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Bromine-polysulfide Redox-flow Battery Design: Cost Analysis

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Abstract

This paper presents the six levels of the standard chemical engineering design procedures as applied to a bromine-polysulfide redox-flow battery. This project is meant to provide skills in costing and analysis that are necessary in the chemical engineering industry. Based on the potential annual profits of the design it is concluded that the capital costs of this design would make it difficult to construct and operate.

1.0 Introduction

The objective of this report is to document a study-level design and economic analysis of an electrochemical energy storage unit, the bromine-polysulfide redox-flow battery (BPSRFB), at a power capacity of 3 MW. The redox flow battery is a recent renewable innovation used by electric companies to store electric energy during periods of high and low demand. Hydroelectric, solar, and wind systems are the traditional methods of energy generation that have high capital costs. However, they are intermittent, and unpredictable systems that require energy storage for effective incorporation into the electrical supply grid. Redox flow batteries (RFBs) are the subjects of wide scale development activities due to their ability to store large amounts of electrical energy relatively cheaply and efficiently. The BPSRFB is thought to have economic advantages over other energy storage battery concepts. The BPSRFB utilizes sodium bromide as the positive electrolyte and sodium polysulfide as the negative electrolyte.[1] In this system, all of the electroactive species are anions, therefore, a cation-exchange membrane is needed to prevent mixing of the anolyte and catholyte streams.[2] Charge is carried via sodium ions through the membrane.

The design objectives of this project are (1) to develop a flowsheet of a grid-size BPSRFB process, (2) to provide estimates of capital and operating costs and (3) compare the estimated economics of the BPSRFB with other comparable battery techniques. The power level of this project is expected to be 3 MW and charge/discharge times of up to 12 hrs. The charge discharge cycle is performed at least once a day with an expected on-stream efficiency of 77.2%. The economic estimates are in 2014 US dollars. Details of important calculations are found in Appendix A-D. This project is supported by the Electric Power Research Institute in Palo Alto California (USA) and the Tennessee Solar Conversion and Storage using Outreach, Research and Education (TN-SCORE) project (NSF EPS 1004083). Advisors for this project are D. S. Aaron and R. M. Counce. Liaison with EPRI is provided by Chris Trublood.

2.0 Synthesis Information for Processes

2.1 Overall Process Design

Figure 1 shows the schematic representation of a generic RFB. For this study, the catholyte is sodium bromide, and the anolyte is sodium polysulfide. Two pumps are needed in order to push the electrolytes to the battery from the storage tanks. Nickel foam and carbon felt were used for the positive and negative electrodes, respectively. For the purposes of this analysis, it was assumed that there were no side reactions and no crossover.[3] The cells were connected in series, and grouped in stacks. There were a total of twenty stacks used for the entire system.

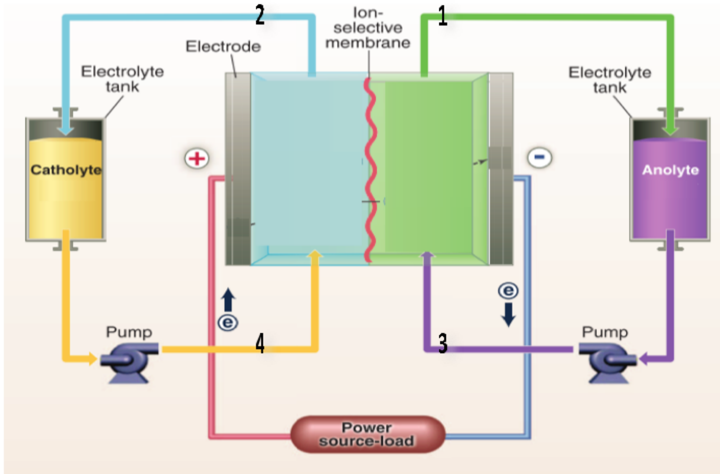

Figure 1: Schematic diagram of experimental set-up for BPSRFB [4]

In addition, the system being used for this analysis uses a bipolar plate with a serpentine flow field as depicted in Figure 2. This is due to the fact that the serpentine flow field claims to have a more uniform distribution of reactant flow than other RFBs and ergo better performance.[5] However, this flow field does come with its own disadvantages. The restricted flow of the electrolytes causes a very high pressure drop over the bipolar plate and increases the cost of the pump. This will be accounted for in later calculations.

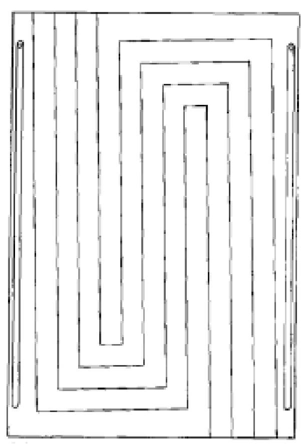

Figure 2: Schematic of serpentine flow field[5]

2.2 *Brief Literature Summary*

In order to perform a cost analysis, a clear understanding of the system is required. In this particular case, knowledge of redox-flow batteries is necessary. By reading Skyllas- Kazacos' research in the article, "Progress in Flow Battery Research and Development," a general understanding of how RFBs work can be obtained. While she does compare a few of the early RFBs that were developed, she quickly delves into the promising vanadium redox-flow batteries (VRBs). However, even with the promise shown by these developments Skyllas-Kazacos stipulates that there are still several obstacles that need to be overcome. One such obstacle is the expense associated with the batteries. While the

membrane expense is present in both the VRB and BPSRFB, the high price of the vanadium electrolyte can be avoided by researching alternative batteries.

Another great source for a comparison of RFBs is Ponce de Leon's journal article "Redox flow cells for energy conversion." In this article a basic understanding of the pros and cons to seven different batteries can be obtained including the BPSRFB. However, before analyzing any particular battery, Ponce de Leon discusses different factors of all RFBs including electrode properties, membrane considerations, and flow distribution. He specifies that the electrode reactions must be reversible and that the costs of the reactants must be reasonable. In addition, he states that a typical membrane in sodium salts should allow Na^+ transport. Lastly, he notes that the flow distribution of the electrolyte should be a constant mean linear flow to prevent stagnant zones.

For this study, the electrode and membrane choices correspond with the Ponce de Leon's specifications. However, in any non-idealized case, the flow field will yield stagnant areas. In Xianguo Li's article "Review of bipolar plates in PEM fuel cells: Flow-field designs" this design choice is explored further. Li stipulates that since the bipolar plate constitutes 30% of the total cost in the fuel cell stack, the choice of flow field can considerably reduce the cost of the design. Throughout the article, the pros and cons of six different flow fields are discussed: pin-type, series-parallel, serpentine, integrated, interdigitated, and flow field designs made from metal sheets. Based on the information, a serpentine flow field was chosen for this study. It is stated that the serpentine flow field produces a uniform distribution of reactant flow and stack compression. However, a disadvantage of this particular design is the high pressure drop created by the bends in the serpentine pattern.

Finally to obtain a full understanding of the schematic of the BPSRFB, Scamman's "Numerical modelling of a bromide–polysulphide redox flow battery: Part 1: Modelling approach and validation for a pilot-scale system" can be utilized. In this article, Scamman models redox flow battery systems for energy storage application design. These models are then used to predict cell performance, species concentration, current distribution, and electrolyte deterioration for the system. Furthermore, Scamman was able to confirm the accuracy of this system of modelling using data obtained from the pilot-scale BPSRFB system that Regenesys Technologies (UK) Ltd. commercialized. Through Scamman's analysis of this pilot-scale BPSRFB, a better understanding of the design of the battery can be acquired.

Lastly, Weber's article, "Redox Flow Batteries: A Review," provides not only an overview of a general RFB and its ability to be used for energy storage but also offers specifics on RFBs such as iron-chromium, bromine-polysulfide, vanadium, and vanadium-bromine. Most importantly, Weber notes that the BPSRFB system is prone to crossover and mixing of the electrolytes. If this were to occur it could lead to precipitation of the sulfur. Therefore, the battery being studied in this paper will use an idealized case in which there is no crossover and no mixing of the electrolytes.

With an understanding of the mechanics of the BPRSFB, Ulrich's book *Chemical Engineering Process Design and Economics: A Practical Guide* provides necessary correlations and constants for the cost analysis of the battery. In addition, Douglas' "A hierarchical decision procedure for process synthesis," ran through the general process of the five levels of cost analysis. With the gathered information, a thorough cost analysis of the bromine polysulfide redox-flow battery can be performed in which the effect of the membrane costs on the overall capital cost of the system is examined.

2.3 Tables of product and raw material costs and specifications

Below are the cost and specifications of the materials used for the BPSRFB.

Materials	Costs	Specifications
Sodium Polysulfide	\$0.34 per kg	1 M, viscosity of 0.018 Pa-s [10]
Sodium Bromine	\$4.90 per kg	5 M, viscosity of 0.003 Pa-s [11]
Carbon Felt	\$20 per m^2	1 m^2 per cell
Nickel Foam[12]	\$20 per m^2	1 m^2 per cell
Current Conductor	\$51 per m^2	1 m^2 per cell
Membrane	Varied	1 m^2 per cell
Electricity (Purchased)	\$0.01/kW-hr	
Electricity (Produced)	\$0.16/kW-hr	

Table 1: Product and raw materials costs and specifications

2.4 Table(s) of relevant properties

Table 2 describes the parameters of the plant. Table 3 describes the various parameters of the cell. These values were used in the calculations of the cost analysis.

Overall Plant Parameters	
Maximum rated output power	3000 kW
Energy storage capacity	18000 kW-hr
Charge Cycle	6 Hours
Discharge Cycle	6 hours
Cycles Per Year	≤ 365

Table 2: Properties related to the overall plant

Cell Parameters	
Membrane	Varied
Cell Voltage	0.8 volts per cell
Electrolytes	NaBr and Na_2S_4

Table 3: Properties related to the individual cell

2.5 Input information for base-case

2.5.1 Aqueous solutions of Sodium Bromide and Sodium Polysulfide at 5 M and 1 M, respectively.

2.5.2 The power capacity is 3MW

2.5.3 Cost of membrane varied to test profitability sensitivity to membrane cost.

2.5.4 All costs are to be in 2014 dollars, unless otherwise specified (ChE Index = 567.7)

2.5.5 Sale cost of electricity is \$0.16/kW-hr

2.5.6 Purchased electrical cost energy is \$0.01/kW-hr

2.5.7 One complete charge/discharge cycle is assumed to take 24 hours

2.5.8 State of charge (SOC) limits are 0.1 and 0.9 [7]

2.5.9 The temperature and pressure of the battery is assumed to be non-disruptive

2.6 Identification of design variables

For this study, the primary design variable being analyzed is the membrane cost. By fluctuating the price of the ion exchange membrane used in the cell, the overall cost estimate was monitored. This allowed for a direct correlation to be made between the membrane cost and the overall capital cost of the system. In addition to this, other variables were chosen and remained constant throughout the study. For example, a maximum output power of 3 MW, or 3000 kW, and a total of twenty stacks were used to complete the calculations needed. For a complete list of design variables, please refer to Tables 1, 2, and 3 as well as to section 4.

3.0 Method of Approach

In order to complete a study-level design and economic analysis of the bromine-polysulfide redox-flow battery (BPSRFB), at a power capacity of 3 MW a walkthrough of the step by step methodology laid out in J. M. Douglas' paper "A Hierarchical Decision Procedure for Process Synthesis" was used. Level 1 simply lists all input information required in the design of the BPSRFB. Level 2 determines the economic potential of the battery by doing an input-output analysis of energy. Level 3 then considers power capacity by calculating the costs of the two pumps and cell components. Level 4 calculates energy capacity which includes the cost of sodium bromide, sodium polysulfide, and storage. Level 5 contains an overall balance of the plant. Finally, level 6 summarizes all capital cost information into a final table.

4.0 Results

The first level of this analysis includes all input information required for the design of the battery and the costs for specific design components. All variables listed below in addition to variables listed in Tables 1 through 3 will be used throughout the rest of the design process unless otherwise indicated.

4.1 Reactions

4.1.1 Stoichiometry

$3Br^- + S_4^{2-} \leftrightarrow Br_3^- + 2S_2^{2-}$

4.1.2 Temperature: Room temperature (25^{o}C)

4.1.3 Stack Pressure: 1 atm

4.1.4 State of charge considerations: Min = 10%, Max = 90%

4.2 Design Details

4.2.1 Design current density of cell: 50 mA/cm^2 (500 A/m^2)

4.2.2 Cell Voltage: Charge Efficiency = 77.2% [13]

4.2.3 Cell Voltage : Discharging Efficiency = 100% [6]

4.2.4 Materials of construction for tanks: fiberglass and carbon steel

4.3 Cost Information

4.3.1 Cell construction materials

- Cost of Assembly: 10% of cell component cost
- Actual Bare Module Factor: 1.4

4.4 Other Information

- 4.4.1 Power Conditioning: $100/kW
- 4.4.2 Transformer Costs: $37/kW
- 4.4.3 Breakers, Contacts, Cabling, etc.: $18/kW
- 4.4.4 Pressure of Stacks: 1 bar

When looking at the input-output analysis for the battery, level 2, the only input and output being considered is energy. Based on the costs and profits of the electricity used and produced by the battery, a maximum economic potential can be calculated using the following equations. See Appendix A for example calculations.

$$Economic\ Potential = cost\ of\ product\ (electricity) - cost\ of\ raw\ materials \tag{1}$$

$$EP_2 = \left(E_{discharging} \times {}^{\$}/_{kW\ h_{discharging}} - E_{charging} \times {}^{\$}/_{kW\ h_{charging}} \right) \times {}^{cycles}/_{year} \tag{2}$$

Using the above equations a graph could then be generated relating the economic potential to the number of cycles per year in which the battery was charged and discharged (Figure 3).

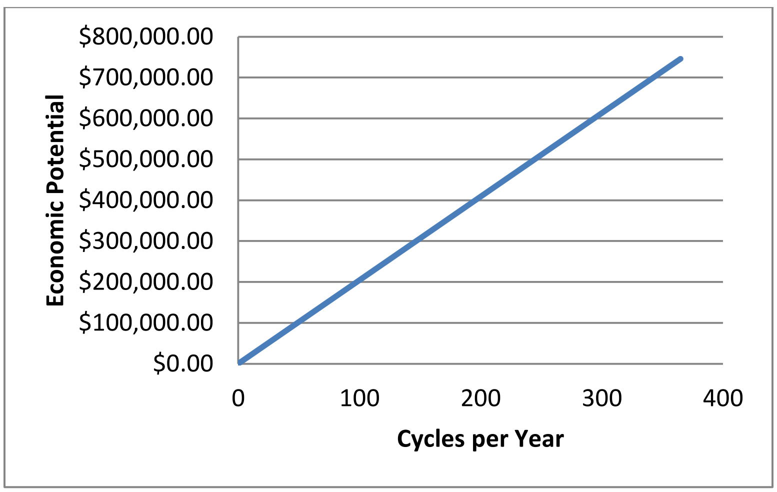

Figure 3: Economic potential for a BPSRFB based on a level 2 design.

In the next level of the design, level 3, all power capacity considerations must be incorporated. This includes the costs of material per cell in addition to the two pumps needed. First, to determine the costs of the cells it is essential to know how many cells are needed to provide an adequate amount of power. To calculate this, the following equations were used assuming that only twenty stacks were needed and each stack carries the same amount of voltage.

$$Total\ Amps = N_s A_s CD \tag{3}$$

Determines the total amps when N_S is the number of stacks, A_s is the area of the stack in m^2, and CD is the current density in mA/cm^2. The voltage per stack (V_S) is then determined by

$$V_s = \frac{Power}{Total\ Amps} \tag{4}$$

where the power is in watts. This can then be used to calculate the number of cells per stack by dividing the voltage per stack by the total voltage per cell, which is equal to 0.8 volts in this model. After finding the number of cells per stack the total number of cells is found by simply multiplying by the number of stacks, assumed to be twenty.

The costs of the raw materials per cell include the cost of the membrane, current collector, carbon felt electrode, and nickel foam electrode. For the initial calculations a membrane cost of $1000 per m^2 will be used. This brings the total cost of materials to $1,155 per cell. The only additional cost outside of raw materials is the cost of assembly which is taken to be 10% of the cell component costs, or $115.50, which brings the total cost per cell to $1,270.50. This cost per cell is then multiplied by the total number of cells and a bare module factor of 1.4 to determine the bare module cost of the cells. An annualized cost can then be determined by multiplying by a factor of 0.24.[8] All other costs calculated throughout the analysis are annualized in the same manner. See Appendix B-1 for example calculations of cell costing.

After determining the total number of cells the cost of the pumps can be defined. In order to cost the pumps, the shaft work, W_S, needed must be calculated using the following equation

$$W_s = \frac{m\Delta P}{\eta} \tag{5}$$

where m is the flow of the solution in m^3/s, ΔP is the change in pressure in Pa, and η is the efficiency of the pump. The flow was determined to be 0.008 m^3/s and 0.02 m^3/s for sodium bromide and sodium polysulfide, respectively. Assuming that the system is at steady state, the flow rate in equals the flow rate out. Thus, referring to Figure 1, streams 2 and 4 are equal to 0.008 m^3/s and streams 1 and 3 are equal to 0.02 m^3/s. To calculate the efficiency the equation below is needed

$$\eta = (1 - 0.12m^{-0.27})(1 - \mu^{0.8}) \tag{6}$$

where μ is the viscosity of the fluid in Pa-s. After determining the shaft work the capital cost is found using Figure 5.49 in Ulrich. For the sodium bromide side of the battery, a cast steel, centrifugal pump was used giving a material factor of 1.4 and a bare module factor of 4.2. Although cast steel is slightly more expensive than cast iron, it is necessary for this design since the corrosion of cast iron by concentrated sodium bromide is a major concern.

On the sodium polysulfide side of the battery, a nickel alloy, centrifugal pump is recommended giving a material factor of 3.5 and a bare module factor of 7. Similarly, nickel alloy was chosen to avoid corrosion of the pump. See Appendix B for example calculations of pump costing. The final calculation for the economic potential for level 3 is done using the following equation:

$$EP_3 = EP_2 - AC_{Cells} - AC_{pumps} \tag{7}$$

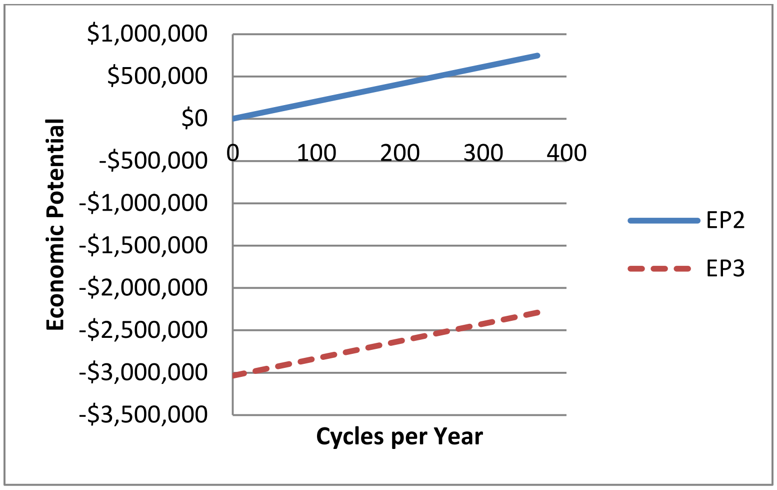

Figure 4: Comparison of economic potential based on a level 2 and level 3 design.

For level 4 of the design, the energy capacity must be determined. The first step to this calculation is finding the mass of the sodium bromide and sodium polysulfide needed for the reaction. Based on the following equation the molar flow of the solution can be determined.

$$Flow\ Rate = \frac{N_{TC}(CD)A_s}{F} \tag{11}$$

In this equation N_{TC} is the number of total cells and F is Faraday's number. Then, using a six hour discharge time and the molecular weight this value can be converted to mass needed in kilograms. This allows costing of the solutions. After finding the mass it can be divided by the density to get the volume needed for the storage tank. However, to avoid having the tank at full capacity all the time a tank with 10% excess volume was chosen. A cone roof, carbon steel tank with a bare module factor of 1.9 was chosen for sodium polysulfide since it will not corrode iron or steel. Similarly, a cone roof, fiberglass tank with a bare module factor of 3 was chosen for sodium bromide due to its corrosive nature. Using this information and Figure 5.61 of Ulrich, the cost of both tanks was found and annualized. The final calculation for the economic potential for level 4 is done using the following equation and displayed in Figure 5.

$$EP_4 = EP_3 - AC_{NaBr} - AC_{NaBr\ Storage} - AC_{Na_2S_4} - AC_{Na_2S_4\ Storage} \tag{12}$$

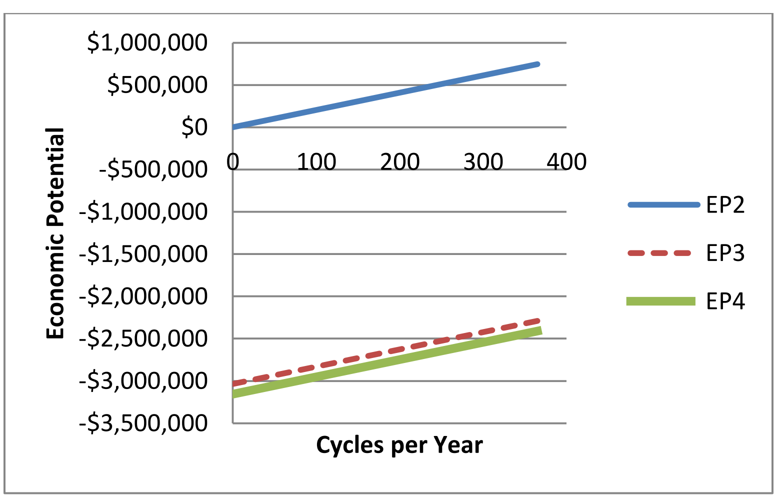

Figure 5: Economic potential for levels 2, 3, and 4.

The balance of plant costs, level 5, includes the costs of construction and site preparation, costs of a control system, and some remaining costs. An average estimate of building and site preparation costs is $900 per square meter (2007).[9] Using an inflation rate of 3% the cost in 2014 would be around $1,107 per square meter. Using an estimate of 500 m^2/MW, the cost of the plant is determined to be around $553,444 per MW. Then, multiplying by the power of this particular model gives a bare module cost of around $1,660,331. The bare module cost of the control system is estimated to be $22,509 and the remaining costs are $56/kW, or $168,000.[9] All costs were annualized before calculating EP_5. See Appendix D for example calculations involved in the level 5 analysis. The final calculation for the economic potential for level 5 is done using the following equation.

$$EP_5 = EP_4 - AC_{BP} - AC_{CS} - AC_{RC} \tag{13}$$

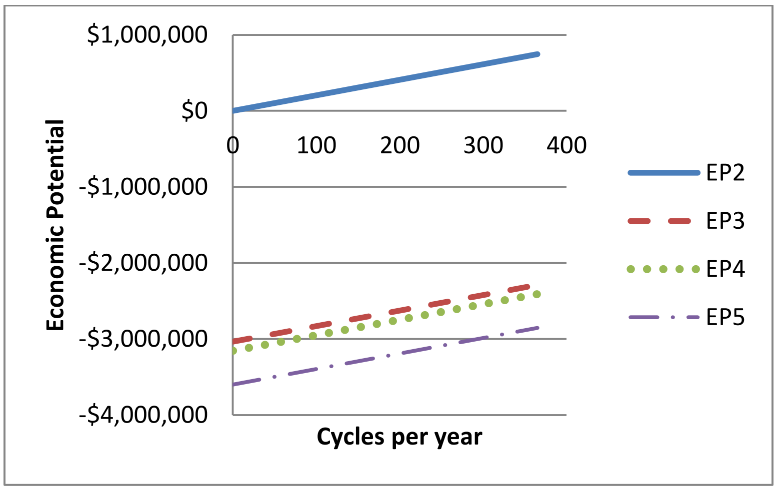

Figure 6: Economic potential for levels 2, 3, 4, and 5.

4.5 Capital Cost Estimates

For level 6, a capital cost summary table is created using the information gathered in levels 1 through 5. A summary of this information is shown in Table 4. Please see Appendix E for the completed capital cost summary table.

Equipment Identification	Total Annualized Cost
Pumps	$14,452
Misc: Solution, Materials, etc.	$3,573,239
Storage Vessels	$13,031
Overall	$3,600,722

Table 4: Overview of Capital Cost Summary

4.6 *Operating Cost Estimates*

In order to determine the operating costs, a capital cost of $3,780,902 was used. Then, the working capital was assumed to be 15% of the fixed capital, which equated to $567,135.27. The summation of those two numbers gave us the total capital investment of $4,348,037.08. Manufacturing expenses were calculated along with operating labor. The operating labor was determined using Table 6.2 in Ulrich. Assuming that each stack is equivalent to a reactor, six operators are required per shift in the plant. After determining the number of operators, utilities, maintenance repairs, operating supplies, laboratory charges, patents and royalties were calculated. Local taxes and insurance were assumed 2% and 1% of the fixed capital respectively. Depreciation was determined to be 10% of the fixed capital. Please see Appendix E for the competed operating cost summary table.

5.0 *Discussion of Results*

The results from the study level design are depicted in Figure 5 and the capital cost estimates shown in Appendix E. In the ideal world, the economic potential should be a positive value at 365 cycles per year or less. For this particular study, only Level 2 yields a positive economic potential. Levels 3, 4, and 5 do not pass into the positive value range at the 365 cycle mark. For the BPSRFB to be economically reasonable at 3 MW energy capacity, some initial capital costs must be reduced or the system must make a higher profit.

For this analysis, the capital cost of the ion exchange membrane will be altered to determine the maximum cost that can be paid per m^2. As shown in Figure 7, even with a membrane cost of $0 per m^2 the economic potential is still about -$70,000. Figure 8 depicts each level of economic potential when no membrane cost is included in the calculation for level 3. In addition, Appendix E shows an alternate capital and operating cost estimate using these values.

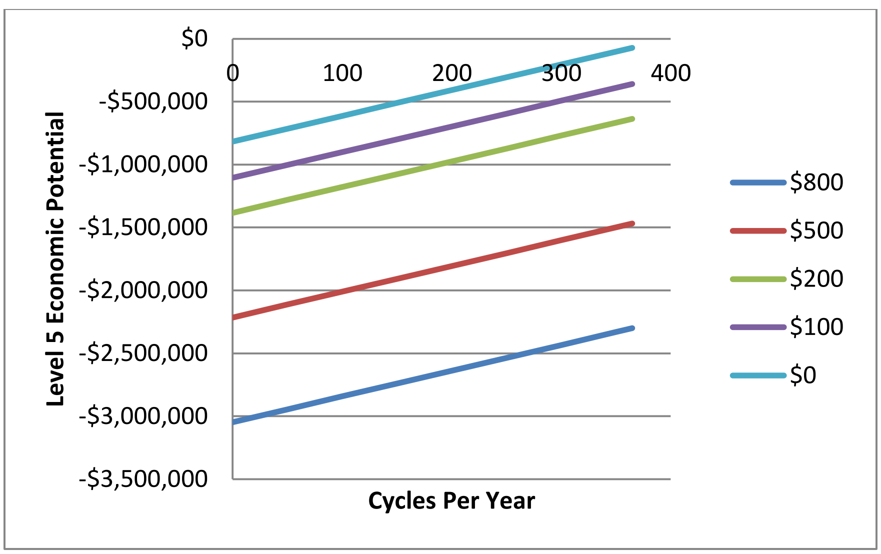

Figure 7: Level 5 costs with varying membrane costs per m^2

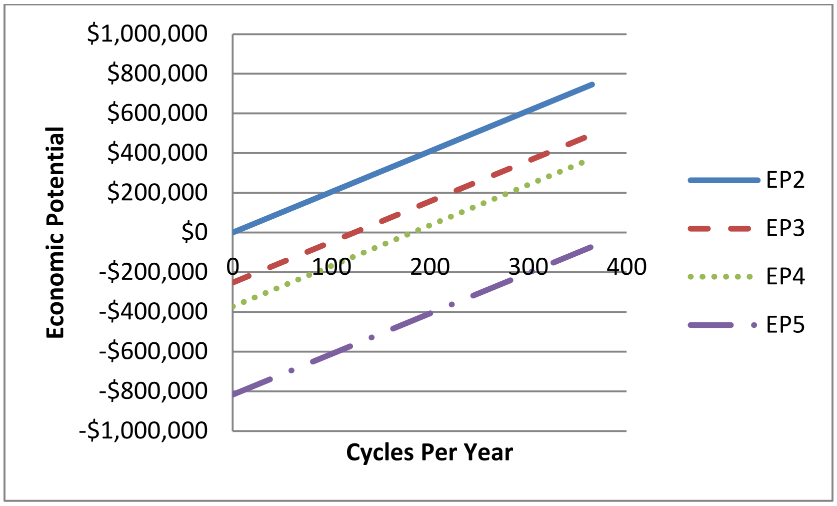

Figure 8: Economic potential for levels 2, 3, 4, and 5 with no membrane cost added.

6.0 Conclusions

Based on the potential annual profits seen in Figure 6 it can be concluded that the capital costs of this design would make it difficult to construct and operate. As seen by the cost summary table (Appendix E) the two largest contributors to the capital cost are the ion exchange membrane and the pumps. As shown in Figure 8 a decreased membrane cost drastically affects the economic potential of this design. However, it is obvious that other modifications need to be made for the design to ensure a positive economic potential. In order for the bromine-polysulfide redox-flow battery to become commercialized in the future the costs must be reduced with innovative improvements to yield a high performance with an optimal budget.

7.0 Recommendations

Since having a lower membrane cost does not produce a satisfactory economic potential, additional modifications must be made to either decrease costs of increase profits. While the membrane cost is a significant factor in the overall economic potential, the costs of the pumps also play a vital role. Even though a cheaper material cannot be used for the pump, due to the specifications of the electrolytes, the cost can be decreased by lowering the pressure drop across the bipolar plate. If the pressure drop is cut in half it lowers the cost by about $2,200. While this amount does not have a large impact when the membrane cost is high, it can have an impact if the membrane cost is lowered. Unfortunately, the only way to decrease this pressure drop over the bipolar plate is by using an alternative design for the flow field. For example, instead of using the serpentine flow pattern described previously, a pin-type flow field could be used as depicted in Figure 9.[5] This design would have a much lower pressure drop than the serpentine flow pattern, but creates stagnant areas where there is little to no electrolyte decreasing the performance of the cell. Therefore, when determining whether to use a different flow pattern the cost saved due to the lower pressure drop needs to be compared to the loss in profits due to the decreased performance.

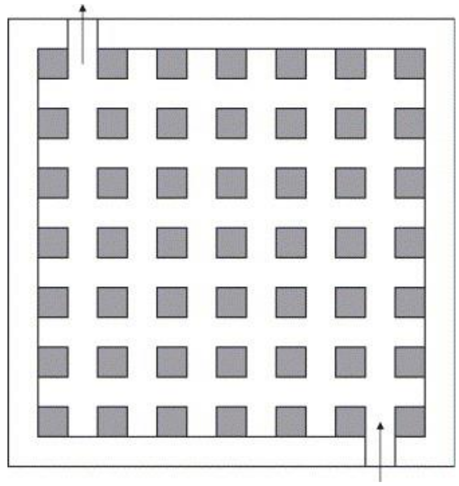

Figure 9: Schematic of a pin-type flow field. [5]

However, if the costs cannot be further decreased, the profits must increase. In order to increase profits, the sale cost of electricity would need to be varied. Table 5 shows the lowest cost at which the output electricity could be sold to still maintain a positive economic potential for multiple scenarios. For example, if nothing was changed about the current design the sale cost of electricity would have to increase to \$0.73/kW·h to earn a profit. However, if the membrane cost was decreased to \$50 per m^2 the cost would only have to be increased to \$0.21/kW·h. As shown, the most cost efficient scenario for the consumer would be if the pressure drop is decreased along with the membrane cost being reduced in which the electricity can be sold for \$0.18/kW·h. However, this once again does not account for the loss of performance that would occur after the flow field has been changed and would need to be studied further.

	Membrane Cost of \$1000 per m^2	**Membrane Cost of \$50 per m^2**
ΔP = 50,000 Pa (Serpentine)	\$0.73/kW·h	\$0.21/kW·h
ΔP = 25,000 Pa (Pin-type)	\$0.72/kW·h	\$0.18/kW·h

Table 5: Necessary sale cost of electricity to obtain a positive economic potential at each of the specified parameters.

8.0 References

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9.0 Appendix

9.1 Appendix A – Level 2 Calculation

The power for this calculation is given to be 3 MW, or 3,000 kW. Then using a six hour discharging time, t_D, and a 100% discharging efficiency, $\mathcal{E}_D$, then

$$E_{discharging} = \frac{(Power) \cdot t_D}{\mathcal{E}_D} = \frac{(3{,}000\ kW) \cdot (6\ hr)}{1} = 18{,}000\ kW \cdot hr$$

Then using a six hour charging time, t_C, and charging efficiency, $\mathcal{E}_C$, of 77.2% the purchased energy is calculated vi

$$E_{charging} = \mathcal{E}_C \cdot P \cdot t_C = (0.772) \cdot (3{,}000\ kW) \cdot (6\ hr) = 13{,}896\ kW \cdot hr$$

Lastly given the price of input and output power of 0.01 $/kW hr and 0.16 $/kW hr respectively, the economic potential calculation for level 2 is completed. Note that the varied value of cycles per year is assumed to be 365 for all calculations.

$$EP_2 = \left(E_{discharging} \cdot {}^{\$}\!/_{kW\ h_{discharging}} - E_{charging} \cdot {}^{\$}\!/_{kW\ h_{charging}} \right) \cdot {}^{cycles}\!/_{year}$$

$$EP_2 = (18{,}000 \cdot 0.16 - 13{,}896 \cdot 0.01) \cdot 365 = \$745{,}826$$

9.2 Appendix B – Level 3 Calculation

9.2.1 Annualized Cost of Cells

To begin the Level 3 calculation the number of stacks, N_s, must be determined. For this calculation it was assumed that there would be 20 stacks and that the area of the area of the stack, A_s, is equal to 1 m^2. The current density, CD, is equal to 100 A/m^2.

$$Total\ Amps = N_s A_s CD = (20)(1\ m^2)(100\ A/m^2) = 10{,}000\ A$$

$$V_s = \frac{Power}{Total\ Amps} = \frac{3{,}000{,}000\ Watts}{10{,}000\ A} = 300\ V$$

where Vs is the voltage per stack. Then given a cell voltage, V_C, of 0.8 V

$$Cells\ Per\ Stack = \frac{V_S}{V_C} = \frac{300\ V}{0.8\ V} = 375$$

$$Total\ Cells = 375 \cdot 20 = 7{,}500$$

Given a cost of $1,199 per cell for materials and assembly

$$Total\ Cost = \ (Total\ Cells) \cdot (Cost\ per\ Cell) = (7{,}500) \cdot (\$1{,}199) = \$\ 8{,}992{,}500$$

$$BMC_{cells} = 1.4 \cdot (Total\ Cost) = \$12{,}589{,}500$$

where BMC is the bare module cost. The annualized cost, AC, is

$$AC_{Cells} = \ 0.24 \cdot BMC_{cells} = (0.24) \cdot (\$12{,}589{,}500) = \$3{,}021{,}480$$

9.2.2 *Annualized Cost of Cells*

Determine the flow rate of sodium bromide

$$m_{NaBr} = \frac{Total\ Cells \cdot A_S \cdot CD}{\text{F}} = \frac{7{,}500 \cdot (1m^2) \cdot (500\ \ A/m^2)}{(96485\ \ C/mol) \cdot (3600\ \ s/hr)} = 139{,}918\ mol/hr$$

Given a concentration of 5 M for sodium bromide

$$m_{NaBr} = \left(139{,}918 \frac{mol}{hr}\right)\left(\frac{1\ hr}{3600\ s}\right)\left(\frac{1\ L}{5\ mol}\right)\left(\frac{m^3}{1000\ L}\right) = \ 0.0078\ \ m^3/s$$

Given a viscosity of 0.003 Pa-s for sodium bromide

$$\eta = (1 - 0.12{m_{NaBr}}^{-0.27})(1 - {\mu_{NaBr}}^{0.8})$$

$$\eta = \ (1 - 0.12 \cdot 0.0078^{-0.27})(1 - 0.003^{0.8}) = 0.55$$

Using an assumed pressure drop, ΔP, of 50,000 Pa

$$W_s = \frac{m_{NaBr}\Delta P}{\eta} = \frac{(0.0078\ \ m^3/s)(50{,}000\ Pa)}{0.55} = 708\ W = 0.708\ kW$$

Use Figure 5.49 in Ulrich to determine cost of the sodium bromide pump. For a centrifugal pump this is found to be $4,936. Then using a bare module factor, F_{BM}, of 4.2

$$BMC_{NaBr\ Pump} = 4.2 \cdot \$4{,}936 = \$20{,}731$$

$$AC_{NaBr\ Pump} = 0.24 \cdot BMC_{NaBr\ Pump} = \$4{,}975$$

Lastly, the annualized cost for sodium polysulfide needs to be determined. The flow rate for sodium polysulfide is found by the stoichiometry of the chemical reaction. Through this it was determined that there should be twice as much sodium bromide. Therefore,

$$m_{Na_2S_4} = \frac{m_{NaBr}}{2} = 69{,}959\ mol/hr$$

Given a concentration of 1 M for sodium polysulfide

$$m_{Na_2S_4} = \left(69{,}959\frac{mol}{hr}\right)\left(\frac{1\ hr}{3600\ s}\right)\left(\frac{1\ L}{1\ mol}\right)\left(\frac{m^3}{1000\ L}\right) = 0.0194\ m^3/s$$

Given a viscosity of 0.018 Pa-s for sodium polysulfide

$$\eta = \left(1 - 0.12 m_{Na_2S_4}{}^{-0.27}\right)\left(1 - \mu_{Na_2S_4}{}^{0.8}\right)$$

$$\eta = \left(1 - 0.12 \cdot 0.0194^{-0.27}\right)\left(1 - 0.018^{0.8}\right) = 0.63$$

Use Figure 5.49 in Ulrich to determine cost of the sodium polysulfide pump. For a centrifugal pump this is found to be \$5,641. Then using a bare module factor of 7

$$BMC_{Na_2S_4\ Pump} = 7 \cdot \$5{,}641 = \$39{,}487$$

$$AC_{Na_2S_4\ Pump} = 0.24 \cdot BMC_{Na_2S_4\ Pump} = \$9{,}477$$

$$AC_{Pumps} = AC_{NaBr\ Pump} + AC_{Na_2S_4\ Pump} = \$4{,}975 + \$9{,}477 = \$14{,}452$$

$$EP_3 = EP_2 - AC_{Cells} - AC_{pumps} = \$745{,}826 - \$3{,}021{,}480 - \$14{,}452 = -\$2{,}290{,}106$$

9.3 *Appendix C – Level 4 Calculation*

9.3.1 *Annualized Cost of Sodium Bromide*

Determine mass of sodium bromide required

$$mass_{NaBr} = (m_{NaBr} \cdot t_D) \cdot MW_{NaBr} = (139{,}918\ mol/hr \cdot 6\ hr) \cdot 0.1\ kg/mol = 86{,}385\text{ kg}$$

Using a cost of \$4.90/kg

$$BMC_{NaBr} = (86{,}385\ kg) \cdot (\$4.90/kg) = \$423{,}289$$

$$AC_{NaBr} = 0.24 \cdot BMC_{NaBr} = 0.24 \cdot (\$423{,}289) = \$101{,}589$$

9.3.2 *Annualized Cost of Sodium Bromide Storage*

Given a density of 3200 kg/m^3 for sodium bromide

$$V_{NaBr} = \frac{mass_{NaBr}}{\rho_{NaBr}} = \frac{86{,}385\ kg}{3200\ \ kg/m^3} = 27\ m^3$$

Assuming 10% excess volume is needed for the storage tank

$$V_{NaBr\ Storage} = 1.1 \cdot V_{NaBr} = 1.1 \cdot 27\ m^3 = 30\ m^3$$

Use Figure 5.61 in Ulrich to determine cost of a storage tank with this volume. For a cone roof storage tank this is found to be \$9,167. Then using a bare module factor of 3 for fiberglass

$$BMC_{NaBr\ Storage} = 3 \cdot \$9{,}167 = \$27{,}500$$

$$AC_{NaBr\ Storage} = 0.24 \cdot BMC_{NaBr\ Storage} = 0.24\ \cdot \$27{,}500 = \$6{,}600$$

9.3.3 Annualized Cost of Sodium Polysulfide

Determine mass of sodium polysulfide required

$$mass_{Na_2S_4} = \left(m_{Na_2S_4} \cdot t_D\right) \cdot MW_{Na_2S_4}$$
$$mass_{Na_2S_4} = \ (69{,}959\ \ mol/hr \cdot 6\ hr) \cdot 0.174\ \ kg/mol = \ 73{,}138\ kg$$

Using a cost of \$0.34/kg

$$BMC_{Na_2S_4} = (73{,}138\ kg) \cdot (\$0.34/kg) = \$24{,}867$$

$$AC_{Na_2S_4} = 0.24 \cdot BMC_{Na_2S_4} = \ 0.24 \cdot (\$24{,}867) = \$5{,}968$$

9.3.4 Annualized Cost of Sodium Polysulfide Storage

Given a density of 968 kg/m^3 for sodium polysulfide

$$V_{Na_2S_4} = \frac{mass_{Na_2S_4}}{\rho_{Na_2S_4}} = \frac{73{,}138\ kg}{968\ \ kg/m^3} = 76\ m^3$$

Assuming 10% excess volume is needed for the storage tank

$$V_{Na_2S_4\ Storage} = 1.1 \cdot V_{Na_2S_4} = 1.1 \cdot 76\ m^3 = 83\ m^3$$

Use Figure 5.61 in Ulrich to determine cost of a storage tank with this volume. For a cone roof storage tank this is found to be \$14,103. Then using a bare module factor of 1.9 for carbon steel

$$BMC_{Na_2S_4\ Storage} = 1.9 \cdot \$14{,}103 = \$\ 26{,}795$$

$$AC_{Na_2S_4\ Storage} = 0.24 \cdot BMC_{Na_2S_4\ Storage} = 0.24 \cdot \$26{,}795 = \ \$6{,}431$$

$$EP_4 = EP_3 - AC_{NaBr} - AC_{NaBr\ Storage} - AC_{Na_2S_4} - AC_{Na_2S_4\ Storage}$$

$$EP_4 = -\$2{,}290{,}106 - \$101{,}589 - \$6{,}600 - \$5{,}968 - \$6{,}431 = -\$2{,}410{,}694$$

9.4 Appendix D – Level 5 Calculation

9.4.1 Annualized Building and Site Preparation Costs

Building and site preparation costs for 2014 are estimated to be \$1,107 per m^2. Also assume about $500 m^2$/MW

$$Cost\ per\ MW = (\$1{,}107/m^2)(500\ m^2/MW) = \$553{,}444/MW$$

Given that the power is equal to 3 MW

$$BMC_{BP} = (Cost\ per\ MW) \cdot (Power) = (\$553{,}444/MW) \cdot (3\ MW) = \$1{,}660{,}331$$

$$AC_{BP} = 0.24 \cdot BMC_{BP} = 0.24 \cdot \$1{,}660{,}330 = \$398{,}479$$

9.4.2 Annualized Cost for Control System

The bare module cost of the control system is estimated to be \$22,509. Therefore,

$$AC_{CS} = 0.24 \cdot BMC_{CS} = 0.24 \cdot \$22{,}509 = \$5{,}402$$

9.4.3 Annualized Cost for Control System

The remaining costs are estimated to be \$56/kW. Therefore,

$$BMC_{RC} = (Cost\ per\ kW) \cdot (Power) = (\$56/kW) \cdot (3\ MW) \cdot (1000\ kW/MW) = \$168{,}000$$

$$AC_{RC} = 0.24 \cdot BMC_{RC} = \$40{,}320$$

$$EP_5 = EP_4 - AC_{BP} - AC_{CS} - AC_{RC}$$

$$EP_5 = -\$2{,}410{,}694 - \$398{,}479 - \$5{,}402 - \$40{,}320 = -\$2{,}854{,}895$$

9.5 Appendix E – Cost Summaries

9.5.1 Capital Cost Summary – For a membrane cost of $1,000 per m^2

Date to which estimate applies: 2014

Job title: **Bromide-polysulfide Redox-flow Battery Design** Location:Knoxville, TN, US By Group 3

Equipment Identification	Capacity or Size Specifications	Purchased Equipment Cost: Year 2004	Purchased Equipment Cost: Target Year*	Base Bare Module Factor, F_{BM}	Base Bare Module Cost, C_{BM}	Material Factor, F_M	Pressure or other Factors, F_P	Actual Bare Module Factor, F_{BM}	Actual Bare Module Cost, C_{BM}	Annulaized Cost Factor	Annualized Cost	Total
Pumps												
Sodium Bromide	Centrifugal, Cast Iron, 0.7 kW, 50,000 Pa	$ 3,500	$ 4,936	4.2	$ 20,731	1.4	1	4.2	$ 20,731	0.24	$ 4,975	
Sodium Tetrasulfide	Centrifugal, Nickel Alloy, 1.5 kW, 50,000 Pa	$ 4,000	$ 5,641	7	$ 39,487	3.5	1	7	$ 39,487	0.24	$ 9,477	
Total Pumps									**$ 60,218**			**$ 14,452**
lisc: Solution, Materials, etc.												
Cell Cost ($1000 Membrane)	1 m2 area, 20 stacks, 300 Volts/amp		$ 8,992,500	1.4	$ 12,589,500			1.4	$ 12,589,500	0.24	$ 3,021,480	
Plant Area Costs	1500 m2		$ 1,660,331						$ 1,660,331	0.24	$ 398,479	
Control System	Control System from Paper		$ 22,509						$ 22,509	0.24	$ 5,402	
Remaining Costs	$56/kW		$ 168,000						$ 168,000	0.24	$ 40,320	
Sodium Bromide	86385 kg, $4/kg	-	-		$ 423,289				$ 423,289	0.24	$ 101,589	
Sodium Tetrasulfide	73138 kg, $.34/kg	-	-		$ 24,867				$ 24,867	0.24	$ 5,968	
Total Misc:									**$ 14,888,495**			**$ 3,573,239**
Storage Vessels												
Sodium Bromide Storage	Cone Roof, Fiberglass, 30 m^3	$ 6,500	$ 9,167	$ 3	$ 27,500			$ 3	$ 27,500	$ 0	$ 6,600	
Sodium Tetrasulfide Storage	Cone Roof, Carbon Steel, 83m^3	$ 10,000	$ 14,103	$ 2	$ 26,795			$ 2	$ 26,795	$ 0	$ 6,431	
Total Storage Vessels									**$ 54,295**			**$ 13,031**
Total Bare Module Cost									**$ 15,003,008**			
Total Annualized Cost												**$3,600,722**

*2004 cost x (Target Year Cost Index/2004 Cost Index)

1.1.1 Capital Cost Summary – Neglecting membrane cost

Date to which estimate applies: 2014

Job title: **Bromide-polysulfide Redox-flow Battery Design** Location: Knoxville,TN, U By Group 3

Equipment Identification	Capacity or Size Specifications	Purchased Equipment Cost: Year 2004	Purchased Equipment Cost: Target Year*	Base Bare Module Factor, F_{BM}	Base Bare Module Cost, C_{BM}	Materia l Factor, F_M	Pressure or other Factors, F_P	Actual Bare Module Factor, F_{BM}	Actual Bare Module Cost, C_{BM}	Annulaized Cost Factor	Annualized Cost	Total
Pumps												
Sodium Bromide	Centrifugal, Cast Iron, 0.7 kW, 50,000 Pa	$ 3,500	$ 4,936	4.2	$ 20,731	1.4	1	4.2	$ 20,731	0.24	$ 4,975	
Sodium Tetrasulfide	Centrifugal, Nickel Alloy, 1.5 kW, 50,000 Pa	$ 4,000	$ 5,641	7	$ 39,487	3.5	1	7	$ 39,487	0.24	$ 9,477	
Total Pumps									**$ 60,218**			**$ 14,452**
lisc: Solution, Materials, etc.												
Cell Cost ($50 Membrane)	1 m2 area, 20 stacks, 300 Volts/amp		$ 750,750	1.4	$ 1,051,050			1.4	$ 1,051,050	0.24	$ 252,252	
Plant Area Costs	1500 m2		$ 1,660,331						$ 1,660,331	0.24	$ 398,479	
Control System	Control System from Paper		$ 22,509						$ 22,509	0.24	$ 5,402	
Extras	From Paper $56/kW		$ 168,000						$ 168,000	0.24	$ 40,320	
Sodium Bromide	86385 kg, $4/kg	-	-		$ 423,289				$ 423,289	0.24	$ 101,589	
Sodium Tetrasulfide	73138 kg, $.34/kg	-	-		$ 24,867				$ 24,867	0.24	$ 5,968	
Total Misc:									**$ 3,350,045**			**$ 804,011**
Storage Vessels												
Sodium Bromide Storage	Cone Roof, Fiberglass, 30 m^3	$ 6,500	$ 9,167	3	$ 27,500			3	$ 27,500	0.24	$ 6,600	
Sodium Tetrasulfide Storage	Cone Roof, Carbon Steel, 83m^3	$ 10,000	$ 14,103	1.9	$ 26,795			1.9	$ 26,795	0.24	$ 6,431	
Total Storage Vessels									**$ 54,295**			**$ 13,031**
Total Bare Module Cost									**$ 3,464,558**			
Total Annualized Cost												**$831,494**

*2004 cost x (Target Year Cost Index/2004 Cost Index)

1.1.2 Operating Cost Summary – Initial input information

	ChE Index	567.7	
	Second/yr	583973.16	
	SUF =	1	
	Operability	0.94	
Process Vessel	D =	2.10	m
	HETP =	1.14	m
	Active H =	18.97	m
	Top Space =	2.10	m
	Botom Space	0.00	m
	Total Height =	21.07	m
	T Volumn =	72.93	m^3
	Packing V=	65.66	m^3
	Vessel Cost=	$ 73.43	
	Packing $/m3 =	$ 3.20	
	Packing costs	210.00	
	Seconds per year =	31557600	
Feed	No 2 Fuel Oil	0.95	$/gal
Bottoms Product	No 6 Fuel Oil	0.75	$/gal
	Electricity	0.12	$/kWh
	Cooling W.	0.07	$/m3
	Steam	0.02	$/kg

1.1.3 Operating Cost Summary -Membrane cost of $1,000 per m^2

Effective Date to Which Estimate Applies: 2014	ost Index Value:	567.7						
Capital								
Fixed capital, CFC	$3,600,722							
Working capital (10-20% of fixed capital), CWC		$540,108	1					
Total capital investment, CTC	$4,140,830							
Manufacturing Expenses								
Direct							$/yr	$/m3
Raw materials	No 2 Fuel Oil	459794232	kg	0.36	$/kgr		$164,171,785	$312,310,263
By-product credits	No 6 Fuel Oil	50429044.8	kg	0.27	$/kgr		-$13,595,300	-$25,862,858
Catalysts and solvents	0						$0	$0
Operating labor	6	Operators	51162.8	$/year			$306,977	$1
Supervisory and clerical labor (10-20% of operating labor)							$46,046	$87,596
Utilities								$0
Steam	1701698	kg/y	33	barg @	$0.019	$/kg	$32,461	$61,752
Electricity	0.01	kWh @	0.12098	$/kWh			$0	$0
Process water	2.82	m3 @		$/m3				$0
Demin water	0.00	m3 @		$/m3				$0
Coolng water	2847758	m3 @	0.07263	$/m3			$206,823	$393,448
Waste disposal		kg @		$/kg				$0
Maintenance and repairs (2-10% of fixed capital)							$216,043	$410,987
Operating supplies (10-20% of maint & repairs)							$32,406	$61,648
Laboratory charges (10-20% of operating labor)							$46,046	$87,596
Patents and royalties (0-6% of total expense)							$0	$0
Total, ADME							$151,463,290	$288,134,406
Indirect								
Overhead (payroll and plant), packaging, storage (50-70% of op. Labor+supervision+ maint.)							$341,440	
Local taxes (1-3% of fixed capital)							$72,014	
Insurance (1-2% of fixed capital)							$36,007	
Total, AIME								
Total manufacturing expense, AME=ADME+AIME							$151,912,751	
General Expenses								
Administrative costs (2% of overhead)							$85,360	
Distribution and selling (1% of total expense)							$3,208,223	
Research and development (5% of total expense)							$1,604,112	
Total general expense, AGE							$4,897,695	
Depreciation (approximately 10% of fixed capital), ABD							$3,600,722	
Total Expenses, ATE							$160,411,168	
Revenue from Sales		kg/yr @		$/kg), As			$161,260,569	
Net annual profit, ANP							$849,401	
Income taxes (net annual profit times the tax rate), AIT							$297,290	
Net annual profit after taxes (ANP-AIT), ANNP							$552,111	

Aftertax rate of return, i = (1.5 ANNP/CTC) x 100 = 20 %

1.1.4 Operating Cost Summary – Neglecting membrane cost

Effective Date to Which Estimate Applies: 2014	ost Index Value:	567.7							
Capital									
Fixed capital, CFC	$831,494								
Working capital (10-20% of fixed capital), CWC		$124,724	1						
Total capital investment, CTC	$956,218								
Manufacturing Expenses									
Direct							$/yr	$/m3	
Raw materials	No 2 Fuel Oil	459794232	kg	0.36	$/kgr		$164,171,785	$312,310,263	
By-product credits	No 6 Fuel Oil	50429044.8	kg	0.27	$/kgr		-$13,595,300	-$25,862,858	
Catalysts and solvents	0						$0	$0	
Operating labor	6	Operators	51162.8	$/year			$306,977	$1	
Supervisory and clerical labor (10-20% of operating labor)							$46,046	$87,596	
Utilities								$0	
Steam	1701698	kg/y	33	barg @	$0.019	$/kg	$32,461	$61,752	
Electricity	0.01	kWh @	0.12098	$/kWh			$0	$0	
Process water	2.82	m3 @		$/m3				$0	
Demin water	0.00	m3 @		$/m3				$0	
Coolng water	2847758	m3 @	0.07263	$/m3			$206,823	$393,448	
Waste disposal		kg @		$/kg				$0	
Maintenance and repairs (2-10% of fixed capital)							$49,890	$94,907	
Operating supplies (10-20% of maint & repairs)							$7,483	$14,236	
Laboratory charges (10-20% of operating labor)							$46,046	$87,596	
Patents and royalties (0-6% of total expense)							$0	$0	
Total, ADME							$151,272,213	$287,770,913	
Indirect									
Overhead (payroll and plant), packaging, storage (50-70% of op. Labor+supervision+ maint.)							$241,748		
Local taxes (1-3% of fixed capital)							$16,630		
Insurance (1-2% of fixed capital)							$8,315		
Total, AIME									
Total manufacturing expense, AME= ADME+AIME							$151,538,905		
General Expenses									
Administrative costs (2% of overhead)							$60,437		
Distribution and selling (1% of total expense)							$3,142,904		
Research and development (5% of total expense)							$1,571,452		
Total general expense, AGE							$4,774,793		
Depreciation (approximately 10% of fixed capital), ABD							$831,494		
Total Expenses, ATE							$157,145,192		
Revenue from Sales		kg/yr @		$/kg), As			$157,341,339		
Net annual profit, ANP							$196,147		
Income taxes (net annual profit times the tax rate), AIT							$68,652		
Net annual profit after taxes (ANP-AIT), ANNP							$127,496		
				Aftertax rate of return, i = (1.5 ANNP/CTC) x 100 =				20	%