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ENERGY BY REVERSE ELECTRODIALYSIS

MASTER

Final Report to the

DEPARTMENT OF ENERGY
Contract EG-77-C-05-5544

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ENERGY BY REVERSE ELECTRODIALYSIS

I. INTRODUCTION

In principle, energy can be derived from the difference between the chemical potentials of concentrated and dilute salt solutions. Energy is thus available wherever a brine and less salty water exist in proximity. Among the potential sources are the brines of oil and natural gas fields, the brines from flooding of salt domes, the brines of salt lakes, and even the waters of oceans and bays, provided less salty water is also nearby.

The potential of these sources of energy has long been recognized, but the economic conversion of the energy to a useful form has yet to be realized.

Of the many approaches to the recovery of this energy, one which appears promising, but inadequately explored, is "reverse electrodialysis", which is the reverse of the well-known process for desalination. In electrodialysis, many cation- and anion-exchange membranes are arranged alternately in a parallel array to form thin solution compartments, or cells, between an pair of electrodes. When a voltage is applied across solutions flowing through the cells, salt passes from alternating cells into adjoining cells, and the result is a desalination of one stream with enrichment of the other.

In reverse electrodialysis, as the term suggests, the process is reversed. When a concentrated salt solution flows through alternating cells and fresh water flows through the others, a voltage is generated across each ion-exchange membrane. The back-emf represented by the electrode reactions is small compared to the sum of the transmembrane voltages, which are additive. Useful power may accordingly be obtained from leads to the electrodes.

II. THE REVERSE ELECTRODIALYSIS PROCESS

Figure 1 is a diagram showing the main components of a reverse electrodialysis (RED) stack. In the stack, many cation-exchange (C) and anion-exchange (A) membranes are arrayed alternately between a pair of electrodes. Brine and dilute solutions flow through the spaces

LEGEND: C - CATION-EXCHANGE MEMBRANE
A - ANION-EXCHANGE MEMBRANE

Figure 1. Diagram of a reverse electrodialysis stack.

between the membranes, with the brine flow alternating with the flow of dilute solutions. Electrode rinse solutions flow through the compartments adjacent to the electrodes. These electrode solutions might be the brine or solutions of selected electrolytes.

Because of the difference between the chemical potentials of the salt ions in the brine and in the dilute solutions, there will be voltages across each membrane, and these voltages are additive. When there are many membranes and solution compartments between each pair of electrodes, the back-emf represented by the electrochemical reactions at each electrode will be small relative to the sum of the additive voltages across the membranes. Each solution and each membrane and other irreversibilities, such as diffusion potentials in the solutions, constitute an electrical resistance. The sum of these resistances and irreversible effects is the internal resistance of the RED battery. The peak power is delivered to an external load when the load impedance equals the internal impedance of the battery.

A. Sources of Brines

A number of brines might be good sources of energy from RED. According to Wick and Isaacs¹, the potential energy in the salt from salt domes of oil wells exceeds that of the oil itself, and vast amounts of energy are also potentially available from salt pans along arid seacoasts, from subterranean brines and salt deposits, and from hypersaline lakes (such as the Great Salt Lake). Hypersaline lakes or aquifers are renewable sources, since they are formed by solar evaporation.

Table I shows normalities of ions in the water from the Great Salt Lake, as calculated from analyses performed by the Utah State Geological and Mineral Survey. This brine contains primarily NaCl and MgCl and is about 4.0 normal in total electrolyte concentration. It also contains low concentrations of Fe^{+2} and Mn^{+2} . These ions can be detrimental to the membranes used in reverse electrodialysis. Thus, the brine from the Great Salt Lake could be an important source of energy when combined with more dilute water in a reverse electrodialysis plant.

A preliminary search of the literature revealed other sources of potentially useful brines: in New Mexico, from the Castile formation (TDS = 324,000 ppm) and the Rustler formation (TDS = 285,000 ppm)², and in North Dakota, from the surface waters of East Devils Lake (TDS = 38,300 ppm) and Eastern Stump Lake (TDS = 68,600 ppm).

Table II shows the concentrations of ions in a brine from a gas well, as calculated from analyses supplied by Schlemmer³. The electrolyte concentration is about 1.7 N. The brine contains only a trace of the possibly detrimental Fe^{+2} , but it contains small amounts of silicates and borates, which might be detrimental to membranes. The composition of the brine given in Table II indicates that brines from oil and gas wells might also be a source of energy.

TABLE I.

IONS IN WATER FROM THE GREAT SALT LAKE
(Typical Compositions of North and South Ends of Lake)

<u>Ion</u>	North End	South End
	<u>Concentrations,</u> <u>normality</u>	<u>Concentrations,</u> <u>normality</u>
<u>Cations</u>		
Ca ⁺²	0.01	0.01
Mg ⁺	0.92	0.77
Na ⁺	3.03	3.07
K ⁺	0.16	0.14
Li ⁺	0.01	0.01
Major Cations *	<u>4.13</u>	<u>4.00</u>
<u>Anions</u>		
HCO ₃ ⁻	0.01	0.01
SO ₄ ⁻²	0.47	0.42
Cl ⁻	3.58	3.46
Major Anions **	<u>4.06</u>	<u>3.89</u>

* The concentrations of Fe⁺² and Mn⁺² are extremely low.

** Also contains F⁻, BO₃⁻³, and SiO₃⁻².

TABLE II.

CONCENTRATIONS OF IONS IN A BRINE FROM
A NATURAL GAS WELL
(From the Delcambre No. 1 Well Near Delcambre, Louisiana)

<u>Ion</u>	<u>Concentration, normality</u>
<u>Cations</u>	
Ca ⁺²	0.08
Mg ⁺²	0.01
Fe ⁺²	Trace
Zn ⁺²	Trace
Sr ⁺²	Trace
Na ⁺	1.62
K	0.01
Major cations	<u>1.72</u>
<u>Anions</u>	
HCO ₃ ⁻	0.01
Cl ⁻	1.61
Major anions*	<u>1.62</u>

* Also contains SiO₃⁻² and BO₃⁻².

Although not mentioned by Wicks and Isaacs, geothermal brines are also potential sources of energy. The composition of a typical geothermal brine⁴ is shown in Table III. The normality of the total salts is about 3.5. This particular geothermal brine contains appreciable amounts of Fe⁺², Mn⁺², BO₃⁻³, NH₄⁺, and H₂S and probably would require pretreatment to remove these before reverse electro-dialysis. Nevertheless, because of the high concentration of ions, geothermal brines also merit consideration as sources of energy.

Such brines are now being studied as a heat source. Klei and Maslan⁵ state that the initial capital cost for a thermal power plant based on flashed steam from geothermal hot brines is from \$180 to \$200/KW, of which the wells and piping amount to about \$65/KW. Fixed charges based on the capital investment (e.g. depreciation, taxes, insurance, and return) account for 80% to 90% of the electrical production costs.

TABLE III.

TYPICAL COMPOSITIONS OF SALTON SEA GEOTHERMAL BRINE

<u>Ions</u>	<u>Concentration, normality</u>
<u>Cations</u>	
Fe ⁺²	0.063
Mn ⁺²	0.041
Zn ⁺²	0.013
Ca ⁺²	1.120
Na ⁺	1.753
K ⁺	0.361
Li ⁺	0.003
NH ₄ ⁺	0.023
	<hr/> 3.377*
<u>Anions</u>	
HCO ₃ ⁻	0.026
Cl ⁻	3.490
BO ₃ ⁻³	0.016
	<hr/> 3.532**

* Also contains SiO₂, As, Pb, Mg, Sn, Ba, Cu, Rb, and Cs in minor amounts.

** Also contains SO₄, F, Br, I, and H₂S in minor amounts.

Additional power could be generated from these plants by reverse electrodialysis, which could use the brine (after flashing of the steam) and the condensed steam from the turbine. The warmth of the two streams is shown later to be of advantage in reverse electrodialysis.

B. Prior and Continuing Research Related to RED

The idea of converting the difference between the chemical potentials of concentrated and dilute solutions to useful energy by reverse electrodialysis was first studied by Pattle^{6,7} in 1953 and 1954. In 1955, Murphy⁸ used reverse electrodialysis as a direct source of energy to desalt brackish water by the osmionic process. Later, Lacey^{9,10} revised the theoretical expressions which had been developed by Murphy; the new equations were used to predict performance for the osmionic process, and the predictions were consistent with the results of experiments performed in bench-scale osmionic demineralizers. Weinstein and Leitz¹¹ developed simplified equations to predict the power output from reverse electrodialysis units, and showed that experimental data obtained with a small reverse-electrodialysis unit agreed with their predictions of gross power outputs.* However, they made a number of simplifying assumptions in the development of the equations, and did not include factors such as concentration polarization, water transfer by osmosis and electroosmosis, and the power required to pump solutions.

In 1965 Lacey¹² and in 1968 Belfort and Guter¹³ developed equations to predict the performance of electrodialysis units. These equations included the effects of concentration polarization at membrane surfaces and of current densities on diffusion potentials, membrane potentials, and internal I^2R losses. Both groups of research workers demonstrated that predicted performances were consistent with experimental results. These equations have been revised by changing the direction of the concentration gradients at membrane surfaces to allow prediction of the gross power output from reverse electrodialysis units.

Later, in 1975, Forgacs^{14,15} developed equations to predict the performance of reverse electrodialysis units for generating power. These equations were identical in form to the revised equations. However, Forgacs made the simplifying assumptions that concentrations (not activities) would be used, that the effects of concentration polarization would be ignored, and that the membranes would be perfectly permselective. Despite these assumptions, the gross power outputs calculated from the equations corresponded fairly closely to data from experiments made with a NaCl brine which simulated the electrolyte concentration of Dead Sea brine and a more dilute solution which simulated seawater. The calculated gross power outputs were slightly higher than the experimental values, as would be expected from the nature of the assumptions.

* Gross power output does not include the power required to pump solutions through the solution compartments.

Forgacs also showed by theory based on the free energy of mixing that 60% to 80% of the available energy is extracted when the mixing of a brine and a more dilute water occurs to the extent of 30 to 40% in a RED cell. He also showed that the volume of dilute water must be ample, relative to the volume of brine available. The following were also discussed by Forgacs:

- The development of ion-exchange membranes with extremely low resistances (0.4 to 0.6 ohm-cm^2) and high transference numbers (0.85 or higher), which can be made for $\$1.00/\text{sq m}$ ($\$0.093/\text{sq ft}$).
- The development of RED stacks with ultrathin compartments by a design amenable to large-scale production.
- An investigation of cell configurations and their effects on power outputs from which the conclusion can be drawn that cells with short path lengths for the flow of solutions (10 to 50 cm) will probably be optimum.
- A study of pumping power from which the conclusion was drawn that the optimum solution velocities would be less than 1 cm/sec .

In 1976, Emren and Bergstrom¹⁶ began work on reverse electrodialysis as an energy source in Sweden. They are working primarily to develop RED units for use in mixing seawater and river water, since there are many rivers entering the sea on the west coast of Sweden. These workers also developed a set of equations to predict performance, and their equations are identical in form to those mentioned previously. In contrast to Forgacs, these workers, like Lacey and Guter and Belfort, use activities of the solutions instead of concentrations. However, the Swedish workers make the simplifying assumption that the effects of concentration polarization at the membrane surfaces are negligible, whereas Lacey introduces the effects of concentration polarization into his theory.

Emren and Bergstrom have developed conceptual designs for RED cells and stacks and have developed special thin membranes which have about 2 ohm-cm^2 resistances, when equilibrated in seawater. They are in the process of developing laboratory models of RED stacks based on their conceptual designs. These workers estimate an energy cost of $\$0.035$ to $\$0.040/\text{Kwh}$, and a capital cost of $\$725/\text{KW}$. These values may be overly optimistic for the costs of a plant in the United States that would use seawater and river water.

C. Equations for Predicting Performance of RED Units

The equations for predicting gross power output are expressed in terms of the voltages and resistances in a cell pair. A cell pair, which is the basic repeating unit in reverse electrodialysis, comprises one anion-exchange membrane, one cation-exchange membrane, one concentrated solution, and one dilute solution, as indicated in Figure 2.

Figure 2 shows the edges of spaced ion-exchange membranes that form a cell pair. When current flows at steady-state, there are concentration gradients in the nearly static solutions in the boundary layers adjacent to the membrane surfaces. For the development of the equations, the Nernst idealization of completely static boundary layers and completely mixed bulk solutions was used. Since this idealization has been used in other research on electrodialysis, the boundary-layer thicknesses at various solution velocities can be taken from those references. In Figure 2, the concentration gradients that exist within the solutions are indicated by lines 1-2-3-4, and 5-6-7-8.

The voltage output from a cell pair is given by Equation 1. On the right side of Equation 1, the first term in brackets is the voltage component, which is composed of four diffusion potentials and two membrane potentials. The second term in brackets is the voltage drop through the resistive elements of a cell pair (*i.e.*, the product of the current density, *i*, and the sum of the resistances of each boundary layer, each bulk solution, and each membrane). A detailed derivation of the equation is given in the Appendix.

$$\begin{aligned}
 E_o = & \left[-\frac{RT}{F} \left\{ (t_d^+ - t_d^-) \ln \frac{\gamma_1 N_1}{\gamma_d N_d} + (t_d^+ - t_d^-) \ln \frac{\gamma_d N_d}{\gamma_4 N_4} + (t_{AM}^+ - t_{AM}^-) \ln \frac{\gamma_4 N_4}{\gamma_5 N_5} \right. \right. \\
 & \left. \left. + (t_c^+ - t_c^-) \ln \frac{\gamma_5 N_5}{\gamma_c N_c} + (t_c^+ - t_c^-) \ln \frac{\gamma_c N_c}{\gamma_8 N_8} + (t_{CM}^+ - t_{CM}^-) \ln \frac{\gamma_8 N_8}{\gamma_1 N_1} \right\} \right] \\
 & - \left[i \left\{ \frac{\delta}{\frac{N_1 + N_d}{2} \Lambda_d} + \frac{\Delta - 2\delta}{N_d \Lambda_d} + \frac{\delta}{\frac{N_d + N_4}{2} \Lambda_d} + R_{AM} \right. \right. \\
 & \left. \left. + \frac{\delta}{\frac{N_5 + N_c}{2} \Lambda_c} + \frac{\Delta - 2\delta}{N_c \Lambda_c} + \frac{\delta}{\frac{N_c + N_8}{2} \Lambda_c} + R_{CM} \right\} \right] \quad (1)
 \end{aligned}$$

C – CATION-EXCHANGE MEMBRANE
 A – ANION-EXCHANGE MEMBRANE
 δ – REPRESENTS IDEALIZED THICKNESS OF BOUNDARY LAYERS
 N – REPRESENTS CONCENTRATIONS OF SOLUTIONS

SUBSCRIPTS 1, 4, 5, AND 8 DESIGNATE THE INTERFACIAL POINTS SHOWN.
 SUBSCRIPTS d AND c REPRESENT THE DILUTE AND CONCENTRATED SOLUTIONS RESPECTIVELY.

Figure 2. Representation of a reverse electrodialysis cell pair and the concentration gradients at membrane surfaces.

where:

E_o = voltage output from one cell pair, volts

i = current density, amp/cm²

R = gas constant 8.314 joules/°K

t_d^+ and t_d^- = transference number of cation and anion in dilute solution, respectively

t_c^+ and t_c^- = transference number of cation and anion in the concentrated solution, respectively

t_{AM}^+ and t_{AM}^- = transference number of cation and anion in the cation-exchange membrane, respectively

t_{CM}^+ and t_{CM}^- = transference number of cation and anion in the anion-exchange membrane, respectively

N_1, N_4, N_5, N_8 = interfacial concentrations at points 1, 4, 5, and 8 in Figure 1 (calculated by Equations 2, 3, 4, and 5), eq/cm³

N_d = concentration of the bulk of the dilute solution, eq/cm³

N_c = concentration of the bulk of the concentrated solution, eq/cm³

$\gamma_1, \gamma_4, \gamma_d, \gamma_5, \gamma_c,$ and γ_8 = activity coefficients for the concentrations designated by the subscripts

Λ_d, Λ_c = equivalent conductances of the dilute and concentrated solutions, respectively, cm²/(ohm) (eq)

Δ = spacing between membranes, cm

δ = thickness of boundary layers, cm

R_{AM} and R_{CM} = resistances of the anion- and cation-exchange membranes, ohm-cm²

The equations to calculate the interfacial concentrations, $N_1, N_4, N_5,$ and N_8 , are Equations 2, 3, 4, and 5.

$$N_1 = N_d + \frac{(t_{CM}^+ - t_d^+) \delta i}{FD} \quad (2)$$

$$N_4 = N_d + \frac{(t_{AM}^- - t_d^-) \delta i}{FD} \quad (3)$$

$$N_5 = N_c - \frac{(t_{AM}^- - t_c^-) \delta i}{FD} \quad (4)$$

$$N_8 = N_c - \frac{(t_{CM}^+ - t_c^+) \delta i}{FD} \quad (5)$$

where: F = the Faraday, 96,500 coul/eq
D = the diffusion coefficient for the salt in solution, cm²/sec and the other symbols have been defined previously.

With Equations 1 through 5, the voltage per cell pair, E_o , can be calculated for any selected value of current density, i . The product of E_o and i gives the gross power output in w/cm².

With this procedure, one can show the sensitivity of E_o and gross power output to current density. By selecting a range of values of i , one can determine the optimum value of current density and power output along with values of E_o and the voltage and resistive components of E_o for each value of i . From examination of the voltage and resistive components of E_o for a range of current densities, the effects of boundary-layer thickness, δ , and current density, i , on the interfacial concentrations, N_1 through N_8 , and thus on the voltage component of E_o and power output can be seen, as well as the effects of spacing between membranes, Δ , and membrane resistances on the resistive component of E_o . These data provide a basis for evaluating the factors that will effect the largest decreases in the cost of power.

The gross power output is:

$$P_g = E_o i \quad (6)$$

To achieve a given thickness of boundary layers, δ , each solution must be pumped through each cell. The power required to pump the solutions, P_p , must be furnished by the reverse electrodialysis device. Therefore, the net power output, P_n , is

$$P_n = P_g - P_p \quad (7)$$

The pumping power required to achieve given thicknesses of the boundary layers is obtained from graphs such as those shown in Figures 3, 4, and 5.

The net power output (gross power minus pumping power) can also be predicted. The pressure-drop through solution compartments varies with the spacing between membranes, the geometry of spacer materials, and the velocity and viscosity of solutions. Data in publications¹⁷⁻²⁷ show the relationships between pressure drops and solution velocities for open channels and a number of different cell-spacer materials. These publications also contain data from which the thickness of boundary layers can be calculated for varying solution viscosities and velocities for types of spacer materials* and spacings between membranes. From a comparison of the thicknesses of boundary layers attainable with each spacer material as functions of pumping power (calculated from pressure-drop data), it was possible to select the spacer material that gave the thinnest boundary layers per unit of pumping power.

The power required for pumping solutions through solution compartments equipped with this one type of spacer material has been introduced into the basic set of equations for each solution velocity studied. Data on pressure drop per unit of length for this type of spacer were obtained from reports on the development of forced-flow electrodialysis.¹⁷⁻¹⁹

Examination of Equation 1 shows that the voltage component of E_o is proportional to the logarithm of the ratio of salt concentrations in the dilute and concentrated solutions. There is a limit on the salt concentration usable in the concentrated solution because the transference numbers of counterions in ion-exchange membranes decrease rapidly when the membranes are bathed by solutions with salt concentrations above some value determined by Donnan equilibrium. The concentration at which the rapid decrease begins depends on the fixed-charge density of the membrane, which varies from one type of membrane to another. Membranes with high fixed-charge densities are desirable,

* In conventional electrodialysis stacks, the membranes are held in proper position relative to each other by mesh-like spacer materials, which not only support and space the membranes, but also aid in lateral mixing of the solutions.

Figure 3. The effect of solution velocity on the thickness of boundary layers in the Aqua-Chem 15/5 Forced-Flow Electro-desalination Stack.

Figure 4. The effect of solution velocity on the pressure drop in the Aqua-Chem 15/5 Forced-Flow Electrodialation Stack.

Figure 5. The pumping power required to obtain various thicknesses of boundary layers in the Aqua-Chem 15/5 Forced-Flow Electrodialysis Stack.

but even with these there is a maximum value of brine concentration that would be desirable. Since there is a limit on the concentration of the brine, and the value of E_o depends on the ratio of concentrations, the use of extremely dilute solutions would be desirable to achieve the highest values of E_o , if it were not for the extremely high resistances of such dilute solutions.

Temperature affects several variables in Equations 1 through 5, including activity coefficients, transference numbers of counterions in membranes, transference numbers in solutions, resistances of membranes, diffusivity, and solution viscosity. Changes in solution viscosity will affect the power required for pumping solutions. The effects of temperature on equivalent conductances have been calculated by the use of Onsager equations and shown to be consistent with experimental results. Information has also been published on the effects of temperature on membrane resistances, and the transference numbers of counterions and water in a number of ion-exchange membranes.²⁸⁻³¹ With data from the preceding references, the effects of temperature on the power output from reverse-electrodialysis units were estimated.

Some of the data we used on membrane resistances and the transference numbers of counterions through membranes were measured in our laboratories and are presented in Figures 6 and 7.

D. Effects of Variables on Power Output From RED Cells

Equation 1 relates the output voltage from an RED cell, E_o , to the total voltage developed by the difference between the chemical potentials of the brine and dilute solutions (the first term in brackets) and to the voltage drop across the internal resistances of the cell (the second term in brackets). This internal voltage drop is a function of current density, and the total voltage developed by a cell is primarily a function of the interfacial concentrations at the opposite surfaces of each membrane and of the transference numbers of the counterions (*i.e.* the ions with charges opposite to the fixed charges within the membrane). The interfacial concentrations depend on the concentrations in the bulk of the solutions, N_d or N_c , and on the transference numbers (t_{CM}^+ , t_{AM}^+ , t_d^-), current densities (i), and thicknesses of boundary layers (δ), as described by equations 2 through 5. In turn, the thicknesses of the diffusive boundary layers depend on the mixing characteristics of cell spacers and on the solution velocities.

It is important to assess the effects of interactions of these various process parameters on the total voltage developed in RED cell, the IR drop within the cell, and the output voltage from a cell (E_o), which is the total voltage component minus the voltage drop across the internal resistance, because the gross power output of a cell is the product of the output voltage and the current density.

Figure 6. Areal resistance of anion- and cation-exchange membranes (AV-3T and CL-2.5T) made by Tokuyama Soda Co., Ltd.

Figure 7. Transference number of counterions in anion- and cation-exchange membranes (AV-3T and CL-2.5T) made by Tokuyama Soda Co., Ltd.

As an initial step, therefore, we have plotted changes in the total voltage component and the voltage drop across internal resistance as functions of various controllable variables. These plots were developed for a particular set of conditions. They are intended only to illustrate the effects of a change in each controllable variable and do not necessarily represent optimum conditions.

Figure 8 shows the effects of increasing current densities on the total voltage component, the iR voltage drop, and the voltage output. The assumptions are given in Table IV.

TABLE IV.

CONDITIONS ASSUMED FOR THE FIRST ILLUSTRATIVE EXAMPLE

Brine compartment thickness	=	0.10 cm
Dilute compartment thickness	=	0.05 cm
Concentration of brine solution	=	4.00 N
Concentration of dilute solution	=	0.10 N
Solution velocities	=	0.50 cm/sec
Solution temperatures	=	298°K
Resistance of each membrane	=	1.0 ohm-cm ²
Current density	=	Varies

The total voltage component decreases linearly with increasing current density. This decrease results from changes in the interfacial concentrations (N_1 , N_4 , N_5 and N_8) as described by equations 2 through 5. The increase in the iR voltage component with current density (in the negative direction) is not quite linear primarily because the average concentrations in the diffusive boundary layers in the dilute compartments increase with current density. Thus, their resistances decrease slightly.

Figure 9 shows the change in the gross power output and the net power output (i.e. gross power output minus pumping power) for the conditions specified in Table I. With the assumed low solution velocities of 0.50 cm/sec, the power required for pumping solutions is very low (0.048 milliwatts/cm²). Therefore, the curves for gross and net power output are almost superimposed. With a decreasing net output voltage at increasing current densities, the peak output power depicted in Figure 9 is to be expected.

Figure 10 shows the effects of changes in the thickness of the dilute compartment on the total voltage component, the iR voltage component, and the net voltage output in an RED cell. The conditions were those given in Table I, except that a current density of 10 ma/cm² was assumed and the thickness of the dilute chamber was a variable.

Figure 8. Relationship between the current density and the total voltage component, the iR voltage component, and the net voltage output for conditions specified in Table I.

Figure 9. The effects of current density on gross and net power output from a RED cell at conditions specified in Table I.

Figure 10. Effects of variations in the thickness of the dilute compartment on the total voltage component, the iR voltage component, and the net voltage output of a RED cell for specified conditions.

The total voltage component is invariant, but the iR voltage component increases (in a negative direction) rapidly with increasing thickness of the dilute chamber. Thus, the net voltage output decreases. The thickness of the dilute compartment is a critical factor in the design of RED cells. These compartments should be made as thin as possible if low solution velocities are used so that pumping power is not exorbitant.

Figure 11 shows the effects of changes in the thickness of the brine compartment on the total voltage component, the iR voltage component, and the net voltage output from an RED cell. The conditions assumed for this illustrative example are the same as given in Table I except that the current density is assumed to be 10 ma/cm^2 and the thickness of the brine compartment is assumed to be varied.

The total voltage component is invariant because with the assumed constant solution velocities and current densities the interfacial concentrations are invariant. The iR voltage component increases very slightly (in a negative direction) with increasing brine compartment thickness, since the brine solution is highly conductive and incremental increases in thickness do not add much to the internal resistance of an RED cell. Thus, the net output voltage decreases only slightly with increasing thickness of brine compartments. In contrast to the thickness of the dilute compartment, the thickness of the brine compartment is not a critical factor in the design of RED units.

Figure 12 shows the effects of the velocity of the dilute solutions on the total voltage component, the iR voltage component, and the net voltage output in an RED cell. The conditions assumed are those given in Table I, except the current density is assumed to be 10 ma/cm^2 and the velocity of the dilute stream varies. The total voltage component increases slightly with increasing velocity because of the effect of velocity on the interfacial concentrations. The iR component increases slightly (in a negative direction) because the resistances in the dilute diffusive boundary layers increase slightly with increasing velocity. The combined effect is that the net voltage output remains almost constant.

Figure 13 shows that the gross power (E_i) remains virtually constant with solution velocity. However, the net power output declines sharply because of increases in pumping power. Similar effects can be shown for variations in the velocity of the brine solution.

Figure 14 shows the effects of increasing membrane resistances on the total voltage component, the iR voltage component, and the net voltage output from an RED cell. The conditions are those in Table I, except the current density was assumed to be constant at 10 ma/cm^2 , and the membrane resistance was a variable. The total voltage component

Figure 11. The effects of the thickness of the brine compartment on the total voltage component, the iR voltage component and the net voltage output from a RED cell under specified conditions.

Figure 12. The effects of the velocity of the dilute solution on the total voltage component, the iR voltage component and the net voltage output in a RED cell for specified conditions.

Figure 13. Effect of the velocity of the dilute solution on the power output from a RED cell for specified conditions.

Figure 14. The effects of membrane resistances on the total voltage component, the iR voltage component, and the net voltage output in a RED cell for specified conditions.

is invariant, and the iR voltage component increases linearly (in a negative direction) with increasing current density, as expected. Therefore, the net voltage output decreases linearly. Low-resistance membranes are indicated to be desirable.

Figure 15 shows the effects of variations in the brine concentration on the total voltage component, the iR component, and the net voltage output when the concentration of the dilute solution is 0.1 N , the current density is 10 ma/cm^2 , and the other conditions are those given in Table I. The increase in the total voltage component reflects both changes in brine concentration and changes in transference numbers through the membranes. The change in the iR voltage component primarily reflects the decreasing resistances of the increasing brine concentrations, and the changes in both voltage components show up in the variations of the net voltage output.

Figure 16 shows the effects of variations in brine concentration on the gross power and net power outputs from an RED cell. As might be expected the net power output is very close to the gross power output because the power required for pumping is low at the low velocity assumed (0.5 cm/sec). It is also obvious that the assumed concentration of 0.1 N for the dilute solution and a current density of 10 ma/cm^2 are not advantageous for use with dilute brines, such as seawater (about 0.6 N).

In summary, some design factors and process variables are of extreme importance to the net power output from RED cells, others are of moderate importance, and still others are not critical. The thickness of the dilute compartment is highly important. The thinnest compartments possible combined with low solution velocities are desirable. In contrast, the thickness of the brine compartments has little effect on the output voltage, when the brine concentration is 4.0 N , as it was in the illustrative example. As the brine concentration decreases, the resistance of the brine increases. Therefore, with lower concentrations of brine the importance of the thickness of the brine compartment will increase.

As shown in Figure 13 the velocity of the dilute solute through 0.05-cm compartments is an important process variable. Although the net voltage and gross power outputs are not affected much by the velocity (Figures 12 and 13), the net power output decreases markedly because the pumping power increases with solution velocity. With thinner dilute compartments, it will be even more important to use low solution velocities. Similar considerations hold for the velocity of the brine solution, although the effects of brine velocity on net power output might be less critical because thicker compartments with lower hydraulic resistance can be used.

Figure 15. The effects of variations in brine concentration on the total voltage component, the iR voltage components, and the net voltage output from an RED cell for specified conditions.

Figure 16. The effects of changes in brine concentrations on the gross and net power output from an RED cell for specified conditions.

The membrane resistance is a relatively important variable, since the resistances of the membranes contribute to the internal resistance of the cell, and thus to the portion of the total voltage dissipated in internal resistance. Membranes with as low resistances as possible are desired, if the selectivities, strengths, and handling characteristics are acceptable.

The ratio between the concentration of brine and that of dilute solution should be high to achieve high net voltage outputs and thus high gross power outputs. However, because of the decrease in transference numbers with increasing brine concentrations, there is some brine concentration at which the gross and net power outputs are optimum (about 4.5 N in Figure 16). For maximum power output, it is advantageous to have a high ratio of the brine concentration to the dilute concentration when both solutions are relatively concentrated and therefore low in resistance. However, useable power outputs might be obtained from more dilute solutions (e.g. seawater and river water) if the compartments are thin and membrane resistance is low.

Based on the design factors and process variables indicated to be important by studies such as those discussed above, we studied the effects of various important parameters on net power output to determine interactions between variables.

Figure 17 shows the variation of net power output with current density with the resistance of the membranes varied from 0.1 ohm-cm² to 1.5 ohm-cm². The conditions assumed for the illustration in Figure 17 are stated in the figure. As the resistances of the membranes are decreased, the peak net power output increases, and the current density at which the peak outputs occur increases, as indicated by the dotted line in Figure 15. Both the increase in peak power outputs and the current densities at the peaks are important factors in determining the areas needed for an RED plant of specified output. These increases result from decreases in the internal resistance of the RED cells, of course.

Figure 18 shows the variation of net power output with current density and the spacing in the dilute compartment for the other assumed conditions specified in the figure. As the thickness of the dilute compartments is decreased, the internal resistance of the RED cells decreases and the net power output increases. Again, the current densities at which the peak output powers occur increase with decreasing thickness of cells (i.e. decreasing internal resistance).

If Figures 17 and 18 are compared, it is seen that decreases in the thickness of the dilute compartments result in greater increases in net power output than do decreases in membrane resistance. This is true for the conditions assumed but it is not necessarily universally true, since both factors contribute to internal resistance.

Figure 19 shows the variation of net power output with current density and the concentration of the dilute solution for the other assumed conditions specified. As the concentration of the dilute solution is increased from 0.05 N to 0.1 N the peak power output

Figure 17. Net power output as a function of current density and membrane resistance for specified conditions.

Figure 18. Net power output as a function of current density and spacing in the dilute compartments for specified conditions.

Figure 19. Net power output as a function of current density and concentration in the dilute compartments for specified conditions.

increases. There is a decrease in both the total voltage component and the internal resistance of the RED cells. For the conditions assumed, the decrease in the internal resistance is more important than the decrease in the total voltage component. However, as the concentration of the dilute solution is varied from 0.1 N to 0.2 N there is a slight decrease in net power output. In this range the change in the total voltage component just about balances changes in internal resistance. As the concentration is increased from 0.2 N to 0.4 N (dotted line in Figure 19) the peak net power output decreases. In this range, for the conditions assumed, the decrease in the total voltage component is more important than the decrease in internal resistance.

Figures 20, 21, 22 and 23 show the variations of net power output with current density and temperature for 1.0 N, 2.0 N, 3.0 N, and 4.0 N brines, respectively. The non-variable conditions assumed are stated in each figure. In all four figures the effects of increased temperature in reducing internal resistance and increasing net power output can be seen.

Comparison of Figures 20, 21 and 22 shows that as the concentration of the brine increases from 1.0 N to 3.0 N the peak net power outputs increase considerably. However, as the concentration of brine is increased from 3.0 N to 4.0 N, the increase in net power output is very slight. From a perturbation in which the brine concentration is increased to 5.0 N, there is an actual decrease in net power output as the brine concentration is increased from 4.0 N to 5.0 N. This leveling off and subsequent decrease are mainly because the decrease in transference numbers of counterions in the membranes at high brine concentrations results in an attendant decrease in the total voltage component.

From the preceding considerations, we chose what we believe to be optimum or near-optimum conditions for operation of RED cells for specific selected brines.

From graphs such as those in Figures 12, 13 and 14, we took the peaks of net power outputs and the current densities at those peaks. We assumed in each instance that the brine concentration would be decreased by 30% during its passage through the RED cells. A decrease in concentration of 30% will result in an extraction of about 60% of the total potential energy initially represented by the brine. We then used the average concentration of the brine to calculate the power output from RED cells. For example, in the first specific case (a brine from a salt dome) we assumed that during steady-state pumping through the salt bed in the dome, the brine entering the RED cells would be 4.7 N NaCl. The brine leaving the RED cells would be 3.3 N, and the average normality would be 4.0 N. The concentration used in the calculations was 4.0 N.

Figure 20. Net power output as a function of current density and temperature for specified conditions (1 N brine)

Figure 21. Net power output as a function of current density and temperature for specified conditions (2N brine).

Figure 22. Net power output as a function of current density and temperature for specified conditions (3.0 N brine).

Figure 23. Net power output as a function of current density and temperature for specified conditions (4.0 N brine).

Figures 24, 25 and 26 show the variations with membrane resistance of the peak net power outputs and the current densities at which the peak outputs occur. Figure 24 gives data for 298°K, Figure 25 gives data for 323°K, and Figure 26 gives data for 338°K. The spacing in the dilute compartment assumed for these calculations was 0.01 cm, which we believe is about the thinnest practical spacing. Forgacs^{13,14} has developed cells with this spacing. In each figure it can be seen that the peak power output and the current density at peak output both increase with decreasing membrane resistance. When Figures 24, 25, and 26 are compared, the peak power outputs and the current densities at the peak outputs increase with increasing temperature. Both the increases with decreasing membrane resistances and with increasing temperature are because of decreasing internal resistances. The maximum temperature we studied was 338°K (65°C) because the electrochemical properties of the currently available membranes begin to degrade at temperatures of 70°C and higher.

By perturbations of variables such as those shown in Figures 22, 23 and 24, we studied four specific cases: (1) brine from salt domes, (2) brine from the Great Salt Lake, (3) brine from a natural gas well, and (4) sea water. The initial normality assumed for brine from a salt dome has already been discussed. The composition of brine from the Great Salt Lake was given in Table I. The composition of the brine from a natural gas well was given in Table II, and the concentration of salt in sea water was assumed to be 0.6 N. For each specific case we assumed a 30% reduction in the concentration of the brine (about 60% of the potential energy recovered). For each case, we assumed a spacing in the brine compartments of 0.1 cm and a spacing in the dilute compartments of 0.01 cm. The velocity of the brine was assumed to be 0.5 cm/sec, and the velocity of the dilute solution was assumed to be 5 cm/sec.

We selected three membrane resistances for these calculations: 1 ohm-cm² to represent the best membranes commercially available, 0.5 ohm-cm² since Forgacs^{13,14} has developed membranes on an experimental basis with this resistance, and 0.3 ohm-cm² as a reasonable research goal.

Table V shows the peak net power outputs and the current densities at which the peak outputs occur for the three selected membrane resistances and for 298°K, 323°K and 338°K for sea water. The optimum concentration of the dilute solution is also shown.

Table VI shows the same data for brine from a natural gas well, Table VII for brine from the Great Salt Lake, and Table VIII for brine from a salt dome.

These data were used in estimating the costs for energy given in the next section of this report.

Figure 24. Variation of peak power output and the current density at which the peak output occurs with membrane resistance at 298°K.

Figure 25. Variation of peak power output and the current density at which the peak output occurs with membrane resistance at 323°K.

Figure 26. Variation of peak power output and the current density at which the peak output occurs with membrane resistance at 338°K.

TABLE V. Variation of net peak power output and the current density at the peak output with temperature for three selected membrane resistances and seawater as the brine.

(Concentration of brine = 0.51 N, thickness of brine compartment = 0.1 cm, thickness of dilute compartment = 0.01 cm, velocity of brine = 0.5 cm/sec, velocity of dilute solution = 10.0 cm/sec)

Temperature, °K	Membrane resistance, ohm-cm ²	Approximate concentration of dilute solution at peak power, N	Net power output at peak, mW/cm ²	Current density at peak power, ma/cm ²
298	0.3	0.01	0.51	7
	0.5	0.01	0.49	7
	1.0	0.01	0.45	7
323	0.3	0.01	0.84	11
	0.5	0.01	0.81	10
	1.0	0.01	0.75	10
338	0.3	0.02	1.11	17
	0.5	0.02	1.06	16
	1.0	0.02	0.97	15

TABLE VI. Variation of net peak power output and the current density at which the peak output occurs with temperature for three selected membrane resistances and a brine from a natural gas well.

(Concentration of brine = 1.45 $\underline{N}$, thickness of brine compartment = 0.1 cm, thickness of dilute compartment = 0.01 cm, velocity of brine = 0.5 cm/sec, velocity of dilute solution = 10.0 cm/sec)

Temperature, °K	Membrane resistance, ohm-cm ²	Approximate concentration of dilute solution at peak power, $\underline{N}$	Net power output at peak, mW/cm ²	Current density at peak power, ma/cm ²
298	0.3	0.05	1.18	18
	0.5	0.05	1.06	16
	1.0	0.05	0.85	13
323	0.3	0.05	1.93	29
	0.5	0.05	1.76	26
	1.0	0.05	1.43	21
338	0.3	0.05	2.48	37
	0.5	0.05	2.27	34
	1.0	0.05	1.89	28

TABLE VII. Variation of net peak power output and the current density at the peak output with temperature for three selected membrane resistances and brine from the Great Salt Lake.

(Concentration of brine = 2.97 N, thickness of brine compartment = 0.1 cm, thickness of dilute compartment = 0.01 cm, velocity of brine = 0.5 cm/sec, velocity of dilute solution = 10 cm/sec)

Temperature, °K	Membrane resistance, ohm-cm ²	Approximate concentration of dilute solution at peak power, N	Net power output at peak, mW/cm ²	Current density at peak power, ma/cm ²
298	0.3	0.05	1.81	25
	0.5	0.05	1.59	22
	1.0	0.05	1.22	16
323	0.3	0.05	2.68	37
	0.5	0.05	2.39	33
	1.0	0.05	1.88	26
338	0.3	0.1	3.34	55
	0.5	0.05	2.96	42
	1.0	0.05	2.38	34

TABLE VIII. Variation of net peak power output and the current density at which the peak output occurs with temperature for three selected membrane resistances and brine from salt dome.

(Concentration of brine = 4.0 N, thickness of brine compartment = 0.1 cm, thickness of dilute compartment = 0.01 cm, velocity of brine = 0.5 cm, velocity of dilute solution = 10 cm/sec)

Temperature, °K	Membrane resistance, ohm-cm ²	Approximate concentration of dilute solution at peak power, <u>N</u>	Net power output at peak, mW/cm ²	Current density at peak power, ma/cm ²
298	0.3	0.15	1.95	29
	0.5	0.10	1.61	26
	1.0	0.10	1.12	17
323	0.3	0.10	2.73	46
	0.5	0.10	2.32	39
	1.0	0.10	1.61	27
338	0.3	0.10	3.33	57
	0.5	0.10	2.86	49
	1.0	0.10	2.22	37

Table IX shows the striking effect on peak-power outputs caused by differences in the dilute-solution compartments. Other than the thicknesses of those compartments, the conditions assumed were identical. At all three temperatures, the peak-power outputs are much higher with the thin (0.01-cm) spacing. At the elevated temperatures, the use of thin compartments results in 2.5 to 3 times the power outputs attainable with the thicker compartments (0.05 cm). The reason for this is a reduction in internal resistance.

It is clear that one of the most critical research priorities is to develop method of manufacturing membrane modules with very thin spacings between membranes in the dilute compartments.

TABLE IX. Comparison of peak-power outputs from RED cells with identical conditions other than the spacings in the dilute compartment.

(Concentration of brine - 2.97 N, velocity of brine = 0.5 cm/sec, thicknesses of dilute compartments = 0.05 and 0.01 cm, velocities of dilute solutions = 0.5 and 5 cm/sec)

Temperature, °K	Membrane resistance, ohm-cm ²	Peak power output, mW/cm ²	
		with 0.05-cm spacing in dilute compartment	with 0.01-cm spacing in dilute compartment
298	0.3	0.69	1.81
	0.5	0.64	1.59
	1.0	0.53	1.22
323	0.3	1.00	2.68
	0.5	0.93	2.39
	1.0	0.80	1.88
338	0.3	1.22	3.34
	0.5	1.14	2.96
	1.0	0.97	2.38

E. Estimates of Capital and Operating Costs of RED Power Units

The Stone and Webster Engineering Corporation³² has made an extensive study of the costs of very large (100 mgd to 300 mgd) electrodialysis plants, which we have used as guidelines for estimating costs of RED plants. In broad outline, the systems upon which

they based their cost calculations consist of a raw-water intake section, a pretreatment section, an activated-carbon filtering section to remove organic contaminants, and an electrodialysis section complete with polishing filters (5-micron), rectifiers, interstage tanks and pumps, and transmission lines, tanks, and pumps for product water storage and disposal of brine.

The Stone and Webster estimates of plant cost were done on the basis of engineering drawings and specifications at particular sites. Some of the costs are site-specific. For example, with the water at Croton Point, New York, extensive pretreatment of the water to remove possibly detrimental inorganic contaminants and activated-carbon filtration to remove organic contaminants are necessary. The combined costs for these two operations amount to 65% of the estimated total plant costs. In contrast, only activated-carbon filtration is needed to treat the water at Indian Point, New York. Since that water had only traces of possibly detrimental inorganic contaminants and only a small amount of organic contaminants, the cost of the activated-carbon portion of the plant was only 39% of the total cost.

Similarly, for RED the particular elements of cost needed for specific waters will vary since the requirements for pretreatment, organics removal, and brine disposal will vary with the composition of the brine and water being used, and with the necessity and requirements for disposal of spent brine. For example, the geothermal brine used as an illustration (see Table III) will require extensive pretreatment to remove Fe^{+2} and Mn^{+2} and other detrimental contaminants. Therefore, the cost of the pretreatment portion of the plant may be expected to be high. In contrast, brine from many salt domes is virtually pure NaCl, and would require little or no pretreatment. Investment costs for the wells to flood the salt domes will be a part of the total cost, of course.

The Stone and Webster cost estimate for individual battery-limits portions of the plants included the individual elements of cost. An example of the specificity of their cost estimates is given in the tables in Appendix II. For individual pieces of equipment they used quotations from original equipment manufacturers. For example, for the electrodialysis equipment, they append quotations of specifications and costs furnished by Ionics, Inc., the manufacturers of the equipment. Because of the attention to detail and overall completeness of the Stone and Webster cost estimates, we believe they serve as good guidelines for estimating the costs for RED plants that depend on water throughput, such as raw water intake, and pretreatment facilities.

We calculated that the throughput of solutions for a 20 megawatt RED plant using 4 N brine and 0.1 N dilute solution with a 30% decrease in the brine concentration will be about 100,000,000 gal/day. With a 30% decrease in brine concentration, about 60% of the potential energy in the brine can be recovered. Therefore, we used the Stone and Webster estimates of cost for a 100,000,000 gal/day plant as the basis for estimating the costs for a 20 megawatt RED plant.

The total investment costs from the Stone and Webster estimate are given in Table X.

TABLE X.
COSTS FOR ELECTRODIALYSIS PLANT TO TREAT WATER
AT INDIAN POINT, NEW YORK

Portion of Plant	Investment Cost
Raw water intake section	\$ 2,787,800
Activated-carbon filtration section	18,578,900
Electrodialysis section	26,328,900
Total Cost	\$ 47,695,600

This cost estimate was made in 1970. If the plant investment costs are updated by the ratio of the plant cost indices given in Chemical Engineering magazine for 1970 and 1978, the following updated costs are obtained for the individual sections of the plant.

$$\text{Cost in 1978} = \text{Cost in 1970} \times \frac{\text{CE index for 1978}}{\text{CE index for 1970}}$$

The plant cost index in 1970 is 125.7; that in April, 1978 is 215.7. The ratio of the two is 1.716. The updated costs are:

Raw water intake section	=	2,787,900 x 1.716	=	\$ 4,784,000
Activated-carbon filtration	=	18,578,900 x 1.716	=	31,881,000
Electrodialysis section	=	26,328,900 x 1.716	=	45,180,000
Total				\$ 81,848,000

The way in which we used some of these costs to estimate the costs for a 20 Megawatt RED plant and the costs for energy is illustrated below.

The estimated lifetimes of the raw water intake, and activated-carbon filtration sections of the plant may be taken to be 20 years. With 8% interest and a sinking fund method of amortization, the annual costs would be:

$$[\text{Total cost}] \times [\text{Sinking fund factor for 20 yr at 8\%}]$$

The factor for 20 yr at 8% is 0.10185

Then, annual costs are:

Raw water intake	= (\$ 4,784,000) (0.10185) = \$ 487,250
Activated-carbon section	= (31,881,000) (0.10185) = <u>3,247,710</u>
Sub-total	\$3,734,960

We have calculated that the cost of the membranes used by Stone and Webster in their estimates was \$1.80/sq ft and the cost of the electrodialysis modules amounted to \$13.90/sq ft of membrane used. With these numbers the cost of power from RED units is exorbitantly high. However, Emren and Bergstrom¹⁶ estimate costs for RED modules at \$0.86/sq ft in 1977 dollars with the manufacturing method they are developing. We believe this figure may be optimistic. Therefore, for our cost estimates we assumed values of \$1.00/sq ft and \$1.50/sq ft for the installed cost of RED modules.

Table VIII shows that with membranes that have resistances of 0.5 ohm-cm² the net power output from RED cells is 2.86 mW/cm², when the cells are operating at 338°K (65°C). This is equivalent to 2.66 W/ft². Thus, a 20,000,000 W plant would require

$$20,000,000 \div 2.86 = 6,993,000 \text{ sq ft of membrane pairs, or}$$

$$2 \times 6,993,000 = 13,986,000 \text{ sq ft of membranes}$$

$$13,986,000 \text{ sq ft} \times \$1.00/\text{sq ft} = \$13,986,000 \text{ in 1977.}$$

To update these 1977 cost to 1978, we used the ratio of the Chemical Engineering plant cost indices.

$\$13,986,000 \times \frac{215.7}{204.1} = \$14,782,000$ in 1978 for the RED membrane cells. However, the membrane cells are not the only costs in the electrodialysis section of the plant. There are several costs in that section that depend on the throughput of water. Such costs were taken from the 1970 Stone and Webster report on costs of large-scale electrodialysis and updated to 1978. A list of the updated costs of other elements in the electrodialysis section follows.

Filters	\$ 240,200
Feed Pumps	1,812,000
RED Pipeing and Heat exchangers	2,492,000
Instruments, controls	784,200
Building, services	<u>223,000</u>
Sub-total	\$ 5,551,400
RED modules	<u>14,782,000</u>
Cost of RED section	\$20,333,400

Klei and Maslin⁵ estimate the cost for the wells for a 20 megawatt hot brine geothermal energy facility to be about \$65/Kw in 1976 dollars. Updating this figure to 1978 by the ratio of plant cost indices:

$$\$65 \times \frac{215.7}{192.1} = \$73/\text{Kw}$$

For a 20,000 Kw plant, the wells might cost:

$$20,000 \times \$73 = \$1,460,000$$

For an RED plant using brine from a salt dome, two sets of wells would be needed; one for pumping fresh brine and one for re-injecting spent brine. Thus, the cost of wells might be

$$\$1,460,000 \times 2 = \$2,920,000$$

The annual costs for amortization would be

RED section* at 10 yr, 8%	= (\$20,333,400) (0.14902)	= \$ 3,030,100
Raw water intake at 20 yr, 8%	= (\$ 4,784,000) (0.10185)	= 487,300
Activated-carbon section at 20 yr, 8%	= (\$31,881,000) (0.10185)	= 3,247,000
Brine and re-injection wells at 20 yr, 8%	= (\$ 2,920,000) (0.10185)	= 297,400
Total annual costs		\$ 7,061,800

The cost of amortization per Kwh at an 80% load factor is

$$\frac{\$ 7,061,800}{20,000 \text{ Kw} \times 24 \text{ hr/day} \times 365 \text{ day/yr} \times 0.8} = \$0.050/\text{Kwh}$$

Forgacs^{14,15} estimates the cost of amortization of plant investment to be 80% to 90% of the total cost of energy from RED installations. On this basis, the cost per Kwh would be

$$\$0.050 \div 0.8 = \$0.0625/\text{Kwh}$$

The total investment cost would be

RED section	\$ 20,334,000
Activated-carbon section	31,881,000
Raw water intake section	4,784,000
Brine-wells section	2,920,000
Total	\$ 59,919,000

$$\$59,919,000 \div 20,000 \text{ Kw} = \$2995/\text{Kw}$$

* The lifetime of the electrodialysis section has been estimated by Forgacs to be 10 years, which seems reasonable. The sinking fund factor for 10 years at 8% is 0.14902.

The above estimated cost is based on the assumption that the brine from a salt dome would not need pretreatment, but that the surface water would need fairly extensive pretreatment by activated-carbon filtration to remove organic contaminants.

If fresh water that needed little pretreatment were used to make up the dilute solution, the following, more optimistic, costs might apply.

The cost of pretreatment facilities might be about 1/3 of the costs listed above, or about \$10,000,000 instead of \$31,881,000. If so, the annual costs for the pretreatment costs would be \$1,018,000 instead of \$3,247,000, and the total annual costs for amortization would be \$4,833,800. The cost of amortization per Kwh would be

$$\frac{\$4,833,800}{\$7,061,800} \times \$0.050 = \$0.034/\text{Kwh}$$

and the total cost per Kwh might be

$$\$0.034 \div 0.8 = \$0.043/\text{Kwh}$$

Estimates of the cost of energy (\$/Kwh) and of the installed cost for power (\$/Kw) based on the concentrations of electrolytes in brines from salt domes, the Great Salt Lake, a natural-gas well, and sea water are summarized in Table XI.

These costs were based on assumptions of \$1.00/sq ft and of \$1.50/sq ft for the installed costs of RED modules. The costs of pretreating the fresh waters to remove materials that would be detrimental to the membranes has been included. For brine from natural-gas wells and for sea water, additional costs for pretreatment of the brines have been included. An estimated cost for brine wells is included for the brine from salt domes. We have calculated that no more than 3% of the total energy will be required to pump the brine from the wells.

The resistances of membranes used for the estimates are 1 ohm-cm² (which represents the best currently available membranes), 0.5 ohm-cm² (which represents membranes available in Israel on an experimental basis), and 0.3 ohm-cm² (which appears to be a reasonable goal for future research).

Table XI shows that if brine from salt domes is used, the estimated cost of energy varies from \$0.060/Kwh to \$0.083/Kwh depending on the membrane resistances and the cost of the membrane modules. As shown previously, if the surface water needs little pretreatment, the cost might be as low as \$0.043/Kwh. The installed costs of the RED power plants vary from \$2880/Kw to \$3735/Kw, depending on membrane resistances and the cost of the membrane modules. More than half of the estimated total installed cost is for pretreatment facilities.

TABLE XI.
SUMMARY OF ESTIMATED COSTS FOR ENERGY AND
POWER WITH BRINES FROM FOUR SOURCES

Brine Source	Assumed cost of RED modules, \$/sq ft of membrane	Estimated cost of energy (\$/Kwh) and (power, \$/Kw) with membrane resistances of:		
		0.3 ohm-cm ²	0.5 ohm-cm ²	1.0 ohm-cm ²
Salt domes	1.00	0.060 (\$2889/Kw)	0.063 (\$2995/Kw)	0.069 (\$3060/Kw)
	1.50	0.065 (\$3195/Kw)	0.076 (\$3220/Kw)	0.083 (\$3735/Kw)
Great Salt Lake	1.00	0.058 (\$2756/Kw)	0.060 (\$2840/Kw)	0.065 (\$3020/Kw)
	1.50	0.065 (\$3080/Kw)	0.070 (\$3200/Kw)	0.077 (\$3470/Kw)
Natural-gas Well	1.00	0.092 (\$4560/Kw)	0.094 (\$4640/Kw)	0.100 (\$4820/Kw)
	1.50	0.104 (\$4980/Kw)	0.107 (\$5100/Kw)	0.114 (\$5380/Kw)
Sea Water	1.00	0.120 (\$5610/Kw)	0.123 (\$5700/Kw)	0.128 (\$5890/Kw)
	1.50	0.139 (\$6270/Kw)	0.160 (\$7110/Kw)	0.168 (\$7390/Kw)

If brine from the Great Salt Lake is used, the estimated costs are slightly lower, \$0.058/Kwh to \$0.077/Kwh, mainly because no brine wells are needed. The estimated installed costs of the facilities are also slightly lower - \$2756/Kw to \$3470/Kw.

When brine from a natural-gas well is used, extensive pretreatment facilities are needed to treat both the fresh water and the brine. This is reflected in Table XI as higher costs for energy (\$0.092/Kwh to \$0.114/Kwh) and higher installed costs (\$4560/Kw to \$5380/Kw).

With sea water, both the sea water and the fresh water must receive extensive pretreatment. In addition, the lower concentration of sea water compared to the other brines makes it necessary to use much more area of RED membrane modules. Both of these factors show up in Table XI as increased estimated costs for energy (\$0.120/Kwh to \$0.168/Kwh) and for the installed costs of facilities (\$5610/Kw to \$7390/Kw).

III. CONCLUSIONS AND RECOMMENDATIONS

The experiments of Weinstein and Leitz¹¹ and of Forgacs^{14,15} have shown that reverse electrodialysis (RED) is technically feasible. Forgacs also showed that equations identical in form to those used here did predict experimental performance in RED units. Our studies have shown that for RED to be economically attractive, steps should be taken to minimize the internal resistance of the RED cells and to maximize the net output voltage from the cells. The main steps that can be taken include:

- The use of large ratios of the concentration of brine to the concentration of the dilute solution.
- With these large ratios, the selection of a concentration of dilute solution that will maximize the gross voltage output (i.e. total voltage component minus internal iR drop).
- The use of materials to maintain the minimum practical thickness for the dilute-solution compartments combined with the proper separation between membranes that will give the minimum thicknesses of diffusive boundary layers at minimum cost for pumping.
- The use of membranes with low electrical resistances and high selectivities combined with long service lifetimes, acceptable strengths and dimensional stabilities.

- The development of methods of manufacturing low-resistance membranes that can be used to make millions of square feet of membranes annually at low-cost.
- The development of methods of manufacturing RED modules that will be much cheaper than the present estimated cost of \$13.90/sq ft, which is based on costs for electrodialysis units.

We recommend that the major thrust of future research efforts should include:

- Experimental studies to determine whether solutions that contain mixtures of ions will behave differently in RED cells than the behaviors predicted by the expressions used in the present study, which were based on the properties of NaCl solutions.
- Studies to maximize the voltage outputs and minimize internal resistances when RED cells are used with solutions of mixed ions, if solutions that contain mixtures of ions do exhibit differences in behavior in RED cells.
- Experimental studies to develop membranes with low resistances, high selectivities, and long service lifetimes.
- Development of manufacturing methods for mass producing RED modules at much lower cost than the hand-assembled electrodialysis units now being made.
- Surveys to determine the geographic location of sources of brine that exist in proximity to sources of water of lower salinity, and the compositions and amounts of each available. The sources of brine might include industrial wastes, such as the waste brines from magnesium manufacture or ethylene glycol manufacture, as well as naturally occurring brines.
- Studies of the environmental impacts of the use of RED power plants and of ways of eliminating any detrimental effects on the environment.
- Studies of the costs of producing energy by the RED method to reflect developments stemming from other areas of research.

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APPENDIX

Derivation and Use of Equations Describing Performance of Reverse Electrodialysis

Figure A-1 shows schematically the concentration profiles in a cell pair (*i.e.*, the repeating unit) of reverse electrodialysis. At steady-state concentration, gradients will exist in the nearly static solution at the membrane-solution interfaces, as indicated. The Nernst idealization of completely mixed bulk solutions and completely static solutions in the diffusive boundary layers has been used for development of the equations to describe the output voltage to be expected from a cell pair, since that idealization has been used in almost all of the research in which the thicknesses of boundary layers were determined.

Between points 1 and 2 (and between points 3 and 4, 5 and 6, and 7 and 8) there will exist a diffusion potential because of the difference in solution concentrations. In each of these zones there will also be a voltage drop that is the product of the current and resistance of the solution within the zone (an IR drop). Across each membrane (points 4 to 5 and 8 to 1) there will be a membrane potential and an IR drop. These diffusion potentials, membrane potentials, and IR drops are expressed in terms of solution concentrations and the properties of the solutions and membranes.

The expressions derived here are for a system containing a uni-univalent electrolyte. The nonthermodynamic assumption is made that single-ion activities can be replaced by concentrations and mean activity coefficients.

The algebraic sum of the individual voltages given in Equations 1 through 8 is the total voltage obtainable from a cell pair, E_o .

The first terms on the right sides of Equations 1, 3, 4, 5, 7, and 8 represent the voltage components of E_o . The second terms on the right sides of those equations and the right sides of Equations 2 and 6 represent the voltage drops due to the resistances of the individual segments of the cell pair (*i.e.*, the resistive components of E_o).

C – CATION-EXCHANGE MEMBRANE
 A – ANION-EXCHANGE MEMBRANE
 δ – REPRESENTS IDEALIZED THICKNESS OF BOUNDARY LAYERS
 N – REPRESENTS CONCENTRATIONS OF SOLUTIONS

SUBSCRIPTS 1, 4, 5, AND 8 DESIGNATE THE INTERFACIAL POINTS SHOWN.
 SUBSCRIPTS d AND c REPRESENT THE DILUTE AND CONCENTRATED SOLUTIONS RESPECTIVELY.

Figure A-1. Representation of a reverse electro dialysis cell pair and the concentration gradients at membrane surfaces.

The net voltages between the individual points in Figure A-1 are given below.

$$-E_{1-2} = \frac{RT}{F} (t_d^+ - t_d^-) \ln \frac{\gamma_1 N_1}{\gamma_d N_d} + \frac{i\delta}{\frac{N_1 + N_d}{2} \Lambda_d} \quad (A-1)$$

$$-E_{2-3} = \frac{i(\Delta - 2\delta)}{N_d \Lambda_d} \quad (A-2)$$

$$-E_{3-4} = \frac{RT}{F} (t_d^+ - t_d^-) \ln \frac{\gamma_d N_d}{\gamma_4 N_4} + \frac{i\delta}{\frac{N_d + N_4}{2} \Lambda_d} \quad (A-3)$$

$$-E_{4-5} = \frac{RT}{F} (t_{AM}^+ - t_{AM}^-) \ln \frac{\gamma_4 N_4}{\gamma_5 N_5} + iR_{AM} \quad (A-4)$$

$$-E_{5-6} = \frac{RT}{F} (t_c^+ - t_c^-) \ln \frac{\gamma_5 N_5}{\gamma_c N_c} + \frac{i\delta}{\frac{N_5 + N_c}{2} \Lambda_c} \quad (A-5)$$

$$-E_{6-7} = \frac{i(\Delta - 2\delta)}{N_c \Lambda_c} \quad (A-6)$$

$$-E_{7-8} = \frac{RT}{F} (t_c^+ - t_c^-) \ln \frac{\gamma_c N_c}{\gamma_8 N_8} + \frac{i\delta}{\frac{N_c + N_8}{2} \Lambda_c} \quad (A-7)$$

$$-E_{8-1} = \frac{RT}{F} (t_{CM}^+ - t_{CM}^-) \ln \frac{\gamma_8 N_8}{\gamma_1 C_1} + iR_{CM} \quad (A-8)$$

where:

E_{i-j} = total voltage between points i and j , where i and j refer to points 1-8 in Figure A-1

R = gas constant, 8.314 joules/°K

T = temperature, °K

F = Faraday's constant, 96,500 coul/eq

t_d^+, t_d^- = transference numbers of the cation and anion in the dilute solution, respectively

t_c^+, t_c^- = transference numbers of the cation and anion in the concentrated solution, respectively

t_{AM}^+, t_{AM}^- = transference numbers of the cation and anion in the anion-exchange membrane

t_{CM}^+, t_{CM}^- = transference numbers of the cation and anion in the cation-exchange membrane

N_1, N_4, N_5, N_8 = concentrations at the membrane-solution interfaces at points 1, 4, 5, and 8 (calculated by Equations A-11 through A-14), eq/cm³

N_d = concentration of the bulk of the dilute solution, eq/cm³

N_c = concentration of the bulk of the concentrated solution, eq/cm³

$\gamma_1, \gamma_d, \gamma_4, \gamma_5, \gamma_c, \gamma_8$ = activity coefficients of the solutions designated by the subscripts.

Λ_d, Λ_c = equivalent conductances of the dilute and concentrated solutions, respectively, $\frac{\text{cm}^2}{(\text{ohm})(\text{eq})}$

Δ = spacing between membranes, cm

δ = thickness of boundary layers, cm

R_{AM}, R_{CM} = areal resistances of the anion-exchange and cation-exchange membranes, respectively, ohm-cm²

i = current density, mA/cm²

Equation A-9 is obtained by summing Equations A-1 through A-8.

$$\begin{aligned}
 E_O = & \left[-\frac{RT}{F} \left\{ (t_d^+ - t_d^-) \ln \frac{\gamma_1 N_1}{\gamma_d N_d} + (t_d^+ - t_d^-) \ln \frac{\gamma_d N_d}{\gamma_4 N_4} + (t_{AM}^+ - t_{AM}^-) \ln \frac{\gamma_4 N_4}{\gamma_5 N_5} \right. \right. \\
 & \left. \left. + (t_c^+ - t_c^-) \ln \frac{\gamma_5 N_5}{\gamma_c N_c} + (t_c^+ - t_c^-) \ln \frac{\gamma_c N_c}{\gamma_8 N_8} + (t_{CM}^+ - t_{CM}^-) \ln \frac{\gamma_8 N_8}{\gamma_1 N_1} \right\} \right] \\
 & - \left[i \left\{ \frac{\delta}{\frac{N_1 + N_d}{2} \Lambda_d} + \frac{\Delta - 2\delta}{N_d \Lambda_d} + \frac{\delta}{\frac{N_d + N_4}{2} \Lambda_d} + R_{AM} \right. \right. \\
 & \left. \left. + \frac{\delta}{\frac{N_5 + N_c}{2} \Lambda_c} + \frac{\Delta - 2\delta}{N_c \Lambda_c} + \frac{\delta}{\frac{N_c + N_8}{2} \Lambda_c} + R_{CM} \right\} \right] \quad (A-9)
 \end{aligned}$$

The interfacial concentration, N_1 , can be found from consideration of the electrical and diffusive transport at steady state. At steady state, the rate of cation transport through the cation-exchange membrane, and thus the rate of arrival of cations at point 1, can be written

$$\frac{i}{F} (t_{CM}^+)$$

The electrical transport of cations into the bulk of solution is

$$\frac{i}{F} (t_S^+).$$

Since $t_s^+ < t_{CM}^+$, the rate of electrical transport of cations from point 1 to point 2 is less than the rate of arrival of cations at point 1 through the membrane, cations accumulate until at steady-state a concentration gradient, $(N_1 - N_d)/\delta$, is developed such that

$$\frac{i}{F} (t_{CM}^+) = \frac{i}{F} (t_d^+) + D \frac{N_1 - N_d}{\delta} \quad (A-10)$$

Rearrangement of Equation 10 gives Equation 11 by which N_1 can be determined.

$$N_1 = N_d + \frac{(t_{CM}^+ - t_d^+)}{DF} \delta i \quad (A-11)$$

If the boundary-layer thickness, δ , is known, N_1 can be determined for any value of i .

Similarly,

$$N_4 = N_d + \frac{(t_{AM}^- - t_d^-)}{DF} \delta i \quad (A-12)$$

$$N_5 = N_c - \frac{(t_{AM}^- - t_c^-)}{DF} \delta i \quad (A-13)$$

$$N_8 = N_c - \frac{(t_{CM}^+ - t_c^+)}{DF} \delta i \quad (A-14)$$

where: D = diffusion coefficient of the electrolyte in the solutions, cm^2/sec . The other terms have been defined previously.

The gross power output from a cell pair is

$$P_g = E_o i \quad (\text{A-15})$$

where: P_g = the gross power output per unit area of cell pair, w/cm^2

However, to achieve a given thickness of boundary layers, δ , each solution must be pumped through each cell pair at some velocity that is characteristic of the cell geometry and the characteristics of the spacer material used. The power required to pump the solutions, P_p , must be furnished by the reverse electrodialysis device. Therefore, the net power output, P_n , is

$$P_n = P_g - P_p \quad (\text{A-16})$$

The pumping power required to achieve given thicknesses of the boundary layers is obtained from graphs such as those shown in Figures A-2, A-3, and A-4.

Some of the data we used on membrane resistances and the transference of counterions through membranes were measured in our laboratories and are presented in Figures A-5 and A-6.

Figure A-2. The effect of solution velocity on the thickness of boundary layers in the Aqua-Chem 15/5 Forced-Flow Electro-desalination Stack.

Figure A-3. The effect of solution velocity on the pressure drop in the Aqua-Chem 15/5 Forced-Flow Electrodessalination Stack.

Figure A-4. The pumping power required to obtain various thicknesses of boundary layers in the Aqua-Chem 15/5 Forced-Flow Electrodesalination Stack.

Figure A-5. Areal resistance of anion- and cation-exchange membranes (AV-3T and CL-2.5T) made by Tokuyama Soda Co., Ltd.

Figure A-6. Transference number of counterions in anion- and cation-exchange membranes (AV-3T and CL-2.5T) made by Tokuyama Soda Co., Ltd.

APPENDIX II

ESTIMATE DETAIL FOR A 100 MGD PLANT USING ACTIVATED CARBON FILTRATION LOCATED AT INDIAN POINT, NEW YORK NEW YORK STATE DESALINATION OPTIMIZATION STUDY

RAW WATER INTAKE PORTION OF PLANT

NOTE: The raw water intake portion of the plant includes the screen well and any services normally provided in the screen well, ultrachlorination of the raw water and raw water delivery piping to the pretreatment intake distribution flume.

CONSTRUCTION COSTS

	<u>Material Costs</u>	<u>Labor Costs</u>	<u>Total Costs</u>
<u>Direct Costs</u>			
Site Development	\$ 12,200	\$ 31,000	\$ 43,300
Structures and Foundations	248,800	408,000	656,800
Trash Racks, Screens and Trash Rake	120,700	21,000	141,700
Raw Water Pumps and Motors	498,000	14,400	512,400
Ultrachlorination Equipment	18,000	2,400	20,400
Raw Water Piping	95,600	262,300	357,900
Misc. Piping and Instrumentation	15,000	14,000	29,000
Electrical	143,100	73,100	216,200
Building Service	19,400	7,800	27,200
Screen Wash Pumps and Piping	26,500	13,700	40,200
Total Direct Costs	\$1,197,300	\$ 847,800	\$2,045,100
<u>Distributable Costs</u>			
Office at Works	\$ 3,000	\$ 43,400	\$ 46,400
Construction Equipment, Temporary Construction, Construction Supplies	45,100	28,500	73,600
Federal and State Taxes and Other Distributables	500	72,500	73,000
Total Distributables	\$ 48,600	\$ 144,400	\$ 193,000
TOTAL CONSTRUCTION COSTS	\$1,245,900	\$ 992,200	\$2,238,100

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<u>Indirect Costs</u>	<u>Material Costs</u>	<u>Labor Costs</u>	<u>Total Costs</u>
Engineering and Design, Fee and Miscellaneous Headquarters Office Allowance for Indeterminates	\$ 130,200 <u>29,600</u>	\$ 63,000 <u>49,400</u>	\$ 193,200 <u>79,000</u>
Total Indirect Costs	\$ 159,800	\$ 112,400	\$ 272,200
TOTAL CONSTRUCTION AND INDIRECT COSTS	\$1,405,700	\$1,104,600	\$2,510,300
(Not including an allowance for labor productivity or change in base wage rates)			
Allowance for Labor Productivity		None	None
Allowance for Differences in Base Wage Rates	<u> </u>	<u>None</u>	<u>None</u>
TOTAL CONSTRUCTION AND INDIRECT COSTS INCLUDING LABOR PRODUCTIVITY AND WAGE RATES FOR SITE	\$1,405,700	\$1,104,600	\$2,510,300
OWNER'S COSTS			
Land - 7.3 Acres @ \$25,000 per Acre			\$ 182,500
Interest During Construction			69,300
Start-up Labor and Materials			8,800
Spare Parts			10,900
Miscellaneous Including Legal, Travel and Public Relations			<u>6,000</u>
Total Owner's Costs			\$ 277,500
TOTAL INVESTMENT COSTS			\$2,787,800

APPENDIX II - Page 3

ESTIMATE DETAIL FOR A 100 MGD PLANT USING ACTIVATED CARBON FILTRATION LOCATED AT INDIAN POINT, NEW YORK NEW YORK STATE DESALINATION OPTIMIZATION STUDY

PRETREATMENT PORTION OF PLANT

NOTE: The pretreatment portion of the plant includes that portion of the plant from the raw water intake distribution flume to and including the treated water clear well.

CONSTRUCTION COSTS

	<u>Material Costs</u>	<u>Labor Costs</u>	<u>Total Costs</u>
<u>Direct Costs</u>			
Site Development	\$ 18,300	\$ 46,700	\$ 65,000
Clarification			
Structures and Foundations	1,046,700	1,658,500	2,705,200
Clarifiers	1,900,000	570,000	2,470,000
Chemical Storage and Feed			
Equipment	85,000	1,800	86,800
Misc. Piping and Instruments	192,100	76,400	268,500
Electrical	62,100	141,700	203,800
Activated Carbon Filtration			
Structures and Foundations	1,265,600	1,545,200	3,810,800
Activated Carbon Filters	1,600,000	480,000	2,080,000
Misc. Piping and Instruments	465,100	174,700	639,800
Electrical	2,400	5,300	7,700
Activated Carbon	1,267,200	21,100	1,288,300
Carbon Regeneration Equipment	200,000	15,000	215,000
Sludge Disposal	360,500	52,500	413,000
Building Service	49,700	81,600	131,300
Total Direct Costs	\$8,514,700	\$5,870,500	\$14,385,200

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APPENDIX II - page 4

CONSTRUCTION COSTS (CONT'D)

<u>Distributable Costs</u>	<u>Material Costs</u>	<u>Labor Costs</u>	<u>Total Costs</u>
Office at Works	\$ 14,400	\$ 292,200	\$ 306,600
Construction Equipment, Temporary			
Construction, Construction Supplies	215,000	191,700	406,700
Federal and State Taxes and Other			
Distributables	<u>3,200</u>	<u>502,200</u>	<u>505,400</u>
Total Distributables	<u>\$ 232,600</u>	<u>\$ 986,100</u>	<u>\$ 1,218,700</u>
TOTAL CONSTRUCTION COSTS	\$8,747,300	\$6,856,600	\$15,603,900
<u>Indirect Costs</u>			
Engineering and Design, Fee and			
Miscellaneous Headquarters			
Office	\$ 677,200	\$ 423,700	\$ 1,100,900
Allowance for Indeterminates	<u>228,400</u>	<u>341,600</u>	<u>570,000</u>
Total Indirect Costs	<u>\$ 905,600</u>	<u>\$ 765,300</u>	<u>\$ 1,670,900</u>
TOTAL CONSTRUCTION AND INDIRECT COSTS	\$9,652,900	\$7,621,900	\$17,274,800
(Not including an allowance for labor			
productivity or change in base wage rate)			
Allowance for Labor Productivity		None	None
Allowance for Differences in Base			
Wage Rates	<u></u>	<u>None</u>	<u>None</u>
TOTAL CONSTRUCTION AND INDIRECT COSTS			
INCLUDING LABOR PRODUCTIVITY AND			
WAGE RATES FOR SITE	\$9,652,900	\$7,621,900	\$17,274,800

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APPENDIX II - page 5

OWNER'S COSTS

Land - 27.6 Acres @ \$25,000 per Acre	\$ 690,000
Interest During Construction	476,800
Spare Parts Inventory	65,300
Start-up Labor and Materials	54,400
Miscellaneous Including Legal, Travel and Public Relations	<u>17,600</u>

Total Owner's Costs	\$ 1,304,100
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TOTAL INVESTMENT COSTS	\$18,578,900
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APPENDIX II - page 6

ESTIMATE DETAIL FOR A 100 MGD PLANT
USING ACTIVATED CARBON FILTRATION
LOCATED AT INDIAN POINT, NEW YORK
NEW YORK STATE DESALINATION OPTIMIZATION STUDY

ELECTRODIALYSIS PORTION OF PLANT

NOTE: The electrodialysis portion of the plant includes that portion of the plant from and including the E.D. feed pumps to and including the product water sump.

CONSTRUCTION COSTS

	<u>Material Costs</u>	<u>Labor Costs</u>	<u>Total Costs</u>
<u>Direct Costs</u>			
Site Development	\$ 24,300	\$ 62,200	\$ 86,500
Structures and Foundations	877,700	1,556,200	2,433,900
Electrodialysis Stacks	8,100,000	648,000	8,748,000
Electrodialysis Filters	128,000	12,000	140,000
E. D. Feed Pumps	880,000	176,000	1,056,000
E. D. Stage Pumps	450,000	90,000	540,000
Brine Recirculation Pumps	1,200,000	240,000	1,440,000
Brine Waste Pumps	45,000	9,000	54,000
Chemical Storage and Feed Equipment	332,500	26,900	359,400
Waste Gas Equipment	7,300	600	7,900
E. D. Piping and Valves for Fixed Staging	1,170,300	281,300	1,451,600
Additional E.D. Piping and Valves for Mixed Staging	424,800	59,800	484,600
Head Tank, Pump and Filter	5,600	600	6,200
E. D. Structures	360,000	126,000	486,000
Instruments and Controls	360,000	97,200	457,200
Electrical for Cells	2,350,300	698,100	3,048,400
Electrical Other Than Cells	123,800	36,700	160,500
Building Services	70,700	58,700	129,400
Total Direct Costs	\$ 16,910,300	\$ 4,179,300	\$21,089,600

APPENDIX II - page 7

CONSTRUCTION COSTS (CONT'D)

	<u>Material Costs</u>	<u>Labor Costs</u>	<u>Total Costs</u>
<u>Distributable Costs</u>			
Office at Works	\$ 35,000	\$ 214,200	\$ 249,200
Construction Equipment, Temporary			
Construction, Construction Supplied	520,500	140,500	661,000
Federal and State Taxes and Other			
Distributables	6,200	357,600	363,800
Total Distributables	\$ 561,700	\$ 712,300	\$ 1,274,000
TOTAL CONSTRUCTION COSTS	\$17,472,000	\$ 4,891,600	\$ 22,363,600
<u>Indirect Costs</u>			
Engineering and Design, Fee and			
Miscellaneous Headquarters Office	\$ 1,500,300	\$ 310,600	\$ 1,810,900
Allowance for Indeterminates	456,100	243,700	699,800
Total Indirect Costs	\$ 1,956,400	\$ 554,300	\$ 2,510,700
TOTAL CONSTRUCTION AND INDIRECT COSTS	\$19,428,400	\$ 5,445,900	\$ 24,874,300
(Not including an allowance for labor productivity of change in labor base wage rates)			
Allowance for Labor Productivity		None	None
Allowance for Differences in Base Wage Rates		None	None
TOTAL CONSTRUCTION AND INDIRECT COSTS INCLUDING LABOR PRODUCTIVITY AND WAGE RATES FOR SITE	\$19,428,400	\$ 5,445,900	\$ 24,874,300

APPENDIX II - page 8

OWNER'S COSTS

Land - 20.6 Acres @ \$25,000 per Acre	\$ 515,000
Interest During Construction	686,500
Spare Parts Inventory	108,500
Start-up Labor and Materials	90,400
Miscellaneous Including Legal, Travel and Public Relations	54,200

Total Owner's Costs	\$ 1,454,600
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TOTAL INVESTMENT COSTS	\$26,328,900
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