8	416.9
9	7		229.1		290.2	422.5
10	8		349.0		296.8	309.2
11	9		231.9		304.7	434.2
12	10		255.8		312.9	418.5
13	11		234.9		321.4	447.9
14	12		236.4		330.2	455.2
15	13		238.1		339.3	462.6
16	14		239.7		348.7	470.4
17	15		805.1		358.5	(85.2)
18	16		361.8		368.6	368.2
19	17		245.1		379.0	495.3
20	18		247.1		389.8	504.1
21	19		249.1		401.0	513.3
22	20		273.6		412.6	500.4
23	21	(1822.3)	0	0	0	1822.3
						Σ 4,902.3

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TABLE VII-3

CALCULATION OF INTERNAL RATE OF RETURN

REVERSE OSMOSIS PLANT, 150 KW TURBO-GENERATOR

Elapsed Year	Operating Year	P.W. Factors at Given Discount Rates			Present Value of Net Annual Income at Given Discount Rates		
(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
		2%	5%	7%	2%	5%	7%
1	-	1.0000	1.0000	1.0000	(2583.4)	(2583.4)	(2583.4)
2	-	0.9804	0.9524	0.9346	(2532.9)	(2460.5)	(2419.5)
3	1	0.9612	0.9070	0.8734	444.6	419.5	403.9
4	2	0.9423	0.8638	0.8163	404.1	370.4	350.0
5	3	0.9238	0.8227	0.7629	394.8	351.6	326.0
6	4	0.9057	0.7835	0.7130	367.4	317.9	289.3
7	5	0.8880	0.7462	0.6663	345.4	290.3	259.2
8	6	0.8706	0.7107	0.6227	338.7	296.3	259.6
9	7	0.8535	0.6768	0.5820	360.6	285.9	245.9
10	8	0.8368	0.6446	0.5439	258.7	199.3	168.1
11	9	0.8203	0.6139	0.5083	356.2	266.6	220.7
12	10	0.8043	0.5847	0.4751	336.6	244.7	198.8
13	11	0.7885	0.5568	0.4440	353.2	249.4	198.9
14	12	0.7730	0.5303	0.4150	351.9	241.4	188.9
15	13	0.7579	0.5051	0.3878	357.6	233.7	179.4
16	14	0.7430	0.4810	0.3624	349.5	226.3	170.4
17	15	0.7284	0.4581	0.3387	(62.0)	(39.0)	(28.9)
18	16	0.7142	0.4363	0.3166	262.9	160.6	116.6
19	17	0.7002	0.4155	0.2959	346.8	205.8	146.5
20	18	0.6864	0.3957	0.2765	346.0	199.5	139.4
21	19	0.6730	0.3769	0.2584	345.5	193.5	132.6
22	20	0.6598	0.3589	0.2415	330.2	179.6	120.8
23	21	0.6468	0.3418	0.2257	1178.7	622.8	411.3
					Σ 2651.1	Σ 472.2	Σ(-)500.5

IRR (interpolated) = 6.2%

TABLE VII-4

CAPITAL COSTS

(R.O. Plant with Full-Power Turbine-Generator)

I.	Solar Pond:	
	Pond construction with liner	$1,061,932
	Initial salt cost	838,237
	Pumps, piping & appurtenances	651,752
	Total, Solar Pond	$2,551,921
II.	Seawater intake, with pumps & piping, installed	133,000
III.	Power generator (Ormat turbine-generator)	695,880
IV.	Switchgear & electricals	83,000
V.	Reverse Osmosis Plant, including pretreatment & initial spares	1,067,000
VI.	Plant building; storage tanks, equipment pads	90,000
VII.	Erection & Installation @ 30% of equipment cost	538,764
VIII.	Engineering & Administration @10%	447,922
	Total Capital Cost	$5,765,382

TABLE VII-5

CALCULATION OF NET "SHADOW" INCOME
REVERSE OSMOSIS PLANT, FULL POWER TURBO-GENERATOR
(IN THOUSAND $)

		Costs		Revenues		
Elapsed Year (1)	Operating Year (2)	Investments (3)	Operation & Maintenance (4)	Water Sales (5)	Subsidy Removal (6)	Net Income (7)
1		2882.7				(2882.7)
2		2882.7				(2882.7)
3	1		116.5	361.4	250.0	494.9
4	2		153.0		256.3	464.7
5	3		174.9		262.8	449.3
6	4		181.9		269.6	449.1
7	5		204.4		276.6	433.6
8	6		181.9		283.8	463.3
9	7		328.8		290.2	322.8
10	8		181.9		296.8	476.3
11	9		204.4		304.7	461.7
12	10		181.9		312.9	492.4
13	11		181.9		321.4	500.9
14	12		181.9		330.2	509.7
15	13		181.9		339.3	518.8
16	14		861.8		348.7	(151.7)
17	15		328.8		358.9	391.5
18	16		181.9		368.6	548.1
19	17		181.9		379.0	558.5
20	18		181.9		389.8	569.3
21	19		204.4		401.0	558
22	20		181.9		412.6	592.1
23		(2156.6)				2156.6
						Σ 5494.1

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TABLE VII-6

CALCULATION OF INTERNAL RATE OF RETURN
REVERSE OSMOSIS PLANT; FULL POWER TURBO-GENERATOR

Elapsed Year	Operating Year	P.W. Factor at Given Discount Rates			Present Value of Net Annual Income at Given at Discount Rates		
		2%	5%	7%	2%	5%	7%
(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
1		1.0000	1.0000	1.0000	(2882.7)	(2882.7)	(2882.7)
2		0.9804	0.9524	0.9346	(2826.2)	(2745.5)	(2694.2)
3	1	0.9612	0.9070	0.8734	475.7	448.9	432.2
4	2	0.9423	0.8638	0.8163	437.9	401.4	379.3
5	3	0.9238	0.8227	0.7629	415.0	369.6	342.8
6	4	0.9057	0.7835	0.7130	406.7	351.9	320.2
7	5	0.8880	0.7462	0.6663	385.0	323.6	288.9
8	6	0.8706	0.7107	0.6227	403.3	329.3	288.5
9	7	0.8535	0.6768	0.5820	275.5	218.5	187.9
10	8	0.8368	0.6446	0.5439	398.6	307.0	259.0
11	9	0.8203	0.6139	0.5083	378.7	285.9	234.7
12	10	0.8043	0.5847	0.4751	396.0	287.9	233.9
13	11	0.7885	0.5568	0.4440	394.9	278.9	222.4
14	12	0.7730	0.5303	0.4150	394.0	270.3	211.5
15	13	0.7579	0.5051	0.3878	393.2	262.0	201.2
16	14	0.7430	0.4810	0.3624	(112.7)	(73.0)	(55.0)
17	15	0.7284	0.4581	0.3387	285.2	179.3	132.6
18	16	0.7142	0.4363	0.3166	391.5	239.1	173.5
19	17	0.7002	0.4155	0.2959	391.0	232.0	165.3
20	18	0.6864	0.3957	0.2765	390.8	225.3	157.4
21	19	0.6730	0.3769	0.2584	375.5	210.3	144.2
22	20	0.6598	0.3589	0.2415	390.7	212.5	142.9
23	21	0.6498	0.3418	0.2257	1401.3	737.1	486.7
					$\sum$ 2958.9	$\sum$ 469.6	$\sum$(-)626.8

IRR (interpolated) = 5.8%

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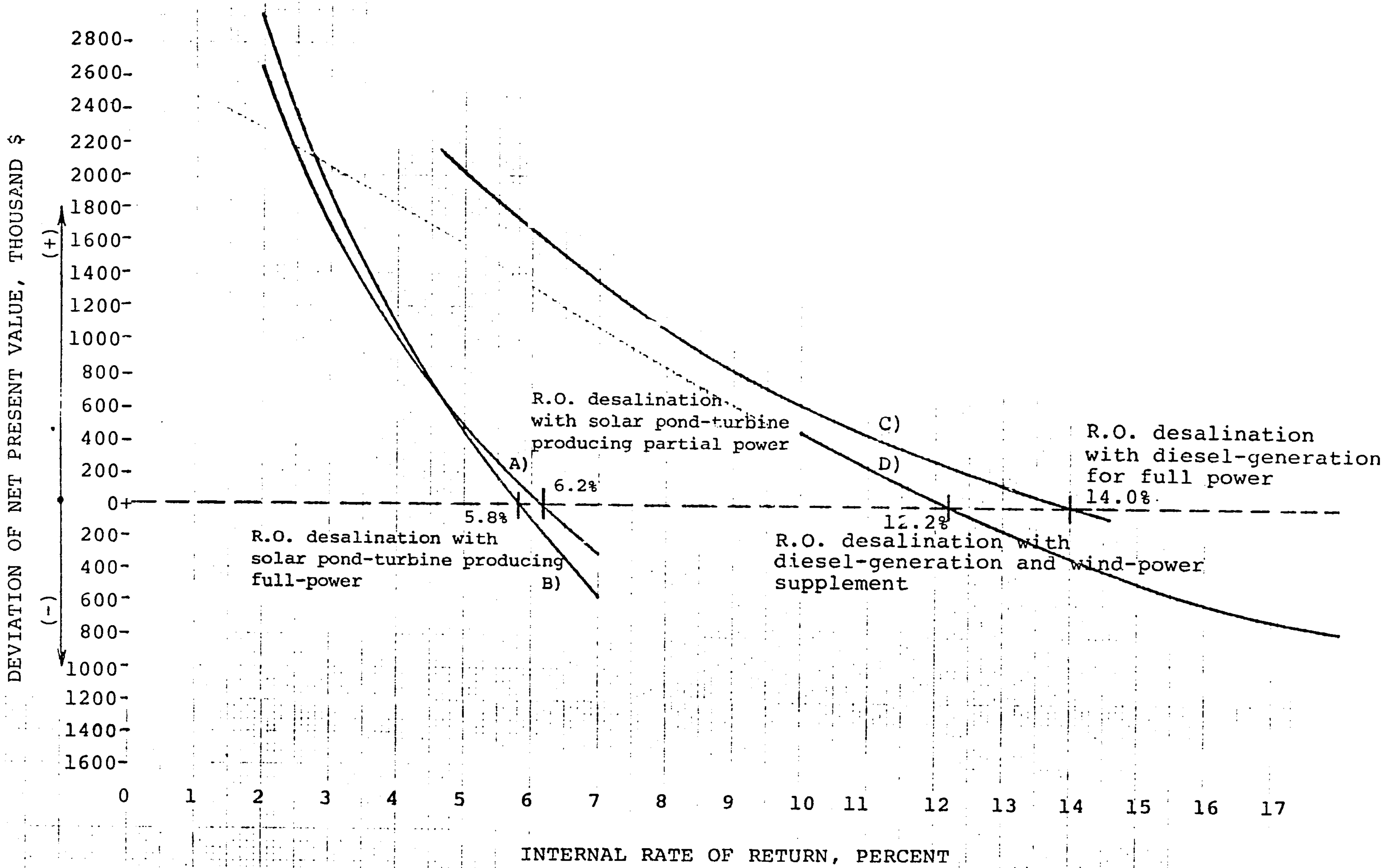

INTERPOLATION GRAPHS SHOWING RETURN ON INVESTMENT FOR VARIOUS CASES
FIGURE VII-1

a) Total capital cost for this system is estimated as $2,050,200. Breakdown of this cost is shown on Table VII-7.

b) Year by year net annual income has been calculated by taking the difference of revenues (shadow price) and costs (investment and O&M, including fuel cost) as shown on Table VII-8.

c) Present values of year by year net annual income has been calculated for 5%, 10%, and 13% discount rates as shown on Table VII-9. Extrapolation yields IRR value of 14.0% as shown on curve c) on Figure VII-1.

iv) Case iv) R.O. plant with diesel-generator and maximum wind-power supplement.

a) Total capital cost for this system is same as in case iii) with additional cost for wind-generator*, estimated as $2,716,580, as shown on Table VII-10.

b) Year by year net annual income has been calculated by taking the difference of revenues (Shadow Price) and costs (investment and O&M) as shown on Table VII-11. Annual fuel savings of equivalent to 370,000 KWH* (considering a Mehrkam 225 KW machine at 90% utilization factor).

c) Present values of year by year net annual income has been calculated for 10%, 13% and 18% discount rates as shown on Table VII-12. IRR is interpolated as 12.2% as shown on curve D) on Figure VII-1.

*Source: "Study-Wind Power Generation for Sal Island," dated August, 1980.

TABLE VII-7

CAPITAL COSTS

(R.O. Plant with Diesel Generator)

I.	Reverse osmosis plant, including pretreatment and initial spares	$ 1,067,000
II.	Seawater intake, with pumps and piping, installed	170,000
III.	Diesel-Generator	95,200
IV.	Switchgear and electricals	83,000
V.	Plant building; storage tanks, equipment pads	90,000
VI.	Erection and Installation @30% of equipment cost	358,600
VII.	Engineering and Administration @10%	186,400
		$ 2,050,200

TABLE VII-8

CALCULATION OF NET "SHADOW" INCOME
REVERSE OSMOSIS PLANT, DIESEL-GENERATOR
(IN THOUSANDS $)

		Costs		Revenues		
Elapsed Year (1)	Operating Year (2)	Investments (3)	Operation & Maintenance (4)	Water Sales (5)	Subsidy Removal (6)	Net Income (7)
0		1000.0				(1000.0)
1		1050.2				(1050.2)
2			232.2	361.4	250.0	379.2
3	1		273.2		256.3	344.5
4	2		299.8		262.8	324.4
5	3		311.9		269.6	319.1
6	4		339.4		276.6	298.6
7	5		342.2		283.8	303
8	6		327.5		290.2	324.1
9	7		333.1		296.8	325.1
10	8		338.8		304.7	327.3
11	9		367.3		312.9	307
12	10		351		321.4	331.8
13	11		377.4		330.2	314.2
14	12		364		339.3	336.6
15	13		370.8		348.7	339.3
16	14		400.4		358.5	319.5
17	15		385.2		368.6	344.8
18	16		392.8		379.0	347.6
19	17		420.7		389.8	330.5
20	18		408.8		401.0	353.6
21	19		439.7		412.6	334.3
22	20	(205)				205
23	21					Σ 4759.3

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TABLE VII-9

CALCULATION OF INTERNAL RATE OF RETURN
REVERSE OSMOSIS PLANT, DIESEL GENERATOR

Elapsed Year	Operating Year	P.W. Factor at Given Discount Rates			Present Value of Net Annual Income at Given Discount Rates		
		5%	10%	13%	5%	10%	13%
(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
1		1.0000	1.0000	1.0000	(1000.0)	(1000.0)	(1000.0)
2		0.9524	0.9091	0.8850	(1000.2)	(954.7)	(929.4)
3	1	0.9070	0.8264	0.7831	343.9	313.4	296.9
4	2	0.8638	0.7513	0.6931	297.6	246.4	238.8
5	3	0.8227	0.6830	0.6133	266.9	221.6	199.0
6	4	0.7835	0.6209	0.5428	250.0	198.1	173.2
7	5	0.7462	0.5645	0.4803	222.8	168.6	143.4
8	6	0.7107	0.5132	0.4251	215.3	155.5	128.8
9	7	0.6768	0.4665	0.3762	219.4	151.2	121.9
10	8	0.6446	0.4241	0.3329	209.5	137.3	108.2
11	9	0.6139	0.3855	0.2946	200.9	126.2	96.6
12	10	0.5847	0.3505	0.2607	179.5	107.6	80.0
13	11	0.5568	0.3186	0.2307	184.8	105.7	76.5
14	12	0.5303	0.2897	0.2042	166.6	91.0	64.2
15	13	0.5051	0.2633	0.1807	170.0	88.6	60.8
16	14	0.4810	0.2394	0.1599	163.2	81.2	54.2
17	15	0.4581	0.2176	0.1415	146.4	69.5	45.2
18	16	0.4363	0.1978	0.1252	150.4	68.2	43.2
19	17	0.4155	0.1799	0.1108	144.4	62.5	38.5
20	18	0.3957	0.1635	0.0981	130.8	54.0	32.4
21	19	0.3769	0.1486	0.0868	133.3	52.5	30.7
22	20	0.3587	0.1351	0.0768	120.0	45.2	25.7
23		0.3418	0.1228	0.0680	70.0	25.2	13.9
					Σ 2029.7	Σ 615.4	Σ 142.7

IRR (extrapolated) = 14.0%

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TABLE VII-10

(R.O. Plant with Diesel Generator and Supplementary Wind Power Generation)

I.	Reverse Osmosis Plant, including pretreatment and initial spares	$ 1,067,000
II.	Seawater intake, with pumps and piping, installed	170,000
III.	Diesel-Generator	95,200
IV.	Switchgear and Electricals	83,000
V.	Plant buildings, storage tanks, equipment pads	90,000
VI.	Erection and installation @30% of equipment cost	358,600
VII.	Engineering and administration @10%	186,400
VIII.	Complete wind power generation system, installed	666,380*
		$ 2,716,580

*Source: "Study-Wind Power Generation for Sal Island", dated August, 1980, Part IV, page 1, escalated by 12.5%.

TABLE VII-11

CALCULATION OF NET "SHADOW" INCOME
REVERSE OSMOSIS PLANT, DIESEL-GENERATOR
WITH WIND-POWER SUPPLEMENT
((IN THOUSAND $)

		Costs		Revenues		
Elapsed Year (1)	Operating Year (2)	Investments (3)	Operation & Maintenance (4)	Water Sales (5)	Subsidy Removal (6)	Net Income (7)
1		1300.0				(1300.0)
2		1416.6				(1416.6)
3	1		191.6	361.4	250.0	419.8
4	2		231.0		256.3	386.7
5	3		256.0		262.8	368.2
6	4		266.4		269.6	364.6
7	5		293.2		276.6	344.8
8	6		288.1		283.8	357.1
9	7		276.5		290.2	375.1
10	8		280.1		296.8	378.1
11	9		283.8		304.7	382.3
12	10		311.2		312.9	363.1
13	11		291.7		321.4	391.1
14	12		310.9		330.2	380.7
15	13		300.1		339.3	400.6
16	14		304.5		348.7	405.6
17	15		332.6		358.9	387.7
18	16		313.9		368.6	416.1
19	17		318.8		379.0	421.6
20	18		339.0		389.8	412.2
21	19		329.1		401.0	433.3
22	20		358.0		412.6	416.0
23		(271.7)				271.7

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TABLE VII-12

CALCULATION OF INTERNAL RATE OF RETURN
REVERSE OSMOSIS PLANT, DIESEL GENERATOR WITH
WIND-POWER SUPPLEMENT

Elapsed Year	Operating Year	P.W. Factor at Given Discount Rates			Present Value of Net Annual Income at Given Discount Rates		
		10%	13%	18%	10%	13%	18%
(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
1		1.0000	1.0000	1.0000	(1300)	(1300)	(1300)
2		0.9091	0.8850	0.8475	(1287.8)	(1253.7)	(1200.6)
3	1	0.8264	0.7831	0.7182	346.9	328.7	301.5
4	2	0.7513	0.6931	0.6086	290.5	268.0	235.3
5	3	0.6830	0.6133	0.5158	251.5	225.8	189.9
6	4	0.6209	0.5428	0.4371	226.4	197.9	159.4
7	5	0.5645	0.4803	0.3704	194.6	165.6	127.7
8	6	0.5132	0.4251	0.3139	183.3	151.8	112.1
9	7	0.4665	0.3762	0.2660	175.0	141.1	99.8
10	8	0.4241	0.3329	0.2255	160.3	125.9	85.3
11	9	0.3855	0.2946	0.1911	147.4	112.6	73.1
12	10	0.3505	0.2607	0.1619	127.3	94.7	58.8
13	11	0.3186	0.2307	0.1372	124.6	90.2	53.6
14	12	0.2897	0.2042	0.1163	110.3	77.7	44.3
15	13	0.2633	0.1807	0.0985	105.5	72.4	39.4
16	14	0.2394	0.1599	0.0835	97.1	64.9	33.9
17	15	0.2176	0.1415	0.0708	84.4	54.9	27.4
18	16	0.1978	0.1252	0.0600	82.3	52.1	25.0
19	17	0.1799	0.1108	0.0508	75.8	46.7	21.4
20	18	0.1635	0.0981	0.0431	67.4	40.4	17.8
21	19	0.1486	0.0868	0.0365	64.4	37.6	15.8
22	20	0.1351	0.0768	0.0309	56.2	31.9	12.9
23		0.1228	0.0680	0.0262	33.4	18.5	7.1
					Σ 450.9	Σ(-) 154.3	Σ(-) 759.1

IRR (interpolated) = 12.2%

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F. USE CHARGE FOR WATER

It is feasible to collect a use charge for water on Sal Island. Currently, desalinated water is sold to the public, generally by the bucketful, at the storage tank. Water is also delivered by trucks to domestic users, as well as commercial facilities (hotels, shops, etc.). Current average domestic charges are $2.50/m^3, and $3.00/m^3 is charged for commercial use. For any new desalination installation, charging for water use could continue as the demand for water continues to rise.

G. DISCUSSIONS

The foregoing results are based on certain assumed values of cost variables. While the assumptions are logical and site-specific for Sal Island, it is useful to verify the sensitivity of the cost variables to the value of IRR.

1) Capital Cost:

In the solar pond applications, a ten percent capital cost decrease would signify an increase of IRR by approximately 1.5 percent.

For example, the initial salt cost is a significant component of the capital costs for a solar pond (approximately 14.5% in case ii) on page VII-5). If the salt price is halved, the IRR increases by over one percent.

Secondly, the pond construction cost can be reduced if a suitable natural depression is existing at site. Reducing pond construction cost by 50 percent would increase the IRR by 0.4 percent.

The Rankine cycle power apparatus is also a major cost item. It is possible that with increased sales in the coming years the production cost of these generators could be reduced. A 20 percent cost reduction would denote an increase in IRR value of 014 percent.

The net efficiency of the solar pond is conservatively estimated at about 15 percent. With proper operation and controls, it is likely to increase the efficiency to up to 20 percent.

If the pond efficiency is estimated, say, at 18 percent, this would decrease the solar pond area by 17 percent, thereby reducing the pond cost and increasing the IRR by 0.5 percent.

2) Fuel Costs:

The relative advantage of diesel-powered R.O. systems over solar pond powered systems would reduce significantly if the differential cost growth of the diesel oil increases. For example, if the oil price is escalated at a rate of 7% p.a. (instead of 3.5% p.a. considered in the study), its effect on the IRR would be to reduce it by 3.5 percent.

3) Shadow Price:

For Sal Island, actual dollar values of annual subsidy by the local government have been used to account for the shadow price. This value may be widely different in other countries or locations. Shadow price will increase with increasing fuel costs. Higher shadow price for water would render the solar pond application more attractive. (IRR for all cases considered would be reduced significantly if the current water sales price alone is used.)

4) Pond Operation:

Correct pond operation and control is of the utmost importance. To restart an upset salt pond would signify a heavy replacement salt cost and loss of potable water production. Fortunately, considerable experience has been gathered with laboratory and commercial solar pond applications and techniques have been developed for appropriate pond operation.

H. CONCLUSIONS

The financial and economic analysis show that the solar pond powered desalination systems have a positive return on investment. Under the economic considerations adopted for the study, IRR's for solar pond powwered systems, however, are inferior to those for the diesel-powered or wind-powered systems. This would economically not justify constructing the solar pond project at this time. However, as explained earlier, a possibility exists of reducing costs for solar pond application, thereby increasing its commercial attractiveness. On the other hand, ever-increasing costs of imported fuel could render the conventional diesel power less attractive. These changes in the cost picture are time-dependent. It is, therefore, concluded that the economic merits of a solar pond project specifically for seawater desalination should be further reviewed at a later date, probably within a period of three years.

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FIGURES