

Economic Implications of Nuclear Desalination on Seawater Chemical Recovery*

William C. Yee
Oak Ridge National Laboratory
Oak Ridge, Tennessee

ABSTRACT

An economic analysis is made of several possible plant complexes that recover chemicals from the brine discharge of a nuclear desalination plant producing 1000 mgd of fresh water with a seawater concentration factor of 2. Primary product recovery involves using solar energy at an arid desert site to separate out sequentially gypsum, salt, potash and magnesium chloride by fractional crystallization. Secondary product production involves using electrical and steam energy from the desalination plant to obtain cement and sulfuric acid, caustic soda and chlorine, and magnesium metal. Caustic soda and chlorine (as hydrochloric acid) can be used to remove bicarbonate ion from seawater prior to fresh water production.

Land requirements for solar evaporation are reduced by the availability of evaporator effluent as starting material. Starting with two times concentrated (2x) seawater there is a 40% reduction in acreage for the production of solar salt; with 3x seawater, a 60% reduction. At a 10% cost of money, the cost of salt production is reduced 15% and 25%, respectively; this assumes that the cost of the undeveloped land is negligible in comparison to the land improvement cost. The processing of 3% of the brine discharge (2x seawater concentration) from a 1000 mgd plant would consume 400 mw(e) of electric power, would require a capital investment of \$95 million exclusive of the nuclear reactor station, would generate an income of \$40 million and would have a break-even power cost of 3 mills/kwhr. Seawater treatment by the NaOH-Cl method would compete with the conventional sulfuric acid method of treatment at a power cost

of 4 mills/kwhr when sulfur costs \$50/ton or 2.5 mills/kwhr when sulfur costs \$40/ton. A cost analysis is also given for a salt complex that utilizes 6% of the evaporator effluent from the nuclear desalination unit.

INTRODUCTION

In June 1967, the Oak Ridge National Laboratory began a study of the technical and economic feasibility of "nuclear-powered industrial and agro-industrial complexes," primarily as an avenue to industrial, agricultural, and general economic advancement in developing countries. Such a complex, shown schematically in Figure 1, might consist of a large nuclear reactor station producing both electricity and desalted water. In the absence

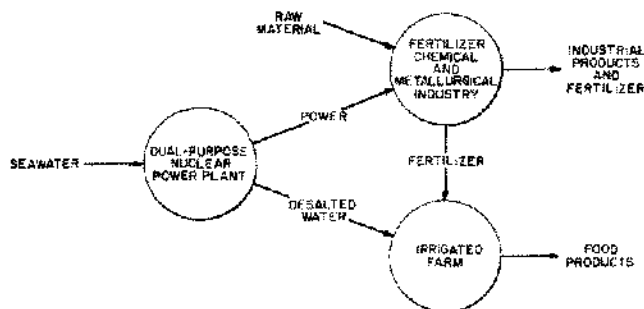


Figure 1. Nuclear-powered Agro-industrial complex.

*Research sponsored by the U.S. Atomic Energy Commission under contract with the Union Carbide Corporation.

of a large power grid the electricity would be consumed principally in adjacent industrial processes and for pumping water, while the desalted water could be used either for municipal and industrial purposes in an industrial complex or in an irrigated agricultural complex located in a coastal desert region. An artist's conception of such a complex (i.e., nuplex) located in a coastal area is illustrated in Figure 2.

To make this feasibility study, the Laboratory brought together a full-time study group staff of 16 engineers, economists, scientists, and agricultural experts during the summer of 1967 under the direction of Professor E.A. Mason of the Massachusetts Institute of Technology. This staff was assisted by six consultants who worked on special topics and by an advisory panel of 13 distinguished consultants from industry, government, and academic institutions. Experts from nine industrial

organizations provided information concerning capital and operating costs for various industrial processes, while a large number of other contributors provided information on various other aspects of the project. The detailed results of this study project are published in a series of reports (ORNL-4290, 1968; Young and Michel, ORNL-4291, 1968; Stout, ORNL-4292, 1969; Tamura and Young, ORNL-4293, 1969; Squires and Lobo, ORNL-4294, 1969; Ritchey, ORNL-4295, 1969; and Goeller and Mrochek, ORNL-4296, 1969).

This paper considers one aspect that was not fully covered in the general study, namely, the recovery of seawater chemicals from the brine concentrate that is discharged as a waste from the nuclear desalination unit. Large dual-purpose nuclear desalination units might produce 2000 megawatts of electrical power [mw(e)] and 1000 million gallons of desalted water daily (mgd) from

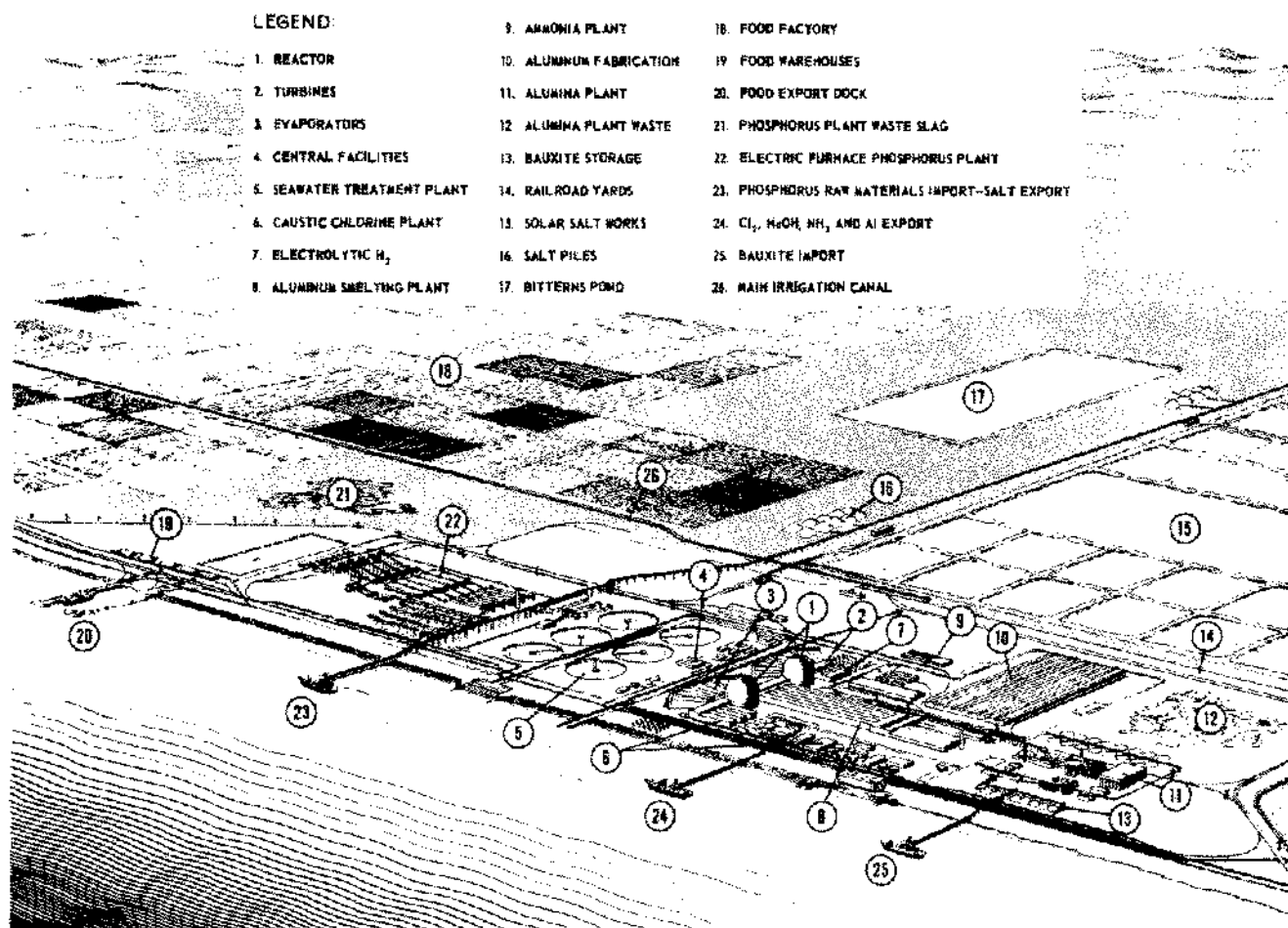


Figure 2. Conceptual layout of a nuclear-powered Agro-industrial complex.

2000 mgd of seawater. A portion of the 1000 mgd of waste brine effluent (2x concentrated seawater), could be concentrated further by solar evaporation for primary product recovery since solar energy is abundant in an arid desert location. Secondary product production would utilize the electrical and steam energy available at the nuclear reactor station. Various combinations of these separate industrial processes or building blocks would form a salt complex in which waste effluent brine is the source for the recovery of a number of basic chemicals.

One of these products, magnesium metal, involves comparatively new production technology which is expected to result in a 400% growth in the consumption of the metal in the next decade (*Chemical Week*, May 17, 1969). Thus the product mix of chemicals recoverable from seawater should revert from one characterized by the production of bulk commodities like salt and potash, which have low unit prices, to one that includes the production of a high unit priced metal.

This presentation focuses on the economics of the process and technical details are kept to a minimum. Process building blocks are developed first followed by two examples of a salt complex.

PROCESS BUILDING BLOCKS

In an arid coastal area, solar energy can be utilized to concentrate evaporator effluent and to recover a number of chemical compounds by fractional crystallization; further processing can lead to a variety of product chemicals. One schematic example of such a complex is shown in Figure 3. Solar evaporation of the brine permits the sequential separation of a number of primary products: gypsum at 20 to 26°Be, salt at 26 to 30°Be,

potash at 33 to 36°Be, and magnesium chloride concentrate (solution) at 40°Be.

Gypsum can be the raw material for sulfuric acid and cement manufacture or for the production of elemental sulfur alone. Salt can be refined and sold. It can be used also as the raw material for the production of caustic soda and chlorine, which are basic chemicals to a whole family of organic and inorganic end products. For the nuplex in particular, both caustic soda and chlorine (as hydrochloric acid) can be used to treat seawater before nuclear desalination or to manufacture pulp from the large amounts of wheat straw that remain as a waste of the nuplex food factory (ORNL-4290, 1968). Caustic soda can be used to produce aluminum oxide (Al_2O_3) from bauxite, rayon from pulp, soap from animal fat, or detergents from petroleum residues when they are available. Chlorine can be used to produce a multitude of chlorinated hydrocarbons or to manufacture polyvinyl chloride (PVC) plastic. Potash can be recovered, refined, and sold as the third component of a balanced fertilizer. Magnesium chloride can be dehydrated and reduced by fused salt electrolysis to magnesium metal and chlorine. Other products that are also retrievable but are not shown in the figure include bromine at 32°Be and lithium, a by-product of the purification of magnesium chloride concentrate. The major use of bromine, of course, is for the production of ethylene dibromide, an antiknock additive in gasoline. Lithium additive to molten alumina reduces power consumption requirements for electrolytic reduction of the alumina to aluminum metal (Smith and Andrews, 1967).

This paper is concerned only with the primary and secondary products and the treatment of seawater with caustic soda and hydrochloric acid. The economic building blocks are discussed in the increasing order of their sensitivity to the cost of electric power; i.e., solar salt, potassium chloride, caustic and chlorine, sulfuric acid and cement, and magnesium metal. All costs and income are based on conditions prevailing in mid-1967, with no allowance for escalation. Capital investment for these calculations include only battery limit costs; offsite costs are added in the salt complex examples given in the following section.

For all of these illustrations the indirect costs include recovery of investment, return on investment, and interest on working capital. Taxes and local insurance considerations are omitted. Recovery of investment is based on a service life of 15 years for all industrial plants except the solar salt operation which is 35 years. It is calculated by the

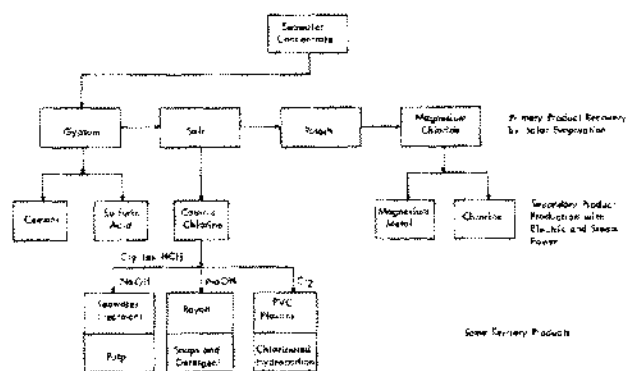


Figure 3. Schematic example of a salt complex.

method that rapidly depreciates the plants in the later years of the plant lifetime, namely, by the sinking fund formula,

$$SFDF = \frac{i}{(1+i)^n - 1}$$

where

SFDF = sinking fund depreciation factor

i = cost of money as an annual interest rate

n = investment lifetime (years).

Return on investment is considered at a cost of money of 2.5, 5, 10, and 20%. Working capital is based on the cost of 60 days of production at gross manufacturing cost.

Solar Salt.

Solar salt production, obviously, is least dependent on the cost of power, requiring only 4 kwhr/ton. It is assumed that 40,000 working acres are required to produce 3000 tons/day (1 million tons/yr) of product from ocean water and that the cost of the undeveloped land is negligible compared with land improvement costs. This land area can be divided into three categories: the initial ponds (16,000 acres), intermediate ponds (14,000 acres), and crystallizer ponds (10,000 acres). The initial ponds could be eliminated if the starting material is evaporator effluent that is 2x seawater; at 3x seawater about one-half of the intermediate ponds could also be eliminated. In the latter case 1 million tons could be produced annually on less than 50% of the land required if ocean water is the starting material.

Figure 4 shows the effect of power cost, cost of money, seawater concentration factor, and plant capacity on the total manufacturing cost of solar salt production. Figures 4(a) and 4(b) show that the cost of power has practically no effect on the product cost; however, product cost is sensitive to the cost of money because of the high capital investment for the process. For example, the manufacturing cost for a 3000 ton/day (1×10^6 tons/yr) plant at a 10% cost of money is about \$1.87/ton of NaCl at power costs ranging from 1 to 8 mills/kwhr; as the cost of money goes from 2.5% to 20%, however, the manufacturing cost jumps from about \$1.40 to \$2.65/ton of NaCl, an increase of almost 90%.

The seawater concentration factor has a significant effect on the manufacturing cost. As illus-

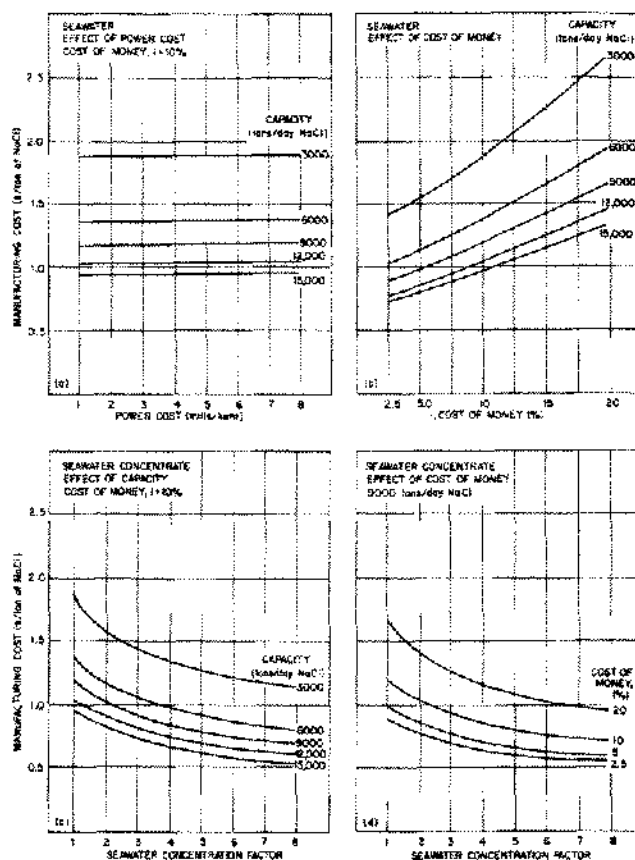


Figure 4. Manufacturing cost of solar salt from seawater and seawater concentrate.

trated in Figure 4(c) for a 10% cost of money, there is a 15% drop in cost when 2x seawater is the starting material, and there is a cost reduction of 25% with 3x seawater as the starting material. When using ocean water, solar salt can be produced for \$1/ton or less at plant capacities of 12,000 tons/day (4×10^6 tons/yr) or more. If seawater concentrates are used instead of ocean water, solar salt at \$1/ton can be obtained at lower plant capacities: at 2x seawater, a salt plant of 9000 tons/day (3×10^6 tons/yr) or more is required; at 3.5x, a plant size of only 6000 tons/day (2×10^6 tons/yr) is needed.

Over a period of time it is not unreasonable to expect that the soil base of the solar ponds will become increasingly impervious, that the leakage losses will be smaller and that the yield of salt per acre should be larger. This factor is difficult to evaluate in a general discussion and its effect on salt production costs is not included here (ORNL-4290, 1968, p. 211).

potassium chloride.

The recovery of potash from an all-chlorine item involves the simultaneous crystallization of nallite ($\text{KCl} \cdot \text{MgCl}_2 \cdot 6\text{H}_2\text{O}$) and halite (NaCl) in salt bitterns in the NaCl to KCl ratio of about 1 to 1, the separation of carnallite from halite by conventional amine flotation, and the leaching of nallite with salt bitterns recycle liquor, which solves MgCl_2 and leaves KCl as solid crystals. Power requirements for this process are relatively all at 70 kwhr/ton of KCl .

The manufacturing costs decrease with increased plant capacity and are not significantly influenced by power costs or the cost of money. The capital investment for this process includes only the separation and purification of potassium chloride without any ponding costs, which are minimal compared with the ponding costs of recovering solar salt. As shown in Figure 5, at a 10% cost of money the cost of potash production changes very little, i.e., from about \$11 to about \$11.50/ton of Cl for power costs ranging from 1 to 8 mills/whr. Increasing the plant capacity tenfold from 50 tons/day (50,000 tons/yr) to 500 tons/day (500,000 tons/yr) could drop the cost from \$15 to 7/ton of KCl .

Canadian-produced potash costs \$23/ton of potassium chloride F.O.B. port of embarkation. This includes a production cost of \$14 F.O.B. Saskatchewan, and the minimum charge of \$9/ton (*Chemical Week*, 1968, p. 47) for shipment by rail from Saskatchewan to Vancouver, British Columbia, on the west coast. The price to an underdeveloped country like India, including additional shipping charges, is about \$40/ton. Thus potash produced locally by solar evaporation of evaporator effluent at a cost of \$11 to \$12/ton would provide a considerable saving over the imported product. More importantly, foreign exchange supplies would be conserved. At present there are no known large supplies of natural potash in the Asian countries, and they rely on American and European supplies to furnish their needs.

Caustic soda and chlorine.

Electrolysis of a saturated salt solution in a diaphragm cell* yields chlorine, hydrogen, and a caustic soda solution containing an equimolar quantity of unelectrolyzed salt. When produced for sale, caustic is concentrated to at least 50% NaOH by evaporation; in the process, salt is recovered and recycled to electrolysis. Power requirements are 3200 kwhr/ton of Cl_2 ; an additional 100 kwhr/ton are required for caustic concentration.

The manufacturing cost of caustic-chlorine production is shown in Figure 6 as a function of power cost, cost of money, salt cost, and plant capacity. For a plant capacity of 1000 tons/day of Cl_2 , a reduction in power cost from 6 to 4 mills/kwhr reduces the manufacturing cost from about \$47 to \$39/ton of Cl_2 when the cost of money is 10% and the cost of salt is \$3/ton. The manufacturing cost is the combined cost of producing 1.0 ton of chlorine and 1.1 tons of caustic but is expressed in the conventional manner of cost per ton of chlorine.

Caustic-chlorine production offers the possibility of using caustic soda, hydrochloric acid, or any combination of these reagents for seawater treatment prior to nuclear desalination (Yee, 1968). Distillation of seawater (Office of Saline Water Report NP-16361, 1965, p. 218) without treatment

*Mercury cells are also used for brine electrolysis; here, concentrated caustic soda is obtained directly along with chlorine. In this paper, the diaphragm cell method is used in the cost analysis because caustic soda concentration is optional. When dilute caustic soda can be used, as, for example, in seawater treatment prior to nuclear desalination, no capital investment in a caustic concentrator is required, and the capital cost of the diaphragm cell process is cheaper than the mercury cell process.

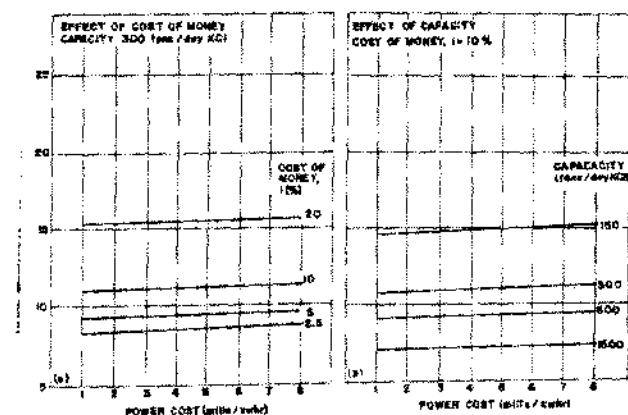


Figure 5. Manufacturing cost of potassium chloride from solar salt bitterns.

In view of the current large excess supply of potash, it is necessary to include a statement about the advantages of producing potash at seaside. Logistics play an important role in potash economics. The largest sources of supply are the inland areas of Canada and the United States.

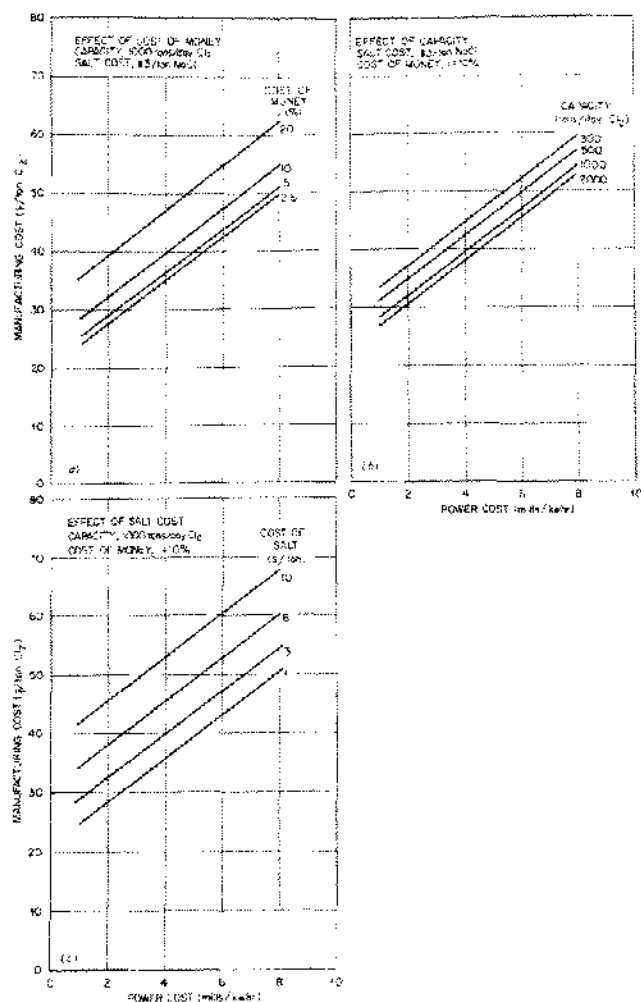
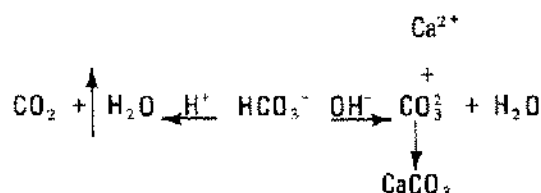


Figure 6. Manufacturing cost for production of caustic soda and chlorine.

limits the efficiency of fresh water recovery because of the formation of alkaline scale [CaCO_3 , $\text{Mg}(\text{OH})_2$] at 170 to 180°F and the deposition of this scale on heat transfer surfaces of the desalination unit. Removal of the bicarbonate ion in seawater prior to desalination allows the evaporator temperature to go to a maximum (Yee, 1968 and Office of Saline Water Report NP-16361, p. 218) of about 260 and 270°F. This can be done either by acid treatment, which converts the bicarbonate ion to carbon dioxide gas, or by caustic treatment which yields the carbonate ion and, in turn, combines with the calcium ion present in seawater to precipitate calcium carbonate. Removing a fraction of the calcium ion in seawater in addition to complete bicarbonate ion removal allows an even

higher evaporator temperature (Marshall and Slusher, 1968, p. 83). The mechanism for these reactions is



The freedom to choose either method of seawater treatment offers economic advantages in marketing situations where product demand is not in balance with the coproduced caustic and chlorine (1.1 ton NaOH/1 ton Cl_2). For example, in industrialized nations (e.g., the United States) the demand for chlorine is greater than that for caustic soda (*Chemical Engineering*, 1967, p. 128) so that NaOH treatment of seawater is a sink for surplus caustic. In a developing nation like India, the reverse is true (United Nations Publication E/CN.11/635, 1963, p. 356) so surplus HCl could be used initially to treat seawater. As the nation becomes more industrialized, the ratio of NaOH to HCl used for seawater treatment could be varied according to the marketability of caustic soda and chlorine. In the extreme, all of the chlorine could be marketed and surplus caustic could be used for treating seawater, as in the U.S. case. Alternatively, all of the caustic and chlorine could be utilized in a closed-cycle process for seawater treatment.

Nuclear desalination in a coastal desert region offers the advantage of abundant solar energy for recovering salt, cheap power to convert this raw material into caustic soda and chlorine, and a degree of freedom to select a method of seawater treatment that utilizes either surplus chemicals not marketed or the entire plant output in a closed-cycle process. Later in the paper a cost analysis of the closed-cycle process is made for two specific salt complexes which include the treatment of 2000 mgd of seawater to produce 1000 mgd of fresh water.

Cement and sulfuric acid.

Gypsum ($\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$), recovered by solar evaporation of salt bitterns, can be reacted with sand, clay, and coke at 2300°F in a kiln or fluidized bed to produce cement clinker; the accompanying sulfur dioxide off-gas can be converted to concentrated (96%) sulfuric acid by the contact process. Power requirements are small at 200

kw/hr/ton of H_2SO_4 if fossil fuel is available to furnish about 11 million Btu/ton of H_2SO_4 ; in an all-electric system, the power requirement is about 3400 kw/hr/ton of H_2SO_4 .

The manufacturing cost is shown in Figure 7 as a function of power cost, cost of money, and plant

reduced from 8 to 2 mills/kw/hr. The breakeven power cost for the all-electric process to be competitive with fossil fuel is 2.5 mills/kw/hr. Unit operating costs are minimum at 1000 tons/day because labor requirements are the same even if capacity is lower; larger plant capacities require constructing a two-unit plant (personal communication, Simon-Carves, Chemical Engineering Ltd., March 1968).

Magnesium metal.

Magnesium metal can be produced from MgCl_2 -rich solar salt bitterns by spray drying and dehydrating at 1200 to 1650°F to produce $\text{MgCl}_2 \cdot x\text{MgO}$, by chlorination to reconvert the oxide to MgCl_2 , and by fused-salt electrolysis to produce magnesium metal and chlorine (*Chemical Engineering*, 1969, p. 60 and ORNL-4290, 1968, p. 204). It is assumed that 4.4 tons of anhydrous MgCl_2 is required per ton of metal produced. This brine concentrate- MgCl_2 process produces chlorine whereas the traditional seawater- $\text{Mg}(\text{OH})_2$ process consumes chlorine. Utility requirements include 115 million Btu/ton of magnesium for fossil fuel (assuming 65% fossil-fuel utilization) and 20,000 kw/hr/ton for electricity. In an all-electric system (assuming 95% power utilization), 42,000 kw/hr/ton is required.

The manufacturing cost is shown as a function of power cost, cost of money, and plant capacity in Figure 8. A cost comparison is also made between using fossil fuel at \$0.50/10⁶ Btu as a heat source and using electric power. For a plant capacity of 150 tons/day (50,000 tons/yr) and a 10% cost of money, the manufacturing cost drops from \$570 to \$310/ton of magnesium metal as the power cost decreases from 8 to 2 mills/kw/hr. If a credit of \$50/ton is allowed for the coproduct chlorine, the cost of magnesium is cheaper and drops to \$460 and \$210 for electricity costs of 8 mills/kw/hr and 2 mills/kw/hr, respectively. The market price for magnesium in the United States is about \$700/ton while in Germany where semi-competitive conditions prevail, the price is about \$520/ton (Jackson, 1969, p. 116). The cost calculations assume a net yield of 2.2 tons of chlorine per ton of magnesium metal produced; 0.5 ton of the total 2.7 tons of chlorine produced is used to reconvert MgO to MgCl_2 . The breakeven power cost for the all-electric process to become competitive with the process including the use of fossil fuel is 2.4 mills/kw/hr.

Initially, the demand for magnesium in a developing nation may not be sufficient to warrant

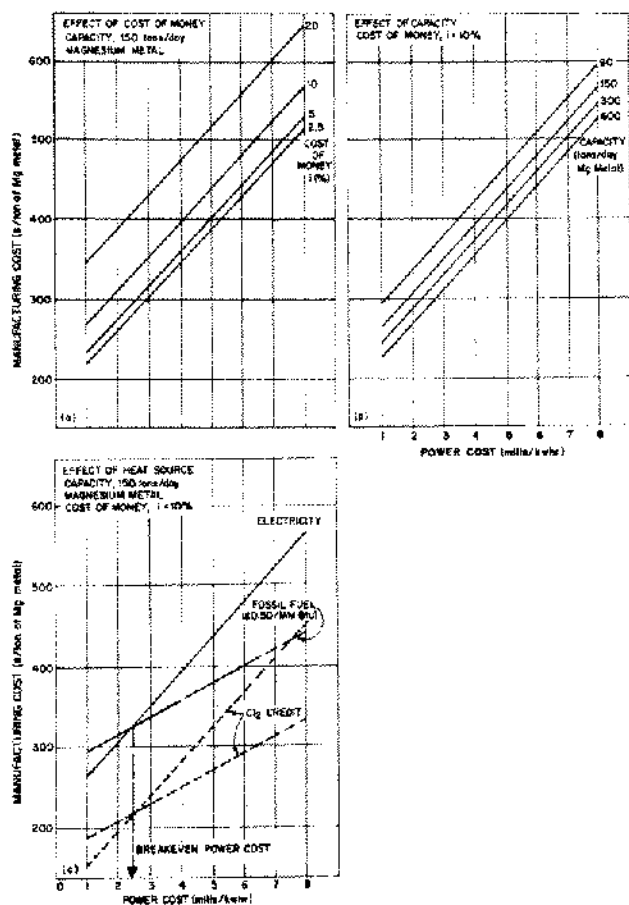
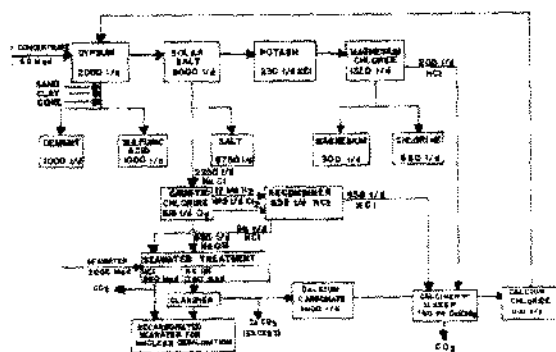


Figure 7. Manufacturing cost of sulfuric acid and cement from solar salt bitterns.

capacity. A comparison is also made between fossil fuel and electricity as a source of heat; utilization of fossil fuel is assumed to be 65% (Moditz et al, 1965, p. 2405) and that for electricity, 95%. The gypsum requirement is 2 tons for each ton of cement clinker and sulfuric acid produced.

For a 1000 ton/day plant and at a 10% cost of money, the manufacturing cost of producing 1 ton of cement and sulfuric acid is reduced from about \$40 to about \$20/ton when the cost of power is



10. Flowsheet of a salt complex utilizing 6% of the 2x concentrated brine effluent from a 1000-mgd nuclear desalination

obtain a two-fold increase in each of the products except for caustic-chlorine. This plant capacity has only to be increased to 615 tons of Cl_2 per day to obtain adequate hydrochloric acid for calcium chloride production and NaOH and HCl for treatment of 2000 mgd seawater. Out of a total of 600 tons of salt produced daily (3×10^6 tons/year) about 2250 tons/day is required for caustic-chlorine production while the balance is sold.

olar evaporation in conjunction with nuclear dination is most feasible in remote arid coastal rt regions where the cost of undeveloped land is significant compared with the cost of land im- vement. Locations such as the Magdalena Plain Baja, California, or the Kutch Peninsula in thwest India are examples of such remote es. Practically all coastal land in the United es is relatively expensive compared with the en coastal deserts of developing nations.

capital cost of each plant is assumed to be greater than that in the United States by a factor of 1.2; i.e., 60% of the capital items is assumed to be imported to the remote complex site and the cost assumed to be 1.2x that in the United States (or other supply country) because of transportation costs, while the remaining 40% is indigenous and conservatively costs the same as in the United States ($1.12 = 0.6 \times 1.2 + 0.4 \times 0.1$) (ORNL-40, 1968, p. 19).

The cost of offsites is added to the total battery plant costs to allow for the cost of such support facilities as maintenance shops, administrative facilities, railroads, roads, etc. Two functions are used for determining the offsite cost (ORNL-40, 1968, p. 82): the first is for the range of 10^5

to 10⁸ dollars of total battery limits capital investment or

$0.25/10 \times \text{sum of battery limits plant costs in dollars} \times 10^{-6})^{0.997}$; while the second is for the range of 10^8 to 10^9 dollars or

$0.128/(\text{sum of battery limits plant costs in dollars} \times 10^{-6}/100)^{0.329}$. Thus offsite capital costs are \$1.5 million for plant investment of 10^7 dollars, \$13 million at 10^8 dollars, and \$60 million at 10^9 dollars.

For the salt complex that processes 3% of the waste brine effluent, about \$40 million dollars of product value would be generated annually, about 400 mw(e) of electricity would be required and the capital cost of the complex excluding investment for power generating facilities would be about \$95 million including \$11 million for offsites. If 6% of the waste brine is processed, the annual product value would be \$82 million, the amount of electricity would be 750 mw(e) and the capital cost would be \$148 million including \$16 million for offsites.

A cost analysis of each salt complex is given in Figure 11 as a function of the cost of power. The volume of brine effluent utilized is purposely kept low in these illustrations to maintain reasonable plant capacities. With 3% of the waste brine effluent used and with \$40 million of annual income, the break-even power cost is about 5.3 mills/kwhr at a 10% cost of money. Doubling the amount of effluent used makes the break-even power cost even more attractive at 7.2 mills/kwhr. The annual labor costs used in this analysis are assumed to be one-half of those in the United States; that is to say, the individual labor cost is assumed to be

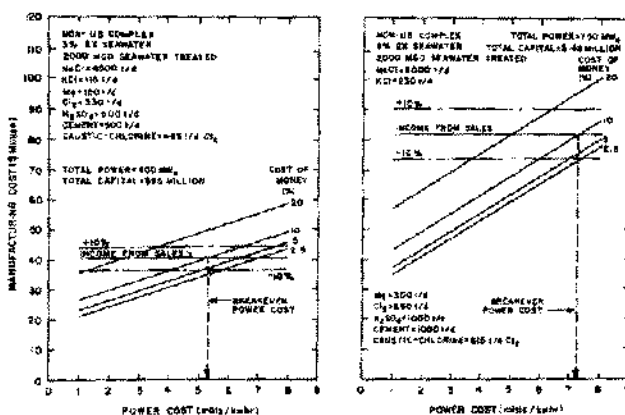


Figure 11. Non-U.S. salt complex utilizing 3% and 6% of the 2x concentrated brine effluent from a nuclear desalination unit producing 1000-mgd of fresh water.

Table 1. Summary of Product Value, Power Consumption, and Capital Cost of Two Salt Complexes

	Unit price	3% 2x Seawater				6% 2x Seawater			
		\$/ton	Product, tons annually	Product values, \$ million	Power consumption, Mw(e)	Capital cost, ^a \$ million	Product values, \$ million	Power consumption, Mw(e)	Capital cost, ^a \$ million
NaCl ^b	4		916,000	3.66	0.7	7.5	2,260,000	9.04	11.1
KCl	16		37,500	0.60	0.3	2.1	75,000	1.20	3.9
Caustic-chlorine ^c			(products used internally)		65.0	20.6	(products used internally)	82.0	25.5
Cement	17		165,000	2.81	72.0	12.3	330,000	5.61	19.0
H ₂ SO ₄	15		165,000	2.48			330,000	4.95	
Mg metal	500		50,000	25.00	263.0	41.2	100,000	50.00	73.1
Chlorine	50		110,000	5.50		11.0	220,000	11.00	15.6
Offsites Charge				40.05	401.0	94.7		81.80	148.0
Totals									

^aForeign capital costs are assumed to be 1.12 x U.S. costs. See ORNL-4290, p. 19.

^bFor 3% 2x seawater: 1.5 million tons produced annually of which 0.9 million is sold; 6% 2x seawater: 3.0 million tons produced annually of which 2.26 million is sold.

^cCapital cost includes the cost of clarifiers for seawater treatment with NaOH. For 3% 2x seawater: 0.5 million tons of NaCl electrolyzed annually to produce NaOH and Cl₂ (as HCl) for treatment of 2000 mgd of seawater and for production of CaCl₂. Plant size equals 485 tons of Cl₂ per day. For 6% 2x seawater: 0.74 million tons of NaCl electrolyzed annually to produce NaOH and Cl₂ (as HCl) for treatment of 2000 mgd of seawater and for production of CaCl₂. Plant size equals 615 tons of Cl₂ per day.

sixth of the U.S. labor rate of \$4.00/man-hr or 7/man-hr but the efficiency of non-U.S. labor assumed to be one-third of the U.S. worker. Overhead is assumed to be 60% of operating cost (ORNL-4290, 1968, p. 20).

Each of the examples given, seawater treatment in connection with nuclear desalination has included as part of the complex. The capital requirements for auxiliary equipment vary for seawater treatment by the NaOH-HCl method: the greater the fraction of seawater treated with caustic the greater the investment in clarifying equipment to separate out the precipitated calcium carbonate solids. Acid treatment requires an investment in a hydrogen-chlorine recombiner to obtain hydrochloric acid, but the cost is small compared to the cost of the clarifiers. For 2000 mgd of water, caustic treatment alone would require a total investment of about \$11.5 million in clarifier equipment; for acid treatment alone only \$0.2 to 0.3 million would have to be spent on a hydrochlorine recombiner. In the salt complex utilizing 3% of the brine effluent, caustic treatment is limited to about 70% of the seawater, and the clarifier equipment costs about \$8 million. In the complex utilizing 6% of the effluent, caustic treatment is limited to about 85% of the seawater, and the clarifier equipment costs \$10 million.

The NaOH-HCl method, which is sensitive to power cost, is compared with the conventional method of adding sulfuric acid (Yee, 1968 and ORNL-4290, 1968, p. 205), in Figure 12. Sulfuric acid is sensitive to the cost of elemental sulfur, a material used for acid production. Cost estimates are also given for two other methods of seawater treatment that are in the development stage: CO_2 suppression (Cadwallader, 1967, p. 42) and lime-magnesium carbonate (LMC) process (Office of Saline Water Report 192, 1966).

At a 10% cost of money and 3% utilization of the effluent, the break-even power cost is 4 mills/kwhr when the cost of sulfur is \$50/ton and 3 mills/kwhr when sulfur is \$40/ton. For the case utilizing 6% of the brine effluent, a greater volume of seawater is treated with caustic, there is a higher power cost for clarifier equipment, and this lowers the break-even power cost to 2 mills/kwhr when sulfur is \$50/ton and less than 1 mill/kwhr when sulfur is \$40/ton. In general, NaOH-HCl treatment is cheaper than CO_2 suppression and the LMC process.

The economics of using the NaOH-HCl method for treating seawater has been restricted in this discussion to the case in which the total caustic-

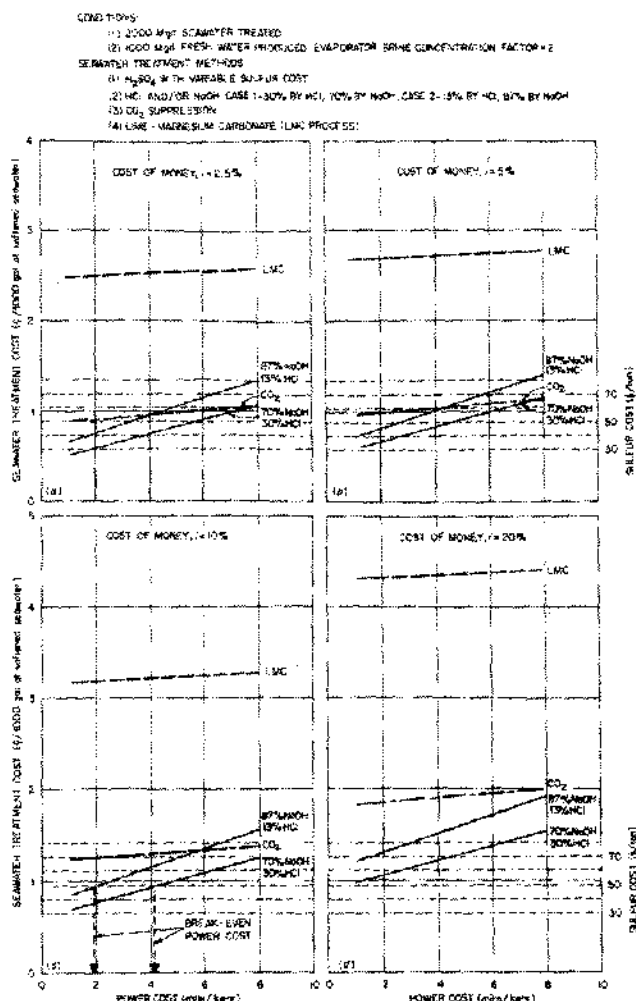


Figure 12. Cost of seawater treatment in a salt complex using caustic soda and hydrochloric acid and a comparison with other treatment methods as a function of power cost and cost of money.

chlorine capacity is consumed internally within the complex, specifically for seawater treatment and CaCl_2 production. If there is an alternative use for caustic other than for seawater treatment, such as in rayon pulp production from wheat straw, a greater fraction of the seawater could be treated with HCl, and the capital investment in auxiliary clarifier equipment together with the cost of seawater treatment would be reduced. For example, with equimolar treatment, NaOH treatment of one-half of the 2000 mgd seawater and HCl treatment of the other half, the clarifier cost would drop to \$5.7 million, the break-even power cost would be 4.5 mills/kwhr at a sulfur cost of \$50/ton and 3 mills/kwhr at \$40/ton of sulfur. Sulfur costs in developing nations like India are higher than in

the United States and range from \$60 to \$80/ton (ORNL-4290, 1968, p. 62).

In summary, the salt complex by utilizing the waste evaporator effluent stream, the abundantly available solar energy, and the electrical and steam energy from the nuclear plant can furnish basic chemicals to a developing nation including magnesium metal, can supply potash to provide a balanced fertilizer for the nuplex food factory, and can provide the caustic and hydrochloric acid for economic treatment of seawater prior to nuclear desalination.

REFERENCES

- Adwallader, E.A., 1967, Carbon Dioxide—The Key to Economic Desalination: *Ind. Eng. Chem.*, v. 59, p. 42.
- Chemical Engineering*, 1967, Caustic Soda—Too Much of a Good Thing: v. 74, no. 23, p. 128.
- Chemical Engineering*, 1969, Magnesium Push Is On: v. 76, no. 18, p. 60.
- Chemical Week*, 1968, Potash Problems on the Prairie: v. 102, no. 15, p. 47.
- Chemical Week*, 1969, Magnesium Booster: Budding Auto Market: v. 104, no. 20, p. 36.
- Coeller, H.E., 1969, Tables for Computing Manufacturing Costs of Industrial Products in an Agro-Industrial Complex: ORNL-4296.
- Cockson, W.H., 1969, 1968 Mineral Review—Aluminum and Magnesium: *Can. Mining J.*, v. 40, no. 2, p. 116.
- Marshall, W.L., and Slusher, R., 1968, Aqueous Systems at High Temperature. Solubility of Calcium Sulfate and Its Hydrates in Seawater and Saline Water Concentrate and Temperature Concentration Limits: *J. Chem. Eng. Data*, v. 13, p. 83.
- Moditz, H.E., Hein, H.F., and Adamec, A.W., 1967, Electric Heating and Development of Electric Energy Consumption With Special Reference to Modern Industrial Thermal Processes: in *World Power Conference on Problems of Future Years in Energy Utilization*, Tokyo, Japan, Oct. 16-22, 1966, v. 5, p. 2405.
- Rock Ridge National Laboratory, 1968, Nuclear Energy Centers: Industrial and Agro-Industrial Complexes: ORNL-4290.
- Office of Saline Water, 1965, Saline Water Conversion Report, 1965: Report NP-16361, p. 218.
- Office of Saline Water, 1966, LMC Process Development of Precipitation Processes for Removal of Scale Formers from Seawater: Research and Development Progress Report 192.
- Ritchey, J.A., 1969, Problems in Implementation of an Agro-Industrial Complex, ORNL-4295.
- Simon-Carves Chemical Engineering Ltd., England, personal communication, March 1968.
- Smith, E.E., and Andrews, H.J., 1967; Mining Great Salt Lake—A \$75 Billion Reserve of Lithium, Magnesium, Potash, Sodium Salts: A.I.M.E. Annual Meeting, Feb. 19-23, 1967, Los Angeles, Calif., American Institute of Mining, Metallurgical and Petroleum Engineers, Inc., New York.
- Squires, A.M., and Lobo, W.E., 1969, Nuclear Energy Centers: I. Steelmaking in an Industrial Complex and II. Acetylene Production From Naphtha By Electric Arc and By Partial Combustion: ORNL-4294.
- Stout, P., 1969, Potential Agricultural Production From Nuclear-Powered Agro-Industrial Complexes Designed for the Upper Indo-Gangetic Plain: ORNL-4292.
- Tamura, T., and Young, W.J., 1969, Data Obtained On Several Possible Locales for the Agro-Industrial Complex: ORNL-4293.
- United Nations Publication E/CN.11/635, 1963, The Alkali Industry in India by P.K. Seshan in *Proceedings of the Seminar on the Development of Basic Chemical and Allied Industries in Asia and the Far East* (held from 3 to 13 October 1962), p. 356.
- Yee, W.C., 1968, Caustic-Chlorine Production, Seawater Treatment and Nuclear Desalination: paper presented at 133rd National Meeting of the Electrochemical Society, Boston, Mass., May 5-9, 1968.
- Young, G., and Michel, J.W., 1968, Nuclear Energy Centers: Industrial and Agro-Industrial Complexes—Summary Report: ORNL-4291.

The author is grateful to a number of individuals who supplied the technical information for this paper: J.E. Currey, Hooker Chemical Company; V. deNora, Oronzio deNora Impianti Electrochimici, Italy; S. Fougner, Magnalith Corporation, New York; J. Holtzapfel, Kaiser Aluminum and Chemical Corporation; J.D. McFarlane, Simon-Carves Chemical Engineering Ltd., England; J. Nichols, Monsanto Company; A.Z. Richards, Caldwell, Richards and Sorenson, Inc., Salt Lake City; and J.L. Ryon, International Salt Company.