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Denver, Colorado, U.S.A.

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23-25 March 1981

Sponsored jointly by
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Coordinated by
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بالتنسيق مع
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VOLUME I

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PRINCIPLES OF DESALINATION

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ABSTRACT

A review of the principles of the major desalination processes is presented, with emphasis on the energy required for the separation process. The types of energy, heat, mechanical, and electrical, used in these processes are described and the method for comparing them on a common basis is discussed, both from the first-law and second-law aspects. The way to determine the minimal separation energy for desalination is described, and typical values are presented. In real processes, the energy required is about 10 to 50 times higher than the minimal. Solar energy conversion to thermal energy at different temperature levels, and to mechanical work, is described, and values of typical conversion efficiencies are given. It appears that for seawater desalination, solar stills are approximately competitive with more advanced systems where the sun is used to generate higher temperature heat, or power, for use in an efficient distillation, freezing, RO, or ED plant. This would change to the advantage of the latter systems as their efficiencies and effectiveness increase.

DESALINATION PROCESSES

Separation of water from salts in an aqueous salt solution can be obtained by a number of different processes. A brief review of the principles and energy-characteristics of the major ones is provided below. Much more detailed information can be found in [1-3].

In distillation, the solution is brought to a thermodynamic state in which the more volatile component (H_2O) is separated by evaporation from the less volatile salts. This state can be attained either by raising the temperature of the solution at constant pressure, or by lowering the pressure at constant temperature. The first process can be produced either by supplying heat directly, or by mechanically compressing the emanating water vapor to a higher pressure and, consequently, temperature, so it could transfer heat by condensation to the evaporating brine. The second process can be produced by condensing the vapors by using coolant which has a lower temperature, in a vessel essentially evacuated of non-condensables. This process can, for example, be obtained by evaporating water from the ocean's surface in evacuated vessels, where the condensers are cooled by colder seawater from the ocean's depths, such as envisaged in the ocean-thermal energy conversion concept.

The most common commercial desalination processes using direct heat are multi-stage flash evaporation (MSF), multi-effect distillation such as in the horizontal tube multi-effect (HTME) and vertical-tube evaporator (VTE) concepts, or single-effect evaporation, such as in basin-type solar stills. Commercial distillation processes typically operate at temperatures below 120°C, dictated by the need to minimize $CaSO_4$ scaling.

In freezing, the solution is brought to a thermodynamic state in which the component with a higher freezing point (H_2O) separates itself by crystallizing within a saline solution which thus becomes increasingly concentrated and which consequently has a progressively lower freezing point. The crystals are then washed from the salt and consist of pure water. The separation of the water from the salt during freezing can be clearly seen on the phase diagram of seawater, where, for salt concentrations below the eutectic ones, the water component is the first to separate by freezing as the temperature is lowered. Naturally, lowering the temperature further below a given level, would start also the precipitation of salts, which is undesirable in this process. Typically, the process operates between ~-5°C (freezing temperature of seawater) and the ambient air temperature (melting of product).

Two desalination processes utilizing the selective transport properties of membranes are Reverse Osmosis and Electrodialysis. In both, a membrane which has a high rejection to the passage of one of the components (either water or salts) separates the highly saline salt solution from the more dilute one. To overcome the electrochemical potential difference which arises from this difference in concentration and which tends to create flows that reduce it, an opposing thermodynamic potential is applied. This potential is made large enough to generate a flux in the direction which increasingly separates the two components. In reverse osmosis this potential is pressure applied to the more concentrated solution, and it needs to be larger than the osmotic pressure difference between the two solutions to provide a positive net flux of H_2O through the membrane, towards the side which has the lower salt concentration (the product side). In electrodialysis, the potential is DC electric voltage which makes the salt ions move towards potentials of opposite sign. By placing the solution which needs to be desalted between a cation-permeable membrane and an anion-permeable one, it will become gradually depleted from the salt ions which increase in concentration in the channels on the opposite sides of the membranes, and will hence be able to supply desalted water.

In both of these "selective transport" processes, it is not necessary to have a phase change of the solution to desalt it, as is required in destillation and freezing, and they operate at essentially ambient temperature. An exception is the possibility of electrodialysis of saline water at elevated temperatures. This was shown to require less energy than operation at ambient temperature mainly because of higher electric conductivity and lower viscosity of the solution [4]. This process, however, is not widely used.

Phase change needs a relatively large energy investment, the magnitude of which is almost independent of the concentration of the solution.

To make these processes commercially acceptable, this energy needs to be recovered as best as possible, by techniques such as feedwater preheat by condensing the product, in the MSF and vapor-compression processes, and by using this heat of condensation for both feedwater preheat and evaporation at a lower pressure, as done in multi-effect installations. These techniques, however, increase the cost of the plants and their complexity. As indicated above, the magnitude of the driving potential in the selective transport processes is proportional principally to the concentration gradient between the concentrated and dilute solutions, i.e. to the salinity of the feedwater. The energy required for desalination is thus proportional to the salinity of the feed and to the desired flux of the product, and, for low-salinity feeds, is typically lower than that required to change its phase.

Table 1 below summarizes the different types of energy inputs for these major desalination processes.

PROCESS	ENERGY TYPE	
	HEAT	MECHANICAL
DISTILLATION: MSF, MEF	at 70-120°C	
VAPOR COMPRESSION	steam ejector, or	electric motor, I.C. engine, turbine
FREEZING	absorption cooling cycle, or	vapor compression cooling; motor-, engine-, or turbine-driven vapor compressor
REVERSE OSMOSIS		electric motor-, engine-, or turbine-driven high pressure pump
ELECTRODIALYSIS		DC electric current

Table 1: Types of energy used in desalination processes

Before discussing the energy requirements of desalination processes, it is thus necessary to realize that the "quality" of the exergy (2d law) must be considered in addition to the commonly presented quantity or efficiency (1st law). This quality can be expressed as the exergy (or "essergy", [5]), or even as the actual cost of a unit of energy. It allows us thus to evaluate the different potentials to do work, or the different costs of the same unit of energy (say, Joule) when obtained in the form of mechanical shaft power, or electrical energy, or heat supplied at different temperatures.

Heat is usually generated at temperatures far above those needed for the desalination process (typically for distillation), since it originates from combustion (or nuclear fission) at temperatures of about 1000°C or more. If the temperature of the steam is degraded down to about 120°C or less needed for desalination, a large potential to do work (say, generate power) between these two temperatures is lost. It is thus desirable to discount the heat used for desalination by the amount of work it can produce from the origination temperature of the steam (say about 500°C) and down to the temperature at which it is used for this desalination process. Thermodynamically, the criterion for the quality of the heat source is its exergy, ε:

$$\varepsilon = E + P_0V - T_0S - (E_0 + P_0V_0 - T_0S_0) \tag{1}$$

where $E \equiv$ energy $P \equiv$ pressure
$V \equiv$ volume $T \equiv$ absolute temperature
$S \equiv$ entropy

The subscripts $_0$ denotes the state of equilibrium.For an open system, the steady-flow exergy ε_f:

$$\varepsilon_f = H - T_0S - \Sigma\mu_{i0}Ni = H - T_0S - (H_0 - T_0S_0) \tag{2}$$

for $N_i = N_{i0}$, where
$\mu_i \equiv$ Gibbs chemical potential of component i
$N_i \equiv$ amount of component i(say in moles)

From the economic standpoint, the above discount can be applied, allocating the full cost of steam generation to the highest temperature steam from the boiler (~500°C), and assuming that steam (or hot water) at ambient condensation temperatures (say 40-50°C) has no economic value.

When mechanical energy is applied, it is directly interpretable as exergy, however each unit of work requires in real, conventional plants the expenditure of about 2.5 to 4 units of heat, depending on the efficiency of the power plant, and in the case of electric power, also on the efficiency of the electric generation and distribution system.

The least amount of heat needed to produce a unit of work would occur if an ideal (reversible) thermodynamic cycle would be employed between the temperatures of the heat source and the ambient (T and T_0, respectively), where

$$W_{rev} = Q\left(1 - \frac{T_0}{T}\right) \tag{3}$$

In summary, the exergy of heat and work sources, as expressed in eqns (1-3) serves as the common thermodynamic base for the comparison of energy sources needed to operate desalination systems. Reversible processes are implicitly assumed and the results are thus an upper limit of performance. A more realistic criterion for the comparison of commercial processes is the heat source's potential to do work via real (irreversible) processes, i.e. based on actual efficiencies.

Finally, desalination processes need auxiliary energy for purposes such as pumping and instrumentation. Although pumps may be driven by waste-steam turbines or internal combustion engines (both operating from a heat source), typically the energy is electrical. The amount of the auxiliary energy is of the order of up to 3% of the total and hence has a heat equivalent of the order of 10% of the total. The impact of the auxiliary energy requirements needs careful scrutiny, especially where solar energy is to be utilized: whereas the plant may operate well and economically with solar heat at relatively low temperatures, the need for shaft of electrical power by the auxiliaries may force the provision of significantly more expensive systems for that purpose.

In closing, it is noteworthy that another criterion for the comparison of different types of energy inputs, which is used when a major objective is the substitution of one energy resource for another (say, to conserve oil), is the "Resource Energy" value of each type (cf. [6]). All energy values there are expressed in terms of the energy-value of the resource to be replaced.

THEORETICAL MINIMUM ENERGY REQUIREMENT FOR DESALINATION

The reversible, isothermal, isobaric, differential (no mixing of solutions of different concentrations) process with zero yield (infinitesimal amount of water is separated out) requires the least amount of work, W_{rev}, to proceed, which is equal to the change in the Gibbs Free Energy between the final and initial states of the system [7-9].

$$W_{rev} = (\Delta g)_{P_0,T_0} \tag{4}$$

where g ≡ Gibbs Free Energy per mole of product, and

$$G_i = \sum_i \mu_i N_i = g_i N_w \tag{5}$$

where

μ_i ≡ electrochemical potential of component i

N_i ≡ moles of component i

N_w ≡ moles of water,

and if heat is applied to the process too, it is converted to its work-equivalent via eqn (3), and thus eqn (4) becomes

$$q_{rev}\left(1 - \frac{T_0}{T}\right) + W_{rev} = (\Delta g)_{P_0;T_0} \tag{6}$$

Work needed for the differential process, removing a small increment of water dN from the solution at each unit of time, is

$$dW = (\Delta G)dN \tag{7}$$

or assuming water vapor at low pressure to be an ideal gas:

$$W = \int_{N_1}^{N_2} RT \ln a_B \, dN = \int_{N_1}^{N_2} RT \ln\left(\frac{P}{P_0}\right) dN \tag{8}$$

where N_1 and N_2 are the initial and final number of moles of salt solution, R is the universal gas constant, a_B is the activity of water in the salt solution, and P is the vapor pressure of pure water at the process temperature T. For desalination of a 3.50% NaCl solution of an ambient temperature of 25°C, and for zero yield, the very least separation energy is shown by (7) to be 2.98 kW-hr /1000gal. water, or 2.83 kJ/kg water. For 50% recovery of water, this energy increases to 4.12 kW-hr/1000 gal. (3.91 kJ/kg), and for 90% recovery it is 8.40 kW-hr/1000 gal. (7.98 kJ/kg).

Relaxing the restrictive assumption of ideal gas, the separation work for real solutions was computed in [8] by means of an extended Debye-Hückel equation, and is presented in Table 2.

Temperature °C	% recovery of fresh water 0	25	50	75
25	2.55 (2.68)	2.91 (3.06)	3.52 (3.70)	4.80 (5.04)
75	2.96 (3.11)	3.43 (3.60)	4.16 (4.37)	5.72 (6.01)
125	3.31 (3.48)	3.82 (4.01)	4.63 (4.86)	6.34 (6.66)

Table 2: Minimal Energy of Separation of Pure Water from Normal Sea Water of 3.45% wt. % Salt, kJ/kg fresh water (kW-hr/1000 gal. fresh water), [8].

In distillation, the minimal driving force is the boiling point elevation, as seen in eqn (8), however the separation energy in real processes increases above the values of Table 2 due to irreversibilities, such as mixing of solutions of different concentration (which is increasingly aggravated with the difference in concentration), nonisothermality due to the need to change phase or transfer heat, and transport and rejection losses. Polarization losses occur in membranes processes, and solution and membrane resistance (in electrodialysis) add to the work of separation. The separation energy increases with the concentration of the desalted solution: it almost triples when the concentration is doubled to 6.9%.

In practice, the energy needed for desalination is about 12-to 50-fold larger than the minimal energy of separation, which only serves as the lower thermodynamic limit. In addition to the energy of separation, auxiliary energy is needed for pumping and other purposes. As a matter of fact, although it may be energetically desirable to operate at the lowest recovery ratios, this increases the pumping power. The total energy (separation and pumping) thus goes through a minimum for a certain recovery ratio.

Typical energy consumption for major desalination processes is shown in Table 3.

PROCESS	FEATURE	PERFORMANCE FACTOR, lb Water/MBtu	kJ/kg Water	PUMPING ENERGY, kJ/kg Water
DISTILLATION	Single-Effect Waste-Heat	0.95	2443	14-33
DISTILLATION MSF	90°C	8	290	8-10
DISTILLATION MSF	120°C	12	193	6- 8
DISTILLATION MEF	71°C	8	290	4- 6
DISTILLATION MEF	120°C	15	155	4- 6
VAPOR COMPRESSION	65 kWhr/kgal	38	61	included
FREEZING	50 kWhr/kgal	49	47	incl.
REVERSE OSMOSIS	seawater feed 45 kWhr/kgal	54	43	incl.
REVERSE OSMOSIS	5000 ppm feed 12 kWhr/1000gal	200	12	incl.
ELECTRODIALYSIS	5000 ppm feed 12 kWhr/1000gal	200	12	incl.

Table 3: Typical Energy Consumption in Desalination Processes (cf. [10])

SOLAR ENERGY CONVERSION

Solar energy radiation consists of electromagnetic waves located principally in the wave band of 0.2 to 2.0μm , with a maximum at about 0.45μm . Its intensity is higher in summer than in winter and generally varies throughout the world from about 1 to 8 kW/m^2 day. The average in the U.S. is about 4.7 kW/m^{2}day and in Saudi Arabia above 6.3 kW/m^{2}day. Solar radiation can be converted to heat by its absorption in a solar-thermal collector, or to electricity by a photovoltaic or thermo-electric device. Increasing the incoming energy flux by concentration allows the attainment of higher temperatures. Solar energy is also the source of wind, which can be used to generate mechanical power.

Solar energy can be used for water desalination in several ways. The most common one is via a solar still which serves both as a converter of solar energy to heat, and as a distiller [2,3,11]. These stills are typically single-effect, because they operate best at low temperatures (~<65°C). An alternate thermal method is to collect solar heat without or with the aid of concentrating devices and to use that heat to operate a conventional distillation plant, such as MSF or MEF. As indicated above, such plants operate economically in the top-temperature range of 70-120°C, hence requiring a solar heat source at 80-135°C. Such temperatures may presently be obtained with reasonable collection efficiency by using flat-plate evacuated collectors with or without moderate concentration, CPC collectors, trough-type and similar concentrators, and solar ponds.

Thermal energy at such temperature levels can also be used to operate a thermal engine and generate shaft power. Power cycles using organic working fluids are typically used with an overall-cycle efficiency of about 10%, and a hybrid-source steam cycle ("SSPRE") using an auxiliary heat source to superheat from the 100°C obtained from solar collectors, to 600°C, (at approximately atmospheric pressure) has an efficiency of about 20% [6]. Higher concentration devices, such as the central tower concept (cf. [12]) can produce temperatures (and hence also efficiencies) similar to those obtained in conventional power plants. The shaft power from these cycles can be used to drive pumps or compressors directly, or to generate electricity.

Electricity to provide power for desalination processes or their auxiliary equipment can be obtained directly by photovoltaic conversion devices, such as silicon or cadmium-sulphide cells.

Finally, as mentioned above, the surface of the oceans serves as a solar energy collector. Where the ocean is sufficiently deep and the mixing isn't excessive, the temperature in the depths is about 20°C lower than that at the surface. This temperature difference can be utilized to generate power via a Rankine cycle. In the open-cycle concept (cf. [13]), the surface seawater is flashed in an evacuated chamber, the generated steam drives a turbine, and condenses on the condenser which is cooled by the colder water from the depths (~600m) and which maintains the low pressure in the system. The condensate is fresh water which is presently not contemplated to be used, but which could provide large quantities of desalted water as a by-product of the OTEC power plant. Typically a 100 MW plant produces $(1.43)10^5$ m^3/day (37.8 MGD) of fresh water.

Due to the transient nature of both solar energy and the ambient weather conditions which affect collection efficiency, most solar energy-based processes require some form of energy storage. This can be obtained by storing heat (sensible or latent, or thermochemical), electricity (batteries), or mechanical power (pumped storage, compressed air, flywheels), but it complicates the overall system and increases its cost. Seemingly, one could operate intermittently and store fresh water, but both the undesirability of plant transients and the limitations of water storage capacity, do not always allow this method.

A summary of solar energy conversion efficiencies and other features relevant to desalination is provided in Table 4. Two comparisons need to be made: one with conventional-fuel systems, and the other among the different solar candidates. A noteworthy observation is that the higher temperatures which the collectors provide, as compared to the still, allow operation at a better plant Performance Ratio, since the solar energy conversion efficiency of the collectors is typically at least as good as that of the still. Thus, (Table 3), the P.R. of the still is 0.95 (lb/MBtu), of an MEF plant operating with collectors at 80°C it is about 8, and of the one at 135°C it is 12-18. Hence, the productivity of collector-based distillation plants can be about 9-20 fold higher than that of the solar still. Coincidentally, the collector systems also cost approximately that much more than stills*, and the distillation plant's cost and higher maintenance expenses need to be yet added to that.

PROCESS	Efficiency, %			
	Solar Stills	Solar Collectors & Concentrators 80°C	135°C	Central Tower 500°C
Conversion to Heat	30-50	40-60	30-50	
Thermal Power	--	<10%	4(12 for hybrid)	20-25
Photovoltaic Conversion	9-15	(at ~ambient temperature)		

Table 4: Typical Solar Energy Conversion Efficiencies (from solar source to final use)

As to the production of electricity and mechanical power for desalination purposes, examination of Tables 3 and 4 indicates that vapor compression, freezing and seawater RO have energy demands which are about 1/4 of those for MSF and MEF distillation. However, for operation under 135°C, where reasonably economical, proven, and reliable collectors exist, the power conversion efficiency is less than 1/4, and the systems may be only marginally competitive. Higher conversion efficiency, such as of the SSPRE cycle [6] would improve the situation of the power conversion schemes. On the other hand, operation of brackish water RO and ED by solar-generated power has a clear advantage over distillation of the same. In any case, the costs added by the power cycle need to be considered, too. Power production from a high temperature solar source, such as the Power Tower, can only be compared by an economic analysis, since heat at these temperatures is not directly applicable to desalination processes.

SOME PROJECTIONS AND CONCLUSIONS

As energy costs keep rising at a rate above overall inflation, and as demand for water keeps growing, there would be an increasing trend to

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*cost of land not included

design more energy-efficient desalination plants. In distillation it would mean that Performance Ratios of 20 and above will be sought, and further improvements would be made in most of the other processes. Since solar energy is still not fully competitive with conventional sources, this trend would create more favorable conditions for its use and would remove the single-effect collector-cum-still from competition for most areas, except where land is cheap and where sophisticated maintenance and operation are not available. Solar energy would become an increasingly desirable source for the replacement of conventional fuels, and for allowing their use in more appropriate applications.

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AN OVERVIEW OF SOLAR DESALINATION TECHNOLOGIES

by

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ABSTRACT

Providing adequate water is becoming a serious problem in arid areas as the population of the world increases. In many arid areas saline or brackish water is available, but desalination is necessary to make it potable or useful for irrigation. In 1977 the worldwide desalination capacity was 4 million m^3/day, provided by 1,500 plants of over 100 m^3/day capacity. The capacity in the United States was 400,000 m^3/day, provided by 510 plants. At present most of these plants are powered by conventional fossil fuels directly or by electric energy from a grid system; only a few small systems are powered by solar energy. However, renewable solar resources can be used in a variety of ways to supply energy for desalination. This paper discusses the most important solar desalination methods and the solar technologies which can supply the heat and power necessary for desalination.

1 OVERVIEW OF SOLAR TECHNOLOGIES

Solar energy technologies can be divided into two broad classifications [1]: Natural Collection Systems and Technological Conversion Systems (see Fig. 1).

In natural collection technologies the biosphere—earth, wind, or water—provides storage and collection free. Since no collectors need to be built, the cost of useful energy for natural systems is determined by the converter, e.g., the wind turbine. For technological conversion systems, the amount of collectable energy is proportional to the total amount of incident solar energy on the area of the collector at a given geographical location. Therefore, the cost of useful energy for a technological conversion system depends on the cost and efficiency of the collectors as well as the convertors.

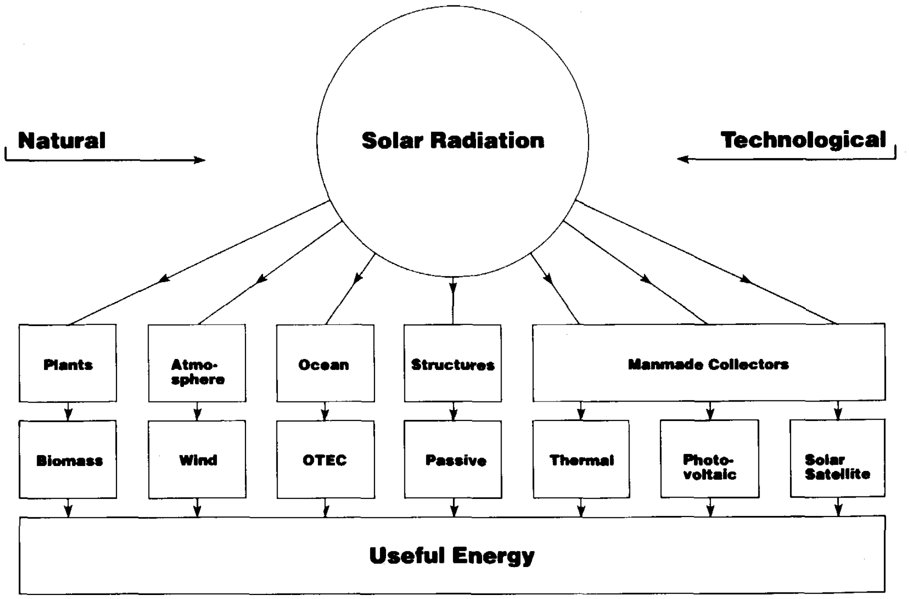


Figure 1. Classification of Solar Technologies

1.1 Solar Thermal Conversion

Two distinct but complementary approaches are used to convert sunlight to heat [1]: passive and active systems.

A passive system is generally defined as one in which the energy and mass flow is by natural means (e.g., conduction, convection, radiation, and evaporation), but without a prime-mover such as a pump. An active system is one in which the mass flow is by means of fans or pumps. A hybrid system is one where at least one significant flow is by natural means and at least one is by forced means. For most desalination processes active systems are required.

The simplest type of collectors is the solar pond. A shallow pond is heated by the sun during the day, but covered by a transparent sheet of plastic to prevent heat losses at night. A somewhat more sophisticated type is the so-called salty pond, in which a salt is dissolved to counteract the natural tendency of warm water to rise. If the concentration of salt is sufficient to inhibit convection, the sun can heat the bottom layer to temperatures as high as 85° C and the cooler layer on top acts as a buffer to reduce heat losses to the surroundings. Heat can be extracted from the bottom layer and the body of the pond can provide storage for times when there is no sunshine. The potential of such ponds for desalination will be discussed later in this paper.

Simple flat-plate collectors can typically achieve temperatures up to 90° C. Figure 2 shows a typical flat-plate collector. Slightly higher temperatures, say 200° C, can be attained by enclosing the tubes carrying the cooling fluid in a vacuum. This type is called an evacuated tube collector [2]. For higher temperatures it is necessary to concentrate the solar radiation. Optical concentration is necessary to achieve temperatures sufficient to drive a heat engine efficiently and convert solar radiation to power. Optical concentration generally necessitates tracking the sun with the concentrator or the receiver.

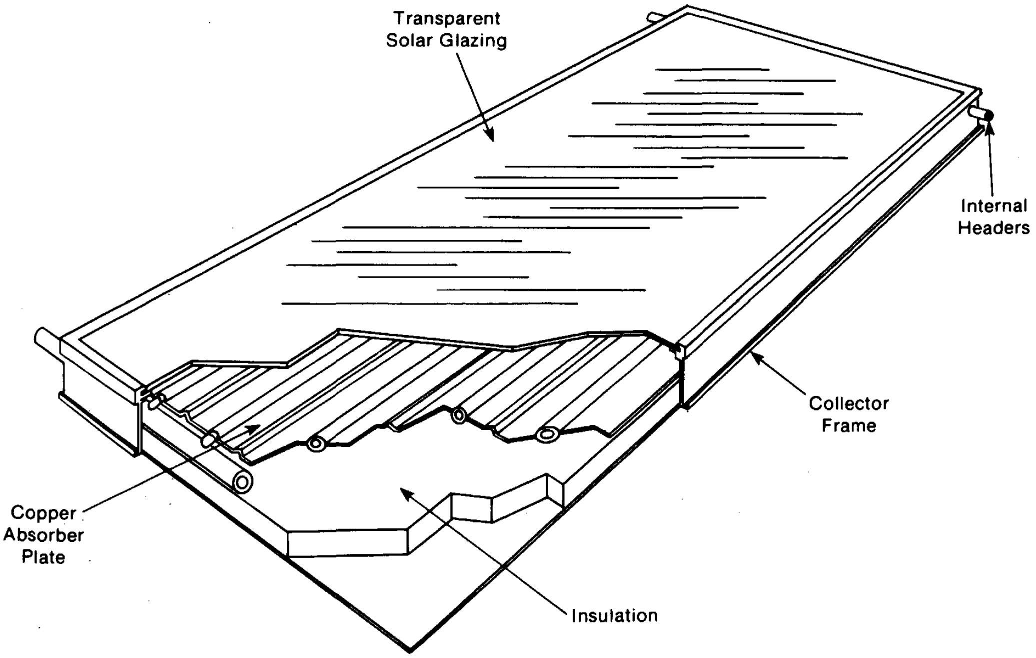
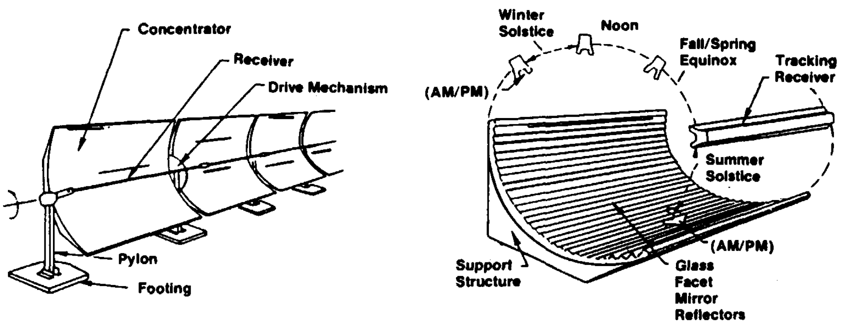


Figure 2. Schematic Diagram of Liquid Flat-Plate Collector

There are two basic receiver concentrator categories: point focusing and line focusing. Concepts of each type are shown in Fig. 3. The most common type, suitable for intermediate temperature, uses a single-axis tracking parabolic trough as the concentrator to focus sunlight on a linear receiver. As shown in Fig. 3a, the concentrator rotates about its axis to follow the sun's apparent motion in the sky. A fixed mirror collector like the one shown in Fig. 3b requires no concentrator movement. Instead, the receiver moves as the sun's motion changes the focal line of the fixed concentrator. Linear-segmented array concentrators (see Fig. 3d) have fixed receivers and moveable reflectors. The mechanically linked reflectors rotate so that a line focus from the receiver is maintained stationary. Parabolic dish concentrators are dual-axis tracking point-focusing devices that can achieve temperatures as high as 1200°C. The concentrator tracks the sun and reflects the radiation onto a point-focus receiver (see Fig. 3c). Fresnel-lens collectors can either be line focusing or point focusing, depending on the type of lens. Fresnel lenses are refracting types because the radiation passes through the transparent lens and is directed towards the focal point or focal line by refraction. The lens structure tracks the sun and refracts sunlight onto the receiver located behind the lens. Figure 3e illustrates a fresnel lens linear-concentrator system.

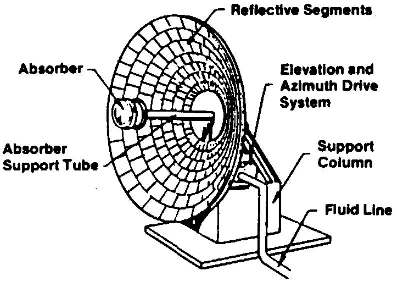
Figure 4 shows the operating temperatures for different types of collectors. The proper choice of collector depends of the temperature needed for the process.

The selection is called end-use matching and is described in [3]. Figure 5 illustrates a closed-loop distributed solar thermal conversion system suitable for providing process steam to a distillation plant. Normally no storage is required for solar distillation since the process can be carried on when the sun is shining and turned off at night.

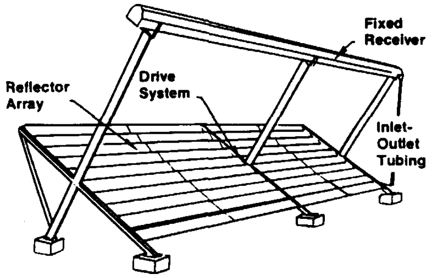


(a) Parabolic Trough

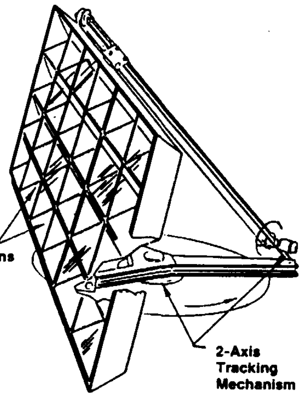
(b) Fixed Mirror



(c) Parabolic Dish



(d) Linear-Segmented Array



(e) Fresnel Lens Collector

Figure 3. Five Basic Distributed Receiver-Concentrator Modules

There are two broad classes of solar thermal conversion systems: central receiver and distributed receiver systems. In a central receiver system mirrors called heliostats reflect solar energy to a single central collector in which the working fluid is heated. In a distributed system a receiver is placed at the focal zone of each concentrator, and a hydraulic distribution network transports the hot working fluid to a central point. Figures 6 and 7, respectively, show schematic diagrams for a typical central receiver thermal conversion system and a typical distributed receiver system.

1.2 Bioconversion

Bioconversion processes convert biomass into useful energy. Biomass is material derived from growing organisms—wood, sugarcane, corncobs, or seaweed. It is a direct and natural form of solar energy and can be produced where the sun shines on land and oceans. Direct combustion of wood has been the principal form of energy for most of man's existence. In the United States, wood provided 75% of

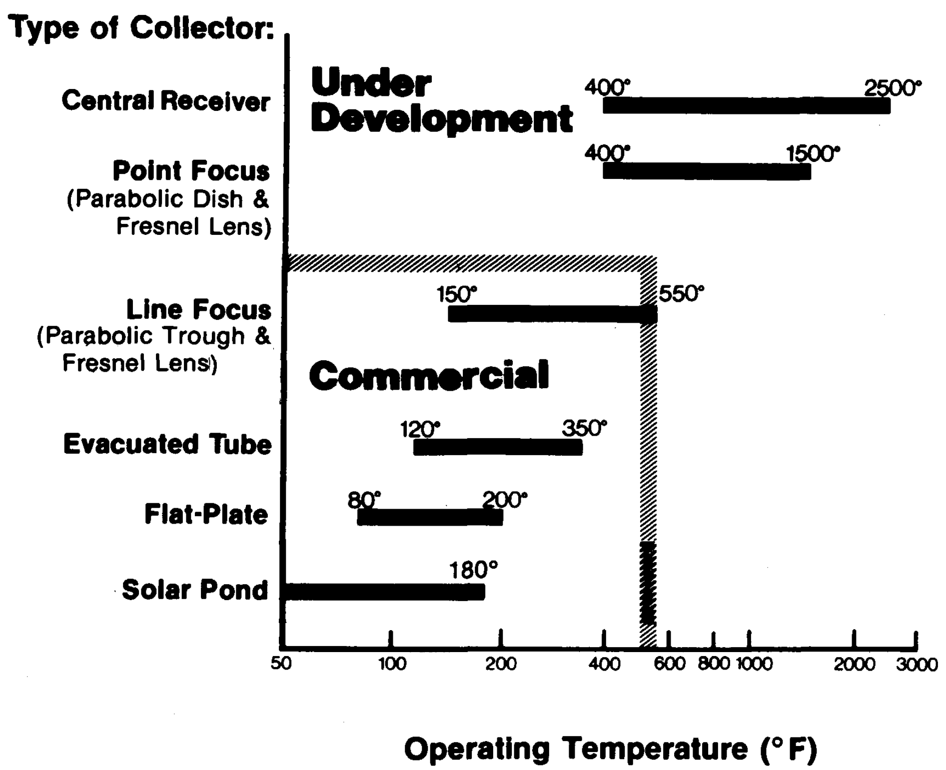


Figure 4. Operating Temperature Ranges for Different Types of Concentrating Collectors

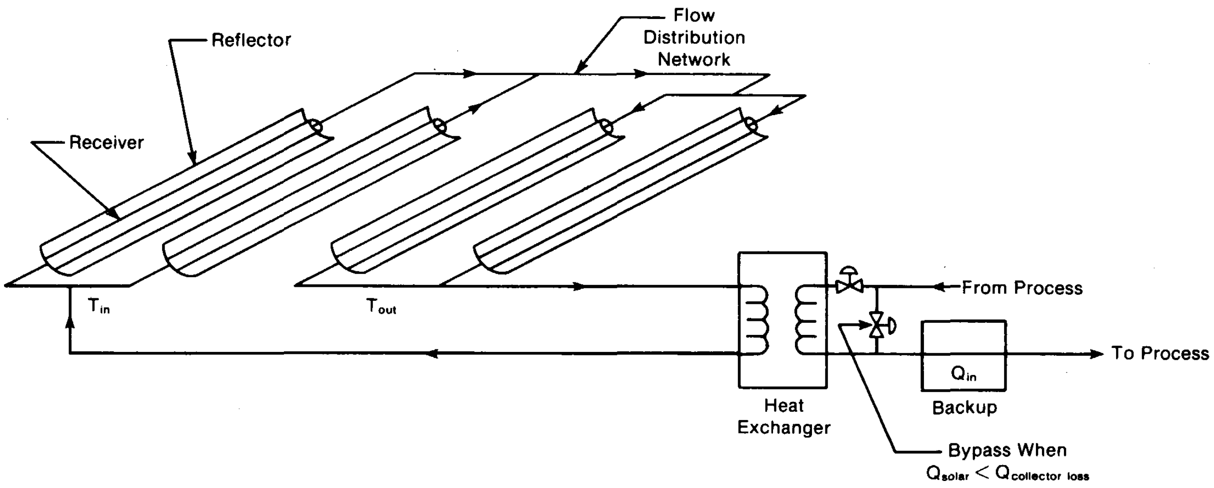


Figure 5. Closed Loop Distributed Solar Thermal System with Fossil Fuel Backup

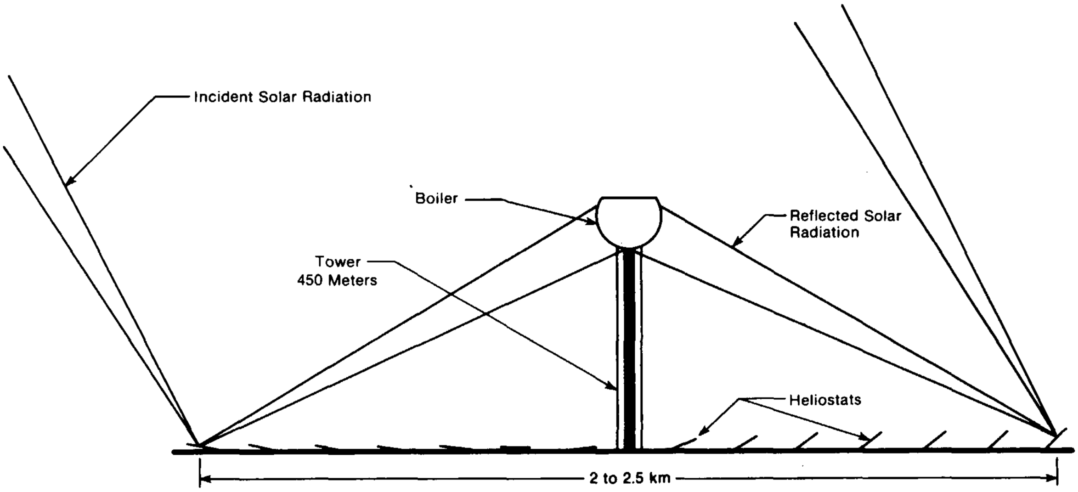


Figure 6. Cross Section of a Central Receiver System

Sunlight is reflected from mirrors supported on heliostats that are actively steered to image the sunlight onto the absorbing surface of the boiler. Here the energy is converted to heat which is carried to the ground (e.g., as high pressure steam) and used as process heat or to drive a turbine generator set. The system depicted can power a 200-MW electric generator if the solar elevation is above 15°. The waste heat could be used for desalination in a dual or total energy system.

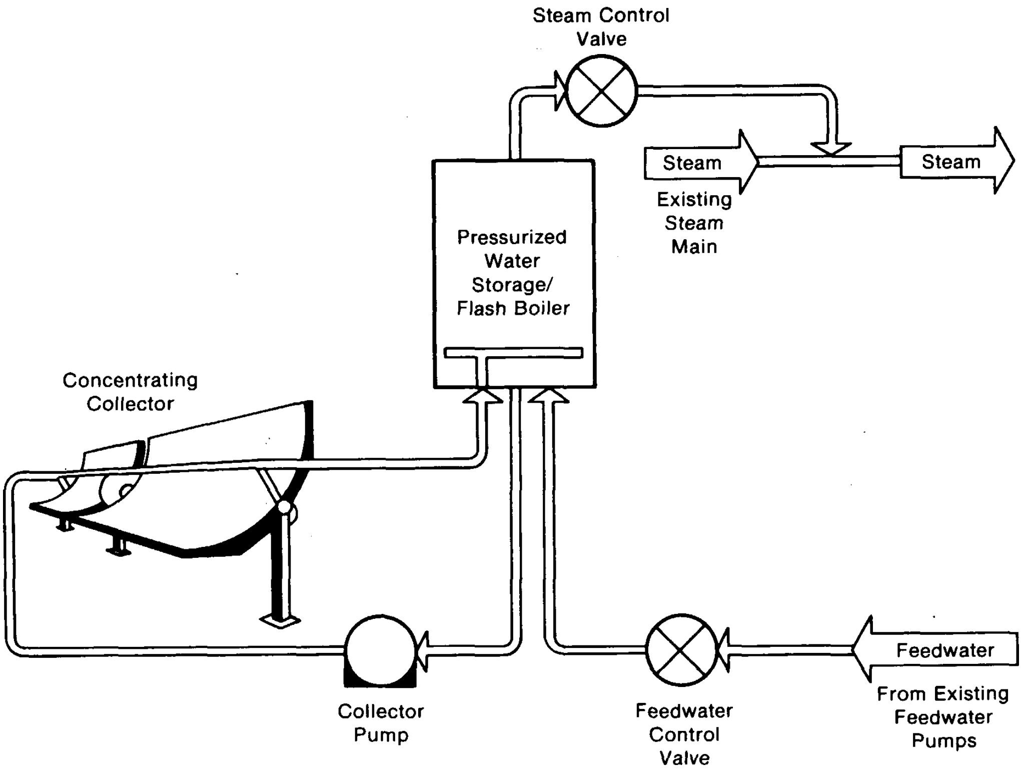


Figure 7. Solar System for Producing Industrial Process Steam

the total energy a hundred years ago. Biomass is also an attractive source for making alcohol fuel. Possible fuels from solar biomass processes are: (1) methanol—wood alcohol derived from wood or municipal wastes, and (2) ethanol—grain alcohol reduced by fermentation from agricultural products. Alcohols are in some ways superior to gasoline as a fuel: they are higher in octane rating and therefore burn cleaner than gasoline. One current use of alcohol is as a mixture with gasoline called "gasohol" which can power internal combustion engines for pumping in desalination. In 1980, Brazil substituted alcohol derived from sugarcane for a full 20% of the fuel requirements that would formerly have been met with gasoline.

1.3 Wind Energy

A fraction of the solar radiation incident on the Earth is converted by the atmosphere into the kinetic energy of winds. A wind turbine can convert the kinetic energy of moving air into mechanical energy. Wind energy is a natural solar technology that has provided power to man from the beginning of civilization. Wind power was also used extensively in the United States until recent times. More than 6 million small windmills (each less than 1 kW) were used in this country before the 1950's to pump water and to generate electricity on farms and in small communities. Modern designs to harness wind power cover a wide range of sizes and technologies. Deployment plans range from units for a single home to plants with the capacity of small fossil fuel plants. A typical wind generator capable of 1 megawatt output is shown in Fig. 8. Wind can be used for pumping or to generate electricity for those desalination processes that require electrical power.

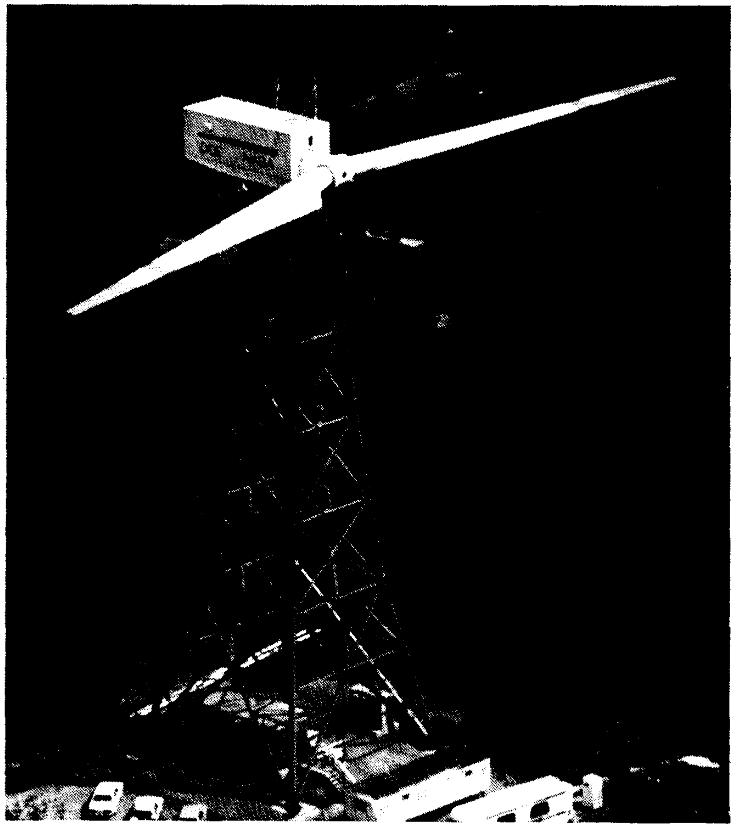


Figure 8. Wind Generator

1.4 Ocean Thermal Energy Conversion (OTEC)

Ocean Thermal Energy Conversion (OTEC) is another natural form of solar energy technology. OTEC systems tap the thermal gradients between various ocean depths. In a closed cycle, surface water heated by the sun vaporizes a working fluid such as ammonia. The pressurized ammonia vapor drives a turbine to produce electricity, which is then transmitted to land via cable. The vapor is condensed by colder water from the deep regions and again converted to a liquid. It is then ready to begin the cycle again [1,2].

The available amount of thermal energy within the ocean is large, but converting this energy in a closed cycle requires development of efficient, durable, and cheap heat-exchange equipment for evaporators and condensers that can withstand rough and corrosive conditions. The first closed-cycle demonstration plant began operation in 1980.

It is also possible to design OTEC systems operating on an open cycle, i.e., one in which seawater is the working fluid. Such a scheme can eliminate the need for one or both heat exchangers by using direct-contact heat exchangers. As shown in the flow diagram of Fig. 9, in this type of cycle the production of electric power can be combined with desalination in the condensation process. However, at present the cost of water is not enough to justify the efficiency penalty imposed on the power system and the frequently excessive cost of transporting the desalinated water.

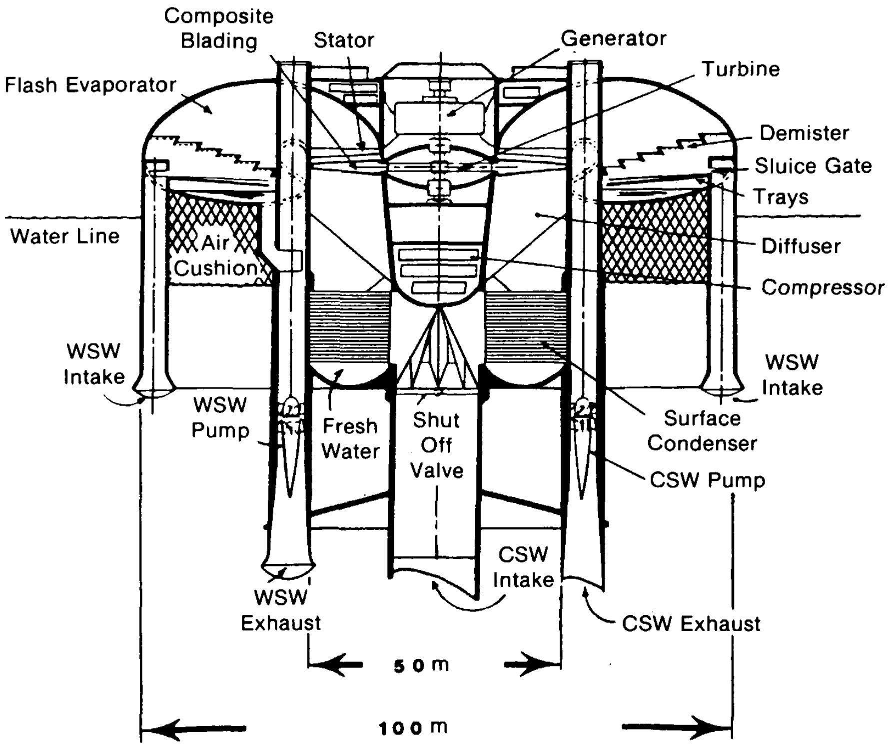


Figure 9. 100-MWe Open Cycle OTEC Plant Cross Section with Direct-Contact Evaporation and Condenser Desalination

1.5 Conversion of Solar Energy to Electricity

Conversion of solar energy into electricity can be achieved by a variety of methods. One is solar thermal conversion. Solar thermal conversion systems use concentrating solar collectors to generate steam which drives a turbine-generator to produce electricity. The most promising type of solar collecting system for electric power production is the central receiver, discussed previously. It consists of a large tower surrounded by a field of tracking mirrors (heliostats) which concentrate the sun's rays onto a boiler located in the top of the tower (see Fig. 3). Solar thermal conversion has some potential for centralized production of electricity, but for small-scale distributed installation e.g., for individual houses, direct conversion by photovoltaic means is much more promising.

Photovoltaic cells convert solar radiation directly into electricity. The conversion of solar radiation into electrical energy by solar cells is a direct spinoff of the space program. This simple, solid-state device requires no moving parts and holds the promise of long operating life with little need for servicing. It is expected that by 1985 solar cells will be used to generate electricity in a wide range of applications.

The principle of operation of a solar cell is illustrated in Fig. 10. Photovoltaic solar cells are made by combining two very thin layers of semiconductor material. Each layer is prepared by adding different materials to the major constituent, such as silicon, so that one layer has negative electrical properties and the other layer has positive properties. Terminals of an external electrical circuit are attached to the front and the back of the cell. Sunlight knocks electrons loose from some of the silicon atoms, leaving "holes" in the atomic structure. Because of the electrical properties of the cell, the increment of free electrons creates a voltage in the cell, and a current can be drawn through the external circuit. If a load, such as a DC electrical motor, is placed in the circuit, it will be operated by the current. Thus, sunlight (specifically, photons) is directly converted into electricity.

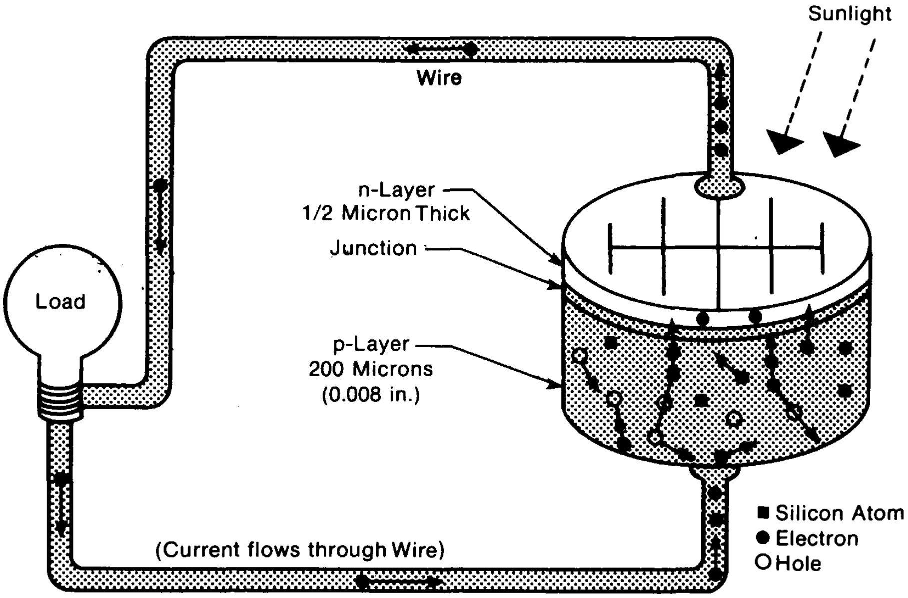


Figure 10. Principle of Photovoltaic Generation of Electricity

Solar cells are connected and placed into a sealed glass or plastic unit called a module. A number of these modules, each of which might deliver 12 watts, are then positioned into a rigid frame, called an array, to provide a specific power rating (Fig. 11). But since solar cells only deliver power during daylight hours, a photovoltaic system requires storage (usually a battery) if power is to be supplied at night or during cloudy weather conditions. Also, the power requires conditioning by an inverter to change the DC current to AC current and to regulate and keep the voltage output constant. Perhaps one of the most immediate applications for this technology is providing power in remote areas where there is no existing power grid.

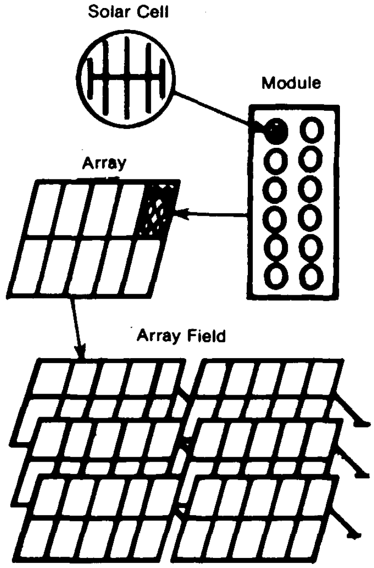


Figure 11. Schematic Diagram Showing an Array Field and Its Components

2 OVERVIEW OF DESALINATION METHODS

Water desalination can be attained by four main technologies: membrane, distillation, crystallization, and chemical [4]. In 1977, 23% of the capacity was handled by membrane processes and 77% by distillation. However, it is expected that in the next 20 years the majority of the plants will use either membrane or crystallization technology, because they are more energy-efficient than distillation.

2.1 Membrane and Crystallization Processes

Electrodialysis and reverse osmosis are the most advanced membrane processes. Additional membrane processes under study are transport depletion and piezodialysis. Transport depletion, a variation of electrodialysis that uses nonselective membranes, has lower membrane cost, but requires more energy. Piezodialysis uses pressure to drive concentrated brine through membranes. High production rates are anticipated, but demonstrations have so far been only at a bench scale.

There are four crystallization processes being studied: (1) vacuum freezing/ vapor compression, (2) secondary refrigerant, (3) eutectic freezing, and (4) hydrate formation. However, none has reached a commercial stage, and economic viability has not been demonstrated.

2.2 Distillation Processes

Distillation processes include: (1) single-effect, multistage flash distillation, (2) multieffect, multistage flash distillation, (3) vertical tube distillation, (4) horizontal tube evaporation, (5) vapor compression distillation, (6) single- and multi-effect solar stills, and (7) critical point separation. The first five systems show promise for the economical production of fresh water in relatively large systems.

The key factors to be considered in water desalination are the composition and the temperature of the feedwater, the product water quality desired, the water recovery ratio that must be achieved, the methods available for brine disposal, the plant utilization factor, the performance factor, and finally, the type of energy supply available [5,6,7].

The feedwater can range from brackish water having a total of dissolved solids of 1,500-20,000 mg/L to seawater having a total dissolved solids of 35,000-42,000 mg/L. The cost for membrane desalting is strongly dependent on the specific impurity composition and the salinity of the feedwater, but this is not the case for distillation processes. Conventionally distilled water costs from \$1.20 to \$1.50/m^3 in 1979.

The product water recovery ratio (i.e., the ratio of purified water output to feedwater input) affects the selection and design of desalting processes as well as the method for control of scale formation. Also, if an evaporation pond is used, its size is determined by the water recovery ratio.

The performance factor (PF) is an important number, because it indicates the energy required for desalination. It is defined as the mass of desalted water in kg per 2,320 kilojoules of energy input (the energy required to vaporize 1 kg of water is 2,320 kJ). Typical energy requirements and performance factors are shown in Table 1 for various processes when desalting brackish water. Desalting seawater requires more energy than desalting of brackish water. For water desalination with solar energy, the peak temperature required for the distillation process is important because it affects the type of collector needed. Moreover, the efficiency of solar collectors decreases as the temperature increases. For most distillation processes the minimum required temperature ranges from 60°-150° C.

Large-scale solar-powered desalination systems do not differ greatly from conventionally powered desalination systems, except that the fuel storage and boiler used in fossil-fuel-powered systems have to be replaced with solar collectors and storage subsystems. Large-scale systems are made up of subsystems that consist of feedwater pretreatment, solar collectors, an electric power generator, a distillation or membrane unit, an energy storage subsystem, a brine disposal subsystem, and product water storage and delivery subsystems. These subsystems must be combined in a system that results in the lowest levelized cost per cubic meter of desalinated product water. At present it seems that, the most economical systems are powered by a combination of solar and conventional energy. The solar energy fraction that provides the lowest product water cost ranges from about 1/4 to 1/2, depending on fuel cost and plan utilization factor.

Table 1. ENERGY REQUIREMENTS AND PERFORMANCE FACTORS FOR DESALTING BRACKISH WATER

	kJ/kg	PF
Reverse osmosis	30	77
Electrodialysis	50	46
Absorption vacuum freeze	63	37
Vacuum freeze/vapor compression	102	25
Multieffect, multistage distillation	125	19
Multistage, flash/vertical tube distillation	175	13
Vapor compression distillation	186	12
Vertical tube evaporator distillation	204	11
Multistage flash distillation	251	9
Single-effect solar still	4350	0.53

Feed water, 3500 mg/liter; product water, 500 mg/liter recovery ratio, 70%; and thermal-to-electric energy conversion efficiency, 25%.

Source: A Feasibility Study on the Use of Solar Energy for Desalination of Water. Prepared for The Texas Water Development Board, State of Texas. Contract No. 14-70028, 31 May 1977.

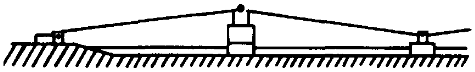
2.3 Solar Stills

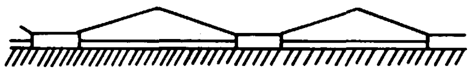
Solar stills [5] are not widely used for large-scale solar energy water desalination because they have a low output per unit area, and the land area available can be an important constraint. The production of good solar stills ranges from 1 to 4 liters/m^2-day as compared to 20-77 liters/m^2-day for other processes. Solar stills also suffer water losses because vapor and distillate leaks invariably occur in the system and, in addition, considerable maintenance is required to keep them operating. It is estimated that approximately 10 manhours are required per year for each 100 m^2. Also, the productivity of solar stills decreases over time for a variety of reasons. Finally, the cost per unit output is relatively high, particularly in industrial countries having high labor costs. The output of existing solar stills ranges from 0.24 to 2.5 liters/m^2-day, depending on location and on whether a single stage or several stages are employed. Despite these negative features, the solar still approach is attractive for certain small-scale operations, especially in remote locations.

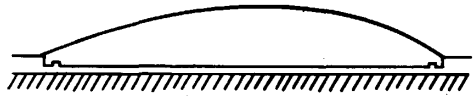
Water distillation systems may be single-effect or multiple-effect. In the simple, single-effect system, seawater is heated in shallow troughs, condenses on the sloped glazing and flows to a collection tank. Single-effect systems require an energy input equal to that necessary to vaporize the water because they do not recover any of the heat released by condensation. Multiple-effect systems recover a portion of the heat of condensation and are therefore able to provide more fresh water with a given collector area.

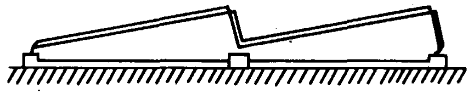
Figure 12 illustrates a number of different still designs for single-effect solar distillation. All of these use stationary solar collectors to supply the heat. The basic requirements of a good solar still are that the unit:

- be easily assembled in the field,
- be constructed of materials available in or near the region so that transportation costs will be small,
- be lightweight for ease of handling and shipping,
- have an effective life, with normal maintenance, of 10 to 20 years,
- not require any external power sources,
- can also serve as a rainfall catchment surface,
- is able to withstand prevailing winds,
- be manufactured of materials that will not contaminate the collected rain water or the distillate, and
- meet standard civil and structural engineering standards.

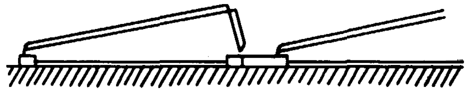

Basin-Type Solar Still


Greenhouse-Type Solar Still


Inflated Plastic Cover Design Still


Inclined Glass Cover Design Still


V-Shaped Plastic Cover Design Still


Variation of Inclined Glass Cover Still Design


Single-Sloped Cover Design Still

Figure 12. Approaches to Single-Effect Solar Distillation

A typical multi-effect distillation system is shown schematically in Fig. 13. Vapor from boiling brine in the first effect is condensed on the surface of heat exchanger tubes. The heat of condensation is then used to produce boiling in the cooling brine stream on the other side of the tubes, at a somewhat reduced temperature and pressure. The vapor produced by this secondary boiling is condensed on the surface of the second-effect tubes, causing additional vapor production in the third-effect cooling brine. The latent heat is thus recovered from effect to effect for additional distillation. The plant performance factor, i.e., the amount of distillate product obtained per unit heat input, is roughly proportional to the number of effects in the plant.

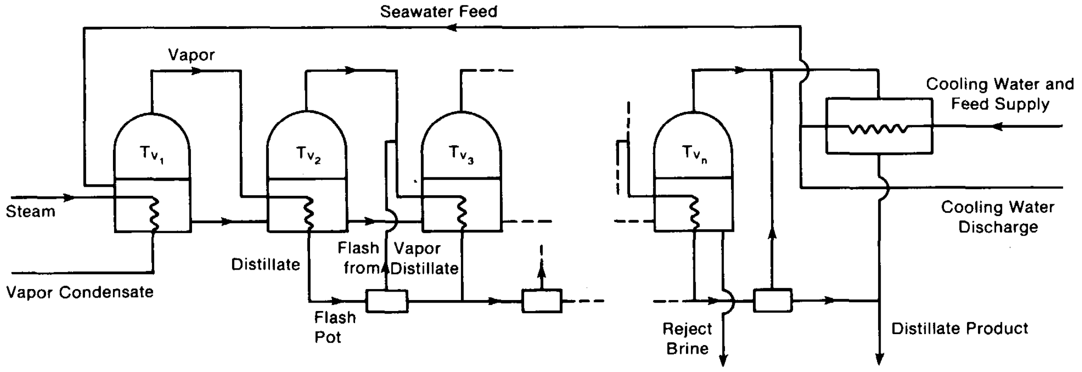


Figure 13. Schematic Diagram of Multieffect Design Scheme

Figure 14 shows operating data of some representative multieffect distillation plants. In a conservative design (Case I) brine is boiled at 93°C in the first effect. The latent heat is then thermodynamically degraded through 10 effects, thus producing 4.3 kg of desalted water per MJ (10^4lb per M Btu). The heat of condensation in the last effect is finally rejected to cool salty water.

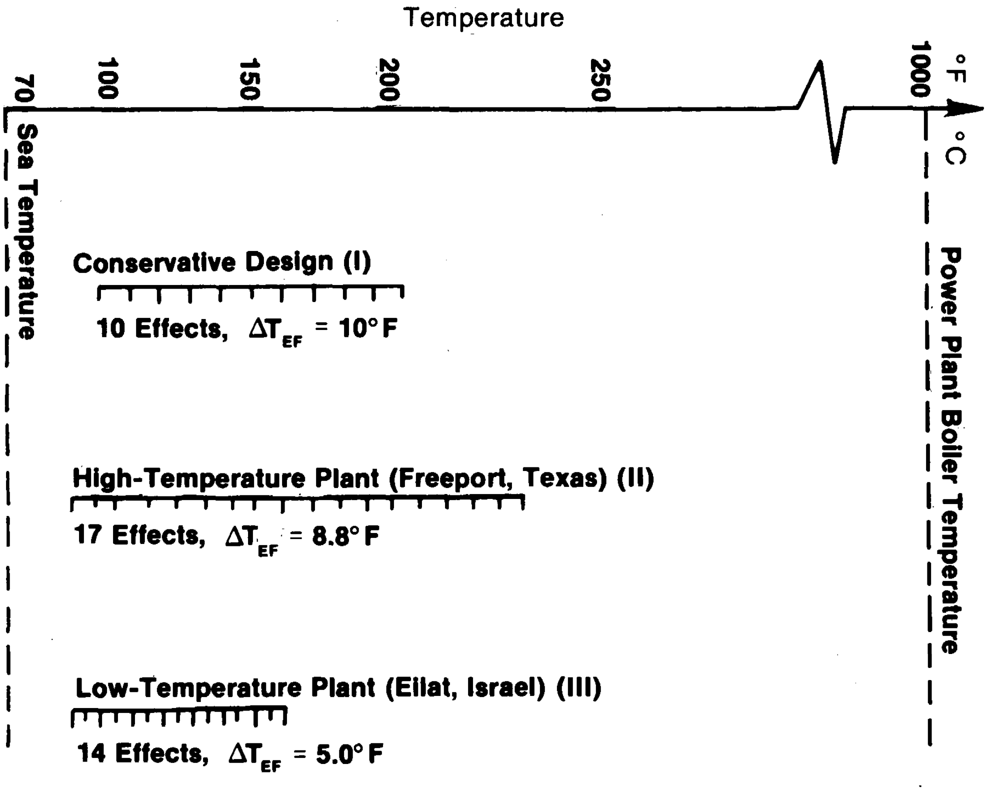


Figure 14. Operating Data of Some Multi-Effect Desalination Plants

The upper limit of temperature in the first effect is determined by scale deposition on the heat exchange surfaces. Chemical treatment can inhibit scale formation and permit operation at somewhat higher temperatures, allowing an increase in the number of effects. The number of effects can be further increased by designing for a reduced temperature drop per effect. This can be achieved by increasing the heat transfer area. Case II represents the multieffect demonstration plant in Freeport, Texas, which operated at a maximum temperature of 115° C. A 5° C temperature drop per effect permits a 17-effect design.

A further increase in steam economy ratio can be achieved by using stainless steel or Titanium tubes and by eliminating scale deposition on the tubes through scale seeding techniques. These techniques could permit a maximum brine temperature, of about 150° C with an economy ratio close to 20. This design requires, however, a complicated plant with exacting maintenance requirements.

Burning of fossil fuel for the sole purpose of heating brine to 120°-150° C in a desalination plant is very wasteful. Most of the seawater distillation plants have therefore been integrated with electric power plants. In such dual-purpose plants, high-temperature steam produced by fossil fuels is first passed through a turbine where energy is extracted for electricity generation. The thermodynamically degraded steam at the turbine exit is then condensed in the first effect of the desalination plant. The same approach could of course be used in with a solar power plant (see Fig. 3).

The combination of desalination and power generation into a dual-purpose plant changes profoundly the design strategy of the desalination plants. In a dual-purpose plant there is less economic advantage in increasing the maximum brine temperature in the desalting plant, since it leads to an increase in turbine backpressure with a corresponding penalty in power production. Conversely, the decrease in performance factor brought about by a reduction of maximum brine temperature in a dual-purpose plant is accompanied by a reduction in cost of the steam and a simplification in plant design and maintenance.

The Eilat plant in Israel, represented by Case III in Fig. 14, is an example of a plant that incorporates advantages obtainable with a low brine temperature. It operates at a maximum brine temperature of 85° C, which permits the use of cheap aluminum tubes and minimizes the need for chemical treatment of makeup seawater to prevent scale deposition. A relatively high economy ratio is obtained since the low cost of the heat exchange surfaces permits the use of a large heat-transfer area with a low temperature degradation for each effect.

Low-cost, low-temperature heat can also be obtained from solar energy in salty solar ponds. Figure 15 shows a desalination plant coupled with a solar pond. A stratified pond that is thermally stabilized to suppress convection currents has a

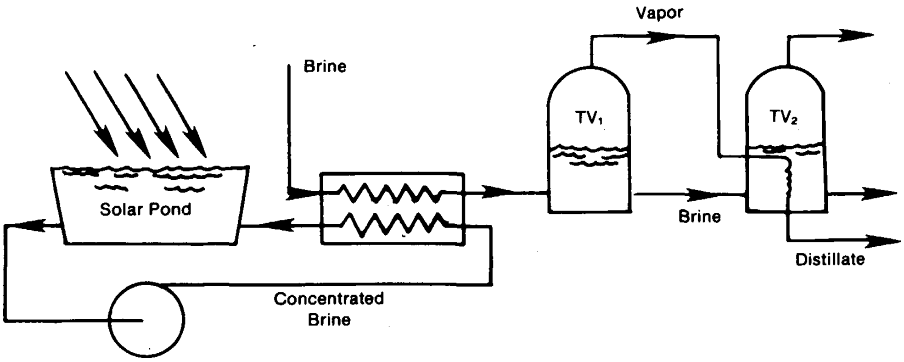


Figure 15. Schematic Diagram of Desalination System Using a Solar Pond as Heat Source

temperature profile as shown in Fig. 16. It can be seen that solar radiation trapped in the deep layer of the pond can produce appreciable temperatures. Brine from these layers can be extracted from the pond for heat delivery in a heat exchanger and then be returned to the pond at the appropriate depth.

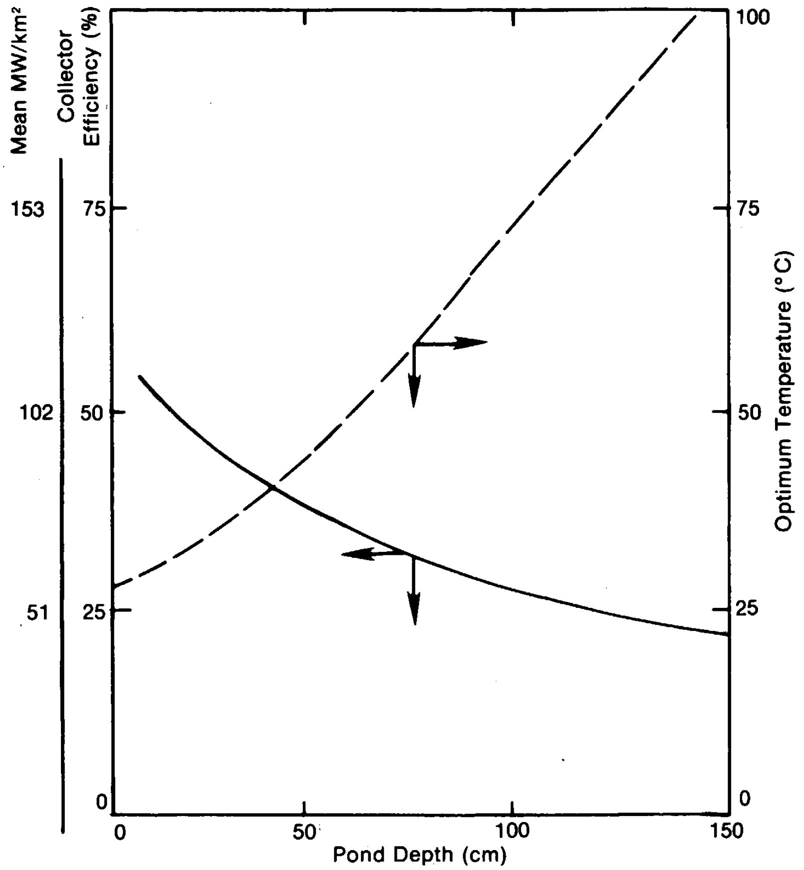


Figure 16. Temperature Profile in a Salt-Stabilized Solar Pond

3 SOLAR DESALINATION SYSTEMS

In a recent U.S. procurement activity for a commercial-sized solar energy water desalination plant having a capacity of 6000 m^3/day, 31 industrial teams representing 116 different companies responded. The resulting proposals contained combinations of 8 solar energy technologies with 8 desalination technologies, as shown in the matrix in Table 2. The various proposals were for 13 brackish water and 23 seawater plants.

Inspection of Table 2 shows that the main 8 solar technologies discussed in Sec. 1 can be combined with the principal 8 desalination technologies to form 64 possible solar desalination systems. Although not all 64 are technically viable, there are several systems that look potentially attractive, but without more R&D it is not possible to select the preferred operating mode. This presents a problem to any community that wants to build a solar system now.

Five contract awards have been made for system definition studies from the combinations in Table 2. They are expected to be completed by late 1981. It is then planned to construct one or two pilot plants either in the United States or in

Table 2. COMBINATION OF DESALINATION AND SOLAR ENERGY TECHNOLOGIES IN SERI PROPOSALS REVIEW*

Desalination Technology	Solar Energy Technology Central Receiver	Point Focus	Line Focus	Evacuated Tube	Photovoltaic	Wind	Solar Pond	OTEC
Electrodialysis					4	1		
Reverse Osmosis—Seawater	3	4	6		1	2	1	
Reverse Osmosis—Brackish	1	2	5		2	1		
Multieffect Distillation	1		2					
Multistage Flash Distillation	1		2		2	2	1	2
Vapor Compression		1	2	2				
Freezing		1			1		1	
Solar Still					1			

Note: Numbers in matrix are the number of proposals in each category.

*Source: W. Luft, Solar Energy Water Desalination Proposal Review Board Report to SOLERAS Project Selection Committee, Solar Energy Research Institute, SERI/SP-761-684, Golden, CO, May 1980.

Saudi Arabia for the systems with the lowest cost per unit of product water. The pilot plants will have capacities of 100 to 400 m^3/day of product water. It is anticipated that at least one pilot plant will be constructed by 1983. The desalination technologies selected for the initial system definition study are freezing and reverse osmosis—either alone or in combination with vertical tube distillation, multieffect distillation, or electrodialysis. The solar energy technologies included are central receiver systems, line- and point-focus collectors, wind energy, and photovoltaics. No solar pond system has been selected in this procurement, but for developing countries a pond with a multi-effect still may be the best near-term solution. Three of the five systems under study are for seawater desalination and two are for brackish water desalination. After these studies are complete it should be possible to make a selection of the most appropriate desalination system to meet any particular set of conditions.

ACKNOWLEDGEMENT

The authors would like to acknowledge the help of Prof. A. Kogan, the Technion, Israel, in the preparation of this paper. Section 2, Overview of Desalination Methods, has been adapted from a forthcoming book entitled A Global Assessment of Renewable Energy: Resources, Technologies and Applications (by F. Kreith, R. Larson, and F. Arnold) to which Professor Kogan contributed on the topic of desalination.

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NOTES

COMMENTS ON WESTERN UNITED STATES WATER RESOURCES

J. W. O'Meara
Executive Vice President
National Water Resources Association

ABSTRACT

The paper provides brief comments about the water resources of Saudi Arabia and discusses in general terms the Western United States water problems. Emphasis is placed on the Colorado River Basin and the salinity problems of the region. The background information leading to the decision to build a 100 million gallon per day desalting plant at Yuma, Arizona, is reported.

The author also recalls some of the early solar humidification research and development sponsored by the Office of Saline Water.

WATER PROBLEMS OF SAUDI ARABIA

For a good many of the years that I was employed by the Office of Saline Water, HRH Prince Mohamed al Faisel was the Governor of the Saline Water Conversion Office of the Kingdom of Saudi Arabia.

Prince Mohamed once told me that the Saudi Arabian Peninsula was about the same size as the United States east of the Mississippi River, but, he noted, in that vast area there was not a single flowing stream of any size.

The average rainfall in the United States is about 30 inches per year, ranging from about 11 inches in Nevada, Arizona, and Southern California, to just over 50 inches per year in Florida, Georgia, and the Carolinas.

The total water supply available in the United States is 1200 billion gallons per day (gpd). I do not have available comparative data for Saudi Arabia, but I think everyone in the room is aware that there is a water problem in the Kingdom. Every time they drill a water well, they hit oil. Saudi Arabia is about in the center of a huge desert that extends eastward from the Atlantic coast of Africa to India. Parts of Saudi Arabia receive only two or three inches of rainfall in an average year and, after that, needed moisture can come in a single storm. However, the Asir Mountains on the southeast coast rise to an elevation of about 3000 meters, and, in this area, the rainfall reaches about 20 inches per year. Unfortunately, the Asir is mostly volcanic rock, the mountains are steep and close to the sea, which results in high runoff and very little ground water recharge. Some of the streams that are formed by this seasonal rainfall actually evaporate before they reach the sea.

That brings me to an interesting situation. The Bible says: "All the rivers flow into the sea, yet the sea is not full." Since I have just reported that the rivers of the Asir do not reach the sea, perhaps we could conclude the Bible is wrong: Someone suggested to me that the Bible is not a proper citation since the people in that region read and live by the teachings of the Koran!

WATER RESOURCES OF THE COLORADO RIVER BASIN

I would cite an interesting parallel. The Colorado River, the major river in the southwest United States, which has average flow of almost 13 billion gpd, no longer flows to the sea.

On the Colorado and its tributaries, the Department of the Interior has constructed six major dams. These dams collect and store the precious waters that flow down the mountains in the spring as the warming sun melts the winter snowpack. There is an old saying here in this part of the country: "It doesn't rain in the West, it snows." Thus, if man does not collect and save the spring runoff, there would be very little productive agriculture in this region.

The water stored in these reservoirs is used to generate electricity; it is used for irrigation and recreation; and it is used to enhance fish and wildlife habitat. Some is transported nearly 250 miles for part of the municipal supply of Los Angeles and San Diego. In fact, the aqueduct from the Colorado River to Los Angeles transports about 700,000,000 gpd. In a few years, with the completion of the Central Arizona Project, an additional 1 billion gpd will be delivered from the Colorado to the thirsty cities of Tuscon and Phoeniz, Arizona.

UNITED STATES-MEXICO WATER PROBLEMS

The United States has a treaty with Mexico which was signed in 1944. In that treaty, the United States agreed to guarantee a flow of 1.5 billion gpd across our border into Mexico. If you have visited Yuma, Arizona, and I know many of you have for reasons I will discuss in just a moment, you probably have visited Morelos Dam, which blocks the flow of the Colorado River just across the border in Mexico. Morelos is a diversion dam, and it now marks the end of the Colorado River. It no longer flows into the Gulf of California. Every remaining drop of water in the river is diverted to the Mexicalli Valley for irrigation. There is no return flow. Below the dam, what was once the bed of a mighty river, is now dry land and covered with scrub trees and brush.

The Bible is wrong--the Colorado River no longer reaches the sea, but it represents a marvelous example of 100 percent utilization of a precious natural resource.

The vast region of the West, which the Colorado River drains, has an abundance of various kinds of salt-laden land and some rather large salt springs.

For example, LaVerkin Springs on the Virgin River, a tributary of the Colorado, has a daily flow of about 8 million gpd of 10,000 parts per million (ppm) water. Here in Colorado the Glenwood Dotsero Springs contributes about 11 million gpd of 14,000 ppm water.

The construction of reservoirs on the Colorado, and the extensive use of this water for irrigation, has had a double adverse impact on the salinity problems of the river. Irrigation return flows to the river are more saline than when they were diverted because they pick up salts as they flow across the saline soils. There is also increased evaporation from reservoir surfaces which, of course, means the remaining water is higher in salinity.

Water diverted from the river at Imperial Dam, the last diversion point on the river in the United States, has an average salinity of about 900 ppm. Some of this water is delivered to the Wellton-Mohawk Irrigation District near Yuma, Arizona. The return flows from the irrigation district to the river are about 3,200 ppm. This slug of salinity flowing into the river just above the United States border increases the overall salinity of the river as it flows into Mexico to approximately 1,250 ppm.

RESOLUTION OF UNITED STATES-MEXICO DISPUTE

Although the treaty between the United States and Mexico determined the volume of water that would flow into Mexico, it was silent on the question of water quality. As developments continued north of the border, the quality of the water flowing into Mexico continued to deteriorate.

The quality of the water they were receiving became a major issue with the Government of Mexico, and, during a state visit to Washington, D. C., by President Echiverria, it was an issue that was listed high on the agenda of the subjects to be discussed with President Nixon.

As a result of that meeting, President Nixon pledged to the Mexican President that the United States would take steps to improve the quality of water flowing into Mexico so it would be equal in quality to the point of last diversion in the United States, and that point, as I mentioned earlier is Imperial Dam located about 15 miles up river from Yuma, Arizona.

Without going into all the studies the United States undertook to achieve this promise, suffice to say, it resulted in the decision to construct a 100 million gpd desalting plant near Yuma to reduce the salinity of the Wellton-Mohawk return flows to a low enough salinity to meet the terms of the new agreement with Mexico.

This plant, which will use reverse osmosis technology, is now under construction and is scheduled for start up in 1984.

The Yuma plant will be the major United States desalting plant, and it will continue to be our major plant for many years to come because, as of now, there is nothing on the drawing boards in the United States that will challenge its size.

COMMENDATION FOR THE EUROPEAN FEDERATION OF CHEMICAL ENGINEERING

The European Federation of Chemical Engineering (the counterpart of the American Chemical Society) established a Working Party on Fresh Water from the Sea in the early 1960's. This past September, it conducted the 7th International Symposium on Fresh Water from the Sea in Amsterdam. I recall, with considerable pleasure, that I had the opportunity, at the 4th Symposium which was held in 1973 in Heidelberg, Germany, to announce the United States' plan to build the Yuma desalting plant.

What I recall most clearly was my subsequent discussions with the delegation from the USSR. They were convinced I had slipped my decimal point one or two places. A 1 million gpd plant they were ready to accept, a 10 million gpd plant was incredible, and a 100 million gpd plant was sheer propaganda!

In passing, I would like to say that the seven International Symposia, sponsored by the Working Party on Fresh Water from the Sea of the European Federation of Chemical Engineering, have contributed substantially to the development and dissemination of saline water conversion research throughout the world through stimulating exchanges of information. I want you to know that the motivating force behind that important effort is the team of Dr. Anthony and Dr. Eurodyke Delyannis of Athens, Greece, who are with us today. Not only has Dr. Delyannis chaired all seven symposia, he and his lovely wife have conducted some of the world's most valuable solar humidification research.

SOLAR DISTILLATION PROGRAM OF OSW

Along with the Delyannis' effort, I would like to mention a few other researchers who contributed substantially to the solar humidification program conducted by the Office of Saline Water (OSW) in the late 1950's and early 1960's. Even if you have not met these people, but are interested in solar humidification, I am sure these names will be familiar to you for you no doubt have read their reports.

George Löf, whose home is right here in Colorado, conducted extensive developmental work; Dr. Maria Telkes of New York University; Everett

Howe of the University of California at Berkley; Carl N. Hodges from Arizona; Frank Edland from DuPont; Bjorksen, Duffy, and Smith from Wisconsin; and Jim Eibling from Battelle Memorial Institute in Ohio. Joseph J. Strobel was the leading staff proponent of solar research and development in the Office of Saline Water. In the early 1970's Joe awarded a contract to Battelle Memorial Institute to prepare a manual incorporating a summary of the best results obtained from the research sponsored by OSW, and to recommend the most cost-effective system that could then be designed.

The general conclusion of the report--Research and Development Progress Report No. 546--was that solar distillation was competitive with other desalting systems in plants up to 20,000 gpd in areas where skilled operators were not available and fossil fuels were expensive or unavailable.

I don't know what has happened to advance solar technology in the decade that has passed since OSW wrapped up its solar R&D program, but I do know what has happened to energy costs. I suspect the economic viability of solar processes has substantially improved, which, in turn, may underscore the feasibility of this meeting.

TECHNOLOGY TRANSFER

Time has a way of changing things. The sale of OSW R&D Progress Reports on solar work had dropped off to almost zero. Then the United States suddenly got interested in the development of energy sources other than oil, and the old OSW solar reports suddenly became best sellers. OSW had more solar data in its reports than existed anywhere else. More money is being invested annually in solar energy development than OSW spent in 15 years on its solar humidification program.

Just as those who began a crash program on the development of solar energy were able to learn much from the work conducted in the desalting program, so too may the desalters be able to take advantage of advances in the solar energy program.

CONCLUSION

I congratulate the officials of the Government of Saudi Arabia and the U.S. Department of Energy for arranging this program. I think it is timely, and I hope it proves fruitful.

I would like to conclude with an old adage which I think is timely and appropriate. If I have a dollar and you have a dollar, and we exchange dollars, we each have one dollar. But, if I have an idea and you have an idea, and we exchange ideas, then we each have two ideas.

* * * * *

BRACKISH WATER DESALINATION

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BRACKISH WATER DESALINATION

The use of desalination processes for treating brackish groundwater has been actively pursued for more than 25 years in the United States. Although a strong effort has been made by the U.S. Government through the Office of Saline Water, later OWRT, and other agencies, an equally strong effort has been made by American industry. As a result, hundreds of brackish water desalination plants now exist throughout the United States and the world. The desalination processes have developed to the point where brackish water desalination is now considered a reliable "state-of-the-art " technology.

In the initial phases of developing a brackish water resource, the need for comprehensive advance planning is extremely important. Careful planning to guide the project from conception to final treatment and distribution will result in a project with minimum difficulty. In developing a good plan, one must first audit surface and groundwater resources, energy available to the area, water distribution system, and quality of water required. The water quality necessary to meet the local requirements should be considered carefully since local requirements, particularly in underdeveloped countries, may differ slightly from World Health Organization standards.

Thorough investigation of potential alternative supplies is a key planning task. In order to properly conduct this phase of the study, one must first obtain a good working knowledge of the area hydrology and water quality. Both the quantity and quality of surface and groundwater vary considerably from place to place throughout the world. During a recent evaluation of 45 potential sites for desalting plants sponsored by the U.S. Office of Water Research and Technology, a wide spectrum of water problems was examined. The potential solutions to these water supply and quality problems were remarkably varied and quite site specific. Although desalination could have been a solution at any of the 45 locations, it may not have been the most economic in all cases.

To illustrate the wide variety in water quality which exists in the United States, the compositions experienced during the study ranged from mildly brackish (for example 1,000 ppm) of predominately sodium chloride, to several thousand ppm predominately calcium sulfate (Table 1). The sites examined varied from coastal to inland and included both rural and urban areas. It was further determined that localized areas were stricken with specific problems such as an excess of fluorides or nitrates. Because of this wide variety of potential problems, good water quality data must be obtained which includes complete analysis for major species as well as problem species such as fluorides and nitrates. In addition, certain ions are known to create additional problems for desalination processes and these ions should also be analyzed.

Because of the age-old controversy of riparian rights, regional or area-wide solutions for water problems are becoming more and more difficult, especially here in the United States. These problems, in fact, are even more pronounced in the eastern states than they are in the west. Desalination processes are particularly adaptable to potable water production at specific site locations. One of the strong advantages of membrane desalination processes in particular, is that the desalting plant can be constructed in a relatively short period of time. Consider for example, the recent construction of a 3 mgd plant to desalt seawater at Key West, Florida. The plant took approximately 7 months to construct and install. It began operation in December 1980. Generally speaking, seawater plants are somewhat more difficult to design and construct than are brackish water plants. It is reasonable to expect that a brackish water plant could also be constructed in this short time period. In fact, for smaller size desalting plants, many manufacturers have so called packaged systems that are predesigned and can be manufactured in an extremely short period of time.

A key function for a responsible consulting engineer is to achieve a minimum cost solution for a particular problem for the client. Some of those minimum-cost solutions for water treatment are:

1. Minimum treatment (for example chlorination plus simple filtration).
2. Intermediate treatment (for example iron and manganese removal plus filtration and chlorination).

In areas where only brackish water is available however, these simple solutions will not result in a satisfactory potable supply. For these situations, various desalination processes can be utilized.

Desalination Processes

The commonly-used desalination processes for brackish water are ion exchange, electrodialysis and reverse osmosis. The selection of one of the desalting processes depends on several factors which include:

1. Salt concentration
2. Required product water quality
3. Brine disposal facilities
4. Pretreatment required
5. Presence of bacteria, viruses, or other similar material
6. Availability of energy and chemicals

Ion Exchange

Although the ion-exchange process has been widely used in the United States for many years, it has been primarily utilized for softening municipal water supplies. It is rarely used however, for desalting a municipal supply. There are numerous reasons for the limited use of ion exchange for desalination, but the most significant is that other processes can accomplish the same objective at lower total cost. This is particularly true at higher TDS levels since the cost of operating ion-exchange systems (primarily regeneration costs) is directly related to the quantity of salt removed. In addition, the brine created when regenerating the resin has a very high salt content which creates a disposal problem.

New ion-exchange resins are constantly being developed however, which have specific uses in municipal treatment. One such resin was recently developed during conduct of an EPA sponsored development program in McFarland, California for removal of nitrates. A new resin specific to nitrates will remove nitrates to acceptable levels at minimal cost.

Electrodialysis

Electrodialysis is perhaps the oldest process utilized for desalting municipal water supply. There are several municipal plants in operation in the United States desalting brackish water. A 3-mgd plant located at Foss Reservoir in Oklahoma began operation in 1975. One of the oldest plants was constructed at Buckeye, Arizona with a capacity of 650,000 gpd in 1962. Many other electrodialysis plants are in operation throughout the world including a 5-mgd plant in operation in Benghazi, Libya.

Electrodialysis is generally conceded to be most effective when the mineral content of the water ranges from 1000 to 3500 ppm TDS. It is the best established of the membrane processes and in 1977 approximately 42 mgd of purified product water were being produced by electrodialysis desalting plants.

Although the basic process of electrodialysis has been well described in various technical journals and in previous workshops, perhaps a brief description is in order. The process is an electro-chemical separation which utilizes pairs of membranes, one selective to cations and the other selective to anions. The membranes are placed alternately to form compartments. A direct-current source is placed across the resulting compartment. As saline feedwater is introduced to the compartment, the cations travel toward the cathode and the anions travel toward the anode. Since the membranes are alternating anion or cation selective, every other cell becomes concentrated with the ions, while the alternate compartments become depleted resulting in demineralized product water.

The amount of energy required for the process is directly related to the quantity of ions removed from the saline solution. The conductance of saline water increases with temperature, and the process becomes more efficient with increasing temperature up to the present commercial limit which is about 110 to 160° F. Since process efficiency is improved by thorough mixing in the compartments, a plastic spacer is introduced to the compartment which causes water to flow through the compartment with hydraulic turbulence.

Minimal pretreatment is normally required. Certain materials, however, when present in the feedwater, adversely affect the process and should be removed with some type of pretreatment. Iron or hydrogen sulfide at greater than 0.3 ppm, manganese at greater than 0.1 ppm, and free chlorine should be avoided. In addition, the turbidity should be maintained at a level less than 2 JTU.

Recently, a significant improvement in the process has been developed which is called the electrodialysis reversal process referred to as EDR. The EDR process achieves cleaning and descaling of electrodialysis membranes several times per hour by periodically reversing the direct-current flow through the electrodialysis membrane compartments. Simultaneously, the product and waste streams are interchanged.

The EDR process has a unique advantage since it can result in prevention of calcium carbonate and calcium sulfate scale formation without the addition of acid or complexing chemicals to the feed stream. Acid or chemical addition is normally required for electrodialysis or reverse osmosis for water containing calcium and sulfate. The elimination of acid or complexing chemicals addition to the feed stream is a significant improvement in membrane processes and should increase the reliability of the process. Furthermore, in many remote locations, acid and chemical transportation is difficult and usually expensive.

The state-of-the-art of electrodialysis systems has sufficiently progressed to the point that "packaged" units are readily available. Typical operation and maintenance costs for packaged plants may range from 35 to 52 cents per 1000 gallons of product water for a 1500 ppm water producing 150 ppm TDS. About 48 to 63 cents is a typical range for a 3000 ppm feedwater prodcing 300 ppm TDS. Capital equipment cost for packaged systems range typically between 55 cents to $2.00 per gallon of installed capacity. The capital costs for large plant installations are somewhat less.

Reverse Osmosis

The reverse osmosis process is one which has been widely acclaimed as elegant in its simplicity. A driving force of hydraulic pressure is utilized in forcing water molecules through a semipermeable membrane. The membrane allows water molecules to pass, but rejects most of the dissolved ions. Multivalent ions are better removed than monovalent ions. A wide variety of membranes exist which allow the user to obtain the desired efficiency of demineralization.

The first semipermeable membrane with truly commercial potential was developed at the University of California at Los Angeles in 1961. Since that time, membranes have improved remarkably. Commercially available membranes today are capable of desalting seawater as well as selectively desalting certain water supplies. Low-pressure membranes which operate at nominally 200 psi have been developed for brackish water desalting applications.

The traditional cellulose acetate membranes are still widely used today and will continue to have good utility in desalination operations. Polyamide membranes in hollow-fiber form have now been used commercially for more than a decade and have proven to be reliable and efficient. The newest technology to reach the industry is the thin-film composite membrane wherein a very thin polymer film is deposited onto a porous substrate at which interface the two polymers are interacted. The result is a very thin desalination film which allows improved desalination characteristics including higher productivity.

Reverse osmosis systems are basically simple. A high-pressure pump is required to raise the pressure within the system to a range of 200 to 1000 psi depending on the feedwater source and membrane type. The pressurized feedwater is introduced to a reverse osmosis membrane element which provides the desired separation. A throttling valve or energy recovery system is provided on the concentrated stream to maintain pressure within the system. This basic unit is then duplicated to provide multistaged systems to allow higher water recovery, or alternatively, water may be recycled from the concentrated side to the feed side to provide higher recovery of potable water.

The type of pumps, materials of construction, and membranes selected, are dependent upon the choice of either brackish water or seawater for the feed source. The instrumentation and controls for reverse osmosis plants can be as basic or sophisticated as the client requires. Some plants are completely manually operated, whereas a recent municipal plant installed in the United States of America is almost totally automated.

Most reverse osmosis systems are susceptible to fouling by suspended materials in the feed source. The type of pretreatment required for the system will depend greatly upon the source of the feedwater. Common sources are surface supplies, groundwater wells, and seawater. Special considerations must be taken when certain problem materials are present in the feedwater.

Posttreatment of reverse osmosis product water is generally simple and normally requires degasification to remove carbon dioxide formed by feedwater acidification, passivation of the product water to prevent corrosion in the distribution system, and chlorination to prevent bacteria growth.

Several considerations must be taken in properly selecting the site. Generally, reverse osmosis systems are quite compact and little land area is required. The membranes need only to be protected from natural elements such as the sun, wind, and freezing. Reverse osmosis depends on the utilization of electrical energy to provide power to the high pressure pumps. The availability of electrical energy as well as consumable chemicals and membranes, should be considered in site selection, as well as disposal of the brine which will result.

Few environmental and health problems are expected when using a reverse osmosis system. The environmental effects caused by degasifying carbon dioxide and possibly hydrogen sulfide are minimal. High-pressure pumps generally exhibit a fairly high noise level and precautions should be taken to muffle this noise, particularly if diesel generation is utilized in combination with the pumps. In certain locations in the world, the environmental effects of brine disposal should be carefully considered. Generally, reverse osmosis membranes will remove bacteria and viruses. However, bactericide addition to the product and periodic cleansing of the elements is recommended.

The economics of the reverse osmosis process are generally conceded to be attractive for desalination. The actual range of capital costs ranges considerably depending upon the salinity and source of the feedwater. Recent desalination plants have been as low as 50 cents per gpd installed capacity, to as high as $11.00 per gpd installed capacity for the recent seawater desalination plant built in Jeddah, Saudi Arabia. Operation and maintenance costs can be as low as 60 cents per 1000 gallons product water, to as high as $4.00 to $6.00 per 1000 gallons product water for seawater systems. Energy consumption also varies depending upon the salinity of the incoming feed source. Consumption can be as low as 7 kwh per 1000 gallons, to 33 kwh per 1000 gallons. Recent innovations in energy recovery systems will reduce the net energy consumed in the process.

Reverse osmosis is a particularly versatile process and can be used for a wide variety of applications. Desalination of brackish water and seawater are the most obvious of its many uses. In the United States, perhaps the most widely accepted industrial utilization of reverse osmosis is production of high-purity water for the semiconductor industry. Other applications include industrial water reuse, cooling tower blowdown concentration, various industrial wastewater processing and municipal wastewater processing.

There are several mechanical designs which provide the support for the semipermeable membranes. They are tubular, spiral wound, and hollow-fine fiber. Each of these designs has its advantages and disadvantages depending upon the particular feedwater composition. The tubular system possesses an open-flow channel through which suspended solids may pass without becoming fouled within the system. So called "dirty" feed streams can therefore be processed with minimal or no pretreatment. Its principal disadvantages are that it has a very low membrane surface area to volume ratio compared to other designs and is relatively expensive. The spiralwound design is similar to a "jelly roll" and utilizes membrane in sheet form. This configuration allows the utilization of many different types of membranes that are manufactured in sheet form. The design has a much higher membrane surface area to volume ratio and is much less expensive to produce the same volume of water. One of its disadvantages however, is that the flow passage is much smaller and contains a plastic turbulence promoter which further restricts flow. The hollow-fine fiber design is made of a bundle of hollow-fine fibers of "hairlike" diameter. This design has an extremely high membrane surface area to volume ratio and is also inexpensive. Its principal disadvantage is that the saline water passages (between fibers) are quite small and excellent pretreatment is required.

There are obviously other advantages and disadvantages of the three processes which have been well described in industry publications. The most significant advances made in the development of reverse osmosis components in the last 10 years has been the development of new membranes. Approximately 10 years ago, the choice of membranes was either cellulose acetate (or cellulose triacetate or blends), or the polyamid hollow-fine fiber.

Today, a wide variety of membranes exist including new cellulosic formulations, aramid, polybenzimidazolone, and other polymers. The most recent development is the thin-film composite membrane wherein a thin polymer coating (for example polyamide) is placed on a microporous polysulfone support layer. At least two companies commercially offer thin film composite membranes. The membranes exhibit improved salt rejection (up to 99.5 percent), higher flux, and have a wider useful operating pH range (for example 4 - 11). It is believed that the membranes resist biological attack and resist compaction. Feed temperature up to 60°C can be used without membrane damage. Although some thin-film composites are damaged by chlorine, newer formulations have been developed which tolerate chlorine.

Energy

In addition to the cost of membrane replacement, the other major cost factor in operating electrodialysis and reverse osmosis plants is the cost of electrical energy. It is in this critical area of operation that solar energy could very well play a significant future part for both processes. The SOLERAS program has taken a bold step toward merging solar technology with desalination technology. The results of these study programs and resulting pilot programs will be closely studied by the entire desalination world.

Other efforts are ongoing on a smaller scale. It was reported at the 1979 meeting of the International Congress on Desalination and Water Reuse, that a small 2.5 kw solar generator was combined with a 1.5 m^3/d reverse osmosis system. The project is a joint effort between the Mexican organization, DIGAASES and the German organizations, AEG-TELEFUNKEN and GKSS. This pilot work is being carried out as a result of a scientific and technical cooperative agreement between Mexico and Germany.

As a practical matter, small units utilizing solar energy are a needed requirement today in various parts of the world. This type of "self-supporting" system could be used effectively for example, in the southern parts of Tunisia where water is brackish and electrical energy is not available. A solar-powered process is being considered for use in arid regions of the state of Texas. It takes little imagination to estimate the vast areas of the world which could take advantage of solar powered desalination processes.

It should be noted that wind energy is also a consideration in many parts of the world and a combined solar and wind-powered desalination plant is certainly a possibility.

Summary

Providing water supply of adequate quality and quantity for the 1980's will require innovation as well as the utilization of many different disciplines. These disciplines may include hydrology, geotechnics, water rights law, water distribution, large project financing, water treatment, waste disposal, alternative energy technology, desalination and others. The role of solar energy combined with desalination techniques for desalting brackish groundwater will be a significant one in the next decade. Good state-of-the-art expertise already exists in both solar energy and desalination technologies. The current effort to combine the two into a meaningful and practical working process is a most important one. As our population continues to grow and potable water becomes even more scarce, thorough evaluation of all possible water supply and quality improvement alternatives will become exceedingly important.

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OTHER IMPAIRED WATERS:
OPPORTUNITIES FOR COST EFFECTIVE APPLICATIONS OF DESALTING PROCESSES

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ABSTRACT

Industrial wastewaters can be reused with varying degrees of treatment back in the industrial operation. Reuse in this manner frees other water sources in an area for other use, effectively increasing the high quality water resource. Treatment of industrial wastes for discharge usually involves oxidation of organic materials, precipitation or absorption of metals for handling with the other waste solids, and neutralization of the acids and bases. Desalting processes can treat the same wastes and yield a higher quality product, if necessary, or provide a specific removal function if no more is needed for recycle. Since desalting processes are usually more expensive than conventional treatment, before these processes will be adopted they must produce some treatment.

Industries were selected for analysis to define how the desalting processes produce this added benefit. Applications for desalting processes that have been proposed, evaluated, and/or implemented are discussed. Waste characteristics from the industries are discussed to illustrate the advantages offered by desalting processes. Limitations of various processes are described to illustrate how technical development would make desalting processes more adaptable to the needs of industrial waste renovation.

INTRODUCTION

Impaired Waters

An "impaired water" is generally not suitable for direct reuse by municipality or industry. Conventional treatment provides a quality suitable for return to the environment where further natural purification and dilution allow its reuse. This results in sequential degradation of the total flow. In areas where the water flow to usage ratio is too small the ability to naturally purify and dilute are exceeded. This is illustrated for dissolved salts, where there is no destruction and dilution ability has been exceeded, by the Colorado River basin. Examples of natural water sources whose ability to oxidize biodegradable wastes have been exceeded are numerous.

Industry Selection

One way to expand an areas water supply is to remove or restrict some of the users. Another is to pretreat all discharges so that any return doesn't degrade the receiving body. The largest users, where most significant consumption decreases can be made, are industries with either a large water use per unit output or a large output, or both.

The industries discussed here were selected on several factors:
- volume used by the total industry.
- quality of discharge, either requiring extensive treatment for disposal or having high value by-products.
- basic industries of interest to nations of diverse development.

The U.S. Department of Commerce develops statistics on resource utilization in different industies. One of those statistics is water use. The largest industrial user is the inorganic chemical industry[1]. The industry is so diverse that a generalized discussion is not possible and it therefore is not discussed here. The next largest water users that were selected are:
- iron and steel making.
- petroleum refining.
- pulp and paper making.

Electric power generation was also chosen, even though it is not treated directly in the Commerce Standard Industrial Code (SIC Code). It was selected because of the large consumptive use for cooling, a need common on a smaller scale to most industries, and the fact that it was the first industry to be forced to zero discharge on a large scale. Metal finishing was also selected. It was the first industry to thoroughly evaluate and adopt all commercial desalting technologies. Finally, mineral and non-ferrous metal mining was selected. This is an industry that is common to most countries.

Economic Considerations

Environmental restrictions require treatment of wastes before returning them to a natural water body. They are designed to use conventional technology to optimize the cost and benefit relationship. Application of desalting equipment in these applications will generally be more expensive. To be selected they must provide some higher benefit that can be quantified. Factors to be considered include:
- by-product recovery, either directly or by concentration for recovery by another process.
- concentration of pollutants into a smaller volume for impounding rather than destruction.
- higher quality water for reuse in the plant that would otherwise not be possible.
- selective treatment for recycle in the plant for a use that doesn't require the quality of fresh make-up.

The various desalting processes provide these benefits to varying degrees. Their costs vary as well, so any analysis of new applications

TABLE 1.

DESALTING PROCESS SUMMARY

PROCESS	PRETREATMENT	PROCESS CAPABILITIES				ENERGY CONSUMPTION			RELATIVE CAPITAL COST (0 to 10)
		max. brine conc.	can it cryst-al'ze?	PROD. QUALITY		low TDS	high TDS	equiv't primary fuel[1]	
				elec-'lyte	organ-ics				
EVAPORATION									
-multistage	deaeration	25%	yes	5 mg/l	varies	12[3]	12[3]	85/85	10
-multieffect	deaeration	25%	no	5 mg/l	varies	5[3]	5[3]	200/200	8
-vapor comp.	deaeration	25%	limited	5 mg/l	varies	50[4]	90[4]	60/105	10
REVERSE OSMOSIS / ULTRAFILTRA'N	-filter, 5 u -lime soft'n or clarify -precipitation inhibitor -pH control	2%-brac 6%-sea	no	95%[2] 99+%[2]	varies varies	10[4]	40[4]	15/45	4 to 7
ELECTRODIALYSIS	same as RO	15%	no	varies[2]	0% sep'n	10[4]	50[4]	15/60	4 to 7
RO/ED	same as RO	15%	no	97%	varies	-	50[4]	-/60	8
ION EXCH'GE	-filter -act. carbon or org. trap	-	limited	1 mg/l	0% sep'n	-	-	-/-	5
FREEZE CRYSTALL'N									
-waste heat	-deaerate	25%	yes	100mg/l	99.5+%	10[4]	10[4]	15/15	10
-convention	-deaerate	(unverified)	yes	100mg/l	99.5+%	45[4]	60[4]	55/70	10

1. Equivalent fuel at .35 generation efficiency, for electrical, and 1.0, for steam
2. ED effects about a 50% reduction in electrolytes per stage. It does not separate undissociated organics.
3. LBS. WATER/1000 BTU Steam. Generally, the quantity given is that found economically and/or technically practical.
4. KW-HRS/1000 GAL PRODUCT
5. Freezing reduces electrolytes and organics equally, is capable of achieving over 200 to 1 separation ratios, and produces a product of less than 100 mg/l without large wash losses.

requires definition of both the cost and benefit factors. The abilities of the various desalting processes and characteristics that effect their costs are outlined in Table 1.

PROCESS TECHNOLOGY

The desalting processes are being discussed in other parts of this program. Further information is available in the literature. These processes do have certain characteristics, in varying degrees, that make them especially suited to reclamation of impaired waters, including:

- they all possess high energy efficiency in separating a pure water from the dissolved solutes.
- they can concentrate the solutes to high concentrations in a correspondingly small volume of water.
- they have high separation factors (ratio of feed to product concentration).

Since these characteristics, as well as cost factors vary between the processes, all such characteristics must be considered in the process selection for a given application. The differences between processes are addressed by Table 1.

In some cases the processes can be used more efficiently together. For example, when a water source has a relatively low concentration of electrolytes a membrane process will do the bulk of the water removal more efficiently than will a thermal porcess. However, if it becomes necessary to concentrate to saturation or even to produce salt crystals then a thermal process is necessary. Where both are required then each can be used where it is most effective. Where high concentrations are needed and membranes would not normally be used, an alternative to thermal processes is an RO/ED combination. Since RO is limited in upper concentration it can't be used alone. Where a high quality product is needed the power consumption for ED increases due to the high resistivity of the product cell solution. The two can be used very effectively together, the RO to produce a high quality effluent and concentrate the solute so that energy consumption in the ED concentrator is minimized.

IRON AND STEEL MAKING

Wastewater Characteristics

The iron and steel industry is the single largest industrial user of water[1]. The average use has been reported at nearly 40,000 gal/ton product[2]. When recycle is forced upon a mill it has been reported that the individual polution problems can be solved more economically[3]. An analysis of zero discharge for the industry[4] showed that such action would add 4 to 5% to the cost of producing steel, would require nearly $.5 billion to implement, and would add over 1100 MWe of electrical demand for the industry.

Water use in the different processes in a steel mill are summarized in Table 2, with the flow regulations proposed in the EPA Effluent Guideline.

TABLE 2

STEEL MILL WASTEWATERS

PLANT OPERATION	WATER DISCHARGES GAL/TON STEEL		WASTEWATERS & CHARACTERISTICS
	conventional	EPA limit	
Coking	2,500 to 8,000	175	•condensates and direct contact quench waters. •critical contaminants phenol, H_2S, NH_3, CN^-, oil, grease, BOD.
Sintering	100 to 350	50	•indirect cooling and scrubber qaters. •suspended solids and oil are major contaminants.
Blast Furnaces (iron making)	1,500 to 5,000	125	•indirect cooling is recycled. •slag cooling is direct; picks up suspended solids. •scrubber waters contain suspended solids, phenol, NH_3, CN^- of major importance.
Steel Furnaces	500 to 12,000	0 to 50	•indirect cooling requirment is high. •scrubber waters pick up suspended solids and fluorides.
Vacuum Degassing	?	25	•little contamination or waste flow.
Continuous Casting	1,800	125	•has both direct and indirect cooling. •direct cooling waters pick up suspended solids and oil.
Primary Hot Rolling	300 to 2,000	25	•some indirect cooling of equipment. •sprays for descaling and cooling and flume waters pick up scale and oils.
Secondary Hot Rolling	1,300 to 16,000	0 to 150	•direct contact waters are used for cooling and descaling.
Pickling	50 to 250	0 to 80	•rinse waters contain dilute acid and iron salts. •scrubbers are used to keep acid fumes down.
Cold Rolling	?	25 to 1000	•alkaline cleaning is usually part of the cleaning operation. •waters are contaminated with dirt, scale, and oil.

Many steel mill wastewaters can be recycled for consumptive reuse with minimal treatment. Cooling requirements, both direct and indirect, are quite high, giving the mill a natural concentrator for some if its waste streams. Recycle of mill wastewaters will change their quality which may affect steel product quality. There is little information on the quality of wastewaters in a steel mill and the quantities vary widely between mills. The wastewater sources and characteristics can be discussed qualitatively.

1. Indirect cooling water blown down from cooling towers and boiler blowdown contain usually only dissolved salts and treatment chemicals, unless oily wastes are used for makeup.
2. Direct contact cooling waters have traditionally been clarified and discharged. Recycle will produce a concentration of dissolved salts and contaminants leached from the coke and steel that they cool.
3. Descaling waters are contaminated predominantly by suspended solids and oil.
4. Scrubber waters from coking ovens and from steel-making furnaces are contaminated with dissolved organics, ammonia, hydrogen sulfide, cyanide and dissolved metals in significant concentrations.
5. Pickle liquors have 1 to 8% free acid and up to 10% dissolved iron.
6. Rinses from pickling, hot dipping, and finishing operations are a diluted correlary of their respective baths.
7. Rolling operations produce a water-oil emulsion, either used directly for lubrication or a result of cleaning operations on the formed product.

Desalting Applications

1. Coking Wastewaters. Phenol extraction was developed for treatment of coking concensates after the ammonia and hydrogen sulfide gases are stripped. Even after stripping and extraction there is a high concentration of other organics and volatilized metals left in the stream. Concentration will make the stream suitable for solvent extraction to recover the other organics. Acids could be concentrated by either RO or freezing, but their volatility would probably prevent significant concentration by an evaporative process.

2. Scrubber Waters. The scrubber solution from steel furnaces contain reduction products, ammonia and hydrogen sulfide, and a mix of volatile organics from cracking and condensing reactions. Fluorides from iron ore are also absorbed in the scrubber. While the organics are not as concentrated as those from the coking operation they do require treatment. Effluent from a sour water stripper could be concentrated and the organic stream combined with that from coking to recover phenols, oil and grease, and organic acids.

3. Oily Wastes. Besides condensing oils in some of the scrubbing systems, other areas in the steel mill also generate an oily waste. In rolling operations the lubricant is often a water emulsion. Cleaning solutions and spent rolling lubricants are major sources of oily wastes. The common method of breaking these emulsions is with acidification to pH 2 or lower. An alternative is to concentrate it with an ultrafiltration system.

4. Pickle Rinses. The rinses from pickling are dilute solutions of the acids used for pickling. The acid value does not justify recovery but has to be neutralized for reuse. Pickling is usually done on a hot form and the rinse water has to be cooled between uses, building up the salt concentration. Chlorides in the rinse waters in excess cause quality problems in the rolled product. Recycle pickle rinses require side stream bleed and/or treatment to control product quality. Since chloride is the main offender an ED or RO process would seem ideally suited for the application.

5. Pickle Liquors. During a pickling operation the acid is consumed by reaction with the iron oxides, leaving a reduced iron solution. The acid value must be maintained or capacity of the pickling line declines unacceptibly. With HCl this level is .5 to 1.% free acid with 10% ferrous ion. With sulfuric acid there must be about 8% free acid with an equal amount of dissolved iron. About 20 to 22 pounds of dissolved iron are produced per ton of metal pickled with 2 pounds of HCl or 20 pounds of sulfuric acid in the spent stream.

Evaporation of the sulfuric acid waste with crystallization of the ferrous sulfate is a common practice in pickling operations. The volatility of HCl prevents use of the technique in hydrochloric pickling operations. Freeze crystallization at the water-ferrous salt eutectic would be an alternative method of reclaiming spent pickle acids.

6. Direct Contact Cooling. Extensive reuse of water in a steel mill will likely have some quality effects, depending on the original makeup and the extent of reuse. In rolling mills the presence of excessive chlorides in the rinse waters leads to pitting that then forms a hole when the material is rolled into a sheet. High chlorides in the water used to mix the sinter is also known to lead to inferior quality[5]. Recycling may thus lead to the need for a side stream desalting treatment to remove chloride from the cooling waters.

7. Boiler Feed. The high cost of fuel does not affect the steel industry to quite the extent that it does other industries because they have large quantities of medium BTU gas as a by-product. Even so, boiler efficiency is still a way to reduce costs significantly. Boiler feed pretreatment by a membrane process is one way to improve this efficiency, reducing energy losses in boiler blowdown significantly.

PETROLEUM REFINING

Wastewater Characteristics

In 1973 the petroleum refining industry used about 2.5 billion gallons of water per day in their operations[6]. There is little quantitative information on where this water is used or on how it becomes contaminated before discharge. The need for this information was cited at an industry conference on wastewater management in 1978[7].

The water used in a refinery can be divided into three general categories:

-boiler makeup, with the generated stream used for process heat,

cooling of high temperature processes, or for stripping volatile components in distillation columns.
-direct injection of the liquid, for cooling or as the solvent in an extraction process.
-cooling tower makeup.

The 2.5 billion gallons a day of water is about 200 gallons per barrel of oil refined. A topping and cracking refinery with a good water management program has the following design criteria for the water systems[8]:

Evaporative cooling	40 gal/bbl
Cooling tower make-up	45-50 gal/bbl
Cooling tower blowdown	5-10 gal/bbl
Steam	15 gal/bbl

The difference between makeup quantities for the whole industry and the consumption and discharge quantities indicates that most of the industry can still do a lot to optimize their water management and in so doing decrease their consumption of fresh water by some 10 billion gallons per day.

A process schematic for a topping and cracking refinery is shown in Figure 1. The makeup requirements for both water and steam are shown as well as the source of the individual wastes. Not shown are miscellaneous but often significant water users and waste generators such as pump seals, floor drains, storm sewers, and spills.

There is little quantitative information on the composition of individual waste streams. Refiners have not developed this information as they have treated the combined stream. Collection systems in most refineries also make it difficult to sample individual operations. The best information on the individual streams discusses them qualitatively only[9]. These are summarized in Table 3.

Desalting Opportunities

1. Organic Recovery. The crude oil distillation columns and the catalytic cracking operations produce effluents high in phenolic and other organic materials. Based on similarity to other cracking operations a large fraction of the other organics would be organic acids and alcohols. It has been shown[10] that concentration of a similar waste stream from coal conversion operations with subsequent incineration is cheaper than biological oxidation. Concentration also raises the possibility of recovery of the orgaincs by solvent extraction.

The referenced study compared freeze crystallization and reverse osmosis for concentration of coal conversion wastes. Evaporation was not considered because of the similar volatilities of the organics and the water. No present RO membrane has suitable organic rejection properties to make it an effective concentrator of the organic acids and alcohols. The BOD concentrations of the two wastes would indicate that concentration for recovery of destruction would favor the organically contaminated refinery wastes even more than the process condensate from coal conversion.

2. Zero Discharge Refineries. The EPA Effluent Guidelines development

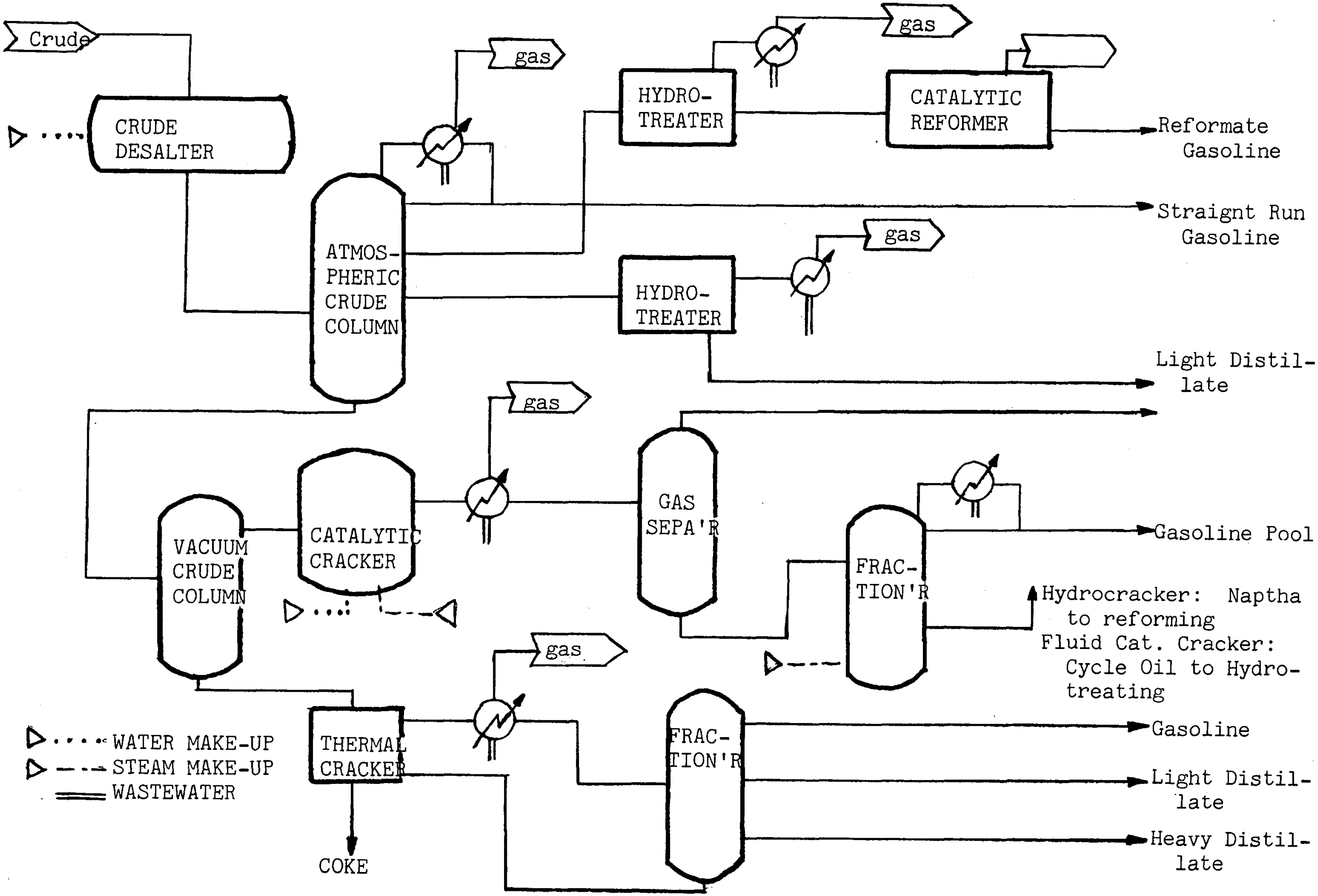


FIGURE 1

PETROLEUM REFINERY SCHEMATIC

TABLE 3

REFINERY PROCESS WASTEWATERS

PROCESS	OIL	NH_3	PHENOL	$S^=$	TSS	TDS
Desalting	x	x	x	x	x	x
Crude Distillation						
-atmospheric	x	x	x	x		x
-vacuum	x	x		x		
Thermal Cracking	x	x	x	x		alk.
Catalytic Cracking	x	x	x	x		
Hydrocracking		x	x	x		
Polymerization						
-pretreatment		x		x		acid
-gas scrubbing		x		x		
Alkylation						
-product wash	x			x	x	x
-accumulators	x			x		
Isomerization						
Reforming						
-pretreatment		x		x		
-accumulator	x	x		x		alk.
Solvent Refining						
Hydrotreating		x	x	x		
Asphalt Production	x		x			

document[11] cites 55 refineries that proctice some form of zero discharge. In fact, 36 of these have evaporation/percolation ponds, 7 have injection wells, and 4 use contract disposal. In many areas those facilities with ponds will have to begin monitoring the groundwater around them for contamination. All of these facilities may become prime candidates for alternative zero discharge technology. In these instances the biggest problem will be management of the dissolved salts that can't be destroyed. Concentration will be the prime need. Recovered water can be used in the cooling towers. See the discussion on waste management in the zero discharge power generation industry for more information on alternatives for this category of waste.

3. Boiler Feed Treatment. Large quantities of steam are generated for process use by this industry. Much of it is generated in low to intermediate pressure boilers that require little pretreatment but have large blowdown fractions. Energy savings and reliability improvements in the boiler system might justify a membrane process for improving feed quality and decreasing blowdown.

ELECTRIC POWER GENERATION

This was the first major industry in the U.S. that had to implement zero discharge technology. Desalting technology has been a major block in the scheme to recycle process waters in the plant. The technology is presented here because it illustrates three aspects of recovery of impaired waters better than any other industry:

-all of the desalting processes have been proposed for use in this application and have been tested or implemented to varying degrees.

- most industries have a large cooling load which is the predominant part of the power industry water management problem. Knowledge here will apply to other industies.
- it illustrates the importance of defining what the "products" are. The major value comes from decreasing the volume and cost of final disposal of the dissolved materials in the plant streams.

Wastewater Characteristics

By far the greatest water user in a power plant is the cooling system and blowdown is the largest source of wastewater in the plant. A general schematic is shown in Figure 2 of the water uses and wastewater sources in a typical power plant. Power plants built in zero discharge areas in the past decade have found it necessary to quantify flows associated with such a schematic[12]. A nuclear power plant is a special version of this diagram where some of the wastes are collected in a separate system and treated for decontamination. A fossil fired plant with scrubbers is another special case. The scrubber system can assimilate large quantities of low quality water into its opeations, evaporating the water and precipitating the dissolved solids for ultimate disposal with the fly and scrubber ashes.

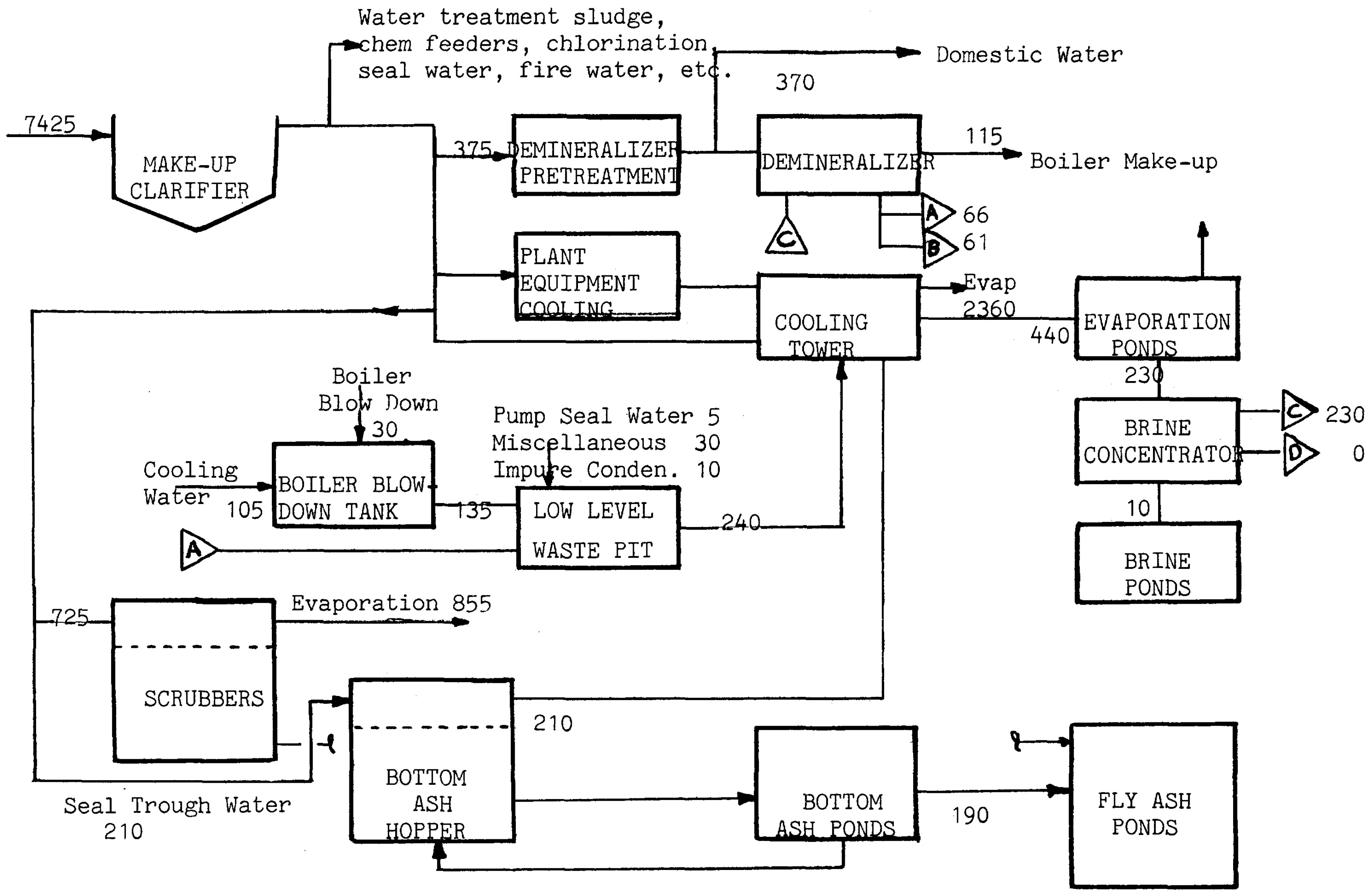
Blowdown from cooling towers alone will amount to over .5 mgd per 1000 megawatts of capacity. Compared to this flow, all other waste streams are relatively minor. The quality of the cooling tower blowdown can be generalized:

- the stream will be nearly saturated with calcium sulfate and often silica.
- the TDS of the stream will vary with the quality of the make-up, ranging from a few thousand to as high as 20,000 mg/l.
- pH is usually just about neutral and the water usually has corrosion and precipitation inhibitor additives and, intermittently, biocides.

Some of the other wastewaters from the plant can produce small changes that affect any subsequent treatment of the combined stream. Floor drains that contain any significant amount of oil interferes with suspended solid removal. Washwaters from economizers, especially in coal fired plants, can contain fine suspended matter as well as large amounts of dissolved material and require pretreatment before most of the desalting processes. The demineralizer regenerant and flush waters about neutralize themselves and are a problem only to the extent that they add to the solids balance in the plant.

Modern power plants use high pressure boilers that require an ultrapure feedwater. This is provided by ion exchangers often preceded by an RO unit. A high quality makeup can decrease the cost of the make-up through purchase of less expensive demineralizers and fewer regeneration chemicals. These costs can be substantial and the basis for more expensive recovery systems that produce a higher quality recycle water. Any reclaimed water that exists beyond the boiler makeup requirement replaces feed to the cooling towers and receives a credit accordingly.

In a zero discharge plant the solids in the water entering the plant,



FLOWS: GPM for a 1000 MWe Plant

FIGURE 2

ELECTRIC GENERATING STATION
SIMPLIFIED WATER MANAGEMENT SCHEMATIC

in the coal that aren't combustible, and any water treatment additives must be stored in an environmentally sound manner. The method used has always been on-site ponds with impervious liners. These ponds have a surface that can be used for evaporation and zero discharge stations have depended heavily on these for disposal of the water used to convey the solids. The capital cost for all ponds beyond those needed for solids storage is at least $17.50 per GPD of water evaporated. This is a very expensive "brine concentrator". Replacing this is the chief economic incentive for use of desalting equipment in the power industry.

Desalting Applications

1. Demineralizer Pretreatment. Ion exchange is an expensive way to remove dissolved solids in waters. Partial desalting by a membrane process allows the demineralizer to work at its optimum condition. There is strong evidence that RO improves long term operations of the demineralizer and extends resin life.

2. Dissolved Solids Concentration. While the bulk of the wastes needing concentration come from the cooling towers, the rest of the plant wastes are often combined with the blowdown for treatment.

The first process used to supplement evaporation ponds was a crystallizing vapor compression evaporator[13]. These units have been installed in numerous zero discharge power plants, produce a crystal slurry, and recover 95% or more of the wastewater feed. Part of the justification for the process is the high quality product that lessens the load on the demineralizers. Demineralizer flow is less than 25% of the cooling tower blowdown, so only part of the flow can be credited with the high quality.

Reverse osmosis is being incorporated into the reuse schemes to preconcentrate the feed to the evaporators. The feed to the RO system is lime softened and then acidified. Brackish water membranes have been used with recovery usually limited to 75 to 80%. In some cases recovery is limited by concentration, the osmotic pressure exceeding the abilities of the membrane package.

Electrodialysis was also investigated for this concentration application. It is not limited by TDS and can compete with the high recovery thermal processes. Another advantage is that the costs for treatment are not affected significantly by the volume to be treated. Costs are mainly a function of the amount of salt that has to be removed. This process would also require lime softening pretreatment. A series of articles[14, 15] described the technology and how it would fit into this application. A major conclusion was that the plant would run most economically with a cooling tower TDS of around 12,000 mg/l. Above that level the cost of condenser leaks would exceed the cost of the concentration.

Freeze crystallization was tested for this application in laboratory equipment. The results were quite promising[16], showing that the unit could work in a eutectic mode, crystallizing both water for recovery and the salts in the feed.

Power plants have a lot of waste heat available. Even though it is of a low quality, it can be used to run a waste heat evaporator-crystallizer[17].

METAL FINISHING

Wastewater Characteristics

Most of the wastewater volume from metal finishing comes from operations that prepare the metal for the final plating step. Refinishing operations have a first step of removing the old plated material. Before actual plating occurs, the surface is cleaned in a series of operations such as:

- -pickling for scale removal.
- -solvent degreasing.
- -detergent cleaning, often with an electrolytic assist.
- -a final acid dip to dissolve any remaining oxidation products.
- -rinses are interspersed between the cleaning operations.

Only after the substrate has been suitably prepared is the plating operation done. Sometimes several metals are plated sequentially if a special finish is needed. Other operations can also follow electroplating to give special properties.

Since organic cleaners don't enter the waste stream to any great degree, there are three basic categories of wastes generated in an electroplating shop. A fourth is generated in shops that refinish used parts.

1. Cleaning solutions. These are generally alkaline sodium salts with surfactants that are used in electrolytic cleaning. Different formulations have been found to be effective on different metals[18]. The objective of such washes is to remove caked soil and dirt. When the solution abilities become depleted the bath is dumped and replaced. The solution usually contains 3 to 10% dissolved salts, usually predominantly hydroxides and metasilicates of sodium.
2. Plating baths. These solutions are very diverse but can be discussed qualitatively. Generally, a plating bath contains materials that perform each of the following functions:
   - -the plating metal is in the solution as a soluble salt. The concentration of different plating materials will vary from 3 to over 30%. The acid-alkaline nature of various plating baths is summarized in Table 4.
   - -electrolytes are added in cases where the metal alone doesn't provide sufficient electrical conductivity. If the solution doesn't have sufficient pH buffering capacity a weak acid or base is added, usually boric acid and ammonia, respectively.
   - -other additives that give special properties to the plated material. Surfactants and sugars are used for deposit uniformity and brightness. Chelating agents are added to prevent crystallization of metal salts or impurities that enter the plating bath.
3. Rinses. To prevent contamination of the various cleaning and plating baths, there is always a rinse between operations. With counter current rinsing there is typically a 1000 to 1 dilution of the plating or cleaning solution in the rinse[19].

TABLE 4

METAL PLATING BATH CHARACTERISTICS

METAL PLATED	BATH			TYPICAL ANION	METAL CONC gm / l.	COMMON ELECTROLYTES	BUFFER[1]	
	ACID	ALK	CN^-				HBO_3	NH_4
Copper, acid	x	x	x	SO_4, BF_4	300-400	-	15-30	-
Copper, alk.				CN, P_2O_7	15-75	NaOH, Na_2CO_3	-	2
Cadmium		x	x	CN	30-60	KCN, KOH	-	-
Chromium	x			SO_4, Cl	250-400	-	-	-
Brass		x	x	CN	60-80	NaCN, NaOH,	-	30-40
Gold, acid	x			CN	5-40	HPO_4=	-	-
Gold, alk		x	x	CN	5-30	KOH, K_2CO_3	-	-
Silver, alk		x	x	CN	30-50	KCN, KNO_3	-	-
Nickel	x			BF_4, SO_4	200-400	NiCl,	30-40	-
Zinc, acid	x			SO_4	300-500	-	-	15-30
Zinc, alk		x	x	CN	60-75	NaOH	-	-
Iron	x			SO_4, BF_4	250-450	-	-	-
Tin	x			SO_4, BF_4	50-100	-	-	-
Lead	x			BF_4	300	HBF_4	135	-

[1] buffer concentration in grams/liter

Desalting Applications

1. Cleaning Baths and Rinses. In operations where the purity of the rinse water is critical, as in semiconductor manufacturing, the demineralized water used for rinsing is quite expensive. A desalting process could recover this high quality rinse for recycle, lowering the load on the ion exchange units. Organics used in photoresists, cleaning solutions, or for stripping photoresists will have to be rejected from the product water or an adsorption process added. The value of cleaning chemicals is low but a side-stream ultrafiltration operation would extend their life and maintain a higher quality.

2. Plating Rinses. Evaporation, RO, and ion exchange have all been used in recovery of metals from various plating line rinses. Ion exchange is limited in its ability to return all components to the plating bath. Before the rise in energy prices, evaporators were shown to economically concentrate rinse waters if counter current rinses were used to minimize the volume[20]. RO has been used quite extensively and shown to economically concentrate rinse streams for return of the drag-out to the plating baths[21]. A review of treatment methods[22] lists those plating operations with rinse water that require other methods in addition to or in place of RO. Freeze crystallization has also been shown to economically concentrate rinse waters from electroplating operations[23]. It has the advantages of higher concentration and equal recovery of all components. Care must be taken to maintain bath chemistry when recycling metals if some of the bath components are not concentrated to the same extent as the metal ions.

3. Pickling Baths. Pickling from metals other than steel can usually be handled the same way as in the iron and steel making industry, described above. The basis for pickling is that the metal oxide is reduced, making it soluble in the acid. The basis for recovery then becomes removal of the metal and anion residual from the acid and reconcentraion of the remaining acid. Alternatives for treating the baths directly include:

- -evaporation and cooling crystallization of the metal salt where it shows a significant temperature-solubility relationship.
- -crystallization alone if the metal salt that precipitates is sufficiently hydrated or the make-up acid is suitable concentrated.
- -with volatile acids, such as hydrochloric, the waste pickle liquor can be processed in a roasting operation where the metal is oxidized and the diluted acid is vaporized and recaptured in a scribber for recycle to the operation.

4. Stripping Wastes and Rinses. When refinishing includes removing the previous deposits, the wastes and rinses are a complex mixture of metals and oxidizing acids. The high acid concentrations make neutralization and precipitation impractical, even for rinses. Membranes are not suitably stable. Evaporators would require exotic material of construction and are therefore uneconomical. Freeze concentration seems to be ideally suited for concentrating the rinses for readdition to the process solution or for electrolytic recovery of the metals.

PULP AND PAPER MAKING

Fiber for making paper is obtained by dissolving the non-fiberous materials from a cellulosic material. The result is a water solution of the chemicals used to dissolve those materials and the organics themselves. Most pulping operations recover the chemicals and dissolved organics and reuse them in some way.

Figure 3 shows the basic operations in an integrated pulp and paper mill with the major wastewater sources. Many of the mechanical operations necessary to make paper are not shown. Pulp mills minimize wastes from the pulping operation by counter-current washing of the pulp and concentration of the pulping liquors for reuse. Wastes are generated from bleaching operations, from the dilute pulp wash waters, from spills, and from the volatile materials that are evaporated with the water in the chemical recovery evaporators.

Wastewater Characteristics

The pulp and paper industry in the U.S. uses just under 6 billion gallons of water a day. Only about 6 or 7% of this is actually consumed in making the paper, most of that in the cooling towers. Wastewaters in pulp and paper mills come from process and cooling operations. A typical integrated mill will consume about 1250 gpm for cooling, producing 100 to 400 gpm of blowdown. The volume and quality will depend directly on the source of the make-up water and on water treatment employed.

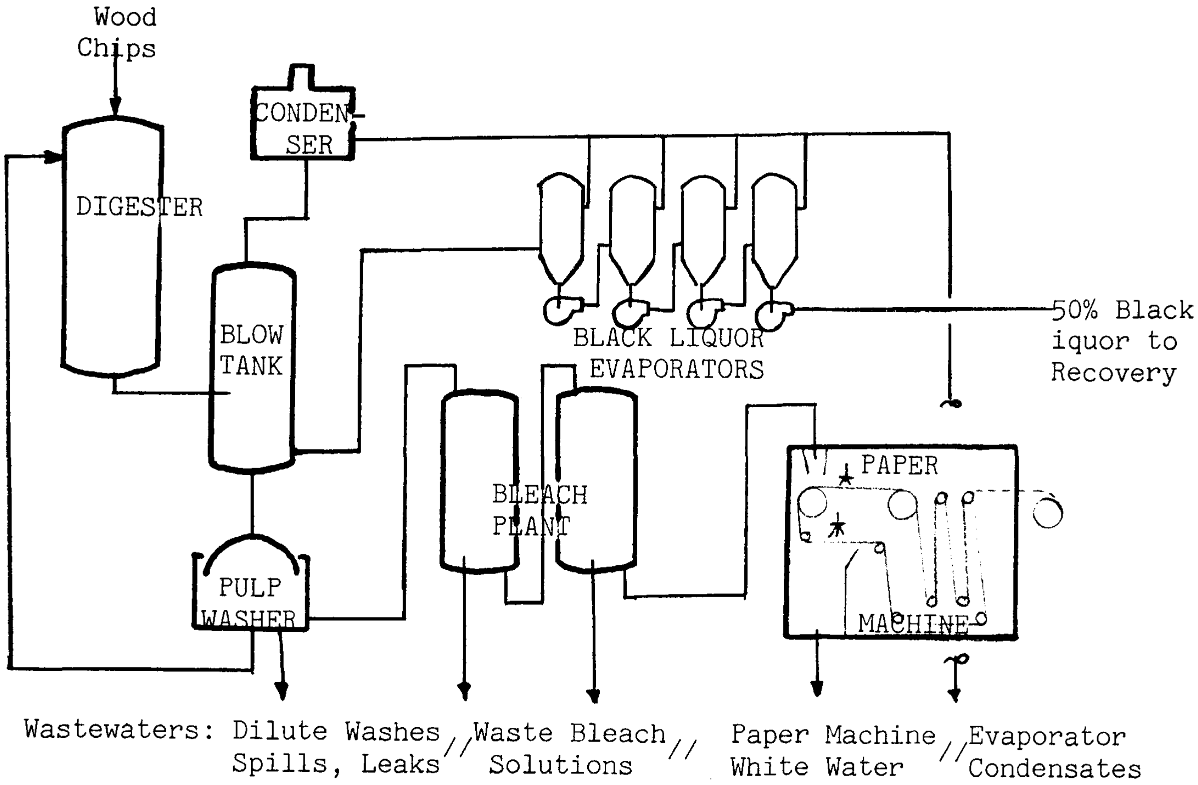


FIGURE 3

INTEGRATED PULP/PAPER MILL
SIMPLE PROCESS SCHEMATIC

Process wastes arise from four major areas, and can be qualitatively described.

1. Weak black liquor, is that flow from final pulp washing operations, or from spills contaminated with the chemical mix (black liquor) after digesting wood chips to make pulp. Typically these wastes contain less than 3% total solids. The composition of the 1 to 3% solids in this waste is about evenly divided between organic and inorganic components. Most of the organics are lignins but 5 to 10% are alcohols, organic acids, organic sulfur and other low molecular weight organics. The inorganic fraction reflects the chemical used in cooking. These dilute streams would thermally and hydraulically over load the black liquor evaporators if added directly to them.

2. Bleach plant effluents contain the by-products of the operations used to remove color from the cellulose fiber. This is usually a multi-stage process, using different bleaching agents in the various stages. Conventional agents are chlorine, chlorine dioxide, caustic, and hypochlorite. Conventional bleach plants produce an effluent containing about .25% solids. Counter-current and continuous bleaching can increase this to about 1%. Different cooking processes, wood feed, and bleach cycles can increase the total solids in the effluent from the bleaching stages. Of the solids in this waste stream, 40 to 50% are salts of the bleaching compound and 10 to 20% are organics, which are not very biodegradable.

3. White water is the water collected in a paper machine that is either drained from the pulp, used to wash the pulp, or used to wash the fiber from the paper forming screen. In mills where this water is not recycled the dissolved solids content is less that 1%. Many mills have begun to recycle this water. The equilibrium solids level has been reported to be 3.5% or greater. Most of these solids are organic.

Volatile organics account for about 5% of the organics. Sodium salts account for 25% or less of the solids.
4. Evaporator condensates contain the volatile components that are vaporized with the water in the black liquor evaporators. In alkaline plants the average concentration of volatile organics in the condensate is over 250 mg/l. In neutral and acid plants, where the organic acids are also volatile, the condensate can have 1000 mg/l. Digester blow tanks produce a much more concentrated waste stream that contributes nearly 10 times the volatile organics[24]. Several operators have found it more economical to recover these volatile organics then to treat them in the wastewater[25, 26].

Several other operations occur in some mills that generate a waste stream of importance. Barking produces a wastewater contaminated with readily soluble wood chemicals. Specialty papers are coated with various materials, predominantly titanium dioxide and clays. The colloidal materials present a separation problem. Mills that reclaim waste paper have to go through a de-inking operation that generates a wastewater with up to .5% contaminants, chiefly organic.

Desalting Applications

This industry, where all of the wastes contain significant quantities of organics, many of them low molecular weight alcohols and acids with significant market value, and all of them with high BOD levels, seems to be an ideal place for desalting processes. Most mills have some spare evaporator capacity so the concentrate from any dilute mill waste can be added to the evaporators. Evaporator concentrate is fed to recovery boilers where the organic content is combusted and provides a significant part of the mills energy.

1. Black Liquor Evaporator Enhancement. Mills recently have been increasing evaporator capacity by adding another effect, which downgrades following effects somewhat but does not add any steam load. A study for DOE showed that freeze concentration for preconcentration of black liquor to 25% solids could significantly decrease energy requirements in a mill[27].

2. Weak Waste Concentration. Concentration of the weak wastewater discussed above would reduce the load on the biological treatment plant as well as recover organics for combustion or by-product value. The Institute of Paper Chemistry has tested RO and evaluated freeze crystallization in the laboratory for these applications[28]. Significant technical development is needed to develop RO membrane and freeze crystallizers to concentrate these wastes economically. Existing RO modules require pretreatment to prevent fouling. Some ultrafiltration equipment would be able to concentrate and remove over 75% of the BOD from the waste streams, but that would not address the problem of salt build up if the wastewater were recycled. Since the greatest by-product value is in the volatile organics a process that would concentrate them for subsequent recovery would have even greater economic incentive. Freeze Crystallization seems to meet those needs.

3. Dissolved Solids Control. Recycle in the bleach processes is limited by build up of salts. Closed loop white water systems have led

to excessive corrosion in some instances. In plants with recovery boilers the salts from various sources could be recycled thru the boiler if they were adequately concentrated.

MINERAL AND NON-FERROUS METAL MINING

This industry is extremely diverse, including not only actual mining but also on-site beneficiation processing that is necessary for transportation and/or recovery of the primary metal. Mine drainage is not discussed here as most places have found it more economical to treat conventionally. While many of the wastewater problems in this industry are solved, there are some that persist. In surveying the industry the EPA has documented the major remaining problems[29].

Wastewater Characteristics

The following beneficiation processes can generate an aqueous effluent:
- wet grinding and screening operations.
- centrifugal classification.
- flotation and aqueous extraction concentrations.

Usually the mine stores these wastes on-site with the tailings or lets them overflow the tailings pond into a receiving stream. In many mills where the mineral is extracted and refined on site there are other wastes that are added to the tailings ponds as well.

There are generally two kinds of wastes generated from mining and beneficiation processes and a third when refining is practiced on-site. Wastes from the initial grading and classifying operation are contaminated with soluble minerals in the ore. They present a problem with ground water contamination since the tailings ponds are seldomly lined. Flotation uses chemicals that enhance separation. In extraction processes a chemical is added that chelates, complexes, or in some other way reacts with the particular mineral of interest. With, for example, gold and silver mining this extractant is cyanide, usually at 200 to 500 mg/l. Excess flotation aids and extractants pass out with the process waters and contaminate the ground water or receiving stream.

Mining operations with current problems include:
1. Copper, where cyanide is used as a flotation aid, and contaminates ground waters around the tailings ponds.
2. Lead and zinc mining, where the flotation agents pass into the plant effluents and form a surface film on tailings and clarification ponds and pass on through to the receiving stream.
3. Gold and silver mining, where the extraction cyanides create a ground water problem.
4. Uranium mining where the tailings still contain a few percent uranium that eventually leaches into the water in the ponds. Recycle of this water builds the chloride content to the point that further reuse is not feasible. Ground water contamination is becoming such a problem here and in the solution mining of uranium that ponds will have to be lined with synthetic liner in the future if they are allowed at all.

Desalting Applications

Pretreatment requirements for many of these wastes present more of an impedement to desalting processes than does the actual cost of the equipment. For instance, in the uranium milling operation an acid extraction is used that results in a high acidity in the tailing pond water. Softening and neutralization is more expensive than the desalting that follows. Solution mines require desalting during restoration to remove all traces of the mining chemicals from the aquifers in which they have operated. More than likely other mining operations will also find that they have to go to similar measures to minimize the wastewater that they have to impound.

Desalting processes can address several needs that the mining industry is likely to have.

1. Water Recycle. Storage in many operations will become excessively expensive and water recycle will be adopted. A bleed of a concentrated dissolved solids stream will minimize storage requirements.

2. Chemical Recovery. Concentration of waste streams will allow more economical recovery and recycle of the flotation and extraction chemicals. A loose RO membrane should concentrate organic flotation agents for either direct recycle or for economical extraction from the water by another organic.

3. Direct Recovery. Many of materials in this category have a strong temperature - solubility effect. Since single effect evaporator crystallizers are used, energy savings would usually justify conversion to a vapor compression cycle to a eutectic freeze crystallization process.

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DESALINATION OF SEAWATER

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ABSTRACT

Desalination of seawater is vital to the development and growth of many countries in the world including Saudi Arabia. Numerous processes have been proposed for this purpose, but only a few of these have attained important commercial status. Multiple-effect distillation (MED), multiple-stage flash distillation (MSF) and reverse osmosis (RO) are being used now for seawater desalination. Vapor compression, freeze separation and solar distillation have been less developed and are limited to very special cases. Saudi Arabia has gained a valuable experience, with large M.S.F. and R.O. plants, which is proving to be useful in designing new ones. This paper discusses the various seawater desalination methods, the Saudi Arabian experience with seawater desalination and the prospects of utilizing solar energy in seawater desalination.

INTRODUCTION

Desalination can be defined as the science of producing fresh potable water from saline supplies reliably and at a reasonable cost. The World Health Organization has specified the maximum total dissolved solids concentration permissible in public water supply not to exceed 500 ppm. Therefore, the salinity of seawater (35,000 ppm or more) and of brackish water (about 3,000 ppm) must be reduced to less than 500 ppm. This paper will deal only with seawater desalination which is expensive and more complicated than that of brackish water.

Numerous processes have been proposed for the desalination of seawater, but only a few of these have attained important commercial status. Distillation and reverse osmosis are being used now for seawater desalination. Vapor compression, freeze separation and solar distillation have been less developed and are limited to very special cases. In this paper, we will briefly review the major seawater desalination methods, then examine how two of them are being used now in Saudi Arabia, a country which has a desalination capacity in excess of 181,800 CMD

(48 MGD). The Saudi Arabian experience with multistage flash (MSF) and reverse osmosis (RO) seawater desalination plants is important to the rest of the world. Finally, we will comment on the energy consumptions and the prospects for solar energy utilization in seawater desalination.

PROPERTIES OF SEAWATER

Seawater Composition

There are probably over 70 chemical elements that have been detected in seawater, and it is probable that it contains traces of every naturally occurring element. The majority of the constituents exist in ionic form-often as pairs or in complexes rather than as single ions. Seawater has been analyzed in many parts of the world and has been found to have a fairly uniform combination of chemicals in most of the seas. The total concentration of dissolved salts, however, varies appreciably from one sea to the other. In the Middle East, for example, the salinity of seawater varies from 13,000 ppm in the Black Sea to over 60,000 ppm in the Arabian Gulf. The salinity of some seas varies with season. The Arabian Gulf salinity varies between 45,000 ppm and 60,000 ppm depending on the tide and wind currents, while the Red Sea salinity does not vary that much. Table 1 shows the ionic composition, the total dissolved solids (TDS), pH and temperature of water in certain locations on the Red Sea and the Arabian Gulf.

All the atmospheric gases are found in solution in seawater. Oxygen and nitrogen are mostly physically dissolved. Carbon dioxide partially reacts with water to form carbonic acid and its dissociation products. A wide variety of dissolved organic compounds exist in seawater. These include aromatic hydrocarbons from hydrocarbon spillages and other substances derived from industrial wastes as well as substances such as carbohydrates derived from marine organisms.

Properties of Water and Aqueous Solutions

The following properties of water are important to desalination. Water is a dipolar molecule. Therefore, it is a good solvent for ions. It forms ions itself; its boiling point varies significantly with small changes in pressure and it has relatively high specific and latent heats. As far as aqueous solutions, the important properties are:

- **Solubility Limits** : Most salts have a limited solubility in water. This limit depends on temperature and sometimes, to a lesser extent, on pH, and the presence of other ions. The majority of salts become increasingly soluble with rising temperature but one or two do the opposite and these often cause problems in desalination. The three most troublesome salts tend to be $CaCO_3$, $CaSO_4$ and $Mg(OH)_2$. All of them have relatively low solubilities and often occur in nearly saturation quantities in seawater feed.

- **Boiling Point Elevation** : The presence of a non-volatile salt in water will raise its boiling point.

TABLE 1. SEAWATERS OF SAUDI ARABIA

Location	Ionic Composition							TDS	pH	Temperature °C	
	Na	Ca	Mg	SO_4	Cl	HCO_3	SIO_2			Min.	Max.
Al-Khobar, Arabian Gulf	19,100	617	2069	4523	32100	127	2.0	58500	8.1	15	35
Jubail, Arabian Gulf	12,998	517	1594	3282	23588	141	2.5	42528	8.2	15	35
Al-Khafji, Arabian Gulf	13,460	500	1665	3100	23100	135	5.0	42000	8.3		
Jeddah, Red Sea	14,000	520	1476	2960	22000	200	0.4	41200	8.0	22	32
Yanbu, Red Sea	13,626	750	1600	3700	23500	120	0.5	43300	8.0	24	30.5

- Vapor Pressure Lowering : Similarly the presence of salt in solution will lower the water vapor pressure at any given temperature.

- Freezing Point Depression : The presence of salt in solution will lower the freezing point of the water in the solution.

- Osmotic Pressure : If a semi-permeable membrane that allows water to pass but not salt, is used to separate pure water from a salt solution, then the osmotic pressure is that pressure to which the saline solution must be subjected in order to halt the flow of the pure water through the membrane into the saline solution. This is the process of osmosis.

- The Membrane Potential : If two saline solutions of different concentrations are separated by an ion selective membrane that will allow either only anions or only cations to pass and is impermeable to water, an electrical potential develops between them.

- Minimum Energy of Separation : The minimum energy input to achieve the desired separation is the free energy difference between the outgoing and incoming water flows where they all enter and leave at the same temperature and pressure. In this case the desalination plant would be thermodynamically reversible.

In order to obtain an idea about the order of magnitude of the above properties, they were estimated for a 35000 ppm solution of sodium chloride which is roughly equivalent to standard seawater. The estimated results are 1 :

1) Boiling point elevation at 1 atm is 0.63°C.
2) Vapor pressure loss at 100°C is 0.0228 bar.
3) Freezing point depression at 1 atm is -2.28°C.
4) Osmotic pressure at 300K is 30 bar.
5) Membrane potential relative to a 0.05% NaCl solution at 300K is 0.11 volts.
6) Minimum energy of separation for a 50% recovery ratio is 77 KJ/Kg-mole or 4.3 KJ/Kg.

SEAWATER DESALINATION METHODS

Several desalination methods exist nowadays and for convenience they can be classified by energy consumption or by considering the band of feedwater salinities for which they are applicable. Another useful way to classify desalination methods is to separate the processes into those which involve phase change to separate the pure water from the saline water and those which do separation without phase change.

In the first category the processes are:

1) Multi-stage Flash Distillation (MSF)
2) Multiple-Effect Distillation (MED)
3) Vapor Compression

4) Solar Distillation
5) Freezing

While those in the second category are :

6) Reverse Osmosis
7) Electrodialysis

The basic principles of each of the above processes are explained in the literature [1,2,3]. But the suitability of the various methods of desalination is illustrated in Figure 1. From this figure, one can see that the phase-change processes are capable of producing potable water from high salinityfeed waters at an economical rate while those classified as non-phase change methods are only economically viable with low salinity feeds. Due to recent technological advances, reverse osmosis is now being applied to the desalination of seawater although it is still too early to measure its success.

This paper is mainly concerned with seawater desalination, therefore, only those methods applicable to such type of feed, and of importance to Saudi Arabia, will be reviewed here. The methods are multistage flash distillation, multiple effect distillation, vapor compression, freezing, and reverse osmosis.

Multi-Stage Flash Distillation [1,2]

The multi-stage flash distillation (MSF) method is illustrated in figure 2. In this process seawater feed is taken into the plant, after screening, and is fed through the heat rejection section whose function is to reject thermal energy from the plant to allow the product water and the brine blow down to leave the plant at the lowest possible temperatures. A portion of the warm feedwater (known as the makeup water) passes through the deaerator and the decarbonator then is combined with the recirculating flow to form the feed to the heat recovery section. This water passes through a series of heat exchangers, its temperature being raised as it does so. After passing through the last of these exchangers, the warm dilute brine enters the brine heater. Here its temperature is raised until it is approximately equal to the saturation temperature for the system pressure. The hot dilute brine then enters the first stage or flashing chambre through an orifice and in so doing has its pressure reduced. Since the water was already at saturation temperature for a higher system pressure, it will become superheated and give off water vapor. This is known as flashing. The formed vapor in stage 1 passes through a wire mesh (demister) to remove any entrained brine droplets and onto the heat exchanger tubes where the vapor is condensed and drips into a collector or distillate tray. The latent heat of condensation is thus transferred to the flowing inlet seawater inside the tubes. This is the reason this section of the plant is called the heat recovery section. The process is then repeated all the way down the plant as both brine and distillate enter the next stage which is at a lower pressure.

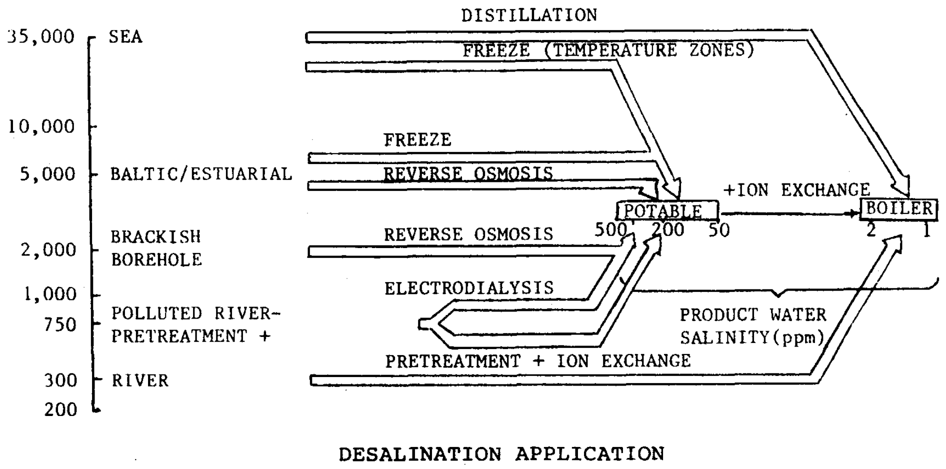

FIGURE 1. FEED-WATER SALINITY AND PROCESS APPLICABILITY

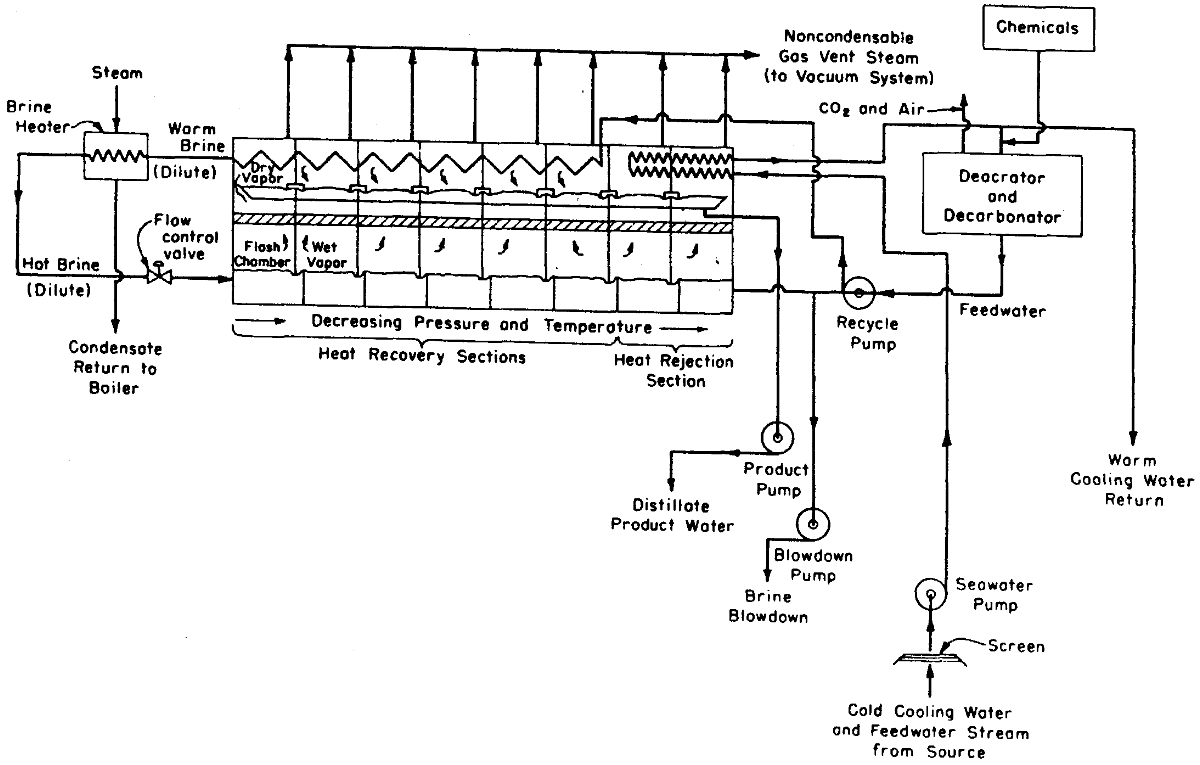

FIGURE 2. MULTI-STAGE FLASH DISTILLATION PLANT SCHEMATIC

The multi-stage flash process has an important advantage over other distillation processes because the number of stages used is not strictly tied to the performance ratio (defined as the number of pounds of distillate produced per 1000 Btu of heat input) required from the plant. The minimum possible should be, roughly, greater than the performance ratio, while the upper limit is imposed by the boiling point elevation, defined earlier as the elevation of boiling point due to the presence of non-volatile solute above that of the pure solvent at the same pressure. The minimum interstage temperature drop must exceed this for flashing to occur. Within these limits, we are free to vary the number of stages. This is advantageous because as we increase the number of stages, at a cost of extra partitioning, we increase the temperature difference over the heat exchangers and hence require less heat transfer material with obvious saving in plant capital cost. The tubes of the heat exchangers or evaporators could be arranged either parallel to the direction of brine flow (long-tube design) or perpendicular to it (cross-tube design). In the latter case, the tubes could be either horizontal or vertical. An example of a seawater MSF desalination plant is Jeddah II which will be discussed in the next section.

Multiple Effect Distillation [1,2]

In multiple effect distillation vapor is produced by either pressure reduction (flashing) or by heat input (boiling). Unlike MSF process, the MED process operates on a once through system having no large amount of brine recirculation. Therefore pump sizes are smaller.

Similar to the MSF process, the inlet seawater passes through a series of heaters (condensers for the vapor) but after passing through the last of these, instead of entering a brine heater, the seawater enters the first effect where heating steam raises its temperature to the saturation temperature for the system pressure, thereafter the thermal input to the effect produces evaporation. Part of the formed vapor goes to heat the inlet feed and the other part provides the heat supply for the second effect, which receives brine from the first effect. The process is then repeated all the way down the plant. The distillate also passes down the plant, flashing as it does due to the progressive reduction down the plant of the system pressure.

There are several plant arrangements possible with multiple effect distillation. Modern plants may have vertical tubes, Figure 3, or horizontal tubes with steam condensing on the outside of the tube and boiling on the inside, or, conversely with steam condensing on the inside and boiling on the outside of the tube. These arrangements eliminate hydrostatic head suppression of boiling.

The performance ratio of an MED plant is rigidly tied to the number of effects for that plant. This condition is the reverse of that for an MSF plant. For a performance ratio of ten, an MED plant will have about 13 effects while an MSF plant could have from about 11 to approximately 35. The choice for the latter plant depends on factors such as relative cost of material. A comparison of the MED process against the

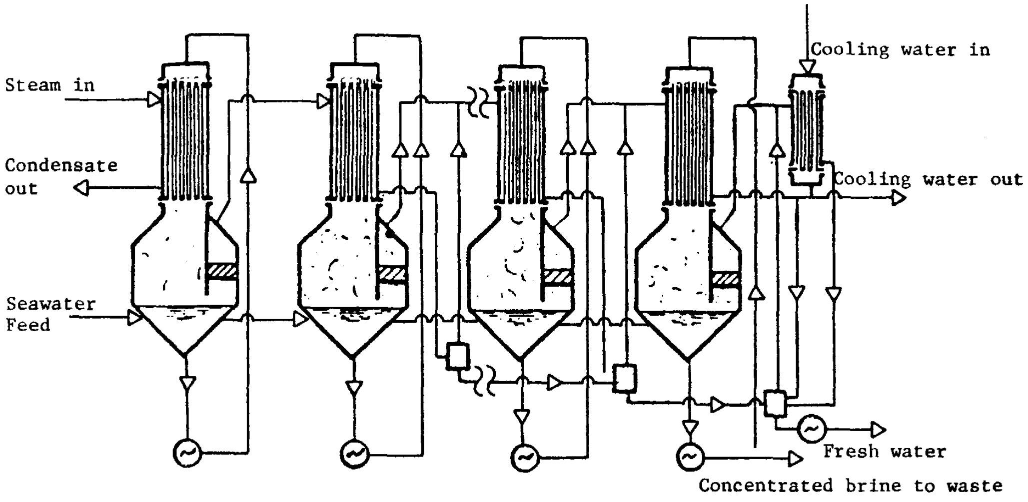

FIGURE 3. MULTIPLE EFFECT VERTICAL TUBE DISTILLATION

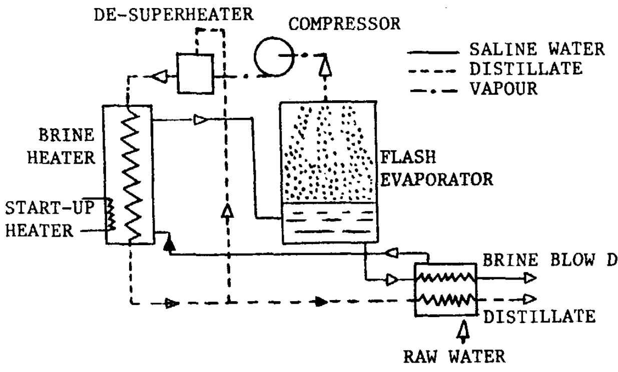

FIGURE 4. VAPOR COMPRESSION DESALINATION

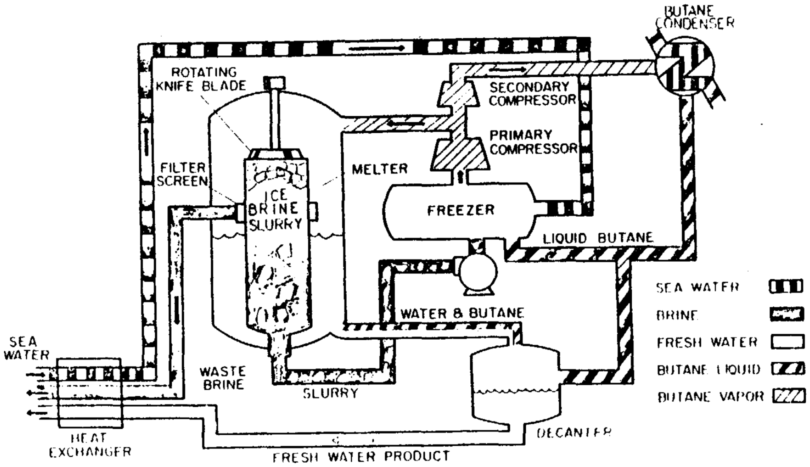

FIGURE 5. SECONDARY REFRIGERANT FREEZING PROCESS

MSF and vapor compression processes, that includes operational conditions and costs, are available in the literature [4].

Vapor Compression Desalination [1,2]

In distillation processes the energy input to the system is accomplished by heating, whereas in vapor compression the input is accomplished using a work device, Figure 4. The seawater feed entering the plant is initially heated in liquid/liquid heat exchangers by the brine blow down and the distillate streams. The seawater then enters the brine heater where it is further heated by the thermal rejection from the compressor. The feed then enters the flash evaporator where vapor is formed. The method shown in Figure 4 is based on the horizontal falling film process. Similar to the MED process, vapor produced in this effect is then passed to the vapor compressor where it is compressed; its temperature being raised in the process and then passed back to the first effect. The compressor represents the only energy input to the system and as the latent heat is effectively recycled round the plant, the compressor only has to supply energy for the flash drop plus losses and consequently very high performance ratios may be obtained. For valid comparison with energy consumption in MED process, for example, the electrical energy consumed in the compressor has to be converted into primary fuel energy. This can be done by approximately multiplying the power consumption by a factor of 3.

The economy of this process is favorable and that is due to the regenerative use of heat noted above and to the simplicity of design [4,5]. However, of the total installed capacity of land based seawater distillation plants for domestic use, vapor compression accounts for less than 1%. Presently the VC plants are of small capacity type, often about 200 metric tonnes per day and rarely over 600 tpd. This limitation is due to the capacity of the compressors available now. Also, operational problems with large vapor compressors make people wary about large VC plants.

Freeze Desalination [1,2]

Freezing is attractive from energy point of view because the energy required to freeze a unit mass of water is about one sixth of the energy required to evaporate it. The freezing of a salt solution causes crystals of pure water to nucleate and grow leaving a brine concentrate behind. Figure 5 shows a one commonly proposed freezing method, where a secondary refrigerant, namely, butane is evaporated in direct contact with seawater to remove the latent heat of crystallization. Inlet seawater is precooled by heat exchange with the product water and blowdown brine streams. The seawater then enters the freezer where liquid butane is bubbled through it. Butane vaporises and lowers the water temperature causing the formation of salt-free ice crystals in a more concentrated seawater or brine. About one-half of the seawater is frozen into ice crystals, the ice-brine slurry is then pumped to a washer-melter. The slurry rises in the water and the ice crystals are compacted into a porous bed of ice. The bed of ice is pushed upward by a slight positive pressure caused by the brine flowing through the bed and outward through screens positioned near the middle of the column. The rising ice bed is washed then removed by means of a mechanical scraper into the outer annulus which is called the melter. The butane vapor which contains the heat removed to form the ice, is compressed in the primary compressor and then introduced into the melter where it condenses on the ice. Heat is given up and the ice is melted. The condensed butane and the product water flow together to a decanter where the two liquids are separated.

The product water leaves the decanter to the fresh water storage tanks while the liquid butane is recycled back to the freezer. Butane vapor not required for ice melting is further compressed by the secondary compressor and then condensed in the butane condenser. The condenser is cooled by seawater. The liquid butane is re-cycled to the freezer.

There are other freezing processes that we will not discuss here but are discussed in the literature [2,5]. A major advantage of the freezing process is obviously the low temperature at which desalination takes place. The two main sources of difficulty in the distillation process, chemical scaling and corrosion, will be far less apparent in the freezing process. Because of the minimal corrosion effects in freezing plants, cheaper construction materials such as non-corrosive plastics, protective coated low-alloy steels and concretes may be employed.

However, there are several problems with freezing plants such as heat transfer problems in the heat exchangers, high energy input to the compressor etc., and they have to be solved before the process emerges on the world market. The market is now dominated by MSF method and freezing would have to be proven superior to it before being accepted. This would require large sums of money to build a full scale trial plant.

Reverse Osmosis Desalination [1,2]

Reverse osmosis takes its name from the fact that in this process water is made to pass from the more concentrated solution to the less concentrated one, which is the reverse of the osmosis phenomenon. The force necessary to accomplish this is the application of pressure greater than the osmotic pressure of the saline solution. For the case of seawater, if the required pressure is applied then pure water will be obtained on one side of the membrane and concentrated brine on the other side, Figure 6. The energy input to the system will be a function of the pressure required and hence of the salinity of the feed.

The important factors in reverse osmosis are rejection ratio, flux and membrane life, with usually high salt rejection being achieved at the expense of low flux and vice versa. There are various designs of membranes such as spiral wound, hollow fibers, etc. But the membranes are very sensitive to biological and non-biological fouling. Because of this reverse osmosis plants normally require elaborate pretreatment

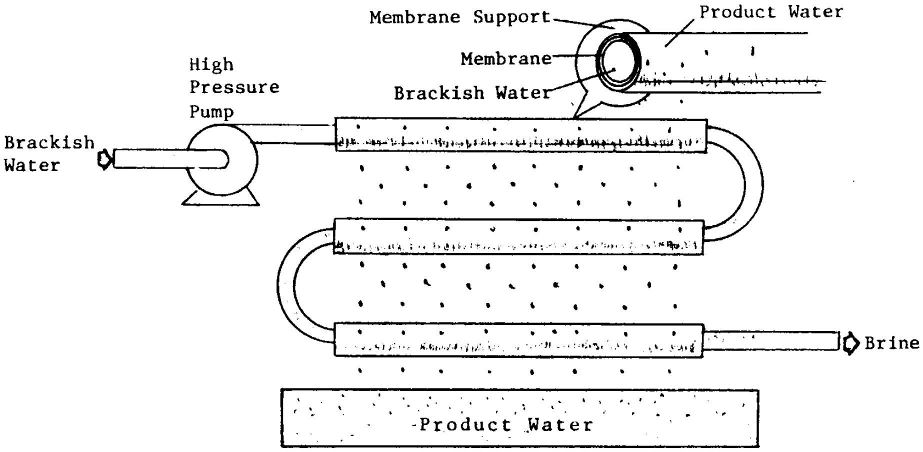

FIGURE 6. REVERSE OSMOSIS METHOD

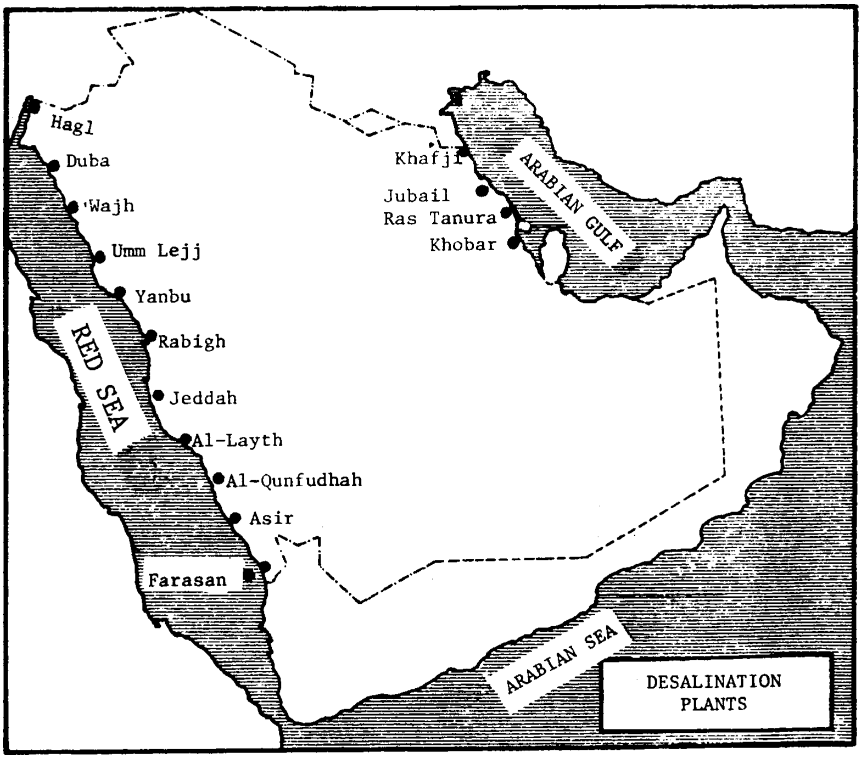

FIGURE 7. LOCATIONS OF SEAWATER DESALINATION PLANTS IN SAUDI ARABIA

units. An important example of seawater reverse osmosis method is Jeddah 12,000 m^3/day RO plant. This plant will be discussed in details in the next section.

SEAWATER DESALINATION IN SAUDI ARABIA

Desalination of seawater is one of the most important engineering tasks facing Saudi Arabia today. Seawater desalination is under the jurisdiction of Saline Water Conversion Corporation (SWCC). Since its creation in 1965, SWCC has established several desalination plants on the Red Sea and the Arabian Gulf. Figure 7 is a map of Saudi Arabia showing the locations of existing and planned plants. Tables 2 and 3 contain information about the existing seawater desalination plants, those that were built by SWCC or by other organizations. The anticipated country's need for desalinated seawater and power in the near future is shown in Table 4. This table, which lists plants under construction and planned plants, is taken from the current Saudi Arabian five year plan [6]. According to Table 4, the total seawater desalination capacity in 1980 is 181,800 CMD while the total generated power for the dual purpose plants is about 350 MW. In 1987 the total water production is estimated to be 1,900,900 CMD while the total power generation is estimated to be 4300 MW. Most of the new plants will be constructed on the Red Sea coast. Desalinated seawater will be pumped to cities like Medina and Riyadh about 190 and 465 kilometers inland respectively. Also cities like Taif and Abha which are on the mountains, about 1500 and 2300 meters respectively above sea level, will obtain desalinated seawater.

Almost all of the desalination plants that are being operated by SWCC now are of the MSF type. Some of them are cross-tube design, but the majority are long-tube design. The antiscaling additive used in the majority of the plants is sulfuric acid which is manufactured locally by Safco in Dammam. The energy source for plants on the eastern coast is natural gas while plants on the Red Sea use liquid fuels. Most of the new plants will continue to be MSF. However, SWCC is willing to establish reverse osmosis seawater desalination plants for small towns like Al-Birk on the Red Sea.

Typical Saudi Arabian Seawater Desalination Plants

Jeddah II MSF Plant

This plant is representative of seawater MSF desalination plants in Saudi Arabia. It is a dual purpose plant and was started in 1978. The plant comprises two oil fired boilers, two passout condensing turbine generator units and four desalination units complete with all auxiliary plant and equipment. Experience with Jeddah I helped make this plant more reliable. Some specific information about the plant is listed below [7].

TABLE 2. MSF DESALINATION PLANTS IN SAUDI ARABIA

Location	Water Production, CMD	Electricity MW	Contractor	Contract Cost $ 10^6	Start-up Date	Contracting Agency
Wajh I	227		Aquachem	0.57	1969	SWCC
Wajh II	454		Weiritam, Sowit	21.24	1979	"
Duba I	227		Aquachem	0.57	1969	"
Duba II	454		Weiritam, Sowit		1979	"
Jeddah I	18900	50	AEG, Contin. Eng. IHI, Aquachem	52.28	1970	"
Jeddah II	37850	80	Brown Boveri, Citoh, Sasakura	210.00	1978	"
Jeddah III	75700	200	Kraftwerk Union Deutsche Babcock Weir Westgarth	525.00	1979	"
Khobar I	28400		Aquachem	47.36	1973	"
Khafji I	454		Sasakura	3.03	1974	"
Khafji (quick flow)	1249		Krupp	5.17	1978	"
Umm Lejj I	454		Krupp	12.00	1975	"
Farasan	500	0.77	Krupp	12.30	1979	"
Hagl I	454		Hitachi-Zosen	13.00	1979	"
Khafji	1890		Sasakura		1966	Arabian Oil Co.
Khafji	2270		"		1971	" "
Khafji	2270		"		1975	" "

Location	Water Production, CMD	Electricity MW	Contractor	Contract Cost $\$ 10^6$	Start-up Date	Contracting Agency
Dammam	1325		Sasakura		1973	SAFCO
Ras Tanura	1100		"		1973	ARAMCO
Ras Tanura	1630		"		1973	"
Ras Tanura	1630		"		1975	"
Ras Tanura	3255		"		1978	"
Ras Tanura	1630		"		1978	"
Jeddah Refinery	6132		Weir Westgarth		1976	Petromin
Jeddah Airport	20800		Sumitomo, Weir Westgarth		1980	International Airport
Yanbu (Barge Plant)	2044		Sasakura		1978	Royal Commission for Jubail and Yanbu
Yanbu Refinery	3785		Weir Westgarth		1979	Petromin
Jubail (Barge plant)	18900		Hitachi		1980	Royal Commission for Jubail and Yanbu

TABLE 3. RO SEAWATER DESALINATION PLANTS IN SAUDI ARABIA

Location	Start up Date	Capacity CMD	Membrane Type	Feedwater	Product Water	Contracting Agency	Contractor	Cost $\$10^6$
Jeddah	1979	12000	SW	Seawater	Potable	SWCC	U.O.P.	30
Yanbu	1980	3000	HF	"	"	Royal Commission for Jubail and Yanbu	Polymetrics	07

1. General Information

Dual purpose plant : Power + Water
Plant capacity power: 80 MW Water : 37,850 m^3/day (10 MGD)
Number of MSF units in plant: 4
Number of modules per unit : 5
Number of stages per unit : 31 (heat recovery) +3 (heat rejection) = 34
Tubes arrangement : Long tube type
Antiscaling additive : H_2SO_4 98% N
Performance ratio : 0.004 Kg/KJ (9.3 lb/1000 Btu)
Energy source : Heavy fuel oil

2. Brine Heater

Brine heater is a cylindrical shell and straight tube type heat exchanger

Water flow rate : 3450 m^3/hr
Temperature of water : $T_{in} = 113^oC$ $T_{out} = 117^oC$
Steam flow rate : 54 m^3/hr
Tube diameter : 19 mm OD Thickness : 0.9 mm
Tube length : 6,670 mm
Number of passes : 1
Number of tubes per pass : 2059
TDS of circulating water : 45000 ppm
TDS of blow down stream : 67000 ppm
Concentration ratio : 1.3

3. Major Units

Make up water vacuum deaerator
Make up water decarbonator
Venting equipment : Two two-stage steam ejector.

Full duty stand-by pumps are provided for the plant.

TABLE 4. ESTIMATED PRODUCTION (WATER AND POWER) OF DESALINATION PLANTS AT FULL CAPACITY

Water Production Thousand CMD	1980	1981	1982	1983	1984	1985	1986	1987
Existing Plants	179.2	179.2	179.2	179.2	179.2	179.2	179.2	179.2
Plants Under Construction								
Yanbu	-	-	95.0	95.0	95.0	95.0	95.0	95.0
Rabigh I	-	-	1.0	1.0	1.0	1.0	1.0	1.0
Jeddah IV	-	-	190.0	190.0	190.0	190.0	190.0	190.0
Jubail I	-	-	114.0	114.0	114.0	114.0	114.0	114.0
Al-Khobar II	-	-	190.0	190.0	190.0	190.0	190.0	190.0
Jubail II	-	-	-	-	665.0	665.0	665.0	665.0
New Plants								
Birk I	1.9	1.9	1.9	1.9	1.9	1.9	1.9	1.9
Mobile Units	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Al-Layth	-	-	0.4	0.4	0.4	0.4	0.4	0.4
Mastoura	-	-	1.9	1.9	1.9	1.9	1.9	1.9
Thool I	-	-	1.9	1.9	1.9	1.9	1.9	1.9
Duba III	-	-	-	3.8	3.8	3.8	3.8	3.8
Hagl I	-	-	-	5.7	5.7	5.7	5.7	5.7
Al-Khafji II	-	-	-	23.0	23.0	23.0	23.0	23.0
Umm Lajj II	-	-	-	3.8	3.8	3.8	3.8	3.8
Al-Qunfudhah	-	-	-	-	3.8	3.8	3.8	3.8
Yanbu II	-	-	-	-	-	76.0	76.0	76.0
Mecca/Taif	-	-	-	-	-	152.0	152.0	152.0
Al-Wajh	-	-	-	-	-	-	3.8	3.8
Asir I	-	-	-	-	-	-	-	94.0
Jeddah V	-	-	-	-	-	-	-	94.0
Total Water Production	181.8	181.8	776.0	812.2	1481.1	1709.9	1712.9	1900.9
Power Generation Capacity (MW)								
Existing Plants								
West Coast	350	350	350	350	350	350	350	350
Plants Under Construction								
West Coast	-	-	750	750	750	750	750	750
East Coast	-	-	750	750	2045	2045	2045	2045

Power Generation Capacity MW	1980	1981	1982	1983	1984	1985	1986	1987
New Plants								
West Coast	-	-	-	35	45	645	655	1155
East Coast	-	-	-	-	-	-	-	-
Total Power Generation	350	350	1850	1885	3190	3790	3800	4300

Jeddah Reverse Osmosis Plant

This world's largest seawater reverse osmosis plant is under the jurisdiction of SWCC; and it is located on the same site as Jeddah I, II, and III, MSF desalination plants. It can be dismantled, however, and moved to other locations, if needed. The plant capacity is 12,000 m^3/day and was put in operation in January 1979. Its process flow diagram is shown in Figure 8. Because it is the first major RO seawater desalination plant, it will be described here in more details [7].

Seawater of temperature 28-33°C is supplied through a pipe type intake. It is lifted then by means of two low speed vertical pumps (a third one is on stand by basis) to a splitter box. It is treated with copper sulfate (1-1.5 ppm) to destroy living microorganisms. It is then diverted to three 32-foot diameter, gravity flow, dual media sand filters. Each filter is divided into free compartments to facilitate backwashing while maintaining full operation of the plant. All sections are backwashed every 24 hours (eight in operation, one in backwash). Backwash is on automatic cycle. Before entering the RO units, the feedwater is treated further with sulfuric acid and sodium hexametaphosphate, followed by micron cartridge filteration (2.5μ). The addition of the acid brings pH down to 6.4. The sodium hexametaphosphate is added in the amount of 2 ppm to inhibit scale formation.

The reverse osmosis plant is made up of first and second stage units (actually they should be called first and second pass units since it is the product water that is processed twice). The plant includes nine first stage units, eight operating and one standby, and three second stage units. The nine first stage units are fed by individual pipelines from the distribution header. Each stream is treated with acid and SHMP (as mentioned before) prior to desalting. The pressure of each of the nine feedstreams is boosted by two multi-stage vertical turbine pumps, operating in parallel. Each pump is driven directly by 425 HP diesel engine, it has a capacity of 500 gpm at a discharge pressure of 1000 psi.

The first stage units run at 30% recovery. Thus each of the nine first stage units is producing 1630 m^3/day. This results in a combined output

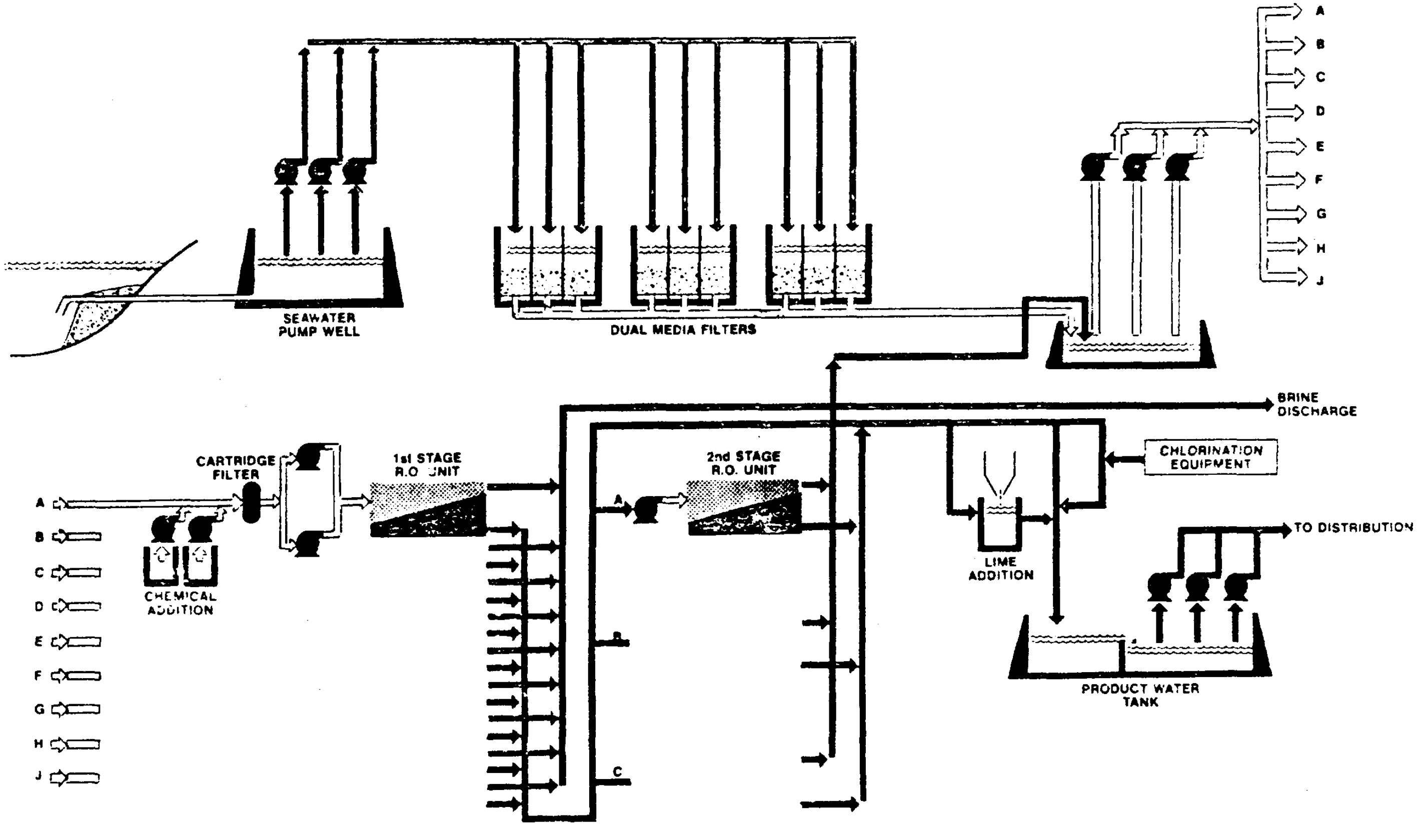

FIGURE 8. JEDDAH SEAWATER RO DESALINATION SYSTEM PROCESS FLOW DIAGRAM

from the eight operating first stage units of 13,000 m^3/day. The permeate from the first stage units flows to a common collection header. The permeate can either be routed directly to the post-treatment system or split for partial flow to the second stage units. The salinity of product water from the first stage may exceed the specified limit of 1000 ppm, while the salinity of water from the second stage should be no more than 100 ppm. In such a case these waters are blended to assure that the total salinity of the product water does not exceed 1000 ppm. The second stage units operate at 85% recovery. The concentrate (brine), from the second stage is recycled through the first stage units. One feed pump, operating at a pressure of 500 psi, is used for each of the second stage units. These pumps are identical to the feed pumps of the first stage units, and are also driven by identical diesel engines. The nine first stage units consists of 56 (7×8) pressure vessels (fiberglass), 150 mm diameter and 6.5 m long. These pressure tubes are fed in parallel. Each vessel contains six TFC Model 1501 PA spiral wound membranes, resulting in a total of 336 membrane elements for each first stage unit. Each of the three second stage units consists also of 56 pressure vessels but, these units are piped in a configuration such that water flows through a three-pass system composed of 32 vessels in the first pass, 16 in the second pass, and 8 in the third pass.

Experience with Seawater Desalination

The Saline Water Conversion Corporation experience with seawater desalination plants will be briefly reviewed here [7].

Multistage Flash Desalination Plants

Wajh I and Duba I Plants. These plants operate at low temperature. That is, the maximum temperature of the seawater in the brine heater is below 95°C. The antiscaling additive is a polyphosphate. There are no major problems with operating these plants. The desalination units are in good condition after more than 10 years of operation.

Jeddah I and Al-Khobar I Plants. The design and materials specifications of these plants were not based on accurate analysis of the very saline waters of the Red Sea and the Arabian Gulf. Therefore corrosion and scaling have been major problems for these plants; and the experience gained was of worldwide importance. In Jeddah I plant, the brine heater tubes were replaced twice in the past ten years. Last year, in addition to replacing the brine heater tubes, the tubes in the first and sixth stages were also replaced. In Al-Khobar I plant, which went on operation at the end of 1973, the brine heater tubes were totally replaced in 1978. Other material replacement and maintenance work was done in both plants. The problems with Jeddah I plant can be summarized as follows:

a. Inadequacy of design. The material of construction was not selected properly. Carbon steel was used for flashing chambers, water boxes, acid pipes, and other places when stainless steel should have been selected. The decarbonator was not adequate. Sampling points locations were not suitable. Pumps selection was not good.

b. Acid treatment and related corrosion and scaling problems. Because the locations of the sampling points or instruments were not suitable, the pH meters and the flow meters readings were not accurate. Therefore, the amount of sulfuric acid injected was off, which meant that either excess amounts were added causing corrosion in brine heater and flashing chambers or little added causing thermal decomposition in flashing chamber and generation of CO_2 which attacks metals. The plant suffers from the high concentration of free CO_2 (between 12 and 16 ppm) because the decarbonator's performance is not adequate.

c. Scaling in brine heater. Calcium sulfate scale formation inside the brine heater tubes is an extremely difficult problem with Jeddah I plant. This is due to the inadequacy of the design which includes bad instrumentation. Occasionally, the temperature of the low pressure steam in the brine heater goes to 310°F, far above the design value of 250°F which causes $CaSO_4$ scale to form inside the tubes. During a six month period in 1978 the percentage of damaged tubes due to corrosion and scaling ranged from 10% in the heat recovery section to 18% in the brine heater. This percentage is very high and it illustrates the magnitude of the problem.

Umm Lejj I Plant. This plant was started in 1975. Its condition has been generally good. The plant was successfully converted from polyphosphate treatment to Belgard EV which caused an increase of 30% in water production while the maximum temperature in the brine heater was 98°C.

Jeddah II Plant. This plant has been performing according to design specifications with no major problems. The acid treatment seems to be adequate. No scale has been formed in the brine heater tubes. This is probably due to the good control on steam temperature and due to the use of circulating balls (Taprogge) that prevents the formation of scale. Other desalination plants in Jeddah and in other cities on the Red Sea are new and so far no major problems have appeared.

Reverse Osmosis Plants

Jeddah RO Plant. Experience with Jeddah Plant, being the largest RO seawater desalination plant, is important to the rest of the world. It can be considered as a big pilot plant. Experience gained from it will certainly improve future plants. The major problems experienced were due to the design and not rigid enough specifications. Corrosion caused the flexible brine hoses which are stainless steel 316 reinforced neoprene to burst. This resulted in destruction of about 50 membranes. All flexible brine hoses have been replaced by rigid 317 stainless steel pipes. The plant generators were replaced by generators with proper windings that can resist corrosion in a humid and tropical environment.

As for membranes, the plant has a total of 4032, 6 in T.F.C. model 1501 PA membranes. When the plant first went on stream, the operating pressure was 58 Kg/cm^2 (825 psi) in one year of operation it had been increased to 70 Kg/cm^2 (995 psi). This increase was due to membrane compactness. A major problem that affected the performance of the plant was the collapse of product water carrying element in the membrane. This was found to be due to improper heat treatment during membrane manufacturing. Corrective actions were taken and newly manufactured membranes seem to perform satisfactorily.

Yanbu RO Plant. The first unit of this plant went on operation April 15, 1980, second on May 15, and third on May 26. The additional two units will be in operation by February 1, 1981. The total design output of the five-unit plant will be 5000 m^3/day, or 1.32 mgd. The feedwater for the three units is taken from five onshore wells. Three more wells will be needed for the additional units. The wells are located about one kilometer from the sea. They are 30 m deep. Drawing water from wells, instead of taking it directly from the sea, provides good filteration, but for some reason its salinity (58,000 ppm) is considerably higher than that from the sea (42,000 ppm). It is still too early to assess the performance of this plant.

ENERGY SOURCES AND OPPORTUNITIES FOR SOLAR SEAWATER DESALINATION

Energy Requirements and Sources

The energy required for the operation of the various seawater desalination methods described in this paper need to be provided in one or more of the following three forms: Mechanical, Electrical, or Thermal Energy. For fair comparison, all of these forms must be related to a common prime energy source. For example, the energy required by the MSF process is comparatively low grade thermal energy and can be obtained as waste heat from thermal power production at a relatively low cost in terms of prime energy consumption. But the energy used by the RO process is in the form of mechanical work and is thus more expensive in terms of prime energy consumption. Since these two methods are the most important seawater desalination methods, we will examine their energy requirements. The typical current energy requirements for each process are of the order of 300 KJ/Kg for MSF and 34 KJ/Kg for RO. With reasonable expectations that with improved technology these might be improved to 176 KJ/Kg and 25 KJ/Kg respectively. Converting these consumptions to prime thermal energy, taking in consideration that the energy coming to the MSF process will be in the form of extraction steam from a turbine and for the RO process the thermal energy conversion efficiency is 38%, we obtain the following:

	Typical Current	Projected Minimum
MSF	120 KJ/Kg	70 KJ/Kg
RO (conventional no energy recovery)	89 KJ/Kg	60 KJ/Kg

Many other considerations come into the choice and design of a desalination plant, but simply in terms of energy utilization there is not much to choose between the prospects of the two processes.

Desalination plants on the east coast of Saudi Arabia are designed to use natural gas as an energy source for the intermediate future. While those on the west coast are designed to use heavy fuel oil. As far as energy sources and costs are concerned, Saudi Arabia is assured of ample supply. However, the country is interested in developing alternative sources. One of them is solar energy.

Solar Desalination

The simplest form of solar desalination is the solar still. Its output is proportional to the intensity of the solar radiation and the area of the still. There are, therefore, virtually no economies of scale to be gained from building large installations, and conversely no particular penalties in building small ones. A typical figure for solar radiation in arid regions is 5000 kcal/m^2 day which is equivalent to 8.9 litres/m^2 day. A well designed solar still is usually about 50% efficient and would produce 4.5 l/m^2 day or about 1 gal/10 f^2 day. The 50% figure may seem to be low, but it is about the optimum economically.

The tendency now is not to use solar stills for desalination but rather to collect solar energy and convert it into a thermal or electrical form for supplying the normal desalination processes discussed earlier. An example of this is the Saudi Arabian-U.S. solar energy water desalination project. In this project three solar seawater desalination systems are being studied right now. They are:

1. Indirect freezing process with point focus solar collector system.
2. Reverse osmosis and electrodialysis system with line focus solar collectors plus flat plate photovoltaic collectors.
3. Reverse osmosis and Multiple Effect distillation system with central solar receivers.

Based on the results of the feasibility studies, one process or more will be selected for construction of pilot plants. The Saudi Arabian-U.S. project will certainly have a major impact on the development of solar desalination.

CONCLUSIONS

Multistage flash (dual purpose) will continue to be the dominant seawater desalination method in the intermediate future. Reverse osmosis will be applied to the desalination of seawater more often, but the rate will really depend on the performance of existing major plants, like Jeddah 12000 m^3/day plant. Other desalination methods have not been commercially developed because of technical problems and lack of funds to solve them and to build major pilot plants. Because of the escalating costs of hydrocarbon fuels, stronger interest is developing among many nations to utilize solar energy in desalination. The most promising approach seems to be in converting solar energy into thermal or electrical form then using it to run hybrid desalination system.

The total production capacity of seawater desalination plants in Saudi Arabia is expected to increase, by more than ten times between now and 1987, to reach 1,900,900 CMD. This dramatic increase represents a major challenge to the country and to the desalination industry.

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ELECTRODIALYSIS

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ABSTRACT

The basic principles of electrodialysis are reviewed with special emphasis on the efficiency of energy utilization. Methods suggested in the literature for minimizing the polarization and its adverse effects are described. This includes the use of polarity reversal, ion-conducting spacers, ion-exchange resins between membranes and heterogeneous membranes. Effect of raising the temperature in decreasing power consumption leading to the development of HTED is discussed. Attempts to desalinate sea water by ED are also mentioned. Turning to solar application it is found that when DC power is being produced by PV generators the use of ED is most appropriate while when both DC and AC are available from solar sources a combination of RO and ED becomes an obvious suggestion.

INTRODUCTION

Electrodialysis, solvent extraction and reverse osmosis are selective transport methods in which salt or solvent is transferred away from the feed solution across some physical barrier without change in the state of the solvent, in contrast to distillation or freezing which depend on a change in the state of the solvent to bring about the desired separation. In electrodialysis partial separation of the components of ionic solution is brought about by placing across the path of an electric current sheets of material (membranes) which are permeable to one type of an ion and not the other. An ED desalination cell contains two different types of ion-selective membranes. One of the membranes is cation-permeable i.e. allows the passage of positive ions (cations) while the other anion-permeable membrane allows the passage of negative ions (anions). If a direct current is established across a stream of saline water passing between a pair of those membranes, ions acting as carriers of electricity will migrate across the stream. The cation-permeable membrane will permit positive ions (e.g. Na^+) to pass through while repelling negative ions (e.g. Cl^-), and the other membrane allows negative ions to go through but not positive ions. The membranes act as one-way check valves thus preventing the reentry of the ions they let through. Hence, the space between the membranes gets desalted while the streams on the electrode sides become concentrated with the penetrating ions. In practical ED desalting devices many pairs of membranes are used between a single pair of electrodes, forming an ED stack. Plastic separators are inserted in the solution compartment to keep the membranes apart and to promote mixing. The cells can be stacked either horizontally or vertically. The saline water flow is divided into many small streams in which most of the current-carrying ions are trapped in half of these streams and desalted water in the other half. The amount of electric current requirement varies proportionately to the amount of dissolved salts to be removed. Increasing the number of pairs of membranes between the electrodes increases the efficiency of current utilization. The practical optimum number of pairs for any given design will depend upon mechanical assembly problems, sealing against water leakage, and on insulation and control of large dc electrical voltages. Most designs vary between 275 and 500 pairs of membranes per pair of electrodes.

Components of ED Unit

The most common ED scheme was described by Meyer & Strauss (1940) and its commercial application was pioneered by Tudor et al (1953). This scheme is illustrated in the simple diagram shown in Figure 1. It comprises a series of alternating cation-permeable (C) and anion-permeable (A) membranes set between two electrodes with saline water being fed in between the membranes. As the dc current passes through the system the water streams become alternately diluted and concentrated giving rise to a product and concentrate. The combination of membranes, spacers, gaskets, electrodes, together with necessary devices for dividing, directing and collecting the streams of water is called an electrodialysis stack. Each pair of membranes sealed around the edges to prevent leakage and separated by spacers to provide for water to flow across the membrane surface is called a cell.

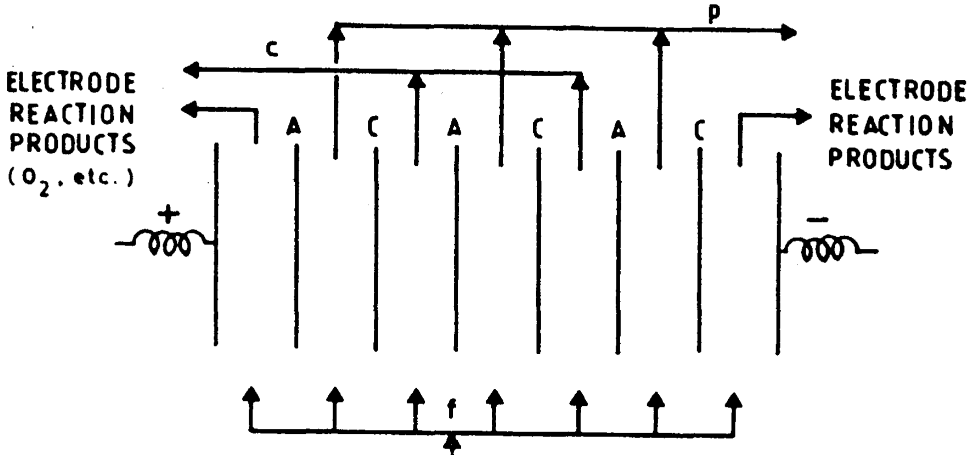


FIGURE 1 CONVENTIONAL ELECTRODIALYSIS

A, anion-permeable membrane; C, cation-permeable membrane; f, feed; p, product; c, concentrate

The thickness of the spacers and gaskets determine the membrane spacing and hence the width of path for the water stream. Spacer and gasket design determines the pattern of flow across the surface of the membranes. There are two types of water flow, namely, sheet flow and tortuous path flow. The tortuous path flow arrangement gives rise to a much longer flow path.

CONCENTRATION POLARIZATION

The movement of ions toward the membrane is governed by a potential gradient, a concentration gradient and by the hydrodynamic characteristic of flow. In the flow channel between the membranes, there is a central turbulent core with two thin viscous boundary layers adjacent to the membrane surfaces. Whereas ions migrate rapidly in the turbulent core their migration through the boundary layers is very slow and is governed primarily by diffusion. In addition, the number of ions at the surface of the membrane is reduced by the withdrawl of ions required to act as current carriers through the membranes. This increases the drop in voltage across the boundary layer to the extent that it will cause ionization of water. The resulting ions will participate as current carriers and decrease the need for the transport of solute ions through the boundary layer. While this is true at all current densities, it is only at high current density that the voltage drop is high enough to cause appreciable water splitting. Since the tendency to polarization increases with current density there are limits to the current that can be used in an ED system. The limiting or critical current density beyond which polarization is certain to occur is very hard to predict due to the complicated nature of the phenomenon. However, there are some indicators which can be experimentally determined to suggest approximate limits. Cowan [1] devised one of these methods in which the ratio of the voltage to current (V/I) is plotted against I^{-1}, the resulting curve passes through a minimum. The current which corresponds to this minimum is taken as the critical value of current.

It is worth mentioning that polarization in limiting the current density to an upper limit, has an adverse effect on the economy of the process. This is because, the lower the current density the larger the membrane area needed and hence the higher the cost.

Polarization takes place more actively at the anionic membrane with the result of OH^- ions migrating through the membrane to the concentrate compartment leaving the H^+ ions in the product compartment. This of course will cause the pH on the product side to decrease and on the concentrate side to increase. The increase in the pH of the concentrate stream tends to cause scale deposition due to the combination of the hydroxide ions with magnesium ions, which are most probably present in the concentrate stream. The resulting magnesium hydroxide is relatively insoluble and tends to form a scale on the membrane.

Limiting Current Density

There is no method by which polarization can be completely prevented[2]

In actual practice, polarization is held to reasonable values by controlling the ratio of the current density to the normality of the dilute stream i/N and by designing the system to make the boundary layer thickness ∂ as small as possible. There are two ways to reduce ∂ namely, high velocities (high flow ratios and/or small channel cross section) or the use of eddy promoters. Both of these measures need more pumping power but they reduce the stack power consumption and shift the polarization point to higher current densities, both of which mean better economy. In principle, the diffusion layer thickness depends on the channel geometry, distance from solution inlet, and flow velocity. In practice approximate average values of ∂ are used which are determined from approximate design equations of the type [3] :

$$\text{polarization parameter} \equiv \left| \frac{i_{\text{lim}}}{C_o} \right| = \left| \frac{F\ D}{\partial(\bar{t}_- - t_-)} \right| = \alpha\ \mu^{\beta} \quad \text{where,}$$

i_{lim}	limiting current density
C_o	bulk electrolyte concentration in g equivalents/cm^3
F	Farady's constant coulombs/g. equivalent
D	the diffusion coefficient of electrolyte in solution cm^2/sec.
∂	local thickness of diffusion layer, cm
$\bar{t}_-, t_-$	transport number of anions in membrane and solution respectively
u	linear velocity in cm/sec.
α β	empirical constants characteristic of each spacer material. For some spacers β is close to 0.6

Lacey [4] presented data from which boundary layer thickness ∂, pressure drop and pumping power can be obtained as a function of velocity from figures such as those given in Figure 2

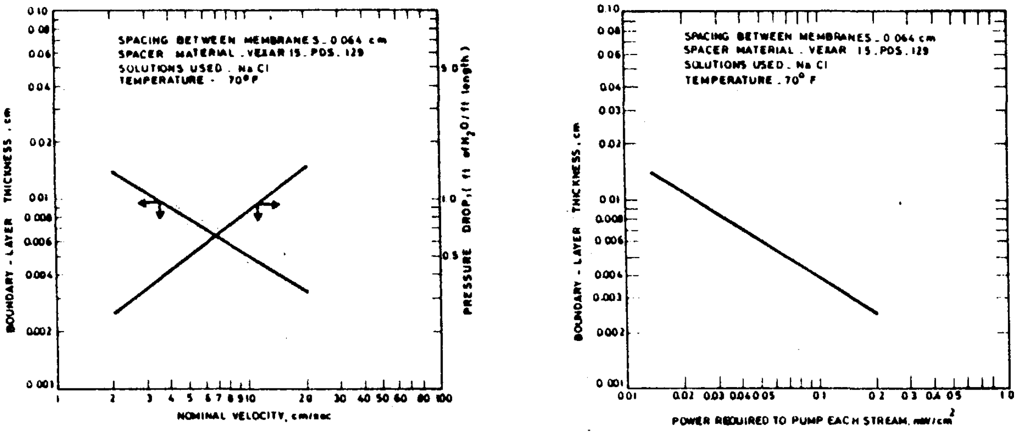


FIGURE 2 EFFECT OF SOLUTION VELOCITY ON BOUNDARY LAYER THICKNESS, PRESS DROP AND PUMPING POWER

Probstein et al in 1972 [5] reaffirmed that the most critical parameter governing the system characteristics is the laminar sublayer diffusion thickness. In 1974 Isaacson and Sonin [6] developed a general scheme for choosing the optimum channel geometry and flow conditions. The best flow conditions were found to be obtained with channels containing eddy promoters, either strips or mesh-like obstructions to create eddying flow. Later on the same authors [7] developed correlations for an empirical Sherwood number and friction factor which showed that eddy promoter spacing of about four channel thicknesses represents the best spacing from the point of view of process economics. They also concluded that the strip-type promoters they tested have a performance equivalent to or better than the best of other types of eddy promotion systems which have been tested in the Reynolds number range of interest in ED systems. They defined a quantity they called J_{lim} which although does not represent an absolute physical upper bound to the current density, yet it represents a practical limit beyond which, not only power increases but water dissociation, pH changes, precipitation hazards and other undesirable effects take place.

The operating current density i must be chosen sufficiently below the limiting current density. That is

$$i = \alpha \, i_{lim}$$

where α is smaller than unity to ensure essentially polarization free operation. The precise value of α is controlled by the specific harmfull effects which one is trying to avoid by avoiding polarization. Values of α as low as 0.7 are used.

EFFICIENCY OF ENERGY UTILIZATION

In practical ED systems, we find that the amount of current required to produce a certain amount of desalting exceeds that calculated for flow through ideal membranes. This is due among other things to:

i) membranes being less than perfectly selective i.e. ions of the wrong electrical charge sign passing through the membranes
ii) current leakages through stack manifolds
iii) osmotic transfer of water may accompany the flow of current across the membrane
iv) water splitting at high current densities.

The ratio of the number of chemical equivalents of salt ions removed to the number of Faradays of electricity used is defined as current, Faraday or coulomb efficiency η_c and can be expressed as follows [8]:

$$\eta_c = \frac{P(C_1 - C_2)F}{I \, N_p}$$

where, η_c is the current or coulomb efficiency
P is the rate of product water in Kg/sec.
C_1, C_2 concentration of incoming and outgoing product streams in g. equivalent/Kg.
F Faraday's constant (96000 coulombs/g.equivalent)

I	current in ampere
N_p	number of membrane pairs.

Current leakage by conduction through the manifold system is a function of the relative areas of manifold cross section to active membrane area. Its probable effect is a reduction in efficiency of less than 5%.

Water transfer by osmosis is due to the osmotic pressure differences of the solutions on either sides of the membrane and proportional to the membrane porosity to water molecules. When water is transferred from the product stream to the concentrate stream this means:

a) more current has to be used to overcome drop in water quality due to this transfer in order to meet product quality specifications
b) reduction in the quantity of product recovered from a certain amount of feed.

This source of inefficiency is only serious when high salinity feed and/or high degree of desalination are involved. Also, water splitting which results from concentration polarization deserves special consideration.

Overall Efficiency

Principal sources of inefficiencies or energy sinks can be summarized as follows:

Current terms	Voltage terms
Permselectivity	Electrode reactions
Manifold leakage	Stack resistance
Water transport	Polarization potential

Hence, the overall efficiency (η) of an electrodialysis stack [2] is the product of six principal efficiency terms:

$$\eta = \eta_s \ \eta_m \ \eta_w \ \eta_e \ \eta_R \ \eta_C$$

$\eta_s \times \eta_m = \eta_C$ the Faraday or coulombs efficiency

$\eta_s \times \eta_m \times \eta_w = \eta_I$ the overall current efficiency

When water transport is small $\eta_F \simeq \eta_I$

While ($\eta_s \times \eta_w$) is a function of membrane selected, η_m is a function of manifold configuration. The efficiency term η_w dealing with water transport is given by:

$$\eta_w = 1 - \frac{\eta \ \bar{T}_w(0.018 \ m_i)}{\eta_s}$$

where T_w = transference number of water in the membrane

m_i = molality of feed water, while the term η_c is given by:

$$\eta_e = 1 - \frac{E_e}{E_T}$$

where E_e electrode potential (2.5-3.5V) and E_T total stack potential. As the number of cell pairs increase η_e approaches unity. The term η_R depends on the resistance of the individual components of the stack (bulk solutions resistances, boundary-layer resistances, membrane resistance). It will also depend on the particular configuration and flow patterns selected, the type of turbulence promoters placed in the compartments and the temperature. The last term η_c depends on the boundary layer concentration profiles and the extent to which they give rise to polarization which results in increase of energy consumption.

The major inefficiency in stack operation is associated with η_R η_C which explains why most of the progress in electrodialysis technology is a result of attempts to decrease total resistance and minimize polarization and its adverse effects.

ENERGY REQUIREMENTS

Processes that relay on a change in phase, such as distillation involve a high rate of energy circulation. The energy needed to operate the system will only be a fraction of the latent heat of the solvent and independent of the amount of solute present. Losses and inefficiencies tend to be proportional to the energy circulation. On the other hand, energy required for desalting processes depending upon selective transport as ED will depend on the salinity of the feed used and product obtained. The theoretical energy required U at various salinities of feed, product and blowdown has been derived by Spiegler in 1956 [2] based on general thermodynamic arguments and independent of the details of the actual separation process.

$$U = 5.21\ (N_f - N_p)\left(\frac{\ln \beta}{\beta-1} - \frac{\ln \alpha}{\alpha-1}\right)\ \text{KWh/Kgal}$$

N_f, N_c, N_p concentration in equivalent/ℓ of feed, concentrate and product respectively.

This theoretical energy corresponds to a case where the driving force is infinitesimal and hence the process takes place slowly and reversibly. In actual systems more energy is required in order to maintain large driving forces necessary to obtain practical production rates. In most ED systems energy required for practical production rates may be 10-20 times higher Shaffer and Mintz [2] gave theoretical and actual energy requirements as a function of the salinity of the feed for ED in comparison to distillation as shown in Figure 4. It is clear from the figure that ED requires less energy than distillation for salinities up to 11,000 ppm. The same authors derived an equation for the calculation of the energy actually required for desalination of F' gallons per hour of feed water having a concentration N_f equivalents/ℓ to a product of N_p equivalent/ℓ with an average diluate concentration N_d equivalent/ℓ in a cell with membrane area A square feet and resistance R ohm and current efficiency η_c

$$\text{KWh/Kgal} = 25.4 \frac{F'}{A}(N_f - N_p)\left(\frac{R\,N_d}{\eta_c^2}\right) \log\left(\frac{N_f}{N_p}\right)$$

For continuous operation of multistack plants where stacks are connected in series and the product of one is the feed for the next stack, when same voltage is applied to each stack, Mintz [9] in 1963 showed that the total d.c. power required Q' is given by

$$Q' = \left[\frac{F^2}{A} \ln \frac{q}{N_p}\left(\frac{R\,N_d}{\eta^2}\right)\right] (F')^2\, N_p (q-1)(1 + q + q^2 + \ldots q^{n-1})$$

where $q = \frac{N_f}{N_p}$ and n is the stacks in series, the rest of notations have the same significance as in the previous equation.

Data on power consumption for some existing ED plants are given in Table 1 which show the effect of some of the parameters such as feed concentration, concentration ratio, plant capacity on power consumption. When plotted in Figure 3 these data fall in expected regions and follow more or less within the expected pattern of variation.

TABLE 1

REPORTED POWER CONSUMPTION OF SOME ED PLANTS

Concentration (ppm) feed	Concentration (ppm) product	Concentration ratio feed/product	Capacity Kgal/day	Power Consumption KWh/Kgal	Type of process	Ref.
845	190	4.45	500	6.9	---	10
1400	500	2.8	4000	5.1	EDR	11
2358	573	4.1	2100	7.5	---	12
2400	500	4.8	500	13.6	---	13
3000	550	4.45	170	5.0	EDR	14
3000	500	6.0	516	2.7	---	15
35000	500	70.0	98	32.9	HTED	16
35000	500	70.0	38.7	61.3	---	17

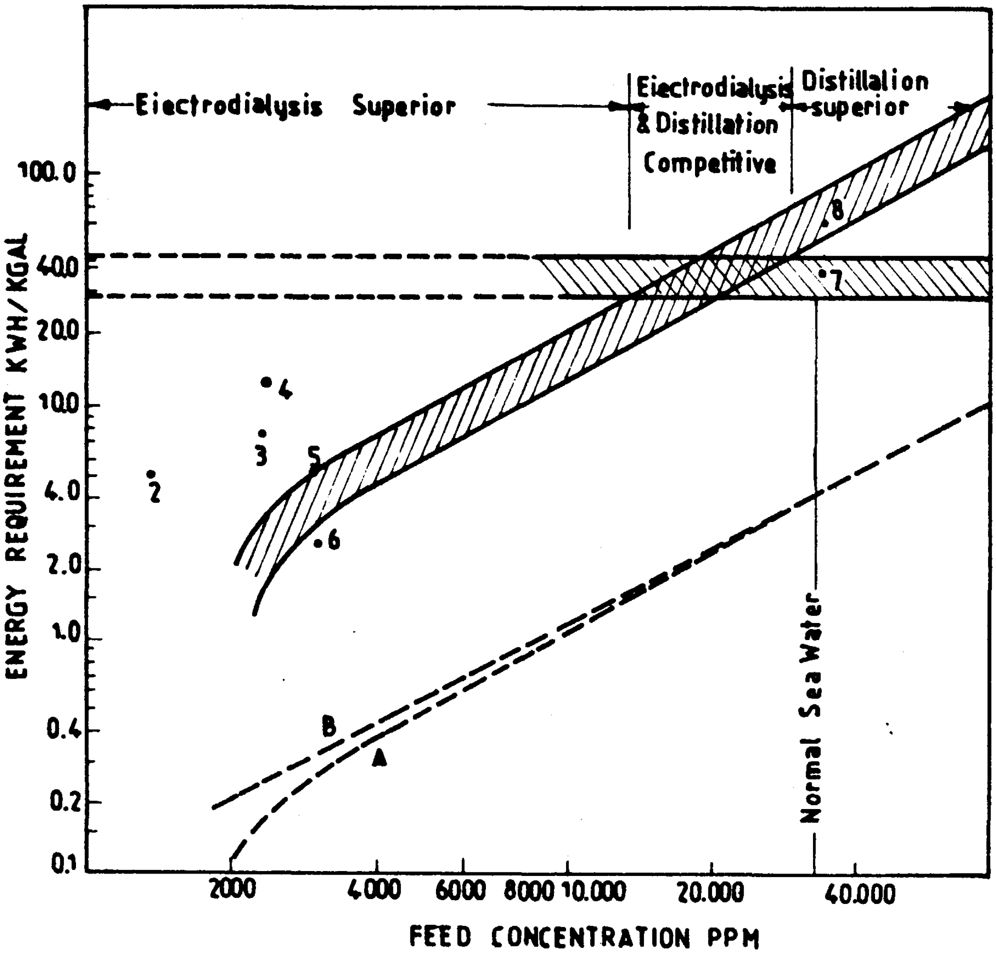


FIGURE 3 ENERGY REQUIREMENTS FOR DISTILLATION AND ELECTRODIALYSIS OF SALT WATER.

Basis: Average equivalent weight of salt = 60; blowdown concentration twice feed concentration; distillation to produce pure water; and electrodialysis to produce 0.005-N (300 ppm) product. (A) Theoretical energy for electrodialysis; (B) theoretical energy for distillation; (C) estimated actual energy for electrodialysis; and (D) estimated actual energy for distillation. Plotted points show actual energy consumption for some ED plants, numbers refer to order in which data are given in Table 1.

DEVELOPMENTS AND PROCESS VARIATIONS

The Electrodialysis Reversal "EDR" Process

Although referred to in OWRT reports in 1975 and 1977 the first published information about this process appeared in 1979 "Desalination" by Katz [18]. The EDR process is an improved electrodialysis process in which the surfaces of the membranes are cleaned periodically several times/hr by the reversal of the DC flow through the membrane array. It is the only desalination process having this self-cleaning feature which prevents the scale formation without the use of chemicals. Prior to 1970 acids, polyphosphates or similar agents were the only means with which insoluble salt precipitation could be controlled. However, while effectively controlling calcium carbonate scale, these chemicals have less control over calcium sulfate scale. Although experimentation started in the 50's and small commercial units (500 to 3000 gpd) were tried and proven in the 60's, it was only in 1970 that the polarity reversal principle was applied on larger commercial units. By 1974 all Ionics ED units were being designed using the EDR principle. The largest of these units is a 15000 m^3/day (4 mgpd) plant at Corfu in Greece.

Description of the EDR Process: The EDR process uses a standard ED array of alternating cation and anion membranes, separated by alternating product and brine compartments. The array is operated in the standard ED manner for a fixed period of time around 20 minutes and then the process is reversed by an automatic timing circuit in the following sequence:

1. The direction of the DC field is reversed by reversing the polarity of the electrodes. This polarity reversal immediately begins converting the product compartments into brine compartments and vice versa, by reversing the direction of flow of the ions.

2. Automatic valves interchange the feed to and discharge from the product and brine compartments.

3. There is a 1-2 minute period immediately following the polarity reversal when water from both sets of compartments is "off-specification" and both streams are automatically diverted to waste. This "purge" of the brine and product compartments every 20 minutes breaks up polarization films, carries off loose scale and seeds of scale.

The Corfu EDR Plant: The membrane stacks of the Corfu EDR plant are arranged in six parallel modules [11]. There are ten stacks in each module, and each stack contains 380 cell pairs. The total membrane area in the six modules is approx. 21,000 m^2. Each individual module is further divided into four parallel desalting lines. Two of these lines have two stages and the other two have three stages. Approx. 45% of the feed flow passes through the 3-stage line and is desalted from 1400 to 462 ppm TDS. The other 55% passes through the 2-stage line and is desalted from 1400 to 532 ppm TDS. The two product water are blended to give a 500 ppm final product. Power consumption was reported to have been 1.36 KWh/m^3 (some 20% lower than contractual value).

High Capacity Unconventional Electrodialyzer

Electrodialyzers can be classified as sheet flow type or tortuous path type. Sheet flow uses relatively low linear velocity resulting in low press.drop. However due to limitations of path length, it is not possible to attain the necessary concentration ratio normally higher than 3 except by batch operation or by using multistage systems. Tortuous path type electrodialyzers are meant to compensate for the disadvantages of the sheet flow type. Even with this type it is not always possible to attain a satisfactory desalination ratio using a single stage. Consequently, it is usual to employ two or three electrodialyzers in multi-stages. In addition, due to the use of high linear velocities pressure losses are large.

Experience in sea water concentration using large electrodialyzers led to the use of membranes having large unit area and effectively utilizing the advantages of both sheet flow and tortuous path types to design an electrodialyzer having the following features [19]:

- High concentration ratio (13.3), i.e. 92.5% demineralization.
- Large production capacity (product approximately 5000 m^3/day per unit.)
- High water recovery (up to 85% with scale-free operation).
- Low electric power consumption (1.1 KWh/m^3 to desalinate water from 3000 ppm to 500 ppm). Low power consumption results from reduction of current leakage since the same water is fed as dilute and concentrate. In addition the flow rate of the concentrate is considerably lower than that used in the circulation method leading to lower pumping energy.

Results of measurements of operating performance and continuous operation are given in Figure 4 as reported by Kawahara et al [19].

Polarization Reduction Using Ion-Conducting Spacers

Preliminary investigation for the possibility of using ion-conducting spacers in an attempt to reduce polarization was conducted by Kedem [20]. This work is based on the idea of polarization reduction being achieved by increasing the surface of contact between solution and resin phases. Kedem suggested different configurations for the ion-conducting spacers. He considered his results only preliminary but demonstrating that ion-conducting spacers reduce resistance and increase current density without sacrificing current efficiency with polarization being reduced in the usual working range.

Use of Ion-Exchange Resins Between Membranes

The effect of introducing ion-exchange resins between permeselective membranes was tested by Korngold [21] in a laboratory scale electrodialysis unit. His results indicated that polarization is substantially decreased, and high electrical efficiency can be obtained even at high i/C values of 1000-5000 mA cm^{-2}/g.eq.lit^{-1} and at low linear velocity of solution (1.2 cm/sec) in the cell. Different ion exchange resins were tested. Among the features he quotes for the system he proposes, are:

1. It can operate with thick cells of up to about 10 mm. Operation is simpler in these cells and leakage between diluate and brine cells can easily be eliminated. Flow distribution between the cells can be maintained evenly so that the danger of salt precipitation in the cells is substantially decreased.

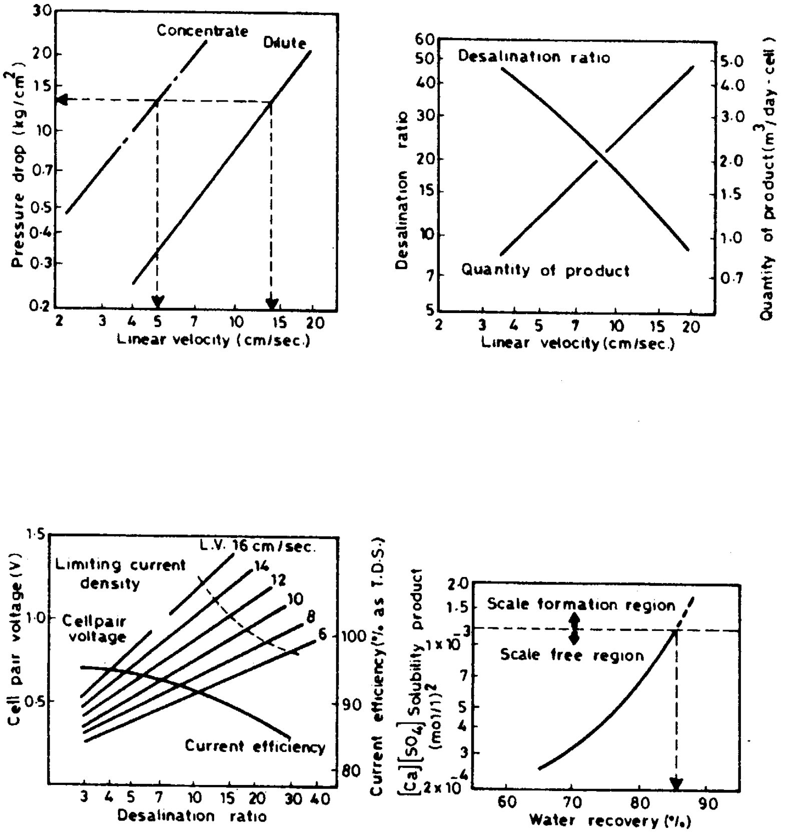


FIGURE 4 PERFORMANCE CHARACTERISTICS OF HIGH CAPACITY ELECTRODIALYZER AS REPORTED BY KAWAHARA ET AL.

2. Ion exchange resin placed between the membranes reduces the electrical resistance of the stack. For brackish water desalination, 10 mm cells normally require 6-7 times more energy than 1.3 mm cells. However a resin-filled 10 mm cell requires only about twice the energy of an unfilled 1.3 mm cell.

3. The limiting current density can be increased 3-4 times when ion exchange resin is introduced between the membrane.

4. In choosing the type of resin, interaction between bivalent and trivalent ions and the resin must be taken into account. Such interaction can dramatically increase the electrical resistance of the stack.

Heterogeneous Membranes

Kishi et al reported in 1977 [15] that they have developed heterogeneous membranes which consist of ion exchange resin particles (diameter less than 20) dispersed in polypropylene matrix. After dispersion of the resin particle, micro cracks in the propylene matrix and cavities between resin and the matrix are formed. Table 2 gives properties of their membranes.

TABLE 2

PROPERTIES OF HETEROGENEOUS MEMBRANES

Property		Cation	Anion
Thickness	mm	0.36	0.38
Bursting Strength	Kg/cm^2G	3.1	2.3
Specific Resistance	Ω cm	360.0	170.0
Ion Transport Number	-	0.93	0.92
Ion exchange Capacity	meq/g.dry resin	2.5	1.0

Exposure to temperature up to 50°C has no effect on Ion Transport Numbers. The membranes show high durability and stability of properties in strong alkalis, acids and oxidizers. These make it possible to employ effective hard chemical cleaning for fouling. A test plant using such membranes was run to determine the degree of fouling and find the most adequate method of cleaning. From the results of the cleaning tests 3% caustic soda cleaning was selected and 10,000 hr operation without disassembling has been carried out successfully. Based on the results of this test plant a large scale desalination plant was contructed in Kohl Al Zubair for the Government of Iraq having a capacity of 208 $m^3/hr \times 2$ (1.32×10^6 GPD x 2). A fully automatic chemical cleaning system has been designed to be used in the Kohl Al Zubair plant. The chemical cleaning method is fundamentally the same as the regeneration procedure of ion exchanger bed. Electrodialyzer section has two trains; each train has 10 (1 is stand-by) stacks.

Bubble-Cleaning Type Electrodialyzer

As reported in 1977 by Kuroda et al [22] a new type of electrodialyzer has been developed in which air bubbles mixed with water are introduced continuously between the membranes. This gives rise to the following features: (a) the agitation of the solution in the dialyzer by the air bubbles prevents suspended solids from fouling the membrane i.e. prevents soft scale, (b) the turbulence produced by the air bubbles diminishes concentration polarization, (c) flow takes place by an air lift. Kuroda et al report that turbulence promotion caused by air bubbles make it unnecessary to use spacers which act as turbulent promoters. Instead they used gaskets provided with several rectangular compartments. Results obtained from a 10 m^3/d test unit indicate that air bubbling is quite effective in turbulence promotion and hence the use of current densities twice that of conventional dialyzers are suggested. Pressure drop is reported to be slight when compared to that encountered with other types of electrodialyzers. In addition, since solutions are air-lifted, power consumption is expected to be lower.

HIGH TEMPERATURE ELECTRODIALYSIS

The fact that conductance of salt solutions increases rapidly with temperature indicates that the use of elevated temperatures in ED will result in lowering ohmic power losses and hence higher efficiencies. Investigations on HTED were first reported in 1965 by Forgacs during The First International Symposium on Water Desalination and again at The Second International Symposium Held in Athens, Greece in 1967 . Their initial calculations indicated that by increasing the temperature of operation from ambient to 90°C the energy consumption is expected to be reduced as much as 60-70%. The variation of energy consumed with temperature as published by Bejerano et al [23] show that decrease in power consumption is much more pronounced below 70°C than above it . This results from two factors: (a) that membrane resistance decreases much more rapidly below the range 70-80°C than above it, (b) that electro-osmotic water transfer through membranes increases, especially in the higher temperature region (70-100°C). Hence, Forgas et al concluded from their work that most of the advantages of HTED will be realized when working in the range of 70°C. At this temperature range ED equipment can be constructed and operated utilizing commercially available materials used for conventional ED. In the same Smposium in 1967, McRae et al reported the results of their bench scale studies of HTED which was later published in "Desalination" [24]. They measured response of membrane properties such as permeselectivity, stability, resistance and water transfer, to high temperatures. Their results indicate that potentials at least as high as 3 volts per cell can be applied at 82°C (180°F) without polarization. Since at 80°C it will be possible to accomplish a degree of mineralization in one stage which would take more than two stages at 25°C using the same potential and superficial velocity, a considerable reduction in pumping energy required for a given demineralization duty should be expected. They concluded from their economic studies that if an ED plant of a capacity of one mgpd or more built in connection with a steam turbine electric generating station, sea water demineralization at 82°C (180°F) is expected to have a cost in the same range as that of dual purpose MSF plants of 100 mgpd capacities.

Experimental bench scale study of sea water desalination using HTED was reported in 1968 by Forgacs et al [24] which show that power consumption decreased by about 63% when operating temperature was increased from 30°C to 70°C. Factors governing scale formation were investigated and measures to eliminate scale formation were suggested. HTED pilot plant tests were conducted [16] at temperatures up to 80°C on sea water and diluted sea water. These tests furnished data required for the design of a 380 m^3/day plant for the desalination of sea water at 65°C. Stack design centered [25] on a large spacer with an effective area of approximately 1.2 sq. meters. A process flowsheet for a continuous flow plant with two concentrating stream recirculation loops has been selected as the most reliable system to achieve the goals of 0.16m^3/day per m^2 effective cell pair area and 8 KWh/m^3 D.C. energy consumption. The flowsheet has been designed to give a maximum concentration ratio of brine to product of 20, with 50% water recovery.

Another HTED sea water desalination plant (50,000 gpd) has been contructed. Its design features and projected performance characteristics as reported by Mattson in 1979 [26] are as follows:

Single membrane stack expandable to 1600 cell pairs (100,000 gpd)
Thin (14 mil) cation and anion membranes
Large (54 x 40 in) membranes with over 80% area utilization
Design operating temperature
    140°F normal
    180°F maximum
Large (65 x 42 in) and thin (23 mil) gasket separators
30 KWh/1000 gals projected power consumption
5 gpd/ft^2 cell pair at 50 percent recovery

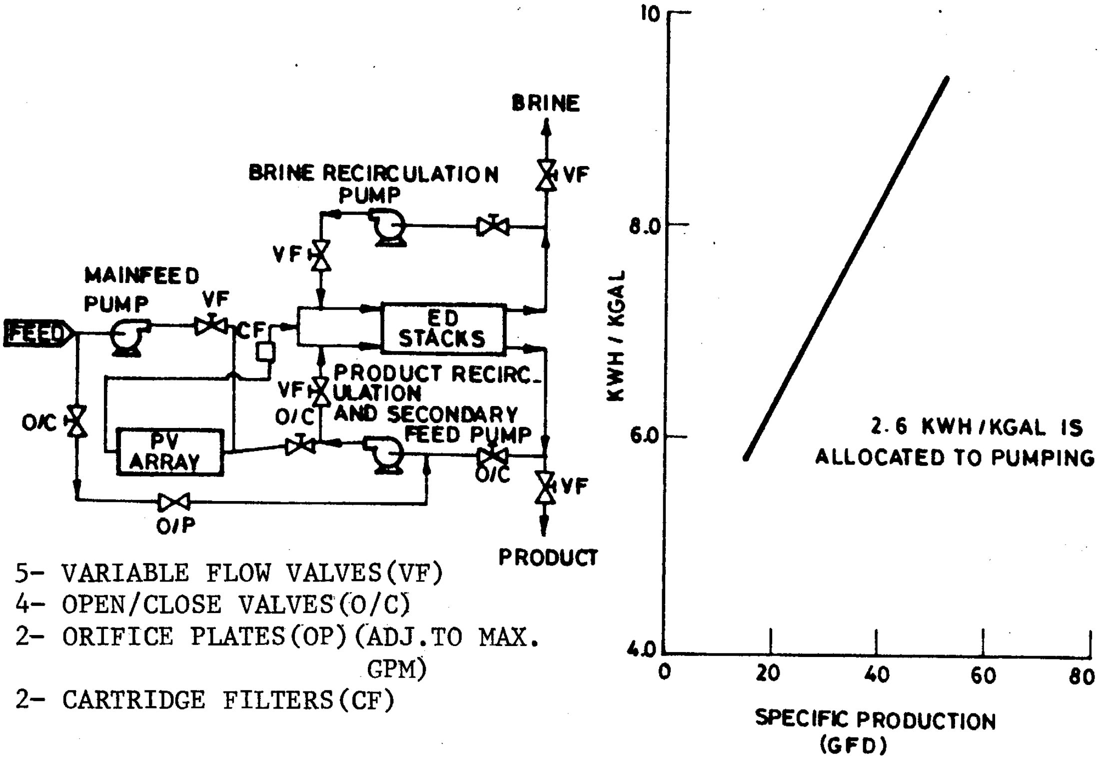
SEA WATER DESALINATION BY ED

The first sea water desalination plant by ED in the world started its operation in September 1974 at Noshima island of Hofu city, Yamaguchi Prefecture, Japan [17] . The capacity of the plant is 120 m^3/day and the plant has been supplying fresh water (TDS 500 ppm) from sea water (TDS 35,000 ppm) to the 800 inhabitants of the island. Although the operation is batchwise, it is fully automatic to supply fresh water continuously using a storage of proper capacity. The stack is a filter press (sheet flow) type containing 400 membrane pairs (Aciplex K-101 and A-101), size 1.3 m x 1.3 m with an active area of 1.4 m^2 per sheet. Average current efficiency is between 80 and 85% while the electric consumption is approximately 16 KWh/m^3 of which 12 KWh for DC power and 4 KWh for AC power. The water cost is relatively high but still lower than that of conventional distillation processes of the same capacity.

Based upon a finding that the capital cost of distillation equipment per gallon per day (GPD) of installed capacity rises rapidly for plants smaller than a few million gallons per day (MGD) while that for ED equipment desalting sea water to drinking water quality remain essentially unchanged for capacities less than 1 MGD. Lunstrom in 1980 [27] projected that for plant capacities of 100,000 GPD the costs of both plants approach $ 7 per GPD installed. For smaller capacities ED plants will have a lower capital when the product is of drinking water quality. Lundstrom developed a method for the rapid assessment of design features and capital cost requirements for ED plants desalting sea water from salinity 35,000 ppm to 4 ppm. The result of his analysis indicate that in the range of 50,000 to 100,000 GPD capacity the preferred plant configuration is that consisting of two major stages (with interstage pumps) and each major stage containing eight electrical stages with only one hydraulic stage per electrical stage. Operating with a water recovery of 33 % a specific production in excess of 3.7 gallons per square foot of effective cell area per day is expected. Pressure drop per major stage within the membrane stacks of 40 psi is reported.

SOLAR POWERED ELECTRODIALYSIS

Many regions which lack fresh water sources have abundance of sunshine, which shows that there is a strong correlation between scarcity of water and availability of solar insolation. The fact that ED uses DC which can be obtained directly from solar insolation by photovoltaic generators suggests the inclusion of the ED stack as an active part of the electrical generation circuit. This unique property of ED as compared to other desalting processes can be used to minimze battery storage and reduce daily losses which are in the order of 15 to 25% of the collected energy. In addition, because the output rate of the system can be adjusted to respond to varying available power levels, the ED system can be run at partial output when this is deemed necessary while keeping the product water quality constant. Since energy efficiency of the ED system increases 1% per ^{o}F rise in inlet feed water, this indicates that reduction in the ED plant power demand is expected by cooling photovoltaic concentrating type collectors with the feed water. Lundstorm [28] indicates moreover that although a combination of a PV generator with an ED desalting plant is marginally cost-effective with PV generators cost running \$ 10 per watt. Continued increase in cost of fossil fuels and anticipated drop in PV generators cost makes solar powered electrodialysis SPED more promising. A process flowsheet is suggested by Lundstorm as seen in Figure 5 in which ED stacks are incorporated as an active part of the electrical circuit.



5- VARIABLE FLOW VALVES(VF)
4- OPEN/CLOSE VALVES(O/C)
2- ORIFICE PLATES(OP)(ADJ.TO MAX. GPM)
2- CARTRIDGE FILTERS(CF)

FIGURE 5 SOLAR POWERED ED FLOWSHEET AND POWER REQUIREMENT VARIATION WITH SPECIFIC PRODUCTION RATE @ 70 ^{o}F.

By varying the flow rate through the stack and the amount of recirculation, the resistance of the stack can be varied. Similarly, adjusting the flow from the feed pump, the voltage at which the plant operates for a given product salinity can be matched to the voltage of the PV generator and the current adjusts accordingly. At full flow, the specific production (GFD) is selected in such a way that the capital equipment costs of the ED plant and PV generator are balanced.

Whenever AC power is generated through the use of line focussing thermal collectors and DC power produced directly using solar photovoltaic cells, this suggests the use of a combination of RO which is run by AC and ED which is run by DC. The obvious arrangement in a case where sea water desalination is considered would be the use of RO for the primary reduction of the salinity of sea water followed by final desalination using ED. In addition, using the feed water to the ED stacks as a coolant increases its temperature with the added advantages of higher efficiency of energy utilization.

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CURRENT DEVELOPMENTS IN FREEZING PROCESSES

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ABSTRACT

The vacuum freezing process is presently being developed in two forms, both of which are designed to eliminate the large water vapor compressor which has limited module size and perhaps acceptance of the process in the past. A $95M^3$/day (25,000 gpd) AFVC process pilot plant is under test which uses a low temperature sodium chloride absorber loop in conjunction with a closed cycle mechanical refrigeration system in place of a vapor compressor. A $225M^3$/day (60,000 gpd) VFEA process pilot plant design has been completed that uses a sodium hydroxide absorbent cycle driving a low pressure steam ejector to eliminate the vapor compressor in a thermally driven vacuum process. New thrusts have begun for development of Indirect Freezing Processes using controlled conditions to prevent ice formation on heat transfer surfaces. An Eutectic freezing pilot plant is being constructed whose effluent is to be only salts and potable water. Freezing processes offer many possibilities for combinations of Rankine or Photovoltaic solar cycles using the reject heat for low temperature absorption refrigeration.

INTRODUCTION

Interest in freezing is accelerating because of its modest energy requirement and universality of application and in the present context because of its versatility for accepting solar energy in many forms. In addition to desalting process development efforts, increasing interest is emerging for a wide range of commercial applications in the food processing and chemical industries as well as water reuse applications.

ADVANTAGES OF FREEZING

High resistance to corrosion and fouling may, in the final analysis, be the principal advantages of freezing.

Corrosion is exponentially reduced with decreasing temperature. Several years of pilot plant operation [1,2] has shown that corrosion is negligible. In this respect, ironically, corrosion is more severe on the external surfaces of the process equipment due to ambient atmospheric conditions than it is to the internal surfaces.

Fouling of surfaces is not a significant factor in freezing, thus pretreatment is not a prerequisite of the direct contact processes. The actual concentration takes place in the bulk, i.e., existing ice crystals grow while in direct contact with subcooled brine. Sufficient work has not been done on Indirect freezing to indicate whether this is also true; however, the temperature, concentration change and salt solubilities at the process conditions tend to preclude fouling.

Freezing is relatively insensitive to the nature of the feed solution in that the separation (crystallization) step occurs within the solution itself. Thus, the nature of the solution, i.e., salinity, organic, inorganic, affects only the energy requirements because of freezing point depression, or the crystal size obtained. It does not affect the purity of the ice that is formed. The principal limitation as to the universality of freezing is perhaps in terms of viscosity. As solutions tend to become more viscous with concentration, such as with sugar solutions, it becomes increasingly difficult to separate the pure ice from the viscous concentrate, thus a limit is reached.

Energy requirements for freezing are demonstrably lower than in distillation processes. Mature freezing processes can confidently project energy consumptions in seawater desalting between 6 and 9 KW-Hr/M^3 (25 - 35 KW-HR/1000 gal.) The principal reason that freezing can project such low energy requirements vis-a-vis distillation is that the driving forces to effect crystallization and melting are a much smaller fraction of the intrinsic freezing point depression temperature difference than are the driving forces for evaporation and condensation in relationship to boiling point elevation. In addition, the heat to be transferred in freezing is approximately one-seventh (1/7) of that involved in distillation, thus any driving force penalties are partitioned over a smaller absolute energy transfer.

The comparison with membrane processes is not quite so advantageous. Here the practical energy requirements can be quite comparable. Even though the driving forces for reasonable fluxes in membrane processes are quite large on a percentage basis as compared to freezing, because of the absence of a phase change, the absolute energy requirements converge.

In this latter respect freezing in comparison to membrane processes relies principally on its insensitivity to fouling, minimal corrosion, absence of pretreatment, and longer equipment life.

CURRENT DEVELOPMENT

Vacuum Freezing and Indirect Freezing are the focus of attention in freezing process development at the present time. Both processes were the recipients of the earliest efforts in freezing in the 1950's. Vacuum freezing advanced with steady development until the late 1960's when attention was diverted to secondary freezing development. Work on Indirect processes was abandoned very early.

Carrier Corporation [1] and Colt Industries [2] built promising vacuum freezing pilot plants though neither effort saw successful commercial adoption. Colt's process employed a large vapor compressor which restricted the process to small modules [400M^3/day (100,000 gpd]. Carrier substituted an absorber for the compressor. This eliminated the intrinsic module size limitation; however, the advantages of this approach were never fully exploited.

Early investigations into Indirect Freezing by the University of Washington [3], Carrier Corporation [4], Applied Science Lab.[5], and Curran [6], among others, explored the possibilities of sheet and block freezing. Although it was established that these techniques could produce salt free ice, the projected economies,based on the high heat transfer resistance of sheets and blocks,as well as projected equipment costs,extrapolated from commercial ice producing machines were prohibitive.

The vacuum process has always held a certain fascination because of the direct contact heat transfer and the use of water itself as a refrigerant. Two evolutions of the vacuum process are currently being developed, the Absorption Freezing Vapor Compressor Process (AFVC) and the Vacuum Freezing Ejector Absorption Process (VFEA). Neither process requires a mechanical water vapor compressor. Work on Indirect Freezing resumed in recent years with emphasis on bulk as opposed to film freezing.

AFVC Process [7]

This variation of vacuum freezing has been pilot planted at the OWRT test facility in Wrightsville Beach, North Carolina at a nominal capacity of 95M^3/day (25,000 gpd.). This process substitutes a novel absorption system for the compressor whereby heat is transferred from the absorber to the generator by a closed cycle refrigeration system. The absorbent used is sodium chloride which has the advantage that any entrainment of brine from the crystallizer has only a negligible effect on absorbent contamination. Since the refrigeration systen can be optimized independently of all considerations other than temperature span, compressor size limitations do not inhibit module size as in the early vacuum (Colt) system.

Figure (1) is a schematic of the AFVC process which details the principal functions. The saline feed is first passed through a deaerator to strip out most of the dissolved air. After being chilled by the outgoing product and brine streams, it is introduced to the vacuum crystallizer. In the crystallizer a portion of the water is evaporated. This refrigeration effect freezes a portion of the feed in the ratio of the latent heats of vapor and ice, less a certain amount of evaporative cooling required to remove thermal intrusions. These result from the difference in temperature between the chilled feed and the concentrated brine and ice in the crystallizer, and sensible heat input from internal pump work. In the nominal case the feed is at -0.5°C (31°F) and in the case of a 2 to 1 concentration or normal seawater, the brine and ice are at -4°C (25°F). This results in a nominal ratio of ice to vapor evaporated of 6.7 to 1. The process

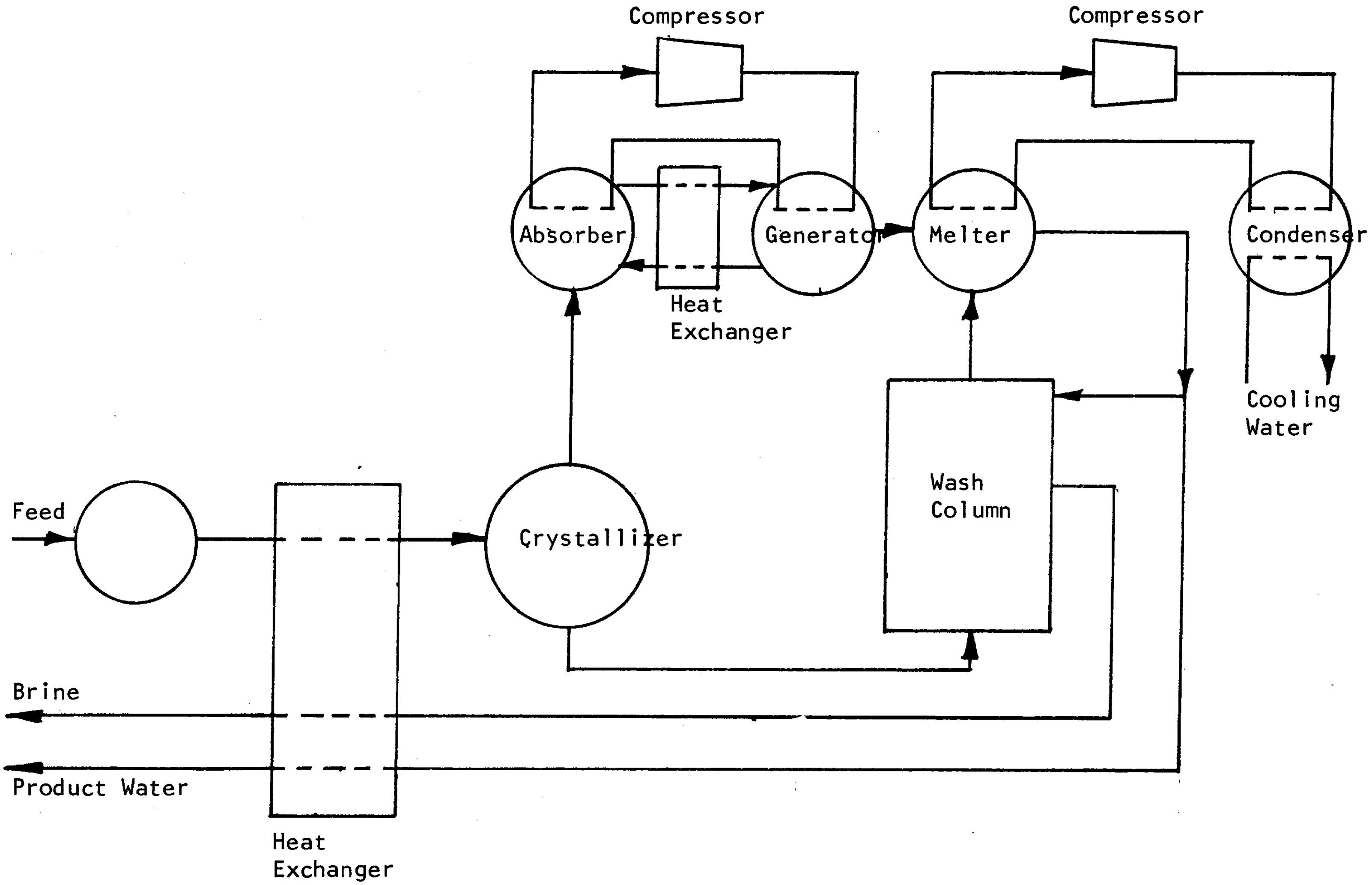

FIGURE 5
AFVC FREEZING SCHEMATIC

continues at a rate determined by the pressure difference between the vapor and the water phase in the absence of air. Evaporation rates range from 2300 to 4500 CAL/HR-CM^2-°C (5000 to 10,000 B/HR-FT^2-°F).

The slurry of small ice crystals, a typical major dimension is 250 microns, and brine is directed to the bottom of a countercurrent wash column. The ice and brine flow to drains located midway up the column which allow the brine to pass but retain the ice. The ice compacts into a permeable bed on which additional ice is continually deposited at the bottom as the brine flows through the bed towards the drainage screens. The flow through the porous bed produces an hydraulic force that propels the bed upward at the rate that ice is added at the bottom; this piston-like action forces the ice above the screens and through the brine level maintained by the screens, resulting in drainage of a portion of the brine entrained in the interstices of the ice bed. Potable water in a net quantity of approximately 5% of the resulting product is distributed over the top of the ice plug and percolates through the bed to the drainage screens. This action results in displacement of the residual brine leaving a bed of emerging ice crystals and adhereing potable water.

Following washing, the ice is melted by direct contact (in the absence of air) with vapor extracted for the crystallizer. The resulting melt, potable water, is extracted from the system, along with the brine from the wash column screens, through the feed heat exchangers.

Water vapor from the crystallizer is absorbed on a solution of sodium chloride at a nominal concentration of 15% and maintained at a temperature slightly above that of the brine in the crystallizer. This establishes a vapor pressure of the absorbent several tenths of a millimeter of mercury below that of the crystallizer brine, ensuring an adequate driving force for evaporation. The absorbent is maintained at this temperature through a refrigerated heat transfer surface; actually a tube bundle over which all the absorbent is percolated.

The diluted absorbent is continuously removed to a generator/concentrator and replaced by enriched absorbent to maintain the required average concentration. The dilute absorbent is concentrated (enriched) by transferring the heat, extracted by the refrigeration system, in the absorber at a higher temperature in heat pump fashion to the generator, again through a heat transfer tube bundle over which the dilute absorbent is percolated. The temperature of the generator is maintained by the refrigeration system at a nominal temperature of 3.3°C (38°F) which establishes a vapor pressure for the concentrated absorbent of 5 mm Hg. Under these conditions the water of absorption is evaporated from the dilute absorbent and condensed on the melted ice which has a vapor pressure of 4.6 mm Hg.

The concentrate is returned to the absorber after heat exchange with the incoming dilute stream.

The refrigeration system is a closed cycle mechanical compressor system, and is the primary work function in the process.

The water vapor evolved in the generator/concentrator has a larger thermal content than the ice can absorb, because of sensible heat extracted in the crystallizer due to the feed/concentrate temperature difference, internal pump work, and the work of the refrigerant heat pump. Consequently, a portion of the generator vapor is condensed on a separate heat transfer surface, chilled by a separate closed cycle refrigeration system, which rejects the heat of condensation to an ambient cooling stream.

The crystallizer, absorber, generator, wash column and melter are continuously purged to extract air from leakage and residual air from the feed stream.

The AFVC process has been conceived as a totally mechanical system; however, a combination mechanical/thermal variation will be discussed later.

VFEA Process [8]

Figure (2) defines a vacuum freezing system that utilizes thermal energy to completely eliminate the need for any type of mechanical vapor compressor.

The salient feature of this process is a combination absorption vapor ejector system which pumps the vapor from the crystallizer to the melter.

The process can operate efficiently with a temperature difference of 30°C (55°F) but as with distillation and other thermal cycles, the yield increases with increasing temperature differences up to a practical limit of a generator temperature of 175°C (350°F).

The absorber-ejector system has been studied on a laboratory pilot scale. The rest of the system, feed heat exchanger, crystallizer, wash column and melter, is identical to that described for the AFVC process and piloted by Colt and Carrier. A 225M^3/day (60,000 gpd) pilot plant design was commissioned by the Office of Water Research & Technology and completed in 1980.

As indicated in Figure (2), a portion of the vapor from the crystallizer is absorbed in a solution of sodium hydroxide (NAOH), at a nominal concentration of 50%, percolated over a tube bundle similar to that employed in the AFVC process. The heat of absorption is removed through the heat transfer surface by ambient cooling water. The concentration of the absorbent is selected to maintain the vapor pressure of the absorbent a few tenths of a millimeter of mercury below the vapor pressure of the crystallizer brine, while still maintaining the absorbent at a temperature above ambient, typically 32°C (90°F). Thus only pump work is expended in cooling the absorber.

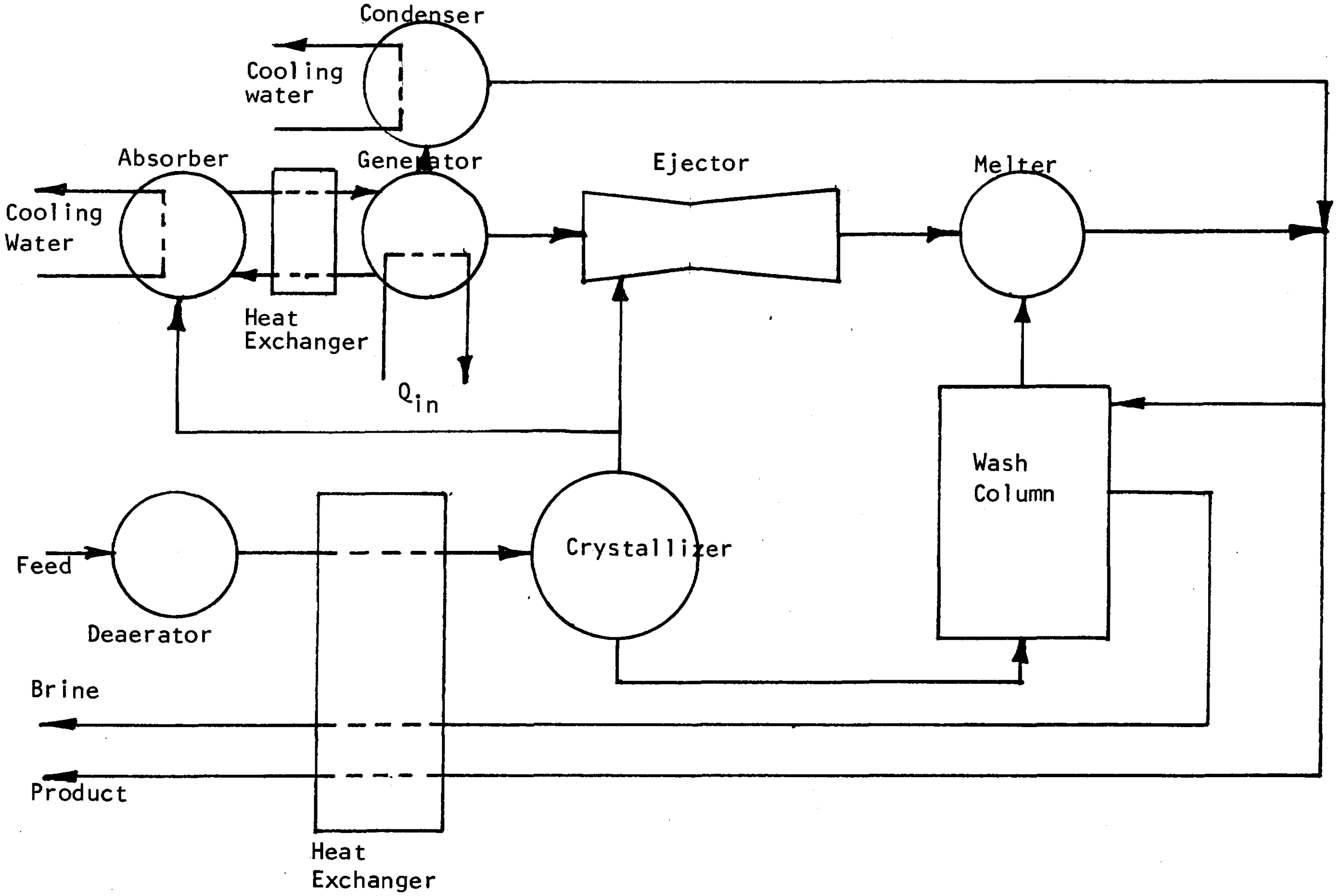

FIGURE 2
VFEA SCHEMATIC

Dilute absorbent is enriched in a generator/concentrator by external heat. The generator temperature can be varied as shown in Figure (3) with varying effect on process yield. The temperature in the generator is selected such that the concentrate will have a vapor pressure sufficient to allow the evaporated water to condense in the condenser at ambient cooling water temperature. This is the minimum generator temperature; however, any further increase in generator temperature will increase the enthalpy of the evolved water vapor which in turn results in improved ejector efficiency.

The concentrated absorbent is returned to the generator after heat exchange with the incoming dilute stream as in the case of the AFVC process.

The major portion of steam from the generator is used as the motive steam for an ejector. It compresses the greater portion of the vapor from the crystallizer to the required melter pressure. Typically from 3.2 to 5 mm Hg.

The efficiency of the ejector increases with increasing steam temperature as expressed by mass ratio and shown in Figure (4). Mass ratio is the ratio of motive steam to driven vapor, thus the smaller this ratio the less driving steam required and consequently the smaller the absorber and generator.

That portion of the vapor, evolved in the generator, which is condensed by ambient cooling water represents the excess thermal content or heat reject load, as defined in the description of the AFVC process.

Indirect Freezing

Research during the past two years, has shown the feasibility of freezing ice in the bulk of a saline solution on a continuous basis without adherence of ice on the heat transfer surface.

These efforts are leading to the construction of a $23M^3$/day (6000 gpd) pilot plant for extended evaluation of this method of freeze desalination on seawater at the government test facility at Wrightsville Beach, North Carolina.

The motivation behind this work is a simplification of the freezing process and an enhancement in the scaling of desalting plants to larger sizes. The total separation of the crystallization of ice from the refrigeration system eliminates air, refrigerant removal complications and entrainment problems associated with direct contact processes. It is also simpler to scale metallic heat transfer surfaces to large sizes in the manner of shell and tube heat exchangers, than it is to scale direct contact crystallizers. In addition, they are considerably more compact.

Though Indirect freezing introduces heat transfer surfaces that were not required in direct contact vacuum process as originally conceived,

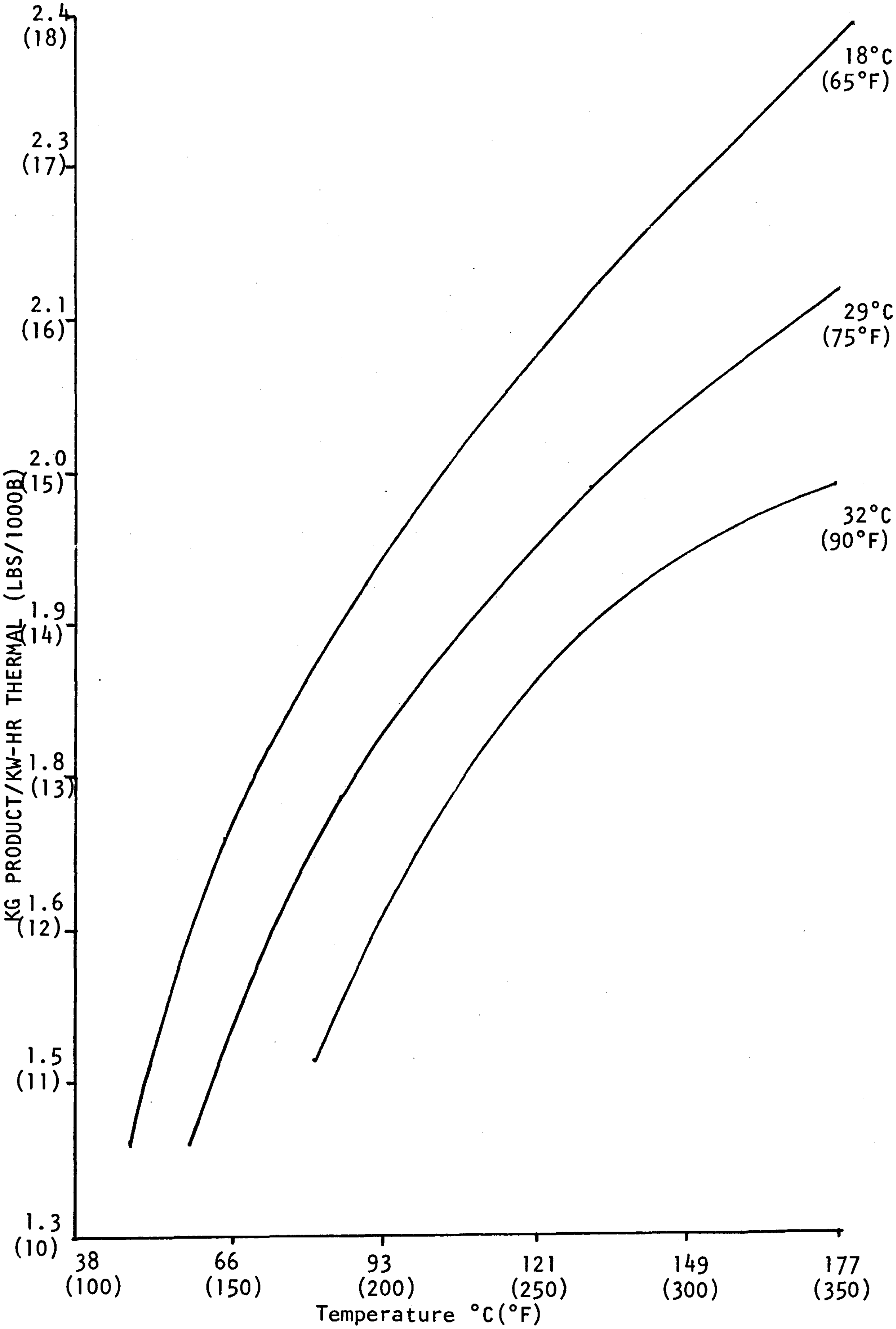

FIGURE 3 VFEA PERFORMANCE RATIO

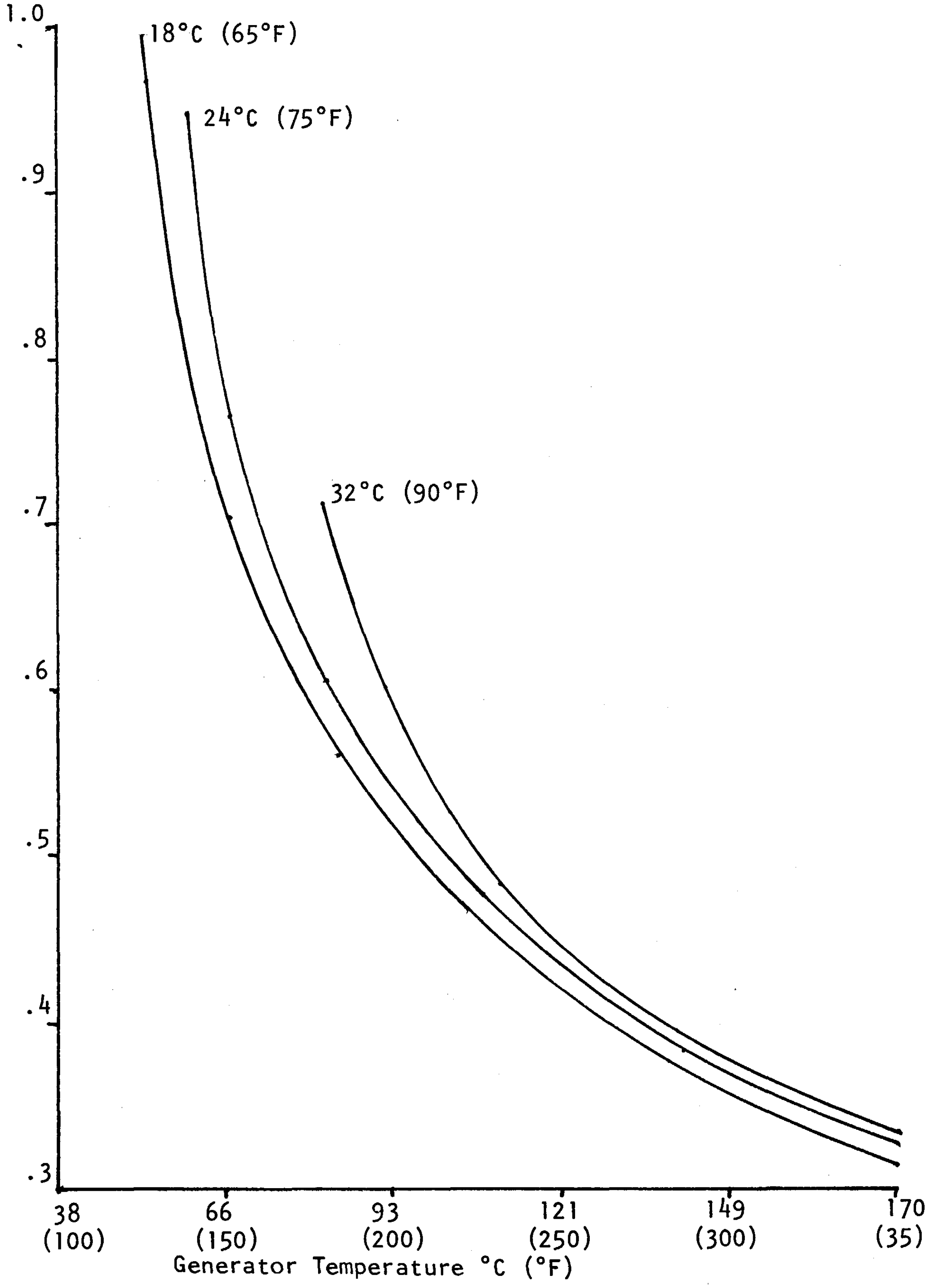

FIGURE 4

VFEA EJECTOR MASS RATIO vs TEMPERATURE

current utilization of absorption systems in place of vapor compressors in vacuum processes require these surfaces also.

Laboratory Studies

Research efforts on bulk freezing [9] were initially predicated on the analogy to controlled seeding in distillation plants to prevent scaling of the heat transfer surfaces. Early efforts to prevent ice growth on refrigerated surfaces by varying the quantity of ice in the feed stream, the flow velocity of the stream over the surface. The salinity of the solution and the refrigerant to brine temperature difference were singularly unrewarding.

The character of the heat transfer surface itself, it was learned, is the dominant factor in preventing ice adherence. In this respect it was also surprising to learn that the material of the surface was also a negligible factor, i.e., plastic as well as metal surfaces were equally susceptible to icing.

Icing of that surface does not occur if the surface is smooth and free of sharp discontinuities. The surface does not have to be free of undulations, however, very small surface scratches served as nucleation sites.

Electropolished metal, deposited coating such as paints, or bonded plastic films without seams are satisfactory. The ability to resist icing on smooth surfaces is also a function of salinity. Concentrations above 3.5% have been shown to resist surface icing. The higher the salinity the greater the temperature difference permissible before incipient icing.

Proprietary research is continuing on other approaches to indirect bulk crystallization. They promise to further decrease the sensitivity of surfaces to icing and to minimize the requirements for surface preparation.

The Indirect Process

The application of indirect crystallization to desalination is shown in Figure (5). The simplicity of this process is apparent in the comparison to the previous schematics.

The feed heat exchanger and wash column are identical to those required in the vacuum processes. The crystallizer, however, is composed of a multiplicity of smooth tubes enclosed in a refrigerant jacket, similar to a shell and tube heat exchanger.

Feed from the heat exchanger is pumped through the tubes wherein ice is crystallized in the bulk. The slurry from the crystallizer/heat exchanger is directed to a wash column where the ice and brine are separated. The washed ice is then pumped through another shell and

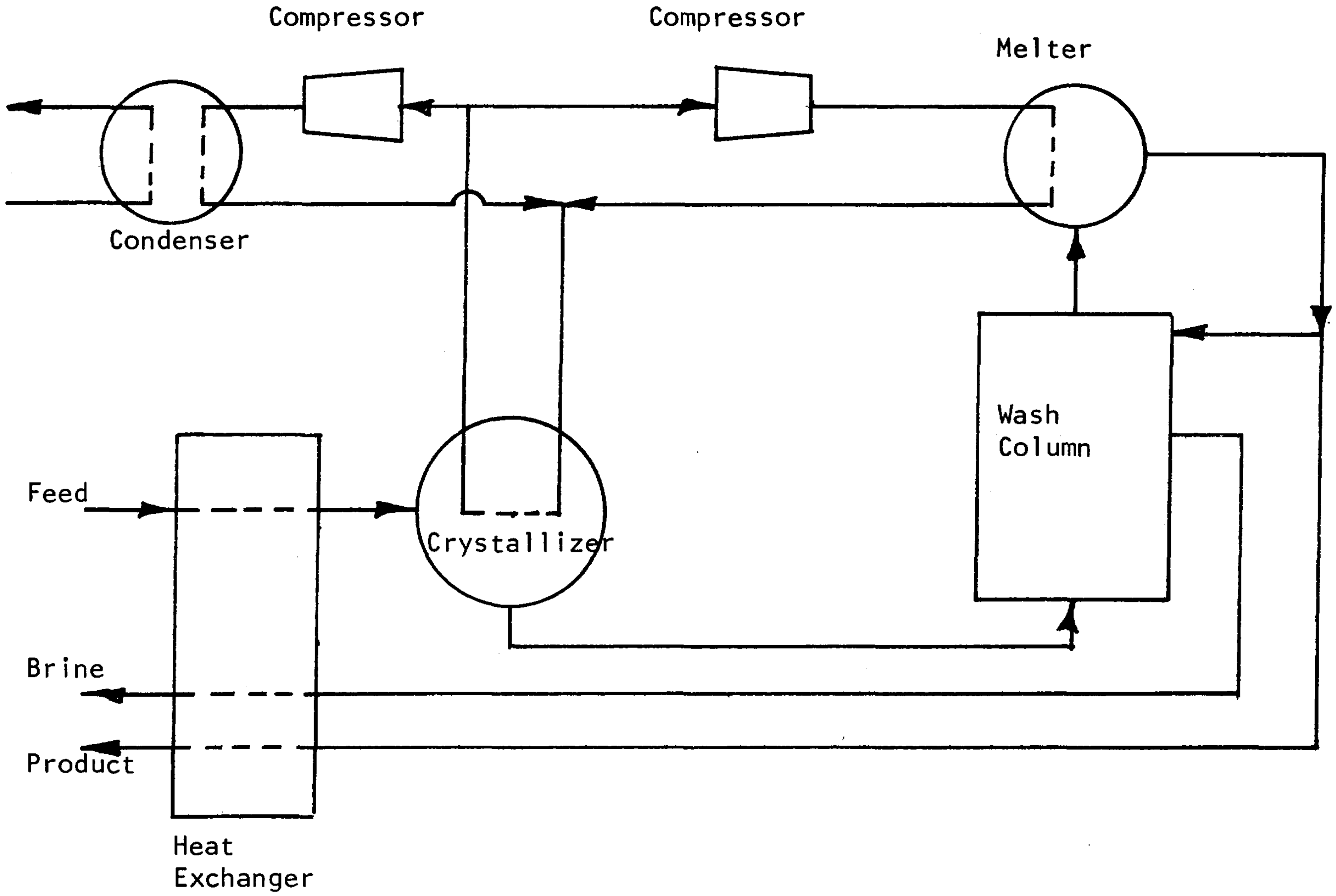

FIGURE 5
INDIRECT FREEZING SCHEMATIC

tube heat exchanger and melted. The melted ice along with the brine extracted in the wash column are removed from the system through the feed heat exchanger.

The shell sides of the crystallizer and the melter are the evaporator and condensor, respectively, for a state-of-the-art refrigeration system, whose function is to remove the heat of crystalization of ice and sensible heat from the feed stream. A temperature difference between the refrigerant and feed from 2 to 5°C (4 to 10°F) is a reasonable compromise between driving force and heat transfer surface cost. The evaporated refrigerant is compressed and partially condensed in the melter where the heat of condensation supplies the heat of fusion to the melting ice. A portion of the refrigerant gas, with a heat content equal to the heat reject load, is further compressed and condensed by ambient cooling water before being returned to the crystallizer with the refrigerant liquid condensed in the melter.

Eutectic Freezing [10]

The Office of Water Research & Technology has recently contracted for construction of a $38M^3$/day (100 gpd) Eutectic Freezing pilot plant.

The Eutectic freezing process is the extension of any freezing process to the point where both ice and salt are crystallized from solution. The effluents are salt and potable water.

The potential applications of this technology are in concentrating and precipitating salts from industrial solutions, control of salt concentration in industrial streams where loss of water is unacceptable, such as cooling tower blowdown in arid areas and similar waste water recycling applications. There is also possible application to brackish water desalting where brine disposal is a problem or undesirable because of the loss of water. In this case, an eutectic freezing process in conjunction with another primary desalting process may be economically feasible.

The process functions on the principal that if sufficient ice is crystallized from a saline solution, the concentration will be increased to the point where the solution will be saturated with respect to the dissolved salts. They will then precipitate from solution as discrete phases if further ice is crystallized from the solution.

The feasibility of the process has been demonstrated with direct contact crystallization [10]. The salts in the brines investigated had a specific gravity greater than the ice and the brine. They are separable by devices such as cyclones or gravity settlers which utilize this difference.

The pilot plant as currently specified [11] will use a cyclone between the crystallizer and the wash column to separate the salt from the ice. The salt underflow is to be collected in a filter. Currently a direct contact crystallizer using R-114 as a refrigerant has been specified;

however, an indirect crystallizer or vacuum crystallizer will be considered.

SOLAR ENERGY AND FREEZING

Freezing processes are unique in relation to solar energy because the principal power consuming elements are refrigeration systems. These refrigeration systems can be mechanical or absorption permitting the option of mechanical or thermal energy, or combinations thereof. Further, thermal absorption cycles allow a relatively wide range of operating temperatures through variations of types of absorbents and absorbent concentrations.

All freezing processes, as do all desalting processes, require moving of fluids, thus mechanical pump work is required; however, this represents a relatively small portion of the total system work, consequently the major portion of the energy requirements can be provided thermally.

A number of solar combinations are possible.

Those processes which require mechanical energy for the primary refrigeration cycle can employ either a high temperature Rankine cycle collector - turbine system to provide the mechanical/electrical energy and use the turbine reject heat to drive an absorption heat reject system; or a Photovoltaic array for the mechanical/electrical energy input and use the cell coolant heat for the absorption heat reject system.

A flat plate collector system or concentration gradient pond could provide the thermal energy for a primary absorption process.

Specific requirements for the processes discussed in this paper are summarized in Table (1).

The numbers given for thermal input are based on a sink or heat reject temperature of 32°C (90°F), a feed salinity of 3.5% and a brine salinity of 7%. These conditions are not necessarily optimal; they are, however, typical.

The temperatures for the thermal inputs shown in Table (1) are based on a 32°C (90°F) reject temperature and can be shifted up or down linearly for different sink temperatures. An interesting aspect of sink temperature is that the localities now being considered for solar desalting are also arid climates, consequently it is possible to consider evaporative cooling of cooling water to insure a minimum sink temperature and thus a minimum solar heat temperature.

Absorption Freezing

A very interesting variation on the vacuum freezing process that has some attractive features vis-a-vis solar desalting is shown in

TABLE 1
FREEZING PROCESS ENERGY REQUIREMENTS

PROCESS	Electrical KW-HR/M^3 (KW-HR/1000 gal)	Thermal KW-HR Thermal/M^3 (B/1000 gal)
AFVC	7.5 (28.5)	14.9 @ 74°C (193,000 @ 166°F)
VFEA	1.48 (5.6)	49.6 @ 82°C (640,000 @ 180°F)
INDIRECT FREEZING	5.8 (22.1)	13.4 @ 74°C (173,000 @ 166°F)
ABSORPTION FREEZING	3.2 (12)	100 @ 44°C (1,300,000 @ 112°F)

Conditions: Feed - 3.5% TDS
Concentration Ratio - 2
Sink Temperature - 32°C (90°F)

Figure (6).

The process is identical to VFEA and AFVC except that the ejector/absorption and mechanical/absorption cycles of the two processes are replaced with a simple absorber system.

The absorber is maintained at ambient, and consequently only a moderate temperature elevation above ambient is required to liberate the water of absorption in the generator. The heat reject vapor load is shown schematically and accounted for in Table 1 as being pumped by a mechanical refrigeration system as in the AFVC process described in the text. This could, however, be accounted for by a separate absorption system operating at 71°C (160°F) for a sink of 32°C (90°F).

The single stage absorption cycle has a considerably lower thermal efficiency in terms of water production per unit of heat but the temperature elevation required is only 12.2°C (22°F) above ambient. A simple open pond could provide the necessary heat.

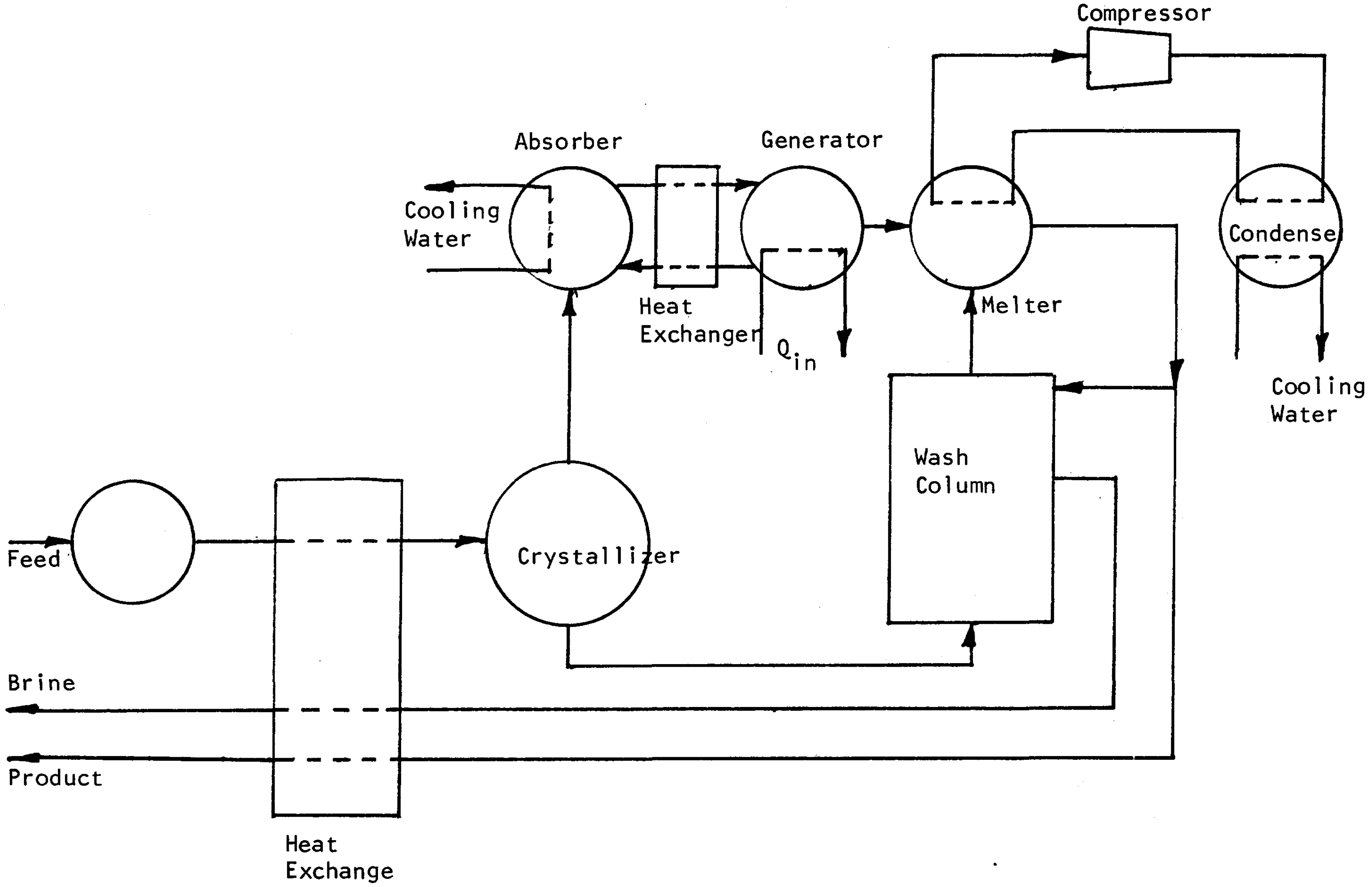

FIGURE 6
ABSORPTION FREEZING SCHEMATIC

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SOLAR DISTILLATION IN GREECE

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ABSTRACT

It is usual in arid locations to collect rain and store the rainwater in reservoirs. This water is then used as fresh water supply. Three different designs of large size plastic covered solar stills have been erected and tested at the islands of Symi and Aegina. In all of them the plastic cover could not withstand heavy rains and strong winds. A glass covered solar still design was developed at the Solar Experiment Station in Symi. This design has been applied in erecting commercial size solar stills at six greek islands and at Gwadar in Pakistan. An operation method of the solar stills to avoid formation of algae and scale consists of periodic complete evacuation of the basin and refill with fresh seawater. Rain collection may double the production of distillate. Preheating the feed water of solar stills increases considerably the output of product water.

INTRODUCTION

Water is a matter of life and death. Life was created in water and there is no life without water. Excessive evaporation of water from the soil gives a picture which is lamentable. Desalination of brackish and seawater is a means of modern technology to offer life to those regions, where lack of fresh water stops development. It is well known, that solar distillation exhibits a considerable economic advantage over other salt water distillation processes, because of cost-free energy and reduced operating costs. Another advantage of the process is its simplicity, as no moving parts are involved.

Since the intensity of solar radiation per unit of evaporating area is determined by nature, the output of a solar distillation unit in a given location can only be increased by increasing the evaporating surface. Capital costs and amortization depend more or less proportionally with the size of projected output and most of the studies in solar distillation are directed toward improving the efficiency and reducing the construction cost through appropriate design. Another parameter might be, the expected life of the plant. Amortization and plant life are the determining factors for the cost of the product water.

Although the advantage of cost-free energy is partly offset by increased amortization cost, distillation with solar energy remains the most favorable process for small-capacity water desalting at remote locations, where there is considerable solar radiation. Most solar distillation plants are being (or will be) erected in less developed countries or in areas where there are limited maintenance facilities.

As the incidence of solar radiation varies over the day, the time of the year, the cloudy weather and the geographic location, solar distillation can never be a steady state operation.

RAIN COLLECTION AS WATER SUPPLY

Before entering into details on solar distillation, I think it is advisable to spent a few words on the water supply situation in some remote areas, where lack of potable water does exist. As an example, in Gwadar of Pakistan, prior of building a large solar still, rainwater was collected in a basin, which was the drinking water supply of the nearby village. Sanitary water conditions were not the ideal. When the basin runs dry, the inhabitants were digging holes in the sand and collected very carefully the upper layers of stratified underground water as long as potable. When the water became salty,they just dig another hole sufficiently apart. Transportation of the water to the village is usually the duty of the women, who carry the jars of water on their heads. When men are involved in the transport of water, they most often use tins and donkeys to carry the water home.

In Mediterranian islands and in places with similar climate, it is usual to collect the rain water during the winter period, store it in underground cisterns and use the stored water during the rest of the year. Rainwater collection is made by shaping accordingly the flat roofs of the houses and draining the rain water to the cisterns. Conveying of the water to long distances is not required, as each house has ussually an individual cistern, but water stored for several months has not necessarily the best taste.

PLASTIC COVERED STILLS

In 1963 a wealthy american philanthropist decided to make a social experiment. The purpose was to investigate the effect of the supply of fresh water to a poor community, deprived from natural water resources. He finnanced the erection of a solar still at the small island of Symi in Dodecanese. He was familiar with the american design of inflated plastic covered solar stills and was anxious to see a prototype of this design operating on a commercial size. Church World Service was charged with the construction of the Symi solar still [1].

Figures 1 to 5 show some of the usual solar still designs. The first of them represents the original deep basin design by Lof and Battelle, as applied in the Daytona Beach Experiment Station. The design applied in Symi (Fig. 2) provided for a plastic cover supported by a small overpressure inside the still. Two gutters on each side of the cover collected the distillate. The water basin was shallow and the seawater layer reached quickly enough a temperature, at which distillation could start. Everything behaved well, until severe atmospheric conditions occured. Heavy rain deflated the stills and the plastic cover went down to lay on the seawater surface. Strong winds and hail damaged the plastic cover. One of the major characteristics of the inflated solar still is the required absolute tightness of the plastic cover to the structure of the

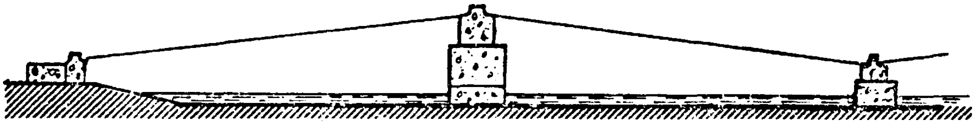

Fig. 1. Battelle-Löf design

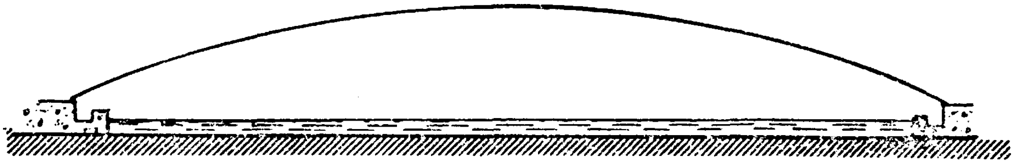

Fig. 2. Inflated plastic cover design

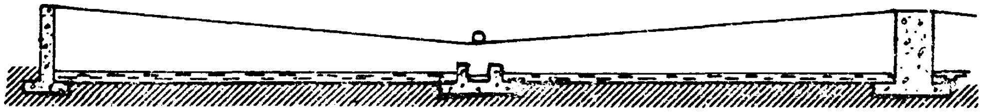

Fig. 3. V-shape plastic cover design

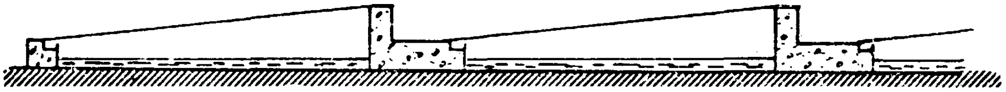

Fig. 4. Stretched plastic or inclined glass cover design

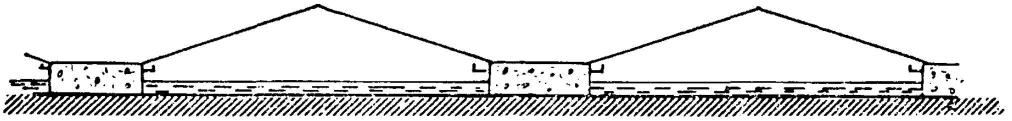

Fig. 5. C.S.I.R.O. - Australia design

basin. In view of the higher than atmospheric pressure, which prevail inside the still, any untight place leads to a loss of precious vapor. Due to various disadvantages revealed during operation the inflated plastic covered design has been abandoned.

A new design, the so-colled V-shape cover design (Fig. 3) was first tested at Symi by rebuilding part of the existing still and erecting two elements in V-shape. In this design the plastic cover is held firm by a pipe ranging along the axis of the still and forming the cover in the shape of a very flat V. One gutter along the axis and in the middle of the still is sufficient to collect the distillate, instead of two gutters at the edges, as required in the inflated still design. The new design was intended to keep the plastic cover tight and protect it from the action of rain and winds, as the cover was supposed to be held firm by the pipe and rain was expected to drain off.

Before having sufficient experience with the V-shape at Symi, a solar still entirely of the V-shape cover design was erected at Perdika in Aegina. During operation rain water did not drain completely and remained partly stagnant on the cover, increasing the load. Wind pushed the pipe to balance right and left on the cover. As a result severe damages occured again to the cover material of both V-shaped stilla, at Symi and Aegina. Consequently this design was also abandoned.

A third design was then applied, the so-called stretched plastic cover design (Fig. 4). The cover is inclined to the south, fixed at the two ends and stretched mechanically, without the burden of the pipe. Even this design could not withstand the atmospheric conditions. Hence, three various designs, all of them based on a plastic cover, were tested in commercial size and, unfortunately, all three failed to meet the expectations. This was the end of the plastic covered solar stills in Greece.

The plastic covered solar stills in Greece of various designs were erected at:

Symi in 1964 with inflated cover and a water surface of	2686 m^2
Aegina in 1965 with V-shape cover and a water surface of	1490 m^2
Symi in 1968 with stretched cover and a water surface of	2600 m^2
Aegina in 1968 with stretched cover and a water surface of	1486 m^2

SOLAR EXPERIMENT STATION AT SYMI

Using a generous grant of the same american philanthropist, who has finnanced the erection of the plastic covered solar stills in Greece, the Technical University of Athens has erected at the island of Symi a Solar Experiment Station [2].

Our purpose was to extend our research to all kinds of solar applications but we started with the development of a solar still. Although we had not omitted to test as well plastic covers, our design was mainly based on a glass cover. We have not used the conventional design with symmetric roof type cover, as for instance the Australian design (Fig. 5), but rather a shed-type cover with one large glass pane looking to south and one small glass pane looking to north. The large glass pane allows the solar radiation to be totally directed to the absorbing water surface, as the length of this glass pane is nearly equal to the width of the water basin. To avoid sagging of the large glass pane, an aluminum supporting structure was introduced. An inclination of no more than 10° was given to the large glass pane for easy drainage of the condensate.

We tested single-stills, twin-stills and triple-stills. Preference was given at that time to the twin-still, because of somewhat reduced investment cost. At the day of inauguration of the first inflated cover commercial solar still in Symi we were already in a position to present a model of our solar still design [3].

We have also tested at the Experiment Station several other solar applications, like water heating for household use, solar cooling, production of salt on a conveyor belt, etc. We also bought a house to be used as demostration of various solar applications, including space heating. At a time, when solar energy was the most neglected source of energy, these activities might had some interest. There is nothing exciting today about them and I will omly put emphasis on some experiments aiming to increase the yield of a solar still. We used first a black orlon fabric to act as black body. The absorption of solar radiation was at the begining excellent. However, with time, the fabrics acted also as excellent nuclei

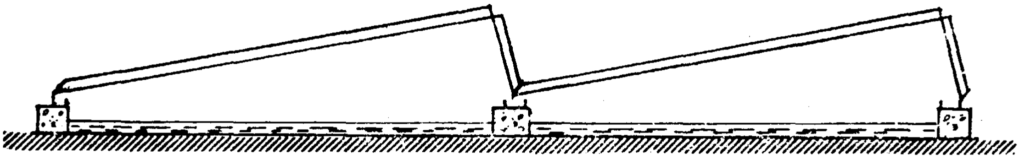

Fig. 6. The authors design

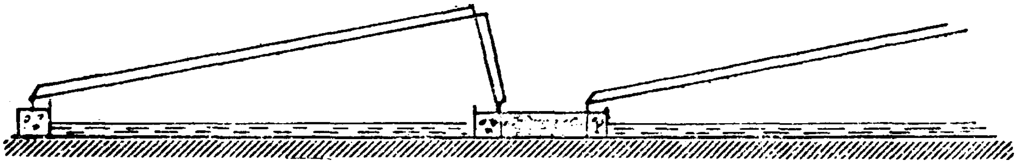

Fig. 7. The authors modified design

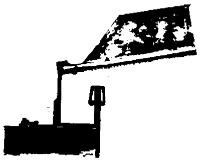

Fig. 8.

Fig. 9.

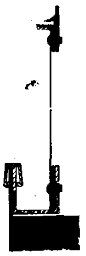

Fig. 10.

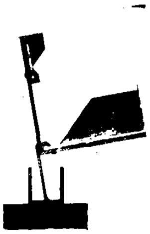

Fig. 11.

for the deposition of scale, especially because stagnant areas in the free flow of seawater were formed underneath the clothing. We therefore abandoned the use of the black fabric.

We then introduced at the back side of the still, inside the glass pane facing north, a serpentine pipe through which the feed water was circulated before entering the still. Our intention was to preheat the feed water and at the same time to create an additional condensing surface. The results were quite satisfactory. Nevertheless, as the heat input to the still by solar radiation remained the same, the total heat balance was not affected and we could only register a displacement of the bulk of the condensation from the large glass pane, as usual, to the condenser pipes. The increase in total yield was not large enough to justify the additional investment cost. To increase the yield in an impressive way, the total heat imput into the still must also be increased when using additional condensing surfaces.

SOLAR STILL DESIGN AND CONSTRUCTION

You might be interested to know more about our solar still design and construction, as well as about its mode of operation. And first on the design. Fig. 6 shows on top a cross section of a twin solar still of our design, as developed at the Solar Experiment Station in Symi. The single basin still is of quite similar in design (Fig. 7). Underneath some construction details. The first shows the distillate collecting gutter to the south side of the still (Fig. 8), connected with the end of the T-shaped aluminum support of the large glass pane. The lining of the basin, usually of butyl rubber sheeting, is air-tight fixed to the distillate collecting gutter. The glass pane is also air-tight fixed to the aluminum supporting structure, with an inclination, as already said, of no more than 10°. A similar construction is given to the distillate collecting gutter facing north (Fig. 9) and the two triangle shape sides, facing east and west (Fig. 10). In the twin basin still, an intermediate gutter is provided to connect the two basins (Fig. 11).

The sequence of construction steps in erecting this type of solar still in the field is as follows. At first a concrete frame is poured to form the basin of the still and the bottom is levelled with a slight inclination for drainage purposes. The basin frame formed in this way is then covered by a single butyl rubber sheeting, ordered exactly to the required dimensions. The butyl rubber sheeting, together with the edges of the concrete frame, forms the basin of the still and serves at the same time as black body. However, special attention should be given to kill any seeds in the soil. Believe it or not, plants have such a living force that they might penetrate the rubber, which has a thickness of about one millimeter, and enjoy life and sunshine.

After the butyl rubber sheeting is in place, it might be proceeded to the erection of the glass supporting structure. The aluminum structure is prefabricated in the workshop by welding in elements of 5 m lenght. These elements are then welded together on the site to form the solar still of the required length. In Patmos the length of the stills was 40 m, but this lenght proved to be excessive. In all other stills, built in Greece, the length was reduced to 15 or 20 m, with a width per basin of about 1.70 m. The last operation consists in placing the glass panes on the supporting structure and seal them air-tight. Adequate piping and proper drainage of the solar still are the items, which should be seriously considered to obtain satisfactory operation.

SOLAR STILL OPERATION

In Australia, in order to save labor, a serious effort was made to adjust the feed of the stills in such a way as to obtain a continuous still operation. In this case the quantity of water fed to the still must be equal to the quantities of distillate produced and brine rejected. A continuous operation is beforehand a very difficult task, because of the complete lack of steady state conditions. Hence, we have developed a mode of operation consisting in the periodic complete evacuation of the brine from the stills and recharge with fresh seawater.

The system requires a specific piping arrangement. A channel is digged between two groups of solar stills, in which the piping is accommodeted. The channel is covered watertight by concrete providing room for the valves. Each still is equiped with a feed water valve and an evacuation tube for drainage. The evacuation tube from inside the still leads to small plastic pipes, placed in the channel and turnable up and down. The length of these pipes are chosen in a way that, when they look up, their open end is at the same height with the desired water level inside the still basin. Any excess water overflows into the channel. In other words the water level inside the still is controlled by the length of these pipes, when they are looking up. When the pipes are turned down, their end is quite lower than the water basin in the still and the basins are entirely evacuated in the channel. To complete the description of the piping system, may I add that two seawater feed pumps are always provided for each solar distillation plant, one for normal operation and one as stand-by.

The mode of solar still operation that we have developed has the following cycles. The stills are filled with seawater to a depth of 2 to 3 centimeters and the level is controlled, as already mentioned, by the height of the open end of the drainage pipes. The still is then left alone for 3 to 4 days to normal operation. Depending on the evaporation rate and the amount of distillate produced, the brine remaining in the basin is concentrated to a certain degree. However, before the critical concentration of calcium carbonate deposition is reached, the drainage pipes are turned down and the solar still basins are evacuated. To avoid heat losses this operation is performed early in the morning. By this way scaling deposits are prevented and no time is left for algae to grow.

After the evacuation is completed the stills are filled again with fresh seawater. The system has behaved well in all commercial solar stills in Greece, provided that the instructions for proper operation were observed.

At this point I wish to emphasize, that a solar still must always be kept in operation, independently if the distillate is required or not, in other words the year around. Unfortunately some communities did not operate the solar stills during winter time, when fresh water from rain collection was available. If the still is not regularly evacuated, over-concentration of the seawater inside the basin is inevitable with the result of forming scaling deposits and algae.

SOLAR STILLS IN GREECE

Based on our design, the Greek Government has built six commercial solar stills in the islands. As a rule the community involved was requested to make available the required land at no charge. In such case the Government has financed the erection of the solar still and the Bank for Industrial Development, which is Government owned has proceeded to the construction. After the solar distillation plant was completed, it was turned over to the Communal Authority for operation and distribution of the produced fresh water to the local consumers. The latter is the weak point of the whole system.

Communal Authorities are not so permanent as a solar still, They might change with the next election and the attitude of the new-comers towards a solar still might also change. Cases have been recorded where the solar stills have been left unattended for long periods of time, especially in winter, with the unavoidable unpleasant results.

The solar stills, which have been built by the Government of Greece, were erected at the islands of:

Patmos in 1967 with an evaporating area of	8600 m^2	[4,5]
Kimolos in 1968 with an evaporating area of	2508 m^2	
Nisyros in 1969 with an evaporating area of	2005 m^2	[6]
Cephalonia in 1971 with an evaporating area of	2200 m^2	

Ithaca in 1971 with an evaporating area of 2400 m^2
Megisti in 1973 wirh an evaporating area of 2528 m^2

All of them are using seawater as a feed and no corrosion has been observed in the aluminum alloy used for the construction of the glass supporting structure. All have operated for years in a satisfactory way. An exception is Patmos, where, after more than a decade of operation the local authorities have recently developed apparently more abundant ways of water supply for this community.

In order to complete the overall picture some slides of various solar distillation plants in Greece will be presented. The first of them shows a general view of Patmos, which is one of the largest solar stills in the world. The next slides show as well Patmos under construction with the provisional workshop in the back, where the elements of the cover supporting structure were assembled, and the Patmos still shortly after the official inauguration in May 1967.

The other pictures show the Kimolos still, the Nisyros still and the Cephalonia solar still. In Nisyros a sufficiently large level terrain was not available and the still was arranged in three levels, instead of expensive leveling of the ground.

Our solar still design has also been used by the Pakistan Atomic Energy Commission to erect at Gwadar two solar distillation plants. The first, a preliminary one, was rather a prototype and covered an area of 306 m^2. The local contractor made first a mistake orienting the large glass pane to the north, instead to the south. This mistake was corrected but gave for some time a unique opportunity to compare the productivity of the still versus orientation. Hence, the wrongly oriented part of the still exhibited a lower productivity by about 30 %. The final solar distillation plant at Gwadar covered an area of 9072 m^2 and is, together with Patmos, one of the largest solar distillation plants in the world [7]. Unfortunately no picture is available.

RAIN COLLECTION

There are two more items which I wish to briefly discuss before closing my talk. The first is rain collection in a solar distillation plant. Glass covers and concrete paved path-ways between the stills are excellent rain catchment areas. Usually the period of the rains does not coincide with the time of the year at which intense solar radiation prevails. Therefore these two water reclaiming methods complement each other. An extensive report on the importance of rain collection in a solar distillation plant has been published by us some years ago [3].

Accordingly the average daily output of distillate over the year reached less than 3 liters per square meter per day and the addition of collected rainwater increased this amount to nearly 5 liters per square meter per day, under the prevailing at that year rainfall. This means that distillate plus rain brings the average daily productivity over the year of the solar still quite close to the highest distillate output, as recorded in the hottest summer months.

When the rainfall is sufficiently high, it might therfore be concluded that a solar still is an excellent rain collecting area, in which incidently some distilled water may be produced.

The second item, which I wish to briefly discuss, is the effect of preheating the feed water to the solar still. Solar collectors have a higher efficiency than solar stills. It appears therefore beforehand of interest to preheat the feedwater by means of solar collectors and proceed to its distillation in a separate device. Combination of solar heated feedwater with conventional distillation or vapor compression has been proposed on several occasions, but there is no practical experience of this process as far as I know. Our experiments were directed rather in combining heating of the feedwater by means of solar collectors with a conventional greenhouse solar still, under continuous water circulation.

Preheating of the feedwater to a solar still rises quicker than usual the water temperature in the basin and hence increases the rate of distillate output. The still was operated both as shallow basin and as deep basin. Whereas the major output in the shallow basin occurs during the day hours, in the deep basin the operation is prolonged during the night hours. However, the total output of the two systems remains more or less equal, when the maximum temperature of the basin is in both cases about the same under stagnant conditions. The nocturnal output of the deep basin still, allthough several times that of the shallow basin, is offset by the reduced output during the day.

Our experimental device consisted of a greenhouse type solar still, as applied in the greek islands, with a water evaporating surface of 9.85 m^2 assisted by a radiation absorbing area of solar collectors of 5.4 m^2. A typical run under stagnant conditions, with no water circulation and no preheating of the feedwater, gave the following results:

Basin	Radiation		Max. temp.		Min. temp.		Output		
	$\frac{W}{m^2}$	$\frac{Btu}{ft^2.day}$	°C	°F	°C	°F		lit./day	
shallow	7402	2346	65	149	20	68	day	8-20h	41.36
							night	20- 8h	5.04
							total		46.40
deep	6769	2146	52	125	29	84	day	8-20h	15.16
							night	20- 8h	23.37
							total		38.53

Combining preheating in the solar collectors and continuous circulation of feed water, the results obtained were:

Basin	Radiation		Max. temp.		Min. temp.		Output		
shallow	7664	2430	75	167	23	73	day	8-20h	66.50
							night	20- 8h	2.93
							total		69.43
deep	7708	2444	64	147	37	99	day	8-20h	41.52
							night	20- 8h	27.56
							total		69.08

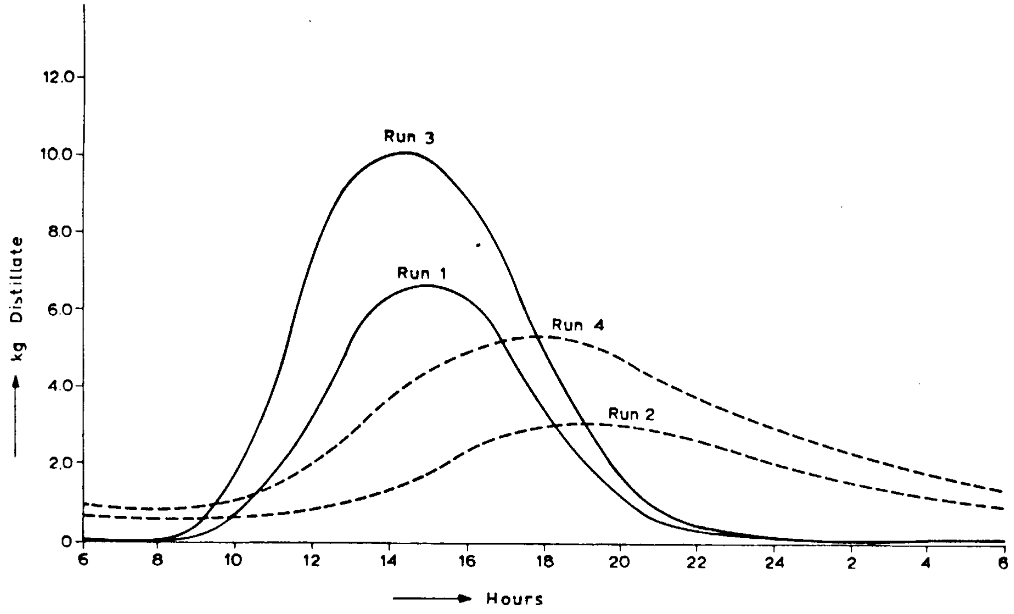

Fig. 12.

Variation of the circulation rate from 120 to 340 l/h showed no significant effect on the water temperature in the basin and hence on the distillation output. Fig. 12 shows typical distillate porduction on various conditions.

Run 1 refers to shallow basin with stagnant water at 2.5 cm depth.
Run 2 refers to deep basin with stagnant water at 15 cm depth.
Run 3 refers to shallow basin with 340 l/h circulating water at 2.5 cm depth and
Run 4 refers to deep basin with the same circulation at 15 cm depth.

CONCLUSIONS

The operation of commercial solar stills in the greek islands confirmed the expectations that a solar still is a reliable method to supply fresh water to small isolated communities with restricted technical facilities. The communities involved, usually poor, are released from the expense for energy, which reduces the operation cost to bearable limits. Hence the erection of a solar still is the only possible way of water supply, from the economic point of view, for these communities, provided that there is no charge for amortization. Taking in consideration the continuously rising oil prices a serious effort should be made to promote the erection of solar stills for the benefit of those communities, which have no other means to assure their potable water.

Criticism has been made about the healthiness of drinking distilled water. I do not dare to consider stabilization of distillate in such small communities, but I wish to remind that for centuries many communities had rain water as their main water supply. And the inhabitants have survived quite well.

AKNOWLEGDGMENT

It is gratefully aknowledged that the Wemyss Foundation, Wilmington, Delaware, financed the erection of the plastic covered solar stills at Symi and Aegina, as well as the solar Experiment Station at Symi.

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DESALINATION FOR MUNICIPAL SUPPLY

by

H.D. Altinbilek* and M. Akyurt**

ABSTRACT

Water requirements in municipalities as well as factors affecting demand are reviewed. Consumption and its relation with demand are discussed. Variations in demand are explained, using examples from Saudi Arabia. Various ways of meeting the water demand and the design of municipal systems for meeting the variations in demand are briefly reviewed.

Water demand and supply in Saudi Arabia is studied, including projections into the future. The role of desalinated water in meeting the present and future demand in the Kingdom is stressed. The plans of the Kingdom to build two solar plants, one for desalination of sea water and the other for brakish water, are reported.

It is observed in conclusion that solar desalination has a promising future in the Kingdom as well as in other areas with acute water shortages, provided that solar technology has matured sufficiently to meet the challenges involved.

GENERAL

Water is becoming a limiting factor on the economic growth in many parts of the world. Modern civilizations are far more dependent on water than past civilizations. As the population of the world continues to increase rapidly, modern standards of personal cleanliness call for vastly more water than was used a century ago.

It follows that water requirements of contemporary towns are very large. Complex municipal supply systems are required for providing cities with sufficient quantities of potable water. Water delivered by municipal works must be hygienically safe, esthetically attractive and palatable, and economically satisfactory for its intended uses. Sanitary standards set by the World Health Organization for

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water for human consumption are given in Table 1.

More often than not, the entire water demand of a community cannot be met from a single source. It becomes necessary hence to augment a main water source by additional sources such as rivers, lakes, ground water and desalination plants. The planning of a municipal water supply system in general involves the following |1|:

TABLE 1

WORLD HEALTH ORGANIZATION
STANDARDS FOR DRINKING
WATER

Substance	Maximum permissable concentration (ppm)
Carbon dioxide	20
Carbonates of sodium and potassium	150
Chlorides	250
Chlorine (free)	1.0
Copper	3.0
Detergents	1.0
Flourine	1.5
Iron	0.3
Lead	0.1
Magnesium	125
Nitrates	10
Phenols	0.001
Sulphates	250
Zinc	15
Total solids in suspension	500
Maximum NaCl	250

a. Estimation of the future population of the community.
b. Determination of the quantity, quality, areal and timewise distribution of water to be provided.
c. Location of reliable sources of water.
d. Designing of the necessary storage and transmission systems.
e. Designing of water treatment facilities to bring the physical, chemical and biological characteristics of water to desired levels.
f. Designing of the distribution system.
g. Planning of an organization which will operate and maintain the supply, treatment and distribution facilities.

WATER REQUIREMENTS

Uses of water supplied to municipalities may be broken down into four classes.

Domestic use is that water which is consumed in private residences for such functions as drinking, cooking, washing, bathing, laundering, house cleaning, heating, air conditioning, garden watering, sewage disposal, and car washing. The use of water in an average home in Akron, Ohio in 1970 is given in Table 2 |2|.

Commercial and industrial use is that water which is used by commercial and industrial establishments such as hotels, laundries, canneries, manufacturing plants, etc.

TABLE 2

WATER USE IN AN AVERAGE HOME
IN AKRON, OHIO |2|

Function	Share of Total (%)
Washing	1
Drinking and food preparation	5
Bathing	37
Lawn watering	3
Toilet flushing	41
Dish washing	6
Household cleaning	3
Clothes washing	4

Public use includes water required for parks, mosques, public buildings, schools, public toilets, public pools and fountains, and for street washing and fire protection.

Loss and waste represent all unaccounted-for water through leaks, breaks at pipe joints and in house service pipes as well as through unauthorized connections.

Water demand, expressed commonly as liters (ℓ) or gallons (gal) per capita per day, varies much and is influenced by many factors. Average values of water demand for the United States is given in Table 3 |3|.

The total sum of water demands is termed the "daily average demand" per capita, which is equal to the total annual supply divided by 365 times the population served. The daily average demand varies widely all over the world. While for cities in North America it is generally taken to be 568 ℓ/cap/day (150 gal/cap/day) as in Table 3, its

TABLE 3

WATER DEMAND IN THE UNITED STATES |3|

Use	Demand-ℓ/cap/day	(gal/cap/day)	% of Total
Domestic use	277	(60)	40.0
Commercial use	76	(20)	13.4
Industrial use	189	(50)	33.2
Public use	38	(10)	6.7
Loss and waste	38	(10)	6.7
	568	150	100.0

accepted values per capita per day are 415 ℓ (110 gal) in Scotland |4|, 225 ℓ (60 gal) in Turkey |5| and 245 ℓ (65 gal) in Saudi Arabia |6|. In certain parts of the world where water has to be carried, the daily average demand has been observed to fall below 40 ℓ /cap/day (11 gal/cap/day). Some of the factors that influence the daily average water demand are listed below:

- Climate
- Socioeconomic characteristics of the city
- Size of the city
- Types of industrial and commercial activities
- Extent of sewerage
- Quality of the water
- Abundance of public and private water supplies
- Completeness of meterage
- Cost of water
- Pressures in the distribution system, and
- Management.

It is important to differentiate between "demand" and present "consumption". In areas where the supply is less than the demand, consumption is naturally below the actual demand for water. One would expect a discrete jump in consumption in such cases, following the addition of a new supply to the system. A sudden increase of 25% was measured in Bursa, Turkey (pop. 330 000 in 1975) when the supply was sufficiently increased |5|. In like manner a 50% jump in water use is anticipated in major towns in the Kingdom if the supply is abundant |6|.

A high degree of wastage can occur when water is supplied on an unmetered basis or is very cheap. This would result in an artificially high demand. Likewise, wastage through leaks can reach 40% or more of the total supply |4|.

VARIATIONS IN DEMAND

Water demand shows annual, monthly, daily and hourly variations. The hourly variations in demand affect the sizing of the distribution mains and the service reservoirs. The consumption during the hour of maximum use lies between 1.7 to 3.25 times the average hourly water demand |5|. The smaller the community, the more variable the demand. Industrial use, on the other hand, shows less variations.

Variations in daily demand throughout the year reflect seasonal and climatic changes. The ratio of maximum daily use to average daily demand varies usually between 1.2 and 2.0, with an average design value of 1.5 |7|.

Figure 1 illustrates the seasonal variations of the daily consumption in Jeddah, Riyadh and Medinah |6|. Per capita consumption is expected to be higher in Jeddah due to a hotter and humid climate, larger city size, availability of water and a higher standard of living. The demand in Medinah is significantly lower during the winter months. There is, however, a pronounced increase in demand during the Hajj season (pilgrimage to Mecca) occurring in the present case during January and February. It is interesting to note that this increase in demand in Medinah is accompanied by a corresponding drop in water use in Riyadh, caused by the migration of people and transfer of many governmental functions to Mecca and Medina during the pilgrimage. The peak in consumption of water in Mecca during pilgrimage is bounded due to limited availability.

MEETING THE VARIATIONS IN DEMAND

Water supply systems are designed to meet community requirements over a reasonable period of time. The seasonal, monthly, daily and hourly departures from the average demand naturally affect the design of the system. Municipal water systems generally comprise collection works, purification works, transmission works and distribution works. The capacity of individual system components is set by what is expected of them and by economics constraints.

Distribution systems must have sufficient capacity to meet maximum hourly demands expected at the end of the design period. Transmission works, on the other hand, are designed for the maximum daily demand. Desalination plants and lesser purification works, which can be located at either end of the transmission works, are also designed in general for the maximum daily rate.

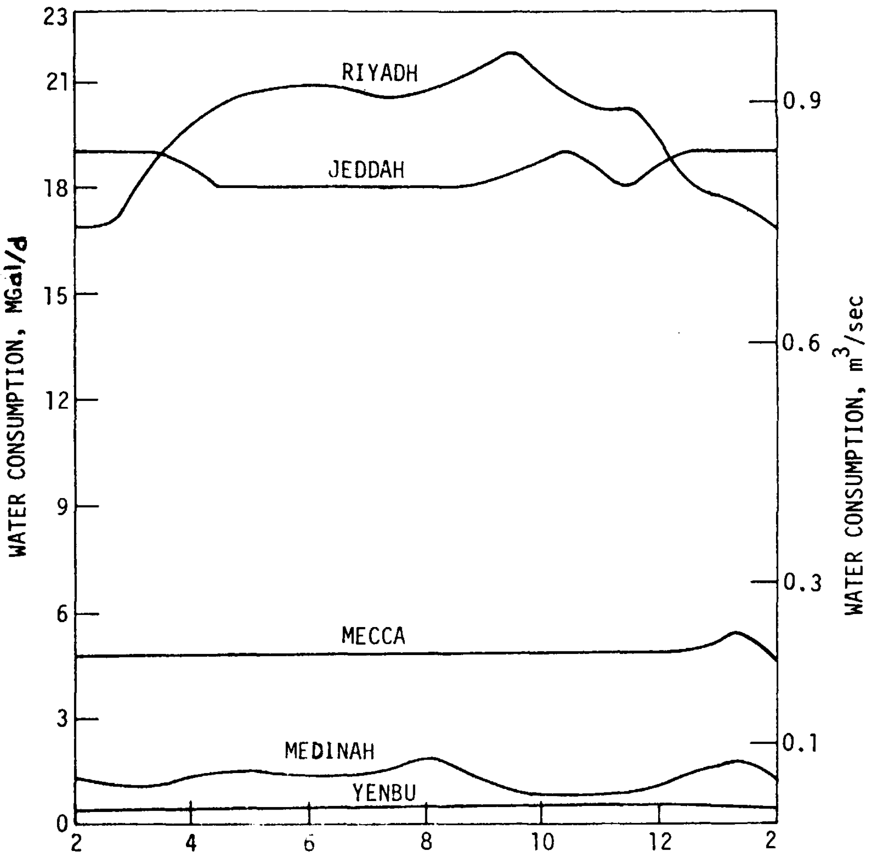

Fig. 1. WATER CONSUMPTION PATTERNS IN MAJOR SAUDI TOWNS IN 1972|6|.

Differences between the maximum hourly demand and the maximum daily demand are compensated by service reservoirs which are provided throughout the distribution works. Service reservoirs store enough water to

a) Compensate for hourly variations in water consumption;
b) Deliver water for fire fighting purposes, and,
c) Provide emergency storage for minor shut downs.

Annual and seasonal variations in water demand are met by providing the collection works and supply systems as well as the transmission and purification works with sufficient capacity. Here again final sizing is the result of economical considerations. The pipes and other hydraulic elements of municipal water systems demonstrate an "economy of scale" property such that the cost per unit capacity decreases as the capacity increases. For example, a pipe 2 m in diameter may be 2.5 times more expensive than a pipe one meter in diameter, but will carry approximately six times more water. Accordingly, pipe components are generally sized for demands expected at the end of the design period.

Desalination plants, on the other hand, do not seem to possess such a pronounced economy of scale property. For this reason, they are designed generally as staged constructions. The scheduling of a desalination plant in stages to augment the ground water supply is illustrated in Figure 2.

CURRENT PROGRAMS IN SAUDI ARABIA

Meeting the water requirements of municipalities in Saudi Arabia is a real challenge. Surface waters are not available in general; there are no rivers in the Kingdom. The average annual rainfall is less than 10 cm (4 inches). Only in the south west the rainfall may reach 30 cm (12 inches) a year. Over twenty dams have been built and are being built in this region for catching run-off |8|. Water which becomes available in this manner is used to supplement ground water reservoirs as well as to supply domestic and irrigation needs.

Underground aquifers are present in several parts of the country, especially in the east, ranging in depth from 75 to 250 m (250 to 850 ft) |6|. Numerous deep wells have been drilled throughout the Kingdom for utilizing ground water. Surveys indicate that proven reserves are about 337,500 million cubic meters (89,000 billion gallons) |9|. Eventual depletion of these reserves is expected since sufficient replenishment of ground water reserves is

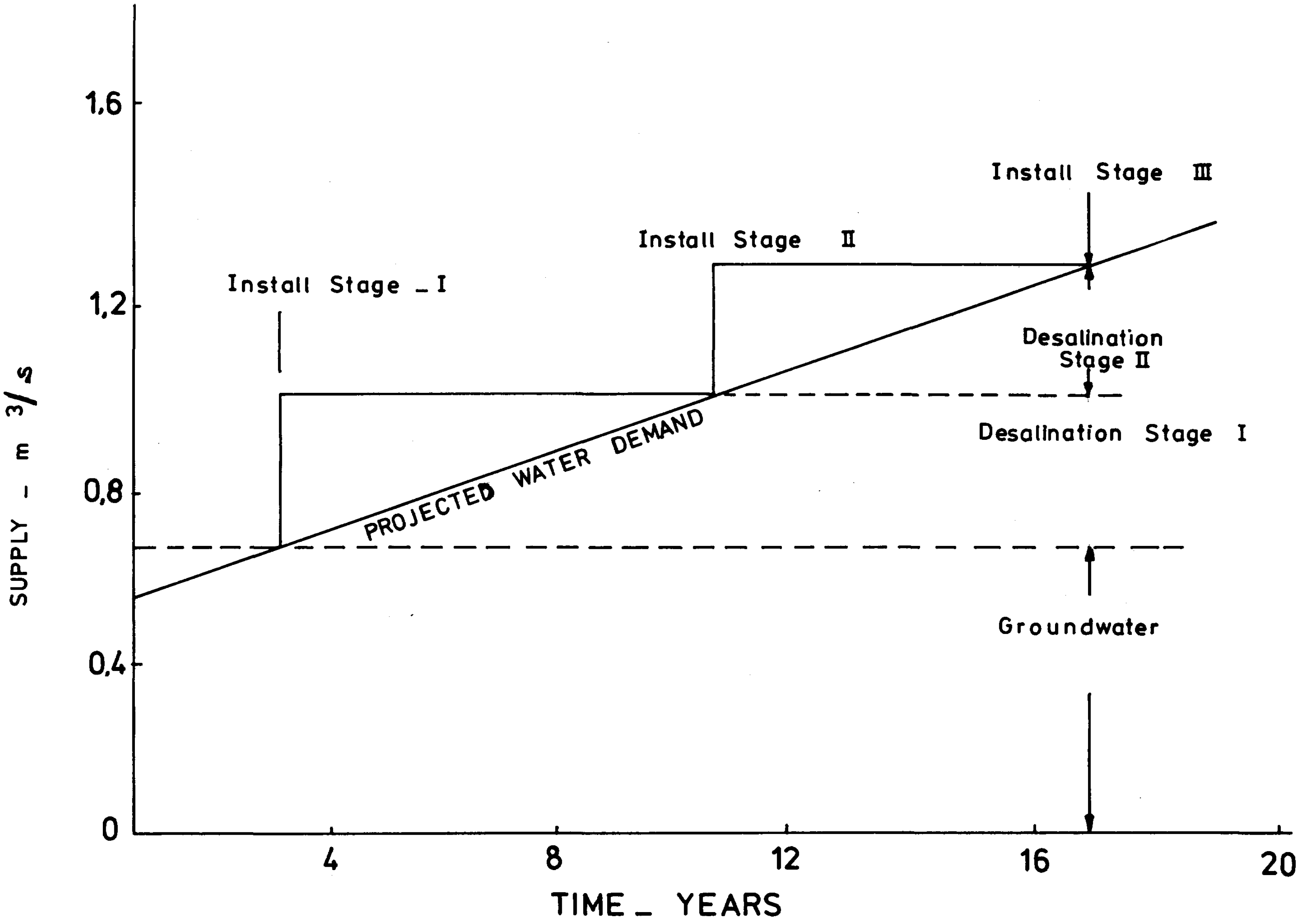

FIG. 2 _ STAGING OF DESALINATION FOR MEETING THE DEMAND unlikely.

The demand for potable water, on the other hand, has been rising at increasing rates in the Kingdom. In 1975 the annual water demand of the country was estimated as follows (in million cubic meters/in billion gallons) |8|.

Irrigation	1900	500
Oil well injection	390	100
Urban	170	45
Total	2460 mil. m^3	645 billion gallons

The projection of annual demand for water over the next twenty years is given in Table |9|. Figure 3 illustrates the growth of agricultural, municipal and total demand over the same period. As may be observed from this figure, sharp rises in the demand for water are anticipated, starting with the year 1985.

TABLE 4

WATER DEMAND IN SAUDI ARABIA (1980-2000)|9|

Year	Demand	
	$10^6 m^3$/year	billion gal/year
1980	4658	1230
1985	5340	1420
1990	5724	1512
2000	6523	1723

In the past ground water has met the major portion of the Kingdom's combined agricultural, industrial and urban demand, while surface water satisfied a smaller share. In addition to these conventional resources, increasing amounts of desalinated water has also been supplied in recent years. In Jeddah, for instance, desalination plants currently supply 160 000 m^3/day (42.3 million gallons/day) along with two ground water sources, Wadi Fatima (average yield 60 000 m^3/day) and Wadi Khulays (30 000 m^3/day).

It is apparent that demands in the Kingdom will grow rapidly with rising population, increased urbanization and industrialization, and the spreading of irrigated agriculture. Uncertainities about the availability of ground water, forecasts of rates of growth in urban water consumption of over 10% annually, awareness of waste and misuse of water resources, and the knowledge that

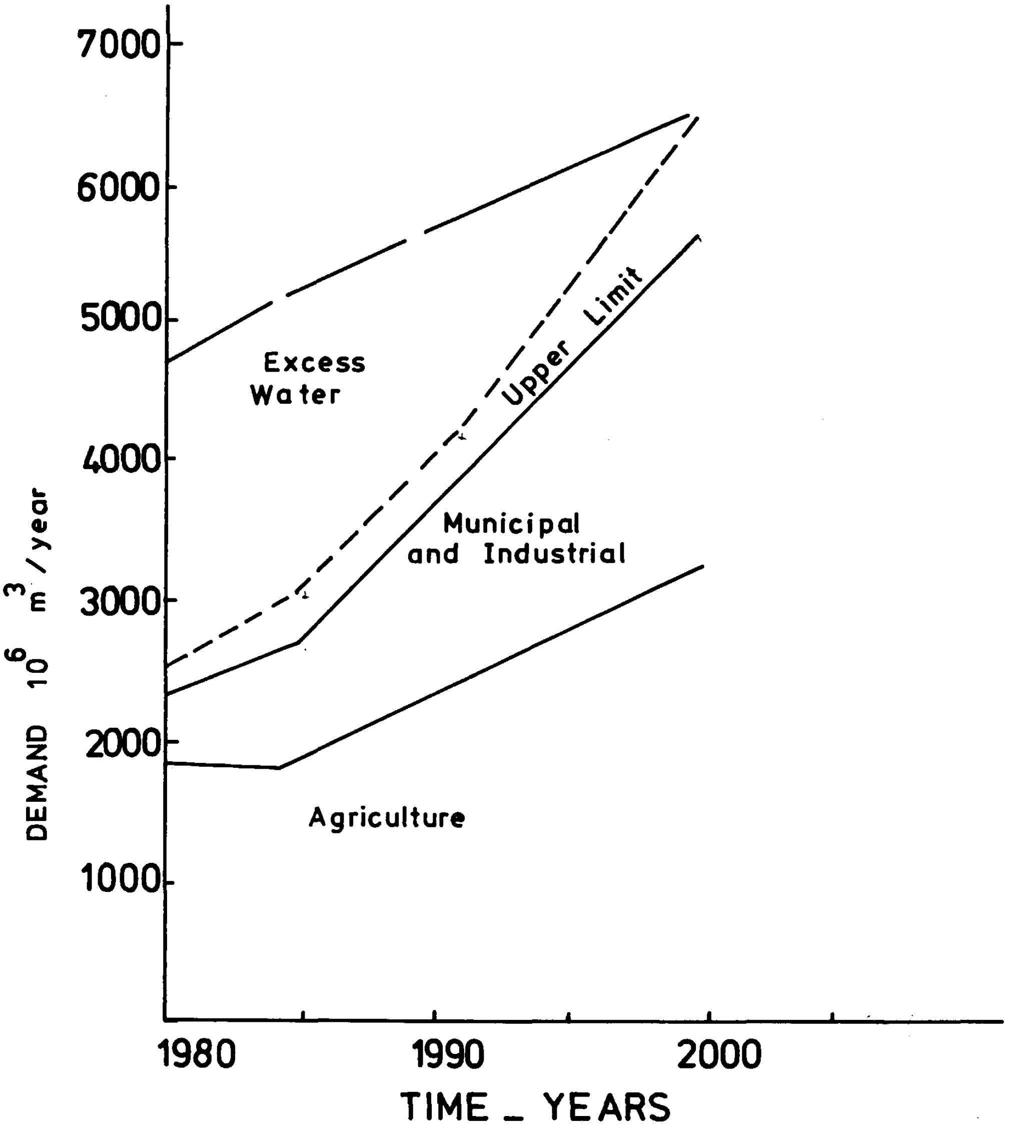

FIG. 3 _ GROWTH OF DEMAND FOR WATER (1980 _ 2000) [9] development of industry, mining and agriculture is dependent on abundant and assured supplies of water have led to the adoption of the following national policies |8|

- Continue to develop ground water resources to meet immediate urban, industrial and agricultural demands in locations distant from the sea coast, and,
- Accelerate development of the supply of desalinated sea water to meet urban and industrial demand in locations at or near the sea coast, and plan to use desalinated water to meet demands in selected locations away from the coast.

Thus present long range plans include programs for having desalinated water replace natural resources as the main water supply for municipal and industrial use, so that natural resources can be used for irrigation |6|.

Water supply forecasts have been made in the light of these considerations such that desalted water assumes considerable importance by the year 1985. Figures 4|9| and 5|6| show, respectively, the projections on the annual supply of water and the increase in capacity of desalination. It is clear from Figure 4 that in the future as in the past ground water will provide the bulk of the water requirements of the Kingdom. Desalted water will, however, play an increasing role in the near future, and it will become a very significant source of water soon. Likewise, moderate amounts of recycled water will be used in non-municipal applications after 1985.

Desalination plants in the country (Figure 6) utilize the multistage flash process, with the exception of several experimental units. Most plants use heavy oil for fuel. Desalination plants located along the east coast, however, rely on natural gas |6|.

As a result of the genuine interest in and the emphasis laid on research on solar energy applications, the Kingdom is planning to build during the current Third Five Year Development Plan two desalination pilot plants running on solar energy |9|. One of these solar plants will desalt sea water and the other brakish water, and they will each have a daily desalination capacity of 6,000 m^3 (1.6 million gal).

CONCLUDING REMARKS

Although ground water will continue to be the dominating factor in meeting this country's water demands for some

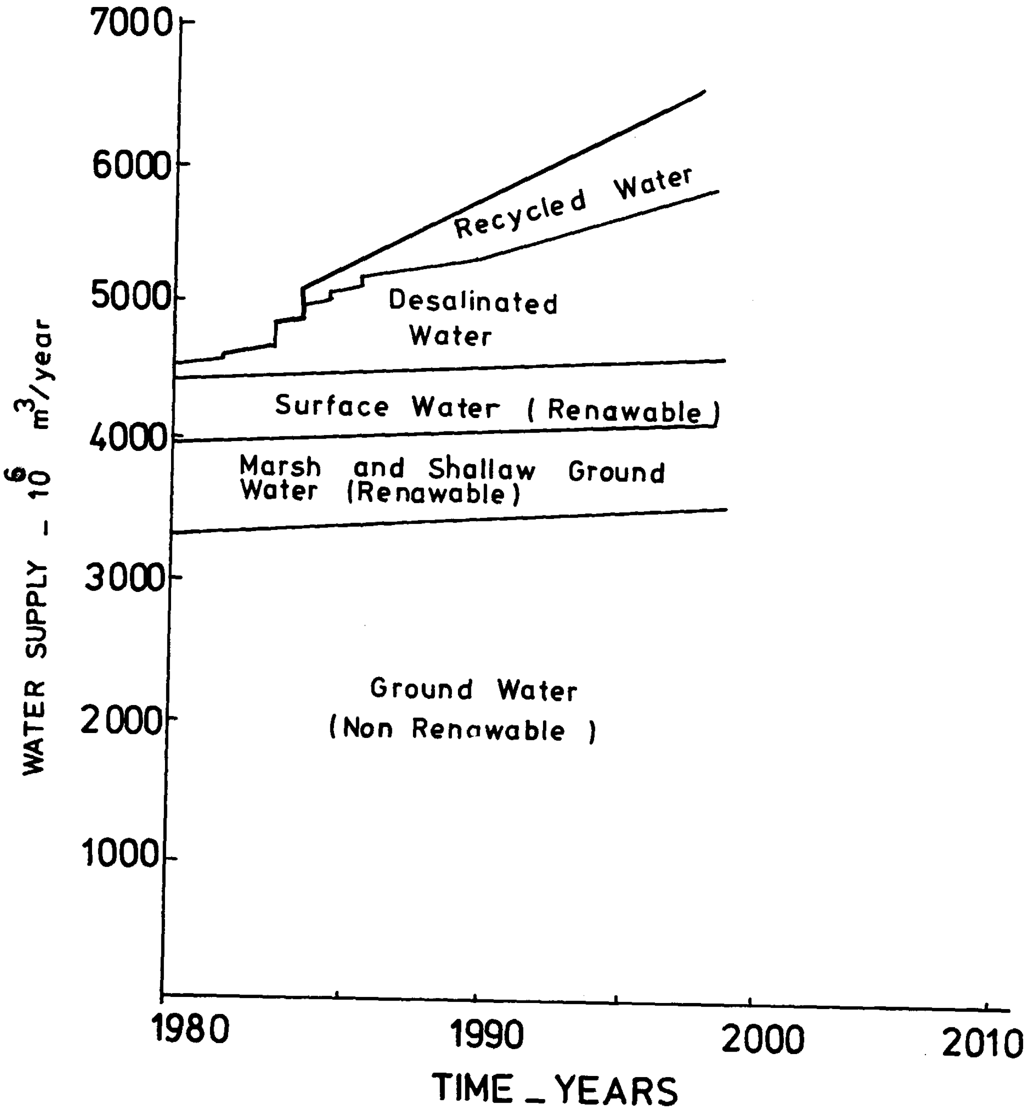

FIG. 4 _ ANNUAL SUPPLY OF WATER (1980 _ 2000) [9]

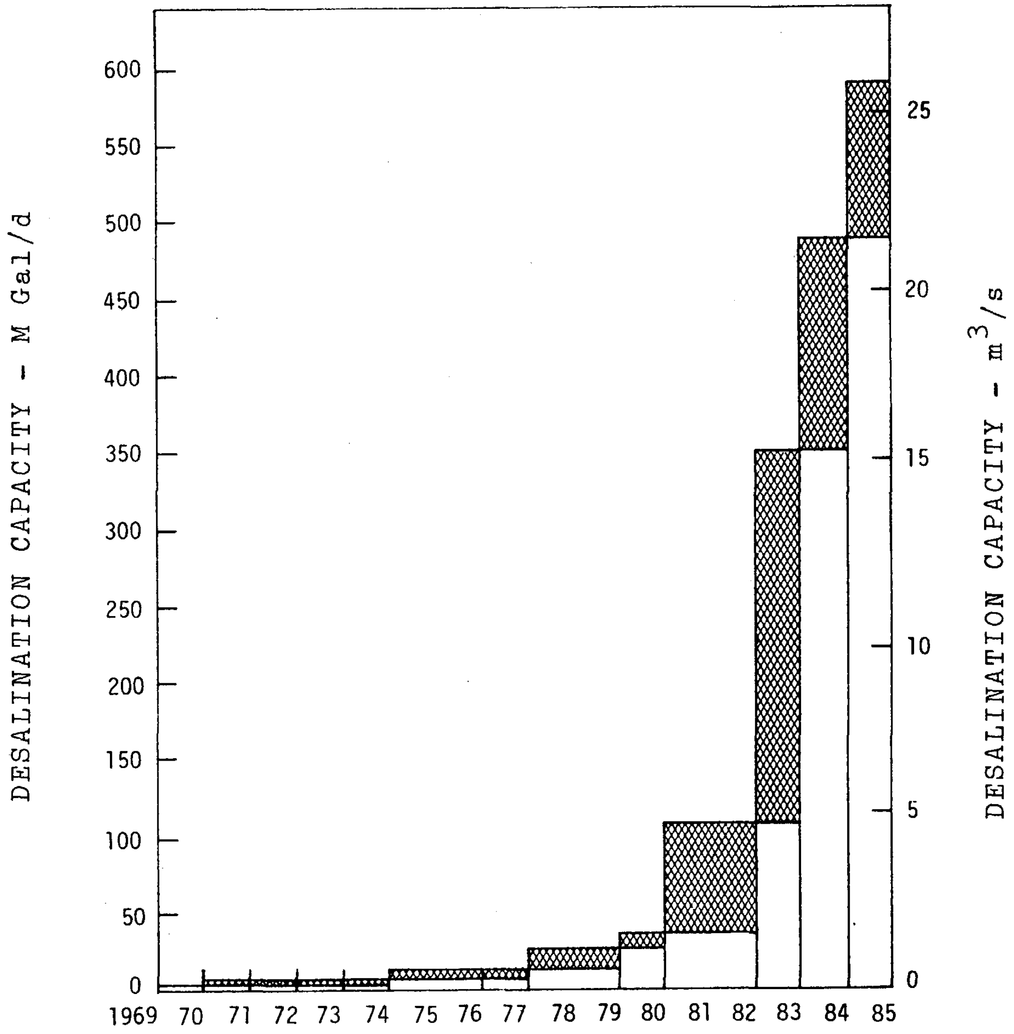

FIG. 5. VARIATION OF DESALINATION CAPACITY WITH TIME |6 |.

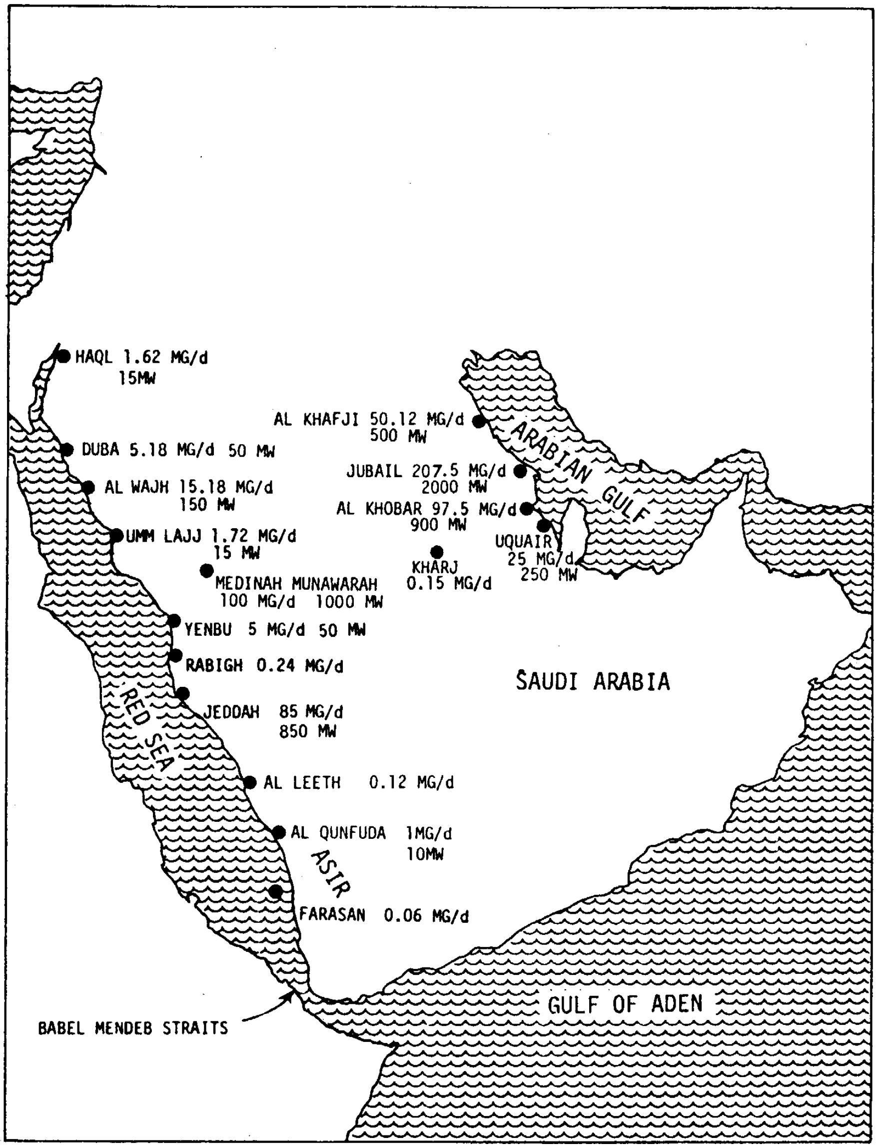

Fig. 6. PROJECTED WATER DESALINATION IN SAUDI ARABIA|6|.

time to come, desalinated water is and will continue to assume a very significant role. By the year 1985 the annual supply of desalinated water is to be equivalent to 68% of the Kingdom's total supply of water from renewable sources. By the turn of the century this figure will reach 114%

In the face of dwindling fossil and ground water reserves, and increasing fuel costs it is almost certain that solar energy will have a significant role to play in the years to come in supplying desalinated water in this country as well as in others facing the same water shortage problem. The pace of acceptance and utilization of solar desalination will be set, to a large extent, by the maturity of available solar technology,

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DESALINATION FOR INDUSTRIAL USES

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ABSTRACT

A previous work [1] predicted a massive (29.6 billion gallons per day) market for desalination (desalting) in the United States by the year 2000. Most of this market (85 percent) is expected to result from industry's compliance with the Federal Water Pollution Control Act of 1972 (FWPCA). The broad intent of the FWPCA is to achieve fishable and swimmable waters by 1983, and to achieve zero discharge of pollutants by 1985. The obvious implications for industry in that they must look for every possible way to reuse (recycle) the water within their own facility. Desalting will be required for a portion of this recycled water in order to make it suitable for its intended use. What is finally to be discharged must also be cleaned up. This cleanup will frequently including desalting. How these two activities, desalting of some recycle streams and some return flow streams, contributed to the estimated total desalting market is discussed. Industries which are high water users are listed. Specific examples of present and proposed use of desalting to solve these problems are provided. Energy sources for these processes are also discussed.

INTRODUCTION

In any developing technology, it is necessary to determine the current state-of-the-art before deciding the future course of action. The Office of Water Research and Technology (OWRT) sponsored state-of-the-art studies for the three desalting processes: distillation, freezing, and membrane processes. Each of these studies provides valuable technical information and has considerable background and history of the process with which it is concerned. Each report is based on a specialist's point of view in that area. As a result, they contain differences in cost bases and assumptions. Also, interactions among these technologies were not considered by the individual investigators. An effort was therefore needed to adjust the market and cost data, presented by each report, to a common basis, compare the

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present and the future developments of each technology, consider the interactions among them, and make recommendations for the future development of these three desalting technologies.

OWRT awarded a contract to Fluor Engineers and Constructors, Inc. to carry out this task. The objective of this project was to develop recommendations regarding the research and development effort that should be expended in each of the major desalting techniques - distillation, membranes, and freezing. These research and development recommendations were to be a total program required to bring each process to a "fully developed" state, regardless of who carries it out and whether it is carried out nationally or internationally either by governments, or private industry. The recommendations were to be based on the estimated future needs for water from desalting, the fully developed potential for each technique, and the time and effort required to reach full development. The technical approach to perform this work was carried out in six steps: review and adjust market data, review and adjust process economics, estimate "fully developed" economics of each process, estimate cost and time required to reach full development of each process, consider economic interactions among the processes and make the recommendations for development work.

We conclude that full development of the processes could result in savings on water costs of one-third to three-quarters (depending on the process, water type, etc.) of current costs - not too surprising a result. But, the results of the desalting market estimate contained some startling surprises. First of all, the market was so huge that it caused us to re-examine our assumptions and methods more than once. Admittedly, our methods were very broad brush, but refined figures were not necessary to our purpose. All that we needed to know was that the potential market was large enough so that the estimated savings in costs would justify the proposed research and development budget. As it turns out, a market of 10 percent (or two percent for that matter) of the estimated size would be large enough to satisfy that requirement. The greatest surprise was that 85 percent of the expected market would result from industry's compliance with the Federal Water Pollution Control Act of 1972 (FWPCA).

Concurrently, we, and many others, had been doing some work in applying desalting to these problems. This paper provides a discussion of:

- The nature of the problems
- The potential size of the market
- The industries involved
- Specific examples of desalting solutions
- Energy sources

THE NATURE OF THE PROBLEMS

Industry faces numerous problems with their waste streams, in order to comply with the FWPCA. Some of these problems can be solved, or partly solved, by the proper application of various desalting techniques. Knowledge of the objective of the FWPCA will aid in understanding these problems.

Objectives of the FWPCA

The broad intent of the FWPCA is to achieve fishable and swimmable waters by 1983, and to achieve zero discharge of pollutants by 1985. These objectives are to be met by application of Best Practical Technology (BPT), by July 1, 1977, and by application of Best Available Technology (BAT), by July 1, 1983. Reports indicate that the July 1, 1977 deadline has been met by about 90 percent of the industrial units. However, there is an ongoing debate on whether BAT is needed because BPT has produced better water quality than was expected. Obviously, any relaxation of the law could have a drastic impact on the projected desalting market.

Implications For Industry

Almost all industrial return flows are contaminated by pollutants, toxic and nontoxic. In order to comply with the law, a first obvious step is for each installation to reuse (recylce) as much water as possible within their own facility. This will at least make their waste stream smaller. It will also cut down on the amount of raw (make-up) water required. In some instances, the water to be reused will need to be desalted in order to be suitable for its intended use. Finally, regardless of how much the return flow may have been shrunk in size, it must comply with the law. These return flows would be concentrated by extensive recycling. Some of these streams will require desalting.

THE POTENTIAL SIZE OF THE MARKET

As stated above, one of the requirements of the state-of-the-art study was to make a rough estimate of future desalting needs so that we would have some measure of potential savings that could result from an aggressive research and development program. We examined seven different sectors for their potential desalting requirements:

- National water shortage
- Local water shortage
- Manufacturing return flow clean-up
- Steam electric reuse flow clean-up
- Steam electric reuse water clean-up
- Manufacturing reuse water clean-up

- Drinking water clean-up

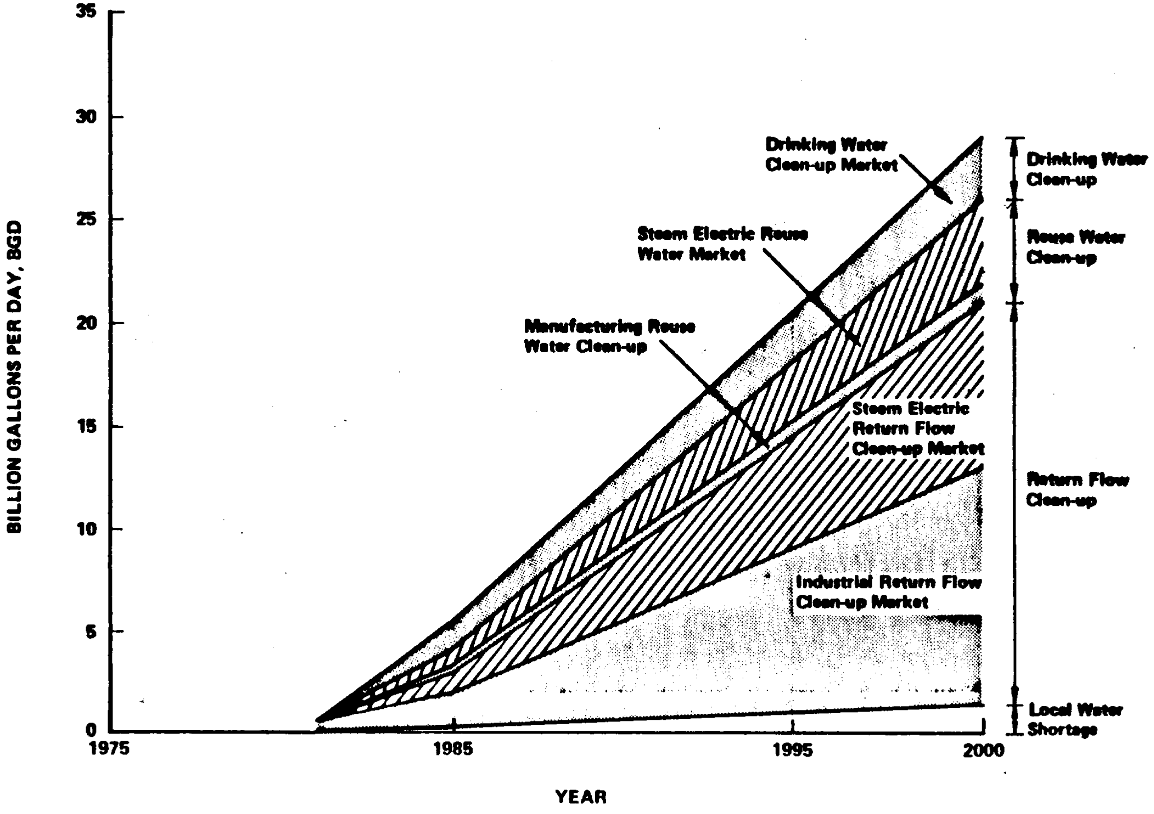
There appeared to be no impending shortage on a national basis. Unfortunately, surplus water in one place cannot always make up for shortages in another, because of distance. As a result, there is a relatively small (about five percent of the total) potential desalting market due to local water shortage. Some potential for desalting appeared in each of the other sectors (See Figure 1). The amount due to drinking water clean-up and local water shortage are of no interest in this paper except to show how small they are when compared to the industrial sector (The steam electric and manufacturing combined are regarded as the industrial sector in this paper). The industrial sector aggregates a whopping 85 percent of the potential market.

Return Flow Clean-up

From Manufacturing. The 1975 projections [2] of future water consumption (see Figure 2) indicate an increase from 105 billions of gallons per day (BGD) in 1975 to 125 BGD by the year 2000, while the water withdrawal is projected to decrease from 347 BGD to 292 BGD during the same period. These projections indicate that the water consumption in 1975 was expected to be 30 percent of the withdrawal and could be about 41 percent of the withdrawal in the year 2000. The return flow water, or the difference of withdrawal and consumption projections for the year, is estimated to be 167 BGD. The increasing water consumption, coupled with reduced return flow, would cause a higher concentration of solids and contaminants in the return flow which will have to be eliminated because of the FWPCA requirements of zero discharges by 1985.

These return flows are projected to be 14.1 BGD for the year 2000. We estimated that by the year 2000, 80 percent of the return flows could provide a potential desalting market of 11.3 BGD.

From Steam Electric Generation. The 1975 water withdrawal for steam generation was 92.6 BGD with 2.1 BGD of consumption, and a return flow of 90.5 BGD. The overall concentration factor of the return flow was only (92.6/90.5) 1.023, indicating the cooling to be mostly once through. By the year 2000, the water withdrawal is anticipated to decrease to 70 BGD, and the water consumption to increase to 9.1 BGD. The projected return flow would be 60.9 BGD. The overall average concentration factor of return flow water would be approximately (70.0/60.9) 1.14 indicating more water recycling. This would require clean-up by desalting in some cases, particularly in regions where high salinity problems already exist, or where the water supply is already short. Return flows from the South Atlantic Gulf, Texas Gulf, Rio Grande, Colorado, Great Basin, and Pacific Southwest regions are expected to shrink from 15.1 BGD in 1975 to 10.4 BGD by the year 2000, due to extensive recycling. If 80% of these streams are desalted the desalting technology market from this source would be 8.28 BGD.



POTENTIAL U.S. DESALTING TECHNOLOGY MARKET

FIGURE NO. 1

Reuse (Recycling) Clean-Up

This market estimate was based on the following principal assumptions:

- Total water reuse was taken as the difference between the 1968 and 1975 projections [2] of water requirements (see Figure 2).

- Of this total water reused, it was assumed that only one percent of the steam electric reuse, and two percent of the industrial reuse would be desalted.

The information presented above was for the market in the United States only. Undoubtedly, similar problems exist or will exist in other countries. In general, the proposed solutions should be applicable to these problems as well.

THE INDUSTRIES INVOLVED

In one of their recent publications [3] OWRT provided the following information:

"Industrial requirements and discharges are estimated for the following manufacturing and minerals Standard Industrial Classifications (SIC):

Manufacturing Industries

Primary Metals - SIC 33
Chemicals and Allied Products - SIC 28
Paper and Allied Products - SIC 26
Petroleum and Coal Products - SIC 29
Food and Kindred Products - SIC 20
Transportation Equipment - SIC 37
Textile Mills Products - SIC 22
Other Manufacturing

Minerals Industries

Metal Mining - SIC 10
Non-Metal Mining - SIC 14
Fuels
  Anthracite Mining - SIC 11
  Bituminous Coal and Lignite Mining - SIC 12
  Oil and Gas Extraction - SIC - 13

"Primary metals (SIC 33), chemicals (SIC 28) and paper (SIC 26) industries accounted for about 77 percent of the total fresh water requirements for manufacturing industries in 1975. The total fresh water requirement of these same three industry categories in 2000 is projected to be about 14 BGD or about 72 percent of the total industrial manufacturing requirement of 19.7 BGD. Most of the water use for manufacturing is in the Great Lakes, Ohio, Mid-Atlantic, South Atlantic-Gulf, Lower Mississippi and Texas-Gulf regions with the Great Lakes Region accounting for over 20 percent of all industrial use.

"Water use in manufacturing industries is primarily for three purposes: (1) cooling, (2) boiler feed, and (3) processing. About 70 percent of all industrial use is for cooling and this represents a potential for wastewater reuse and recycling. Boiler feed water and

U.S. WATER SUPPLY AND DEMAND

RETURN FLOW = WITHDRAWAL - CONSUMPTIONS

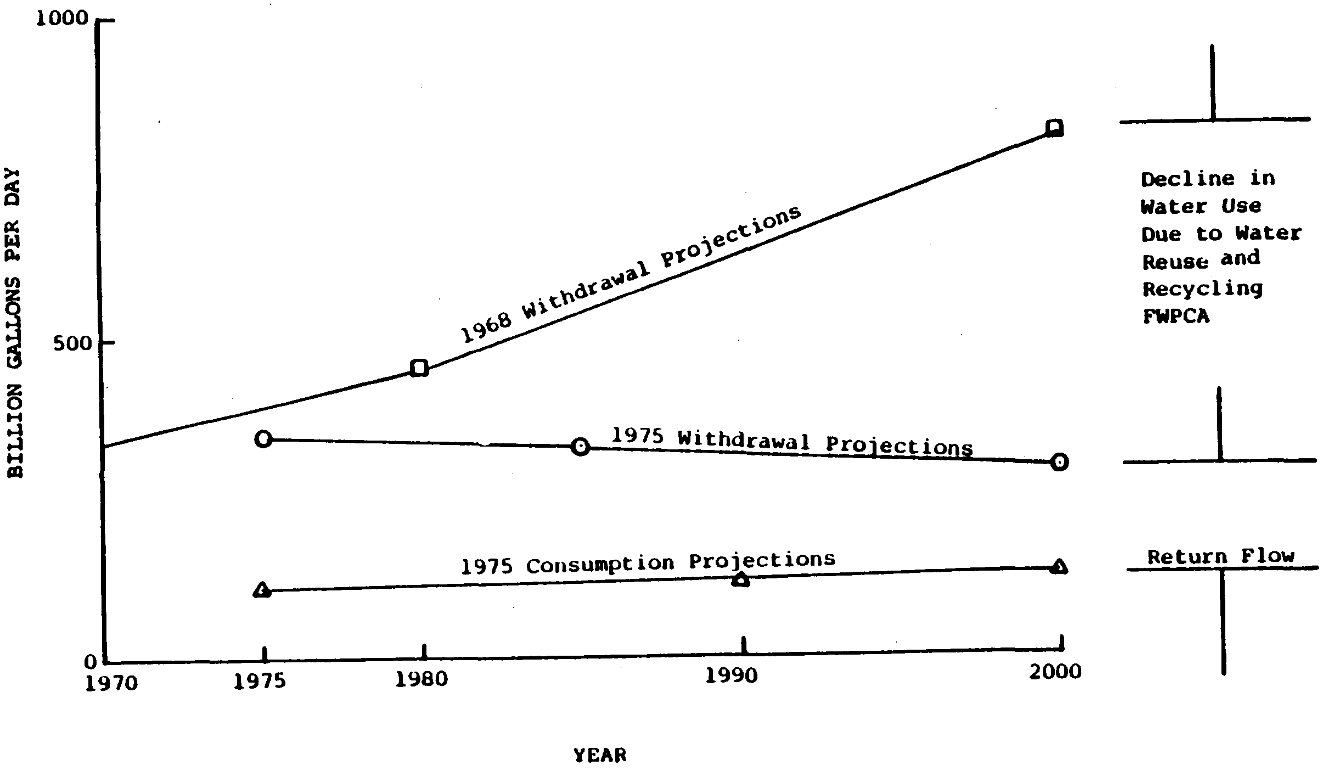


FIGURE NO. 2

process water in general require higher quality water and the potential for reuse of wastewater for these purposes is more limited.

"The mineral industries include metal mining (SIC 10), fuels (SIC 11, 12 and 13), and non-metal minerals (SIC 14). The largest water withdrawal is by the non-metal mining industry, but the largest consumer is the fuels industry with more than 60 percent of the total. The principal water using non-metal mineral industries are chemical and fertilizer mining (SIC 147), 51 percent of the total and sand and gravel (SIC 144), 30 percent of the total. Over 90 percent of the water withdrawn by non-metal minerals industries was for coal processing with very little used for cooling. Non-metals will continue to be the major water using segment of the minerals industries with an increase from 3.7 BGD in 1975 to 6.1 BGD in 2000. Reclaimed wastewater could be used for several purposes, including processing, in some minerals industries."

Interest in this area is confirmed by the fact that these industries are on the list of "high water-using industries" in the "OWRT Priorities for Fiscal Year 1981 Support - WATER REUSE RESEARCH AND DEVELOPMENT".

SPECIFIC EXAMPLES OF DESALTING SOLUTIONS

Early research and development efforts in desalting were focused on saline water conversion - fresh water from seawater or brackish water. Enactment of the FWPCA broadened that view to include the clean-up of waste streams either for reuse (recycling) or legal discharge. The problems are different, primarily because some of the pollutants in these waters interfere with some of the desalting processes, especially the membrane processes. However, while additional pretreatment, or a combination of processes may be required in some instances, the technologies of saline water conversion are generally transferable to waste stream clean-up. In the examples that follow, it will be seen how the technology was transferred to another discipline (waste stream clean-up).

Desalting Plant Brine Disposal Methods - Inland Locations

The Need for Low Cost Concentrators. The need to convert the effluent from desalting plants at inland locations to solids (ultimate disposal) has been recognized for some time. Return of the effluent to the source stream is almost always environmentally unacceptable, and probably illegal. Also, the technical problem of contaminating the source becomes significant to the desalting plant itself if a subsurface source is utilized. Deep well injection is only slightly better, and in the long run would also prove environmentally unacceptable. As a result, the effluent must be transported by some means (pumping, truck, etc.) to a suitable disposal site. Needless to say, the more concentrated the effluent is, the better, and in most cases, a solid would be preferred to a concentrated liquid. Higher concentrations also produce more by-product water (if the process used produces water. Some, like solar ponds, do not).

But the need for lowest possible cost concentrators is not limited to inland desalting plants. Concentration and water recovery from industrial process and waste streams is becoming increasingly important, and will soon become critical if the objectives of the FWPCA are to be met. Virtually every industry is faced with problems in this area.

The Office of Saline Water (OSW) awarded contracts to five equipment manufacturers to study the problem of concentrating desalting plant effluents to solids. Each proposed their best solution to the problem with an estimate of capital and water costs. The range of estimated costs was extremely wide. Four different types (representatives of about 70 percent of all inland brackish waters) of waters had been considered in the studies, and much of the differences could be attributed to the fact that some waters are more difficult to concentrate than others. But, even for the same water type, the range was wide. Part of the spread, of course, could be attributed to real difference in the processes, but unfortunately, these differences were magnified by different assumptions used by the manufacturers. Subsequently, Fluor Corporation was requested to put all of these studies on the same basis, so that the real differences could be analyzed. During this study [4], it became apparent that two basic flow sheets emerged, one for low boiling point rise (BPR) waters, and one for high BPR waters. Both flow sheets utilized multiple effect evaporators to perform the initial concentration. The final step was different in that crystallizers were used for low BPR waters, and fluidized bed dryers for high BPR waters. But more importantly, it became apparent that both processes were quite costly. A better process was needed.

The Yuma Desalting Plant - 91 Percent Recovery

Subsequently, OSW called upon us to prepare a budgeting cost estimate [5] for a 200 MGD, 91 percent recovery plant operating on the Wellton-Mohawk Drain (now known as the 100 MGD Yuma Desalting Plant). The method selected to achieve 91 percent recovery was to use ion-exchange for partial calcium removal between stages in a three-stage reverse osmosis plant. This method was chosen because it was believed to be technically feasible and could be designed and estimated in the four-weeks time available to prepare the estimate. As far as we know, no one associated with the project thought this method to be optimum or even desirable, but it did allow us to get a handle on the costs. The costs were excessively high, especially the incremental cost of the extra water produced above 70 percent recovery. After the project was completed and more time was available, we began to search for a better way to reach 91 percent recovery.

One of the best ideas to come out of the brine disposal studies (see above) was that of using seeding techniques as a means of preventing scaling. It occurred to us that a better approach to the problem of 91 percent recovery would be to use reverse osmosis to achieve the percent recovery allowed by the calcium sulfate, say 70 percent, and an evaporator with calcium sulfate seeding to control scaling at the desired total recovery. Steam turbines could be used to drive the high-pressure reverse osmosis feed pumps, and the exhaust steam could be used in the evaporator. A quick look at the economics of this combination showed great improvement over the reverse osmosis-ion exchange combination but the incremental cost of the water between 70 and 91 percent recovery was still too high to be attractive in that situation.

The RCC Process

By about this time Resources Conservation Company (RCC) had developed an advanced process which not only took advantage of seeding techniques to control scaling, but vapor compression to save energy. Since the RCC process was developed to concentrate waste streams, rather than produce fresh water from saline water (fresh water is a by-product), this is an excellent example of two concepts that were originally developed for saline water conversion being applied to water reuse by industry. RCC summarized the needs of the electric utility industry in their report [6] entitled "Brine Concentrators Providing Zero Liquid Discharge". They state that "Blowdowns and wastes may include cooling tower blowdown, demineralizer waste, boiler blowdown, boiler cleaning waste, scrubber waste and ash sluice waters".

Freezing Processes - The Low Cost Concentrators of the Future

At that time, many of those interested in desalting held the view that freezing processes had either been abandoned or at least put on the back burner because of inability to wash the ice.

Then OSW announced a Freezing Symposium. We were among those who were surprised and impressed by what we saw. Freezing processes were alive and well, and making good progress. They could wash the ice after all. As a matter of fact, we learned later that an impressive list of demonstration plants had successfully washed ice on a continuous basis. At this point it was apparent that the energy savings potential of freezing processes might actually be realized. Unfortunately, not everyone interested in desalting attended the meeting or read the paper. As a result, many people today still hold the opinion that we did then.

Those taking the tour of AVCO Laboratories had an opportunity to see many outstanding displays. The one which really triggered our interest was the eutectic freezing process bench-scale demonstration unit. Raw seawater was fed in and product water and wet salt came out; no pretreatment was required. Such simplicity is highly appealing. After that symposium, we realized that the combination for high concentration should be reverse osmosis-freezing rather than reverse osmosis-evaporation. For a feed stream like that of the Yuma Desalting Plant, 96 percent overall recovery is equivalent to only about two times concentrated seawater, quite a normal concentration for a freezing process.

The RCC Process represented a significant advance. However, it appeared that the freezing process could have still more favorable economics, and lower energy requirements. A comparison of the flow sheets and operating conditions would lead one to expect lower capital cost for the eutectic freezing process. And, while vapor compression is a great energy saver when compared to other distillation processes, freezing should be still lower. These assumptions needed to be confirmed and quantified. We undertook such a study [7] for OWRT. The principal conclusion drawn from this work was that once they have been developed (this work is ongoing) and demonstrated, freezing processes will be superior concentrators. Also, they will become more attractive from both the energy and concentration cost viewpoints as the desired concentration of final effluent increases. This is particularly evident when the desired final effluent is a solid. The freezing process, when operated at its eutectic point (the so-called "eutectic freezing process"), is the only process which can produce fresh water and solid salt simultaneously.

The assumed feeds in this study were approximately the same as those in the "Brine Disposal Methods" study mentioned above, except that the original saline water plant from which they came was assumed to be a reverse osmosis plant operating at near calcium sulfate saturation, instead of a multistage flash plant. The study did not look at the overall picture, however, but only at the concentrators. A study which did look at the overall picture, and which was based on actual conditions at an electric utility follows.

Alternate Methods of Cooling Tower Blowdown Concentration

In response to a request by the Water and Waste Equipment Manufacturers Association we produced a paper [8] on alternate methods of cooling tower blowdown concentration in which we examined the comparative economics and energy consumption of the various methods.

Cases Considered. The particular water systems selected for study was the water system of the Huntington Station of Utah Power and Light Company, as reported [6] by RCC. Figure 3 is a simplified version of their water system diagram. In the original scheme, cooling tower blowdown was to be evaporated in a waste pond, which also was to receive waste waters from a mixed bed condensate polisher, and from a cold lime softener. Instead, a brine concentrator by RCC was installed to concentrate the cooling tower blowdown. The condensate from the evaporator is used for the cooling tower and demineralizer makeup. The waste from the RCC concentrator is dumped into the waste pond, the size of which was only 11 acres compared to 530 acres which would have been required by the original scheme.

Comparisons of the energy comsumption and costs were made for three cases. In Case I, the RCC unit is replaced by an eutectic freezing plant. The schematic of this arrangement is presented in Figure 4.

In the second scheme, described as Case II, the cooling tower blowdown is first cold lime and soda ash treated to remove calcium hardness so that higher recovery can be achieved in the reverse osmosis unit. Product water recovered from the reverse osmosis unit is used as makeup to the cooling tower. The reject stream of the reverse osmosis unit is fed to a eutectic freezing plant. The product recovered from

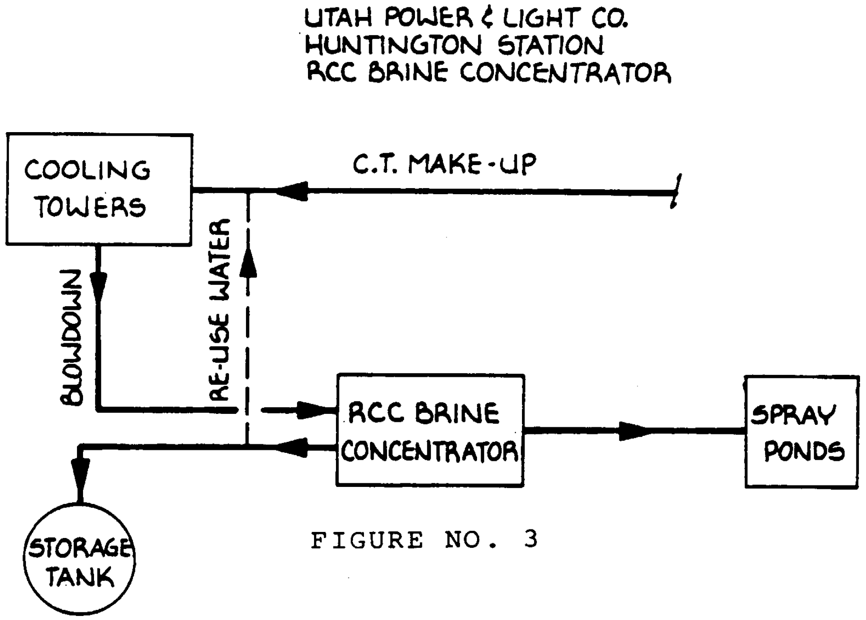
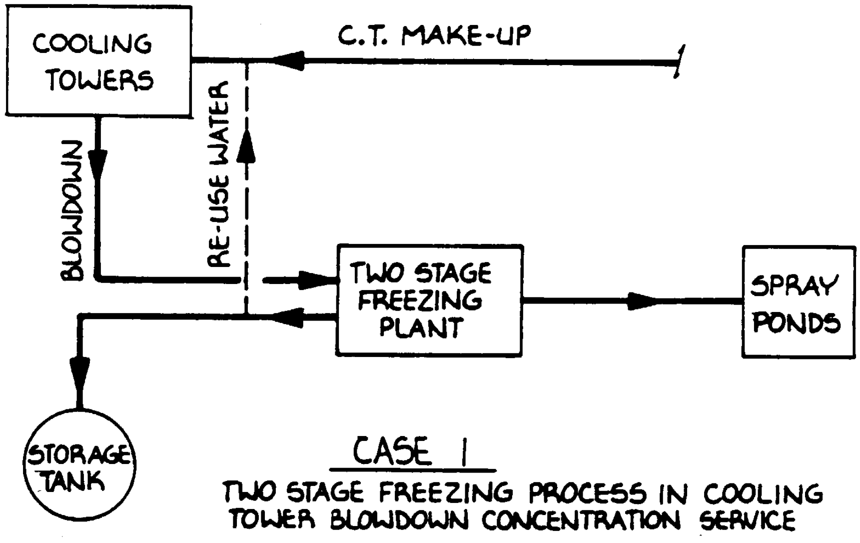

UTAH POWER & LIGHT CO.
HUNTINGTON STATION
RCC BRINE CONCENTRATOR

FIGURE NO. 3



CASE I

TWO STAGE FREEZING PROCESS IN COOLING TOWER BLOWDOWN CONCENTRATION SERVICE

FIGURE NO. 4

this plant is also used for cooling tower makeup. The final concentrate from the eutectic freezing plant is mixed with the underflow from the chemical pretreatment unit, and is pumped to a pond or spray dryer. The process schematic for this case is presented in Figure 5.

Alternatively, the cooling tower blowdown could be directly desalted in a reverse osmosis unit, but at a lower recovery, due to higher hardness in the cooling tower blowdown. The reject stream from the reverse osmosis unit could be further concentrated in a eutectic freezing plant. The product from the reverse osmosis and the eutectic freezing units are combined and used as cooling tower makeup. This scheme is described as Case III. A schematic of this case is presented in Figure 6.

Basis of Comparison. Comparison of the energy comsumption, capital costs, direct operating costs, and total costs of the RCC brine concentrator with the alternative processes identified as Cases I, II, and III was carried out for both a low and a high final concentration.

The lower range of final concentration for comparison was selected to be a brine containing 10 percent salts, including dissolved and suspended solids, because it was the design concentration for the RCC Brine concentrator at the Utah Power and Light Co. The higher concentration was selected to be about 32 percent salts, which would be the limit for the RCC brine concentrator because of $MgSO_4$ and Na_2SO_4 present in the cooling tower blowdown. At this concentration, the effluent from the eutectic freezing plant in Cases I, II and III would be around 60 percent solid hydrate crystals and 40 percent eutectic brine. Since the hydrate salts in each case are stable up to 90°F, the concentrate from the eutectic freezing plant would be 60 percent solids, but only 32 percent salt, because of water in hydrate crystals.

As described earlier, the flow to the RCC plant alone, and in Cases I, II and III would be 225,000 GPD (156 GPM) of cooling tower blowdown at 3469 ppm TDS.

In Case II, the result of feed pretreatment is a reduction of TDS to 2949 ppm, and SiO_2 to about 30 ppm. This makes an 85 percent recovery possible in the reverse osmosis unit. The reverse osmosis reject stream would have 18,493 ppm TDS. The product from the reverse osmosis unit would be 127.5 GPM, and the feed to the two stage eutectic plant would be only 22.5 GPM or 33,750 GPD.

In Case III, the reverse osmosis unit has a much lower recovery of 35 percent, which results in a reject stream TDS of 5145 ppm, at a flow of 101 GPM or 146,000 GPD. This reject stream is fed to the eutectic freezing plant.

Results. Estimates of capital cost, direct operating costs, and total costs of concentrating the cooling tower blowdown are presented in Figure 7. All energy consumption and costs are calculated per thousand gallons of feed (which is the cooling tower blowdown). In Fig-

CASE II

COMBINATION CHEMICAL TREATMENT – REVERSE OSMOSIS & 2 STAGE FREEZE CONCENTRATOR

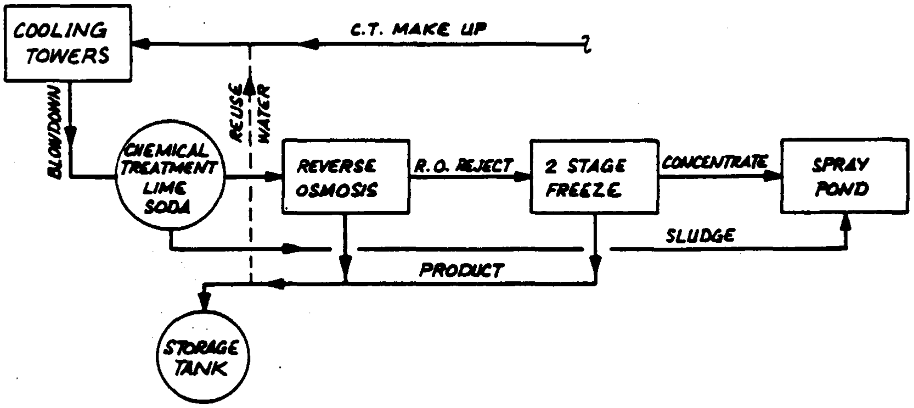


FIGURE NO. 5

CASE III

COMBINATION REVERSE OSMOSIS & 2 STAGE FREEZE PLANT AS CONCENTRATOR

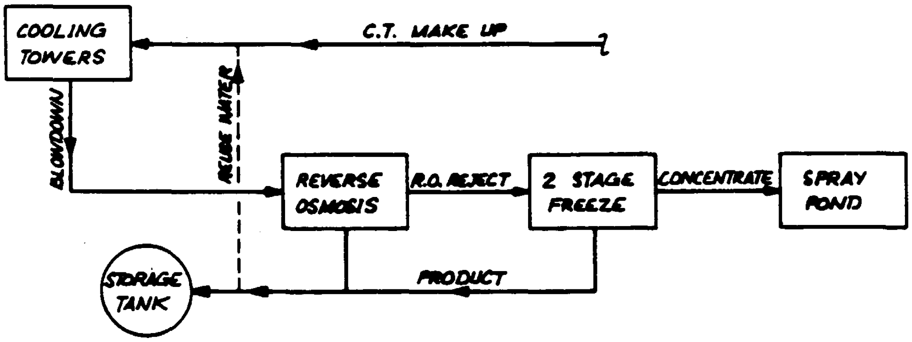


FIGURE NO. 6

ure 7, estimates were made only for the low and high concentrations. The lines joining through two concentrations have been shown straight in the absence of additional points.

The results indicate that at a final concentration of 10 percent salts by weight, the energy consumption of the combination of processes in Case II is lowest, being about 12.3 KW-Hr/1000 gallon compared to 70 KW-Hr/1000 gallons for the RCC brine concentrator. The energy consumptions for Cases I and III at this final concentration were 32.7 and 30.0 KW-Hr/1000 gallons, respectively.

The direct operating costs, which include chemicals, were \$0.71/1000 gallons, in Case II, compared to \$1.75/1000 gallons for the RCC brine concentrator. The direct operating costs for Cases I and III were \$0.89 and \$0.86 per thousand gallons, respectively.

The energy consumption at a higher final concentration of 32 percent was 14.60 KW-Hr/1000 gallons for Case II, compared to approximately 150 KW-Hr/1000 gallons for the RCC brine concentrator. The energy consumption for Cases I and III were 35.5 and 32.2 KW-Hr/1000 gallons, respectively. The direct operating costs for Case II would be \$0.77/1000 gallons compared to \$3.75/1000 gallons for the RCC brine concentrator. The direct operating costs for Cases I and III were \$0.96 and \$0.92 per thousand gallons, respectively.

For either low or high final concentration, the installed cost of the RCC brine concentrator on a prepared site would be of the same order of magnitude as a eutectic freezing plant, around \$1.4 million. The cost of a combination chemical pretreatment plant, reverse osmosis unit, and eutectic freezing plant (Case II process) would be approximately \$654.000. Case III, which is a combination of eutectic freezing and reverse osmosis without chemical pretreatment of the cooling tower blowdown would cost approximately \$1.2 million.

The overall or total cost, including direct operating cost, depreciation, and capital recovery of concentrating cooling tower blowdown to 32 percent salts would be \$1.80/1000 gallons for the combination lime soda, reverse osmosis, and two-stage eutectic freezing processes compared to \$6.00 for the RCC brine concentrator. The overal costs of alternative processes, Cases I and III, would be around \$3.27 and \$3.01, respectively.

Of course, it would also be possible to use a combination of reverse osmosis and an RCC concentrator with significantly better results than using the RCC unit alone. So the fact that we compared the RCC Unit alone with the combination reverse osmosis and the eutectic freezing plant could be construed as comparing apples and oranges. However, the intent was to compare present commercial practice with where we could be. If we had had sufficient data and time, we could have calculated and presented the case reverse osmosis and RCC. We did not. However, we believe that the costs and power consumption would have been higher for this combination than for reverse osmosis and the eutectic freezing plant based on our direct comparisons of RCC with the eutectic freezing plant. These cases are more like "apples and

DIRECT OPERATING & TOTAL COST COMPARISION

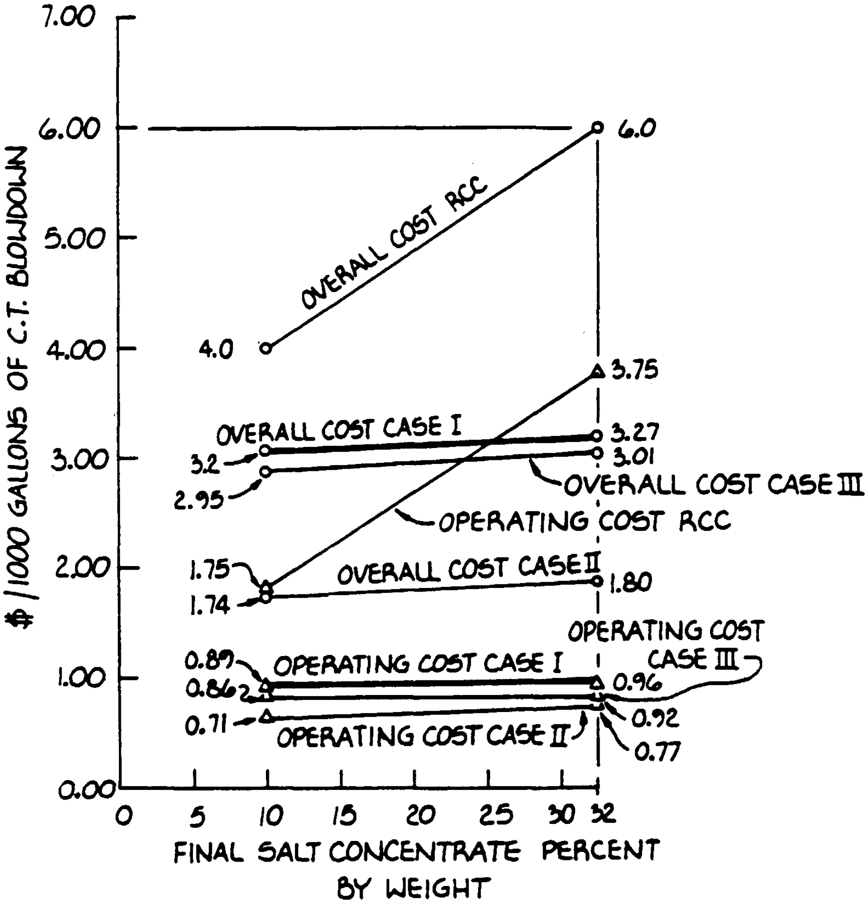


FIGURE NO. 7

apples". Power consumption is estimated at 150 KWH/KGal for the RCC unit and 35.5 KWH/KGal for the freezing case.

Other Examples

When one considers the number of industries that use cooling towers, it becomes clear that the above example of industrial use of desalting technologies is of extremely widespread application. Obviously, this can be easily extended to any industry that has boiler blowdown, demineralizer waste or any of the other waste streams mentioned under the paragraph "The RCC Process" above. To these can be added the following other specific examples of industrial use of desalting technologies that are either in operation or planned:

- Concentration of acid mine water
- Preliminary demineralization by reverse osmosis prior to ion-exchange demineralization
- Recovery of high quality water and valuable components from metal finishing photographic electroplating, cheese whey processing, petrochemical, pulp and paper, and textile industries.

ENERGY SOURCES

The desalting processes do not know or care where the energy comes from as long as there is enough of the correct quality. One could further develop the concept of using wind power (windmills) to generate electricity. At the same time he could do further development work on reverse osmosis or electrodialysis. But since the motors or cells do not know where the electricity is coming from, these must be regarded as two completely independent development projects that are carried out simultaneously. A similar case would be the simultaneous demonstration of gas from coal and a new distillation process.

On the other hand, a good case can be made for the fact that a dual purpose power-distilled water plant can be more economical than two single purpose plants. Development and demonstration of improvements in the connecting links between these two units can be justified.

Location is another important factor. Some processes can be run on either heat or electricity. Steam turbines can drive pumps for reverse osmosis or vapor compressors. Selection will depend on relative costs for different types of energy at each location.

Some thought is being given to simultaneous solutions to independent problems. For example, one area that has a sewage disposal problem, and a water shortage is considering using sewer gas and ultimately sewage as fuel for a distillation plant.

Another unique energy source can result from combining a liquified natural gas (LNG) vaporizer with a freezing process desalting plant.

Such a plant would require virtually no power for the process except that which is required to bring the seawater into the plant, and remove the brine and product water. One could suggest other uses for the "cold", but, at least at some locations, alternative uses are either economically or environmentally unacceptable [9].

CONCLUSION

Desalting technologies will find ever-widening application in the solution of industry's waste-stream problems of today and tomorrow.

NOMENCLATURE

As frequently happens in a relatively new discipline, no uniform system of nomenclature has been established for the subject matter of this paper. The author would like to emphasize two of the more important differences:

"Reuse"

As herein used, the term "water reuse" means water that has been impaired within a certain facility, and is then reused at some point within that facility. In some instances, it will be possible to "reuse" this water in a place in the facility where the quality requirements are such that desalting of the impaired stream is not required. In other cases, desalting will be required. What we refer to here as "reuse" is referred to by others as "recycle".

"Waste Water Clean-up"

As herein used, the term "waste water clean-up" means the final clean-up of a stream before it is discharged. Desalting may or may not be required, depending on its salt content. Others refer to such a stream as "reuse" on the basis that it will be "reused" by others.

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SOLERAS SOLAR ENERGY WATER DESALINATION PROJECT

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Introduction

In October 1977 Saudi Arabia and the United States signed a Project Agreement for Cooperation in the Field of Solar Energy (SOLERAS) under the auspices of the United States-Saudi Arabian Joint Commission on Economic Cooperation. The objectives of the agreement are:

- to cooperate in the field of solar energy technology for the mutual benefit of the two countries, including the development and stimulation of solar industries within the two countries,
- to advance the development of solar energy technology in the two countries, and
- to facilitate the transfer between the two countries of technology developed under this agreement.

The Solar Energy Research Institute (SERI), as the Operating Agent is responsible for implementing SOLERAS in accordance with directives of the SOLERAS Executive Board.

A five-year technical program plan for SOLERAS has been approved by the Executive Board. As part of this technical program plan, an area of Industrial Solar Applications for solar technology has been identified. The objectives of the Industrial Solar Applications program area are to introduce solar energy technologies into industrial applications and to foster the establishment of domestic industries using renewable energy sources, thereby lessening industrial dependence on fossil fuels and minimizing deleterious effects on the environment. A specific objective is to demonstrate the utilization of solar energy to desalinate water.

There is a need for water desalination both in Saudi Arabia and in the United States. In Saudi Arabia water is needed principally for municipal and agricultural applications. In the United States desalination is mainly required to control river salinity and to provide potable water to selected communities that have critical water quality problems or that have water shortages.

Conventionally powered desalting plants have been in operation for several years. At the beginning of 1977, about 1500 land-based, fossil fuel or electric powered, desalting plants with a minimum capacity of 100 m^3 per day were in operation or under construction throughout the world. These plants are capable of producing nearly four million cubic meters of fresh water daily for municipal or industrial uses. Distillation processes account for 77 percent of total plant capacity. The balance is almost entirely in membrane processes.

In 1977, Saudi Arabia had a conventional desalination plant capacity of 87,000 m^3/day. Projects are underway for plants to be constructed during the next 5 years with a total capacity of 1.36 million m^3/day of desalted water. Most of these plants are located along the Arabian Gulf and the Red Sea coast.

In the United States at the beginning of 1977, 510 conventional desalination plants provided about 0.4 million m^3/day of fresh water.

Despite this widespread desalination activity, no significant efforts have been made to replace the conventional power plants associated with desalination of water with solar energy systems. Most solar desalination work in the world has been restricted to simple solar stills of limited output.

The SOLERAS Solar Energy Water Desalination Project has been conceived in order to advance the technical and economic feasibility of large scale solar powered desalination of brackish and seawater.

Because of differences between seawater and brackish water desalting, two distinct systems will be developed under this project. It is expected that both systems will find applications, both in the United States and in Saudi Arabia.

To accomplish the objective of the SOLERAS solar energy water desalination project, a 3-phase activity is planned. The phases are as follows:

Phase 1: Preliminary System Design and Cost Analysis
Phase 2: Detailed Pilot Plant Design and Construction
Phase 3: Pilot Plant Operation and Training of Personnel

Phase 1: System analyses and economic analyses will be performed by several companies on a solar energy desalination system of their choice for either sea water or brackish water desalination. The systems will each be for an average daily product water capacity of 6000 m^3. The main criterion for the analysis will be the product water cost. Each system will be designed for a specific site and application. The site, application, and technology will have broad applicability to general water desalination needs in either the United States or Saudi Arabia. It is the intent of this project to encourage innovation without unduly affecting performance and reliability. Subsystems and their interfaces will be defined during Phase 1 and product-water cost projections will be made for commercial plants of a range of capacities.

Finally, a development plan for Phase 2 will be generated including detailed cost estimates for the design and construction of a pilot plant with a capacity of 100 to 400 m^3/day using the technology of the baseline system.

Phase 2: Of the several systems designed in Phase 1, one system in each category (brackish and seawater desalination) will be chosen for pilot plant construction. The criteria for selection will include levelized cost per unit of product water for the commercial sized plant, design and construction cost for the pilot plant, consistency in cost between the commercial sized plant and the pilot plant, maturity of system design and projected plant reliability. Each pilot plant will have a product-water output capacity of 100 to 400 m^3/day. The pilot plants will be designed in detail and constructed on specific sites.

The size of the pilot plant was selected to be within the budget limitations of the SOLERAS program and is of a capacity that provides useful technical and economic data for the planning, design, and construction of a commercially-sized plant. A pilot plant delivering 400 m^3/day of desalted water would provide water to 2,000 people or could provide irrigation water for about 8,000 m^2 of greenhouse agriculture. If the ratio of the ultimate plant capacity to the pilot plant capacity becomes too great, less useful technical and economic information for application to the full scale plant can be extracted from the pilot plant construction and operation.

Phase 3: The pilot plants will be operated and performance measurements made to provide the information essential for designing commercial-sized desalting plants. Local personnel will be trained in the operation and maintenance of the plant so they can make performance measurements.

The schedule for Phase 1 is from October 1980 to July 1981. Phase 2 is expected to start in October 1981 with the pilot plant construction completed by July 1983. Phase 3 will start at the completion of Phase 2 and will continue until the end of 1983.

Five companies have been awarded contracts for Phase 1. These companies and their team members are shown in Table 1. The technologies involved in the five systems represent six different desalination technologies (seawater and brackish water reverse osmosis are regarded as two different processes, and five different solar energy technologies. Two of the systems are for brackish water to be located in the United States, and three of the systems are for seawater to be located on the Red Sea coast in Saudi Arabia.

During the Phase 1, particular attention will be directed to defining the systems in detail and to generating cost estimates for the commercial size plant providing 1.8 million cubic meters per year. All cost estimations will be made to a common methodology to assure that the resulting levelized cost projections will be comparable. Of course, there are some differences in the systems between the brackish water systems and seawater desalination systems that make cost comparisons difficult. However, within each type of plant (brackish water and seawater plants) the costs should be comparable. To determine the sensitivity of the levelized product water cost to plant capacity, cost analysis will be performed for plant capacities ranging from 1,000 to 10,000 cubic meter per day of desalted water. These cost analyses will be made both for plants constructed in the United States and in Saudi Arabia. In addition to the cost estimate for the commercial-sized plants, preliminary designs of smaller pilot plants will be made. The pilot plants are expected to be in the capacity range of 100 to 400 cubic meter per day.

The results from Phase 1 will be analyzed for selection of contractors and systems for the Phase 2 activities, namely, the detail design and construction of pilot plants. The criteria that will be used in making that selection are:

- Projected levelized product water cost from a commercial size plant.
- Design and construction cost of a pilot plant.
- Consistency in cost between the commercial size plant and the pilot plant.
- The completeness of the Phase 2 implementation plan.
- Maturity of system design and pilot plant reliability.
- Contractor experience and capabilities for plant construction.
- Ability of the system to meet the system requirement imposed.

Table 1. CONTRACTORS FOR PHASE 1

Prime Contractor	Team Members
Boeing Engineering & Construction Co.	Resources Conservation Co. International
Catalytic, Inc.	Science Applications, Inc.
Chicago Bridge & Iron Co.	Foster-Miller Associates Inc. Arabian Chicago Bridge & Iron Co.
DHR, Inc.	Science Applications, Inc. Ionics, Inc. Al-Radwan
Exxon Research & Engineering Co.	Permutit Co., Inc. Martin-Marietta Badger Energy, Inc. Saudi Investment Development Center Ecodyne-Unitec Div.

SYSTEM DESIGN FOR A COMMERCIAL SOLAR BRAYTON CYCLE CENTRAL RECEIVER WATER DESALINATION PLANT

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ABSTRACT

The system design for a future commercial solar energy brackish water desalination plant is described. Key features of the plant are discussed along with its configuration selection rationale, design objectives, operation, and performance.

The water treatment technology used in the plant is ion exchange pretreatment and single stage reverse osmosis desalination utilizing high-flux membranes. Electrical power needed for plant operation is provided by a solar energy system which is based on the Brayton cycle having air as the working fluid. Primary solar system components are: heliostat field, central cavity-tube receiver, receiver support tower, thermal energy storage, and a commercial gas turbine generator set. The thermal energy storage subsystem is of the sensible heat brick type and provides a capability for continuous day/night power generation during most weather conditions.

This system design was selected in a study of various system alternatives and their life cycle product water costs for a representative site in western Texas.

INTRODUCTION

Boeing Engineering and Construction (BEC) has been awarded a contract by the Solar Energy Research Institute (SERI) to prepare a design for a solar energy water desalination system that would transform brackish well water into potable water for a community in southwest Texas. BEC is performing overall project management, system engineering and solar subsystem design; Resources Conservation Company, a partly-owned subsidiary of BEC, is providing designs for the water-related subsystem.

The 10 month contract which began in October, 1980 covers Phase 1 of a 3-phase program that is (1) sponsored jointly by the governments of Saudi Arabia and the United States as part of the SOLERAS agreement and (2) administered by the Solar Energy Research Institute (SERI) to advance the development of solar energy technology in the two countries. Phase 2 will involve construction of a pilot plant and Phase 3 will cover pilot plant operation and training of personnel. Operation of the pilot plant will provide verification of the design features and performance of the large-scale commercial plant.

This paper reports work done early in Phase 1 under Task 3, System Analysis, dealing with selection of a system configuration for a commercial SOLERAS solar energy water desalination plant. The objectives of the configuration selection study were to define a system that has near-term practicability, operational flexibility, and overall system simplicity.

GENERAL SYSTEM CONCEPT

The system design is based on Brayton cycle solar power generation and reverse osmosis water desalination. All electrical power needed for normal plant operation is provided by a Brayton cycle engine (gas turbine) having compressed air as a working fluid. Primary solar system components are: heliostat field, central cavity tube heat exchanger receiver, receiver support tower, thermal energy storage and a commercial gas turbine generator set. Most of these components utilize technology being developed in separate U.S. Department of Energy and Electric Power Research Institute projects. A simplified schematic is shown in Figure 1.

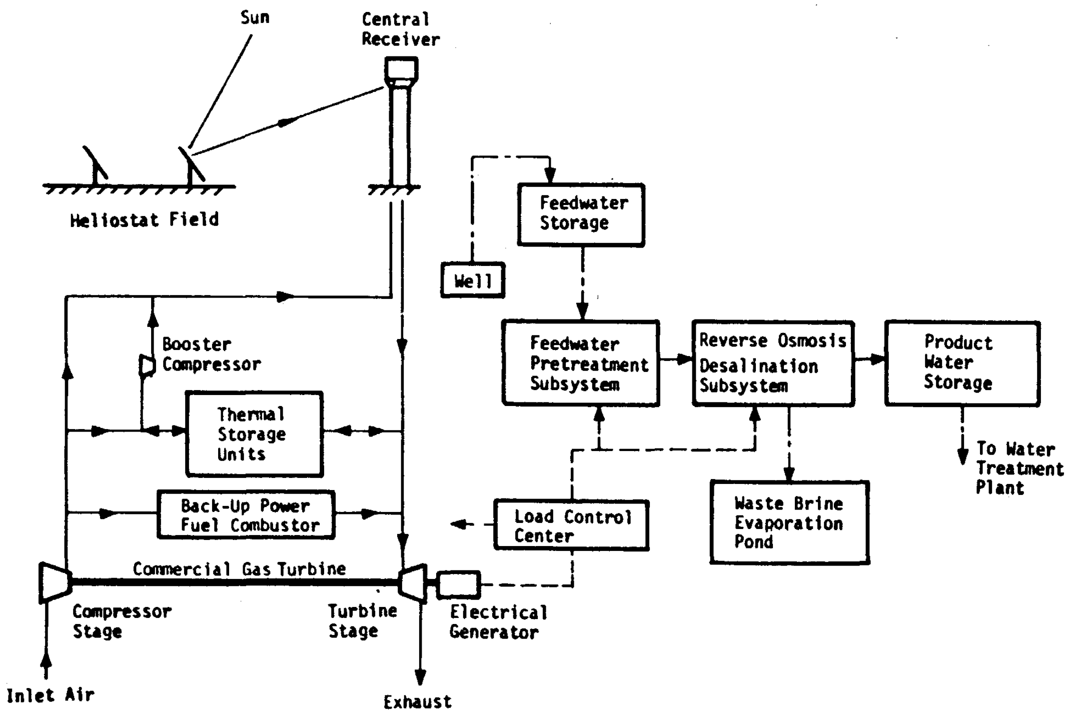

Figure 1 General System Schematic

Air is used as a solar system working fluid in order to (1) simplify design of the receiver, thermal storage units, fluid lines and interfaces (no corrosion, low cost, etc.), (2) allow use of commercial turbomachinery, and (3) simplify interfacing of these components because of the common working fluid. Also with this type of system, fossil fuel can be burned in a modified turbine combustor to provide emergency standby power at low additional system cost.

An important operational aspect of the system is that the water subsystems should be run as continuously as possible to minimize product water cost. This, combined with a SERI requirement that normal operation shall not use fossil fuel, suggests that some form of thermal energy storage be used to accomplish continuous day/night power generation. For thermal storage, we have selected simple sensible heat units having pressure shell/insulated brick checkerwork construction (a storage scheme that is very compatible with the air Brayton cycle). With this type of energy storage, charging is accomplished by routing heated air from the receiver to the thermal storage units during daytime conditions. When the solar resource becomes inadequate, the compressor flow is diverted away from the receiver and is routed through the thermal storage units where it is heated to the required turbine inlet temperature.

Reverse osmosis is selected as the basic water desalination method. Compared to other desalination technologies, the reverse osmosis process has low power demand which minimizes the solar subsystems size and costs. Also the reverse osmosis process has a simple electrical interface with the solar side of the system.

Based on this general system concept, several system variations (all using reverse osmosis) having different water recovery ratio and solar subsystem sizes were defined. These candidate systems, as discussed in the following sections, were evaluated in terms of levelized product water cost and other factors to arrive at a final commercial SOLERAS plant selection.

SYSTEM REQUIREMENTS

The general system requirements that govern the system design are listed below [1,2]:

Brackish water desalination based on
    SERI's before/after chemistry standards
1,800,000 m^3/year product water production rate
Minimum levelized product water cost
    using SERI economic parameters
20 year plant life
High reliability/maintainability
Independent from fossil fuel in normal operations
Technical readiness
Low risk

Minimum overall water recovery factor: 0.70
Minimum plant availability factor: 0.82
Feedwater storage: 3 days
Product water storage: 10 days
Fuel storage for backup power: 7 days
Minimum process chemical storage: 30 days
Design life: 20 years
Zero leakage evaporation pond

In addition to these requirements, the system is designed to meet several other conditions. First, water production is based on electrical power generation capability. This condition offers independence from having to design the system to meet unusual local water demand profiles and allows easier assessment of plant performance in various locations.

The plant is designed for a location in the Midland-Odessa area of West Texas. SOLMET and 30 year average weather data is available for Midland-Odessa and is suitable for plant design and performance analysis. This region has good solar insolation and has wide spread water quality problems and thus could benefit from commercial availability of solar water desalination plants.

Several conditions were specified that influence product water cost. Feedwater is assumed to be groundwater and is pumped to the plant at no cost. After desalination, the consumers will draw product water from plant storage for purification in their own treatment facilities. Finally, the plant is assumed to be the "Nth" constructed which implies initial design and development costs have already been amortized.

CANDIDATE SYSTEM CONCEPTS

Four candidate system concepts were defined that all employ reverse osmosis (RO) desalination technology and satisfy the requirements of product water production rate and before/after water chemistry. These candidate systems, shown in Figure 2, differ in (1) the method of feedwater pretreatment, (2) number of reverse osmosis stages, (3) use of standard flux vs. high flux reverse osmosis membranes, and (4) use of multiple effect vapor compression for increased water recovery. The systems consequently have varying water recovery ratios, electrical power demands, and water subsystem costs.

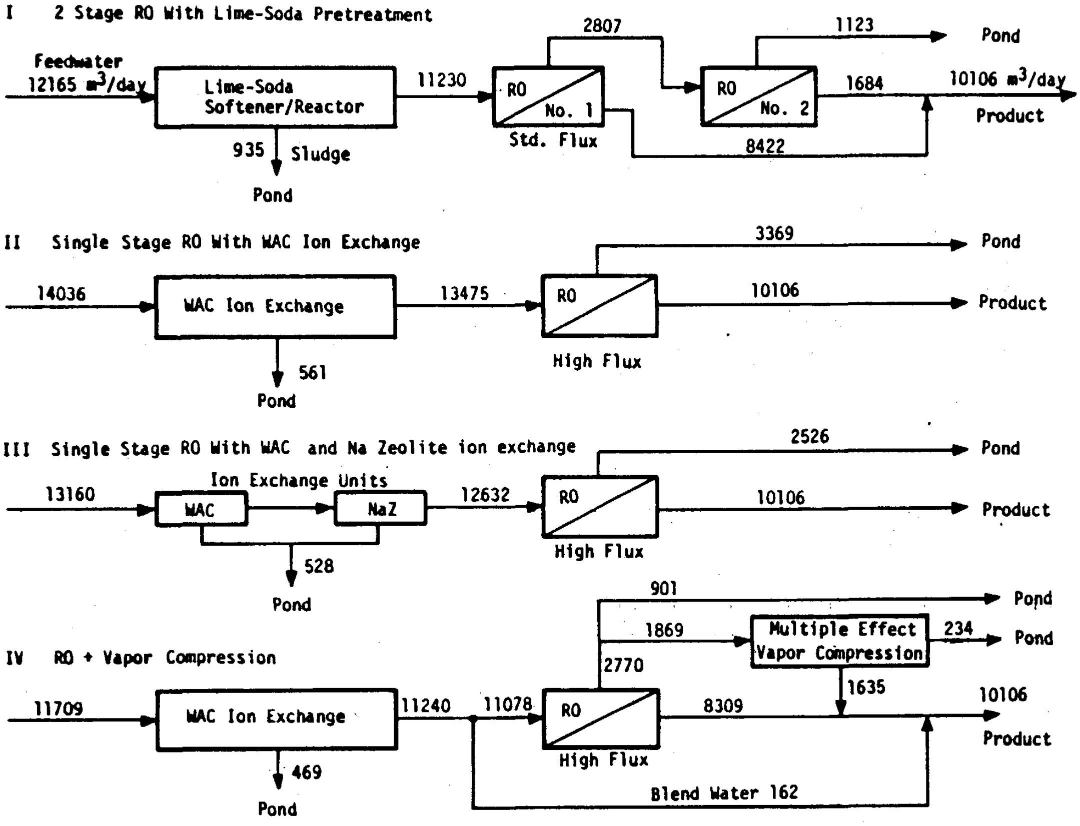

Figure 2 Candidate Water Subsystems

Configuration I (2 stage reverse osmosis with lime-soda pretreatment) served as a baseline configuration in a preliminary systems analysis study that defined sizes of the primary solar subsystem components. This baseline system was analyzed for performance on a seasonal basis using 30 year averaged weather data for Midland-Odessa, Texas [3]. The DELSOL code [4] was used to design the heliostat field and predict seasonal energy inputs to the central receiver. A general plant performance computer model was used to predict daily electrical power production of this design on an hourly basis.

Table I lists the predicted seasonal electrical output for the baseline system. this performance, together with sun availability (weather factor) and a conservative assumption for plant availability, forms a basis for defining the peak water production design point for the water subsystems. The peak rate of 10106 m^3/year was used to size the water related components of each candidate system.

Table I Daily Design Water Production Rate

		March	June	Sept	Dec	Average
Net plant electrical output (MW_e-hr)		22.2	18.3	17.9	15.4	18.45
Clear day plant performance						
Baseline system						
Relative electrical output	E	1.21	0.99	0.97	0.83	
Sun availability	S	0.72	0.80	0.79	0.74	
Plant availability	A					0.82
Annual water production rate (m^3/day)	Vo					4945
Required daily production (m^3/day)						
Vo(E)/(AS)		10106	7472	7400	6796	
Basic assumption:		Peak Design Flow				
Water-production follows daily energy production in each season						

After establishing pump and plant parasitic power requirements for each system, the respective solar system components could then be sized. Also, evaporation pond sizes were determined for each system based on their respective water recovery ratios and a mean pond evaporation rate of 1.35 m /year. A summary of the candidate system size parameters appears in the following table.

Table II Candidate System Features

		Water Recovery Ratio %	Evaporation Pond Size Ha	Peak Power Demand kW_e		Turbine	Number of Heliostat	Thermal Storage Media Mass Gg
				Water Systems	Total Plant			
I	2-Stage RO with lime-soda pretreatment	90	13.3	724	1034	Simple Cycle Centaur	600	2.5
II	Single stage RO with WAC ion exchange	72	51.9	512	731	Simple Cycle Saturn	436	1.68
III	Single stage RO with WAC and NaZ ion exchange	76.5	41.0	607	867	Combined Cycle Saturn	436	1.68
IV	RO + vapor compression with WAC ion exchange	86.3	21.2	1459	2083	Combined Cycle Centaur	600	2.5

A survey of commercial gas turbines indicated the Saturn and Centaur turbines, produced by Solar Turbines International, are a good match to the candidate systems. These turbines are currently available, are in widespread use, and have sufficiently high pressure ratios (of importance to receiver design). To meet the higher power demands, the turbines for systems III and IV require combined cycle equipment in which the gas turbine reject heat is used to produce steam for a turbine-generator set. The combined cycle turbines, since they utilize reject heat, do not require larger heliostat fields or more massive thermal storage than their respective simple cycle systems.

Costs for the candidate systems are listed in Table III. These costs are engineering estimates for an "Nth" plant. Major equipment costs are based on information obtained from suppliers and cost estimates prepared for a recent Department of Energy project for industrial process heat involving similar heliostat field and central receiver concept[5]. Since the plant is not near-term, the heliostat status costs are assumed to be $200/m^2 for system comparison purposes. Water subsystem costs are based on information from suppliers obtained by RCC for prepackaged skid-mounted assemblies. A sizeable portion of the operation and maintenance costs are due to plant personnel (14 total) and costs related to the water subsystems (reverse osmosis membrane replacement).

Table III Cost Model Data for Candidate Systems

	Cost Account	I 2-Stage RO ($1000)	II 1-Stage RO	III 1-Stage RO	IV RO & Vapor Compression
5101	Site	184	220	220	184
5102	Facilities & Enclosures	400	416	450	400
5103	Solar Energy Collection	6,828	5140	5,140	6,828
5104	Energy Storage	2,900	210	2,050	2,900
5105	Energy Delivery	1,528	600	900	2,528
5106	Back-Up Power	134	95	100	175
5107	Feedwater Pretreatment	6,570	1,600	600	1,600
5108	Desalination	9,550	3,240	3,000	15,740
5109	Water Storage & Delivery	2,600	2,600	2,600	2,600
5110	Waste Disposal	3,483	13,000	10,240	5,300
5111	Controls & Instrumentation				
	Solar Subsytems	350	343	410	350
	Water Subsytems	350	296	400	450
5112	Data Acquisition	150	150	150	150
5120	Maintenance Support	50	50	50	50
5130	Technical Data	20	20	20	20
	Total Capital Costs	35,274	29,870	26,159	39,334
	Annual Operating Costs	1,070	563	670	960
	Annual Maintenance Costs	815	351	387	635
	Annual Back-Up Fuel Costs	25	18	22	51

SYSTEM EVALUATIONS

In evaluating the candidate systems, four areas were considered:

Levelized (life cycle) product water costs
Technology readiness
Complexity
Application limitations
Implementation risks

Product water cost is a dominant evaluation factor and could be treated quantitively. The other areas are of a qualitive nature but are never-the-less important to system evaluation and selection.

The levelized product water costs for 1.8 million m^3 annual production were computed by the method for sum-of-the-years-digits depreciation given in reference [6] using economic parameters specified by SERI [1]. The primary parameters are:

System operating lifetime	20 years
Accounting lifetime	16 years
Cost of capital (and rate of return on capital)	0.086
Base year for constant dollars	1980
Price year for cost information	1980
First year of commercial operation	1983
Rate of general inflation	0.060
Escalation rate for capital costs	0.060
Escalation rate for operating costs	0.070
Escalation rate for maintenace costs	0.070
Escalation rate for fuel costs	15%
Insurance + "other tax" fraction	0.020
Investment tax credit	0.100
Tax rate	0.5
Raw land cost	$\$1.25/m^2$
Cost for lined evaporation ponds	$\$25/m^2$
Cost for fuel oil (31 GJ/m^3)	$\$157/m^3$

The resulting computed cost parameters that were used to calculate levelized product water costs are:

Capital recovery factor (8.6%, 20yrs)	0.1064
Present value of sum-of-the-years-digits depreciation	0.6376
Fixed charge rate	0.1437

It should be noted that the study ground rules presented by SERI were chosen to enable comparison of widely varying system concepts. Consequently, the levelized product water costs may not represent actual realizable values, but are useful for evaluating the candidate systems.

Table IV presents levelized product water costs for the candidate systems using the respective system cost data given in the previous table. Systems II and III, while having lower water recovery ratios, have levelized product water costs that are relatively low and comparable. System III has a slightly lower product water cost than System II primarily because of the evaporation pond size reduction allowed by the higher water recovery ratio. Systems I and IV have significantly higher product water costs due to higher capital, operation and maintenance costs; these higher costs are not offset by the associated higher water recovery ratios.

Table IV Product Water Cost Comparisons

		Overall Water Recovery Ratio	Costs			Levelized Water Cost
			Capital	Annual Operating	Annual Maintenance	
		%	$M	$M	$M	$\$/m^3$
I	Two Stage RO Lime soda pretreatment Std. membranes Simple cycle Centaur	90	35.3	1.07	0.82	4.86
II	Single Stage RO WAC Ion exchange pretreatment High flux membrane Simple cycle Saturn	72	29.9	0.56	0.35	3.39
III	Single Stage RO WAC + NaZ ion exchange High flux membrane Combined cycle Saturn	76.5	26.2	0.64	0.39	3.22
IV	RO + Vapor Compression WAC ion exchange High flux membrane Combined cycle Centaur	86.3	39.3	0.96	0.64	4.95

In the area of technology readiness, Table V, the water subsystems are all state-of-the-art and commercially available. On the solar system side, development is required for a production receiver design. The Electric Power Research Institute is currently funding BEC in a project called the Full System Experiment; this project is providing technical background in the areas of receiver design, receiver-turbine integration, and operation and control of a hybrid solar/fossil fuel central receiver as a power generation system. Heliostats are in an advanced stage of development at BEC as part of the Department of Energy's Second Generation Heliostat program. The turbomachinery components are currently available except for the combined cycle Saturn turbine configuration which requires development. From an overall viewpoint, all of the candidate solar systems require design development of a similar nature prior to commercialization.

Table V Technology Readiness Evaluations

	Candidate	
I	2-Stage RO with lime-soda pretreatment. Simple cycle Centaur.	S.O.A. water systems-standard membranes. Simple cycle Centaur is available.
II	Singe stage RO with WAC ion exchange. Simple cycle Saturn.	S.O.A. water systems-high flux membranes are now in service. Simple cycle Saturn is readily available.
III	Single stage RO with WAC + NaZ ion exchange. Combined cycle Saturn.	S.O.A. water systems-high flux membranes are now in service. Combined cycle Saturn needs development.
IV	RO + vapor compression with WAC ion exchange. Combined cycle Centaur.	S.O.A. water systems-high flux membranes are now in service. Combined cycle Centaur is available.
	General	All systems require similar solar subsystem design development.

Our assessment of candidate systems with respect to complexity, which relates to reliability, appears in Table VI. System II is the simplest system because of operational ease and flexibility, simple cycle power generation, and small energy collection subsystem size.

Table VI Complexity Evaluations

	Candidate	
I	2-Stage RO Lime-soda pretreatment Simple cycle Centaur	Lime-soda process requires close control
II	Single Stage RO WAC ion exchange Simple cycle Saturn	Simple on/off water systems - Minimal chemical controls smallest power demand - Smallest and simplest solar subsystems
III	Single Stage RO WAC and NaZ ion exchange Combned cycle Saturn	- Requires more controls than above - Increased complexity with combined cycle
IV	RO and vapor compression WAC ion exchange Combined cycle Centaur	- Highest water system complexity due to V.C. - Difficult to turn on/off - Complex combined cycle power generation - Largest solar subsystems

As indicated in Table VII, there are no significant limitations on locations for the candidate systems. Of course, the systems do have site limitations common to many other solar energy applications: high quality solar insolation is needed! System IV, because of its combined cycle power generation and vapor compression features (not a simple on/off system) is judged to be unsuitable for remote sites.

Table VII Application Limitation Evaluations

Candidate	
I 2-Stage RO Lime-soda pretreatment Simple cycle Centaur	- Can treat wide range of feedwaters
II Single Stage RO WAC ion exchange Simple cycle Saturn	- Requires certain concentration of ions in feedwater - Opportunity for continued cycle power generation - Requires largest pond
III Single Stage RO WAC and NaZ ion exchange Combined cycle Saturn	- Requires certain concentration of ions in feedwater
IV RO and vapor compression WAC ion exchange Combined cycle Centaur	- Can treat wide range of feedwaters - Requires skilled operators

Table VIII Risk Evaluations

Candidate	
I 2-Stage RO Lime-soda pretreatment Simple cycle Centaur	- Widely used processes - Lime-soda process not normally used in intermittent operations
II Single Stage RO WAC ion exchange Simple cycle Saturn	- New applications of WAC ion exchange and high flux RO, but suppliers will guarantee - Small solar system, size lessens development risk
III Single Stage RO WAC and NaZ ion exchange Combined cycle Saturn	- New application of WAC and NA ion exchange, but suppliers will guarantee - Combined cycle Saturn needs development
IV RO and vapor compression WAC ion exchange Combined cycle Centaur	- New application of WAC ion exchange, but suppliers will guarantee - Vapor compression not normally used in on/off operation
General	- Solar receiver technology under development in separate programs - Heliostat costs dependent on developing market

The candidate systems having varying implementation risks as shown in Table VIII. System II, because it has the simplest solar subsystems, appears to have the least risk for successful implementation and operation.

SYSTEM SELECTION

The preceding evaluations are summarized in Table IX. Based on these evaluations, overall ratings were determined that rank the candidate systems according to resonsiveness to the system requirements specified by SERI. The resulting ratings indicate that System II, single stage reverse osmosis with weak acid cation exchange, is the preferred system concept. System II was selected for the on-going Task 3 project activities consisting of additional system design definition, operating availability analysis, plant performance analysis and cost analysis of the selected commercial plant concept.

Table IX System Concept Evaluation Summary

	Candidate	Levelized Product Water Cost $\$/m^3$	Technology Readiness	Complexity	Application Limitiations	Risk Factors	Overall Selection Rating Based on SOLERAS Requirements
I	2-Stage RO Lime-soda pretreatment Simple cycle Centaur	4.86	High	Low	Minimal	Low	3
II	Single Stage RO WAC ion exchange Simple cycle Saturn	3.39	High	Low	Minimal (except for large evaporation pond)	Low	1
III	Single Stage RO WAC and NaZ ion exchange Combined cycle Saturn	3.22	Low	Medium	Minimal	Medium	2
IV	RO and vapor compression WAC ion exchange Combined cycle Centaur	4.95	Low	High	Some	High	4

CONCLUSIONS

As a result of the subsystem trade-off study, a system configuration was selected that meets the SOLERAS project requirements for a commercial solar energy water desalination plant. The configuration consists of a heliostat/central receiver energy collection subsystem, air Brayton cycle power generation subsystem, sensible heat thermal energy storage, weak acid cation feedwater pretreatment, single stage reverse osmosis water desalination, and other subsystems. A preliminary systems analysis of the plant indirectly it has the potential to desalinate water with attractive levelized product water cost and would have the simplicity and operational flexibility needed for isolated applications. The plant's economics are enhanced by the simple system configuration and capability for continuous day/night operations during normal weather conditions without using fossil fuel. The selected configuration has a low power requirement and thus allows continuous operation using sensible heat thermal energy storage of reasonable size.

ACKNOWLEDGEMENTS

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Water system design was the responsibility of our subcontractor, Resources Conservation Company, where work was directed by Mr. Deb Mukhopadhyay, Process Engineering Manager.

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BRACKISH WATER SOLAR DESALINATION

SOLERAS Program

Workshop on Solar Desalination

by

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25 March, 1981
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ABSTRACT

Under contract to the Solar Energy Research Institute (SERI), Catalytic Inc. with Scientific Applications, Inc. have engineered a design for a solar powered, brackish water, desalination facility. The design contract is part of the SOLERAS program, a joint United States/Saudi Arabian effort to further the development of solar technology. Brownsville, Texas was selected as the site for application of this design concept. The site is typical of many areas in the southwest United States and the Kingdom of Saudi Arabia.

The system concept combines advanced solar energy systems and reverse osmosis membrane desalination systems. An average daily water production of 6000 m^3/day is obtained from a brackish water with a quality of 6000 $^{mg}/l$ total dissolved solids (TDS). A product water quality of 500 $^{mg}/l$ TDS is obtained from a two stage reverse osmosis membrane system operating at 90% product water recovery. Feedwater pretreatment requirements include: lime softening, filtration, and ion exchange softening. Brine from the membrane units is disposed through solar evaporation ponds.

The electric power required to operate the desalination system is generated by the solar system. Point focus, high temperature collectors combine with parabolic trough, medium temperature collectors to provide thermal energy to a steam turbine generator. The power supply is complemented by a wind energy conversion system. The solar energy system has been designed for a solar fraction of 1.0.

INTRODUCTION

Saudi Arabia and the United States have signed a project agreement for cooperation in the field of Solar Energy (SOLERAS) under the auspices of the United States-Saudi Arabian Joint Commission on Economic Cooperation. The objectives of the agreement are:

- to cooperate in the field of solar energy technology for the mutual benefit of the two countries, including the development and stimulation of solar industries within the two countries,
- to advance the development of solar energy technology in the two countries, and
- to facilitate the transfer between the two countries of technology developed under this agreement.

The Solar Energy Research Institute (SERI), as the Operating Agent, is responsible for implementing SOLERAS in accordance with directives of the SOLERAS Executive Board.

A five-year technical program plan for SOLERAS has been approved by the Executive Board. As a part of this technical program plan, an area of Industrial Solar Applications for solar technology has been identified. The objectives of the Industrial Solar Applications program are to introduce solar energy technologies into industrial applications and to foster establishment of domestic industries using renewable energy sources, thereby lessening industrial dependence on fossil fuels and minimizing deleterious effects on the environment.

There is a need for water desalination both in Saudi Arabia and in the United States. In Saudi Arabia water is needed principally for municipal and agricultural applications. In the United States desalination is mainly required to control river salinity and to provide potable water to selected communities that have critical water quality problems or that have water shortages.

Despite widespread desalination activity, no significant efforts have been made to replace the conventional power plants associated with desalination of water with solar energy systems. Most solar desalination work in the world has been restricted to simple solar stills of limited output.

It is the objective of this project to advance the technical and economic feasibility of large-scale solar-powered desalination plants. "Solar" power is to be understood in its broadest sense so as to include both direct and indirect solar sources.

To accomplish the objective of this solar energy water desalination project, a three-phase activity is planned. The phases are as follows:

Phase 1. Preliminary System Design and Cost Analysis
Phase 2. Detailed Pilot Plant Design and Construction
Phase 3. Pilot Plant Operation and Training of Personnel

Catalytic has been awarded a contract to conduct a preliminary system design and cost analysis for a brackish water desalination project to be located in Brownsville, Texas. System analyses and economic analyses will be performed to define the baseline solar energy desalination system. The baseline system will be for an average daily product water capacity of 6000 m^3. The baseline system will be optimal relative to technological risk, performance, and product water cost. Subsystems and their interfaces will be defined and product water cost projections will be made for commercial plants over a range of capacities.

Scientific Applications, Inc. (SAI) is subcontractor to Catalytic with responsibility for the solar power system.

A development plan for Phase 2 is to be generated including detailed cost estimates for the design and construction of a pilot plant of 100 m^3/day capacity, using the technology of the baseline system.

SITE SELECTION

The goals of this project are to demonstrate the feasibility of solar desalination technology and to stimulate the use of solar energy for this and other applications in the United States (US) and the Kingdom of Saudi Arabia (KSA). This implies that the selected site must possess:

- the resource requirements crucial to efficient plant operation and product water utilization
- the climatic characteristics that represent large geographic regions in both countries
- the flexibility for conducting the demonstration experiment in either or both countries without any substantial changes in the technology and the design of the solar desalination system.

These considerations served as broad guidelines for our approach to site selection and subsequently to the formulation of a cost-effective solar desalination system concept. Our approach to site selection resulted in the selection of a US site, Brownsville, Texas, for the brackish water desalination system. The rational for this decision and the overall approach to site selection are briefly presented below.

The site selection approach involved two decision points. The first issue related to an evaluation of whether the brackish water desalination system should be sited in the KSA or the US. The second concern involved the identification of a geographic region that typified both countries. A set of criteria encompassing system location, design, operation, utilization, and technology transfer considerations was first developed. These criteria were then used in a qualitative but systematic manner to identify the country and region of location of the brackish water demonstration facility.

The question, whether a brackish water solar desalination system should be located in the KSA or the US, was addressed broadly on the basis of the availability and utilization of water resources in the two countries. Specifically, to resolve this issue, the following factors were considered:

- relative availability of water resources

- distribution of population centers and industry versus water resources

- availability of potable water from conventional sources

- projected desalination requirements for the next five to ten years

- representative solar characteristics

In Saudi Arabia, the major water-using population centers and industries are, generally speaking, concentrated along the Red Sea coast and Arabian Gulf coast. The emerging petroleum processing and petrochemical complexes are sited in coastal areas. In the US, this is not the case. Some of the more rapidly growing areas are those around Las Vegas, Nevada; Phoenix and Tucson, Arizona; Albuquerque, New Mexico; and Austin, Texas. These are mostly areas with a brackish water supply.

In both the US and Saudi Arabia, inland sites are hard pressed to serve the needs of the people from conventional sources. Water treatment and blending are required in both countries.

Saudi Arabia has developed long-range plans to meet the basic water needs of its people. A majority of the capacity proposed for Saudi installations is based on the desalination of seawater.

US plans are relatively poorly defined. A driving force toward water desalination may well be drinking water standards promulgated by the US Environmental Protection Agency (EPA). A proposed 500 $^{mg}/l$ standard to total dissolved solids could cause many inland population areas in the US to look to desalination technology to supply users with compliance water.

A US site is preferable for a brackish water solar desalination system considering equipment development and technology transfer. The specific region for such a plant should be selected with full cognizance of the application of the desalination technology to the KSA since growth of population centers and industry is being projected in the inland areas of Saudi Arabia.

A large number of US sites are amenable to siting of a brackish water solar desalination plant. The recent Office of Water Research and Technology Report entitled "Evaluation of Technical Material and Information for Potential Desalting Demonstration Plants (PB-290-338)" was used in the initial screening of sites. Table 1 presents data on the final fourteen sites considered. State and local authorities were contacted in each of the fourteen localities in the final round

TABLE 1 SITE COMPARISON

CRITERIA SITE	SOLAR (KWH/M^2/D)[b][c]	WIND (W/M^2)[a][b]	MAXIMUM ANNUAL WIND SPEED (KM/H)[d]	LAND AVAILABILITY	WATER (PPM TDS)	PROJECTED WATER SHORTAGE (1983-85)	WATER RESERVE	PLANS FOR DESALINATION PLANT AND CAPACITY	SUMMER/WINTER WATER CONSUMPTION RATIO	MAJOR WATER CONSUMPTION	COMPARABILITY TO SAUDI ARABIA	SUMMER/WINTER SOLSTICE	GEO/TECHNICAL HAZARDS	LATITUDE	ANNUAL EVAPORATION RATE (MM)[c]	AVERAGE ANNUAL PRECIPITATION (MM)[c]	MEAN ANNUAL RELATIVE HUMIDITY (PERCENT)[c]
US SITES																	
ARTESIA, NM	6	160	120	GOOD	>800	NO	BRACKISH	NONE	2:1	AGRICULTURAL	SIMILAR	1.57	SMALL	33°	1,813.5	330	45
BROWNSVILLE, TX	5	200	60	GOOD	1,150	YES	BRACKISH/ SEAWATER	UNDER STUDY	1.2:1	RESIDENTIAL/ INDUSTRIAL	"	1.36	HURRICANE	26°	1,511	610	75
DAUPHIN ISLAND, AL	4	115	140	FAIR	1,200	YES	"	NONE	2:1	RESIDENTIAL/ SEAFOOD IND	"	1.48	"	30°	1,248.1	1,650	76
GRAND ISLE, LA	4+	100	160	POOR	NOT AVAIL (BRACKISH)	YES	"	NONE	2:1	INDUSTRIAL/ TOURISM	"	1.46	"	29°	1,287.7	1,580	75
HENDERSON, NV	6	150	80	GOOD	6,000	YES	BRACKISH	SOUTHERN NEVADA ①	1.6:1	RESIDENTIAL/ MANUFACTURING	"	1.67	SMALL	36°	1,981.2	114	25
KEY WEST, FL	5	130	200	POOR	4,000 TO 6,000	NO	BRACKISH/ SEAWATER	NONE	1:1.2	RESIDENTIAL/ TOURISM	"	1.36	HURRICANE	24.5°	1,422.4	1,070	76
MARIN COUNTY, CA	4.7	45	100	RESTRICTED	15,000 (BAY WATER)	NO	BAY WATER	~ YEAR 2000	2.4:1	RESIDENTIAL (80%)	MODERATE	1.81	EARTHQUAKE	38°	997.4	1,050	75
MELBOURNE, FL	5	—	95		1,100	YES	BRACKISH/ SEAWATER	STUDY IN PROGRESS ②	1.5:1	RESIDENTIAL/ TOURISM	SIMILAR	1.43	HURRICANE	28°	1,188	1,330	80
ROSWELL, NM	6	160	120	GOOD	<1,500	NO	BRACKISH	NONE	2:1	AGRICULTURAL	"	1.58	SMALL	33.5°	1,813.5	330	45
SANTA ANNA, CA	6	40	80	RESTRICTED	2,500	YES	"	UNDER STUDY	2:1	INDUSTRIAL/ MUNICIPAL	"	1.6	EARTHQUAKE	34°	1,351	450	65
WICHITA, KS	5	250	160	GOOD	1,600	NO	"	NONE	1.6:1	RESIDENTIAL/ INDUSTRIAL	POOR	1.73	TORNADO	37.5°	1,724.6	645	60
WRIGHTSVILLE BEACH, NC	4	109	140	FAIR	SEAWATER	NO	BRACKISH/ SEAWATER	CONCEPT DESIGN③	1.6:1	RESIDENTIAL/ TOURISM	MODERATE	1.6	SMALL	34°	1,036.5	130	75
WOLSEY, SD	4	200	120	GOOD	N/A	YES	BRACKISH	NONE	2:1	RESIDENTIAL	POOR	2.17	SNOW/WIND	44.5°	865.6	523	70
YUMA, AZ	6	60	80	GOOD	>1,000	NO	"	COLORADO ④	2:1	AGRICULTURAL	SIMILAR	1.55	SMALL	32.5°	2,064.2	114	30
KSA SITES																	
ALHUFUF	5.5	10	40[b]	GOOD	—	YES	BRACKISH	—	—	MUN/AGRIC/IND		1.36	DUST	25°	—	74[b]	36[b]
ALQATIF	5.7	0.1	7[b]	GOOD	42,000[e]⑦	YES	BRACK/SEA	—	—	"		1.36	"	26°	2,677[b]	60[b]	60[b]
ATTAIF	5.1	0.6	15[b]	GOOD	—	YES	BRACKISH	—	—	MUN/AGRIC		1.29	"	22°	2,782[b]	210[b]	40[b]
JEDDAH	4.9	70	80[d]	GOOD	45,000[e] (RED SEA WATER)	YES	BRACK/SEA	85 MG/D[e]	MILD[e] SEASONAL CHANGE	RESIDENTIAL/ MUNICIPAL/ INDUSTRIAL•		1.29	"	22°	—	100[e]	—
MEDINA	6.6	1	—	GOOD	—	YES	BRACKISH	100 MG/D[e]	⑤[e]	"		1.32	"	24°	4,648[b]	40[b]	30[b]
RIYADH	6.4[f]	42[d]	90[d]	GOOD	—	YES	BRACKISH	—	⑥[e]	"		1.32	"	24°	3,000[f]	20[f]	30[f]
TABUK	4.3	14	—	GOOD	—	YES	BRACKISH	—	—	MUN/AGRIC		1.41	"	27°	1,592[b]	35[b]	37[b]

[a] WIND POWER CLIMATOLOGY OF THE UNITED STATES, JACK W. REED, SAND78-0348.

[b] ANNUAL ENVIRONMENTAL REPORT, 1977, MINISTRY OF DEFENSE AND AVIATION, GENERAL DIRECTORATE OF METEOROLOGY, KINGDOM OF SAUDI ARABIA.

[c] CLIMATIC ATLAS OF THE UNITED STATES, US DEPARTMENT OF COMMERCE, ENVIRONMENTAL SCIENCE SERVICES ADMINISTRATION, ENVIRONMENTAL DATA SERVICE, 1977.

[d] N SUMMARY, US AIR FORCE ETAC AIR WEATHER SERVICE (MAC), 1969.

[e] DESALINATION, 25:9-44, 1978, "DESALTING IN SAUDI ARABIA—DEMAND, PRODUCTION, MANAGEMENT, ASSOCIATED POWER GENERATION AND FUTURE PLANS," BY A.F. ABDUL-FATTAH, A.A. HUSSEINY, AND Z.A. SABRI.

[f] "SOLAR ENERGY IN SAUDI ARABIA TECHNOLOGY AND ECONOMICS," BY A.A.M. SAYIGH COLLEGE OF ENGINEERING UNIVERSITY OF RIYADH.

① PHASE II PROJECT IN PROGRESS.
② STUDY IN PROGRESS (25 MCD IN 1990).
③ CONCEPTUAL DESIGN IN PROGRESS (2 TO 20 MCD).
④ CONTRACT IN PROGRESS.
⑤ LOWER IN WINTER EXCEPT FOR HAJJ SEASON.
⑥ LOW DURING HAJJ SEASON & SUMMER (VACATIONS).
⑦ PERSIAN GULF.

of evaluations.

Brownsville, Texas, was selected as the US brackish water site. Strong interest was expressed by both the Brownsville Public Utilities Board and the Brownsville Navigation District. The prominent factors in the selection of the Brownsville site were:

- brackish water must be used to meet future water requirements
- the site latitude and solar energy resource characteristics of the Brownsville area approximate that of Riyadh, which is visualized as the most representative Saudi Arabian center of inland water needs.

Brownsville offers the unique opportunity to blend water from various sources and operate the proposed pilot plant under a range of conditions simulating both US and Saudi Arabian conditions. Thus, with a system operating at this site, the possibilities for technology testing and data acquisition are better than for any other studied.

SYSTEM OVERVIEW

Our solar powered desalination system design philosophy is based on the use of commercially available subsystems and components. The reason for this is that we want to maximize the reliability of the system and be able to meet all the time schedules. We have taken great care in identifying the major subsystems. All have a history of successful operation in the field. Our experience in designing both desalination and solar power systems, and the philosophy of our management allows us to choose the best available equipment on the market. We believe that this is a very important attribute in that we do not have a vested interest in the production and manufacturing of a specific subsystem or component. We are therefore not biased towards the use of any particular piece of equipment, rather we will select that which makes the most sense based on requirements and economic considerations and that which has the full backing of a reliable supplier.

Our technical approach to the design of a solar powered desalination system is based on an innovative use of commercially available equipment. Coupled with the characteristics of the chosen site we have designed a unique system that incorporates many features of interest to the SOLERAS program. Our approach is driven by several primary variables:

- Solar Radiation
- Winds
- Brackish Water Quality
- System Specification
- Minimum Cost
- High Technology Transferability

The radiation characteristics suggested the use of solar collectors, while the availability of wind implied the use of wind energy conversion system.

Our proposed concept is shown in block diagram form in Figure 1. The desalination system, shown on the right-hand side of the figure consists of two reverse osmosis (RO) units. The first unit produces the largest amount of product water using a lower operating pressure, while the second unit operating at a higher pressure, produces added product water from the reject stream of the first unit. Hydraulic turbines are used after the second RO unit to recover as much energy from the final reject stream as possible. Feedwater quality required by the RO units is provided by the pretreatment system which consists of lime softening, filtration, ion exchange softening and pH adjustment. The final reject brine is sent to the evaporation pond for disposal.

Power needed to drive the desalination system is obtained from a steam turbine-generator system and a wind energy conversion system. The steam required is obtained from a combination of two types of solar collectors-medium temperature and high temperature. The medium temperature collectors supply energy to the point where water can be evaporated into steam at a pressure suitable for steam turbine operation. The high temperature collectors superheat the steam to increase turbine efficiency. The steam turbine generator system output is augmented by the power produced by the wind energy conversion system. Appropriate storage has been included to ensure optimum system operation for minimum cost including sufficient electrical storage to allow system startup without the use of a backup energy source.

The system has many unique features which have been built into the concept to ensure maximum reliability and to demonstrate a wide variety of solar energy and desalination technologies. The system is designed in modular form so that many of its elements can function separately and independently when required. For example the desalination system can be operated over a range of capacities powered by wind, medium temperature solar collectors or a combination of

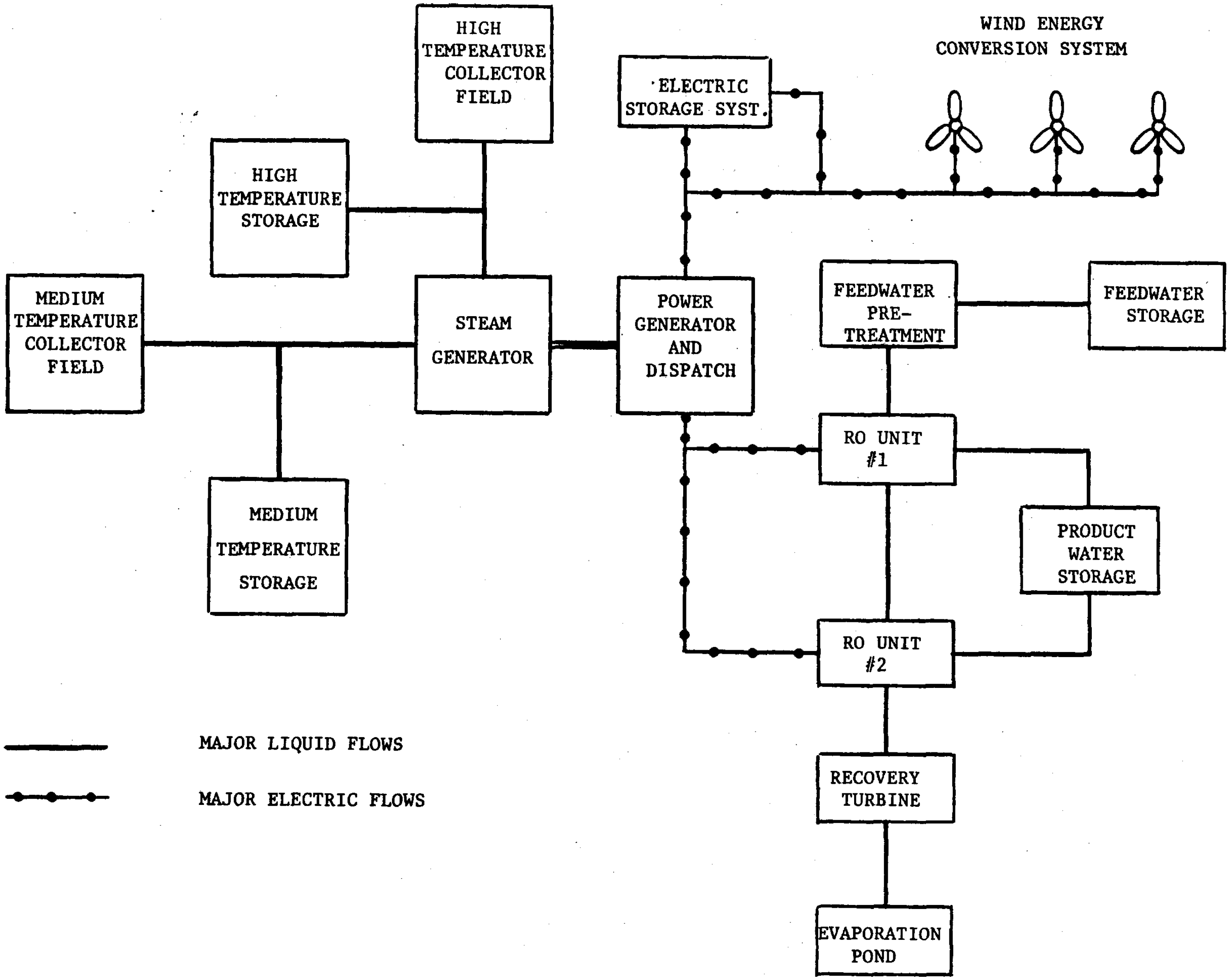

FIGURE 1 BLOCK TO DIAGRAM OF SYSTEM CONCEPT

medium and high temperature collectors. Any excess energy produced by the wind energy conversion system is stored in batteries or placed in thermal storage.

The system, as proposed here, would effectively be operating continuously with its instantaneous capacity dictated by the weather, amount of energy stored, and equipment status. The innovative features of the system can be summarized as follows:

- Both solar thermal and wind energy are used
- Two types of collectors are used to minimize system cost and to provide system availability
- A desalination system consisting of two stages of RO units
- Maximum product water recovery from the brackish feedwater using RO desalination
- Dual use of the land area for both solar and wind generators
- A configuration that can be adapted to a wide variety of feedwater qualities and solar resources
- A system design for a site typical of climatic conditions prevailing in areas of water need.

Economic Framework

The primary system performance specifications are defined by SERI in the request for proposal FJ-9-8123. These ground rules are particularly important in that they define feedwater quality and characteristics product water quality and design requirements. Table 2 lists the feedwater salinity as 6000 $^{\mathrm{mg}}/\mathrm{l}$ total dissolved solids (TDS) and specific chemical constituents. The feedwater temperature is shown as 38°C. Table 3 lists the product water quality as 500 $^{\mathrm{mg}}/\mathrm{l}$ TDS and also lists specific chemical constituents. Table 4 lists plant design requirements including; a nominal production rate of 6000 m^3/d(1.585 MGD), minimum product water recovery of 70%, and a plant life of 20 years. The solar energy fraction required for this design is 1.0 or 100%.

The basic objective of this project is to produce product water at the lowest possible cost. Aside from the performance requirements, economic requirements have also been specified to assure a uniform evaluation of the proposed design concepts. The cost factors are listed in Table 5. The base year for constant dollars is shown to be 1980. Significant factors are an interest rate of 8.6% and an annualized fixed change rate of 0.1437. The cost evaluation

TABLE 2 FEED WATER CHARACTERISTICS

(Ion concentration in mg/liter unless other units are specified)

	Conditions For System B
Total Dissolved Solids	6000
Calcium	500
Magnesium	75
Sodium	1500
Potassum	120
Iron	0.1
Manganese	0.1
Bicarbonate	690
Carbonate	0
Chloride	2000
Sulfate	1100
Nitrate	1
Phosphate	0
Silica	35
Fluoride	4
Suspended Solids	0.5
Color (PK)	-
Temperature (^{o}C)	38
Specific Gravity (unit)	1.010
pH (units)	7-7.5
Total Organic Carbon	-
Turbidity	-

TABLE 3 PRODUCT-WATER QUALITY REQUIREMENTS
(mg/liter)

	Inorganic Chemicals
Arsenic	-
Barium	-
Cadmium	-
Calcium	-
Chloride	250
Chromium (total)	-
Copper	-
Cyanide	-
Fluoride	-
Hardness	-
Hydrogen Sulfide	-
Iron	0.3
Lead	-
Magnesium	-
Manganese	0.05
Mercury	0.002
Nitrate (as N)	10
Selenium	-
Silver	-
Sulfate	250
Zinc	-
TDS	500

	Organic Chemicals
Endrin	-
Foaming Agent	-
Lindane	-
Methoxychlor	-
Mineral Oil	-
Phenols	-
Toxaphene	-
2,4 D	-
2,4 5-TP (silvex)	-

TABLE 4 PLANT DESIGN REQUIREMENTS

	Conditions For System B	Catalytic/SAI System Design
Minimum overall water recovery factor	0.70	0.90
Plant availability factor, minimum	0.82	0.82
Feed-water storage (days)	3	3
Product-water storage (days)	10	10
Fuel storage for backup power subsystem (days)	7	7
Minimum storage capacity for chemicals (days)	30	30
Plant design life (years)	20	20
Nominal product-water rate (m^3/day)	6000	6000
Yearly output of product-water (m^3)	1,800,000	1,800,000
Product-water usage rate, peak (m^3/h)	1,050	1,050
Product-water usage rate, average (m^3/h)	525	525
Product-water usage rate, minimum (m^3/h)	75	75
Solar energy fraction	1.00	1.00
Ratio of brine evaporation from pond to fresh water evaporation	0.7	0.7
Ratio of evaporation pond volume to brine volume	2	2

TABLE 5 CONSTANTS FOR COST MODEL

SYMBOL	DESCRIPTION	VALUE
N	System Operating Lifetime	20 years
k	Cost of Capital (and Rate of Return on Capital)	0.086
$CRF_{k,N}$	Capital Recovery Factor (8.6%, 20 yrs)	0.1064
g	Rate of General Inflation	0.060
g_c	Escalation Rate for Capital Costs	0.060
g_o	Escalation Rate for Operating Costs	0.070
g_m	Escalation Rate for Maintenance Costs	0.070
g_f	Escalation Rate for Fuel Costs	15%
y_b	Base Year for Constant Dollars	1980
y_{co}	First Year of Commercial Operation	1981
y_p	Price Year for Cost Information	1979
	Raw Land Cost	$\$1.25/m^2$
	Cost for Lined Evaporation Ponds	$\$25/m^2$
	Cost for Fuel Oil (31 GJ/m^3)	$\$157/m^3$
$\overline{FCR}$	Fixed Charge Rate, Annualized	0.1437
n	Accounting Lifetime	16 years
$1^{+} 2$	Insurance + "Other Tax" Fraction	0.020
	Investment Tax Credit	0.100
	Tax Rate	0.5
$DPF_{SD,k,n}$	Present Value of Sum-of-the-Year-Digits Depreciation	0.6376

methodology to be used represents a standard, consistant procedure for evaluation (1).

Although solar power systems represent proven technology, it must be appreciated that todays costs of solar equipment reflects a technology that has not yet fully matured. Using todays cost data for solar equipment in a application that would not be realized until 1985 would necessarily distort the relative sizes and capacities of the system components. The appropriate mix of components would be obtained when all technologies are mature. Since such cost data is not available, we selected 1985 solar cost goal for the reference plant design as listed in Table 6.

For the pilot plant, all cost data are to be obtained as 1980 cost estimates. Desalination cost data for the pilot plant and the reference plant design are based on 1980 cost estimates.

DESIGN CONCEPT

The proposed system resulted from an in-depth analysis of various desalination and solar energy conversion processes. This analysis consisted of seven components as follows:

- Process understanding - statement of process operation, performance and its energy and material requirements
- Development of screening criteria - project specified requirements, availability, cost, reliability and performance
- Screening of processes - based on the screening criteria various processes were ranked in order of feasibility
- Process Characterization - detailed evaluation of selected processes
- Systems analysis - based on cost and performance of the various systems
- System snythesis - matches the various project criteria to system performance in a cost effective manner.

The system rationale is described schematically in Figure 2. As a result of application of screening criteria several desalination and solar processes were selected. These promising candidate processes were then further analyzed as shown.

TABLE 6 INSTALLED COST OF SOLAR POWER SYSTEM COMPONENTS
(1980 Dollars)

	ENERGY RELATED[5]	POWER RELATED[6]	NOTES
Evacuated Tubular Collectors	\$100./m^2	---	1,4
Parabolic Trough Collectors	\$230./m^2	---	2,4
Point Focus Collectors	\$580./m^2	---	2,4
Wind Energy Conversion System	\$1,500/kWe	---	3
Thermal Energy Storage	\$40./kWht	\$45./kWt	1
Electrical Energy Storage	\$180./kWhe	\$400/kWe	3
Central Power System	\$420./m^2	---	2
Photovoltaic System	\$1,600./kWe	---	2

Notes:
1) Estimated, unofficial goals.
2) From "cost goals" committee.
3) Current (1980) cost.
4) Includes thermal transport and spares.
5) Costs relate to total energy collected or stored.
6) Costs related to energy collected or stored per unit time.

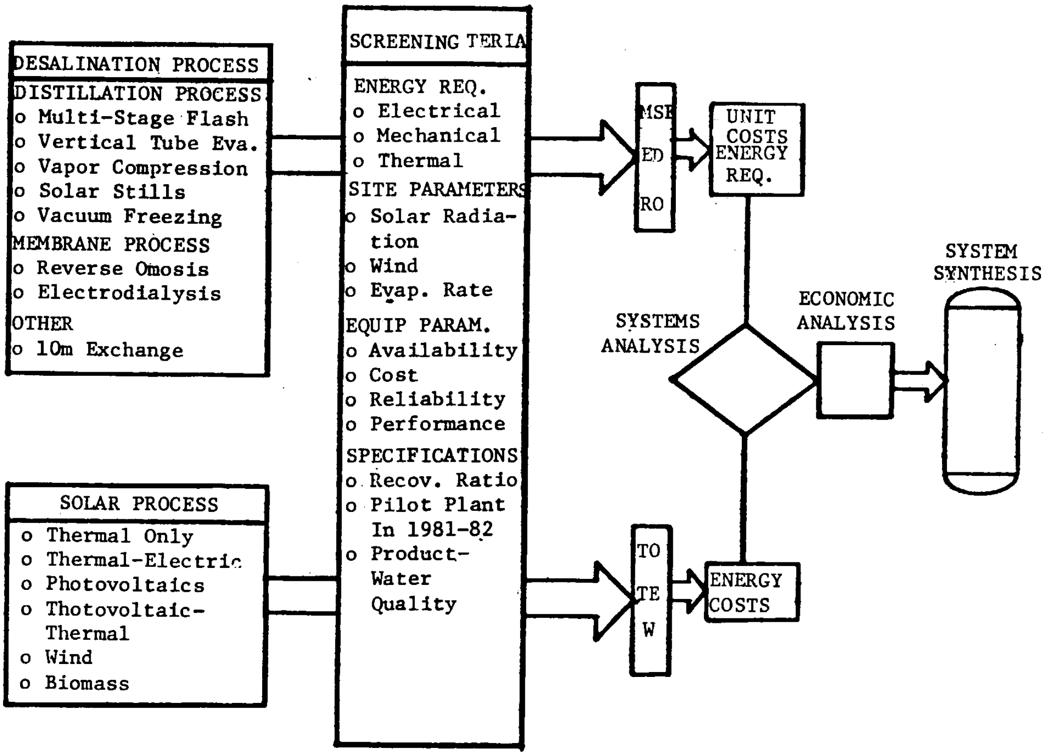

FIGURE 2 SYSTEM SELECTION RATIONALE

PROCESS SELECTION

The major processes were next analyzed by comparing them against a set of screening criteria. These criteria can be categorized into four classes:

- Class I - Energy required - three principal forms of energy are used - the various processes -- electrical, mechanical and thermal. Processes are best matched when their principal energy forms are identical.
- Class II - Site parameters - site parameters influence the type of desalination and solar system to be selected. Insolation, wind, and net evaporation rate effect system design.
- Class III - Equipment parameters - availability of tested and proven equipment; reasonable cost and history; reliable equipment for the 20 year plant life and predictable operation.
- Class IV - RFP specifications, primary consideration are: 70% product water recovery minimum; 20 year plant life; pilot plant construction 1981-82 time frame; product water to meet drinking water standards; and feedwater quality of 6000 $^{mg}/l$ TDS.

The first step in the process selection was the selection of the desalination system. Energy requirements of the various processes are a primary consideration due to the high costs of solar energy plants. Another aspect of the energy requirement is the form of the energy to be used. Electrical energy produced from a solar thermal-electric system requires four to five times the amount of thermal energy per unit of electrical output. By comparing the various desalination processes against the screening criteria, it was concluded that there were three viable processes:

- Multiple stage flash (MSF) evaporation
- Electrodialysis
- Reverse Osmosis

In order to select one particular process, a preliminary design of three desalination plants was made, each having a capacity of 6000 m^3/d. Price quotations were obtained from equipment manufacturers and the energy consumption for each process was determined. As a result of this comparison, the reverse osmosis membrane system was chosen as the desalination system having the lowest energy requirement and cost per cubic meter of product water.

Having selected the desalination process, the solar energy system was then analyzed to determine the most cost-effective means for converting solar energy into electrical energy. Biomass was eliminated on the basis that this resource is too inefficient and incompatible with the arid regions where the need for desalination exists. Photovoltaics were rejected on the basis of cost and low unit efficiency. The screening of solar energy conversion processes resulted in the selection of two viable processes:

- Solar thermal-electric
- Wind

The choice of a particular solar thermal-electric process was determined on a cost and performance basis to be a combination of point focus parabolic dish with an air receiver and parabolic troughs. Wind conversion systems were found to be applicable for the selected plant site. The last step in the process of selecting a system concept is to tie together the desalination and the solar energy processes in a configuration that can meet all of the design criteria.

DESALINATION SYSTEM

Feed Water Pretreatment

The feed water for the desalination processes is assumed to exist at the plant boundary limits. However, it should be noted that specific intake requirements will be incorporated for the selected plant site.

The pretreatment system consists of chlorination to maintain a residual of 0.5 to 1.0 mg/l of chlorine and storage to maintain a three day feed water supply. The following water treatment is required according to the specific requirements of the selected membrane supplier. In general, this treatment consists of the following:

- Addition of lime in conjunction with soda ash and/or ion exchange to remove calcium and magnesium. The extent of calcium removal is determined by the formation of insoluble calcium sulfate at high product water recovery. Silica is also removed in the coprecipitation of calcium and magnesium salt.
- Filtration with dual-media filters for the removal of suspended solids and colloidal matter.

- Addition of acid to prevent hydrolysis of cellulose acetate-type membranes and to maintain a negative Langlier Index which is a measure of a water's tendency to precipitate calcium and magnesium carbonates.

The lime softening pretreatment system consists of a solid contact reactor, dual-bed gravity filters and centrifuge. The reactors are designed for flash mixing and slow mixing of the raw feed water with lime and sludge removal all in one unit. The addition of sufficient lime to achieve reduction of the carbonate hardness present in the raw water increases the pH value to about 10.2. At this pH value, all of the iron is oxidized to ferric hydroxide and the manganese oxidized to manganic hydroxide for removal by sludge takeoff. To aid in agglomeration of the suspended solids into particles large enough to settle quickly, ferric sulfate is added as a coagulant and polyelectrolyte as a coagulant aid. The water from the reactor overflows to the filters for removal of any suspended matter remaining in the reactor effluent.

Prior to entering the filters, acid is added to the effluent from the reactor to reduce the pH value to about 8.0. This prevents any further precipitation from forming deposits in the lines and/or filter beds and results in more efficient filtration and longer filter runs between backwash.

The 5 percent sludge from the solids contact reactor is further dewatered by a centrifuge for disposal. The centrate from the centrifuge is recycled back to the reactor.

From the gravity filters, the water is pumped to the ion exchange columns for additional hardness removal. In the ion exchange process, the feed water is fed through a resin bed and all the cations are exchanged for sodium ions. Periodically, the resin is regenerated with a reject brine and sodium chloride solution. After ion exchange treatment, the feed water flows to the clearwell which has a retention time of 30 minutes. From the clearwell, the water is pumped through the cartridge filters to the reverse osmosis modules.

Reverse Osmosis

One desalination process is utilized to achieve the rated plant production of 6000 m^3/day of potable water. The primary desalting process is a two stage reverse osmosis (RO) membrane system which produces approximately 90% product recovery from the feedwater. The operating pressure for each stage is determined by the osmotic pressure of the brackish water and by the operating characteristics of the selected membrane type. Hydraulic turbines are provided for energy recovery.

The residual, concentrated brine and solids produced in the pretreatment processes are sent to an evaporation pond for final disposal.

The primary RO feed pump pressurizes the pretreated feedwater to about 2760 kPa guage for the first stage product water recovery. Chemicals, such as acid, sodium bisulfite and sodium hexametaphosphate, are injected to protect the membrane and prevent scale formation. The feed water then passes through a cartridge filter and then to the RO modules. The primary RO system is designed to reduce the dissolved solids to 500 mg/l in the product water at 70 percent recovery factor. Modular construction permits easy membrane replacement and convenient system expansion, as required.

The secondary RO system increases the product water recovery to a total of 90% by recovering product water from the primary RO system brine reject stream. The secondary RO system utilizes a high salt rejection membrane equivalent to a seawater RO membrane operating at a pressure of about 5520 kPa guage. System pretreatment requirements for various membrane manufacturers are shown in Table 7 for brackish feedwaters (2). The second stage RO reject brine stream is sent to the pretreatment system to regenerate the spent resin in the ion exchange process.

Evaporation Pond

Waste products are generated primarily in the pretreatment and desalination subsystems. Waste products include pretreatment sludges from the lime softening process and concentrated brine from the regeneration of the ion exchange resin. In addition, various cleaning and storage fluids are produced in maintaining the RO system.

The waste disposal system consists of a lined evaporation pond and the necessary components to ensure disposal in an economic and environmentally acceptable manner. No percolation of salts and water into the soil or ground water is allowed. The evaporation pond incorporates a ratio of brine to fresh water evaporation rate of 0.7 and a ratio of evaporation pond volume to brine of 2.0.

Subsystem Tradeoffs

There are two categories of subsystem tradeoffs; those made as the result of evaluations of each iteration of the systems analysis and those which are made to refine the basic cost models.

The first category is primarily related to the relative sizing, or capacity, among the subsystems. The second category includes changes in the system concept that may be made independently of the system analysis or may be tested in the system analysis to

Table 7
Pretreated Water Quality Requirements

Manufacturer	Cellulose Tri-Acetate Hollow Fine Fiber	Poly Amide Hollow Fine Fiber	Cellulose Acetate/ Cellulose Tri-Acetate Spiral Wound	Cellulose Acetate Spiral Wound
Suspended Matter				
a) Turbidity, JTU	1.0	N. S.	≤ 0.5	< 0.5
b) Plugging Index	≤ 4.0	≤ 3	≤ 4.0	N.S.
c) Plugging Factor, $\%P_{30}$	≤ 60	≤ 45	≤ 60	N.S.
Ionic Content				
a) Iron, mg/l	< 0.7	0.1 max.	≤ 0.5	No Limit Change Flushing Schedule
b) Manganese, mg/l	< 1.3	0.1 max.	≤ 0.5	No Limit Change Flushing Schedule
c) Strontium, mg/l	N.S.	15	N.S.	Depends on SHMP
d) Barium, mg/l	N.S.	0.1	N.S.	No Limit Change Cleaning Frequency
e) Silica, mg/l in brine	< 100	< 150	≤ 150	< 135 ppm
Organics	N.S.	N.S.	N.S.	N.S.
Chemical Additives				
a) Residual Chlorine, ppm	1.0	Zero	0.1 min. 0.5 max. $1.0 < 77^{\circ}F$	2ppm mx. for continuous service
b) SHMP, ppm				
1) Unsoftened	10-20	10 to 15	5-10	5-10
2) Softened	None	0 to 5	N.S.	2-5 in Brine
c) Acidification, pH				
1) Unsoftened	4-7.5	4 to 11	LI + 0.2	3-7
2) Softened		4 to 11	7.5	
Feed Temperature Max., $^{\circ}C$	30°	35°	35°	30°

Langlier Sat. Index ($CaCO_3$) . . . . < 0
Solubility Product ($CaSO_4$) . . . . ($< 1.9 \times 10^{-4}$)
ZETA potential, MV . . . . ($< 10^{-3}$ (SHMP))
0 to -30

N.S. - Not Specified
N.D. - Nondetectable

measure their effect on cost or performance. The tradeoffs in this category are described below:

Choice of RO membrane. In Phase I, a thorough review of all commercially available membranes was made, both for the high and the low pressure type. Each membrane was evaluated from the point of view of:

- Cost
- Reliability
- Operation under extreme conditions of high pressure and temperature
- Suitability for solar application
- Ease of operation and resistance to fouling .

RO Operating Pressure. Under normal temperature conditions, low salt rejection membranes operate at pressures typically around 3,500 kPa and high salt rejection membrane around 6,285 kPa. Most membranes will tolerate these pressures up to a temperature of 35°C. However, our design should be based on 38°C. The tradeoff is made considering the capital cost of the membrane and its replacement cost. The membrane replacement cost is based on estimated membrane life of 2 or 3 years.

Recovery Ratio of the RO System. The recovery ratio for the first RO stage is 70 % and 90% for the combined two stages. This is based on reasonable levels of operating pressures of 2,760 kPa in the primary train and 5,880 kPa in the secondary. The remaining 10 percent, the reject from the RO units, is returned to the ion exchange softener. The water recovery figures are based on the maximum achievable water production considering the brackish water salinity. During Phase I, a tradeoff study was made between the cost of the first stage RO modules, ancillary equipment, and electric pumping power on the one hand and that of the second stage RO modules, ancillary equipment and pumping power on the other hand. Another option was to use a single stage with a high salt rejection membrane.

Pressure Recovery Turbine. An energy recovery turbine to recover the pressure energy from the second stage reject stream is provided. During Phase I, a tradeoff study was made between the cost of the turbine and its associated piping and controls versus the cost of solar power subsystem saved by the recovery turbine.

Normally an energy recovery turbine would not be applicable for a reject flow after 90% product water recovery. However, the inherent high cost of solar energy has indicated the inclusion of the turbine for a cost-effective design.

Feedwater Pretreatment. Feedwater pretreatment has been designed to provide the quality of brackish water as required by the selected membrane and the selected product water recovery. Lime softening combined with filtration and ion exchange for extended calcium removal has been determined to be the most effective manner of pretreating the specified brackish feedwater.

Forced Air Cooled Condenser. The steam from the turbine in the solar power plant, is condensed in an air cooled condenser. This condenser is the final heat sink of the process. A forced air cooled condenser was compared to a wet cooling tower. A wet cooling tower requires makeup water in operation and contributes a waste disposal problem in the blowdown stream. The controlling variables are the average wet and dry bulb temperatures for Brownsville, Texas. As a result of the tradeoff study, the air cooled condenser was found to be the most economic choice.

Vertical Tube Evaporator (VTE). A tradeoff was also made to evaluate the impact of obtaining additional product water recovery from the RO reject brine stream. A crystallizing vertical tube evaporator and low temperature collector system was compared to the cost of solar evaporation. A credit was applied for the additional product water recovery at the cost derived for RO desalination. As a result of this tradeoff it was determined that the VTE and low temperature collector system was not cost-effective for higher product water recovery.

SOLAR POWER GENERATION

The solar energy system was designed to the meteorological conditions prevailing at Brownsville, Texas. The design constraint of a solar fraction of one, the availability of wind, and the less than ideal solar radiation characteristics, led to the choice of a combined solar collector-wind machine power generation system. The various system components are discussed briefly next.

Turbine Generator System

A conventional steam turbine generator system was chosen because of the mature state of the technology and its reasonably, established costs. An inlet temperature 400°C, was chosen to maximize turbine efficiency and minimize collector surface area requirement. The generator produces AC output, which is sent to the dispatcher where it is combined with the power produced by the wind machines and then prioritized to the various loads. The condenser chosen is of the air cooled type. The steam boiler consists of an evaporator, where energy collected by the medium temperature collectors is used to evaporate the water, and a superheater, where energy collected by the parabolic dishes is used to superheat the steam.

Medium Temperature Collectors

The medium temperature collectors gather the bulk of the thermal energy. This energy is stored in a rock-oil thermocline storage system and is used to vaporize the steam in the boiler. The heat transfer medium chosen in this system is SYLTHERM 800, a DOW Corning heat transfer fluid, because it is stable at high temperatures, non-corrosive, non toxic, and is in liquid form at all temperatures of interest.

High Temperature Collectors

The parabolic dish collectors provide energy at a high temperature, compatible with the upper ranges of the steam Rankine turbine systems. They are included to improve system efficiency. The ratio of medium to high temperature collector area is approximately 12 to 1. The heat transfer medium selected for the high temperature collector is air because it can be used over the complete temperature range that will be encountered by the system, it is readily available and environmentally benign. The disadvantage is that higher temperatures are required in the receiver and that power consumption for moving the air around is larger than for an equivalent liquid system. The air is not recirculated, it is drawn from the atmosphere at the receiver, passes through thermal storage to the superheater, preheats the water in the steam cycle, and returns to the atmosphere.

Wind Energy Conversions System

The wind machines convert available wind energy directly to electrical power. Because winds can be highly variable during a day, excess electrical power generated by the wind machines is dumped through resistance heating elements into a high temperature storage system, where its temperature is controlled to be compatible with the operating temperatures of the turbine. Electrical energy is also stored in batteries in order to smooth all power flows. The inverter is sized to handle the peak output of the wind machines.

Wind Loading on Structures

The selection of Brownsville, Texas as the site for the plant requires a careful examination of the extreme wind loads that can be expected on the equipment because of its proximity to the region of hurricanes. Curves of probability exceeding a given windspeed, and hence the mean recurrence interval, were prepared for Brownsville (3). Figure 3 shows these curves for the fastest mile and for windgusts at two elevations. These two elevations, 10m and 24m, are typical for the point focus collectors and wind turbines, respectively. Two design philosophies exist for selecting a design value. One is based on aerospace vehicle design principles that state a 10% risk of exceeding that design value for the given life time. The other one is based on building code standards,which accept a 63 percent risk for an expected life of 50 years. Reference 3 gives a set of curves describing the mean recurrence interval for risk of occurrence within the expected life of the system. For 10 percent risk over an expected system life of 20 years the mean recurrence interval is found to be 230 years; for 63 percent over 50 years, a recurrence interval of 51 years; for a 63 percent risk over 20 years recurrence interval is 26 years. From Figure 3 it is noted that the aerospace design philosophy would require a specification of very high survival windspeeds. The building code philosophy applied at a height of 24m corresponds to approximately a 65m/sec survival windspeed. It is interesting to note that the WTG Energy Systems, Inc. model MP 1-200 wind turbine has a 67m/sec (150 mph) survival wind speed specification (4).

Note that at the 10m level, the extreme wind gust would be 55 m/s (123 mph). However, for the stowed position survival windspeeds for the OMNIUM-G and GE-Shenandoah Concentrators are 35.6m/sec (80 mph) and 40m/s (90 mph), respectively. Thus, to survive the Brownsville environment would require the addition of windbreaks around the point focus collector field.

SYSTEM OPTIMIZATION

Methodology Description

It is obvious from the system description given above that there are many sizes or capacities of subsystems to be optimized. For the solar power system this task is particularly difficult due to the fluctuating nature of solar radiation and wind and the dependency of system performance on the history of both load and weather over the past several hours or even days. Solar systems are gener-

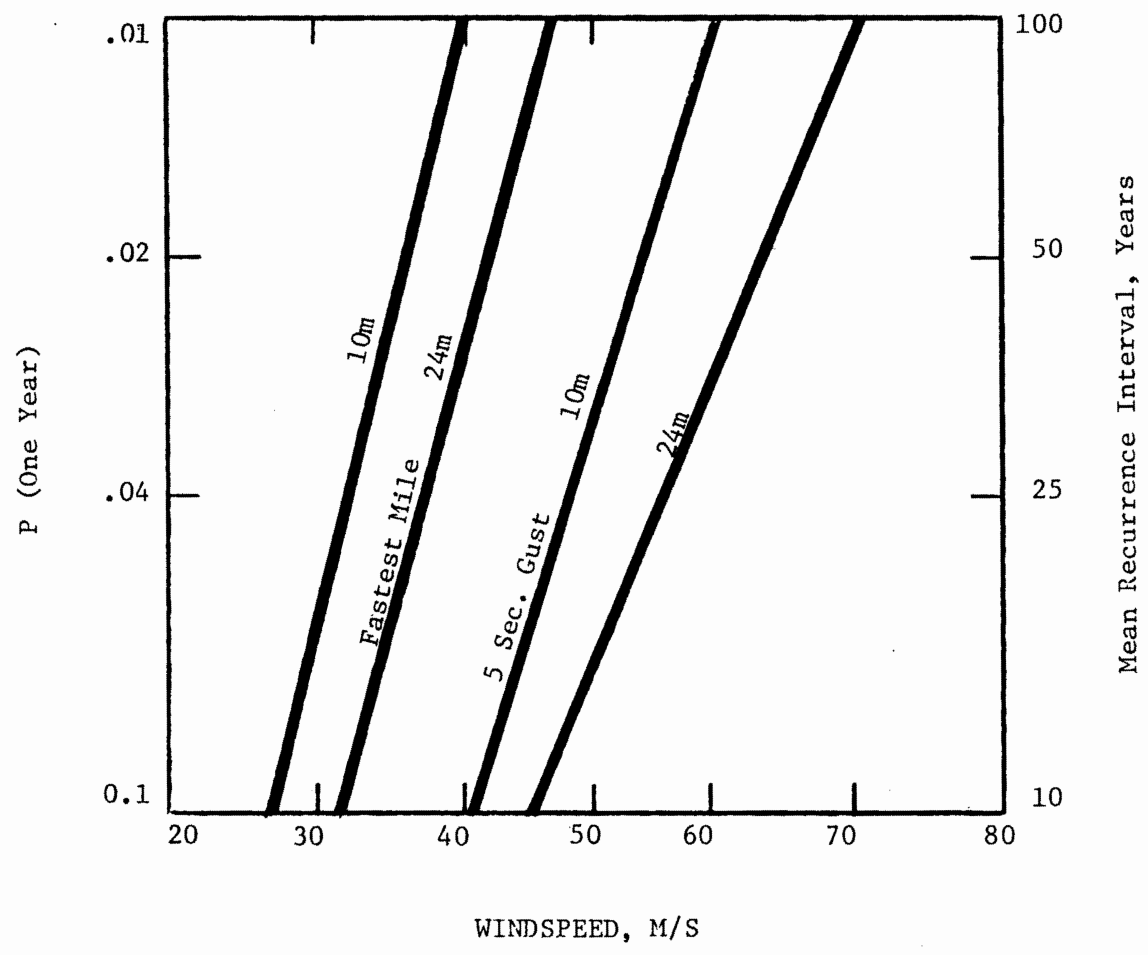

FIGURE 3 EXTREME WINDS FOR BROWNSVILLE, TX

ally seen as satisfying the load only partially leaving the remainder to be satisfied by a conventional backup system. Cost tradeoff analyses are made individually for each component of the system, hopefully to lead to a cost-effective system. However, many variables play a role in defining an optimum system. Frequently, engineering judgement is used to estimate the size, capacity, or operating point of a system, leaving only two or three other variables to a detailed sensitivity and cost tradeoff study. This approach is not feasible for the present system because a large number of variables are involved and very little previous design work has been done on systems requiring a solar fraction of one. The subsystems for which a size or capacity has to be determined are given in Table 8.

There are many methods available in the open literature that can be used to find the minimum or maximum of a mathematical function. For our problem we needed to identify an approach that would satisfy the following criteria:

- The methodology should be able to optimize many (more than 10) variables at the same time.
- These variables or functions of these variables may be subjected to constraining conditions,
- The procedure should provide good accuracy with reasonable computational requirements.

Table 9 provides one overview of the various optimization procedures that are available and shows their relative advantages and disadvantages. As noted in this table, we selected a linear programming technique to optimize the system variables. Although its principal disadvantage is that all performance characteristics have to be modelled by a linear function, the advantages appear to outweigh this shortcoming.

The objective of the optimization process is to minimize the system costs; including capital costs, operation and maintenance costs. The variables to be optimized plus the hourly energy and process flows are needed to define a control strategy. The constraints to be satisfied include the specified meteorological variables at Brownsville, such as solar radiation, wind, and ambient temperature, the equipment performance characteristics, the solar fraction of one, and a nominal daily product water output of 6000 m^3/day with a minimum plant availability factor of 82 percent. A block diagram of the mathematical model is shown in Figure 4 where all electrical energy flows are connected to node n_1 all thermal energy flows to node n_2, and the brine and water flows are as shown. Note that the dish and trough collectors are modelled as a single block, since both function with direct solar radiation

TABLE 8 SUBSYSTEMS TO BE OPTIMIZED

For the Solar Power System	For the Desalination System
1. High temperature collector area	1. Feedwater pretreatment capacity
2. Medium temperature collector area	2. RO desalination capacity
3. High temperature thermal storage capacity	3. Evaporation pond size
4. Medium temperature thermal storage capacity	
5. Number of wind energy conversion systems	
6. Electrical storage capacity	
7. AC-DC inverter capacity	
8. Turbine-generator capacity	
9. Resistance heater capacity	

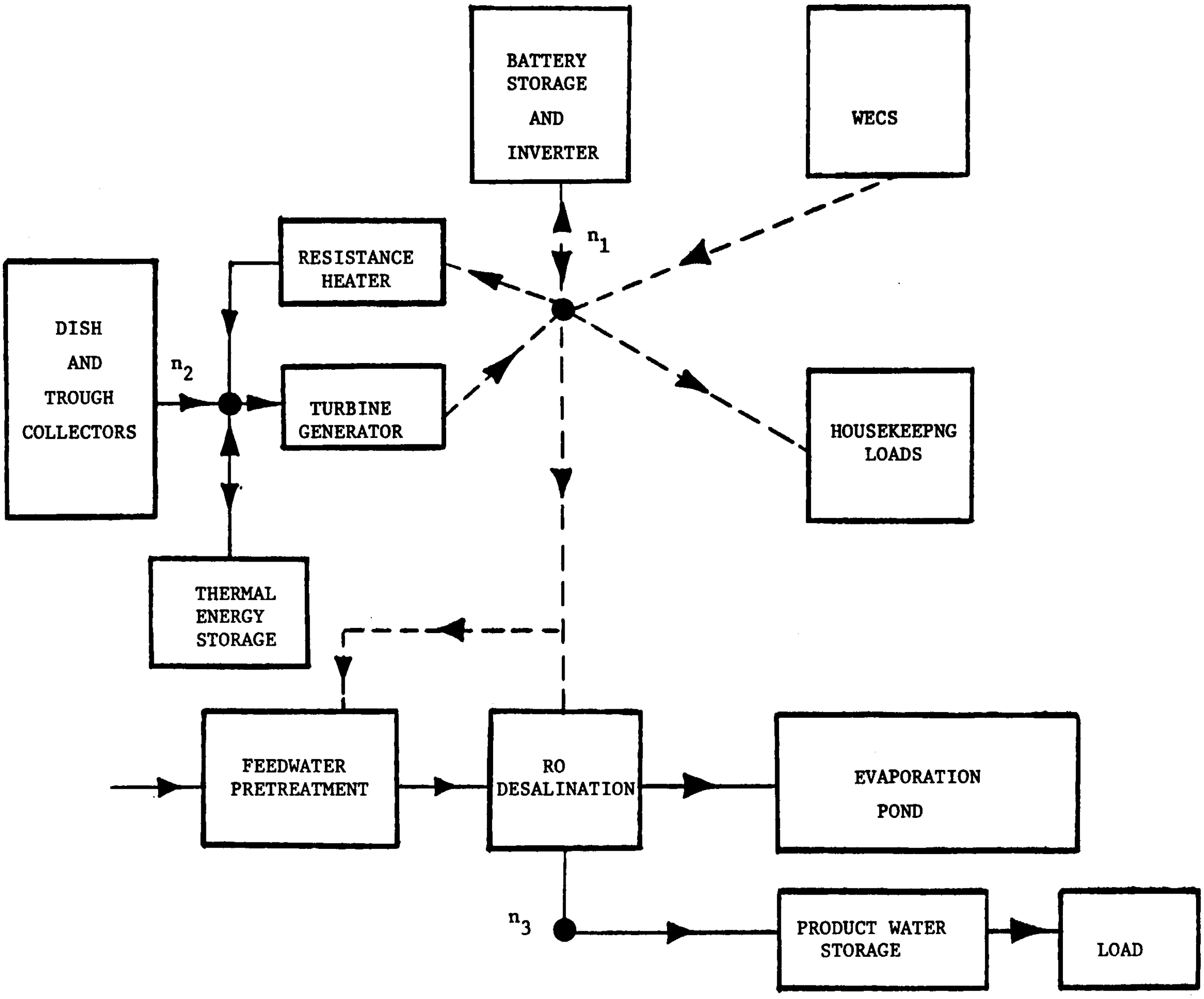

FIGURE 4 MATHEMATICAL MODEL OF THE SYSTEM

only and their relative sizes are fixed.

For each hour, the flow and energy balances were defined at each of the three nodes shown in Figure 4 in terms of the capacities and sizes of the various components. Constraint equations were defined for the total integrated annual product water output and a system control strategy that essentially maintains the RO operating level at a constraint fraction of its rated capacity each day. A matrix generator or pre-processor computer program has been developed that accepts as input the connection of various elements into a system, a meteorological data file, performance characteristics of each of the elements, and the cost functions of these elements. The output of this program is a file in a form appropriate to the linear programming solver APEX (5). Several simple problems were solved with this system and the results verified by hand calculations to ensure proper function of the methodology. A variety of calculations have been performed to investigate system performance and sensitivities to a variety of parameters.

Meteorological Data

The Brownsville site is not ideal from the standpoint of the available solar radiation or wind. It was, therefore, important to select an approximate meteorological data set. Fortunately, Brownsville was one of 26 U.S. sites where long-term solar radiation records were kept so that the Typical Meteorological Year (TMY) hourly meteorological data tape was available. The solar radiation and ambient temperature data on this tape are readily usable for collector performance analyses, but the wind data required additional processing before it could be used to evaluate the performance of wind energy conversion systems. Wind power calculations require windspeeds at the hubheight of the machine. This height is usually significantly larger than the height at which windspeeds are normally measured, the anemometer height. Winds at heights other than the anemometer height are often found from the power law, Equation 1.

$$V_H = V_A \, (Z_H/Z_A)^a \qquad (1)$$

The power coefficient, a, is often taken as 1/7, indicating a particular shape of the wind profile. It is now well known that the wind speed profile depends on many variables, the most important ones being the roughness of the surrounding terrain and the stability of the atmosphere. The appropriate power coefficient was computed for each hour of the day (6). For each Pasquill stability class, which ranges in value from 1 to 7, a second order polynomial was given on log Z_o, where Z_o is the surface roughness.

The Pasquill stability categories were derived from the meteorological data program of one of EPA's air pollution evaluation models, CRSTER (7).

System Evaluation

The optimization methodology as described has been completed and a variety of calculations have been performed to understand the various interactions of the system elements given different kinds of weather sequences and economic performance data. As is generally typical of solar systems we find that the economic optimum is very broad, meaning that different mixes of element sizes and capacities produce approximately the same system cost. The choice of a mixture of solar collectors and wind machines occurs when both resources are readily available. A surprising find was that under certain circumstances the optimization process selects wind machines only.

CONCLUSIONS

- The selected desalination process for this solar project is a two stage reverse osmosis membrane system.
- The product water recovery has been determined to be 90% for a brackish feedwater quality of 6000 $^{mg}/l$ TDS.
- Feedwater pretreatment required to achieve 90% product water recovery consists of lime softening, filtration and ion exchange softening.
- The selected solar energy process is a combination of solar collectors and wind energy conversion systems.
- The solar collectors are two types: high temperature, point focus collectors and medium temperature, parabolic trough collectors.
- The site selected for the baseline plant design is Brownsville, Texas. This site is representative of applicable sites in the United States and Saudi Arabia for solar desalination.

NOMENCLATURE

N	total number of components in system
V_H	windspeed at hubheight
V_A	windspeed at anemometer height

Z_H	height of the hub
Z_A	height of the anemometer
a	power coefficient in power law
Z_o	ground surface roughness

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DESIGN OF A SOLAR ENERGY POWERED FREEZE DESALINATION SYSTEM

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ABSTRACT

This paper presents an efficient and cost effective indirect freeze desalination process using the total solar energy concept. The integrated system consists of modular parabolic dish point focus solar energy collection, molten salt thermal energy storage, and a conventional steam turbine. The shaft power provides electrical energy needs via a generator and vapor compression energy needs via compressors for the system. The vapor compression and turbine exhaust heat provide for the refrigeration required for this unique freeze desalination process. Being a total solar energy concept, efficient energy utilization is of great significance. This is achieved by using high grade heat to produce compression refrigeration and low grade heat to produce absorption refrigeration. The ability of this system to make use of available energy to freeze sea water in an efficient manner makes it very attractive. The paper contains discussions on system advantages, future potential, and system optimization.

INTRODUCTION

The purpose of our project is to establish the technical and economic feasibility of large scale solar powered desalination of seawater.

With a subcontract from SERI under the SOLERAS Agreement, CBI in a joint effort with Foster-Miller Associates (FMA) is pursuing an indirect bulk freeze desalination process that is uniquely integrated with a stand alone solar power system.

The two most prominent items in this project are: 1) the solar power system being designed by FMA; and 2) the indirect freeze desalination system being designed by CBI. Thus, our paper is divided into solar power and freeze desalination sections. The solar concepts will be described first.

SOLAR POWER SELECTION CRITERIA

The combination of freeze desalination and solar power represents an opportunity to provide a solar baseload-type total energy plant. Such a solar power plant can be directly substituted for a fossil fuel plant. The baseline plant is nominally rated at 2 MW.

Selecting the solar power system generic concept involves consideration of the freeze desalination energy needs, desirable power system features and the longer term potential of the generic concepts.

Freeze Desalination Energy Needs

The freezing concept can effectively use both shaft power and waste heat in a cascaded refrigeration system. Compression refrigeration is used to drive the cycle between the melter and the freezer. Waste heat is used in an absorption cycle to aid the compression cycle heat rejection in the melter. The high thermal inertia of the freeze system suggests that the plant run continuously. The capital cost of a continuous duty plant will be less than that of a daytime-only plant. The daytime-only desalination system must be about three times the size of a continuous duty plant, while the solar power system cost is only a little lower for the daytime-only plant. For the continuous duty plant the freeze and solar power systems are about equal in capital cost.

The proportioning of energy utilization is relatively fixed by the optimum freezer and melter temperatures. The nominal process uses two-thirds of the shaft power to drive the refrigeration compressor. The freeze process pumps and overall utility loads consume one-fourth of the power. The average electrical power required for tracking collector drives and thermal fluid circulating pumps add the final 8 percent of the shaft power demand. Near term power conversion cycles at these levels reject sufficient heat at 88 C to satisfy the demand of the matching absorption cycle.

Desirable Power System Features

The generic solar power concept selection can be considered from two viewpoints. First, the concept should be practical and cost competitive for a desalination plant in the near term. Secondly, the solar power concept should have more range and broader applicability in the long term. The second view suggests investing in technologies that can grow in performance, satisfy a wide power range, and match a broad spectrum of processes.

These broad desires lead to more specific desirable features. A modular concept, like dish or trough mounted receivers, allows exact characterization at a pilot scale, increases availability through redundancy, duplicates the more expensive components to match different solar profiles or plant sizes, and adds to the production base of a nearly universal, standardized solar component. The SOLERAS time frame dictates that the solar system be founded in near term development and demonstration projects.

Generic Solar Power Options

The long term view of the solar power options can be seen through comparative studies by others. SERI recently completed a ranking of solar thermal electric plants ranging from 0.1 to 10 MW. From a list of 11 concepts, we have picked four that suit the freeze desalination process and are the most economically competitive.

The four concepts share the features of a central power plant, three provide continuous waste heat to the process, and all are currently under development or operational. The only near miss was the dish mounted Stirling engine; competitive in cost but lacking the centralized power plant feature and probably late in reaching the demonstration phase.

Table 1 presents some of the more important features of these applicable plants. The 5 MW size was closest to the current project, but the ranking was relatively insensitive to size. The highest capacity factor evaluated on a common basis was 0.7. The desalination plant design prefers even higher capacity factors and tend to favor the lower cost systems even more.

TABLE 1. BASIC SOLAR THERMAL POWER OPTIONS

Mature Technology - SERI/TR-351-46 August 1980

5 MW Electric, 0.7 Capacity Factor

Plant Type	Max Temp	Transport	Storage	Annual Efficiency	Capital Cost ($Millions)
Dish/Central Steam Turbine	510 C	Salt	Salt	16%	17.6
Heliostat/Central Steam Turbine	590 C	Salt	Salt	16%	17.9
Heliostat/Central Brayton	815 C	Air	Battery	10%	21.4
Tracking Trough/Central Steam Turbine	295 C	Oil	Oil	11%	21.3

The Dish/Steam concept uses a field of tracking parabolic reflectors to heat molten salt in individual receivers. A piping network transports the hot salt to storage. Salt is used to generate steam for a turbine. The collection system is modular and cost effective.

The Heliostat/Steam concept uses a field of tracking reflectors focused on a central tower mounted receiver to heat molten salt. The salt is stored and used to generate steam for a turbine. Modularity is somewhat compromised by heliostat and receiver specific design optimization for a given power level. The collector area is somewhat larger than a dish system due to shallow incidence angles on many of the field mirrors.

The Heliostat/Brayton also uses a field of tracking reflectors focused on a central tower mounted receiver. In this case, air is heated to directly drive a gas turbine generator set. Electrical storage rather than thermal storage is used as the high temperatures needed for reasonable Brayton engine efficiency cannot presently be stored economically. Electrical storage is relatively expensive and the engine waste heat is also not available continuously without additional storage equipment.

The Trough/Steam concept is similar to the Dish/Steam concept but is limited in cycle temperature by the concentration ratio and efficiency of a line focus system. Thermal transportation and storage are simpler with this oil based thermal transport and storage system but the low cycle efficiency at low temperature requires a much larger and ultimately higher cost solar collection system.

SELECTION AND ADVANTAGES OF DISH/STEAM SYSTEM

Technology readiness favors the Trough/Steam system as several plants are operational. The Dish/Steam concept lacks a molten salt receiver development program but is more conservatively represented in the Shenandoah system, presently on order, using a synthetic fluid at 400 C for transport and storage. The Heliostat/Brayton concept is presently somewhat earlier in the development cycle but no fundamental technical barriers are seen.

From these options, we have picked the Dish/Steam system for its modularity, broad application, and competitive parallel development programs. The demonstrated performance available for the 1983 pilot plant falls short of the mature system goals used in Table 1 due primarily to the missing small hot salt receiver technology. A later section discusses the growth opportunities available to work towards these goals.

INTRODUCTION TO FREEZE DESALINATION

Since freeze desalination is not yet a commercial process, we will digress here to explain the background and attractiveness of freezing.

Freeze desalination is based on the following natural phenomenon. Ice crystals formed by the partial freezing of a salt solution essentially contain only pure water. The salt remains in the residual brine.

Thus, a freeze process is directed toward the making of ice crystals within the brine, separating the crystals from the brine, washing the "contaminants" off the separated crystals, and finally, melting the washed crystals to produce pure water. All freeze processes have the above in common.

ADVANTAGES OF FREEZE DESALINATION

Freeze desalination has the following inherent advantages:

Low Energy Requirement

The heat transfer required to make a pound of steam is about 7 times that needed to make a pound of ice. The latent heat of evaporation for water is approximately 2.26×10^6 joules per kilogram (970 BTU per lb.) while the heat of fusion is only 3.35×10^5 joules per kilogram (144 BTU per lb.). This difference becomes ever more significant as fuel costs continue to escalate.

Negligible Pretreatment

The feed stream does not need treatment to prevent membrane plugging or to retard scaling.

Minimal Corrosion

Metal corrosion is exponentially reduced at the low temperature used.

Low Sensitivity to Fouling

As compared to distillation and membrane processes, fouling of heat transfer surfaces is insignificant.

Ability to Handle Varying Feed Conditions

Pure ice crystals are formed regardless of the feed used.

ECONOMIC INCENTIVE TO PURSUE FREEZE DESALINATION

Comparative analyses of membrane, distillation, and freezing processes in recent years have shown a very favorable advantage for freeze desalination of seawater.

A 1978 desalting state-of-the-art study by Fluor Corporation for the United States Office of Water Research and Technology suggested the following water costs for desalting seawater when computed on an equalized basis:

	$\$/m^3$	\$/1,000 gals.
Reverse Osmosis	1.13	4.28
Distillation (MSF)	0.99	3.75
Freezing	0.58	2.20

These costs were for 19,000 m^3/day (5,000,000 gals./day) plants. The absolute numbers do not reflect present energy or equipment costs; however, the relative rankings vis-a-vis freezing are still valid. The relationship of reverse osmosis to distillation may have transposed due to the escalation of energy costs, since distillation is a more energy intensive process.

INDIRECT FREEZING

In an indirect freeze process, a refrigerant absorbs heat across conductive barrier (heat exchanger) from a flowing salt solution, cooling the brine to its freezing point. It is most desirable and important that ice crystals form in the bulk of the liquid without growth on the heat exchanger surface.

PREVIOUS PROBLEMS WITH INDIRECT FREEZING

The use of indirect freezing for water desalination was an early Office of Saline Water (U. S. Dept. of the Interior) objective, which proved elusive for technical and economic reasons. One process which was investigated froze sheets of ice onto a heat transfer surface; the resulting high thermal resistance required excessive energy for heat transfer. In addition, these early studies did not explore low cost ice production means, relying instead on commercial ice making machines. The equipment costs were discouraging.

CBI'S BREAKTHROUGH

CBI has significantly advanced the status of indirect freezing by successfully demonstrating the feasibility of a simple indirect bulk freeze desalination system at its

Plainfield research facilities. This constitutes a major breakthrough which allows bulk crystallization under steady flow conditions without degradation of the heat transfer surface. It also represents a new direction which materially advances the scalability and reliability of freezing processes while retaining most of the generally recognized assets of freezing.

BASELINE SYSTEMS DESIGN

We will now describe our specific designs for the baseline commercial plant. As before, we will describe the solar system first and the freeze desalination system second. The solar system description will be further enhanced by a discussion of its future potential.

Solar System Design

The current design for the solar power system is constrained by the need to prove performance in a 1983 pilot plant. The available technology is again best represented by the U. S. Department of Energy's Shenandoah project. Table 2 highlights the performance parameters of the baseline plant.

The solar collection system uses aluminized tracking dishes to heat a synthetic fluid to 400 C. A piping network delivers the energy to a molten salt heat exchanger. The thermal tracking power and pumping losses consume 18 percent of the collected energy when the power conversion efficiency is accounted for in the tracking and pumping powers.

Thermal storage is accomplished in a two tank molten salt system. Salt storage is preferred over direct synthetic fluid storage because of the high fluid costs. Backup power is provided by a fossil fuel fired fluid heater in parallel with the solar collection system. Salt pumping power for storage and boiler feed uses 2 percent of the shaft power produced from storage. Because of the consistently high solar insolation levels specified for the SOLERAS site only about one half of the collected energy is stored each day.

Power conversion uses a hot salt to steam boiler. A steam turbine system drives an electrical generator and two refrigeration compressors. One 25 percent capacity compressor maintains high freeze plant efficiency at part load. Running the second, 75 percent capacity, compressor as well provides full load.

TABLE 2. BASELINE SOLAR POWER SYSTEM PERFORMANCE ESTIMATES

COLLECTOR PERFORMANCE

Collector and Receiver Efficiency (Aluminized Relector)	72%
Transport Thermal Efficiency	90%
Tracking and Pumping Efficiency*	91%
Net Collection Efficiency	59%
Receiver Outlet Temperature (Syltherm^{TM} 800 Synthetic Fluid)	400 C

THERMAL STORAGE PERFORMANCE

Storage Thermal Efficiency	95%
Storage Pumping Efficiency (Includes Energy Conversion Loss)	98%
Net Thermal Storage Efficiency	93%
Capacity For Full Power	13 hours
Storage Outlet Temperature (Partherm^{TM} 430 Molten Salt)	375 C

POWER CONVERSION PERFORMANCE

Ideal Cycle Efficiency		22%
Steam Turbine Efficiency		81%
Net Shaft Power Conversion Efficiency		18%
Steam Inlet Temperature		360 C
Steam Inlet Pressure	(203 psia)	1.4 MPa
Absorption (Condenser) Temperature		88 C

OVERALL PERFORMANCE

Solar To Net Power Efficiency	10%
At 25% Production Rate	8%
Net Capacity Factor (Soleras Solar Profile)	82%

*Includes Energy Conversion Loss

The site specific solar profile determines the collector field and storage sizes. Optimization of this sizing requires driving a matrix of systems through a plant performance model using a solar profile of each hour of the year. SOLERAS has provided such a solar profile for sea water desalination plants based on a Saudi Arabian site.

The load control logic used for this plant is based solely on the hot salt storage level. Only two steady state water production rates are used; 25 and 100 percent. Low load is selected when the hot salt storage is reduced to a predetermined level. Minimum power system cost was achieved when a 25 percent hot storage level load reduction point was used.

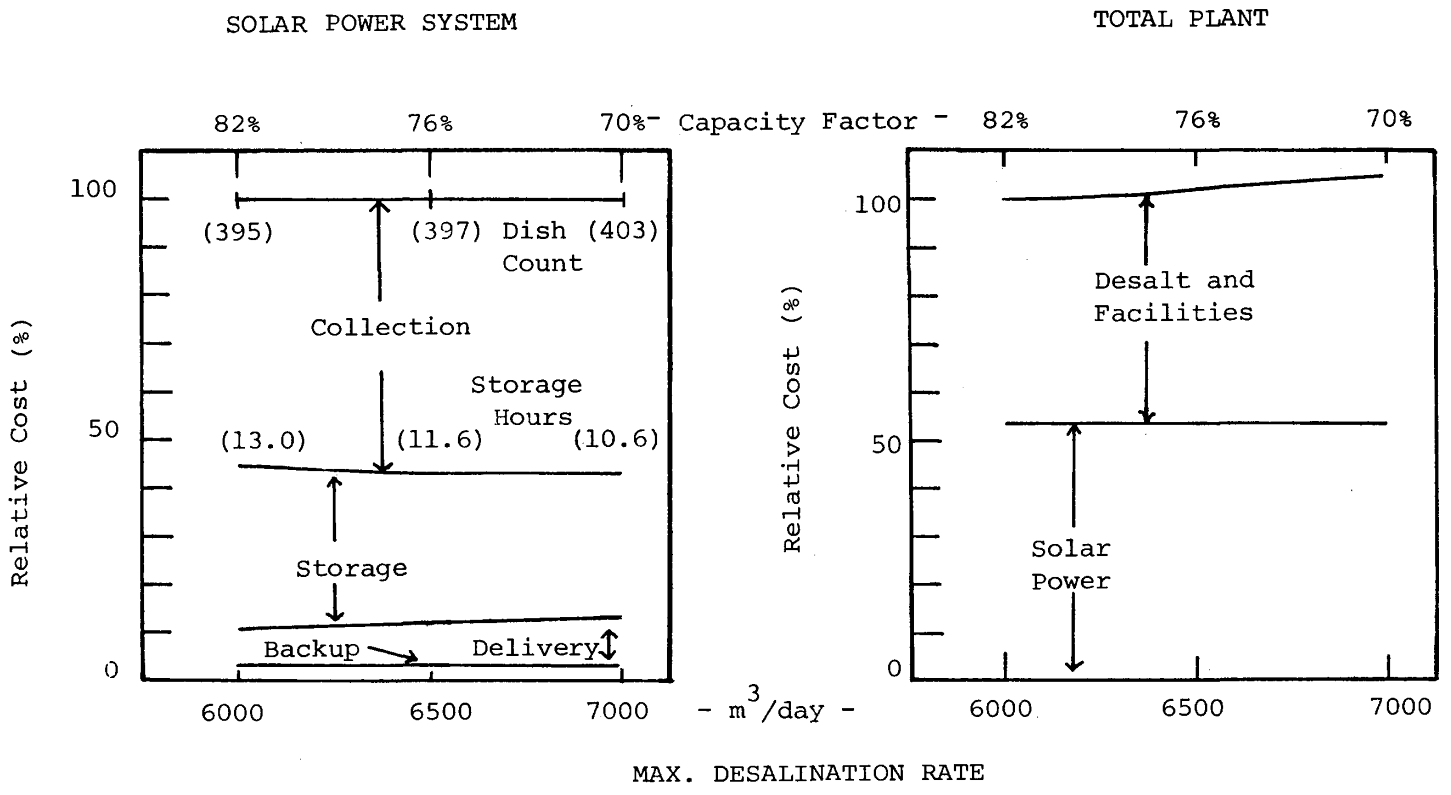
The system cost analysis is tentative at this writing, having completed only the first rough cut on each component. The preliminary results are presented in Figure 1 in a relative cost format. The absolute cost values are subject to change but the trends are believed to be accurate. Surprisingly, the solar power system total costs are insensitive to capacity factor. For a given annual water production the maximum desalination rate i.e., peak power demand has a minimal influence on power system cost.

The solar system must deliver about the same energy each day independent of when it is used during the day. The larger desalination plants need less storage as more water is produced when the sun is up. However, the larger plants run longer at part load at night. The power conversion efficiency at part load is lower due to a higher parasitic loss. On balance, the total energy collection required is nearly constant.

The total plant cost presently appears to increase slightly with maximum desalination rate. Refinement of these estimates may change the relative magnitudes and size sensitivity.

SOLAR POWER SYSTEM GROWTH POTENTIAL

The Baseline design using proven technology results in an attractive but conservative solar power system. Relative to the mature system capabilities of Table 1, the baseline design will require 60 percent more collector area. Given comparable volume production rate economics, the baseline system will not be directly proportional in cost. The mature system for example, uses a higher collection temperature with a stainless steel collection and transport loop and an internally insulated hot storage tank. The baseline is carbon steel throughout.



Average Annual Water Production: 4920 m^3/day
Runs at 25% Max Rate Below 25% Hot Storage
12m Diam. Aluminized Dishes
Storage Hours: Hours if at Max. Rate
Solar Profile: SOLERAS Pilot Plant, Saudi Arabia

FIGURE 1. PRELIMINARY BASELINE PLANT COST TRADEOFF

Improved performance is economically desirable either to reduce the solar system size or to retrofit the desalination plant for higher capacity as improved components become available for direct substitution. The sequence of improvements shown in Table 3 is chronologically likely.

The baseline turbine is typical of current conservative practice. Higher expansion efficiencies, addition of an extraction loop or perhaps reheat could exceed the 17 percent system efficiency improvement listed. This would directly reduce the number of dishes required.

Parabolic dishes with silvered glass reflectors, instead of aluminized films, have been tested and development of a comparable cost version is expected to follow aluminized types. The higher reflectance reduces the number of dishes, and also reduces thermal and pumping losses.

Progress in both higher temperature fluids and salt receivers suggest that a modest collector temperature increase is not far off. A 455 C system using either approach is attractive since carbon steel can still be used throughout the system. The 12 percent increase in overall efficiency might thus translate directly into a cost reduction.

The final step envisions 590 C molten salt receivers. New technology needs include: cost effective piping, tank insulation and freeze protection. The incentives are reduced collector area, much lower salt inventory, and elimination of an intermediate heat exchanger.

In conclusion, the performance of the dish mounted receiver, central steam turbine system can logically be expected to meet or perhaps exceed the long term projections shown earlier. The benefits of higher temperature operation will be the most difficult to achieve but they are also the most significant. The economic trends of these improvements are positive and perhaps obtainable for the future baseline plants.

TABLE 3 . SOLAR POWER SYSTEM GROWTH POTENTIAL

	Improvement	Increase	Cumulative	Net Solar Efficiency	Dishes Required
1.	Baseline, 400 C Oil	-	-	10%	400
2.	Improved/Extraction Turbine	17%	17%	12%	333
3.	Silvered Glass Dish	22%	43%	14%	286
4.	455 C Oil and/or Salt	12%	60%	16%	250
5.	590 C Salt (instead of 455 C)	29%	86%	18%	222

DESCRIPTION OF FREEZE DESALINATION PROCESS CHOSEN FOR THIS SOLAR PROJECT

An artist's view of the solar energy water desalination plant is presented in Figure 2. A block process flow diagram shown on Figure 3 further delineates the system.

Incoming seawater is cooled to approximately + 1.5 C by the outgoing cold brine and cold product water in two separate heat exchangers. The seawater then flows into the separator vessel.

In the freeze exchanger, recirculating ice-brine slurry with feed seawater and recycle brine added is cooled to its freezing point. Both ice fraction and crystal size increase as the feed stream from the separator progresses through the freeze exchanger.

The return flow from the freeze exchangers enters the separator. The separator has several functions: feed supply, ice fraction control, crystal growth, and initial brine separation.

A high ice fraction slurry is pumped from the upper portion of the separator to the counterwasher. This slurry enters the lower portion of the counterwasher where the bulk of the brine separates by gravity. The brine leaves the counterwasher via small drain holes. A portion is recycled to the separator. The remainder becomes the outgoing reject brine. The ice crystals move upward as a porous bed in the counterwasher. Wash water is sprayed downward into the bed to displace the residual brine from the crystals. A rotating mechanical scraper removes the ice from the top of the column.

The ice then cascades into the melter. Here, the ice is melted by heat from condensing refrigerant from the freeze exchanger refrigeration system. Part of the melted ice will be used as wash water in the counterwasher, with the remainder becoming product water.

We have optimized the indirect freezing process by utilizing two refrigeration systems. The primary system is a standard mechanical compression cycle with ammonia refrigerant. The compressed ammonia gas flows into a coil in the melter where it condenses to the liquid state while releasing its latent heat to melt the ice. The liquid ammonia then expands adiabatically through a throttling valve into the freeze exchanger. Heat is absorbed from the ice-brine slurry circulating through the freeze exchanger, vaporizing the ammonia. The ammonia vapor returns to the compressor, completing the cycle.

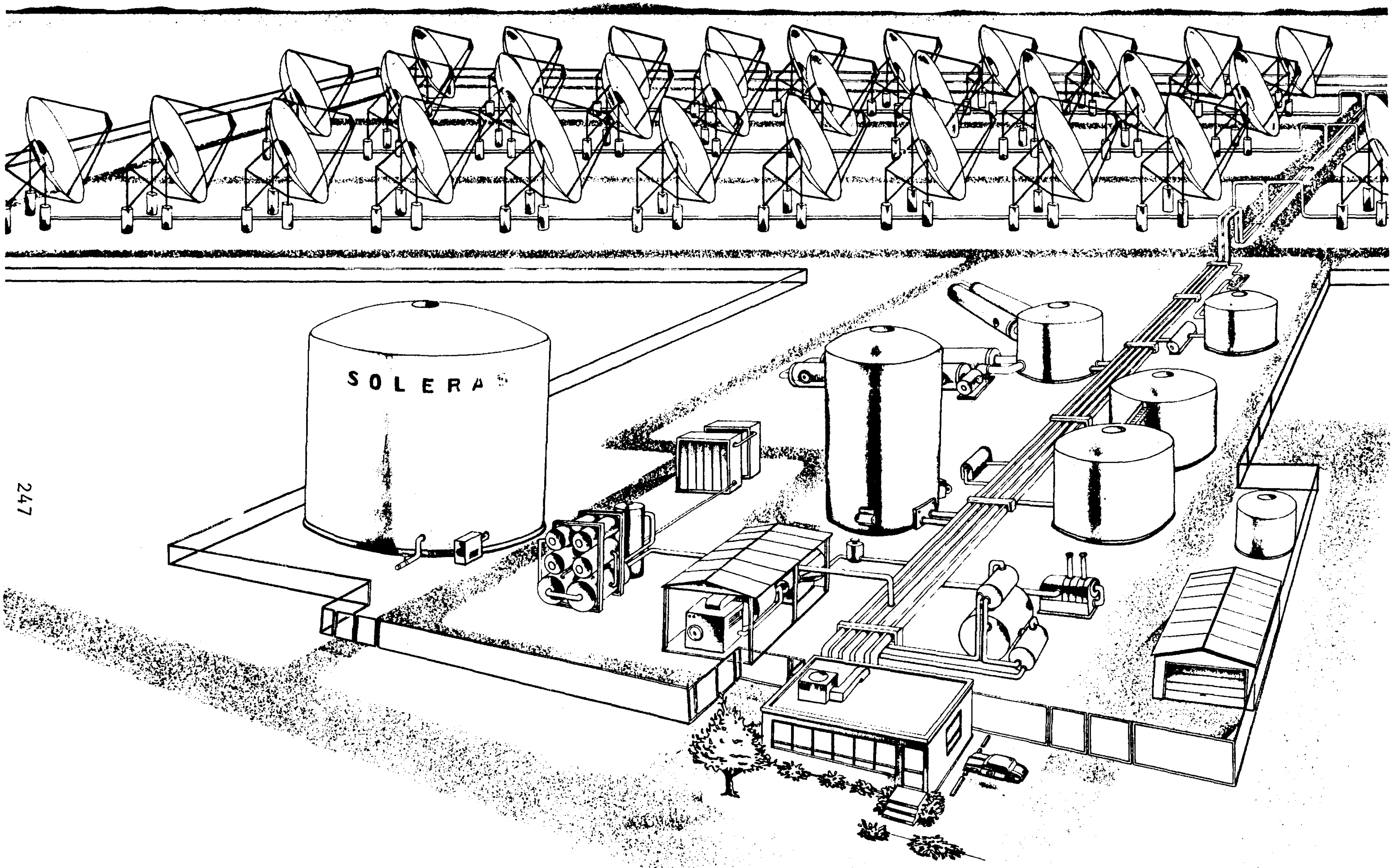


FIGURE 2. SOLAR ENERGY WATER DESALINATION SYSTEM

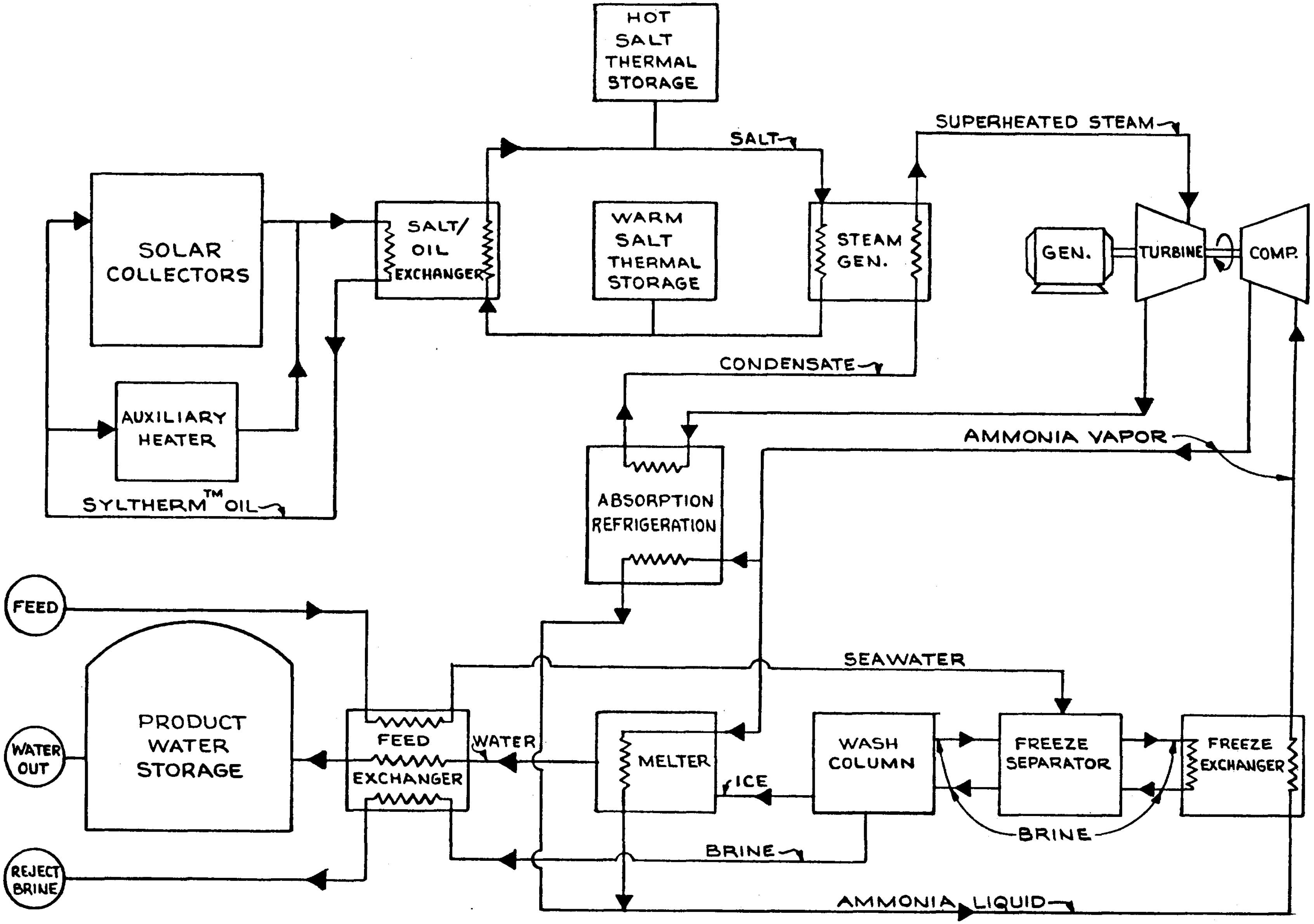


FIGURE 3. SOLAR ENERGY WATER DESALINATION SYSTEM

CBI

However, there is an inherent mismatch between the freezer and melter heat duties. Melting of the ice is insufficient to condense all of the ammonia vapor. Thus, an auxiliary refrigeration system is required to condense the excess vapor.

We plan to use an aqua ammonia secondary absorption cycle as our auxiliary refrigeration system. Exhaust steam from the solar plant electric power generation will drive this absorption refrigerator.

SYSTEM OPTIMIZATION

The total integrated system for seawater desalination is subdivided into distinctly identifiable segments called subsystems. Each subsystem performs certain tasks in making product water from seawater. Thus, optimization of each subsystem, group of subsystems, and their interrelations lends to optimization of the entire system. The system optimization is made based on economics, system and subsystem efficiencies, technology maturity, commercial availability, and ease of maintenance.

The optimization effort usually starts at a critical subsystem component or at a critical functional requirement of part of a subsystem. The optimization is achieved when the effect of variation of the critical item on the entire system is analysed. One broad application of optimization between groups of subsystems may be reviewed by looking at Figure 1. Many other system parameters were optimized to achieve a totally integrated system. For example, selection of aqua ammonia absorption refrigeration over other absorption refrigeration systems, matching low grade heat requirements for absorption refrigeration with turbine heat rejection, establishment of product water recovery ratio, matching number of solar collector dishes with heat storage capacity and product water generation rate, and many others.

ACKNOWLEDGEMENTS

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DESIGN OF A COMMERCIAL SOLAR-POWERED HYBRID DESALINATION PLANT FOR SEAWATER AT YENBO, SAUDI ARABIA

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ABSTRACT

The conceptual design of a commercial solar powered 6,000 m^3/day plant for desalting seawater is described and criteria used in the design and selection of equipment are discussed. The layout of the plant and the process flow diagrams are given with indication of the major performance specifications and a listing of important design parameters. The plant is designed for a generic site of characteristics common to specific sites in Saudi Arabia and the U. S., however, the design can be used for other sites in other countries. Modularity of the design also allows scaling the plant down to lower capacities appropriate for a pilot unit or for areas of low demand. Several measures have been taken to maximize the use of solar insolation and to reduce power consumption. Those include energy recovery, utilization of waste heat, and operating the desalination plant at low flow rate when the solar insolation is low. The plant design was optimized to give the lowest water product price and trade off studies are made to provide high reliability and low cost. A special simulation computer code is used for the cost analysis and for fine tuning the design parameters within the constraints of the performance specifications provided by the SOLERAS program.

INTRODUCTION

The use of solar energy to assist or to drive a reverse osmosis plant has been examined [1]. In one concept, a solar assisted plant is designed with parabolic concentrators to drive a stirling engine. A 72% HaF/25% MgF_2 eutectic salt mixture is employed as a thermal storage medium [2]. Economic analyses are conducted for a wide range of capacities, solar insolation, and solar fraction of the energy required. For small communities in isolated areas, a wind-powered RO system has been designed with energy recovery [3]. For the same application, an integrated system was designed combining a photovoltaic (PV) concentrator that produces DC electric power and collects thermal energy with an electrodialysis (ED) desalting plant [4]. The feedwater is heated in order to reduce the ED power consumption.

As a result of those designs, the combination of RO and ED in design of a plant for sea water desalting is found to be attractive. This concept has been developed further under the SOLERAS program for solar energy water desalination system design. The AC power is to be generated for the reverse osmosis first stage using a solar thermal power system based on line focusing thermal collectors and Rankine cycle power conversion. The DC power for the electrodialysis second stage is to be produced directly from a solar photovoltaic field. Although both the ED and RO desalting concepts are individually inherently more adaptable to solar energy than vapor phase processes because of their energy efficiency, there is considerable merit in the hybrid concept as applied in this design over the choice of either concept for the total desalting process.

The energy intensive nature of desalting processes has been principal in the selection of the hybrid design considered. System components have been selected to provide high degrees of reliability and efficiency in operation to balance the energy requirements and allowances for cost effective water production. Line focusing parabolic trough collectors provide thermal fluid at temperatures close to 300°C for favorable ORC engine efficiency. Steam cycle requirements for pressurization and superheaters under similar conditions are bypassed through the use of an organic cycle turbine. Flat plate photovoltaics provide direct conversion DC power for ED operation, deleting AC to DC inverter losses. In addition, sensible heat thermocline storage capacity is utilized to extend the desalting plant operating day and thereby reduce overall plant capacity and cost.

OVERALL PLANT DESIGN

Figure 1 shows an artist's conception of the commercial solar powered seawater desalination plant under design by the DHR, SAI, IONICS, and Al-Redwan team. As shown in the sketch, the plant comprises three major areas, the largest of which is the solar thermal (ST) collectors, then the photovoltaic collector field, and thirdly, the desalination plant facility that houses the RO unit, the ED unit, the pretreatment unit, the control room and other auxiliary systems. A cutaway shows some of the details of the subsystems enclosed within the facility. Also shown are the underground product water storage, thermal energy storage, energy delivery system, and feedwater well.

The process flow chart of the plant is shown in Figure 2 which delineates the various subsystems. Here, a separation is made between the RO, which operates as a first stage to reduce the salinity from 44,000 ppm to about 1,500 ppm and the second stage ED, which desalts the blend of the low salinity water and pretreated seawater to a drinkable water quality. Both stages are connected by the intermediate tank that stores the RO product water and feeds the ED unit. However, each stage draws on its own energy source. The high pressure pumps of the RO are driven by A-C from an organic Rankine cycle (ORC) turbine which receives thermal energy from a line-focus parabolic trough field. The ED stage draws D-C directly from the photovoltaic field. The two stages are decoupled to provide a high degree of flexibility of operation, that is, one stage can operate in case of outage of the other stage. Provisions are made to heat the ED feed to reduce power consumption in this stage and to recover the energy from the blowdown of the brine from the RO stage. Proper pretreatment and post-treatment subsystems are shown.

Table 1 lists the major performance specifications for a 6,000 m^3/day

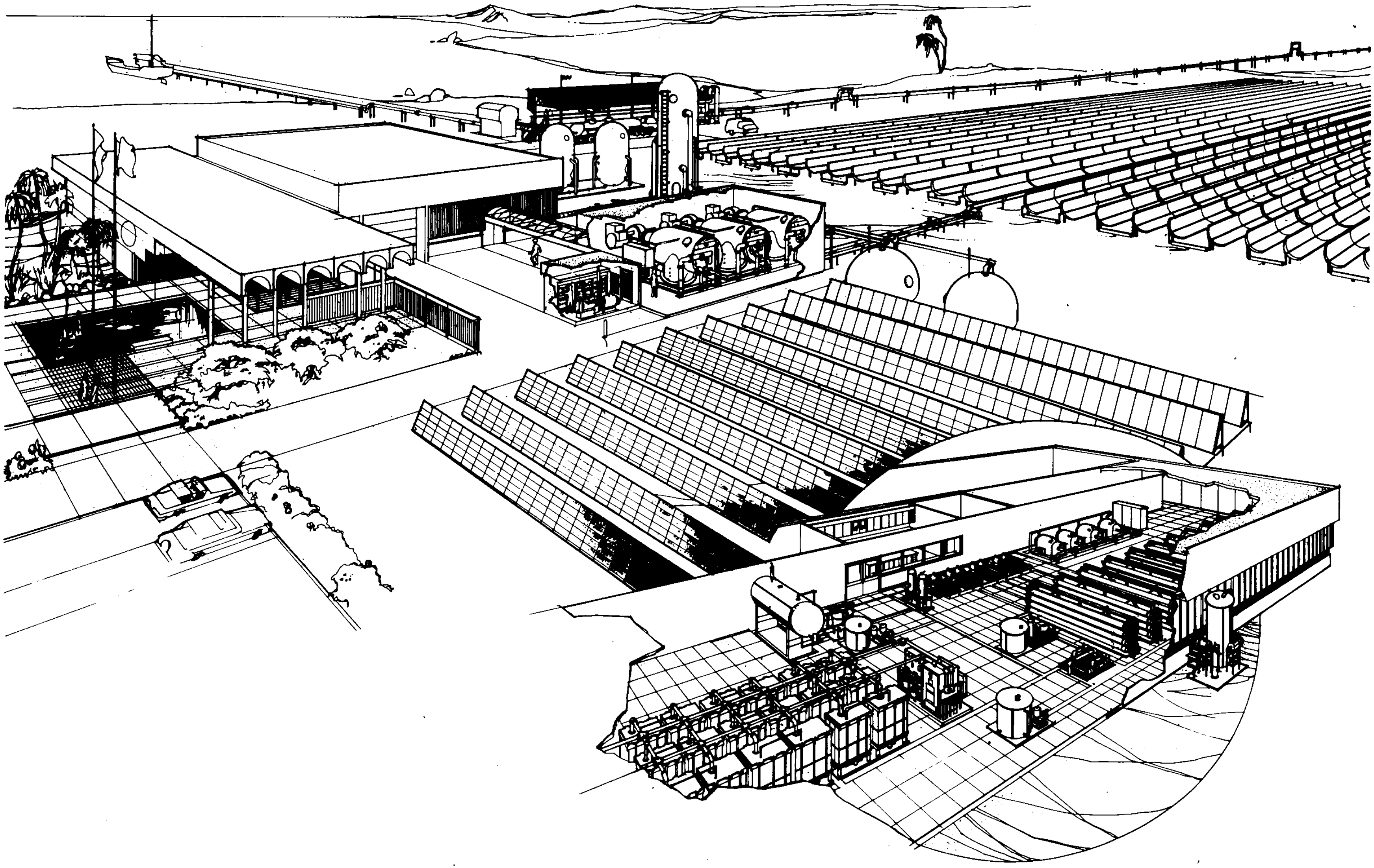
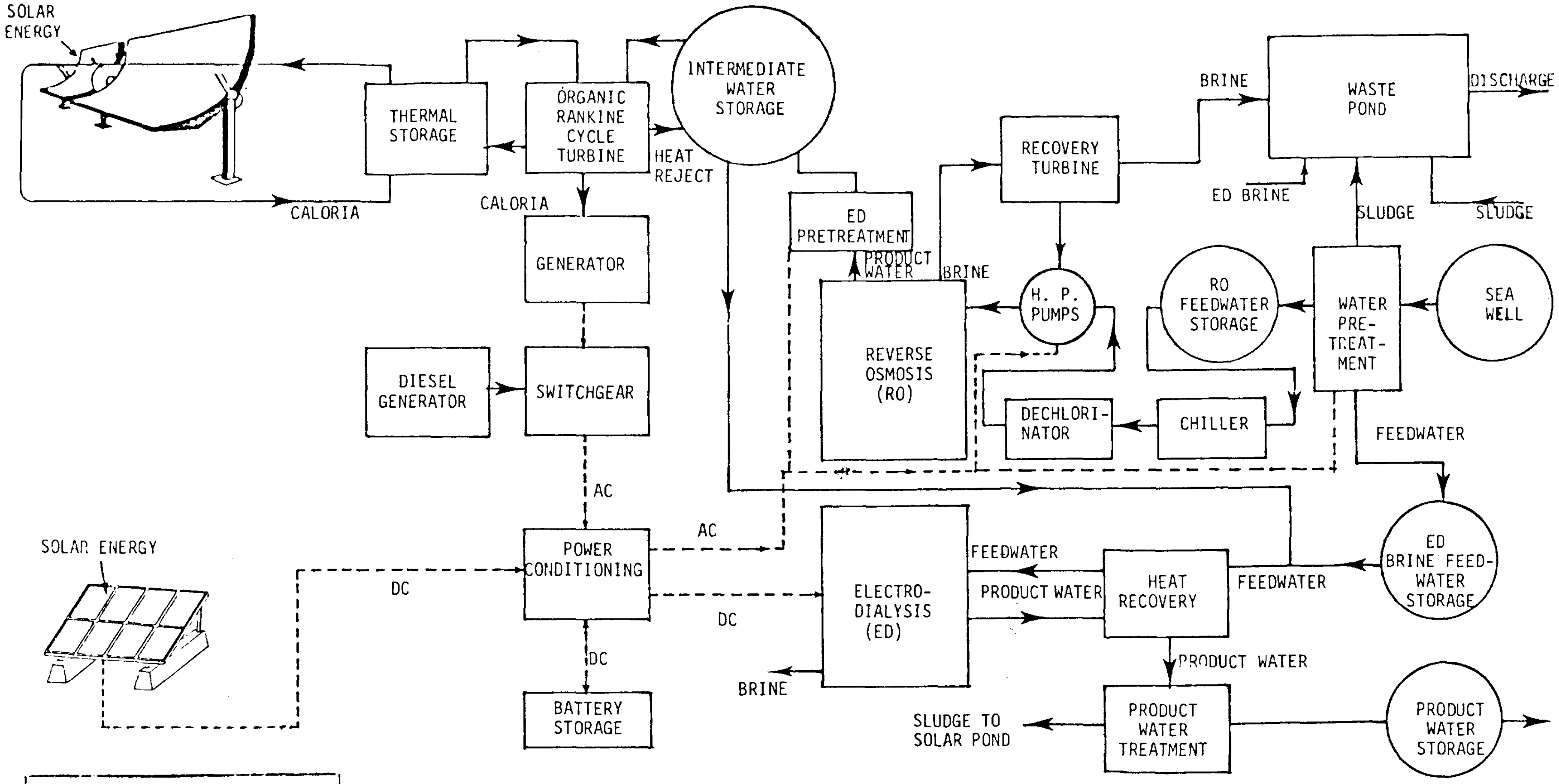


FIGURE 1. ARTIST'S CONCEPTION OF A 6,000 m^3/day SOLAR DESALINATION PLANT



DHR, Inc. Ionics
SAI-Ames Al Radwan
February 3, 1981

FIGURE 2. PROCESS DIAGRAM FOR SOLAR-POWERED DESALINATION PLANT

plant. The design of the plant is such that it can be scaled down or up to other sizes due to its modularity. Since a scaled down version of the plant is to be sited on the Red Sea at Yenbu, the Saudi Arabian codes and standards for buildings, electrical wiring, mechanical equipment, and fire protection are used, (Table 2). Where codes or standards are lacking, U.S. design and construction codes are used, (Table 3), such as those specified by the ACI, AISI, ANSI, ASHRAE, ASME, ASTM, FS, IES, NFiPA AND UL. Also, BOCA, ICBO, national fire and national electrical codes are used except in situations wherein the standards or codes are limited to specific situations or regulations unique to the USA.

TABLE 1. MAJOR PLANT PERFORMANCE SPECIFICATIONS

	CONDITIONS
Minimum overall water recovery factor	None
Plant availability factor, minimum	0.82
Product-water storage (days)	10
Fuel storage for backup power subsystem (days)	7
Minimum storage capacity for chemicals (days)	30
Plant design life (years)	20
Nominal product-water rate (m^3/day)	6,000
Yearly output of product-water (m^3)	1,800,000
Feedwater Storage (days)	3
Solar energy fraction	1.00
Sea Feed water total dissolved solids (mg/liter)	44,000
Sea Feed water maximum temperature (^oC)	35
Product water maximum total dissolved solids (mg/liter)	500
System feedwater	Sea Well

TABLE 2. PLANT PERFORMANCE SPECIFICATIONS
(MINISTRY OF PUBLIC WORKS & HOUSING)

ACTIVITY	SAUDI ARABIAN STANDARDS
Architecture, construction and sanitation	General tech specs, Saudi Arabian Codes and Standards, Sec. 3 Excavation, Sec. 4 Concrete, Sec. 5 Construction, Sec. 6 Sewage, Sec. 7 Metalwork, Sec. 16 Painting, Sec. 17 Landscape
Cooling	Tech. Specs. Air Conditioning and Elevators Secs. 1, 2, and 3
Electricity	Tech. Specs. Electrical Works Schedules 1 through 7

TABLE 2. PLANT PERFORMANCE SPECIFICATIONS (Cont'd)

ACTIVITY	SAUDI ARABIAN STANDARDS
Materials and dimensions	Unified codes for quantities in construction: Structural, Architectural, Sanitary, Mechanical, Electrical, and Decorative Works.

TABLE 3. PLANT PERFORMANCE SPECIFICATIONS, U.S. CODES AND STANDARDS

PERFORMANCE REQUIREMENTS	CODES/STANDARDS
Flammability	Class I (0-15) ASTM E 84, ASTM E 13G
Smoke Production	ASTM E 84
Toxicity	NFiPA
Fire Endurance	ASTM 119
Security	Fence and Computer Controlled Monitoring System
Misuse Resistance	Easy use of controls - fail safe, inhibitors for incorrect operation or sequencing
Static Load, Safety Factor	5
Dynamic Load, Safety Factor	5
Seismic Force	0.2 deadload of equipment
Corrosion Resistance	ASTM B 117
Wind Resistance	70 m/sec
Thermal Resistance	-10 to 70°C
Noise level, not to exceed	60 dBA
Environment, Temperature °C	25
Humidity, %	50
Lighting	ANSI A 11.1
Electrical Wiring	NEC C1 D2
Control Room	Applicable MIL-HDBK-759, FED-STD-186, FED-STD-595

Design Philosophy

The major objective of the conceptual design is to provide a commercially viable solar-powered stand-alone seawater desalination plant that produces water at a minimum cost that is likely to compete with conventional systems as the price of fossil fuel increases and as solar technology advancement improves solar collector efficiency with resultant reductions in cost in both the US and Saudi Arabia. To achieve such objectives, several criteria are used in the design, including:

- High plant availability
- Low power requirements

- Maximum utilization of solar insolation
- Low capital cost
- Universality of the design
- Scalability/modularity of the design
- Compatibility with conditions in the U.S. and Saudi Arabia
- Use of commercially available equipment

High plant availability can be assured by the use of reliable components, providing routine maintenance at times of no solar insolation, avoidance of total forced shutdown of the plant by making partial operation possible, the use of proven and rugged equipment and systems which are reliable in hostile environments, and minimization of maintenance requirements. Review of operation experience records of desalination plants in Saudi Arabia shows that the major causes of shutdown and long maintenance and repair periods are: frequent failure of diesel-generators used to power the plant, poor sea intake due to low sea level, marine growth or oil contamination, and membrane failure due to bursting of brine hoses in RO plants. Solar power systems may be more reliable in operation than diesel generators. The use of a sea well with proper precautions could alleviate the intake problems. Proper selection of hose material would prevent frequent membrane failure.

To meet the second criterion is to select desalination processes of low power consumption. From among the proven technologies of water desalination, RO and ED are more energy efficient than distillation processes. Although freezing has the least power requirements, the technology has not yet reached the commercial stage. Also, the ability to operate at varying power levels is possible in the case of ED. Segmenting the RO units into sections with separate high pressure pumping capacity may allow for flexibility in operation. In contrast, vapor phase and freezing systems are generally unable to respond to high degrees of variability in available power. The flexibility in operation is essential in providing maximum utilization of solar insolation which varies during the day with minimum use of storage. Table 4 displays a comparison of desalination technologies. Such comparison provides the basis for preference of the hybrid RO/ED system as a desalination process.

TABLE 4. DESALINATION TECHNOLOGY COMPARISON FOR SEAWATER PLANTS

PROCESS	COMMERCIAL AVAILABILITY	EFFICIENCY (UNIT OUTPUT FOR SEAWATER PLANTS/ ENERGY)	CAPITAL COST	ECONOMIES OF SCALE	FLEXIBILITY OF OPERATION	RELIABILITY AVAILABILITY & MAINTAINABILITY (RAM)	ENVIRONMENTAL ACCEPTABILITY
DISTILLATION							
MSF*	YES	LOW-MED	HIGH	HIGH	NONE	MED	MED
MEMS*	YES	MED	HIGH	HIGH	LOW	MED	MED
VC*	YES	LOW	MED	HIGH	LOW	LOW	MED
FREEZING	NO	HIGH	MED	LIKELY	NONE	LOW	LOW-MED
MEMBRANE							
RO*	YES	MED-HIGH	MED	LOW	LOW-MED	MED	HIGH
ED*	YES	MED	MED-HIGH	LOW	VERY HIGH	HIGH	HIGH
RO-ED "HYBRID"	YES	MED-HIGH	MED	LOW	VERY HIGH	HIGH	HIGH

*MSF - MULTISTAGE FLASHING; *MEMS - MULTIPLE EFFECT MULTISTAGE; *VC - VAPOR COMPRESSION; *RO - REVERSE OSMOSIS; *ED - ELECTRODIALYSIS

The third criterion of maximum utilization of solar insolation leads also to the selection of a PV system which is likely to provide energy on cloudy days and during haze periods wherein parabolic trough collection is very limited.

Needless to say, low capital cost of equipment is an important criterion for selection of various subsystems. A hybrid RO-ED would meet this criterion. Also, the major equipment, including the solar collectors, energy delivery systems and storage tanks are selected to assure low capital cost while maintaining appropriate conditions compatible with the rest of the design features.

The system design is modular and may be replicated easily in a wide variety of geographic locations both in the US and Saudi Arabia. Thus, site specific features are less important and the size of the collector fields, the power requirements, or the type of pretreatment may be varied with minimum effort. This would assure relatively modest investment levels for commercialization of the plant design. The universality of the design will also ease the transfer of technology between Saudi Arabia and the US and could extend the benefits from applications of solar energy technologies to other countries.

Since the 6,000 m^3/day size is not necessarily the optimum design for specific applications, the plant design is such that it is possible to scale it down to a pilot plant size or to small sizes for areas of lower demand. It should also be feasible for scale-up for future use in areas of higher demand, especially in the commercial stage. System modularity would provide such flexibility without losing the potential economy of size. Maintenance at lower cost is probably the most attractive feature of modular systems.

A special consideration is made to account for factors common between the US and Saudi Arabia. This is done especially in selection of the solar energy source. Indirect solar energy sources; such as wind and biomass, have been excluded since they are less adequate in Saudi Arabia than the direct solar energy source. Wind regimes can not support a wind-powered system anywhere on the seashore whether near the Red Sea of the Gulf. Naturally available biomass is rather scarce in Saudi Arabia.

Another criterion used in selection of equipment is the use of components which are commercially available now or in the near future (by the time of implementation of the design) and selection of off-the-shelf items whenever possible. A specific off-the-shelf item is selected when a minimum of three to four US companies are able to supply the equipment specified in the design and when commercialization programs exist which focus on lowering the cost of such items. Whenever off-the-shelf items are not available, equipment is selected for which at least the performance of the basic technology has been demonstrated for one to two years and commercial projects are in place to develop such equipment. Also, the selected equipment is at least second generation. In addition, options in selection of subsystems and interface components must exist to ensure ability to optimize the economic water production.

Innovative Features

The ED stage is directly coupled to the DC power output of the PV array, and hence the water output of the ED can follow variations in power, minimizing battery storage requirements and maximizing energy use efficiency. Daily storage losses may amount to 15-25 percent of the collected energy if discharge/charge cycles are required. Also, eliminating the need for A-C power invertors, current conditioners and stabilizers avoids losses on the order of 5-10 percent during normal operations.

A second feature of the design is the use of the waste heat from the thermal energy delivery system to increase temperature of the ED feedwater. Power consumption of the ED stacks falls about 1.8% per degree centigrade increase in feedwater temperature, due to an increase in conductivity of the stack. This direct use of wasteheat also eliminates the need of a cooling tower.

The RO and ED stages are decoupled by the use of an intermediate storage tank and the independence of the energy delivery system for each stage. This allows for a great degree of flexibility. Complete shutdown for repairs of either stage can be done without affecting the other stage. The ED (product side only) is blended to accommodate variations (seasonal and tidal) in sea water salinity, degradations of the RO product water with time, and to increase the overall water recovery of the plant.

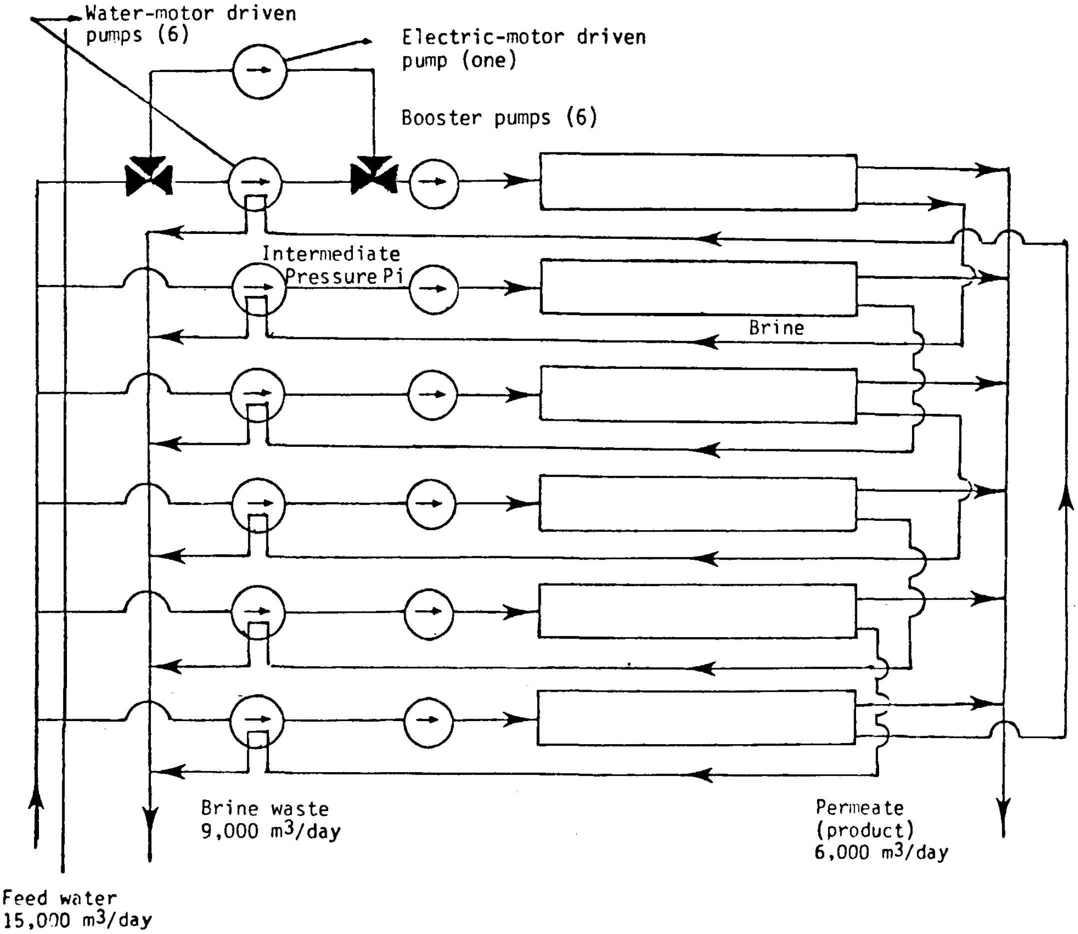


FIGURE 3. SCHEMATIC OF POWER RECOVERY USING DIRECT COUPLING BETWEEN HIGH PRESSURE BRINE AND FEED WATER PUMPING SYSTEM

The high pressure brine from the RO is used directly to operate water motors to drive the high pressure RO pumps (Pelton wheel driven pumps). Booster pumps are also used to reach the required pressure as shown in Figure 3. This arrangement could provide 40% reduction of the load on the thermal collectors not counting for the power requirements for the auxiliary system.

DESIGN MODELLING

The design is simulated by a solar desalination computer code "SOLAR DESAL" developed for the purpose of performing trade-off among various design options and to fine tune the design parameters. A flow chart of the code is shown in Figure 4.

The program has been structured to allow the interchange, substitution, and elimination of major subsystems, a flexibility necessary for trade-off analyses and parameter studies. Reliability considerations are included in the code calculations by the computation of times of failure and of repair. Program logic is based upon the availability of a particular component/subsystem in addition to the usual temperatures, flow rates, etc.

The SOLAR DESAL code simulates the production of water from a plant that includes pretreatment, battery storage, backup diesel generator, and water storage and demand simulations. This is in addition to solar thermal collectors, photovoltaics, a second desalination subsystem, heat engine, waste pond, recovery turbine, and various minor subsystems and components such as heat exchangers, piping, pumps, chiller, etc. Actual insolation data may be read in and patched directly into the solar radiation processor.

The code has the capability to calculate performance of two different types of: (1) solar thermal collectors, (2) different photovoltaic cells, and (3) desalination stages. Solar energy calculations are made for specified basic units of a set number of thermal collectors connected together to provide the necessary temperature and mass flow, or photovoltaic modules connected in series to obtain sufficient voltage. The basic unit can be sized upwards by combining similar units to allow for virtual unlimited array size.

Desalination and heat engine subroutines are constructed to allow the interchange of reverse osmosis and electrodialysis units. Heat engines can be changed simply since performance data in the code is calculated from generalized functions utilizing performance parameters provided by the program user. Similar functions are provided for the recovery turbine, waste pond subsystem, solar collectors, water treatment, RO and electrodialysis subsystems. Thermal energy storage is modelled using modified versions of the TRNSYS 10.1 stratified fluid storage tank and rock bed thermal storage.

Program logic is structured such that system availability is considered first. Each major component/subsystem is examined for available capacity based upon user supplied failure rate and repair time data, or distribution parameters, and user supplied subsystem configuration including re-

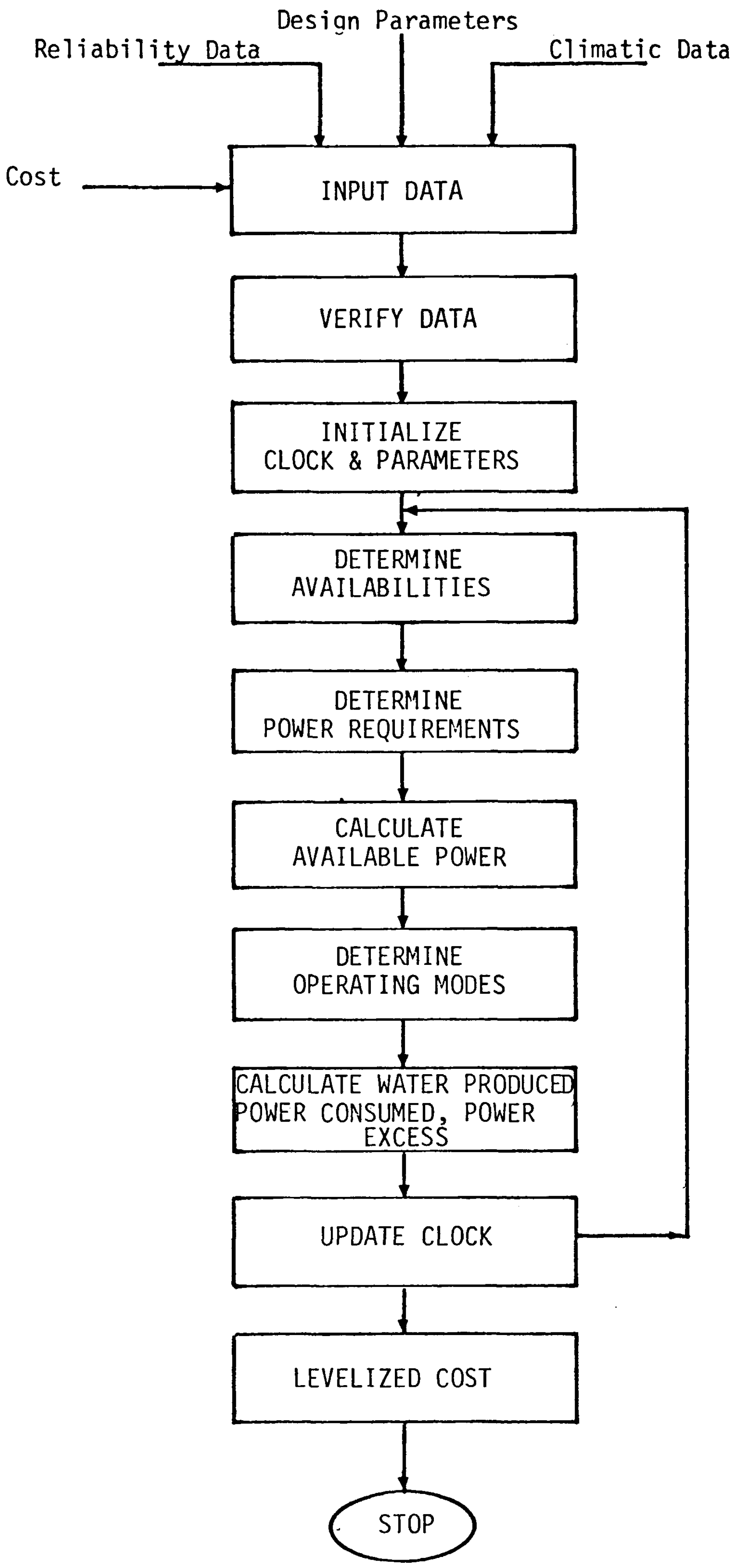


FIGURE 4. SOLAR DESAL FLOW CHART

serve units. After capacities are determined, power requirements and possible flow rates are calculated, available energies from all sources are then determined. These energies are compared with the subsystem requirements to determine water system capabilities, including the proper operating mode for the electrodialysis desalting subsystem. Water production is then found and excess energy is either stored; if possible, or noted. Water storage tank levels are monitored and not used in capacity calculations. The SOLAR DESAL code also includes cost computation based on the levelized cost model [5]. The approach provides cost of energy as well as the cost per unit of product water.

SUBSYSTEMS DESIGN

Desalination System Design

A process flow diagram for a pretreatment system, suitable for the RO plant, is shown in Figure 5. This pretreatment involves chlorination of the sea water at the intake from the sea well, coagulation with clarification, acidification, sodium hexametaphosphate addition, multimedia filtration, and cartridge filtration. Pretreatment of the sea water feed to the ED plant following chlorination is limited to multimedia filtration. The pretreated feed water for the ED plant is sent to a feed water storage tank where the pretreated feed water for the RO plant is used directly. The stored, pretreated feed water is used as required to back wash the multimedia filters.

The first desalting stage is the RO plant (Fig. 6). Its design parameters are given in Table 5. The RO plant discharges, at a minimum, 60% of the high pressure feed water as brine waste. The available energy (exergy) in this pressurized liquid can be recovered by using it to drive an impulse type Pelton turbine. The power produced by the Pelton turbine can be used to drive some of the feedwater pumps so that the ORC generator capacity and thermal field size would be reduced.

In an alternate scheme, the high pressure brine may be used directly to operate "water motors" which are used to raise the feed water pressure to an intermediate value p_i between atmospheric and the supply pressure of 5.52 MPa. Booster pumps are then used to raise the feedwater pressure from p_i to 5.52 MPa, (Figure 3).

As can be seen from Figure 3, one of the RO units (say unit #1) is equipped with both a water-motor pump and an electric motor driven pump. The system operating sequence will be as follows:

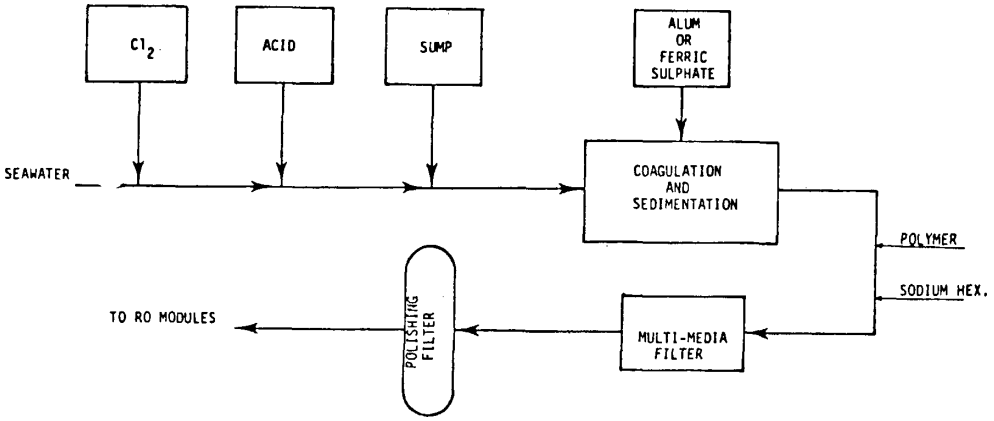
1. Start the ORC generator
2. Start the electric motor driven pump and booster pump for RO module #1
3. Use the brine from module 1 to operate the water-motor driven pump for RO module #2
4. Start the booster pump for module #2
5. Use the brine from module #2 to operate the water-motor driven pump for RO module #3

6. Start the booster pump for module #3
7. Repeat above procedure for modules #4, 5, and 6
8. Use the brine from module #6 to operate the water-motor driven pump for module #1
9. Switch the valving on module #1 so that the feed water is now pumped by the water-motor driven pump and the booster pump
10. Stop the electric-motor driven pump for module #1

Assuming a booster pump efficiency of 0.8, the required ORC generator capacity is 1.68 MW compared to 2.25 MW with no energy recovery. The required capacity using Pelton turbine is 1.74MW.

TABLE 5. RO DESIGN PARAMETERS
(SEA WELL WATER INTAKE)

Design feed temperature, ^oC	20
Maximum feed temperature, ^oC	25
Raw Feed pH	8.2
Operating Pressure, MPa	5.2
Plant capacity, m^3	7,500
Concentration of Product, ppm in ions	1,002
No. of Permeators (B-10 M-6840-055)	968
Specific Power Requirements, MJ/m^3	171
Permeator Product Rate (after 3 years), m^3/hr	0.32
Salt Passage Correction Factor	1.62



ITEM LIST

1. Chlorine addition
2. Acid addition
3. Sodium Hexa-Meter phosphate addition
4. Coagulation and sedimentation
5. Multi-media filter
6. Polishing filter

FIGURE 5. RO PRETREATMENT

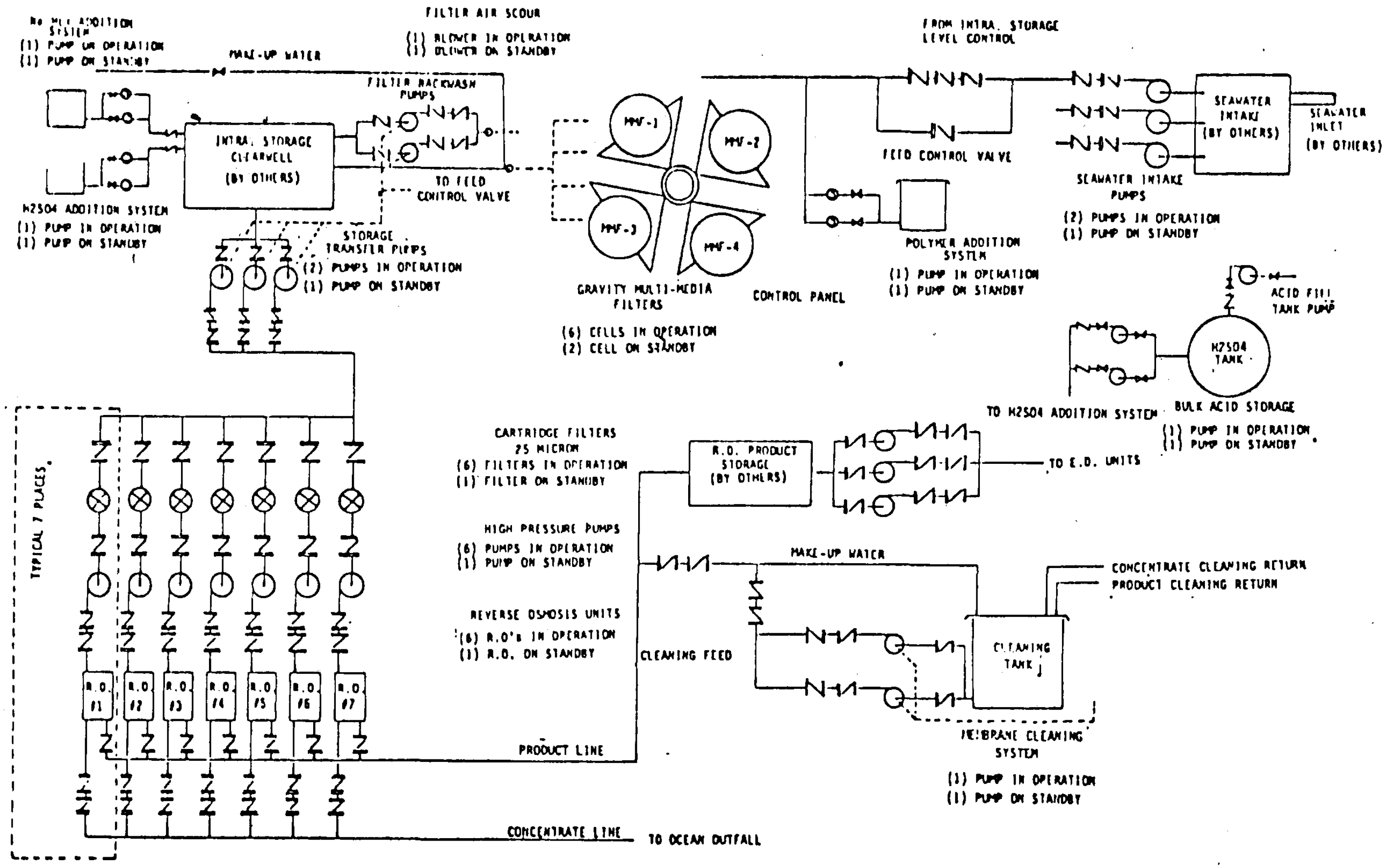
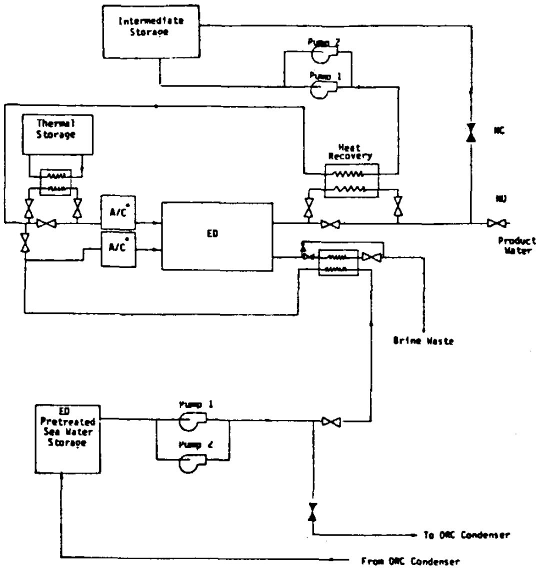


FIGURE 6. PROPOSED FIRST STAGE 6000 m^3/day RO FLOW SCHEMATIC

The second desalting stage is the ED plant (Figure 7). This plant can be operated in two modes - full or reduced flow. At low levels in insolation, it is desirable to reduce the voltage to the ED plant by 50% (through the use of a voltage switch or the battery bank) and to reduce the production of the ED plant accordingly. In the process flowsheet given in Figure 7, the reduced production rate is achieved by using a secondary feed pump (and switching off the main feed pump).



* Active Carbon Filter

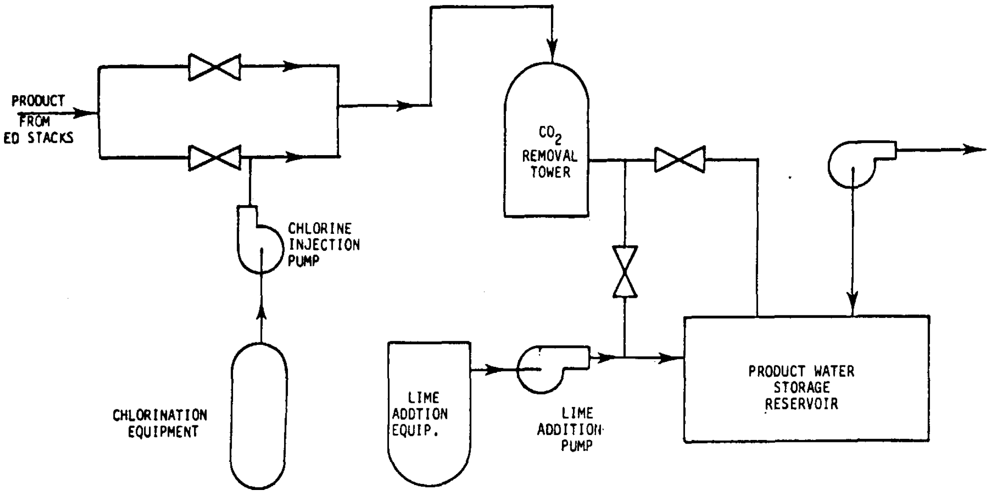
FIGURE 7. PROCESS FLOW SHEET FOR THE PV/ED STAGE

Since the current is also reduced in the reduced flow mode, this mode of operation is more energy efficient. That is, the specific energy consumption is now half of the specific energy consumption (MJ/m^3) at the design capacity and the power demand is significantly reduced. These changes make the reduced flow mode well suited to directly utilize the low levels of energy that are available in the early morning and late afternoon as well as during those days that are partially overcast. This mode is particularly effective in minimizing the loss of scarce energy when the pretreatment (and waste disposal) pump is disconnected and the plant continues to operate at the reduced flow until the feed water (or intermediate product water) storage tank is depleted.

The post-treatment unit is shown in Figure 8. The maximum power requirements before optimization are listed in Table 6 for the component systems in the desalination subsystem. The operational times for the RO and ED plants are assumed to be 24 and 8 hours, respectively. The plant control and data acquistion systems are depicted in Figure 9.

SOLAR ENERGY AND DELIVERY SYSTEMS

Desalination plant energy requirements are provided by a combination of line focusing parabolic trough and flat plate photovoltaic collector fields which can operate in either independent or parallel modes (as shown in Figure 2). The LFTC field provides thermal energy for AC current production to power the RO, pretreatment, post-treatment, and house and auxiliary loads, while the PV field produces DC power for the ED stacks. Power conditioning equipment allows for AC to DC or DC to AC conversion as required, allowing flexibility in desalting plant operation and improved



ITEM LIST

1. Chlorination
2. CO_2 removal
3. Lime addition

FIGURE 8. POST-TREATMENT

TABLE 6. MAXIMUM POWER REQUIREMENTS (kWe)

RO (24 hours)	1629
RO pretreatment	100
Post-treatment	1
Auxiliary Loads	10
Total RO + Pretreatment + Auxiliary	1740
ED (8 hours)	820
Total Plant (24 hours)	2550

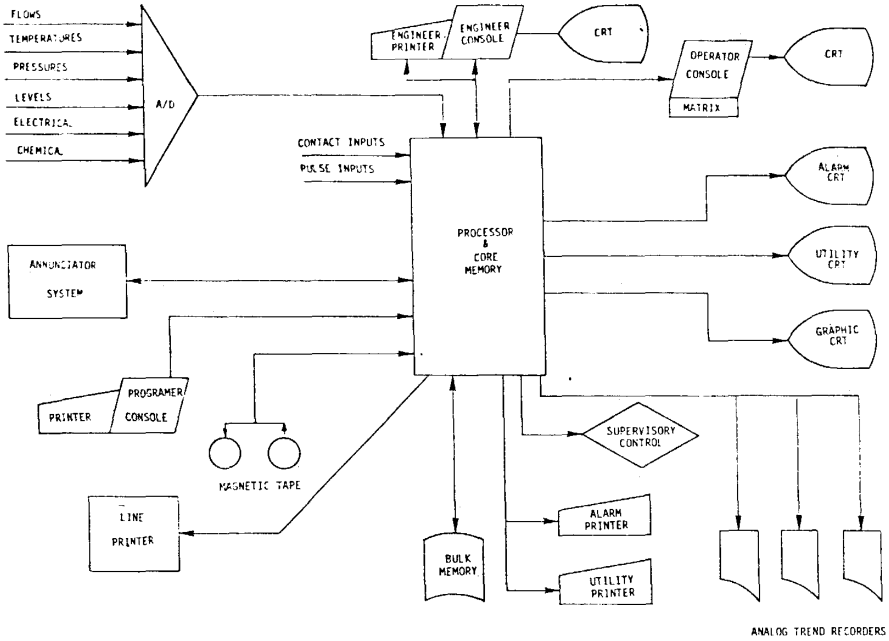


FIGURE 9. PLANT CONTROL AND DATA AQUISITION SYSTEM

power plant reliability through system redundancy.

The AC power requirements are provided by 70,000 m^2 of north-south oriented two-axis parabolic trough collectors, with the capacity for an average eight hour energy delivery of 970 GJ, feeding three 600 kW organic Rankine cycle engines (Figure 10). Caloria, HT-43, exits the collector field at approximately 300°C and is pumped to the thermal storage header and heat exchanger with average energy capacities of 26GJ and 95GJ, respectively. Toluene is pumped through the secondary side of the heat exchanger at 5 kg/sec/ORC providing an ORC turbine inlet temperature of approximately 270°C. Total ORC generation is about 1740 kW which provides all plant AC loads. Condensor outlet (at 59°C) is used to heat the intermediate product water for improved ED operating efficiency. In this mode, the thermal storage system is charging and recycling to the collector inlet.

Two thermal storage tanks are used and comprised of 65% rock and 35% Caloria by volume, providing a heat capacity of .17GJ/m^3. Tanks are then sized at 2250 m^3 with a storage capacity of 380 GJ.

The 440V DC power requirement of the ED stacks is provided by 23,000 m^2 of flat plate photovoltaic collectors. At 30W peak output per PV module, a series connection of 30 modules is required to generate the necessary voltage. An arrangement of 2070 units of 30 modules yields - peak power of 1865 kW (1065 kW average over eight hour solar day) for operation of the ED system during daylight hours. A battery system provides voltage stabilization during fluctuating insolation periods.

CONCLUDING REMARKS

Through the SOLERAS Program the concepts for solar desalination which have been deliberated on for the past four decades, have materialized. The design developed here demonstrates the possibility of commercialization of solar desalination plants using present-day technology. Cost reduction of solar energy systems would allow such efforts to come to fruition.

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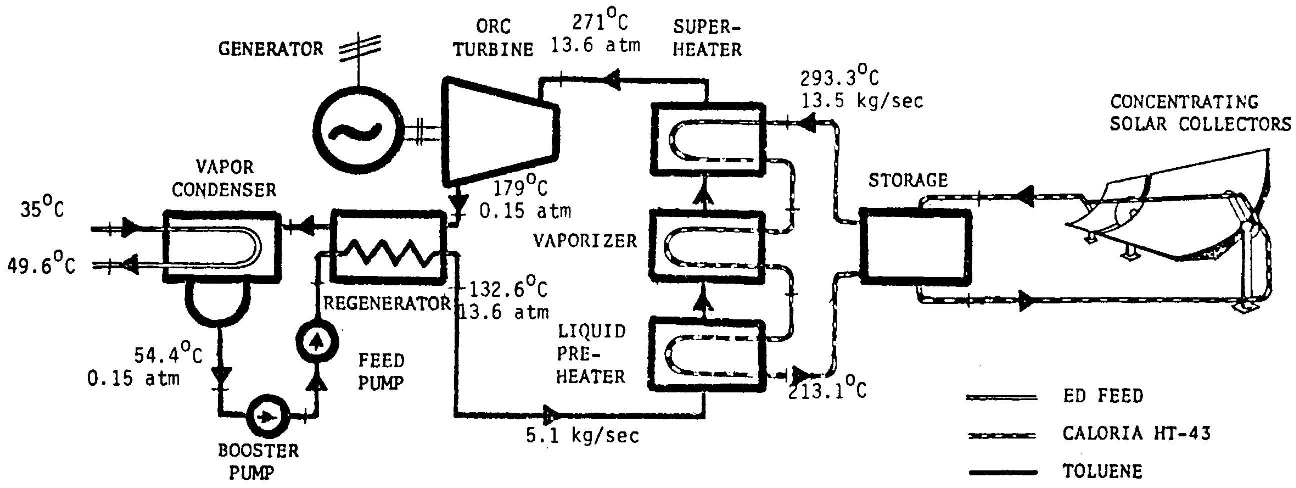

FIGURE 10. SOLAR THERMAL POWER PLANT FOR RO/ED HYBRID (THREE ORC'S ARE CONNECTED IN PARALLEL)

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