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**PRELIMINARY DESIGN AND ANALYSIS OF RECOVERY OF LITHIUM FROM
BRINE WITH THE USE OF A SELECTIVE EXTRACTANT**

**Vi-Duong Dang and Meyer Steinberg
Process Technology Division
Brookhaven National Laboratory
Upton, New York 11973**

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Abstract

Lithium requirements for battery and controlled thermonuclear fusion reactor uses in the next few decades may exceed the current availability of the mineral and brine reserves. It is thus prudent to search for new reserves and resources to satisfy these and other lithium applications in the future. It has been reported that the lithium content of smackover oilfield waters ranges in the order of 100-500 mg/l, and thus could represent a substantial reserve. A method is proposed to extract lithium from this source.

Experimental evidence in the literature indicates that a specific chelating agent of the diketone type, dipivaloylmethane, has a specific selectivity toward lithium in the presence of other metal ions in aqueous solutions. Based in part on this unique property of dipivaloylmethane, a conceptual design of a full size plant is performed to extract lithium from the smackover brine. The study includes alternate flow sheet development, design information on the major units of the process, energy requirement and an economic analysis of a 10^6 Kg Li/yr production facility. The economics of three different process concepts depends on the amount of water evaporated from the initial oilfield feed waters to concentrate the brine. Results as a function of production rates are indicated in a general manner.

Introduction

The requirement of lithium for energy applications such as those for a lithium battery and for the blanket region of controlled thermonuclear fusion reactors has been discussed in a previous symposium.¹ Several of these papers present an assessment of the reserves, resources and requirement of lithium by the year 2000.

The requirement of lithium for the blanket region of controlled thermonuclear fusion reactor is estimated to be between 4.3×10^7 and 10^9 Kg beyond the year 2030.^{2,3} Present lithium reserves are about equal to the estimated demand for lithium required for fusion reactors and battery applications. An ultimate resource of lithium could be sea water which contains a total quantity of lithium of about 2.5×10^{14} Kg. The lithium concentration in sea water, however, is very dilute, 170 ppb.³

It has also been reported that the lithium concentration in smacker brine in the southern part of the United States is about 174 mg/l. This concentration is about three orders of magnitude higher than in sea water. Smacker brine are waters which are usually associated with petroleum oil deposits. The total quantity of lithium in the brine is estimated to be 6.3×10^{10} Kg.⁶ Comparing the extraction of lithium from sea water, the technology of extraction of lithium from brine should be easier because of the existing drilling technology and the higher lithium concentration.

Ion exchange methods have been proposed to extract lithium from sea water.³ These processes suffer from the disadvantage that a larger quantity of ion exchange resins must be used in order to extract the dilute concentration of lithium from sea water. A diketone type chelating agent identified as dipivaloylmethane has been found in the literature to be capable of separate lith-

ium from alkali metal ions in aqueous solutions. A specificity for extraction of lithium from other ions is indicated; an application of this chelating agent for studying lithium transfer across red blood cells has been reported.¹⁴ Although much further additional information is needed for process design purposes, a preliminary scoping design can be undertaken for initial evaluation.

Chemistry of Separation

Guter and Hammond⁷ reported experimentally that lithium can be selectively extracted from aqueous solution by dipivaloylmethane. Dipivaloylmethane (2, 2, 6, 6,-tetramethyl-3,5 - heptanedione) is a chelating agent which is a colorless oil with specific gravity of 0.883. It exists almost entirely in the form of the symmetrical enolic modification I in pure liquid and in organic solvent

When combining with lithium, the chelate probably has the following structure II:

The specificity of DPM for lithium ion in the presence of sodium, potassium and other ions is due to the small radius of the lithium ion (0.60Å). Although sodium ions may interfere with lithium ions in the process of extraction,⁷ careful successive extractions shows that lithium can be recovered to the extent of

97%. Without further experimental results, the following process analysis and design will be based on the work of Guter and Hammond.

Conceptual Process Design for Extraction of Lithium from Spackover Brine

The process concept for the extraction of lithium from brine includes a number of unit operations including evaporation, mixing-settling, electrolysis, distillation and drying. The key step is the liquid-liquid extraction of lithium from the brine by means of dipivaloylmethane in ether. Process analysis depends also on the initial concentration of the constituents in the feed brine. Table 1 lists compositions taken from the work of Collins.⁶ Table 2 lists the analyses of lithium concentration in a number of oilfield and geothermal water determined in various locations around the country which were determined in our laboratory. About 50% of the samples located in Texas contain less than about 10 mg/l Li and are not listed in the table. Several process concepts were developed depending on the amount of water evaporated from original oilfield brine.

Referring to process flow diagram in Figure 1, spackover water is first pumped out from the well and separated from the associated oil and gas at the well head in a separation tank. After the initial separation, further separation can be achieved in skimming tanks. The separated brine is passed into an evaporator to precipitate out the sodium chloride and evaporate a large part of the water. The concentrated bittern is then mixed with ether and DPM in a mixer-extraction vessel. Part of the separated sodium chloride is redissolved and the aqueous sodium chloride is electrolyzed to produce hydrogen and chlorine. The latter are then burned to produce hydrogen chloride which is subsequently used to recover dipivaloylmethane from the metal chelate lithium dipivaloylmethane. Bittern and dipivaloylmethane in ether are then mixed in a mixing tank. After

rigorous stirring, the mix is transferred to a settling tank to separate the water layer and the lithium dipivaloylmethane in the ether layer. Part of the ether may remain in the water layer due to incomplete separation and partial miscibility between ether and water. At room temperature the ether dissolves to the extent of 1 - 1.5% in water and water dissolves to the extent of 7.5% in ether. Therefore, the water layer from the settling tank is transferred to a distillation tower where ether and water are separated and recovered. The top layer which is lithium dipivaloylmethane in ether from the settling tank is fed into the lithium extraction tower together with the recycled water and hydrogen chloride from the C Hl burner. Inside the lithium extraction tower, water and lithium dipivaloylmethane in ether is well mixed in order to ensure improved extraction of lithium from the organic phase into the aqueous phase. The extraction process is further facilitated by the presence of hydrogen chloride. The aqueous and organic phases are then separated in the separation tank. The organic phase will be recycled to remix with the bittern for lithium extraction. The aqueous lithium chloride is transferred to the drying tower to evaporate the water and the lithium chloride is then electrolyzed to produce lithium and chlorine.

The major reactions of the process are

Process Design and Economic Analysis

Process design and analysis are based on the flow diagram of Figure 1. The design calculation is based on the production of W gm Li/yr. Similar to a previous method for extraction of lithium from sea water,³ the first step in the separation process is the separation of other metal ions from lithium. There are particular advantages of separating out the sodium ion first: (1) sodium may interfere with lithium for chelating with dipivaloylmethane, (2) sodium chloride separated from the brine can be used in electrolysis to produce hydrogen and chlorine and subsequently for production of hydrogen chloride which can be used as a neutralization agent for recovery of dipivaloylmethane from its chelating salt. The carbonate precipitation method used in extraction of lithium at Silver Peak, Nevada can not be easily applied here because the concentration ratio of magnesium to lithium in smackover brine is high, about 20 compared to about 1 in Silver Peak. The lithium ions may be adsorbed to magnesium hydroxide when the latter is precipitated which would result in a large bittern loss.

The following discussion will describe the calculation of the important operations of the process. The first two items are the oil tank and the skinning tank, the volume of which has to be at least $2000W \text{ cm}^3$ based on the storage capacity of the brine.

(1) Evaporator

In contrast to solar evaporation used before for crystallization of other salts like sodium chloride, the present method employs evaporators to concentrate the solution because the volume processed is much reduced. The composition of various salts in the bittern is given in Table 3.

Brine of 2000W g/hr flows to the evaporators. Assuming 2 parts per million parts of water are left after evaporation in the evaporator, bittern of 0.00286W g/hr will be obtained after evaporation. Selection of this concentration level is based on a compromise of capital investment and thermal energy requirement for evaporation. The evaporators used are triple effect backward feed evaporators. Heat transfer coefficient of the steam tubes of the evaporators is $0.0678 \text{ cal/sec cm}^2\text{C}$. Saturated steam from nuclear power plant at 9.843 Kg/cm^2 and 196°C is used and water is evaporated as superheated steam at 67.78°C and saturation temperature of 63.89°C . Total energy needed to evaporate the water is 308.37W cal/sec . If we use a steam economy (ratio of total water evaporated to total steam supplied) of three, the actual energy required for evaporation is 102.79W cal/sec . Taking the maximum possible heat flux for the evaporator as $8.665 \text{ cal/cm}^2 \text{ sec}$, one can estimate the total area needed for the evaporator to be 35.58 cm^2 .

(2) NaCl Solution Tank

Using the average concentration of sodium and chloride ions given by Collins,⁶ one can obtain the concentration of sodium chloride in snackover brine as 205,659 mg/l. The requirement for the design of the NaCl solution tank is the volume of the tank. In order to obtain that, one has to know the total flow rate of sodium chloride which is 412W g/hr based on 2000W g/hr of brine. After water is evaporated, the rate of sodium chloride crystallized out is almost complete, about 412W g/hr. To saturate this quantity of sodium chloride, the volume of the solution tank should be at least 1073W cc.

(3) NaCl Electrolyzer

There are many kinds of sodium chloride electrolyzers available commercially. The present design will be based on the information with possibly

some scale up or application of more units. One suitable type of sodium chloride electrolyzer is the De Nora vertical cell with dimensions of $1.2 \times 1.1 \times 3$ cubic meter⁸ which has a capacity of chlorine production of 100 tons/day or 165 tons/day sodium chloride required. Based on several calculations, it is found that the number of electrolyzers needed to produce W g/hr of lithium follows the expression $5.34 \times 10^{-3}W$.

(4) HCl Burner

Determination of a burner or furnace usually depends on the heat requirement of the reaction. The exothermal heat of reaction of hydrogen chloride is 22.1 Kcal/g mole. If the production rate of sodium chloride from the NaCl solution tank is 412 W g/hr, hydrogen can be produced in a rate of 7.043 W g/hr from the sodium chloride electrolyzer. Therefore the rate of hydrogen chloride produced in the hydrogen chloride burner is 257.06 W g/hr. Hence, the total heat released from the burner is 43.23W cal/sec which is an important value for determination of the capital investment of the unit.

(5) Mixing Tank, Settling Tank, Li Extraction Tower and Separation Tank

These four units are similar in the present process in the sense that they serve as storage units. Therefore the determination of these units will be based on their volumes. In order to allow for sufficient capacity, a safety factor of 2.5 to 3 is used here. The volume of mixing tank and settling tank is evaluated as $6.6 \times 10^{-7} W$ gallon while the volume of the lithium extraction tower and lithium separation tank is calculated to be $7.9 \times 10^{-7} W$ gallon.

(6) LiCl Drying Tower

Commercial drying towers are available in several dimensions. One type is 182.9 cm diameter by 60.96 cm deep. One can directly apply a number of these dryers to the present process or scale up the unit to larger dimensions

in order to reduce the number of units required. The number of drying towers needed can be readily calculated by knowing the production rate of lithium which then determines the evaporation rate of water in the drying tower and the dimension of the selected dryer.

The rate of water evaporation in the dryer can be determined by first knowing the flow rate of lithium chloride in the bittern to the dryer as 6.145 W g/hr. In this flow stream, the concentration of lithium chloride is 20.36% by weight which is based on 20% by weight of HCl from the separation tank. The dryer has to be designed so that 3.912 gm H_2O must be evaporated per gram LiCl input. Hence, the total rate of water that has to be evaporated in the dryer is 24.04 W g/hr which is a basic quantity for determining the number of drying towers described in the previous paragraph.

(7) LiCl Electrolyzer

Commercial electrolyzer for LiCl has been reported by Hader et al.⁹ The electrolyzer is rectangular with a graphite anode of 38.1 cm diameter and 883.9 cm long supported from above the cell and steel as cathode and can be scaled to use for the present process. For example, the scale up of the conventional cell with dimensions of $121.9 \times 182.9 \times 91.4 \text{ cm}^3$ can serve the present process well. An electrolyzer cell of $8.7 \times 10^7 \text{ cm}^3$ (3.5 times of the above cell dimension) can handle $1.34 \times 10^8 \text{ gm/hr Li}$. For a higher production rate of lithium, one can increase the number of unit cells.

(8) Distillation Tower for Ether

Ethyl ether can be distilled out from water by means of a tray tower. Solubility of ethyl ether in water is 7% by weight or 0.01798 by mole fraction. Assume the ether concentration is 97% by weight at the top product stream and 98% by weight of water in the bottom stream. The average molecular weight of

the feed is 19.01. Once the feed rate into the tower is specified, the top recycle stream and the bottom reboiler stream can be determined. Using a reflux ratio of 3.5, it is possible to determine the number of plates required by a standard method such as Ponchon-Savarit. As an example, a feed rate of 754 g-moles/hr will give a top product stream of 11.18 g-moles/hr and a bottom stream of 742.82 g-moles/hr. The diameter of the tower can be determined by the Sanders-Brown method.¹⁰ For the present case, the diameter is 30.48 cm and the height of the tower 152.4 cm. The operating pressure can be 2.84 atm.

A summary of the dimensions and number of units needed for any production rate of lithium and an example for a lithium production rate of 1.25×10^5 g/hr is presented in Table 4. The dimensions of equipment chosen are either of a commercial size or a reasonably practical size. Since the capacity of the LiCl electrolyzer can produce more than 1.25×10^5 g/hr of lithium, the number of electrolyzers needed is only one. The dimensions and the number of distillation tower units are not listed because these values usually have to be designed for each production rate of lithium.

Energy Requirement and Economic Analysis

The energy balance of the process and for each piece of equipment is also an important factor in the process. Thermal energy is required for the evaporator in the form of steam to evaporate water. Energy required by the evaporator is 1.291 W kwh(t)/hr. Dissolution of sodium chloride requires the heat of solution and the total heat required for the process is 1.639×10^{-2} W kwh(t)/hr. Electrical energy required for electrolysis of sodium chloride is estimated to be 8.48×10^{-3} W kwh(e)/hr taking into account the inefficiency of the cell. HCl burner will release heat due to the recombination reaction of hydrogen and chlorine. Total heat released in the burner is 0.181 W kwh(t)/hr and this heat

is assumed to be recoverable. The energy required for the dryer to evaporate to dryness is estimated to be 1.551×10^{-2} W kwh(t)/hr. Electrical energy supplied to electrolyze LiCl in the electrolyzer is 1.796×10^{-2} W taking into account the inefficiency. The distillation tower requires a small energy input to the bottom reboiler and the overall energy balance shows that only 4.86×10^{-8} kwh(t)/hr is needed. Mechanical energy requirement for pumping and transporting liquid and/or solid in the process is considered to be small and hence no further detailed calculation is performed. A summary of the energy requirement for the process is shown in Table 5. The energy required for production of one gram of lithium is very small as compared to the energy value of 3400 kwh(e)/gm Li which can be produced in the fusion process.

The preliminary economic analysis was performed as follows. Cost data and information were obtained from Peters¹¹ and Popper.¹² If information was not available from these two references, cost estimation was based on raw material and construction cost of the equipment. Estimates of the cost of the LiCl electrolyzer was obtained in this manner. Correlation of the cost of the equipment with the production rate of lithium has been established for several important pieces of equipment. This cost correlation is based on 1968 dollars and a conversion factor was applied to obtain 1976 dollar.

Cost scaling correlations for the HCl burner was determined to be $184.6 W^{0.85}$ (in dollar units), for the LiCl electrolyzer is 0.3 W, and for the evaporators is $32.25 A^{0.8}$ where A is the heat transfer area needed for evaporation. Cost correlations for the other equipment such as drying towers, mixers, aqueous NaCl electrolyzers and drilling devices are not available so that these were determined analytically.

Capital investment for a 10^6 kg/hr lithium production plant from the brine is given in Table 6. The largest capital equipment cost is the evaporator which has to handle a great deal of brine to produce concentrated bittern. Use of solar ponds could decrease the capital investment, however, it is not always a feasible method for evaporation of water from brine as discussed earlier. Initial inventory and recycling cost of ethyl ether and dipivaloylmethane are also taken into account. In addition to the major pieces of equipment, piping and other unlisted equipment, service facilities, site preparations and electrical instrumentation have also been included.

The production cost for the 10^6 kg Li/yr plant is given in Table 7. Yearly charge, maintenance, electrical and steam cost are taken to be similar to those reported previously³ and as indicated in the table. The production cost for this plant is 5.52¢/gm which is in the order of magnitude of lithium cost of 3.3 cents/gm.¹³ Therefore the present process looks to be a possible process.

Since the most capital intensive item in Table 6 is the evaporator, alternative method for lithium production using the present general process is to evaporate less water or to use the raw brine without evaporation for liquid extraction. If 50% of water in the initial brine is evaporated in the evaporation, one can follow the same procedure of calculation as discussed before for the same process in Figure 1. The capital investment and production cost of this method are given in Tables 8 and 9. The other process is to use the brine directly with dipivaloylmethane in ether for liquid-liquid counter current extraction. Subsequently, lithium is extracted from the ether phase to the aqueous phase in another liquid-liquid extractor. The flow diagram of this process is shown in Figure 2. Process calculations have been performed and the results of

economic analysis are given in Tables 10 and 11. Comparing Tables 6-11, one can see that the most economic process is the one with water evaporated to give two parts of brine from an initial million part of smackover oil field water. Further laboratory process bench scale work will have to be performed to determine operational feasibility especially concerning losses of lithium and the specificity of lithium extraction with a chelating agent such as DPM.

Conclusion

Literature data are used as the basis for a preliminary process design and analysis for the extraction of lithium from smackover oil field brine by means of the chelating agent dipivaloylmethane. Assuming the feasibility of the scale up of the process chemistry, process design calculation was performed as a production rate of lithium. A typical example of lithium production rate of 10^6 kg/yr is tabulated. The major pieces of equipments needed are described. Energy requirement for the process is low, 0.426 kwh(e)/gm. Production cost of the most economic process investigated is 5.52¢/gm (1976 dollar) for a production plant of 10^6 kg/yr. This price is compatible with the present cost of lithium of 3.3¢/gm.

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Table 1. Concentration of Some Constituents in Smackover Brines⁽⁶⁾

<u>Constituent</u>	<u>Average Composition mg/l</u>	<u>Number of Smackover Samples</u>
Lithium	174	71
Sodium	66,973	283
Potassium	2,841	82
Calcium	34,534	284
Magnesium	3,465	280
Strontium	1,924	85
Barium	23	73
Boron	134	71
Copper	1.1	64
Iron	41	90
Manganese	30	69
Chloride	171,686	284
Bromide	3,126	74
Iodide	25	73
Sulfate	446	271

Table 2. Lithium Concentration* in Smackover Oilfield
and Geothermal Water from Various Locations

<u>State</u>	<u>Field</u>	<u>County</u>	<u>Lithium Concentration mg/l</u>
Arkansas	Section 1, Township 18 South, Range 21 West	Columbia	310
Texas	Bryans Mill Unit #1, Bryans Mill	Cass	550
Texas	Currie Plant, Currie	Navarro	370
Arkansas	Dow Chemical Co. Magnolia		280
California	East Mesa 6-1 Holtville	Salton Sea	33
Texas	Chitsey #1, Chitsey	Wood	103
Texas	Buchanan #1, Alba	Wood	72
Texas	Fred Gregg #1A Birthright	Hopkins	190
Texas	Seager #1, East Yantis	Wood	36
Texas	Teas Plant, Teas 8 Carter Bloxom	Freestone	25

*Analysis determined at BNL

Table 3. Estimated Concentrations of the Major
Constituents in Bittern

<u>Constituent</u>	<u>Concentration</u> <u>gm constituent</u> <u>gm water</u>
Sodium	0.151
Calcium	5.794×10^{-5}
Magnesium	0.134
Potassium	0.194
Lithium	0.0136
Chloride	0.176
Sulfate	1.388×10^{-4}

Table 4. Extraction of Lithium From Brine
Summary of Dimensions and Numbers of Major Pieces of Equipment

<u>Equipments</u>	<u>Evaporator</u>	<u>NaCl Solution Tank</u>	<u>NaCl Electrolyzer</u>	<u>Mixing Tank & Settling Tank</u>	<u>Li Extraction Tower and Separation Tank</u>	<u>Drying Tower</u>	<u>LiCl Electrolyzer</u>
Dimensions	$6.35 \times 10^5 \text{ cm}^2$ each	$5.12 \times 5.12 \times$ 5.12 m^3	78 DeNora Cells of $1.2 \times 1.1 \times$ 3 m^3	$5.12 \times 5.12 \times$ 5.12 m^3	$5.12 \times 5.12 \times$ 5.12 m^3	182.88 cm (dia) x 60.96 cm (length)	$426.7 \times 640 \times$ 320 cm^3
Number of Units (N)	$N = \frac{35.58W}{6.35 \times 10^5}$	$N = \frac{1073W}{512^3}$	$N = 5.34 \times 10^{-3} W$	$N = \frac{2.5 \times 10^{-9} W}{5.2^3}$	$N = \frac{3 \times 10^{-9} W}{512^3}$	$N = \frac{4W}{\pi d^2 l}$	$N = 1$
Example Tabulation for $W = 1.25 \times 10^5 \text{ g/hr}$	$N = 1$	$N = 1$	8.56 sets of 78 DeNora Cells	$N = 1$	$N = 1$	$N = 1$	$N = 1$

Table 5. Energy Requirement for the Process

Equipment or Unit	Energy Required	
	Thermal Energy kwh(t)/hr	Electrical Energy kwh(e)/hr
Evaporator	1.291 W	
NaCl Solution Tank	1.639×10^{-2} W	
NaCl Electrolyzer		8.48×10^{-3} W
Hcl Burner	-0.181 W	
Dryer	1.551×10^{-2} W	
LiCl Electrolyzer		1.796×10^{-2} W
Distillation Tower	4.86×10^{-8} W	
Total	1.142 W	2.644×10^{-2} W
		0.426 W (e)

Thermal to electrical conversion efficiency = 35%

Table 6. Capital Investment for a Plant for the Extraction of Lithium from Brine

(Brine is Concentrated by Evaporation of Water to 2 Parts Per Million Parts)

Basis: 10^6 Kg Li/yr plant, 300 days operation, 1976 dollar

<u>Capital Investment</u>	<u>$\\$10^6$</u>
(1) Drilling Device	0.60
Ag NaCl electrolyzer	0.60
Evaporator	15.19
Hcl burner	6.90
Mixing, settling & separation tank	1.50
Drying Tank	0.65
LiCl electrolyzer	0.10
<u>Ether & DPM</u>	<u>3.50</u>
Subtotal	29.04
(2) Piping 29% of item (1)	8.42
(3) Service facilities 56% of item (1)	16.26
(4) Unlisted equipment and misc. 10% of above	5.37
(5) Installation of equipments 35% of above	20.68
(6) Site preparations, buildings 20% of above	15.95
(7) <u>Electrical and instrumentation 20% of above</u>	<u>19.14</u>
Subtotal	114.86
(8) <u>Engineering, Const. Overhead and Contingency</u>	<u>114.86</u>
Total	229.73

Table 7. Production Cost for 10⁶ Kg Li/yr Plant

Evaporating to 2 ppm water, 1976 dollars

<u>Items</u>	<u>Cents/gm</u>
(1) Annual Depreciation 13%	2.98
(2) Electrical Energy Cost (for pumping and electrolysis) 1.5¢ kwh	0.04
(3) Steam Cost ¢ 0.249/1000 gm	0.21
(4) Raw Material	0.00
(5) Maintenance, labor, etc. 10%	2.29
(6) <u>Ether & DPM loss</u>	<u>0.01</u>
Total	5.53

Table 8. Capital Investment for a Plant for the Extraction of Lithium from Brine**(Brine is Concentrated by Evaporation of 50% of Initial Water)****Basis: 10^6 Kg Li/yr Plant, 300 days operation**

<u>Capital Investment</u>	<u>$\\$10^6$</u>
(1) Drilling Device	0.60
Aq. NaCl Electrolyzer	0.60
Evaporator	3.35
HCl Burner	6.90
Mixing, settling & separation tank	5.14
Drying tank	1.46
LiCl electrolyzer	8.60
Distillation Toner	8.00
<u>Ether & DPM</u>	<u>6.50</u>
Subtotal	41.15
(2) Piping 29% of item (1)	11.93
(3) Service facilities 56% of item (1)	23.04
(4) Unlisted equipment & misc. 10% of above	7.61
(5) Installation of equipments 35% of above	29.31
(6) Site preparations, buildings 20% of above	22.61
(7) <u>Electrical & Instrumentation 20% of above</u>	<u>27.13</u>
Subtotal	162.78
(8) <u>Engineering, Cost Overhead & Contingency</u>	<u>162.78</u>
Total	325.56

Table 9. Production Cost for 10^6 Kg Li/yr Plant

Evaporation of 50% of initial water, 1976 dollars

<u>Items</u>	<u>Cents/gm</u>
(1) Annual Depreciation, 13%	4.22
(2) Electrical Energy Cost (for pumping and electrolysis 1.5¢ kwh	0.40
(3) Steam Cost ¢ 0.249/1000 gm	0.05
(4) Raw Material	0.00
(5) Maintenance, labor etc. 10%	3.24
(6) <u>Ether and DPM loss</u>	<u>1.38</u>
Total	9.29

Table 10. Capital Investment for a Plant for the Extraction of Lithium from Brine

(No Concentration of Brine)

Basis: 10^6 Kg Li/yr plant, 300 days operation, 1976 dollar

<u>Capital Investment</u>	<u>$\\$10^6$</u>
(1) Drilling Device	0.60
Liquid-Liquid Extraction tower	59.11
Drying Tank	1.46
HCl Electrolyzer	8.60
<u>Ether & DPM</u>	<u>6.91</u>
Subtotal	76.68
(2) Piping 29% of item (1)	22.23
(3) Service facilities 56% of item (1)	42.94
(4) Unlisted equipment & misc. 10% of above	14.19
(5) Unstallation of equipments 35% of above	54.61
(6) Site preparations, buildings 20% of above	42.13
(7) <u>Electrical & Instrumentation 20% of above</u>	<u>50.56</u>
Subtotal	303.34
(8) <u>Engineering, Cost Overhead & Contingency</u>	<u>303.34</u>
Total	606.67

Table 11. Production Cost for 10⁶ Kg Li/yr Plant

No Concentrations of Brine, 1976 dollar

<u>Items</u>	<u>Cents/gm</u>
(1) Annual Depreciation 13%	17.00
(2) Electrical Energy Cost (for pumping and electrolysis 1.5¢/kwh	0.60
(3) Steam Cost ¢ 0.249/1000 gm	0.05
(4) Raw Material	0.00
(5) Maintenance, Labor etc. 10%	6.04
(6) <u>Ether & DPM loss</u>	<u>2.76</u>
Total	26.45

Figure 1.

FLOW DIAGRAM FOR LITHIUM PRODUCTION FROM SMACKOVER OILFIELD WATER

Figure 2. FLOW DIAGRAM FOR LITHIUM PRODUCTION FROM SMACKOVER OIL FIELD WATER USING DIRECT LIQUID-LIQUID EXTRACTION WITH WATER EVAPORATION