University of Pennsylvania

Department of Chemical & Biomolecular Engineering

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Dear Dr. Lee and Professor Fabiano,

The following document contains the process that we have designed as a solution to the design project proposed by Arthur Rempel and Kyra Berger. This process predominantly treats brine wastewater from desalination plants while simultaneously recovering a significant fraction of the inlet water, and substantial amounts of both $MgCl_2$ and KCl crystals. Our proposed process contains three distinct separation phases; in the first, saturated brine wastewater is passed through multistage flash evaporation vessels, which removes a vast majority of the NaCl in solution alongside other impurities that may be present in the brine. Phase two utilizes crystallizer, centrifuge and filtration units and operates under the premise that carnallite crystals ($KMgCl_3*6H_2O$) will precipitate given the appropriate operating conditions. Subsequently, phase two continues with the crystallization of KCl crystals after re-suspending the carnallite in solution. Finally, phase three embodies the most essential crystallization of the process; $MgCl_2$ crystals are formed via a forced circulation capacity crystallizer. Our process estimates a net annual revenue of $2.6bn from the recovered $MgCl_2$ crystals alone, allowing us to not only compete with, but transform GE's current Brine Concentrator System.

This document contains an intricate process design pertaining to our $MgCl_2$ and KCl separation unit in addition to GE's Brine Concentrator. Furthermore, an in-depth economic and profitability analysis was conducted in comparing our design to GE's system; as a result, we have included recommendations for the implementation of our plant while considering our competitor.

Our proposed plant will be located in San Diego, California, and is suited to treat 1-2mgd. From an economic standpoint, the proposed plant has increased both GE's ROI and NPV by approximately 22.53% and $20,749,200, respectively.

Thank you for your assistance during this project.

Arthur Rempel

Elyssa Gensib

Kyra Berger

Aspen Walker

$MgCl_2$ and KCl Recovery from Brine Wastewater

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Capstone Design Project

Project Submitted to: Dr. Daeyoen Lee, Prof. Fabiano

Project Proposed by: Arthur Rempel, Kyra Berger

Department of Chemical & Biomolecular Engineering

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TABLE OF CONTENTS

Abstract

This project's aim was to design an improved brine wastewater treatment system for desalination facilities. While a multitude of methods exist to do so, General Electric (GE)'s brine concentrator is leading the market by providing a method that not only treats the brine waste, but also recovers anywhere from 60-94% of the water from the feed. However, their brine concentrator is relatively inefficient from both a financial and energetic perspective; our goal was to develop a system to match their results, while limiting costs and energy usage as best possible.

We subsequently designed a system (referred to from here on out as the '$MgCl_2$ Separation Unit') to accomplish the aforementioned objectives. In addition to recovering pure water from concentrated brine, our process also recovers high purity $MgCl_2$ and KCl crystals that are later sold to alleviate the overall process costs. The $MgCl_2$ Separation Unit saw an increased ROI by 22.53% as compared to the GE Brine Concentrator, as well as a surge in NPV of $20,749,200. Additionally, we estimate an aggregate equipment cost of roughly $6,960,000, significantly less than GE's $8,980,000, and utilities costs of $1,590,000 relative to their $6,920,000. The GE process deals with vapor phase water and thus relies on compressors for pressure changes, which are costly in terms of energy consumption and capital costs, whereas our proposed $MgCl_2$ Separator Unit is limited to liquid/aqueous streams, eliminating the need for costly compressors (from both an energy and economic standpoint). Finally, while both systems are fit to treat 1-2mgd, the $MgCl_2$ Separation Unit was optimized to recover roughly 84% of feed water, positioning itself in the upper limit of GE's possible water recovery spectrum.

Introduction

Clean water sources have recently become a scarce commodity across the globe. Brazil has reported its worst drought in the past 50 years [1], while South Africa expects a 32% downfall in maize production this year alone due to a shortage of clean water. And, of course, California is domestically affected, where all 12 of its clean-water reservoirs are at abnormally low levels (and have been, for some time now).

One available method to produce clean water focuses on desalination. Desalination is a process that removes salts and other minerals from untreated water, producing water that is clean for both human consumption and irrigation [2]. Ocean water makes up approximately 97% of the Earth's total water supply [3], suggesting that desalination is an attractive option as a clean water source - especially for coastal regions, like the three aforementioned areas.

Desalination, depicted in Figure 1, is typically conducted either via multistage flash thermal vacuum evaporation, in which saline water is boiled at extremely low pressures with excess heat, or reverse osmosis membrane separation (ROMS), a process by which saline water is forced through a series of semi-permeable membranes at high pressures to filter out salts, minerals, and residual metals. Currently, only 1% of the Earth's clean water supply is produced via desalination, although the UN predicts that by the year 2025, "14% of the world's population [will be] encountering water scarcity [4]." Furthermore, the Western

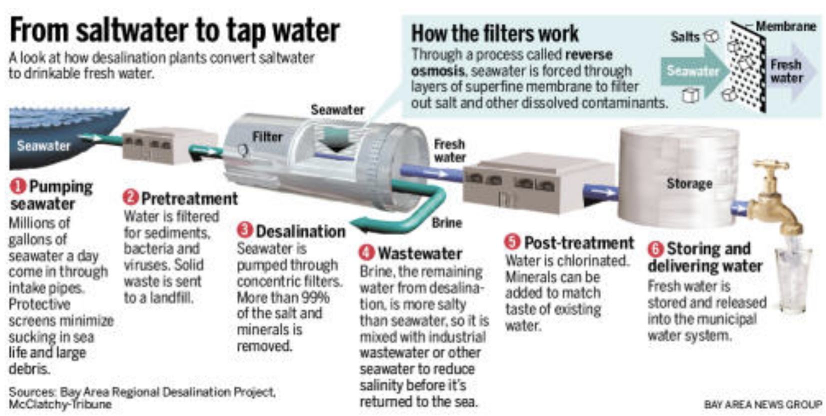


Figure1. General schematic of the desalination process

Hemisphere's largest desalination plant is projected to open in May of 2016 in southern California [5], which will theoretically provide 50 million gallons of drinking water per day to the San Diego county area.

Unfortunately, high costs (as compared to other clean water sources) accompany the desalination process, predominantly due to inefficient energy consumption and ample wastewater by-product, usually referred to as brine water. Brine water is highly concentrated salt water, which also contains other impurities like precious metals that were used to purify the water during the desalination process. Due to its high salt concentrations (typically about 30%, whereas ocean water usually spans the 1-5% range), brine water will sink upon reentering the ocean, potentially affecting marine life and posing major environmental concerns. Currently, a plethora of permits and federal regulations are required for disposing brine water in sewer systems or in the ocean, which contributes to the high economic and energetic costs associated with desalination.

In terms of energy consumption, the desalination process can consume as little as $3kWh/m^3$ (ROMS) [2], although this value can reach anywhere from 13.5 up to $25.5kWh/m^3$, in the case of thermal desalination. For context, fresh water supplies typically require a maximum of $0.2kWh/m^3$ for treatment, clearly much less than any desalination process and immediately rendering the desalination process energetically costly.

With respect to financials, desalinated water costs on average $2,000 per acre-foot (1acre-foot is about $3.26x10^5$gal, or the equivalent amount of water consumed on average, annually by a family of five) [5], which is four times as much as money spent on water from fresh sources; these high costs typically arise due to the difficulty in managing the brine by-product. On average, storage facilities charge between $2.00 and $2.50 to store one cubic meter of brine [8], which, in the case of the aforementioned desalination plant, would equate to somewhere between $378,500 and $473,125 per day, assuming that one gallon of brine is produced for every gallon of purified water [5].

A series of brine management strategies exist and are implemented in desalination plants across the globe, although they are highly dependent on a variety of variables: location, land availability, air moisture content, legal/federal permitting requirements, and economic feasibility.

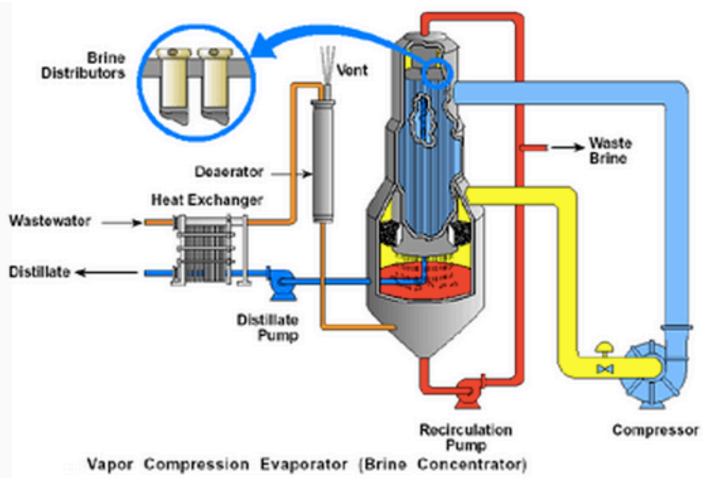


Figure 2. General Schematic of a brine concentrator

Brine concentrators and evaporation ponds are attractive options; both methods allow for additional water recovery from brine wastewater (up to 94%) [7], and completely remove the salts from the environment "by ultimately sequestering the remaining brine in a landfill or in a closed and sealed evaporation pond." Thus, the combination of these brine concentrators and evaporation ponds are deemed environmentally desirable. Figure 2 shows a general schematic of General Electric's brine concentrator [9].

Unfortunately, excessive costs and energy consumption are accompanied by these environmental benefits, rendering these processes infeasible more often than not. Brine concentrators typically consume between 60 and 100kWh per 1000 gallons of brine – assuming a cost of $0.77/kWh, a brine concentrator will generally require $4600-$7700 per day to process 1 mgd (million gallons per day) of brine [8]. Or, in the case of the plant from above, a daily charge between $230,000 and $385,000. Notice that these costs are between 20-40% lower than those of the storage facility mentioned earlier, but they are not feasible or efficient by any means.

Table 1 below further illustrates the high costs associated with treating 10 mgd of brine with these two strategies as compared to other methods [7]. Additionally, Table 2 depicts the energy consumption pertaining to all of the specified methods, clearly highlighting the excess energy required for brine concentrators in comparison to the other methods.

Table 1. Costs of evaporation ponds and brine concentrators compared to other strategies

10 MGD	Pipeline to Yuma	Evaporation Pond	Brine Concentrator	Soften/ RO/ VSEP	Wetlands Surface Discharge	Injection Well
Capital	$266.11	$651.69	$272.71	$286.56	$150.22	$ 114.46
O&M	$ 0.62	$ 3.50	$ 29.75	$ 6.90	$ 1.75	$ 11.31
Annualized	$ 14.92	$ 40.26	$ 44.40	$ 22.30	$ 10.37	$ 17.46

Table 2. Energy consumption of various brine-wastewater treatment methods

10 MGD	Pipeline to Yuma ****	Evaporation Pond	Brine Concentrator	Soften/RO VSEP	Wetlands Surface Discharge	Injection Well
Energy* (kilowatt-hours)	minimal	1,146,000	310,250,000	68,135,000	minimal	143,769,000
Energy Cost**	minimal	$88,000	$23,889,000	$662,000	minimal	$11,070,000
Water Recovered*** (af)	0	0	10,528	9,238	0	0

In addition to energy and management costs, brine wastewater also suffers from economic opportunity costs via the lack of recovering useful byproducts. These byproducts may consist of selenium, nitrates, gypsum and calcium compounds, which "are widely used in the building industry for drywall, plaster, and cement [10]." Other byproducts could consist of magnesium salts, which are useful in the medical industry, and boron, which is a recent hot topic in high-efficiency electronics. Finally, if purified appropriately, any recovered sea salt could be sold as a raw material to further increase profitability.

Brine wastewater regulations are ubiquitous across the United States; while they vary from area to area, this plethora of strenuous legal regulations introduces an extremely costly step to the desalination process. The San Diego Regional Water Quality Control Board oversees the regulation processes for the area in which our plant will be located. They mandate that the brine wastewater salinity limit is less than or equal to 40%, or has a discharge dilution ratio of 7.5:1, in order to be reintroduced into the ocean post-desalination. Additionally, the point of discharge must be greater than 1,000 feet from the shore. [11]

It is inevitable to encounter a monetary loss when dealing with wastewater treatment. Therefore, the overall target of a brine wastewater treatment system, like the GE brine concentrator, is not to generate a profit; rather, the goal is to abide by the environmental regulations and restrictions in the least costly way possible.

The purpose of this design project was to transform the brine concentrator/evaporation pond system by designing a new system that is more efficient in terms of both energy and cost, since the environmental benefits are unprecedented among other common brine-management techniques. If possible, this would not only be economically attractive to investors and desalination firms, but would also give rise to a new, affordable, clean, and abundant water resource.

Objective-Time Chart

The goal of this project was to create a brine wastewater management system that is more economically and energetically sustainable than current systems in the industry. The scope of this project includes completing an analysis of the most profitable recoverables from salt water, the modeling and financial analysis of the one of the leading methods of brine wastewater management created by GE, designing a process to recover $MgCl_2$ and KCl from brine wastewater, and determining the financial feasibility of a plant that recovers $MgCl_2$ and KCl in comparison to the financial obligations associated with GE's system. Project leaders are Arthur Rempel, Kyra Berger, Elyssa Gensib, and Aspen Walker.

Deliverable	**Description**	**Date Accomplished**
Determine location for plant/brine feed composition	Analyzing difference between saline levels and salt compositions of various potential water sources to determine best location and water source for plant	February 2nd
Model GE Brine Concentrator	Determining equipment involved in GE's brine concentrator and creating the process flow diagram in ASPEN with appropriate operating conditions	February 16th
Identify profitable recoverables	Research potential salt recoverables and complete a cost analysis of potential revenue for various compositions depending on water source	February 16th
Model MgCl2 separation from patent	Determine equipment involved in patent and use Excel to calculate material balances for streams	March 22nd
Determine Utility Requirements Needed	Size equipment for both the GE concentrator and the MgCl2 separation system and determine utility requirements for heating and cooling.	March 26th
Complete Financial Analysis	Calculate fixed and variable costs for both systems and determine NPV and ROI	March 30th
Complete Report		April 12th

Market & Competitive Analysis

In a survey conducted across 137 drinking water membrane plants in the United States with a capacity of over 98 m^3/day, 48% discharge the brine to surface waters, 23% discharge to wastewater treatment plants, 12% use land application, 10% use deep well injections, and 6% use evaporation ponds. The only plants that used deep well injection and land application were located in Florida, a reflection of the geological and climate conditions of the area. Wastewater treatment facilities were used primarily by smaller plants, due to the limited capacity of the treatment plants. The cost of these disposal methods ranged between 5-33% of the total cost of desalination. [11]. Figure 3 shows the various methods of wastewater disposal currently in use in Southern California, ranging from ocean outfalls to brine concentrators.

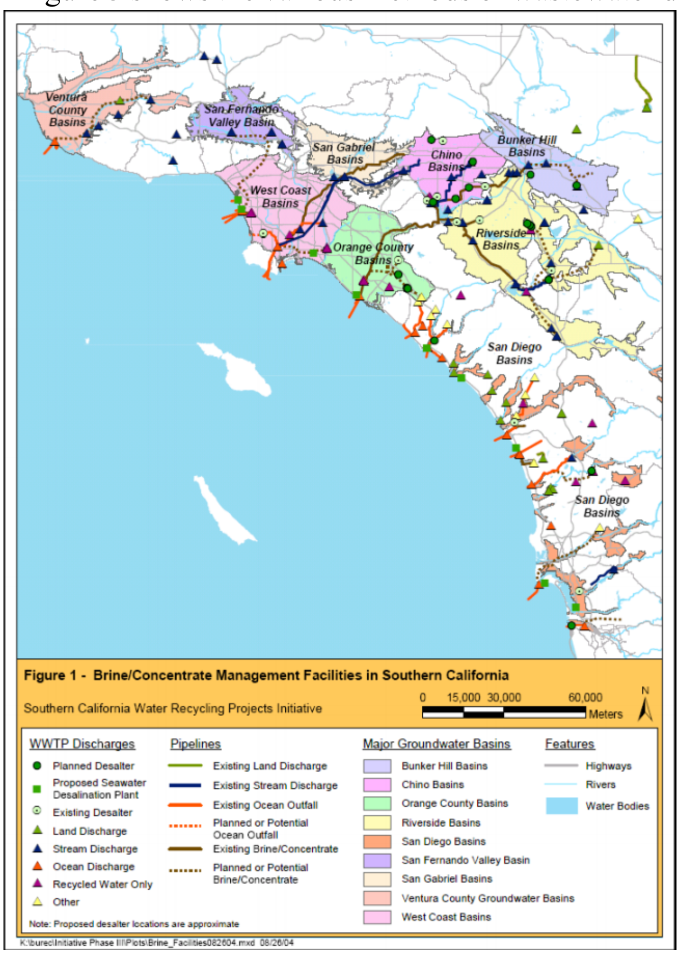


Figure 3. Desalination plants and brine wastewater treatments in Southern California.

Liquid Disposal

This method of managing brine wastewater involves discharging the brine either downstream from the process, either into the ocean or through deep well injection. These methods are desirable due to the lack of treatment of the brine required before disposing the wastewater. Systems involving a downstream wastewater treatment facility of ocean discharge involve a brine line or a sewer to transport the brine which has some disadvantages. First, it can be challenging to find a wastewater treatment plant that has the capacity to treat the brine wastewater depending on the volume of waste produced by the desalination plant. Additionally, there can be issues surrounding regulatory requirements and acquiring appropriate permits for this method of discharge. Ocean discharge is particularly challenging due to the potential effects on the environment as a result of introducing highly concentrated, denser material back into the ocean, potentially harming the ocean floor habitats; however, regulatory steps are put into place to limit the environmental impact of the wastewater, which results in a less environmentally harmful system of disposal. Deep well injection can be a viable alternative. This method has proven to be environmentally safe, does not require pre-treatment of the brine, is simple to design and operate, and has a minimal aesthetic impact on the local community. Deep well injection is complicated due to the limited land availability and various geologic conditions to consider which results in many regulatory requirements as well as the need to obtain a permit. Building these wells can be costly – the classification of brine as an industrial waste requires specific, more expensive construction materials for the wells and an additional well must be drilled to monitor the disposal well. However, it is possible to use abandoned or active oil wells that already exist which can reduce costs associated with construction of this

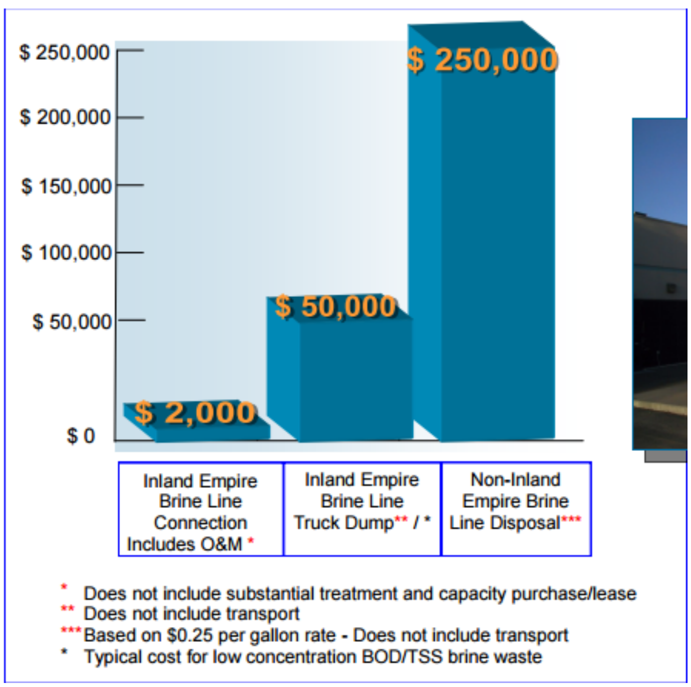


Figure 4. The cost for disposing 1mgd of brine wastewater provided by Inland Empire Brine Line for a brine line connection, a brine line truck dump, and a brine line disposal for inland and non-inland plants [13].

method of brine disposal [12]. An estimate of associated cost with disposing 1 mgd of brine wastewater is expressed in Figure 4.

Landfill Disposal

Disposing the brine waste in landfills can be an feasible method of waste treatment. This method requires little treatment by the desalination plant and transfers the responsibility of managing the waste to the landfill facility. However, there is a large cost associated with transporting the wastewater and using the landfill and there is a loss of potential revenue from disposing of potentially valuable water along with the undesired salts in the brine. Additionally, the classification of the brine as an industrial waste limits the use of incineration as a method of disposal [12]. CDM Smith estimates that it would cost \$230,000 per completed well and the pipeline from the plant to the well for a well that sustains an injection rate of 200 to 400 gpm [14].

Evaporation Ponds

Evaporation ponds involve creating an artificial pond with a large surface area designed to use sunlight to evaporate the water from the waste. This method of disposal has been widely used historically in large part due to its simple design technologically. This process is limited by the availability of land for use as an evaporation pond as well as the high cost associated with building and maintaining this method of waste treatment due to the need to line the ponds to avoid being classified as Class V wells in addition to the need to excavate the salt periodically, as seen in Figure 5. Additionally, there are concerns about potential groundwater infiltration, which can have an effect on drinking water in the area. These evaporation ponds can also have an adverse effect on surrounding habitats and are considered aesthetically unpleasant by community members who live near the plant [12].

Zero Liquid Discharge

Zero Liquid Discharge systems use mechanical or thermal evaporation in combination with crystallization to reduce the volume of the waste stream and recover additional water for use as a commercial product. These systems are desirable in that they have been proven effective in industry, they produce a high-quality stream of water, and require less space than other traditional methods. However, they have high capital cost as well as high operation and maintenance costs, primarily due to the costs associated with the energy requirements for operating the system as seen in Table 3 and Table 4 [15].

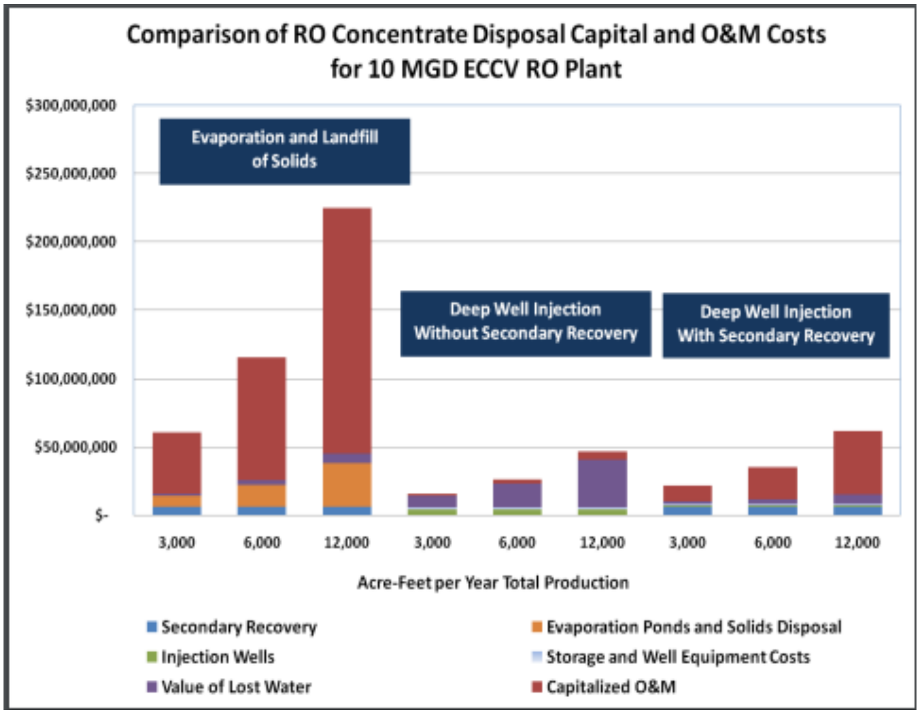


Figure 5. Cost of Capital and Operation and Maintenance Costs for Evaporation and Landfill methods versus Deep Well Injection [14]

Table 3. Capital Costs for a 0.2mgd Zero Liquid Discharge system in 2006

Cost	**0.2 mgd**	**1.0 mgd**	**5.0 mgd**
Total Capital Cost Including Equipment Installation (in USD)	6,170,000	20,681,000	59,826,000

Costs provided by Big Bear Area Regional Wastewater Agency for a 0.2mgd system in 2006. Costs for additional flow rates estimated using the following formula: Cost 2 = (Flow 2/Flow 1)^0.66*Cost 1, where Flow 1 is 0.2 mgd [15].

Table 4. Operation and Maintenance Costs for a 1 mgd Zero Liquid Discharge system

Component	**Operation and Maintenance Cost (USD/year)**
Power	4,844,000
Parts	1,035,000
Chemicals	282,000
Maintenance	621,000
Labor	225,000
Total	7,007,000

Costs provided by GE-Ionics based on a feed flow of 1 mgd [15].

Process Synthesis

Comprehensive research was conducted to analyze the current methods of wastewater treatment and identify the superior system to use as a preliminary model for brine disposal. Zero Liquid Discharge (ZLD), shown in Figure 6 was one of the most widely used methods of treatment and allowed for recovery of water to be used towards additional revenue.

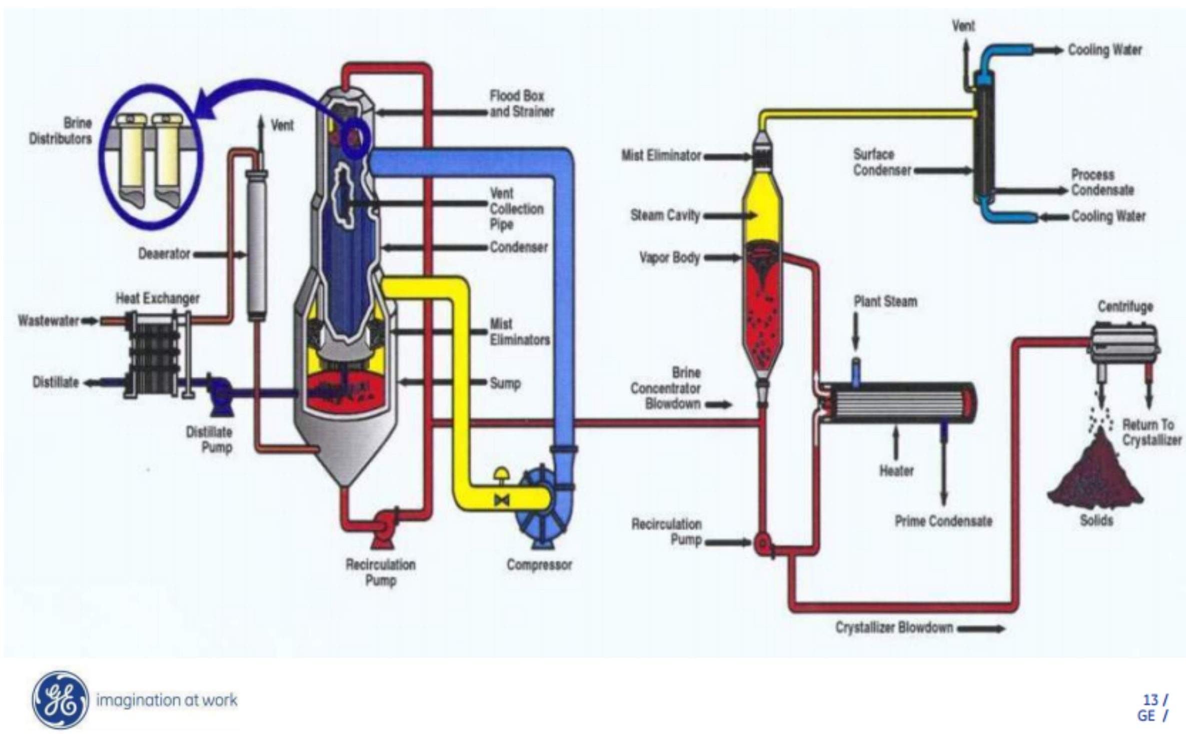


Figure 6. Diagram of GE Brine Concentrator System [21]

Additionally, GE's ZLD equipment is effective in treating a wide variety of waste flows and is more energy-efficient than other single-effect evaporators current in the market. GE's system is limited in that it is not designed to recover specific salts from the original brine waste stream. Various salt recoverables were considered, including $MgSO_4$, Lithium, Bromine, $LiCO_3$, gypsum, $CaCl_2$, and $MgCl_2$. A preliminary analysis of potential revenue from various salts in multiple locations was conducted, and the results are shown in Table 5. $MgCl_2$ was chosen for further research due to its high profitability, its prevalence in ocean water, and its extensive patent literature regarding separation methods.

Table 5. Brine Recoverables with Annual Revenue at Various Locations

Product	Location	Body of Water	Flow Rate (mgd)	Mass Recoverable (metric ton/day)	Annual Revenue (USD/year)
99.5% pure $MgSO_4$	San Diego	Sea	1	13.1	$2,860,000
	Cape may	Groundwater	1	12.4	$2,720,000
$MgCl_2$	San Diego	Sea	1	20.5	$2,610,000
$CaCl_2$	San Diego	Sea	1	3.6	$264,000
	Cape May	Groundwater	1	3.4	$250,000
Lithium	Atacama, Chile	Salt flat	0.05	0.3	$436,000
	Saratoga Springs, NY	Groundwater	0.05	0.2	$374,000
	Green River, WY	River	0.5	0.2	$311,000
	Hombre Muerto, Argentina	Salt lake	0.05	0.1	$162,000
	Rincon, Argentina	Salt lake	0.05	0.1	$103,000
	Great Salt Lake	Salt lake	0.05	0.01	$14,000
Lithium Carbonate	Silver Peak, NV	Dry lake	0.05	0.04	$62,000
Gypsum	San Diego	Sea	1	5.6	$61,000
	Cape May	Groundwater	1	5.3	$58,000
Bromine	Searles Lake	Dry lake	0.05	0.2	$117,000

Magnesium Chloride Market

There is an existing and robust market for $MgCl_2$. Approximately 60% of magnesium compounds produced annually are extracted from seawater and brine [16]. The demand for $MgCl_2$ exists in many industries. One important use of $MgCl_2$ is as a mineral supplement. The production of mineral supplements is a $2.3 billion industry. Nutrition experts state that magnesium is the fastest-growing supplement in the mineral category [17].

The United State Geological Survey estimated that salt production was a $2.2 billion industry in 2014. Approximately 43% of all salt sales are attributed to road de-icing. The state and local governments spend over $2.3 billion to control snow and ice annually, mostly due to the prices for salt. The market is also stable – salt contracts often involve an 80% delivery acceptance rate, guaranteeing demand from state and local agencies. These agencies are compelled to purchase these large quantities of salt due to the reductions in crashes and accidents, which are estimated to be reduced by up to 88% with the addition of salt on the roads. Additionally, a city can spend between $300 million and $700 million per day of unusable roads [18].

The market for $MgCl_2$ can be considered fairly stable. In 2014, the domestic salt market produced $2.2 billion annually, with highway deicing consuming 43% of the total salt produced [19]. In 2015, the production increased by 6%, resulting in $2.3 billion in product, 46% of which went towards highway deicing [20]. This increase in market demand was one reason why we chose $MgCl_2$ as our primary recoverable from brine.

The other consideration for $MgCl_2$ is how far away the demand is from our production site in San Diego. Despite the lack of precipitation in San Diego, the Sierra Nevada mountain range spans most of the state of California and has an average annual snowfall of over 48 inches, as seen in Figure 7 [21]. The areas in this range would provide a stable market with high demand for our $MgCl_2$ product.

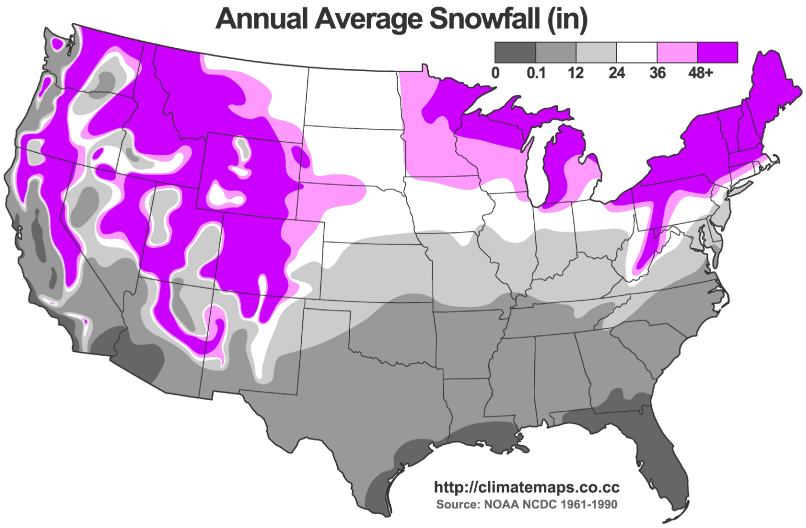


Figure 7. Average annual snowfall in the contiguous United States of America.

Potassium Chloride Market

The main source of consumption of KCl is in the form of potash – a fertilizer comprised of 60-62% KCl. The estimated global consumption of potash in 2013 was 33.4 Mt, a $12.7 billion industry. There are few adequate substitutes for KCl, making it a product with stable demand. The International Fertilizer Industry Association projected that potash consumption will increase at a rate of 3.6% per year through 2018 [22].

Additionally, potassium chloride is a mineral supplement, part of the same $2.3 billion industry as $MgCl_2$. KCl is used to prevent low amounts of potassium in the blood in an attempt to protect the body's cells, kidneys, heart, muscles, and nerves [23].

Patent for Separating $MgCl_2$ and KCl

The separation of $MgCl_2$ from brine is patented under U.S. Patent 2479001 (see Appendix C. for full patent) originally published in August of 1949. This method, shown in Figure 8, uses a multi-effect evaporator to recover significant quantities of water from the brine wastewater and forms carnallite, a critical component needed to produce $MgCl_2$.

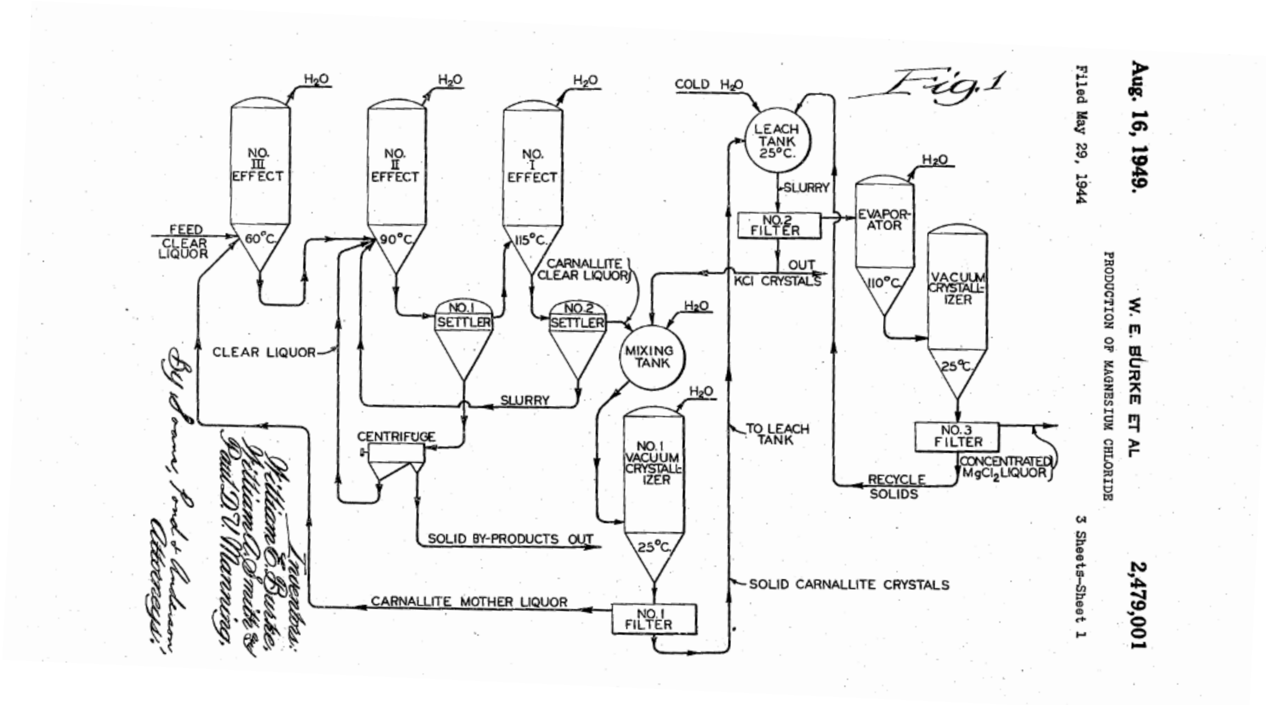
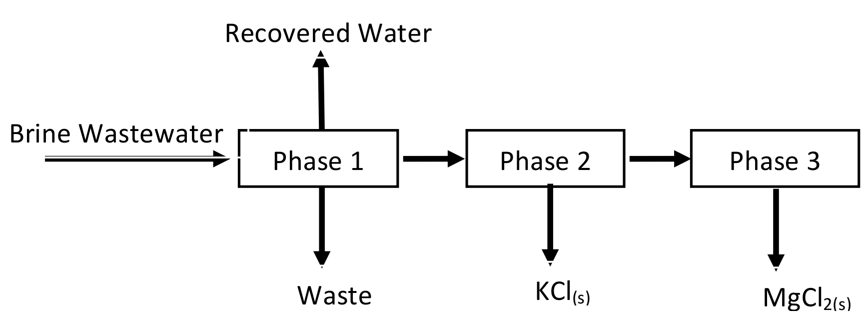


Figure 8. Sketch of process for separation of $MgCl_2$ from U.S. Patent 2479001 [25]

This process consists of three distinct separation phases; the first, which mainly utilizes a multi-stage flash evaporation, is used to recover a large fraction of the inlet water while also removing a highly concentrated waste (mainly composed of $NaCl_{(aq)}$, $NaCl_{(s)}$ and other contaminants.) Phases two and three both use liquid-solid separation units to crystallize and recover KCl and $MgCl_2$ crystals, respectively. The following schematic shows a simple graphical breakdown of this process; note that more intricate design visuals and descriptions are present in the later sections titled ‘Process Flow and Material Balances’ and ‘Process Descriptions.’

Schematic 1. Block Diagram for the $MgCl_2$ recovery process shown in Figure 8.



Assembly Database

Thermodynamic Properties

For the GE Brine Concentrator system, all transport and thermodynamic data was generated using Aspen Plus V8.8 Model's embedded electrolyte package. The ELECNRTL thermodynamic method was used to estimate the thermodynamic properties of all of the streams present in the simulation; ELECNRTL uses NRTL as the standard equation of state when estimating gas properties, and subsequently assumes standard electrolyte chemistry when doing so for liquids and aqueous solutions. The ASPEN simulation generated for the GE brine concentrator can be found in Appendix B in more detail.

For the $MgCl_2$ separation process, ASPEN Plus was used to model all transport and thermodynamic data except solubility data. Figure 9 was generated to determine the relationship between the solubility points of NaCl, $MgCl_2$, and KCl.

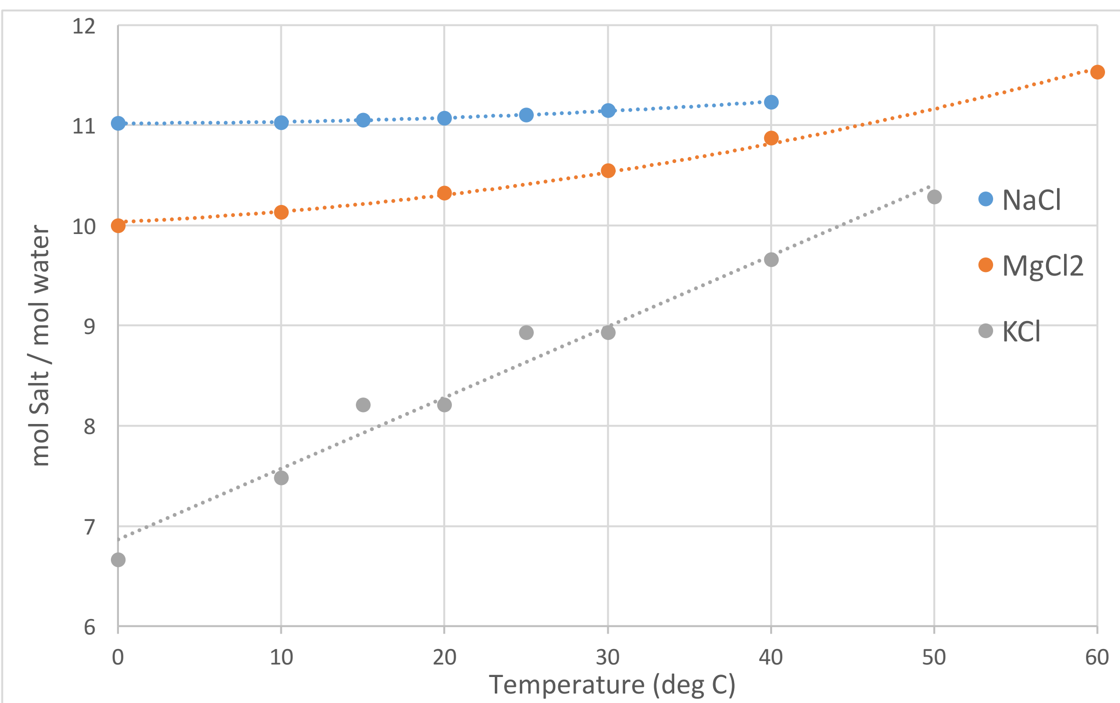


Figure 9. Solubility data generated for NaCl, MgCl2 and KCl

A complete utility database was assembled (shown in Table 6) with steam and cooling water specifications, in addition to electrical data and other general values. This utility database is vital when calculating utility requirements and costing in later sections. Sample calculations for determining utility requirements and utility costs can be found in Appendix A.

Utilities

Table 6. Utility Database for both GE brine concentrator and $MgCl_2$ separation unit

General Utility Data	
Days of operation per year	330
Hours of operation per day	24
Steam Data	
Saturated Steam Pressure	50 psig
Latent Heat of Saturated Steam	912 Btu/lb
Price of Steam	$6/1000 lb
Cooling Water Data	
Inlet Temperature	86°F
Outlet Temperature	110°F
Specific Heat Capacity	1.005 Btu/(lb °F)
Density	62.4 lb/ft^3
Price of Cooling Water	$0.10/1000 gal
Electricity Data	
Price of Electricity	$0.0778/kWh

Products

Costing information for $MgCl_2$, KCl and water in the San Diego area was retrieved from online sources. Medical grade $MgCl_2$ and KCl sell for $800 and $380 per metric ton respectively, while agricultural grade for the same species sells for $180 and $450 per metric ton. Finally, water from the San Diego desalination plant sells at a rate of $0.78 per 1000 lb, as a result of an agreement between the plant and the state government.

Process Flow and Material Balances

Figures 10 and 11 present both of the process flow diagrams generated during this project. Once again, we would like to stress that the premise of this design project was to create a simulation to emulate GE's brine concentrator to the best of our ability (with the available software, Aspen Plus, as shown in Figure 10), and to subsequently design a new system to compete with GE's system in a more energy and cost-efficient manner. The competitive system we have designed (with guidance from the patent available in Appendix C), which we refer to as the $MgCl_2$ Separation Unit, is shown in Figure 11. In this section of the report, we are attempting to juxtapose the two systems beside each other as clearly as possible to highlight both their similarities and differences; that said, more specific descriptions of either design can be found in the following section titled 'Process Descriptions.'

Furthermore, both figures contain all of the participating streams in their respective designs. For legibility purposes, the $MgCl_2$ separator flow sheet (Figure 11) is presented on two pages. Red streams in each figure represent heat (steam) streams, while blue streams represent cooling (water) streams.

Tables 7 and 8 show the mass flows, stream conditions, and significant physical properties for the GE system and the $MgCl_2$ separation unit, respectively. Once more, the $MgCl_2$ stream conditions table extends to two pages as well, in an attempt to make the contents more understandable and aesthetically appealing.

We would like to note that although Aspen Plus was initially used to create a simulation flowsheet for the proposed $MgCl_2$ separation unit (see Appendix B), we found that Aspen Plus lacked the extensive electrolyte chemistry background needed to converge the desired process.

With that, we manually created the material balances on a stream-by-stream basis in Microsoft Excel, using extensive nested conditional loops (see Appendix D) in addition to the solubility data gathered above. This section includes the aforementioned manually generated mass balances in Table 8, while Appendix D has a more complete expansion of the Excel file.

Figure 10. Process flow diagram for the GE brine concentrator. Red arrows symbolize steam heat streams. Blue streams symbolize cooling water streams

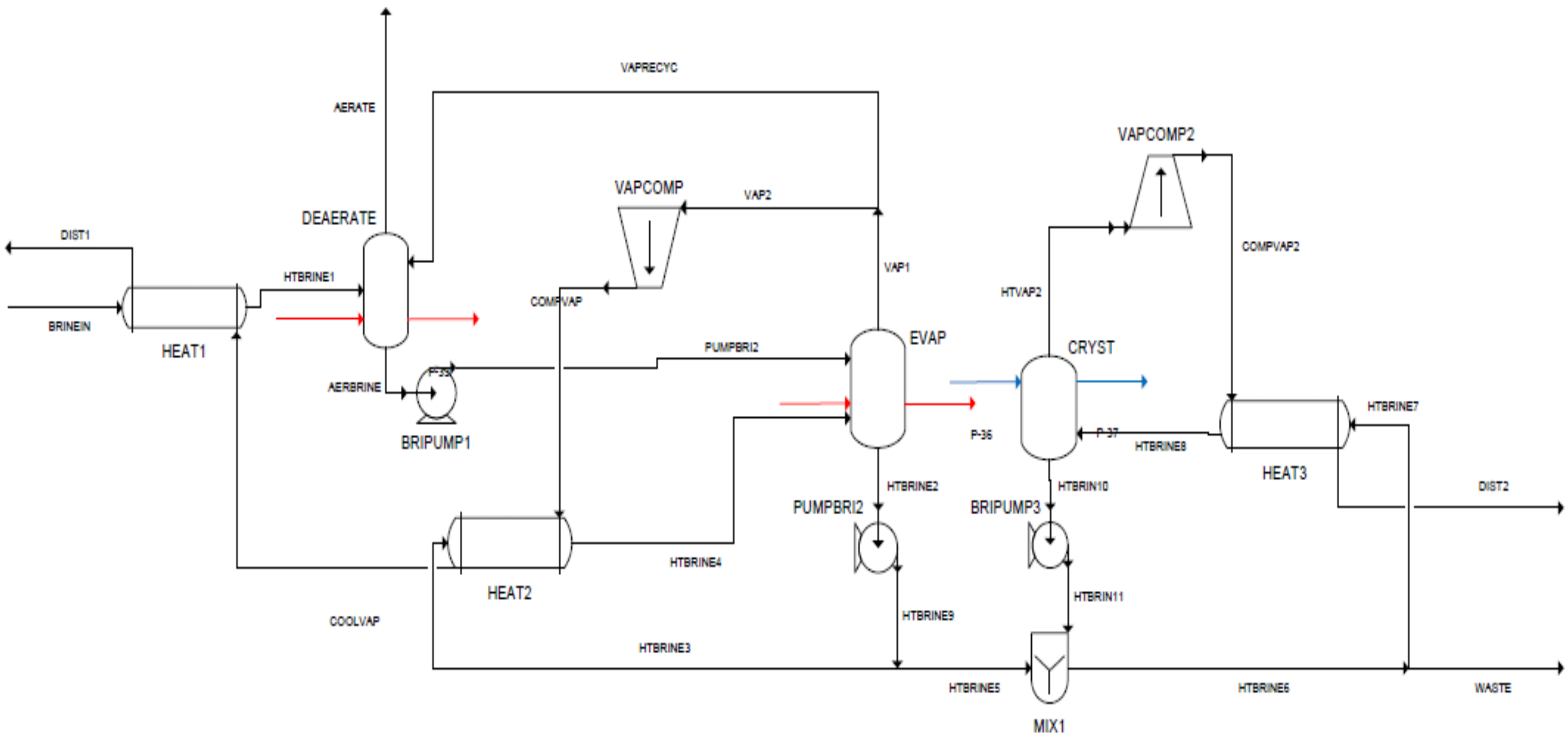
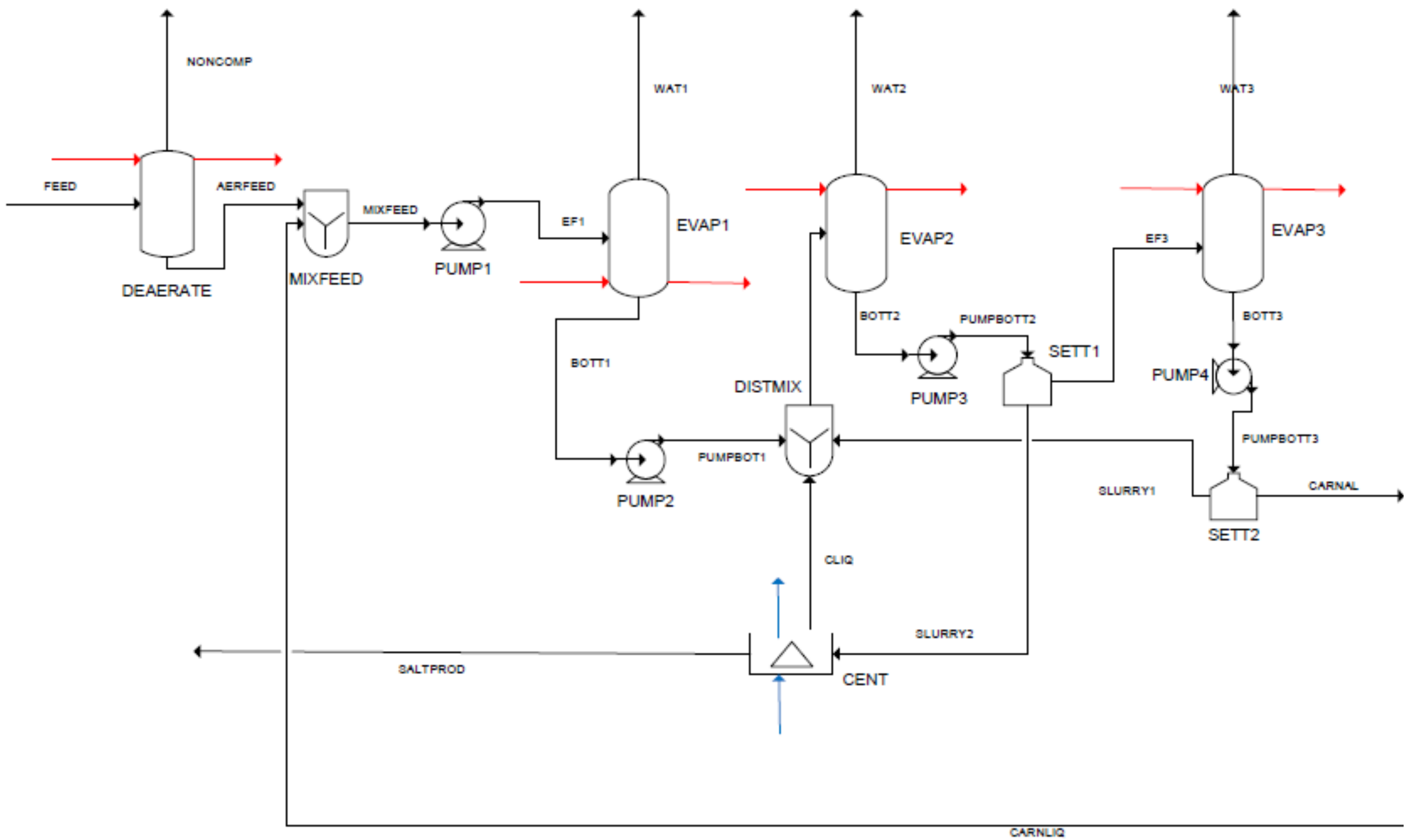


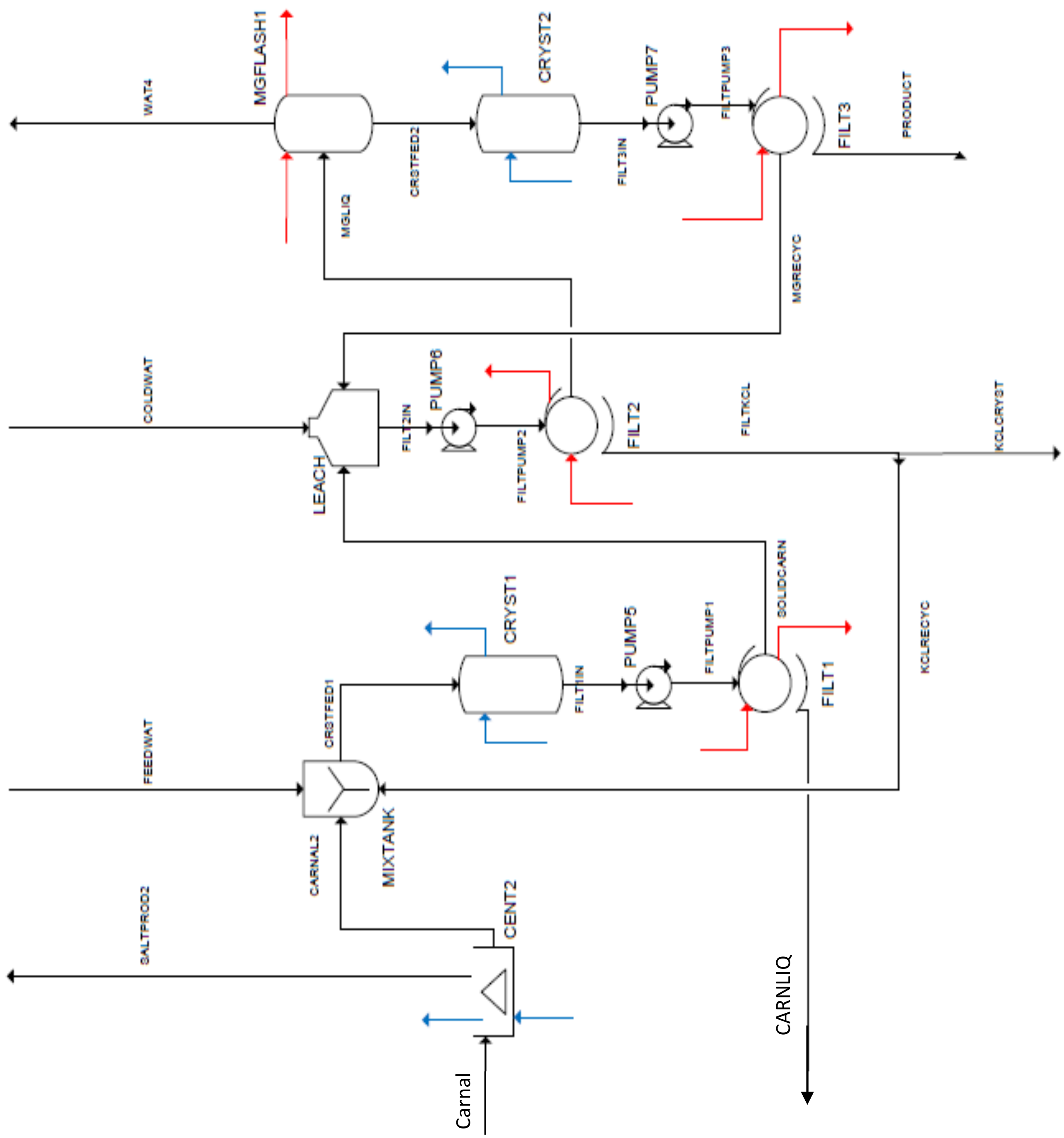
Table 7. GE Brine Concentrator Stream Report

	AERATE	AERBRINE	BRINEIN	COMPVAP	COMPVAP2	COOLVAP	DIST1	DIST2	HTBRIN10	HTBRINE1	HTBRINE2
Stream ID:											
Temperature (C)	86	86	25	395.7	181.2	133.6	95.8	99.6	92.3	94	89.6
Pressure (bar)	0.5	0.5	2	4	1	3	2	1	0.5	0.755	0.5
Vapor Frac	1	0	0.001	1	1	0.4	0	0.05	0	0.01	0
Solid Frac	0	0	0	0	0	0	0	0	0.171	0	0.004
Mass Flow (kg/hr)											
H2O	23150	181152	197870	57887	63207	57887	57887	63207	151299	197870	467333
NA+	0	14602	14602	0	0	0	0	0	8506	14602	55221
NACL(S)	0	0	0	0	0	0	0	0	126836	0	8097
CL-	0	30362	30362	0	0	0	0	0	44494	30362	116536
CO2	232	0	232	0	0	0	0	0	0	232	0
O2	213	0	213	0	0	0	0	0	0	213	0
MG++	0	2326	2326	0	0	0	0	0	9304	2326	9304
MGCL2(S)	0	0	0	0	0	0	0	0	0	0	0
K+	0	1167	1167	0	0	0	0	0	4666	1167	4666
KCL(S)	0	0	0	0	0	0	0	0	0	0	0
Total Mass Flow	23595	229608	246771	57887	63207	57887	57887	63207	345105	246771	661156

	HTBRINE3	HTBRINE4	HTBRINE5	HTBRINE6	HTBRINE7	HTBRINE8	HTVAP2	PUMPBRI2	PUMPBRIN	VAP1	VAP2	VAPREC	WASTE	WASTE
Stream ID:														89.9
Temperature (C)	89.6	145.3	89.6	90.9	90.9	111.7	92.3	86	89.6	89.6	89.6	89.6	90.9	4
Pressure (bar)	0.5	3	0.5	0.5	0.5	1	0.5	1.5	4	0.5	0.5	0.5	0.5	0
Vapor Frac	0	0.024	0	0.001	0.001	0.193	1	0	0	1	1	1	0.001	0.1
Solid Frac	0.004	0.001	0.004	0.108	0.108	0.13	0	0	0.004	0	0	0	0.108	79478.065
Mass Flow (kg/hr)														
H2O	350499	350499	116833	268132	214506	214506	63207	181152	350499	64319	57887	6432	53626	10789
NA+	41416	43241	13805	21586	17269	9889	0	14602	41413	0	0	0	4317	0
NACL(S)	6072	1433	2024	130706	104565	123323	0	0	6080	0	0	0	26141	0
CL-	87402	90216	29134	72508	58006	46625	0	30362	87397	0	0	0	14502	1194
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MG++	6978	6978	2326	11630	9304	9304	0	2326	6978	0	0	0	2326	0
MGCL2(S)	0	0	0	0	0	0	0	0	0	0	0	0	0	0
K+	3500	3500	1167	5833	4666	4666	0	1167	3500	0	0	0	1167	0
KCL(S)	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Mass Flow	495867	495867	165289	510394	408315	408312	63207	229608	495867	64319	57887	6432	102079	43461

Figure 11. Process Flow Sheet Diagram of the $MgCl_2$ Separation Unit. Red arrows symbolize heat (steam) streams and blue arrows symbolize cooling (water) streams





WAT4
MGFLASH1
CRYST2
PUMP7
FILTPUMP3
FILT3
PRODUCT
CRSTFED2
FILT3IN
MGLIQ
MGRECYC
COLDWAT
PUMP6
FILT2
FILTKCL
KCLCRYST
FILT2IN
FILTPUMP2
LEACH
CRYST1
PUMP5
FILTPUMP1
SOLIDCARN
FILT1
KCLRECYC
FEEDWAT
CRSTFED1
FILT1IN
MIXTANK
CARNAL2
CENT2
SALTPROD2
CARNLIQ
Carnal

Table 8. $MgCl_2$ Separation Unit Stream Condition Report

	FEED	AERFED	NONCOMP	MIXFEED	EF1	BOTT1	PUMPBOTT1	WAT1	EF2	BOTT2
Temp (c)	25	50	50	44	44	60	60	60	63	90
Pressure (bar)	1	0.15	0.15	10.15	2	0.15	2	0.15	1	0.5
Solid Fraction	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	2.1	10.8
Sat. Pt (mol Nacl/mol H2O)	11.08	11.27	11.27	11.22	11.22	11.38	11.38	11.38	11.42	11.84
Sat. Pt. (mol KCl/ molH2O)	8.54	10.35	10.35	9.91	9.91	11.07	11.07	11.07	11.29	13.24
St. Pt. (molMgCl2/molH2O)	10.41	11.16	11.16	10.94	10.94	11.56	11.56	11.56	11.70	13.14
H2O	144000.0	129600.0	14400.0	148043.1	148043.1	118434.5	118434.5	29608.6	199855.7	139899.0
NA+	12204.1	12204.1	0.0	12537.3	12537.3	12537.3	12537.3	0.0	29168.3	21160.2
NACL(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	6541.1	26909.8
CL-	18815.0	18815.0	0.0	18815.0	18815.0	18815.0	18815.0	0.0	43646.1	31285.7
CO2	168.7	0.0	168.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O2	155.1	0.0	155.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MG++	1944.0	1944.0	0.0	2511.3	2511.3	2511.3	2511.3	0.0	18602.2	18602.2
MGCL2(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
K+	975.0	975.0	0.0	1529.4	1529.4	1529.4	1529.4	0.0	11328.8	11328.8
KCL(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
KMgCl3*6H2O	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total	178261.8	163538.1	14723.8	183436.1	183436.1	153827.6	153827.6	29608.5	309142.4	249185.6

	PUMPBOTT2	WAT2	EF3*	BOTT3	PUMPBOTT3	WAT3	SLURRY1	SLURRY2	CLIQ***	SALTPROD	CARNAL
Temp (c)	90	90	90	115	115	115	115	90	25	25	100
Pressure (bar)	2	0.5	1.5	1.15	2	1.15	1.5	1.5	1	1	1.5
Solid Fraction	10.8	0.0	10.8	24.3	24.3	0.0	24.3	10.8	0.0	56.5	24.3
Sat. Pt (mol Nacl/mol H2O)	11.84	11.84	11.84	12.35	12.35	12.35	12.35	11.84	11.08	11.08	12.03
Sat. Pt. (mol KCl/ molH2O)	13.24	13.24	13.24	15.05	15.05	15.05	15.05	13.24	8.54	8.54	13.97
St. Pt. (molMgCl2/molH2O)	13.14	13.14	13.14	14.86	14.86	14.86	14.86	13.14	10.41	10.41	13.78
H2O	139899.0	59956.7	97929.3	58757.6	58757.6	39171.7	49943.9	41969.7	31477.3	10492.4	8813.6
NA+	21160.2	0.0	14812.1	9273.5	9273.5	0.0	7882.5	6348.0	317.4	1486.1	1391.0
NACL(S)	26909.8	0.0	18836.8	32924.0	32924.0	0.0	27985.4	8072.9	0.0	19228.2	4938.6
CL-	31285.7	0.0	21900.0	13351.3	13351.3	0.0	11348.6	9385.7	469.3	1902.0	2002.7
CO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MG++	18602.2	0.0	13021.6	13021.6	13021.6	0.0	11068.3	5580.7	5022.6	558.1	1953.2
MGCL2(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
K+	11328.8	0.0	7930.2	7930.2	7930.2	0.0	6740.7	3398.7	3058.8	339.9	1189.5
KCL(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
KMgCl3*6H2O	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total	249185.6	59956.8	174429.9	135258.2	135258.2	39171.8	114969.5	74755.7	40345.3	34006.7	20288.7

	SALTPRD2	CARNAL2**	FEEDWAT	CRSTFED1	FILT1IN	FILTPUMP1	CARNLIQ	SOLDCARN	COLDWAT	FILT2IN	FILTPUMP2
Temp (c)	25	25	25	35	25	25	25	25	5	15	15
Pressure (bar)	1	1	1	1	0.5	3	2	2	1	0.5	3
Solid Fraction	59.6	30.1	0.0	12.8	0.0	0.0	0.0	14.0	0.0	0.0	0.0
Sat. Pt (mol Nacl/mol H2O)	11.08	11.08	11.08	11.15	11.08	11.08	11.08	11.08	11.02	11.04	11.04
Sat. Pt. (mol KCl/ molH2O)	8.54	8.54	8.54	9.26	8.54	8.54	8.54	8.54	7.09	7.81	7.81
St. Pt. (molMgCl2/molH2O)	10.41	10.41	10.41	10.66	10.41	10.41	10.41	10.41	10.08	10.21	10.21
H2O	2644.1	6169.5	15712.8	22769.3	22233.7	22233.7	18443.1	3790.6	15757.9	93358.5	93358.5
NA+	374.5	333.3	0.0	376.6	376.6	376.6	333.3	43.3	0.0	43.3	43.3
NACL(S)	6676.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CL-	1342.9	500.0	0.0	528.1	0.0	0.0	0.0	133.7	0.0	694.7	694.7
CO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MG++	97.7	867.0	0.0	867.0	746.4	746.4	567.3	179.1	0.0	315.4	315.4
MGCL2(S)	0.0	3877.2	0.0	3877.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
K+	59.5	1130.1	0.0	1980.0	1786.6	1786.6	554.4	701.3	0.0	894.7	894.7
KCL(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1014.2	0.0	0.0	0.0
KMgCl3*6H2O	0.0	0.0	0.0	0.0	1376.2	1376.2	0.0	1376.2	0.0	0.0	0.0
Total	11195.1	12877.0	15712.8	30398.1	26519.4	26519.4	19898.1	7238.3	15757.9	95306.6	95306.6

	FILTKCL	KCLRECYC	KCLCRYST	MGLIQ	CRSTFED2	WAT4	FILT3IN	FILTPUMP3	MGREYC*****	PRODUCT
Temp (c)	25	25	25	25	110	110	25	25	25	25
Pressure (bar)	2	2	2	2	1.023	1.023	0.5	3	2	2
Solid Fraction	53.7	53.9	41.4	0.0	0.0	0.0	0.0	0.0	0.0	50.9
Sat. Pt (mol Nacl/mol H2O)	11.08	11.08	11.08	11.08	12.24	12.24	11.08	11.08	11.08	11.08
Sat. Pt. (mol KCl/ molH2O)	8.54	8.54	8.54	8.54	14.69	14.69	8.54	8.54	8.54	8.54
St. Pt. (molMgCl2/molH2O)	10.41	10.41	10.41	10.41	14.49	14.49	10.41	10.41	10.41	10.41
H2O	933.6	886.9	46.7	92424.9	73939.9	18485.0	73939.9	73939.9	73274.5	665.5
NA+	43.3	43.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NACL(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CL-	37.5	28.2	42.2	657.2	657.2	0.0	657.2	657.2	32.9	22.0
CO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MG++	0.0	0.0	0.0	315.4	315.4	0.0	315.4	315.4	15.8	93.5
MGCL2(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	808.5
K+	172.7	164.1	8.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0
KCL(S)	1379.1	1310.1	69.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
KMgCl3*6H2O	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total	2566.2	2432.6	166.5	93397.5	74912.5	18485.0	74912.5	74912.5	73323.1	1589.4

Process Description

GE Brine Concentrator

GE's brine concentrator was modeled using ASPEN Plus simulation, as shown in Figure 10. As an overview, saturated brine (23.3% NaCl by weight) enters the process (desalination facilities typically dispense saturated NaCl solutions as waste), and undergoes a series of heat and separation events; throughout the process, two pure water streams (DIST1 and DIST2) are obtained, successfully recovering approximately 68% of the water fed to the system. Furthermore, a concentrated brine stream exits the system (WASTE), which will not only contain a saturated aqueous NaCl solution, but also solid NaCl crystals and other contaminants.

More specifically, saturated brine (which contains NaCl alongside a variety of other contaminants like $MgCl_2$, KCl, $MgSO_4$, K_2SO_4 etc.) enters the process at approximately 25°C and 1 bar and is immediately passed through a shell-in-tube heat exchanger, which raises the stream's temperature to near boiling conditions (108°C at 1 bar). Although a shell-in-tube exchanger is used in practice, the exchanger was modeled in ASPEN using two HEATER blocks and a heat stream connecting the two, this allowed for the simulation to converge more easily and efficiently. In the process flowsheet diagram shown in Figure 11, however, the more realistic shell-in-tube exchanger is used. The heated brine is then passed through a deaerator, which was modeled with a FLASH2 block and removes any non-compressible fluids from the brine in the AERATE exit stream, such as O_2 and CO_2 gas, as well as a small amount of water in the vapor phase. The deaerated brine (AERBRINE) is directed to the brine evaporator, which is a section of the brine concentrator containing two distinct units: a settler tank and a shell-in-tube heat exchanger. In an attempt to mimic the brine evaporator, a FLASH2 block and an FSPLIT block were used to model the settler tank, and two HEATER blocks alongside a heat stream were used to model the shell-in-tube heat exchanger. The deaerated brine initially enters the flash drum 'EVAP,' which evaporates a portion of the water from the brine, concentrating the brine further; at this point, NaCl crystals will begin forming in the aqueous brine solution as a result of the increased NaCl concentration. The concentrated brine exits as the bottoms product of the flash drum and enters a splitter (SPLIT), which sends majority of the brine through a pump and subsequently into the evaporator's heat exchanger. Finally, the concentrated brine is heated to a temperature above its boiling point (approximately 130°C), and is then recycled into the original flash drum EVAP.

The water vapor distillate from the EVAP flash drum is passed through a compressor, modeled with a COMPR block in ASPEN, increasing its pressure to approximately 4 bar. The compressed water vapor is introduced into the shell-in-tube exchanger as the heating fluid, and is thus used to heat the aforementioned concentrated brine. As a result, the compressed vapor is significantly cooled and mostly condensed to the liquid phase. Finally, the cooled two-phase water stream is passed through the initial shell-in-tube heat exchanger used to heat the feed brine; the water stream is fully condensed and cooled further, exiting the system as the first recovered water stream, DIST1.

As mentioned earlier, the splitter block SPLIT recycles the majority of the concentrated brine from the EVAP flash drum. The portion that is not recycled, however, is redirected to a crystallizer. Rather than utilizing one of ASPEN's crystallizer blocks, which has limited functionality and would be rendered inaccurate in this process, a custom crystallizer was constructed from a variety of blocks. First, the concentrated brine is mixed with a more concentrated brine solution already present in the crystallizer – this was modeled using a MIXER block. The mixed brine stream is passed to a subsequent splitter (SPLIT2), which redirects the majority of the stream to the crystallizer block and ejects the remainder from the process as the waste stream (WASTE). The recycled brine enters a shell-in-tube heat exchanger (once again modeled with two HEATER blocks and a heat stream), increasing its temperature to near boiling; the heated brine enters a FLASH2 drum (CRYST), which vaporizes a portion of the water from the brine stream. The water vapor is compressed to approximately 1.5 bar and is subsequently passed through the aforementioned shell-in-tube exchanger and is used to heat the inlet concentrated brine. After passing through the exchanger, the water vapor both cools and condenses, and exits the system as the second recovered water stream, DIST2.

The CRYST flash drum's bottoms product is a densely concentrated brine solution with a substantial solid fraction (0.30-0.35) as a result of evaporating the flash drum's inlet water content; by removing water from the brine, the CRYST drum is effectively forcing NaCl crystals to form in the extremely concentrated bottoms product. Finally, the bottoms product is recycled to the aforementioned mixer tank and is mixed with the concentrated brine solution from the brine evaporator.

$MgCl_2$ Separation Unit

Figure 11 shows the process flowsheet diagram for the designed $MgCl_2$ separation unit. This separation process occurs in three distinct phases: concentration and impurity removal, KCl recovery, and $MgCl_2$ recovery. As shown, phase one utilizes multi-stage flash evaporation to concentrate the incoming salts by removing majority of the feed water, while phases two and three predominantly use crystallizing and filtering units to remove KCl and $MgCl_2$ crystals as recoverable byproducts.

In phase one, saturated brine wastewater from a desalination facility enters the process at the same conditions as that of the GE Brine Concentrator feed stream. Furthermore, in an attempt to be as consistent as possible, the species compositions of the feed stream to the $MgCl_2$ separator are identical to that of the GE Brine Concentrator, as is shown in Tables 6 and 7. Upon entering the system, the feed brine enters a deaerator, which, in a similar fashion to the GE system, removes any non-compressible fluids from the brine, such as O_2 and CO_2. The deaerated brine is fed into a mixing tank, which slightly lowers its temperature from 50°C to 44°C before it is pumped into the first of three flash evaporators, named EVAP1, which evaporates approximately 20% of the incoming water to stream WAT1. The bottoms product (stream BOTT1), which is now more concentrated and higher in temperature as a result of the flash evaporation, is pumped to a second mixing tank; BOTT1 experiences a slight temperature increase and is mixed with two more concentrated streams SLURRY1 and CLIQ, at which point it enters the second flash evaporator, EVAP2. EVAP2 vaporizes 30% of the incoming water, which is recovered as stream WAT2. The bottoms product from EVAP2 (BOTT2), which is more concentrated and higher in temperature than the feed to the evaporator, is pumped to a settler tank, which sends a stream with high salt concentrations to a centrifuge and a lesser concentrated stream to the third and final flash evaporator. The centrifuge cools the incoming feed to 25°C and performs a liquid solid separation; that is, the majority of the NaCl will crystallize and is removed from the system in the SALTPROD stream in addition to a fraction of the incoming water, $MgCl_2$ and KCl. Conversely, a majority of the water, KCl and $MgCl_2$ exits the centrifuge in the CLIQ stream, which is then recycled to the aforementioned mixer tank DISTMIX that provides the feed stream to the second flash evaporator, EVAP2.

As mentioned earlier, the less concentrated stream leaving the settler tank will be fed to the third and final flash evaporator EVAP3. Here, 40% of the incoming water is vaporized and recovered, and the bottoms product (BOTT3) is pumped to a second settler tank, SETT2. Like the case of the first settler tank, a concentrated stream of salts is recycled to DISTMIX and is re-fed to the second flash evaporator EVAP2. Likewise, a lesser concentrated stream exits SETT2 and is fed to a second centrifuge, CENT2, which removes a majority of the persisting NaCl in stream SALTPROD2, in addition to a small fraction of the water, KCl and $MgCl_2$. As a result, the stream CARNAL2 exits the centrifuge with negligible amounts of NaCl present in solution in addition to majority of the $MgCl_2$ and KCl from the feed stream; this officially ceases phase one of the process.

Phase two focuses on crystallizing KCl. First, the CARNAL2 stream exiting phase one is introduced to a mixing tank, in which it is mixed with a feed water stream (25°C and 1 bar) and a recycled KCl crystalline stream; this mixture serves the purpose of re-suspending any $MgCl_2$ crystals that have formed in the CARNAL2 stream back into aqueous solution. Subsequently, the mixed stream CRSTFED1 exits the mixer and enters the first of two crystallizers, CRYST1, which operates at 25°C and 0.5 bar; under these conditions, carnallite crystals ($KMgCl_3 \bullet 6H_2O$) will form, which is vital for the process to proceed. The carnallite-crystalline stream is pumped through a rotary filter, which sends the solid carnallite crystals and a small fraction of the incoming water to a leach tank, while the remaining aqueous species and a large majority of the water is recycled to phase one and is fed to the original mixing tank with the deaerated feed brine. As a result, a vast majority of the K^+ and Mg^{2+} ions are kept in phase two via the carnallite crystals, while little Na^+ enter phase two.

Cold water (5°C, 1 bar) is fed to the aforementioned leach tank and is used to suspend the carnallite crystals in solution; this re-introduces K^+, Mg^{2+} and Cl^- ions into solution, which, under these operating conditions, will allow KCl crystals to begin precipitating. The mixed solution is then pumped through the second of three rotary filters, which removes the solid KCl crystals in addition to a small amount of water in the stream FILTKCL, while the remaining water and ionic species exit in the MGLIQ stream. FILTKCL is partially removed from the process at this point, successfully recovering $KCl_{(s)}$ crystals; the majority of the FILTKCL stream is redirected in the KCLRECYC stream and is recycled to the first mixing tank in phase two, meeting the CARNAL2 feed stream and feed water. With the completion of this recycle loop, phase two comes to a close.

Phase three, being the final phase of the process, focuses on forming and recovering $MgCl_2$ crystals from solution. The stream MGLIQ is first fed into a flash evaporator, which removes majority (75%) of the water content of the feed. Its bottoms product (CRSTFED2) is introduced to the process' second crystallizer, which operates at 25°C and 1 bar; under these conditions, $MgCl_2$ crystals will precipitate, where Mg^{2+} are notably the limiting reagent under the initial feed stream species compositions. Finally, the $MgCl_2$ crystalline stream FILT3IN is pumped through the third and final rotary filter, which recycles the majority of the water and its relatively negligible ionic composition to the leach tank LEACH, and subsequently recovers the $MgCl_2$ crystals (96% wet-mass purity) in the PRODUCT stream. With this final filtration, the process terminates; overall, four distinct streams recover water from the feed, KCl and $MgCl_2$ crystals are formed in significant amounts (approximately 21tons $MgCl_2$ are formed per day with these flow rates), and the waste streams SALTPROD and SALTPROD2 exit the system at a convenient 31.0% and 35.1% solid, respectively, and collectively contain 100% of the NaCl that enters the system.

Utilities

The three main utilities for both our concentrator and the GE concentrator are electricity, low pressure steam, and cooling water. The main utility for our process was low pressure steam for heating duties, whereas the GE brine concentrator relies most almost exclusively on electricity for utility requirements. For the GE brine concentrator, steam and cooling water costs account for 1.9% and 0.02% respectively, while roughly 98.1% of the cost is due to electricity use. For the Magnesium Chloride Separation Unit, low pressure steam accounts for 91.4% of the overall utilities cost, electricity for 8.2%, and cooling water for 0.4%. Prices for utilities are $0.0778/kWh, $6/1000lb, $0.1/1000 gal for electricity, low pressure steam, and cooling water, respectively [29] [30].

The GE system requires electricity to power a pump for the brine and two vapor compressors. The majority of the total electricity used by the GE system is allocated towards powering the two vapor compressors, VAPCOMP1 accounts for 79% of the annual electricity usage, and VAPCOMP2 is responsible for 21%. In our process, no compressors were used; streams are kept in the liquid phase allowing pumps to be used instead of compressors. This process uses seven pumps, which range in percentage of the total electricity use from 0.8-4.9%. The majority of energy consumed in this process is due to the two centrifuges. ASPEN was unable to give electricity requirements for centrifuge blocks, so a conservative estimate was used to account for electricity use of the centrifuges in our system. For industrial level centrifuges, 90 kW was the maximum electricity requirement for a standard centrifuge, thus 90 kW per centrifuge was used to account for the worst case energy demand of our process [31]. Using this conservative value, the centrifuges are responsible for 79.6% of the total energy usage in the Magnesium Chloride Separation Unit. Note that 90 kW per centrifuge is still significantly less than the 2300 kW and 8700 kW required to power the two vapor compressors in GE's system. The GE system is estimated to use $8.7x10^7$ kWh of electricity per year, while our $MgCl_2$ Separation Unit system uses only $1.7x10^6$ kWh per year. Compressors are very costly when it comes to applying a pressure change, because it is less energy intensive to pump a liquid than to compress a gas. The absence of compressors in our brine concentrator is why our system is more electricity efficient by roughly an order of magnitude.

While our brine concentrator uses less energy than the GE model, we use more cooling water and steam. The GE system requires low pressure steam to provide heat duties for a deaerator and evaporator. The deaerator uses 40% of the total mass of steam while the evaporator uses 60%. The total annual use for the GE system is approximately $2.2x10^7$ lb of steam per year. Most of our annual steam use goes towards heating the flash evaporation units. EVAP1, EVAP2, EVAP3 and MGFLASH1 account for 96% of the total steam usage, with EVAP1 using 18%, EVAP2 38%, EVAP3 25%, and MGFLASH1 15%. The deaerator accounts for the remainder of the required heat duty for the process. Our process does not have a single piece of equipment responsible for the bulk of the energy use as a result of the use of a multi-effect evaporation system. The total annual steam use is $2.6x10^8$ lbs of steam per year, roughly a factor of ten greater than the steam usage for the GE system.

The GE concentrator only uses cooling water in its crystallizer, which requires $1.2x10^7$ gallons of cooling water annually. Our system uses two crystallizers, with the distribution of the cooling water as follows: 3% for CRYST1 and 97% for CRYST2. The second crystallizer requires the most cooling. The total annual use of cooling water is $6.4x10^7$ gallons of cooling water per year, slightly greater than the GE system but on the same order of magnitude. Despite using more steam and cooling utilities in our process, the overall cost of utilities is still less for our system. Because both processes are electricity intensive, being able to decrease our electricity use allows our system to compensate for relatively higher heating and cooling utilities leading to an overall less expensive utility bill by an estimated factor of 4. The GE process requires roughly \$6.9 million in utilities annually, while our process requires \$1.6 million. Additionally, note that because our system recovers 2.2×10^9 lb of water compared to GE's 2.1×10^9 lb annually, our required ratios for steam and cooling water are greater by about an order of magnitude, but our required ratio for electricity is less by about two orders of magnitude. See Tables 9-12 below for all required ratios, totals, and costs on utilities.

Table 9. Utilities Summary.

GE			
Utility	Annual Amount	Required Ratio	Annual Cost ($)
Low Pressure Steam	lb	0.010 lb/lb	132808.42
Cooling Water	gal	0.045 gal/lb	1144.84
Electricity	kWh	0.041 kWh/lb	6791327.65
Recovered Water (lbs)	2,109,941,856		
$MgCl_2$ Separation Unit			
Utility	Annual Amount	Required Ratio	Annual Cost ($)
Low Pressure Steam	lb	0.12 lb/lb	1555472.75
Cooling Water	gal	0.26 gal/lb	6829.21
Electricity	kWh	0.00081 kWh/lb	139335.88
Recovered Water (lbs)	2,203,483,611		

Table 10. Steam Duty Requirements

GE Brine Concentrator			
Block	Duty (gcal/hr)	Annual Use (lbs)	Annual Cost ($)
DEAERATE	3.9791	8742501.553	52455.01
EVAP	6.0954	13392235.42	80353.41
$MgCl_2$ Separation Unit			
Block	Duty (gcal/hr)	Annual Use (lb)	Annual Cost ($)
EVAP1	21.47	47171850	283031.10
EVAP2	44.88	98606084.21	591636.51
EVAP3	28.94	63584226.32	381505.36
FILT2	0.923	2027928.158	12167.57
MGFLASH1	17.80	39112867.89	234677.21
DEAERATE	3.9791	8742501.553	52455.01

Table 11. Cooling Water Duty Requirements

GE Brine Concentrator			
Block	Duty (gcal/hr)	Annual Use (gal)	Annual Cost ($)
CRYST	-1.15	11448380.975	1144.84
$MgCl_2$ Separation Unit			
Block	Duty (gcal/hr)	Annual Use (gal)	Annual Cost ($)
CRYST1	-0.20	2010933.006	201.09
CRYST2	-6.26	62319012.95	6231.90

Table 12. Electricity Requirements

GE Brine Concentrator			
Block	Work (kW)	Annual Use (kWh/yr)	Annual Cost ($)
BRIPUMP	8.66348	68614.8	5338.23
VAPCOMP	8676	68713920	5345942.98
VAPCOMP2	2337.07	18509594.4	1440046.44
$MgCl_2$ Separation Unit			
Block	Work (kW)	Annual Use (kWh)	Annual Cost ($)
PUMP1	11.15	88308.0	6870.36
PUMP2	9.33	73893.6	5748.92
PUMP3	10.43	82605.6	6426.72
PUMP4	3.223	25526.2	1985.94
PUMP5	1.751	13867.9	1078.92
PUMP6	5.637	44645.0	3473.38
PUMP7	4.609	36503.3	2839.96
CENT	90	55455.8	712800.00
CENT2	90	55455.8	712800.00

Unit Descriptions

GE Brine Concentrator

Heat Exchangers

HEAT1 is a shell and tube heat exchanger with tubes constructed from monel, to avoid corrosion due to the brine, and the rest made from carbon steel. The heat exchanger is 10ft long with 47 tubes and raises the temperature of the cold stream, which has a flow rate of 247,000 kg/hr, from 25°C to 94°C while the hot stream, with a flow rate of 58,000 kg/hr, is cooled from 134°C to 96°C. This heat exchanger is designed to heat the incoming brine to a temperature just under boiling in preparation for the deaerator while also cooling the effluent steam to the liquid state to pump out of the plant as a product.

HEAT2 is another shell and tube heat exchanger with tubes constructed from monel and the rest from carbon steel. This exchanger is 20 ft long with 398 tubes and increases the temperature of the cold stream, which has a flow rate of 496,000 kg/hr, from 90°C to 146°C while the hot stream, with a flow rate of 58,000 kg/hr, condenses 60% of the stream and cools from 396°C 134°C. HEAT2 heats the brine before it enters the brine evaporator in order to evaporate and recover the water from the brine stream.

Deaerator

The deaerator is constructed from stainless steel and is 33 feet long and 17 feet in diameter. This vessel is used to remove incompressible gases such as CO_2 and O_2 from the brine feed stream. The column operates at 0 bar and removes all but trace amounts of both CO_2 and O_2 from the brine stream flowing at 247,000 kg/hr. The deaerator stream removes all the gases along with 23,000 kg/hr of water, at a total flow rate of 24,000 kg/hr of vapor. The resulting brine is free of air and flows out at a rate of 230,000 kg/hr.

Brine Evaporator

The brine evaporator is made with stainless steel and is 26 feet long and 13 feet in diameter. This evaporator is used to recover water from the brine stream in order to increase the overall amount of product. Brine enters the evaporator from the deaerator at 86°C with a flow rate of 230,000 kg/hr and there is a recycle stream of brine that enters from HEAT2 at 145°C with a flow rate of 496,000 kg/hr. The evaporator operates at 0.5 bar and 5% of the inlet streams leaves as vapor, with a composition of pure water and a flow rate of 64,000 kg/hr at 90°C. The remaining brine leaves the evaporator at a flow rate of 661,000 kg/hr at 90°C, with 0.4% of the stream as solid NaCl.

Crystallizer

The crystallizer is modelled in ASPEN using an evaporator, a compressor, and a heat exchanger to modify each component of the crystallizer separately in order to achieve optimal separation as opposed to using the traditional crystallizer block available in ASPEN. The evaporator component of the crystallizer is made from stainless steel and is 30 feet long and 60 feet in diameter. It operates at 92°C and 0.5 bar in order to evaporate a portion of the water and concentrate the brine stream, increasing the formation of NaCl crystals. The inlet stream enters at 112°C with a flow rate of 408,000 kg/hr. The effluent pure water stream leaves at 92°C with a flow rate of 63,000 kg/hr. This water stream enters a compressor and is condensed before entering HEAT3. The remaining brine from the crystallizer evaporator, at a temperature of 92°C and flow rate of 345,000 kg/hr with 11% solids in the form of NaCl crystals, is mixed with brine from the brine evaporator in a mixing tank that is 10 feet long and 5 feet in diameter, to form a stream at 91°C with a flow rate of 510,000 kg/hr, which now has a solids content of 11%. Some of this stream is collected as the waste stream while 80% of the stream enters HEAT3.

HEAT3 is a shell and tube heat exchanger with tubes constructed from monel, and the rest made of carbon steel. The heat exchanger is 12 feet long with 2,000 tubes and increases the temperature of the cold stream with a flow rate of 408,000 kg/hr from 91°C to 112°C and cools the hot stream with a flow rate of 63,000 kg/hr from 181°C to 100 °C. This exchanger heats the brine before entering the evaporator component of the crystallization and condenses 95% of the effluent water stream to collect and pump the product.

$MgCl_2$ Separation Unit

Deaerator

The deaerator is 22 feet long and 11 feet in diameter, is made of stainless steel, and is used to remove incompressible gases like CO_2 and O_2, similar to the deaerator in the GE Brine Concentrator system. The deaerator operates at 0.15 bar and 50°C and removes all but trace amounts of CO_2 and O_2 from the inlet stream resulting in a vapor stream comprised of CO_2, O_2 and 14,000 kg/hr of H_2O at 50°C with a flow rate of 15,000 kg/hr. The remaining brine leaves the deaerator at 50°C with a flow rate of 164,000 kg/hr.

Triple Effect Evaporator

The deaerated brine enters a triple effect evaporator made from stainless steel to remove water from the brine stream. The first evaporator is 22 feet long and 11 feet in diameter and operates at 0.15 bar and 60°C to remove 30,000 kg/hr of water. The remaining liquid is pumped to 1 bar to increase the pressure of the stream before entering the second effect and mixed with the recycle streams from the second and third effect to increase the overall efficiency of the system and introduces solid NaCl into the system. The resulting mixed stream, at 63°C with a flow rate of 309,000 kg/hr with a solids fraction of 0.02, enters the second effect which is 28 feet long and 14 feet in diameter, and operates at 90°C and 0.5 bar, removing 60,000 kg/hr of water. The remaining liquid is pumped, increasing the pressure to 2 bar, and enters a settling tank, which is 22 feet long and 11 feet in diameter, where 30% is removed to a centrifuge, CENT, and the remaining 70% enters the third effect. The third effect, which is also 22 feet long and 11 feet in diameter, operates at 115°C and 1.15 bar, which removes 39,000 kg/hr of water. The remaining brine is pumped to a stainless steel second settling tank, which is 26 feet long and 13 feet in diameter, and 85% is recycled and the remaining 15%, operating at 100°C with a flow rate of 20,000 kg/hr and has a solids fraction of 0.24 which is comprised solely of NaCl crystals. The effluent water streams from the triple effect evaporator are collected as product.

<u>Centrifuges</u>

CENT operates at 25°C and 1 bar. The inlet brine stream is 90°C with a flow rate of 75,000 kg/hr and 10% of the stream is comprised of NaCl crystals. The resulting solids stream has a flow rate of 34,000 kg/hr and is 57% solid NaCl. The liquid stream contains no solid and has a flow rate of 40,000 kg/hr. This centrifuge removes the undesired salts, namely NaCl, from the system.

CENT2 operates at 25°C and 1 bar. The inlet brine stream is 100°C with a flow rate of 20,000 kg/hr and a solids fraction of 0.24. The resulting solids stream has a flow rate of 11,000 kg/hr and 60% of the stream is comprised of NaCl crystals. The liquid outlet stream has a flow rate of 13,000 kg/hr and has a solids fraction of 0.3, which is comprised of only $MgCl_2$ crystals. This second centrifuge is designed to remove the remaining NaCl from the system to increase the final purity of the $MgCl_2$ and KCl crystals.

<u>Crystallizers</u>

CRYST1 operates at 25°C and 0.5 bar. The inlet stream is a mix of the outlet stream from CENT2, the recycle stream from the KCl separation at 25°C with a flow rate of 2,000 kg/hr and 54% KCl, and a feed water stream at 25°C with a flow rate of 16,000 kg/hr. The resulting mixed inlet stream has a temperature of 35°C and a flow rate of 30,000 kg/hr, with 13% solids comprised solely of $MgCl_2$. The resulting stream has a flow rate of 27,000 kg/hr, 1400 kg/hr of which is carnallite, $KMgCl_3*6H_2O$, which is necessary to achieve $MgCl_2$ and KCl separation. This stream is pumped to a filter which creates a liquid stream with a flow rate of 20,000 kg/hr and no solids and a solids stream with 1400 kg/hr of carnallite and 1000 kg/hr of KCl crystals, with a total flow rate of 7,200 kg/hr.

CRYST2 is made of carbon steel and operates at 25°C and 0.5 bar. The inlet stream is the resulting stream from a stainless steel evaporation column that is 17 feet long and 9 feet in diameter, and operates at 110°C and 1.0 bar, evaporating 18,000 kg/hr of water. The resulting stream has a flow rate of 75,000 kg/hr and a temperature of 110°C when it enters the crystallizer. The outlet stream from CRYST2 enters a continuous rotary drum vacuum filter which creates a solid stream with a flow rate of 1600 kg/hr, with a solid content of 51%, or 810 kg/hr of $MgCl_2$ crystals. The liquid stream has a flow rate of 73,000 kg/hr containing mostly water with small amounts of Cl^- and Mg^{2+} ions, which is recycled into the leach tank.

Leach Tank

The leach tank is 8 feet long and 4 feet in diameter, is made of stainless steel, and operates at 25°C and 0.5 bar. The inlet stream is a mix of the solid outlet stream from FILT1, with a flow rate of 7200 kg/hr and a solids fraction of 0.14 due to the presence of KCl crystals, the recycle stream from FILT3, with a flow rate of 73,000 kg/hr, and a cold water stream at 5°C with a flow rate of 16,000 kg/hr. The leach tank outlet stream leaves at 15°C with a flow rate of 95,000 kg/hr. This leach tank dissolves the KCl crystals in water in preparation for the KCl separation that occurs in FILT2, where KCl crystals form in a stream at 25°C with a flow rate of 2600 kg/hr, 54% of which is KCl crystals, 95% of which is recycled into the system to maintain appropriate levels of

KCl in the process, and the remaining 5% is collected as product. The liquid stream from FILT2 has a temperature of 25°C and a flow rate of 93,000 kg/hr and is fed into the evaporator to begin the separation of $MgCl_2$.

Equipment List

GE Brine Concentrator

The following tables contain the unit specifications for each piece of equipment that was used in the GE Brine Concentrator system.

<u>Heat Exchangers</u>

Heat Exchanger		
Identification	Item	*Heat Exchanger*
	Item No.	HEAT1
	No. Required	1
Function	Heats feed stream to near boiling point	
Operation	Continuous	
Type	Shell and Tube	
Stream ID	**Tube Side**	**Shell Side**
Stream In	BRINEIN	COOLVAP
Stream Out	HTBRINE1	DIST1
Flow Rate (kg/hr)	247000	57900
Inlet Temperature (°C)	25	134
Outlet Temperature (°C)	94	96
Design Data:	Surface Area (ft^2)	90
	Tube Length (ft)	10
	LMTD (°C)	128
	Heat Duty (MMBTU/HR)	24
	Construction Material	Carbon Steel
Purchase Cost	$	29, 000
Bare Module Cost	$	93, 000

Heat Exchanger		
Identification	Item	*Heat Exchanger*
	Item No.	HEAT2
	No. Required	1
Function	Heats brine to above boiling point for evaporation	
Operation	Continuous	
Type	Shell and Tube	
Stream ID	**Tube Side**	**Shell Side**
Stream In	HTBRINE3	COMPVAP
Stream Out	HTBRINE4	COOLVAP
Flow Rate (kg/hr)	496000	58000
Inlet Temperature (°C)	90	396
Outlet Temperature (°C)	145	134
Design Data:	Surface Area (ft^2)	1570
	Tube Length (ft)	20
	LMTD (°C)	76
	Heat Duty (MMBTU/HR)	9.0
	Construction Material	Carbon Steel
Purchase Cost	$	39, 000
Bare Module Cost	$	125, 000

Heat Exchanger		
Identification	Item	*Heat Exchanger*
	Item No.	HEAT3
	No. Required	1
Function	Heats recycled brine to crystallizer	
Operation	Continuous	
Type	Shell and Tube	
Stream ID	**Tube Side**	**Shell Side**
Stream In	HTBRINE7	COMPVAP2
Stream Out	HTBRINE8	DIST2
Flow Rate (kg/hr)	408000	63000
Inlet Temperature (°C)	91	181
Outlet Temperature (°C)	112	115
Design Data:	Surface Area (ft^2)	4500
	Tube Length(ft)	12
	LMTD (°C)	8.7
	Heat Duty (MMBTU/HR)	24
	Construction Material	Carbon Steel
Purchase Cost	$	73, 000
Bare Module Cost	$	234, 000

Flash Drums

Flash Drum 1			
Identification	Item		*Deaerator*
	Item No.		DEAERATE
	No. Required		1
Function	Remove non-compressible fluids from feed stream		
Operation	Continuous		
Type	Vertical Drum		
Stream ID	**Feed**	**Distillate**	**Bottoms**
	HTBRINE1	AERATE	AERBRINE
Flow Rate (kg/hr)	246771	24000	230000
Pressure (bar)	0.755	0.5	0.5
Temperature (°C)	94	86	86
Design Data:	Operating Pressure (bar)		0.5
	Vapor Fraction		0.1
	Heat Duty (GCAL/HR)		4.0
	Mass Stream (lb/yr)		8.7E+06
	Construction Material		Stainless Steel
Utility Cost ($/year)		$	52, 000
Purchase Cost		$	152, 000
Bare Module Cost		$	630, 000

Flash Drum 2				
Identification	Item			*Flash Evaporator*
	Item No.			EVAP
	No. Required			1
Function	Remove water from incoming feed stream via flash evaporation			
Operation	Continuous			
Type	Vertical Drum			
Stream ID	**Feed 1**	**Feed 2**	**Distillate**	**Bottoms**
	HTBRINE4	PUMPBRI2	VAP1	HTBRINE2
Flow Rate (kg/hr)	495867	230000	64319	661000
Pressure (bar)	3	4	0.5	0.5
Temperature (°C)	145	86	90	90
Design Data:	Operating Pressure (bar)			0.5
	Vapor Fraction			0.05
	Heat Duty (GCAL/HR)			6.1
	Mass Stream (lb/yr)			1.3E+07
	Construction Material			Stainless Steel
Utility Cost ($/year)			$	80, 000
Purchase Cost			$	111, 000
Bare Module Cost			$	461, 000

Flash Drum 3			
Identification	Item		*Crystallizer*
	Item No.		Crystallizer
	No. Required		1
Function	Concentrate incoming brine stream via NaCl crystallization		
Operation	Continuous		
Type	Vertical Drum		
Stream ID	**Feed**	**Distillate**	**Bottoms**
	HTBRINE8	HTVAP2	HTBRIN10
Flow Rate (kg/hr)	408312	63200	345000
Pressure (bar)	1	0.5	0.5
Temperature (°C)	112	92	92
Design Data:	Operating Pressure (bar)		0.5
	Vapor Fraction		0.15
	Heat Duty (GCAL/HR)		-1.2
	Mass Cooling Water (lb/yr)		9.6E+07
	Construction Material		Stainless Steel
Utility Cost ($/year)		$	1,150
Purchase Cost		$	277, 000
Bare Module Cost		$	1,153, 000

Vapor Compressors

Vapor Compressor 1			
Identification	Item		*Single Stage Compressor*
	Item No.		VAPCOMP
	No. Required		1
Function	Compress water vapor from evaporator to heat exchanger		
Operation	Continuous		
Type	Singe Stage Centrifugal		
Stream ID	**Inlet**		**Outlet**
	VAP2		COMPVAP
Temperature (°C)	90		400
Pressure (bar)	0.5		4
Design Data:	Flow Rate (kg/hr)		57900
	Drive Type		Electric
	Electricity Consumed (kWh/year)		68,700, 000
	Construction Material		Cast Iron
	Brake Effeciency		75%
	Motor Efficiency		75%
Utility Cost ($/year)		$	5,345, 000
Purchase Cost		$	2,112, 000
Bare Module Cost		$	3,172, 000

Vapor Compressor 2		
Identification	Item	*Single Stage Compressor*
	Item No.	VAPCOMP2
	No. Required	1
Function	Compress water vapor from crystallizer to heat exchanger	
Operation	Continuous	
Type	Singe Stage Centrifugal	
Stream ID	**Inlet**	**Outlet**
	HTVAP2	COMPVAP2
Temperature (°C)	92	180
Pressure (bar)	0.5	1
Design Data:	Flow Rate (kg/hr)	63000
	Drive Type	Electric
	Electricity Consumed (kWh/year)	18,510, 000
	Construction Material	Cast Iron
	Brake Efficiency	75%
	Motor Efficiency	75%
Utility Cost ($/year)	$	1,440, 000
Purchase Cost	$	1,977, 000
Bare Module Cost	$	2,965, 000

Pumps

PUMP 1		
Identification	Item	*Brine Pump*
	Item No.	BRIPUMP1
	No. Required	1
Function	Pump deaerated brine to flash evaporator	
Operation	Continuous	
Type	Centrifugal, 3600RPM, VSC 75 Hp	
Stream ID	**Inlet**	**Outlet**
	AERBRINE	PUMPBRI2
Pressure (bar)	0.5	1.5
Design Data:	Flow Rate (kg/hr)	230000
	Pump Head (ft)	28
	Brake Horsepower (Hp)	6
	Electricity Consumed (kWh)	69000
	Construction Material	Cast Iron
	Brake Effeciency	72%
	Motor Efficiency	85%
Utility Costs ($/year)	$	5,300
Purchase Cost	$	7,300
Motor Purchase Cost	$	650
Bare Module Cost	$	24,200

PUMP 2		
Identification	Item	*Brine Pump*
	Item No.	PUMPBRI2
	No. Required	1
Function	Pump concentrated brine stream to heat exchanger	
Operation	Continuous	
Type	Centrifugal, 3600RPM, VSC 75 Hp	
Stream ID	**Inlet**	**Outlet**
	HTBRINE2	HTBRINE9
Pressure (bar)	0.5	2
Design Data:	Flow Rate (kg/hr)	661000
	Pump Head (ft)	97
	Brake Horsepower (Hp)	24
	Electricity Consumed (kWh)	69000
	Construction Material	Cast Iron
	Brake Effeciency	73%
	Motor Efficiency	88%
Utility Costs ($/year)	$	5,300
Purchase Cost	$	8,900
Motor Purchase Cost	$	1,700
Bare Module Cost	$	29,000

PUMP 3		
Identification	Item	*Brine Pump*
	Item No.	BRIPUMP3
	No. Required	1
Function	Pump concentrated brine from crystallizer to mixer tank	
Operation	Continuous	
Type	Centrifugal, 3600RPM, VSC 75 Hp	
Stream ID	**Inlet**	**Outlet**
	HTBRIN10	HTBRIN11
Pressure (bar)	0.5	2
Design Data:	Flow Rate (kg/hr)	345000
	Pump Head (ft)	97
	Brake Horsepower (Hp)	39
	Electricity Consumed (kWh)	69000
	Construction Material	Cast Iron
	Brake Effeciency	77%
	Motor Efficiency	89%
Utility Costs ($/year)	$	5,300
Purchase Cost	$	10,600
Motor Purchase Cost	$	2,700
Bare Module Cost	$	35,000

Mixer Tank

Mixer Tank			
Identification	Item		*Mixer*
	Item No.		MIX1
	No. Required		1
Function	Mixes concentrated bottoms product from crystallizer with brine stream		
Operation	Continuous		
Type	Tank		
Stream ID	**Feed 1**	**Feed 2**	**Outlet**
	HTBRINE5	HTBRIN11	HTBRINE6
Flow Rate (kg/hr)	165000	345000	510000
Temperature (°C)	90	92	91
Pressure (bar)	0.5	2	0.5
Design Data:			
	Max Volume (m^3)		100
	Operating Pressure (psig)		7.3
	Shell Thickness (ft)		0.03
	Length (ft)		10
	Construction Material		Carbon Steel
Purchase Cost		$	26,300
Bare Module Cost		$	109,00

$MgCl_2$ Separation Unit

The following tables show the unit descriptions and specifications for all pieces of equipment used in the designed $MgCl_2$ separation unit.

Flash Drums

Flash Drum 1			
Identification	Item		*Deaerator*
	Item No.		DEAERATE
	No. Required		1
Function	Remove non-compressible fluids		
Operation	Continuous		
Type	Vertical Drum		
Stream ID	**Feed**	**Distillate**	**Bottoms**
	FEED	NONCOMP	AERFED
Flow Rate (kg/hr)	178261	14000	164000
Pressure (bar)	1	0.15	0.15
Temperature (°C)	25	50	50
Design Data:	Operating Pressure (bar)		0.15
	Vapor Fraction		0.1
	Heat Duty (GCAL/HR)		4.0
	Mass Steam (lb/yr)		8.7E+06
	Construction Material		Stainless Steel
Utility Cost ($/year)		$	52,500
Purchase Cost		$	87,400
Bare Module Cost		$	363,000

Flash Drum 2			
Identification	Item		*Flash Evaporator*
	Item No.		EVAP1
	No. Required		1
Function	Remove water content from brine inlet via vaporization		
Operation	Continuous		
Type	Vertical Drum		
Stream ID	**Feed**	**Distillate**	**Bottoms**
	EF1	WAT1	BOTT1
Flow Rate (kg/hr)	183000	30000	154000
Pressure (bar)	2	0.15	0.15
Temperature (°C)	44	60	60
Design Data:	Operating Pressure (bar)		0.15
	Vapor Fraction		0.2
	Heat Duty (GCAL/HR)		21
	Mass Steam (lb/yr)		4.7E+07
	Construction Material		Stainless Steel
Utility Cost ($/year)		$	283, 000
Purchase Cost		$	86, 000
Bare Module Cost		$	359, 000

Flash Drum 3			
Identification	Item		*Flash Evaporator*
	Item No.		EVAP2
	No. Required		1
Function	Remove water content from brine inlet via vaporization		
Operation	Continuous		
Type	Vertical Drum		
Stream ID	**Feed**	**Distillate**	**Bottoms**
	EF2	WAT2	BOTT2
Flow Rate (kg/hr)	309000	60000	249000
Pressure (bar)	1	0.5	0.5
Temperature (°C)	63	90	90
Design Data:	Operating Pressure (bar)		0.5
	Vapor Fraction		0.3
	Heat Duty (GCAL/HR)		45
	Mass Steam (lb/yr)		9.9E+07
	Construction Material		Stainless Steel
Utility Cost ($/year)		$	592, 000
Purchase Cost		$	125, 000
Bare Module Cost		$	520, 000

Flash Drum 4			
Identification	Item		*Flash Evaporator*
	Item No.		EVAP3
	No. Required		1
Function	Remove water content from brine inlet via vaporization		
Operation	Continuous		
Type	Vertical Drum		
Stream ID	**Feed**	**Distillate**	**Bottoms**
	EF3	WAT3	BOTT3
Flow Rate (kg/hr)	174000	39000	135000
Pressure (bar)	1.5	1.5	1.5
Temperature (°C)	90	115	115
Design Data:	Operating Pressure (bar)		1.5
	Vapor Fraction		0.4
	Heat Duty (GCAL/HR)		29
	Mass Steam (lb/yr)		6.4E+07
	Construction Material		Stainless Steel
Utility Cost ($/year)		$	381, 000
Purchase Cost		$	92, 000
Bare Module Cost		$	382, 000

Flash Drum 5			
Identification	Item		*Flash Evaporator*
	Item No.		MGFLASH1
	No. Required		1
Function	Remove water from inlet to		
Operation	Continuous		
Type	Vertical Drum		
Stream ID	**Feed**	**Distillate**	**Bottoms**
	MGLIQ	WAT4	CRSTFED2
Flow Rate (kg/hr)	93000	18000	75000
Pressure (bar)	2	1.02	1.02
Temperature (°C)	25	110	110
Design Data:	Operating Pressure (bar)		1.02
	Vapor Fraction		0.2
	Heat Duty (GCAL/HR)		18
	Mass Steam (lb/yr)		3.9E+07
	Construction Material		Stainless Steel
Utility Cost ($/year)		$	235, 000
Purchase Cost		$	59, 000
Bare Module Cost		$	247, 000

Crystallizers

Crystallizer 1		
Identification	Item	*Carnallite Crystallizer*
	Item No.	CRYST1
	No. Required	1
Function	Form carnallite crystals from solution	
Operation	Continuous	
Type	Forced Circulation Capacity	
Stream ID	**Feed**	**Product**
	CRSTFED1	FILT1IN
Flow Rate (kg/hr)	30000	30000
Carnallite Flow (tons/day)	0	34
Pressure (bar)	1	0.5
Temperature (°C)	35	25
Design Data:	Operating Pressure (bar)	0.5
	Heat Duty (GCAL/HR)	-0.2
	Mass Cooling Water (lb/yr)	5.0E+07
	Construction Material	Carbon Steel
Utility Cost ($/year)	$	600
Purchase Cost	$	252, 000
Bare Module Cost	$	518, 000

Crystallizer 2		
Identification	Item	*$MgCl_2$ Crystallizer*
	Item No.	CRYST2
	No. Required	1
Function	Form magnesium chloride crystals from solution	
Operation	Continuous	
Type	Forced Circulation Capacity	
Stream ID	**Feed**	**Product**
	CRSTFED2	FILT3IN
Flow Rate (kg/hr)	75000	75000
$MgCl_2$ Flow (tons/day)	0	27
Pressure (bar)	1.02	0.5
Temperature (°C)	110	25
Design Data:	Operating Pressure (bar)	0.5
	Heat Duty (GCAL/HR)	-6.3
	Mass Cooling Water (lb/yr)	5.2E+08
	Crystal Production Rate (kg/hr)	110
	Construction Material	Carbon Steel
Utility Cost ($/year)	$	6,200
Purchase Cost	$	61, 000
Bare Module Cost	$	126, 000

Centrifuges

Centrifuge 1			
Identification	Item		*Centrifuge*
	Item No.		CENT1
	No. Required		1
Function	Remove NaCl and other impurities from brine stream		
Operation		Continuous	
Type		Continuous Reciprocating Pusher	
Stream ID	**Feed 1** SLURRY2	**Liquid Product** CLIQ	**Solid Product** SALTPROD
Flow Rate (kg/hr)	75000	40000	34000
Pressure (bar)	1.5	1	1
Temperature (°C)	90	25	25
Design Data:	Operating Pressure (bar)		1
	Electrical Consumption (kWh)		55,456
	Construction Material		Carbon Steel
Utility Cost ($/year)		$	712, 800
Purchase Cost		$	375, 000
Bare Module Cost		$	761, 000

Centrifuge 2			
Identification	Item		*Centrifuge*
	Item No.		CENT2
	No. Required		1
Function	Remove NaCl and other impurities from brine stream		
Operation		Continuous	
Type		Continuous Reciprocating Pusher	
Stream ID	**Feed 1**	**Liquid Product**	**Solid Product**
	CARNAL	CARNAL2	SALTPROD2
Flow Rate (kg/hr)	20000	13000	11000
Pressure (bar)	1.5	1	1
Temperature (°C)	100	25	25
Design Data:	Operating Pressure (bar)		1
	Electrical Consumption (kWh)		55,456
	Construction Material		Carbon Steel
Utility Cost ($/year)		$	712, 800
Purchase Cost		$	273, 000
Bare Module Cost		$	554, 000

Filters

Filter 1			
Identification	Item		*Carnallite Filter*
	Item No.		FILT1
	No. Required		1
Function	Remove solid carnallite from solution and recycle liquid stream		
Operation	Continuous		
Type	Continuous Rotary Drum Vacuum Filter, 20rph		
Stream ID	**Feed**	**Liquid Product**	**Solid Product**
	FILTPUMP1	CARNLIQ	SOLIDCARN
Flow Rate (kg/hr)	27000	20000	7200
Pressure (bar)	3	2	2
Temperature (°C)	25	25	25
Design Data:	Operating Pressure (bar)		2
	Diameter (ft)		17
	Cross-Sectional Area (ft^2)		230
	Heat Duty (GCAL/HR)		0
	Mass Steam (lb/yr)		0
	Construction Material		Cast Iron
Utility Cost ($/year)		$	0.00
Purchase Cost		$	214, 000
Bare Module Cost		$	496, 000

Filter 2			
Identification	Item		*KCl Filter*
	Item No.		FILT2
	No. Required		1
Function	Remove solid KCl from solution and recycle aqueous KCl stream		
Operation	Continuous		
Type	Continuous Rotary Drum Vacuum Filter, 20rph		
Stream ID	**Feed**	**Liquid Product**	**Solid Product**
	FILTPUMP2	MGLIQ	FILTKCL
Flow Rate (kg/hr)	95000	93000	2600
Pressure (bar)	3	2	2
Temperature (°C)	15	25	25
Design Data:	Operating Pressure (bar)		2
	Diameter (ft)		17
	Cross-Sectional Area (ft^2)		213
	Heat Duty (GCAL/HR)		0.9
	Mass Steam (lb/yr)		2.0E+06
	Construction Material		Cast Iron
Utility Cost ($/year)		$	12, 000
Purchase Cost		$	207, 000
Bare Module Cost		$	481, 000

Filter 3			
Identification	Item		*$MgCl_2$Filter*
	Item No.		FILT3
	No. Required		1
Function	Remove solid $MgCl_2$ crystals from solution		
Operation	Continuous		
Type	Continuous Rotary Drum Vacuum Filter, 20rph		
Stream ID	**Feed**	**Liquid Product**	**Solid Product**
	FILTPUMP3	MGRECYC	PRODUCT
Flow Rate (kg/hr)	75000	73000	1600
Pressure (bar)	3	2	2
Temperature (°C)	25	25	25
Design Data:	Operating Pressure (bar)		2
	Diameter (ft)		6
	Cross-Sectional Area (ft^2)		30
	Heat Duty (GCAL/HR)		0.00
	Mass Steam (lb/yr)		0.00E+00
	Construction Material		Cast Iron
Utility Cost (/year)		$	0.00
Purchase Cost		$	116, 000
Bare Module Cost		$	269, 000

Settler Tanks

Settler Tank 1				
Identification	Item			*Settler Tank 1*
	Item No.			SETT1
	No. Required			1
Function	Settles EVAP2 bottoms product and sends solid to CENT1			
Operation	Continuous			
Type	Tank			
Stream ID	**Feed**	**Outlet 1**		**Outlet 2**
	PUMPBOTT2	EF3		SLURRY2
Flow Rate (kg/hr)	249000	174000		75000
Temperature (°C)	90	90		90
Pressure (bar)	2	1.5		1.5
Design Data:				
	Internal gauge pressure (psig)			11.3
	Shell Thickness (ft)			0.04
	Length (ft)			22
	Construction Material			Carbon Steel
Purchase Cost			$	86, 000
Bare Module Cost			$	358, 000

Settler Tank 2			
Identification	Item		*Settler Tank 2*
	Item No.		SETT2
	No. Required		1
Function	Settles EVAP3 bottoms product and sends liquid to CENT2		
Operation	Continuous		
Type	Tank		
Stream ID	**Feed**	**Outlet 1**	**Outlet 2**
	PUMPBOTT3	CARNAL	SLURRY1
Flow Rate (kg/hr)	135000	20000	115000
Temperature (°C)	115	115	115
Pressure (bar)	2	1.5	1.5
Design Data:			
	Internal Gauge Pressure (psig)		11
	Shell Thickness (ft)		0.04
	Length (ft)		26
	Construction Material		Carbon Steel
Purchase Cost		$	108, 000
Bare Module Cost		$	448, 000

Leach Tanks

Leach Tank 1				
Identification	Item			*Leach Tank*
	Item No.			Leach
	No. Required			1
Function	Settles EVAP3 bottoms product			
Operation	Continuous			
Type	Tank			
Stream ID	**Feed 1**	**Feed 2**	**Feed 3**	**Outlet**
	COLDWAT	SOLDCARN	MGRECYC	FILT2IN
Flow Rate (kg/hr)	16000	7200	73000	95000
Temperature (°C)	5	25	25	15
Pressure (bar)	1	2	2	0.5
Design Data:				
	Internal Gauge Pressure (psig)			11
	Shell Thickness (ft)			0.02
	Length (ft)			8
	Construction Material			Carbon Steel
Purchase Cost			$	18, 000
Bare Module Cost			$	58, 000

Mixer Tanks

Mixer Tank 1			
Identification	Item		*Mixer Tank 1*
	Item No.		MIXFEED
	No. Required		1
Function	Mixes brine feed stream with the recycle		
Operation	Continuous		
Type	Tank		
Stream ID	**Feed 1** AERFED	**Feed 2** CARNLIQ	**Outlet** MIXFEED
Flow Rate (kg/hr)	163000	20000	183000
Temperature (°C)	50	25	44
Pressure (bar)	0.2	2	1.2
Design Data:			
	Internal Gauge Pressure (psig)		11
	Shell Thickness (ft)		0.04
	Length (ft)		28
	Construction Material		Carbon Steel
Purchase Cost		$	86, 000
Bare Module Cost		$	277, 000

Mixer Tank 2				
Identification	Item			*Mixer Tank 2*
	Item No.			DISTMIX
	No. Required			1
Function	Mixes concentrated brine, centrifuge liquid and settler recycle			
Operation	Continuous			
Type	Tank			
Stream ID	**Feed 1**	**Feed 2**	**Feed 2**	**Outlet**
	PUMPBOTT1	CLIQ	SLURRY1	EF2
Flow Rate (kg/hr)	164000	40000	115000	309000
Temperature (°C)	50	25	115	63
Pressure (bar)	0.15	1	1.5	1
Design Data:				
	Internal Gauge Pressure (psig)			11
	Shell Thickness (ft)			0.04
	Length (ft)			36
	Construction Material			Carbon Steel
Purchase Cost			$	125, 000
Bare Module Cost			$	401, 000

Mixer Tank 3					
Identification	Item				*Mixer Tank 3*
	Item No.				MIXTANK
	No. Required				1
Function	Mixes carnallite solution with feed water before entering carnallite crystallizer				
Operation	Continuous				
Type	Tank				
Stream ID	**Feed 1**	**Feed 2**	**Feed 2**	**Outlet**	
	CARNAL2	FEEDWAT	KCLRECY	CRSTFED1	
Flow Rate (kg/hr)	12000	16000	2400	30000	
Temperature (°C)	25	25	25	35	
Pressure (bar)	1	1	2	1	
Design Data:					
	Internal Gauge Pressure (psig)				11
	Shell Thickness (ft)				0.03
	Length (ft)				9
	Construction Material				Carbon Steel
Purchase Cost				$	16, 000
Bare Module Cost				$	51, 000

Pumps

PUMP 1		
Identification	Item	*Pump*
	Item No.	PUMP1
	No. Required	1
Function	Pump mixed feed to first flash evaporator	
Operation	Continuous	
Type	Centrifugal, 3600RPM	
Stream ID	**Inlet**	**Outlet**
	MIXFEED	EF1
Pressure (bar)	0.15	2
Design Data:	Flow Rate (kg/hr)	183000
	Pump Head (ft)	52
	Brake Horsepower (Hp)	480
	Electricity Consumed (kWh)	88000
	Construction Material	Cast Iron
	Brake Efficiency	89%
	Motor Efficiency	93%
Utility Costs ($/year)	$	6,800
Pump Purchase Cost	$	24, 000
Motor Purchase Cost	$	29, 000
Bare Module Cost	$	176, 000

PUMP 2		
Identification	Item	*Pump*
	Item No.	PUMP2
	No. Required	1
Function	Pump EVAP1 bottoms product to mixer tank	
Operation	Continuous	
Type	Centrifugal, 3600RPM	
Stream ID	**Inlet**	**Outlet**
	BOTT1	PUMPBOTT1
Pressure (bar)	0.15	2
Design Data:	Flow Rate (kg/hr)	154000
	Pump Head (ft)	52
	Brake Horsepower (Hp)	10
	Electricity Consumed (kWh)	74000
	Construction Material	Cast Iron
	Brake Efficiency	89%
	Motor Efficiency	86%
Utility Costs ($/year)	$	5,700
Pump Purchase Cost	$	24, 000
Motor Purchase Cost	$	860
Bare Module Cost	$	82, 000

PUMP 3		
Identification	Item	*Pump*
	Item No.	PUMP3
	No. Required	1
Function	Pump EVAP2 bottoms product to settler tank	
Operation	Continuous	
Type	Centrifugal, 3600RPM	
Stream ID	**Inlet**	**Outlet**
	BOTT2	PUMPBOTT2
Pressure (bar)	0.5	2
Design Data:	Flow Rate (kg/hr)	249000
	Pump Head (ft)	42
	Brake Horsepower (Hp)	10
	Electricity Consumed (kWh)	82606
	Construction Material	Cast Iron
	Brake Efficiency	88%
	Motor Efficiency	86%
Utility Costs ($/year)	$	6,400
Pump Purchase Cost	$	31, 000
Motor Purchase Cost	$	860
Bare Module Cost	$	106, 000

PUMP 4		
Identification	Item	*Pump*
	Item No.	PUMP4
	No. Required	1
Function	Pump EVAP3 bottoms product to settler tank	
Operation	Continuous	
Type	Centrifugal, 3600RPM	
Stream ID	**Inlet**	**Outlet**
	BOTT3	PUMPBOTT3
Pressure (bar)	1.15	3
Design Data:	Flow Rate (kg/hr)	135000
	Pump Head (ft)	24
	Brake Horsepower (Hp)	10
	Electricity Consumed (kWh)	25000
	Construction Material	Cast Iron
	Brake Efficiency	89%
	Motor Efficiency	86%
Utility Costs ($/year)	$	2,000
Pump Purchase Cost	$	15, 000
Motor Purchase Cost	$	860
Bare Module Cost	$	53, 000

PUMP 5		
Identification	Item	*Pump*
	Item No.	PUMP5
	No. Required	1
Function	Pump crystallizer outlet through filter	
Operation	Continuous	
Type	Centrifugal, 3600RPM	
Stream ID	**Inlet**	**Outlet**
	FILT1IN	FILTPUMP1
Pressure (bar)	0.5	3
Design Data:	Flow Rate (kg/hr)	30000
	Pump Head (ft)	70
	Brake Horsepower (Hp)	10
	Electricity Consumed (kWh)	14000
	Construction Material	Cast Iron
	Brake Efficiency	55%
	Motor Efficiency	86%
Utility Costs ($/year)	$	1, 000
Pump Purchase Cost	$	3,000
Motor Purchase Cost	$	860
Bare Module Cost	$	13, 000

PUMP 6		
Identification	Item	*Pump*
	Item No.	PUMP6
	No. Required	1
Function	Pump leach tank outlet through filter	
Operation	Continuous	
Type	Centrifugal, 3600RPM	
Stream ID	**Inlet**	**Outlet**
	FILT2IN	FILTPUMP2
Pressure (bar)	0.5	3
Design Data:	Flow Rate (kg/hr)	95000
	Pump Head (ft)	70
	Brake Horsepower (Hp)	10
	Electricity Consumed (kWh)	45000
	Construction Material	Cast Iron
	Brake Efficiency	79%
	Motor Efficiency	86%
Utility Costs ($/year)	$	3,400
Pump Purchase Cost	$	5,300
Motor Purchase Cost	$	860
Bare Module Cost	$	20, 000

PUMP 7		
Identification	Item	*Pump*
	Item No.	PUMP7
	No. Required	1
Function	Pump crystallizer outlet through final filter	
Operation	Continuous	
Type	Centrifugal, 3600RPM	
Stream ID	**Inlet**	**Outlet**
	FILT3IN	FILTPUMP3
Pressure (bar)	0.5	3
Design Data:	Flow Rate (kg/hr)	75000
	Pump Head (ft)	70
	Brake Horsepower (Hp)	10
	Electricity Consumed (kWh)	37000
	Construction Material	Cast Iron
	Brake Efficiency	77%
	Motor Efficiency	86%
Utility Costs ($/year)	$	2,800
Pump Purchase Cost	$	5,000
Motor Purchase Cost	$	860
Bare Module Cost	$	20, 000

Equipment Cost Summary

The cost of the equipment used to build the GE concentrator System and the $MgCl_2$ Separator Unit are listed in Table 13 and Table 14 respectively. These costs were determined using *Product and Process Design Principles* by Seider et al. The block names refer to the pieces of equipment found in the flowsheets from Figures 11 and 12. The total cost of the plant equipment for the GE concentrator System is $8,980,000 compared to the $MgCl_2$ Separator Unit priced at $6,960,000 By switching to the $MgCl_2$ Separator Unit, we save 23% on equipment cost. Additionally, with an estimated revenue of $2.6million from the recovered $MgCl_2$, we roughly targeted a five-fold cost for the new plant at $12,500,000; we were able to meet our manufacturing goal at only 56% of the anticipated cost.

Table 13. Equipment Cost Summary for the GE Concentrator

GE Concentrator System			
Block Name	Cp	Fbm	Cbm
Compressors			
VAPCOMP	$ 2,114,000	1.5	$ 3,172,000
VAPCOMP2	$ 1,976,000	1.5	$ 2,965,000
Flash Vessels			
DEAERATE	$ 151, 000	4.16	$ 630, 000
EVAP	$ 110, 000	4.16	$ 461, 000
CRYST	$ 277, 000	4.16	$ 1,153, 000
Heat Exchangers			
HEAT1	$ 29, 000	3.17	$ 93, 000
HEAT2	$ 39, 000	3.17	$ 125, 000
HEAT3	$ 74, 000	3.17	$ 234, 000
Mixer			
MIX1	$ 26, 000	3.21	$ 85, 000
Pump			
BRIPUMP1	$ 4,300	3.3	$ 14, 000
PUMPBRI2	$ 6,100	3.3	$ 20, 000
BRIPUMP3	$ 8,000	3.3	$ 26, 000
Total			**$ 8,980,000**

Table 14. Equipment Cost Summary for the $MgCl_2$ Separator Unit

MgCl₂ Separation Unit			
Block Name	Cp	Fbm	Cbm
Centrifuges			
CENT	$ 375,000	2.03	$ 761,000
CENT2	$ 272,000	2.03	$ 554,000
Crystallizers			
CRYST1	$ 252,000	2.06	$ 518,000
CRYST2	$ 61,000	2.06	$ 126,000
Filters			
FILT1	$ 214,000	2.32	$ 496,000
FILT2	$ 207,000	2.32	$ 481,000
FILT3	$ 116,000	2.32	$ 269,000
Flash Vessels			
DEAERATE	$ 87,000	4.16	$ 364,000
EVAP1	$ 86,000	4.16	$ 359,000
EVAP2	$ 125,000	4.16	$ 520,000
EVAP3	$ 92,000	4.16	$ 382,000
MGFLASH1	$ 59,000	4.16	$ 247,000
Leach Tank			
LEACH	$ 18,000	3.21	$ 58,000
Mixing Tanks			
MIXFEED	$ 86,000	3.21	$ 277,000
DISTMIX	$ 125,000	3.21	$ 401,000
MIXTANK	$ 16,000	3.21	$ 51,000
Pumps			
PUMP1	$ 53,000	3.3	$ 176,000
PUMP2	$ 25,000	3.3	$ 82,000
PUMP3	$ 32,000	3.3	$ 106,000
PUMP4	$ 16,000	3.3	$ 53,000
PUMP5	$ 3,900	3.3	$ 13,000
PUMP6	$ 6,200	3.3	$ 20,000
PUMP7	$ 5,900	3.3	$ 20,000
Settling Tanks			
SETT1	$ 86,000	3.21	$ 277,000
SETT2	$ 108,000	3.21	$ 346,000
Total			**$ 6,960,000**

Waste Disposal Methods

The current methods of waste disposal in the Carlsbad facility involve mixing the concentrated brine wastewater with the cooling water from the desalination process and returning the water to the ocean [32]. A smaller plant north of the Carlsbad facility is using an existing outfall that has pinholes to introduce small amounts of the brine back into the ocean [33]. Both of these options can be implemented with no additional variable costs and are more environmentally sound as a method of dealing with the waste stream from the magnesium chloride separation system.

Environmental Hazards

The main environmental hazards directly associated with our process are greenhouse gas emissions associated with electricity consumption and environmental impacts resulting from waste disposal from the process. Additionally, the seawater intake step from the desalination plant is a considerable area of concern; however, this step results from upstream processing from the desalination plant [36].

Water Intake Upstream

In 2006, an Environmental Impact was conducted to assess the potential risk to adjacent water systems, marine life, and air quality prior to construction of the Carlsbad desalination plant. With regard to marine life, the intake of seawater into the plant causes impingement of microorganisms such as fish larvae, while small fish may be caught on intake screens or intake bars. The EIR determined that this danger to the fish was insignificant for two reasons; first, it was concluded that the survival rate of the larvae had no notable effect on the fish population. Additionally, the fish death resulting from sea water intake is less than the fish death resulting from fish being consumed by brown pelicans [36].

Brine Waste Disposal

Brine waste disposal methods determine the potential adverse effects on the environment. If a deep well injection method of disposal is used, there is a risk of infiltration and contaminating nearby aquifers. However, surveying can be done to assess whether or not there are any aquifers susceptible to contamination near a proposed injection site to prevent said contamination.

If a landfill disposal method is selected there is also a possibility that salt constituents penetrate through the soil and contaminate feed water, ponds, or aquifers. Areas at risk include soil that has high cationic transfer capacity, high permeability, or a low clay content. A study conducted in eastern Abu Dhabi concluded that landfill disposal techniques caused traces of brine components to contaminate nearby water sources such as ponds and aquifers, which at times exceeded maximum allowable levels. To prevent infiltration of the brine components into soil, the landfill can be lined. An assessment of the soil may also be considered in determining a feasible location for a landfill, based on soil characteristics and proximity to water sources.

To assess whether a liquid disposal is a suitable method of waste disposal, it is important that the brine output location has a high water velocity to prevent sedimentation from occurring. If sedimentation of brine constituents occurs, the surrounding area becomes at risk of adsorbing these species; however, adsorption does not pose an extreme threat to the marine life. The following species are considered sensitive to salinity: algae, coral reefs, seaweed bays, salt marsh containing marcophytes, and mangal areas. If any are present in close proximity to the disposal site, a different location or disposal method ought to be selected.

Brine salinity is considered to be toxic to a marine environment if it impedes growth, survival, or reproduction. Wet Effluent Toxicity (WET) is a technique used to measure the threshold at which the effluent become both acutely and chronically toxic to surrounding marine life. WET involves the use of bioassays of sensitive marine species such as fish, invertebrates, or plants. WET has been used widely in Australia and has determined the necessity of dilution ratios from 40:1-45:1 to protect 99% or above of marine life.

Salinity Tolerance Evaluation (STE) is a test used to determine impacts on marine life for both the Carlsbad and Huntington Beach desalination plants. It is a long term test spanning 5 months, during which 18 species are exposed to a large range of salinities in a single aquarium. The range of salinities tested is meant to observe the marine life in the extreme and worst case scenarios, to mimic the maximum possible salinity that may be achievable by the liquid discharge system. STE confirmed that while some species were sensitive to salinity increases, none were at risk as all species could tolerate the highest possible salinity of 40 mg/L [30].

Greenhouse Gas Emissions

Greenhouse gas emission are inherent in electricity use associated with plant operation. The EIR concluded that greenhouse gas emissions specifically from desalination processes were not substantially enough to pose an environmental threat [36].

Mitigation Techniques

Poseidon Water is the private owner of the Carlsbad desalination plant. Poseidon has enacted a mitigation plan including the purchase of 66 acres of salt flats with construction and contouring to cause water to wash over and create new wetlands. The creation of the wetlands and facilitation of a new estuary would create a new environment to host fish similar to those harmed by the intake to the plant [35]. To minimize the carbon footprint associated with the plant, Poseidon is installing solar panels to provide a non-fossil fuel alternative power source, while purchasing carbon offsets [36].

Safety Measures

Concentrated brine is not considered a significantly hazardous material. With respect to human safety, brine should be kept away from eyes and skin and not directly consumed. However, exposure of this nature is not fatal and should only cause irritation. Brine is not significantly dangerous unless ingested, in which case nausea, stomach, and bowel irritation is possible. Brine is not flammable, highly explosive, or combustible. If a spill occurs, the leak should be confined as safely as possible and kept away from waterways. A vacuum or pump can be used to remove the brine, and the affected area should be washed with water. Measures can be taken with handling that includes routine inspection of pipes and equipment to check for corrosion, as brine is corrosive over time [37].

Pump Safety

Pumps must be installed and properly grounded in order to prevent any risk of electric shock. Additionally, pumps should be disconnected from its electrical source prior to servicing or handling to prevent risk of electric shock; subsequently, some time for cool down after disconnecting should be allotted prior to servicing and handling. A pump may be dangerously hot if in contact with skin directly after use. After maintenance or handling, safety devices must be replaced prior to operation. If the discharge valve is closed and the pump is operating, the seal may fail or create very high pressure, high temperature steam which can cause an explosion or skin burns. It is recommended that a pressure relief valve be installed on the pump to prevent this malfunction. If a pump is operated at a higher pressure than recommended, the motor may overheat, which can also cause a systematic malfunction.

When the pump is connected to electricity source and in operation, it can be dangerous to put fingers in suction or discharge locations. Loose clothing should not be worn near these openings either [38].

Location & Other Concerns

We have decided to locate our plant in Carlsbad, California, as the necessary infrastructure is already in place for upstream desalination using reverse osmosis and downstream waste disposal into the ocean. The established location for desalination also ensures a market for our product near the facility, which will help to minimize additional transportation costs of the water. Drought-related issues have plagued the San Diego region for many years, providing a persistent demand for water. Additionally, since Carlsbad is an established location for desalination, the city has also shown their support for the construction of similar projects [39]. Building in San Diego, a metropolitan area, creates simplistic avenues for the transportation and distribution of our byproducts, $MgCl_2$ and KCl.

Potential concerns for a brine concentrating plant consist of fouling, biological growth, and the location of process controls. Fouling can occur over time due to the variety of components present in the brine. Some of these components may result in precipitates, which can accumulate over time, so chemical cleaning processes may be considered. Often in desalination plants, the feed stream is chlorinated with either free chloride or NaCl prior to entering the process to prevent biological growth. Because we are receiving a concentrated brine stream from an RO plant, the chlorine content is already substantial enough that we are not concerned with biological growth. Finally, with a process that deals heavily with water, it is important to make sure that the controls are located in a place not at risk of contact with the water as this may cause damage [34].

Furthermore, we have taken into account the solid content of all of the streams in both the GE Brine Concentrator and the $MgCl_2$ separation unit in order to assess whether or not solid-specific pumps are necessary to pump the respective streams from block to block. After doing so, we have determined that centrifugal pumps will suffice all throughout both processes entirely;

solid fractions range from 0 (mainly distillate water streams) to approximately 0.35 (predominantly bottoms product streams from crystallizers and flash vessels), which are within the pumpable limit for a generic, 75Hp centrifugal pump. This analysis was absolutely vital in guaranteeing an accurate utilities assessment and subsequently a profitability analysis; solid-specific pumps would have significantly increased utility (power) consumption for either system, inadvertently raising overall costs and thus diminishing our process' profitability. By controlling the solid fractions throughout the process, we are able to maximize profits while still recovering a substantial amount of water and crystalline products.

Profitability Analysis

The main factors that determine the profitability of each system include the products, byproducts, and utilities. Both systems produce water as the main product; however, the $MgCl_2$ Separator Unit produces additional byproducts of $MgCl_2$ and KCl. The GE system does not yield any byproducts. For every pound of water treated, 0.0055 lb $MgCl_2$ and 0.00047 lb KCl are produced. The combination of revenue from these byproducts generates $1,414,988 per year. The utilities required for both plants include low pressure steam, cooling water, and electricity. Due to the choice of equipment, the $MgCl_2$ Separator Unit requires more pounds of steam and cooling water per pound of water produced than the GE Brine Concentrator System, though it requires less power per pound of water for utilities than the GE Brine Concentrator System. Overall, the utilities for the $MgCl_2$ Separator Unit are less than that of the GE Brine Concentrator System by 75%. With regard to equipment costs, the GE system costs roughly $8.3 million, while the $MgCl_2$ Separation Unit has equipment costs of $6.7 million. Figure 12 shown below gives a summary of major profitability measures. Tables 15-35 give a more thorough profitability analysis.

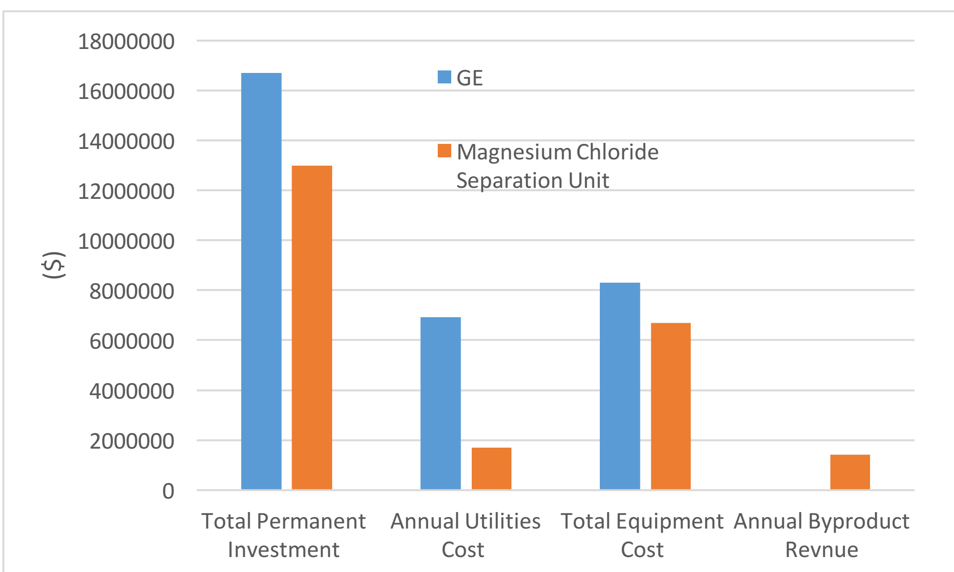


Figure 12. Profitability Comparison between $MgCl_2$ Separation Unit and GE Brine Concentrator

Table 15. General Economic Information GE Brine Concentrator. Note that the price of water is $0.00078/lb which rounds to $0.00 in the spreadsheet.

General Information	
Process Title:	**Brine Wastewater Treatment**
Product:	**Treated Water**
Plant Site Location:	**San Diego**
Site Factor:	**1.25**
Operating Hours per Year:	**7919**
Operating Days Per Year:	**330**
Operating Factor:	**0.9040**

Product Information		
This Process will Yield		
	266,439	lb of Treated Water per hour
	6,394,538	lb of Treated Water per day
	2,109,941,856	lb of Treated Water per year
Price	**$0.00**	**/lb**

Table 16. General Economic Information $MgCl_2$ Separation Unit. Note that the price of water is $0.00078/lb which rounds to $0.00 in the spreadsheet.

General Information	
Process Title:	**Brine Wastewater Treatment**
Product:	**Treated Water**
Plant Site Location:	**San Diego**
Site Factor:	**1.25**
Operating Hours per Year:	**7919**
Operating Days Per Year:	**330**
Operating Factor:	**0.9040**

Product Information		
This Process will Yield		
	278,251	lb of Treated Water per hour
	6,678,033	lb of Treated Water per day
	2,203,483,611	lb of Treated Water per year
Price	**$0.00**	**/lb**

Table 17. Chronology for GE Brine Concentrator

Year	Action	Distribution of Permanent Investment	Production Capacity	Depreciation 5 year MACRS	Product Price
2016	Design		0.0%		
2017	Construction	100%	0.0%		
2018	Production	0%	45.0%	20.00%	$0.00
2019	Production	0%	67.5%	32.00%	$0.00
2020	Production	0%	90.0%	19.20%	$0.00
2021	Production		90.0%	11.52%	$0.00
2022	Production		90.0%	11.52%	$0.00
2023	Production		90.0%	5.76%	$0.00
2024	Production		90.0%		$0.00
2025	Production		90.0%		$0.00
2026	Production		90.0%		$0.00
2027	Production		90.0%		$0.00
2028	Production		90.0%		$0.00
2029	Production		90.0%		$0.00
2030	Production		90.0%		$0.00
2031	Production		90.0%		$0.00
2032	Production		90.0%		$0.00

Table 18. Chronology for $MgCl_2$ Separation Unit

Year	Action	Distribution of Permanent Investment	Production Capacity	Depreciation 5 year MACRS	Product Price
2016	Design		0.0%		
2017	Construction	100%	0.0%		
2018	Production	0%	45.0%	20.00%	$0.00
2019	Production	0%	67.5%	32.00%	$0.00
2020	Production	0%	90.0%	19.20%	$0.00
2021	Production		90.0%	11.52%	$0.00
2022	Production		90.0%	11.52%	$0.00
2023	Production		90.0%	5.76%	$0.00
2024	Production		90.0%		$0.00
2025	Production		90.0%		$0.00
2026	Production		90.0%		$0.00
2027	Production		90.0%		$0.00
2028	Production		90.0%		$0.00
2029	Production		90.0%		$0.00
2030	Production		90.0%		$0.00
2031	Production		90.0%		$0.00
2032	Production		90.0%		$0.00

Table 19. Equipment Costs for GE Brine Concentrator

Equipment Costs		
Equipment Description		**Bare Module Cost**
Heat Exchanger 2.1	Process Machinery	$92,615
Heat Exchanger 2.2	Process Machinery	$124,895
Heat Exchanger 2.3	Process Machinery	$233,670
Compressor 2.1	Process Machinery	$3,172,014
Pump 2.1	Process Machinery	$14,220
Crystallizer (Flash) 2.1	Process Machinery	$1,153,175
Flash 2.1	Process Machinery	$630,272
Pump 2.2	Process Machinery	$20,232
Compressor 2.2	Process Machinery	$2,965,048
Flash 2.2	Process Machinery	$461,479
Pump 2.3	Process Machinery	$26,405
Mixer 2.1	Process Machinery	$84,620
Total		***$8,978,645***

Table 20. Equipment Costs for $MgCl_2$ Separation Unit

Equipment Costs		
Equipment Description		**Bare Module Cost**
Flash 1.1	Process Machinery	$363,766
Flash 1.2	Process Machinery	$358,960
Flash 1.3	Process Machinery	$520,137
Flash 1.4	Process Machinery	$382,500
Flash 1.5	Process Machinery	$247,390
Mixer 1.1	Process Machinery	$276,986
Mixer 1.2	Process Machinery	$401,356
Mixer 1.3	Process Machinery	$50,613
Centrifuge 1.1	Process Machinery	$761,132
Centrifuge 1.2	Process Machinery	$554,170
Settler 1.1	Process Machinery	$276,526
Settler 1.2	Process Machinery	$345,621
Leacher 1.1	Process Machinery	$57,736
Crystallizer 1.1	Process Machinery	$518,342
Crystallizer 1.2	Process Machinery	$125,937
Filter 1.1	Process Machinery	$496,257
Filter 1.2	Process Machinery	$481,069
Filter 1.3	Process Machinery	$268,537
Pump 1.1	Process Machinery	$175,689
Pump 1.2	Process Machinery	$81,500
Pump 1.3	Process Machinery	$106,098
Pump 1.4	Process Machinery	$52,693
Pump 1.5	Process Machinery	$12,980
Pump 1.6	Process Machinery	$20,489
Pump 1.7	Process Machinery	$19,594
Total		***$6,956,076***

Table 21. Utilities for GE Brine Concentrator

Utilities						
	Utility:	Unit:	Required Ratio		Utility Cost	
1	High Pressure Steam	lb	0	lb per lb of Treated Water	$0.000E+0 0	per lb
2	Low Pressure Steam	lb	0.01049	lb per lb of Treated Water	$6.000E-03	per lb
3	Process Water	gal	0	gal per lb of Treated Water	$0.000E+0 0	per gal
4	Cooling Water	lb	0.04528	lb per lb of Treated Water	$1.000E-04	per lb
5	Electricity	kWh	0.04137	kWh per lb of Treated Water	$0.078	per kWh
	Total Weighted Average:				$3.286E-03	per lb of Treated Water

Table 22. Utilities for $MgCl_2$ Separation Unit

Utilities						
	Utility:	Unit:	Required Ratio		Utility Cost	
1	High Pressure Steam	lb	0	lb per lb of Treated Water	$0.000E+0 0	per lb
2	Low Pressure Steam	lb	0.11765	lb per lb of Treated Water	$6.000E-03	per lb
3	Process Water	gal	0	gal per lb of Treated Water	$0.000E+0 0	per gal
4	Cooling Water	lb	0.259	lb per lb of Treated Water	$1.000E-04	per lb
5	Electricity	kWh	0.00081 3	kWh per lb of Treated Water	$0.078	per kWh
	Total Weighted Average:				$7.951E-04	per lb of Treated Water

Table 23. Variable Costs for GE Brine Concentrator

Variable Costs		
General Expenses:		
Selling / Transfer Expenses:	**3.00%**	**of Sales**
Direct Research:	**4.80%**	**of Sales**
Allocated Research:	**0.50%**	**of Sales**
Administrative Expense:	**2.00%**	**of Sales**
Management Incentive Compensation:	**1.25%**	**of Sales**

Table 24. Variable Costs for $MgCl_2$ Separation Unit

Variable Costs			
General Expenses:			
	Selling / Transfer Expenses:	**3.00%**	**of Sales**
	Direct Research:	**4.80%**	**of Sales**
	Allocated Research:	**0.50%**	**of Sales**
	Administrative Expense:	**2.00%**	**of Sales**
	Management Incentive Compensation:	**1.25%**	**of Sales**

Table 25. Working Capital for GE Brine Concentrator

Working Capital			
Accounts Receivable	⇨	**30**	**Days**
Cash Reserves (excluding Raw Materials)	⇨	**30**	**Days**
Accounts Payable	⇨	**30**	**Days**
Treated Water Inventory	⇨	**4**	**Days**
Raw Materials	⇨	**2**	**Days**

Table 26. Working Capital for $MgCl_2$ Separation Unit

Working Capital			
Accounts Receivable	⇨	**30**	**Days**
Cash Reserves (excluding Raw Materials)	⇨	**30**	**Days**
Accounts Payable	⇨	**30**	**Days**
Treated Water Inventory	⇨	**4**	**Days**
Raw Materials	⇨	**2**	**Days**

Table 27. Total Permanent Investment for GE Brine Concentrator

Total Permanent Investment		
Cost of Site Preparations:	**5.00%**	**of Total Bare Module Costs**
Cost of Service Facilities:	**5.00%**	**of Total Bare Module Costs**
Allocated Costs for utility plants and related facilities:	**$0**	
Cost of Contingencies and Contractor Fees:	**18.00%**	**of Direct Permanent Investment**
Cost of Land:	**2.00%**	**of Total Depreciable Capital**
Cost of Royalties:	**$0**	
Cost of Plant Start-Up:	**10.00%**	**of Total Depreciable Capital**

Table 28. Total Permanent Investment for $MgCl_2$ Separation Unit

Total Permanent Investment		
Cost of Site Preparations:	5.00%	of Total Bare Module Costs
Cost of Service Facilities:	5.00%	of Total Bare Module Costs
Allocated Costs for utility plants and related facilities:	$0	
Cost of Contingencies and Contractor Fees:	18.00%	of Direct Permanent Investment
Cost of Land:	2.00%	of Total Depreciable Capital
Cost of Royalties:	$0	
Cost of Plant Start-Up:	10.00%	of Total Depreciable Capital

Table 29. Fixed Costs for GE Brine Concentrator

Fixed Costs		
Operations		
Operators per Shift:	1	(assuming 5 shifts)
Direct Wages and Benefits:	$40	/operator hour
Direct Salaries and Benefits:	15%	of Direct Wages and Benefits
Operating Supplies and Services:	6%	of Direct Wages and Benefits
Technical Assistance to Manufacturing:	$60,000.00	per year, for each Operator per Shift
Control Laboratory:	$65,000.00	per year, for each Operator per Shift
Maintenance		
Wages and Benefits:	4.50%	of Total Depreciable Capital
Salaries and Benefits:	25%	of Maintenance Wages and Benefits
Materials and Services:	100%	of Maintenance Wages and Benefits
Maintenance Overhead:	5%	of Maintenance Wages and Benefits
Operating Overhead		
General Plant Overhead:	7.10%	of Maintenance and Operations Wages and Benefits
Mechanical Department Services:	2.40%	of Maintenance and Operations Wages and Benefits
Employee Relations Department:	5.90%	of Maintenance and Operations Wages and Benefits
Business Services:	7.40%	of Maintenance and Operations Wages and Benefits

Property Taxes and Insurance			
	Property Taxes and Insurance:	**2%**	**of Total Depreciable Capital**

Straight Line Depreciation		
Direct Plant:	8.00%	**of Total Depreciable Capital, less 1.18 times the Allocated Costs for Utility Plants and Related Facilities**
Allocated Plant:	6.00%	**of 1.18 times the Allocated Costs for Utility Plants and Related Facilities**

Other Annual Expenses		
	Rental Fees (Office and Laboratory Space):	**$0**
	Licensing Fees:	**$0**
	Miscellaneous:	**$0**
Depletion Allowance		
	Annual Depletion Allowance:	**$0**

Table 30. Fixed Costs for $MgCl_2$ Separation Unit

Fixed Costs			
Operations			
	Operators per Shift:	**1**	**(assuming 5 shifts)**
	Direct Wages and Benefits:	**$40**	**/operator hour**
	Direct Salaries and Benefits:	**15%**	**of Direct Wages and Benefits**
	Operating Supplies and Services:	**6%**	**of Direct Wages and Benefits**
	Technical Assistance to Manufacturing:	**$60,000.00**	**per year, for each Operator per Shift**
	Control Laboratory:	**$65,000.00**	**per year, for each Operator per Shift**
Maintenance			
	Wages and Benefits:	**4.50%**	**of Total Depreciable Capital**
	Salaries and Benefits:	**25%**	**of Maintenance Wages and Benefits**
	Materials and Services:	**100%**	**of Maintenance Wages and Benefits**
	Maintenance Overhead:	**5%**	**of Maintenance Wages and Benefits**
Operating Overhead			
	General Plant Overhead:	**7.10%**	**of Maintenance and Operations Wages and Benefits**
	Mechanical Department Services:	**2.40%**	**of Maintenance and Operations Wages and Benefits**
	Employee Relations Department:	**5.90%**	**of Maintenance and Operations Wages and Benefits**
	Business Services:	**7.40%**	**of Maintenance and Operations Wages and Benefits**
Property Taxes and Insurance			
	Property Taxes and Insurance:	**2%**	**of Total Depreciable Capital**

Straight Line Depreciation

Direct Plant:	8.00%	**of Total Depreciable Capital, less 1.18 times the Allocated Costs for Utility Plants and Related Facilities**
Allocated Plant:	6.00%	**of 1.18 times the Allocated Costs for Utility Plants and Related Facilities**

Other Annual Expenses

Rental Fees (Office and Laboratory Space):	**$0**
Licensing Fees:	**$0**
Miscellaneous:	**$0**

Depletion Allowance

Annual Depletion Allowance:	**$0**

Table 31. Byproducts for $MgCl_2$ Separation Unit

Byproducts						
	Byproduct:	**Unit:**	**Ratio to Product**		**Byproduct Selling Price**	
1	MgCl2	lb	0.00549	lb per lb of Treated Water	$0.090	per lb
2	KCl	lb	0.000469	lb per lb of Treated Water	$0.225	per lb
	Total Weighted Average:				$5.996E-04	per lb of Treated Water

Table 32. GE Brine Concentrator Cash Flows

Cash Flow Summary

Year	Percentage of Design Capacity	Product Unit Price	Sales	Capital Costs	Working Capital	Var Costs	Fixed Costs	Depreciation	Depletion Allowance	Taxible Income	Taxes	Net Earnings	Cash Flow	Cumulative Net Present Value at 15%
2016	0%		-	-	-	-	-	-	-	-	-	-	-	-
2017	0%		-	(16,316,000)	(184,000)	-	-	-	-	-	-	-	(16,500,000)	(14,347,800)
2018	45%	$0.00	867,800	-	(92,000)	(3,205,600)	(2,826,200)	(2,330,900)	-	(7,494,800)	2,773,100	(4,721,700)	(2,482,900)	(16,225,200)
2019	68%	$0.00	1,334,300	-	(92,000)	(4,808,300)	(2,826,200)	(3,729,400)	-	(10,029,600)	3,711,000	(6,318,700)	(2,681,300)	(17,988,200)
2020	90%	$0.00	1,823,500	-	-	(6,411,100)	(2,826,200)	(2,237,600)	-	(9,651,400)	3,571,000	(6,080,400)	(3,842,800)	(20,185,300)
2021	90%	$0.00	1,869,100	-	-	(6,411,100)	(2,826,200)	(1,342,600)	-	(8,710,800)	3,223,000	(5,487,800)	(4,145,200)	(22,246,200)
2022	90%	$0.00	1,915,800	-	-	(6,411,100)	(2,826,200)	(1,342,600)	-	(8,664,100)	3,205,700	(5,458,400)	(4,115,800)	(24,025,600)
2023	90%	$0.00	1,963,700	-	-	(6,411,100)	(2,826,200)	(671,300)	-	(7,944,900)	2,939,600	(5,005,300)	(4,334,000)	(25,654,900)
2024	90%	$0.00	2,012,800	-	-	(6,411,100)	(2,826,200)	-	-	(7,224,500)	2,673,100	(4,551,400)	(4,551,400)	(27,142,800)
2025	90%	$0.00	2,063,100	-	-	(6,411,100)	(2,826,200)	-	-	(7,174,200)	2,654,500	(4,519,700)	(4,519,700)	(28,427,600)
2026	90%	$0.00	2,114,700	-	-	(6,411,100)	(2,826,200)	-	-	(7,122,600)	2,635,400	(4,487,300)	(4,487,300)	(29,536,800)
2027	90%	$0.00	2,167,600	-	-	(6,411,100)	(2,826,200)	-	-	(7,069,800)	2,615,800	(4,453,900)	(4,453,900)	(30,494,100)
2028	90%	$0.00	2,221,800	-	-	(6,411,100)	(2,826,200)	-	-	(7,015,600)	2,595,800	(4,419,800)	(4,419,800)	(31,320,200)
2029	90%	$0.00	2,277,300	-	-	(6,411,100)	(2,826,200)	-	-	(6,960,000)	2,575,200	(4,384,800)	(4,384,800)	(32,032,900)
2030	90%	$0.00	2,334,200	-	-	(6,411,100)	(2,826,200)	-	-	(6,903,100)	2,554,100	(4,348,900)	(4,348,900)	(32,647,500)
2031	90%	$0.00	2,392,600	-	-	(6,411,100)	(2,826,200)	-	-	(6,844,700)	2,532,500	(4,312,200)	(4,312,200)	(33,177,400)
2032	90%	$0.00	2,452,400	-	367,900	(6,411,100)	(2,826,200)	-	-	(6,784,900)	2,510,400	(4,274,500)	(3,906,500)	(33,594,900)

Table 33. $MgCl_2$ Cash Flows

Cash Flow Summary

Year	Percentage of Design Capacity	Product Unit Price	Sales	Capital Costs	Working Capital	Var Costs	Fixed Costs	Depreciation	Depletion Allowance	Taxible Income	Taxes	Net Earnings	Cash Flow	Cumulative Net Present Value at 15%
2016	0%		-	-	-	-	-	-	-	-	-	-	-	-
2017	0%		-	(12,640,600)	(174,300)	-	-	-	-	-	-	-	(12,814,800)	(11,143,300)
2018	45%	$0.00	906,300	-	(87,100)	(283,100)	(2,468,300)	(1,805,800)	-	(3,650,900)	1,350,800	(2,300,100)	(581,400)	(11,583,000)
2019	68%	$0.00	1,393,400	-	(87,100)	(424,700)	(2,468,300)	(2,889,300)	-	(4,388,800)	1,623,900	(2,765,000)	37,200	(11,558,500)
2020	90%	$0.00	1,904,300	-	-	(566,200)	(2,468,300)	(1,733,600)	-	(2,863,700)	1,059,600	(1,804,200)	(70,600)	(11,598,900)
2021	90%	$0.00	1,952,000	-	-	(566,200)	(2,468,300)	(1,040,100)	-	(2,122,700)	785,400	(1,337,300)	(297,200)	(11,746,600)
2022	90%	$0.00	2,000,800	-	-	(566,200)	(2,468,300)	(1,040,100)	-	(2,073,900)	767,300	(1,306,600)	(266,400)	(11,861,800)
2023	90%	$0.00	2,050,800	-	-	(566,200)	(2,468,300)	(520,100)	-	(1,503,800)	556,400	(947,400)	(427,300)	(12,022,500)
2024	90%	$0.00	2,102,000	-	-	(566,200)	(2,468,300)	-	-	(932,500)	345,000	(587,500)	(587,500)	(12,214,500)
2025	90%	$0.00	2,154,600	-	-	(566,200)	(2,468,300)	-	-	(879,900)	325,600	(554,400)	(554,400)	(12,372,100)
2026	90%	$0.00	2,208,500	-	-	(566,200)	(2,468,300)	-	-	(826,100)	305,600	(520,400)	(520,400)	(12,500,700)
2027	90%	$0.00	2,263,700	-	-	(566,200)	(2,468,300)	-	-	(770,900)	285,200	(485,600)	(485,600)	(12,605,100)
2028	90%	$0.00	2,320,300	-	-	(566,200)	(2,468,300)	-	-	(714,300)	264,300	(450,000)	(450,000)	(12,689,200)
2029	90%	$0.00	2,378,300	-	-	(566,200)	(2,468,300)	-	-	(656,300)	242,800	(413,400)	(413,400)	(12,756,400)
2030	90%	$0.00	2,437,700	-	-	(566,200)	(2,468,300)	-	-	(596,800)	220,800	(376,000)	(376,000)	(12,809,600)
2031	90%	$0.00	2,498,700	-	-	(566,200)	(2,468,300)	-	-	(535,900)	198,300	(337,600)	(337,600)	(12,851,000)
2032	90%	$0.00	2,561,100	-	348,500	(566,200)	(2,468,300)	-	-	(473,400)	175,200	(298,200)	50,300	(12,845,700)

Table 34. Profitability Measures for GE Brine Concentrator

Profitability Measures

The Internal Rate of Return (IRR) for this project is	Negative IRR
The Net Present Value (NPV) of this project in 2016 is	$ (33,594,900)

ROI Analysis (Third Production Year)

Annual Sales	**1,823,505**	
Annual Costs	**(9,237,326)**	
Depreciation	**(1,305,280)**	
Income Tax	**3,226,067**	
Net Earnings	**(5,493,033)**	
Total Capital Investment		**16,683,944**
ROI	**-32.92%**	

Table 35. Profitability Measures for GE Brine Concentrator

Profitability Measures

The Internal Rate of Return (IRR) for this project is	Negative IRR
The Net Present Value (NPV) of this project in 2016 is	$ (12,845,700)

ROI Analysis (Third Production Year)

Annual Sales	**1,904,348**	
Annual Costs	**(3,034,529)**	
Depreciation	**(1,011,246)**	
Income Tax	**792,328**	
Net Earnings	**(1,349,100)**	
Total Capital Investment		**12,989,099**
ROI	**-10.39%**	

Both the GE brine concentrator and the $MgCl_2$ Separation Unit are shown to have a negative IRR, negative ROI, and negative NPV; that said, the GE brine concentrator has a more negative NPV than the $MgCl_2$ Separation Unit (-$34 M versus -$12 M) and ROI (-32.92% as vs. -10.39%). Negative NPV and ROI may seem undesirable at first glance, but are completely expected; to reiterate, any brine wastewater treatment process is not a *profitable* one, but rather is one mandated by legal rules and regulations, as explained in detail in this report's introduction. Thus, the underlying (and realistic) goal for any brine treatment process is not to produce positive ROI and NPV values; rather, it is to generate negative values that are as close to 0 as possible. With that in mind, our $MgCl_2$ process has *increased* GE's ROI by 22.53%, sending the system's ROI closer to the 0% target. Thus, although our process is not *profitable*, it has shown financial improvements to GE's system, making the $MgCl_2$ Separation Unit *less financially costly* than the Brine Concentrator by GE.

A range of prices were available for $MgCl_2$ depending on both purity and intended use. When the profitability analysis was conducted using the price for pharmaceutical grade $MgCl_2$, our process exhibited a positive ROI. However, this ROI does not take into account sterilization procedures that would be required and the equipment that would be purchased for sterilization purposes, which would add to the costs of our plant. Note that the purity of our $MgCl_2$ product is on par with pharmaceutical grade, so the only additional costs necessary to sell pharmaceutical grade $MgCl_2$ would be for sterilization practices. Due to time constraints, we did not include an analysis for these additional costs, so a more conservative estimate for $MgCl_2$ purchase price was used which reflects the price for agricultural grade $MgCl_2$.

Equipment Cost Comparison

The GE system has a higher total cost of equipment despite having fewer pieces of equipment. The GE system totals \$8.3 million while the $MgCl_2$ Separation Unit has a total equipment cost of \$6.7 million. Once again, the use of compressors in the GE system is costly, Compressor 2.1 costs \$3.2 million and Compressor 2.2 costs \$3.0 million. No other piece of equipment in either the GE system or the $MgCl_2$ Separation Unit exceeds \$1 million; thus, the lack of compressors in the $MgCl_2$ Separation Unit single-handedly makes our designed process more financially affordable than the currently marketed GE Brine Concentrator. The most expensive pieces of equipment in the $MgCl_2$ Separation Unit include the centrifuges, crystallizers, and filters. Centrifuge 1.1 costs \$761,000, Centrifuge 1.2 is \$554,000, Crystallizer 1.1 is \$518,000, Filter 1.1 costs \$496,000, and Filter 1.2 costs \$481,000.

Sensitivity Analysis

In a bid to determine the degree of sensitivity that our design has to the market prices of both $MgCl_2$ and KCl, a simple sensitivity analysis was conducted, producing Figure 13 below. We studied the effects of altering $MgCl_2$ selling prices (while holding KCl prices constant at the current market price of $0.225/lb) on the system ROI, and subsequently repeated the analysis by altering KCl prices (while holding $MgCl_2$ prices constant at the current market price of $0.09/lb.)

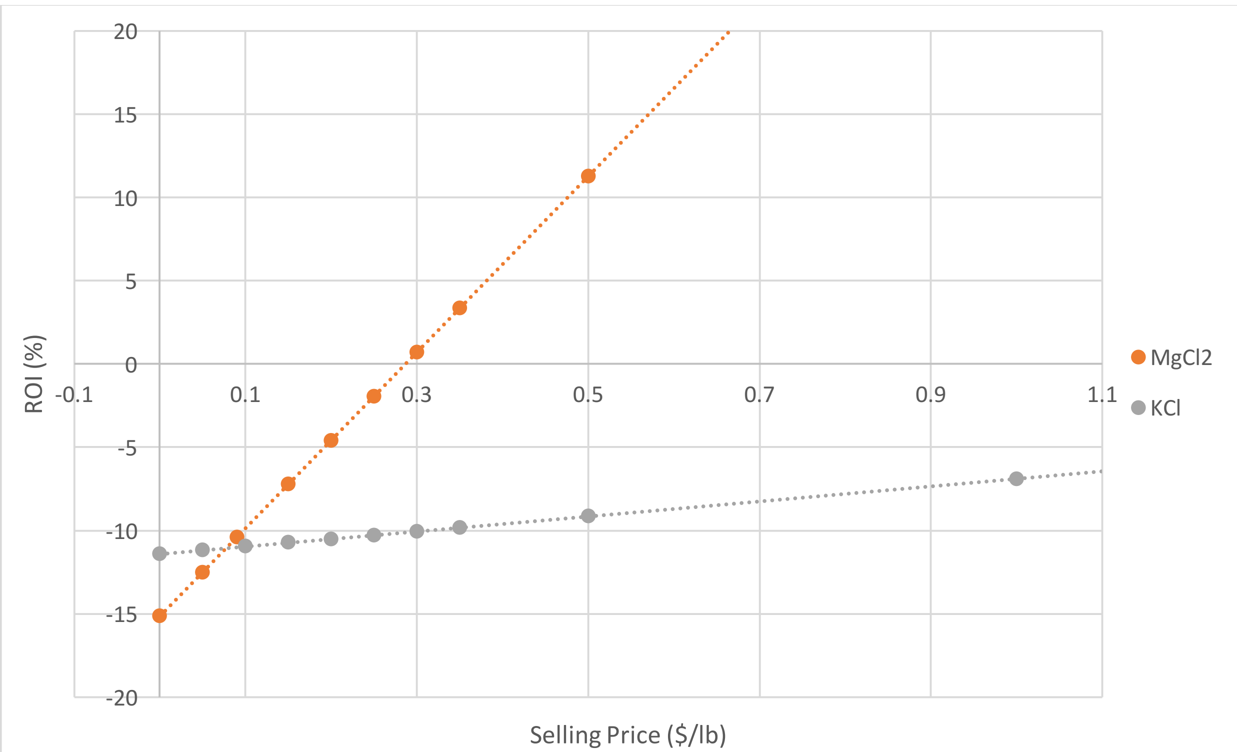


Figure 13. Relationship between system ROI and the selling prices of both MgCl2 and KCl

Figure 13 shows the relationship between our system's ROI and the selling price of both $MgCl_2$ and KCl. Notice that the intersection of both lines represents the current operating conditions of our system (yielding an ROI of -10.39%).

As shown in Figure 13, our system appears to be much more sensitive to fluctuating $MgCl_2$ prices than to KCl prices, which was expected; our proposed design specializes in sequestering $MgCl_2$

crystals from the brine wastewater feed, and as a result, significantly more $MgCl_2$ product is formed than the KCl byproduct.

Furthermore, Figure 13 also illustrates that, assuming all other costs and prices remain constant, our system will return a positive ROI (the system will thereby be making a *profit*) if $MgCl_2$ prices increase from the current \$0.09/lb to approximately \$0.30/lb. Conversely, KCl selling prices would need to increase to approximately \$2.6/lb to yield a positive ROI, assuming the $MgCl_2$ selling price is held constant at the current value of \$0.09/lb.

This sensitivity analysis leads to the important conclusion that small deviations in the $MgCl_2$ selling price could have major implications to our system's profitability; an increase of as little as \$0.05/lb could improve ROI by almost 5%, while a minor drop in its selling price could send profits plummeting.

Conclusions and Recommendations

Treating brine wastewater is a costly process but a necessary expense to comply with environmental restrictions and legal regulations. The equipment necessary to achieve secondary water recovery from the brine waste stream consist of steep fixed costs in addition to low relative revenue (from the recovered water) that inevitably reutnr negative ROI values; that said, these zero liquid discharge (ZLD) systems are more desirable than alternative options due to their ability to recover additional water from saturated brine waste, and the fact that they are less environmentally harmful than other treatment systems. Furthermore, ZLD systems do not isolate salts from the brine as additional sources of revenue, hence encountering a significant economic opportunity cost.

Our $MgCl_2$ Separation Unit recovers both $MgCl_2$ and KCl from wastewater at high purity, 98% and 60% respectively. Additionally, our system recovers 84% of the water from the brine inlet stream, with a flow rate of 149,571 kg/hr. Furthermore, overall equipment costs are considerably lower than the currently marketed GE Brine Concentrator, due to the lack of expensive units like vapor compressors; consequently, our system has an overall equipment cost of $6,668,130. While both our $MgCl_2$ Separation Unit and GE's Brine Concentrator have negative ROI values, our system experiences an increase in ROI by 22.53% as compared to GE's system, concluding that our designed unit is more cost-efficient.

Further research could be conducted to determine if the addition of sterile filtration and use of water for injection (WFI) to comply with FDA regulations would produce food/pharmaceutical grade $MgCl_2$ and KCl, increasing the overall revenue of the system and potentially converting the ROI from negative to positive (producing a unprecedentedly *profitable* treatment system.)

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References

[1] D. Rice, “On four continents, historic droughts wreak havoc,” USA Today, 27-Jul-2015. [Online]. Available at: http://www.usatoday.com/story/news/world/2015/07/24/historic-droughts-wreak-havoc-usa-brazil-n-korea/30513289/. [Accessed: 03-Apr-2016].

[2] “Desalination,” Wikipedia. [Online]. Available at: https://en.wikipedia.org/wiki/desalination#energy_consumption. [Accessed: 03-Apr-2016].

[3] “Oceans & Coasts,” Oceans & Coasts. [Online]. Available at: http://www.noaa.gov/ocean.html. [Accessed: 03-Apr-2016].

[4] “Desalination industry enjoys growth spurt as scarcity starts to bite,” - Global Water Intelligence. [Online]. Available at: http://www.globalwaterintel.com/desalination-industry-enjoys-growth-spurt-scarcity-starts-bite/. [Accessed: 03-Apr-2016].

[5] P. Rogers, “Nation's largest ocean desalination plant goes up near San Diego; Future of the California coast?,” - San Jose Mercury News, 29-May-2014. [Online]. Available at: http://www.mercurynews.com/science/ci_25859513/nations-largest-ocean-desalination-plant-goes-up-near. [Accessed: 03-Apr-2016].

[6] “U.S. Energy Information Administration - EIA - Independent Statistics and Analysis,” Electric Power Monthly, 25-Mar-2016. [Online]. Available at: http://www.eia.gov/electricity/monthly/epm_table_grapher.cfm?t=epmt_5_6_a. [Accessed: 03-Apr-2016].

[7] T. K. Poulson, “Strategic Alternatives for Brine Management in the Valley of the Sun,” Central Arizona Salinity Study , Jan. 2010.

[8] M. Kanellos, “Desalination Without the Ocean,” Greentech Media, 01-Apr-2011. [Online]. Available at: http://www.greentechmedia.com/articles/read/desalination-without-the-ocean. [Accessed: 03-Apr-2016].

[9] “Zero Liquid Discharge (ZLD),” Wastewater Treatment. [Online]. Available at: http://www.gewater.com/zero-liquid-discharge-zld.html#featuredproductssection. [Accessed: 03-Apr-2016].

[10] T. Casey, “Tiny Solar Thermal Power Plant Solves Gigantic Salt Problem,” CleanTechnica, 18-Feb-2014. [Online]. Available at: http://cleantechnica.com/2014/02/18/tiny-solar-power-desalination-plant-solves-big-salt-problem/. [Accessed: 03-Apr-2016].

[11] M. Svensson, “Desalination and the Environment: Options and considerations for brine disposal in inland and coastal locations,” rep.

[12] S. T. Lynch, B. Rohwer, Z. Erdal, and A. Lynch, “Brine/concentrate Management Strategies For Southern California,” Proceedings of the Water Environment Federation proc water environ fed, vol. 2005, no. 12, pp. 3563–3588, Jan. 2005.

[13] “‘A Cost Effective Way to Dispose of Salty Liquid Waste from Your Facility,’” “A Cost Effective Way to Dispose of Salty Liquid Waste from Your Facility.”.

[14] T. Rydners, “Design and Piloting of a Brine Minimization System for Concentrate Disposal,” New Water New Energy, 14-Dec-2011. [Online]. Available at: http://www.wrri.nmsu.edu/conf/conf11/pdf/abstract15_rynders.pdf.

[15] “Zero Liquid Discharge,” Southern California Regional Brine-Concentrate Management Study – Phase I. [Online]. Available at: http://www.usbr.gov/lc/socal/reports/brineconcentrate/6treatmentanddisposal_part2.pdf.

[16] D. A. Kramer, “Magnesium, its Alloys and Compounds ,” U.S. Geological Survey Open-File Report 01-341 .

[17] E. A. Sloan, "Getting Ahead of the Curve: Magnesium," Nutraceuticals World, 02-Jan-2012. [Online]. Available at: http://www.nutraceuticalsworld.com/issues/2012-01/view_trendsense/getting-ahead-of-the-curve-magnesium. [Accessed: 03-Apr-2016].
[18] A. Balakrishnan, "Road salt: Winter's $2.3 billion game changer," Cary Institute of Ecosystem Studies, 20-Feb-2015. [Online]. Available at: http://www.caryinstitute.org/newsroom/road-salt-winters-23-billion-game-changer. [Accessed: 03-Apr-2016].
[19] Mineral Commodity Summaries. 2015.
[20] Mineral Commodity Summaries. 2016.
[21] "Annual Average Snowfall (in)." [Online]. Available at: http://climatemaps.co.cc/. [Accessed: 10-Apr-2016].
[22] S. M. Jasinski, "Potash," 2013 Minerals Yearbook, Aug. 2015.
[23] "potassium chloride oral : Uses, Side Effects, Interactions, Pictures, Warnings & Dosing - WebMD," WebMD. [Online]. Available at: http://www.webmd.com/drugs/2/drug-676-7058/potassium-chloride-oral/potassiumextended-releasedispersibletablet-oral/details. [Accessed: 03-Apr-2016].
[24] "Zero Liquid Discharge Eliminate Liquid Discharge, Recover Valuable Process Water," Zero Liquid Discharge Eliminate Liquid Discharge, Recover Valuable Process Water. GE Water and Technologies, Bellevue, WA, 2009.
[25] W. E. Burke, W. A. Smith, and P. D. V. Manning, "Production of Magnesium Chloride," 16-Aug-1949.
[26] "Sodium Chloride Brine Tables for 60F," *ALKAR*. [Online]. Available at: http://www.alkar.com/download/pdf/sodium chloride brine tables for 60f.pdf.
[27] "Temperature Dependence of the Water Activity of Saturated Salt Solutions," *Decagon Devices*. [Online]. Available at: http://www.aqualab.com/assets/uploads/saturatedsalttempeaturedependence.pdf.
[28] "Solubility, saturation, equilibrium in saturated solutions, supersaturated solutions," Solubility, saturation, equilibrium in saturated solutions, supersaturated solutions. [Online]. Available at: http://www.800mainstreet.com/9/0009-004-solub.html. [Accessed: 03-Apr-2016].
[29] "San Diego, CA Electricity Rates," *Electricity Local*. [Online]. Available at: http://www.electricitylocal.com/states/california/san-diego/. [Accessed: 03-Apr-2016].
[30] W. D. Seider, J. D. Seader, D. R. Lewin, and S. Widagdo, *Product and Process Design Principles: Synthesis, Analysis and Design*, 3rd ed. s.l.: John Wiley and Sons Ltd, 2009.
[31] "CONTURBEX The worm/screen centrifuge," CONTURBEX The worm/screen centrifuge. Siebtechnik, Germany.
[32] "Carlsbad Desalination Project, San Diego, California," Water Technology. [Online]. Available at: http://www.water-technology.net/projects/carlsbaddesalination/. [Accessed: 03-Apr-2016].
[33] D. Potter, "Why Isn't Desalination the Answer to All California's Water Problems?," KQED Science, 18-Dec-2015. [Online]. Available at: http://ww2.kqed.org/science/2015/12/18/why-isnt-desalination-the-answer-to-all-californias-water-problems/. [Accessed: 03-Apr-2016].
[34] S. Masnoon and K. Glucina, "Desalination: Brine and Residual Management," *Water Research Foundation*, 19-Oct-2011. [Online]. Available at: http://www.waterrf.org/resources/lists/specialreports/attachments/2/gwrc_state_of_science_report_on_brine_and_residual_management.pdf.
[35] D. S. Brennan, "Poseidon proposes wetlands project in San Diego Bay," *The San Diego Union-Tribune*, 25-Nov-2012. [Online]. Available at: http://www.sandiegouniontribune.com/news/2012/nov/25/poseidon-proposes-wetlands-project-san-diego-bay/. [Accessed: 03-Apr-2016].

[36] C. Garrett, "Water Law & Policy Monitor - Latham & Watkins," *Water Law & Policy Monitor*, Oct-2014. [Online]. Available at: https://www.lw.com/thoughtleadership/carlsbad-desalination-project. [Accessed: 03-Apr-2016].
[37] Nova Chemicals, "Material Safety Data Sheet: Brine Solution", MSDS ID: NOVA-0087, 31-May-2015. Available at: http://www.novachem.com/Product%20Documents/BrineSolution_MSDS_AMER_CAEN.pdf. [Accessed: 03-Apr-2016].
[38] "Burks Installation and Operation Manual - Centrifugal Pumps," *Burks Installation and Operation Manual - Centrifugal Pumps*. A Crane Co. Compnay, Piqua, OH, 2006.
[39] "S.D. County Water Authority Board Approves Agreement to Purchase Carlsbad Desalination", Nov-30-2012. [Online-. Available at: http://carlsbaddesal.com/s-d-county-water-authority-board-approves-agreement. [Accessed:03-Apr-2016].

Appendix A: Formulae & Sample Calculations

Costing & Sizing

All costing equations from Product and Process Design Principles by Seider et al

Centrifugal Pumps:

$$C_B = \exp\{9.7171 - 0.6019 lnS + 0.0519(lnS)^2\}$$

S=[gpm-ft$^{0.5}$]

$C_p = F_T F_M C_B$

F_T- type factor, found in Table 22.20 [27]

F_M- material factor, found in Table 22.21 [27]

Electric Motors

$$C_B = \exp\{5.8259 + 0.13141 lnP_C + 0.053255(lnP_C)^2 + 0.028628(lnP_C)^3 - 0.0035549(lnP_C)^4\}$$

Where P_c=[hp]

$C_P = F_T C_B$

Centrifugal Compressors

$C_B = \exp\{7.5800 + 0.9 \ln P_C\}$
P_c= actual power consumption [hp]

$200<P_C<30{,}000$ hp

$C_P = F_T F_M C_B$

Draft-tubed baffled Crystallizer

$C_p = 28200 W^{0.63}$

W=[tons of crystals produced per day]

Flash Drum

Assume retention time, τ= 10 min

V= Vessel Volume

Q= Volumetric flow rate of liquid entering vessel

$0.5V \cong Q\tau$

Assume Aspect ratio of 2

$$\frac{L}{D_i} = 2$$

$$V \cong \frac{\pi D^2 L}{4}$$

$$t_p = \frac{P_d D_i}{2SE - 1.2P_d}$$

t_p-calculated thickness of vessel

If the above equation does not exceed the minimum thickness for the vessel operation temperature based on the below table, use the minimum value rather than the calculated value.

Vessel Inside Diameter (ft)	Minimum Wall Thickness (in.)
Up to 4	1/4
4–6	5/16
6–8	3/8
8–10	7/16
10–12	1/2

[27]

P_d- internal design pressure

$P_d = \exp\{0.6068 + 0.91615 lnP_o + 0.0015655(lnP_o)^2\}$

P_o- operation pressure [psig]

S- maximum allowable stress of shell material at design temp, [lb/in^2]

Temperature (°F)	Maximum Allowable Stress (psi)
−20 to 650	15,000
700	15,000
750	15,000
800	14,750
850	14,200
900	13,100

[27]

E-fractional weld efficiency

t_s = shell thickness, choose nearest standard thickness greater than t_p

$W = \pi(D_i + t_s)(L + 0.8D_i)t_s\rho_{metal}$

- Add the density of the stainless steel

$C_V = \exp\{7.0132 + 0.18225 lnW + 0.02297(lnW)^2\}$, Cost of vessel

$C_{PL} = 361.8(D_i)^{0.72960}(L)^{0.70684}$, Cost of platforms and ladders

$C_P = C_V F_M + C_{PL}$

Generalized Bare Module Cost Equation:

$C_{BM} = C_P[F_{BM}+(F_d F_p F_M - 1)]$

F_d- design factor; F_p-pressure factor

For pumps and compressors replace $F_d F_p F_M$ with $F_T F_M$

Pump Brake Efficiency

$$\eta = -0.316 + 0.24015\ln(Q) - 0.01199(\ln(Q))^2$$
$$Q[=]\,gpm$$

Pump Brake Work

$$P = \frac{QH\rho}{3330\eta}$$

$$\rho\frac{[=]lb}{gal}$$
$$H[=]ft$$
$$Q[=]gpm$$

Motor Efficiency

$$\eta = 0.80 + 0.319\ln(Pa) - 0.00812(\ln(Pa))^2$$

Power Consumption

$$P = \frac{Ph}{\eta}$$

Utilities

Steam

$$q = m\Delta H$$

$$m = \frac{\Delta H}{q}$$

where ΔH references the latent heat of vaporization of water, m references the mass of steam, q references the heat required

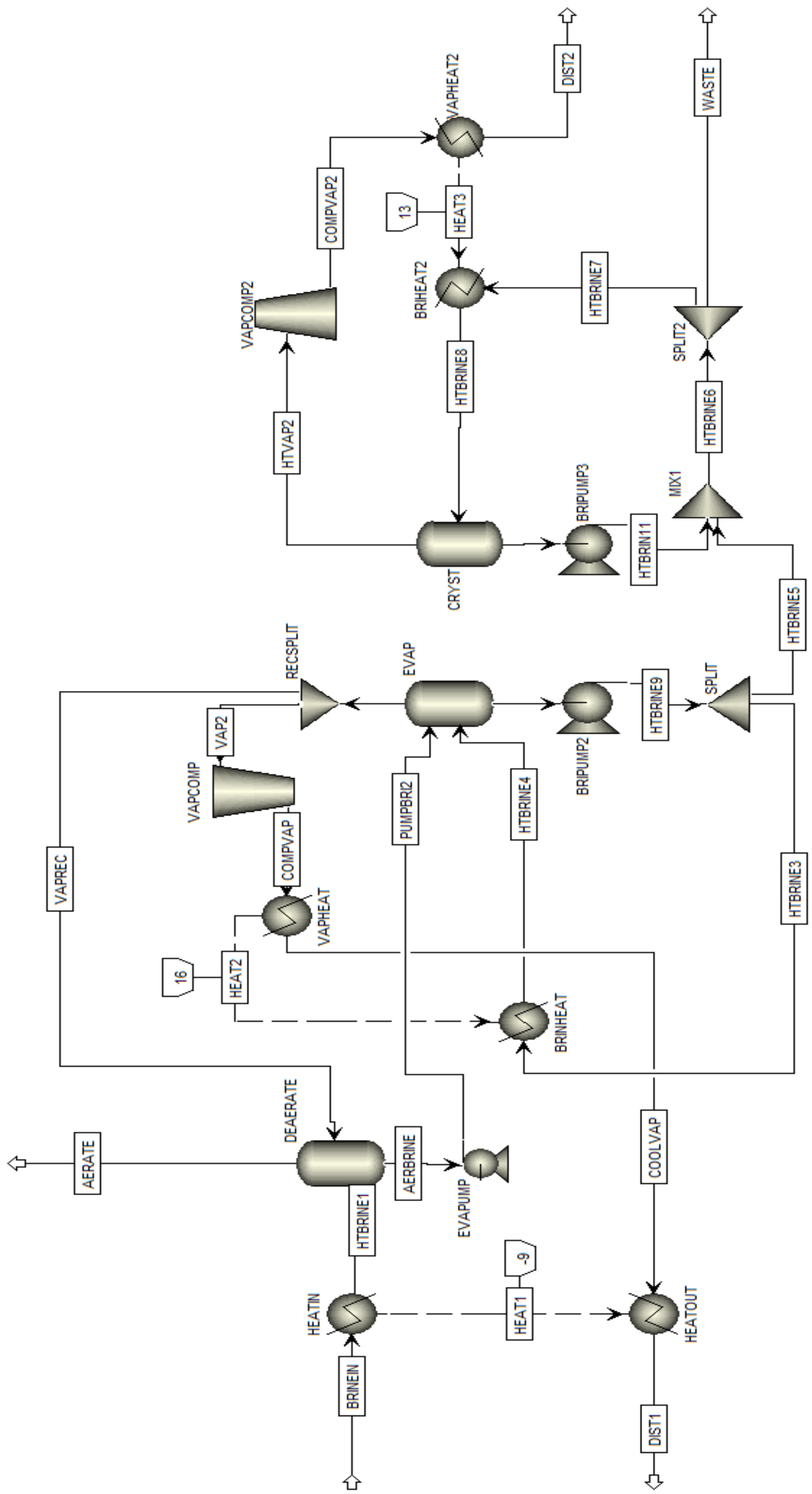
Cooling Water

$$q = m\mathrm{Cp}(\mathrm{To} - \mathrm{Ti})$$

Where Cp references the specific heat value of liquid water, T_o and T_i reference the outlet and inlet temperature of the cooling water in the heating exchange, respective, and m references the mass of cooling water needed.

Appendix B: Aspen Plus Input Summary, Block & Stream Reports, Simulation Results

<u>GE Brine Concentrator Simulation Flowsheet</u>



GE Brine Concentrator Input Summary

```
;Input Summary created by Aspen Plus Rel. 34.0 at 13:26:01 Sun Apr 3, 2016
;Directory S:\Senior Design Rempel\Flowsheets\GEFINALBRINECONCENTRATOR Filename
C:\Users\arempel\AppData\Local\Temp\~ape79f.txt
;
DYNAMICS
   DYNAMICS RESULTS=ON

IN-UNITS MET VOLUME-FLOW='cum/hr' ENTHALPY-FLO='Gcal/hr' &
      HEAT-TRANS-C='kcal/hr-sqm-K' PRESSURE=bar TEMPERATURE=C &
      VOLUME=cum DELTA-T=C HEAD=meter MOLE-DENSITY='kmol/cum' &
      MASS-DENSITY='kg/cum' MOLE-ENTHALP='kcal/mol' &
      MASS-ENTHALP='kcal/kg' HEAT=Gcal MOLE-CONC='mol/l' &
      PDROP=bar

DEF-STREAMS CONVEN ALL
MODEL-OPTION
DESCRIPTION "
   Electrolytes Simulation with Metric Units :
   C, bar, kg/hr, kmol/hr, Gcal/hr, cum/hr.

   Property Method: ELECNRTL

   Flow basis for input: Mass

   Stream report composition: Mass flow
   "
DATABANKS 'APV88 ASPENPCD' / 'APV88 AQUEOUS' / 'APV88 SOLIDS' &
      / 'APV88 INORGANIC' / 'APV88 PURE32'

PROP-SOURCES 'APV88 ASPENPCD' / 'APV88 AQUEOUS' / 'APV88 SOLIDS' &
      / 'APV88 INORGANIC' / 'APV88 PURE32'

COMPONENTS
   H2O H2O /
   SODIU-01 NACL /
   NA+ NA+ /
   "NACL(S)" NACL /
   CL- CL- /
   CO2 CO2 /
   O2 O2 /
   MAGNE-01 MGCL2 /
   MG++ MG+2 /
   "MGCL2(S)" MGCL2 /
   KCL KCL /
   K+ K+ /
   "KCL(S)" KCL /
   SALT1 "MGCL2*4W" /
   SALT2 "MGCL2*2W" /
   SALT3 "MGCL2*W" /
   SALT4 "MGCL2*6W"

CISOLID-COMPS "NACL(S)" "MGCL2(S)" "KCL(S)" SALT1 SALT2 SALT3 &
      SALT4
HENRY-COMPS GLOBAL CO2 O2
HENRY-COMPS NONCOMP O2 CO2
SOLVE
   RUN-MODE MODE=SIM

CHEMISTRY GLOBAL
```

```
    PARAM GAMMA-BASIS=UNSYMMETRIC
    DISS MAGNE-01 MG++ 1 / CL- 2
    DISS KCL CL- 1 / K+ 1
    DISS SODIU-01 CL- 1 / NA+ 1
    SALT "KCL(S)" CL- 1 / K+ 1
    SALT SALT4 MG++ 1 / CL- 2 / H2O 6
    SALT SALT3 H2O 1 / MG++ 1 / CL- 2
    SALT SALT2 MG++ 1 / H2O 2 / CL- 2
    SALT "MGCL2(S)" MG++ 1 / CL- 2
    SALT "NACL(S)" CL- 1 / NA+ 1
    SALT SALT1 MG++ 1 / CL- 2 / H2O 4
    K-SALT "KCL(S)" A=-226.337708 B=2268.976074 C=42.205101  &
        D=-0.092909
    K-SALT SALT4 A=-1410.412964 B=44338.910156 C=237.735703  &
        D=-0.329073
    K-SALT "NACL(S)" A=-203.587494 B=4381.175781 C=35.875179  &
        D=-0.067216

FLOWSHEET
    BLOCK DEAERATE IN=HTBRINE1 VAPREC OUT=AERATE AERBRINE
    BLOCK VAPCOMP IN=VAP2 OUT=COMPVAP
    BLOCK EVAP IN=HTBRINE4 PUMPBRI2 OUT=VAP1 HTBRINE2
    BLOCK SPLIT IN=HTBRINE9 OUT=HTBRINE5 HTBRINE3
    BLOCK HEATIN IN=BRINEIN OUT=HTBRINE1 HEAT1
    BLOCK HEATOUT IN=COOLVAP HEAT1 OUT=DIST1
    BLOCK VAPHEAT IN=COMPVAP OUT=COOLVAP HEAT2
    BLOCK BRINHEAT IN=HTBRINE3 HEAT2 OUT=HTBRINE4
    BLOCK CRYST IN=HTBRINE8 OUT=HTVAP2 HTBRIN10
    BLOCK BRIHEAT2 IN=HTBRINE7 HEAT3 OUT=HTBRINE8
    BLOCK SPLIT2 IN=HTBRINE6 OUT=HTBRINE7 WASTE
    BLOCK MIX1 IN=HTBRINE5 HTBRIN11 OUT=HTBRINE6
    BLOCK VAPCOMP2 IN=HTVAP2 OUT=COMPVAP2
    BLOCK VAPHEAT2 IN=COMPVAP2 OUT=DIST2 HEAT3
    BLOCK RECSPLIT IN=VAP1 OUT=VAP2 VAPREC
    BLOCK EVAPUMP IN=AERBRINE OUT=PUMPBRI2
    BLOCK BRIPUMP2 IN=HTBRINE2 OUT=HTBRINE9
    BLOCK BRIPUMP3 IN=HTBRIN10 OUT=HTBRIN11

PROPERTIES ELECNRTL HENRY-COMPS=GLOBAL CHEMISTRY=GLOBAL  &
        TRUE-COMPS=YES

PROP-DATA HENRY-1
    IN-UNITS MET VOLUME-FLOW='cum/hr' ENTHALPY-FLO='Gcal/hr'  &
        HEAT-TRANS-C='kcal/hr-sqm-K' PRESSURE=bar TEMPERATURE=C  &
        VOLUME=cum DELTA-T=C HEAD=meter MOLE-DENSITY='kmol/cum'  &
        MASS-DENSITY='kg/cum' MOLE-ENTHALP='kcal/mol'  &
        MASS-ENTHALP='kcal/kg' HEAT=Gcal MOLE-CONC='mol/l'  &
        PDROP=bar
    PROP-LIST HENRY
    BPVAL O2 H2O 144.4080745 -7775.060000 -18.39740000  &
        -9.4435400E-3 .8500000000 74.85000000 0.0
    BPVAL CO2 H2O 159.1996745 -8477.711000 -21.95743000  &
        5.78074800E-3 -.1500000000 226.8500000 0.0

PROP-DATA NRTL-1
    IN-UNITS MET VOLUME-FLOW='cum/hr' ENTHALPY-FLO='Gcal/hr'  &
        HEAT-TRANS-C='kcal/hr-sqm-K' PRESSURE=bar TEMPERATURE=C  &
        VOLUME=cum DELTA-T=C HEAD=meter MOLE-DENSITY='kmol/cum'  &
        MASS-DENSITY='kg/cum' MOLE-ENTHALP='kcal/mol'  &
        MASS-ENTHALP='kcal/kg' HEAT=Gcal MOLE-CONC='mol/l'  &
        PDROP=bar
    PROP-LIST NRTL
```

```
  BPVAL H2O CO2 10.06400000 -3268.135000 .2000000000 0.0 0.0  &
    0.0 0.0 200.0000000
  BPVAL CO2 H2O 10.06400000 -3268.135000 .2000000000 0.0 0.0  &
    0.0 0.0 200.0000000

PROP-DATA VLCLK-1
  IN-UNITS MET VOLUME-FLOW='cum/hr' ENTHALPY-FLO='Gcal/hr'  &
    HEAT-TRANS-C='kcal/hr-sqm-K' PRESSURE=bar TEMPERATURE=C  &
    VOLUME=cum DELTA-T=C HEAD=meter MOLE-DENSITY='kmol/cum'  &
    MASS-DENSITY='kg/cum' MOLE-ENTHALP='kcal/mol'  &
    MASS-ENTHALP='kcal/kg' HEAT=Gcal MOLE-CONC='mol/l'  &
    PDROP=bar
  PROP-LIST VLCLK
  BPVAL NA+ CL- 15.14218000 25.28718000
  BPVAL K+ CL- 25.87263000 23.41131000

PROP-DATA GMELCC-1
  IN-UNITS MET VOLUME-FLOW='cum/hr' ENTHALPY-FLO='Gcal/hr'  &
    HEAT-TRANS-C='kcal/hr-sqm-K' PRESSURE=bar TEMPERATURE=C  &
    VOLUME=cum DELTA-T=C HEAD=meter MOLE-DENSITY='kmol/cum'  &
    MASS-DENSITY='kg/cum' MOLE-ENTHALP='kcal/mol'  &
    MASS-ENTHALP='kcal/kg' HEAT=Gcal MOLE-CONC='mol/l'  &
    PDROP=bar
  PROP-LIST GMELCC
  PPVAL H2O ( NA+ CL- ) 5.980196000
  PPVAL ( NA+ CL- ) H2O -3.789168000
  PPVAL CO2 ( NA+ CL- ) 11.67418000
  PPVAL ( NA+ CL- ) CO2 2.879981000
  PPVAL H2O ( MG++ CL- ) 7.872000000
  PPVAL ( MG++ CL- ) H2O -4.147000000
  PPVAL H2O ( K+ CL- ) 6.849537000
  PPVAL ( K+ CL- ) H2O -4.060085000
  PPVAL ( NA+ CL- ) ( K+ CL- ) 1.360000000
  PPVAL ( K+ CL- ) ( NA+ CL- ) -1.023000000

PROP-DATA GMELCD-1
  IN-UNITS MET VOLUME-FLOW='cum/hr' ENTHALPY-FLO='Gcal/hr'  &
    HEAT-TRANS-C='kcal/hr-sqm-K' PRESSURE=bar TEMPERATURE=C  &
    VOLUME=cum DELTA-T=C HEAD=meter MOLE-DENSITY='kmol/cum'  &
    MASS-DENSITY='kg/cum' MOLE-ENTHALP='kcal/mol'  &
    MASS-ENTHALP='kcal/kg' HEAT=Gcal MOLE-CONC='mol/l'  &
    PDROP=bar
  PROP-LIST GMELCD
  PPVAL H2O ( NA+ CL- ) 841.5181000
  PPVAL ( NA+ CL- ) H2O -216.3646000
  PPVAL CO2 ( NA+ CL- ) -2251.456000
  PPVAL ( NA+ CL- ) CO2 1558.967000
  PPVAL H2O ( MG++ CL- ) 1132.700000
  PPVAL ( MG++ CL- ) H2O -412.6000000
  PPVAL H2O ( K+ CL- ) 402.9818000
  PPVAL ( K+ CL- ) H2O -30.93534000
  PPVAL ( NA+ CL- ) ( K+ CL- ) -440.5000000
  PPVAL ( K+ CL- ) ( NA+ CL- ) 331.4000000

PROP-DATA GMELCE-1
  IN-UNITS MET VOLUME-FLOW='cum/hr' ENTHALPY-FLO='Gcal/hr'  &
    HEAT-TRANS-C='kcal/hr-sqm-K' PRESSURE=bar TEMPERATURE=C  &
    VOLUME=cum DELTA-T=C HEAD=meter MOLE-DENSITY='kmol/cum'  &
    MASS-DENSITY='kg/cum' MOLE-ENTHALP='kcal/mol'  &
    MASS-ENTHALP='kcal/kg' HEAT=Gcal MOLE-CONC='mol/l'  &
    PDROP=bar
  PROP-LIST GMELCE
```

```
  PPVAL H2O ( NA+ CL- ) 7.433500000
  PPVAL ( NA+ CL- ) H2O -1.100418000
  PPVAL CO2 ( NA+ CL- ) 16.77602000
  PPVAL ( NA+ CL- ) CO2 21.13943000
  PPVAL H2O ( MG++ CL- ) -5.962000000
  PPVAL ( MG++ CL- ) H2O -.0860000000
  PPVAL H2O ( K+ CL- ) .2065224000
  PPVAL ( K+ CL- ) H2O 1.429560000

PROP-DATA GMELCN-1
  IN-UNITS MET VOLUME-FLOW='cum/hr' ENTHALPY-FLO='Gcal/hr'  &
    HEAT-TRANS-C='kcal/hr-sqm-K' PRESSURE=bar TEMPERATURE=C  &
    VOLUME=cum DELTA-T=C HEAD=meter MOLE-DENSITY='kmol/cum'  &
    MASS-DENSITY='kg/cum' MOLE-ENTHALP='kcal/mol'  &
    MASS-ENTHALP='kcal/kg' HEAT=Gcal MOLE-CONC='mol/l'  &
    PDROP=bar
  PROP-LIST GMELCN
  PPVAL H2O ( NA+ CL- ) .2000000000
  PPVAL CO2 ( NA+ CL- ) .2000000000

STREAM BRINEIN
  SUBSTREAM MIXED TEMP=25. PRES=2. VOLUME-FLOW=7570000. <l/day>
  MASS-FRAC H2O 0.765 / NA+ 0. / "NACL(S)" 0.201 / CL-  &
    0. / CO2 0.002 / O2 0.002 / MG++ 0. / "MGCL2(S)"  &
    0.03
DEF-STREAMS HEAT HEAT1
DEF-STREAMS HEAT HEAT2
DEF-STREAMS HEAT HEAT3
BLOCK MIX1 MIXER
  PARAM
BLOCK RECSPLIT FSPLIT
  FRAC VAPREC 0.1
BLOCK SPLIT FSPLIT
  FRAC HTBRINE3 0.5
BLOCK SPLIT2 FSPLIT
  FRAC WASTE 0.2
BLOCK BRIHEAT2 HEATER
  PARAM PRES=1. DPPARMOPT=NO
BLOCK BRINHEAT HEATER
  PARAM PRES=3. NPHASE=2 DPPARMOPT=NO
  BLOCK-OPTION FREE-WATER=NO
BLOCK HEATIN HEATER
  PARAM TEMP=94. VFRAC=0.01 DPPARMOPT=NO
BLOCK HEATOUT HEATER
  PARAM PRES=2. DPPARMOPT=NO
BLOCK VAPHEAT HEATER
  PARAM PRES=3. VFRAC=0.4 DPPARMOPT=NO
BLOCK VAPHEAT2 HEATER
  PARAM PRES=1. VFRAC=0.05 DPPARMOPT=NO
BLOCK CRYST FLASH2
  PARAM PRES=0.5 VFRAC=0.1
BLOCK DEAERATE FLASH2
  PARAM PRES=0. VFRAC=0.1
BLOCK EVAP FLASH2
  PARAM PRES=0.5 VFRAC=0.2
BLOCK BRIPUMP2 PUMP
  PARAM PRES=4.
BLOCK BRIPUMP3 PUMP
  PARAM PRES=4.
BLOCK EVAPUMP PUMP
  PARAM DELP=1.
BLOCK VAPCOMP COMPR
```

```
   PARAM TYPE=ISENTROPIC PRES=4. SEFF=0.75 MEFF=0.75  &
      SB-MAXIT=30 SB-TOL=0.0001
BLOCK VAPCOMP2 COMPR
   PARAM TYPE=ISENTROPIC DELP=1. SEFF=0.75 MEFF=0.75  &
      SB-MAXIT=30 SB-TOL=0.0001
EO-CONV-OPTI

STREAM-REPOR MOLEFLOW MASSFLOW
```

GE Brine Concentrator Block Reports

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 BLOCK:  BRIHEAT2 MODEL: HEATER
 ------------------------------
   INLET STREAM:          HTBRINE7
   INLET HEAT STREAM:     HEAT3
   OUTLET STREAM:         HTBRINE8
   PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES


                 ***  MASS AND ENERGY BALANCE  ***
                           IN             OUT        RELATIVE DIFF.
    TOTAL BALANCE
      MOLE(KMOL/HR )          13733.9        13651.1       0.602529E-02
      MASS(KG/HR   )          326382.        326384.      -0.470112E-05
      ENTHALPY(GCAL/HR )      -914.228       -914.232       0.420320E-05

                 ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E           0.231054E-04  KG/HR
    PRODUCT STREAMS CO2E        0.231054E-04  KG/HR
    NET STREAMS CO2E PRODUCTION -0.180565E-11  KG/HR
    UTILITIES CO2E PRODUCTION     0.00000      KG/HR
    TOTAL CO2E PRODUCTION       -0.180565E-11  KG/HR

                    ***  INPUT DATA  ***
   TWO    PHASE  PQ  FLASH
   SPECIFIED PRESSURE                 BAR                  1.00000
   DUTY FROM INLET HEAT STREAM(S)      GCAL/HR                 12.6669
   MAXIMUM NO. ITERATIONS                                  30
   CONVERGENCE TOLERANCE                                     0.000100000



                    ***  RESULTS  ***
   OUTLET TEMPERATURE    C                          109.21
   OUTLET PRESSURE       BAR                       1.0000
   OUTLET VAPOR FRACTION                            0.74934E-01



   V-L PHASE EQUILIBRIUM :

     COMP            F(I)          X(I)          Y(I)          K(I)
     H2O            0.81161       0.80816       1.0000        1.2374
     NA+            0.70564E-01   0.70032E-01   0.0000        0.0000
     CL-            0.10206       0.10454       0.0000        0.0000
     CO2            0.42107E-10   0.11646E-12   0.56805E-09   4877.7
     O2             0.64971E-13   0.36327E-17   0.87866E-12   0.24187E+06
     MG++           0.15761E-01   0.17267E-01   0.0000        0.0000
```

BLOCK: BRINHEAT MODEL: HEATER

INLET STREAM: HTBRINE3
INLET HEAT STREAM: HEAT2
OUTLET STREAM: HTBRINE4
PROPERTY OPTION SET: ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
HENRY-COMPS ID: GLOBAL
CHEMISTRY ID: GLOBAL   - TRUE SPECIES

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(KMOL/HR)	4796.24	4655.30	0.293849E-01
MASS(KG/HR)	103536.	103537.	-0.475180E-05
ENTHALPY(GCAL/HR)	-297.157	-297.159	0.622908E-05

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E   0.288561E-04 KG/HR
PRODUCT STREAMS CO2E   0.288562E-04 KG/HR
NET STREAMS CO2E PRODUCTION 0.163056E-09 KG/HR
UTILITIES CO2E PRODUCTION   0.00000   KG/HR
TOTAL CO2E PRODUCTION   0.163056E-09 KG/HR

*** INPUT DATA ***

TWO   PHASE PQ FLASH
SPECIFIED PRESSURE   BAR   3.00000
DUTY FROM INLET HEAT STREAM(S)   GCAL/HR   16.3954
MAXIMUM NO. ITERATIONS   30
CONVERGENCE TOLERANCE   0.000100000

*** RESULTS ***

OUTLET TEMPERATURE   C   145.36
OUTLET PRESSURE   BAR   3.0000
OUTLET VAPOR FRACTION   0.30687

V-L PHASE EQUILIBRIUM :

COMP	F(I)	X(I)	Y(I)	K(I)
H2O	0.81040	0.80236	1.0000	1.2463
NA+	0.78867E-01	0.74340E-01	0.0000	0.0000
CL-	0.10011	0.10698	0.0000	0.0000
CO2	0.14178E-09	0.25878E-12	0.49142E-09	1899.0
O2	0.21895E-12	0.89426E-17	0.75981E-12	84965.
MG++	0.10623E-01	0.16321E-01	0.0000	0.0000

BLOCK: BRIPUMP2 MODEL: PUMP

INLET STREAM: HTBRINE2
OUTLET STREAM: HTBRINE9
PROPERTY OPTION SET: ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
HENRY-COMPS ID: GLOBAL
CHEMISTRY ID: GLOBAL   - TRUE SPECIES

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(KMOL/HR)	9592.51	9592.48	0.283084E-05

MASS(KG/HR) 207073. 207073. 0.00000
ENTHALPY(GCAL/HR) -627.124 -627.105 -0.299210E-04

*** CO2 EQUIVALENT SUMMARY ***
FEED STREAMS CO2E 0.577121E-04 KG/HR
PRODUCT STREAMS CO2E 0.577121E-04 KG/HR
NET STREAMS CO2E PRODUCTION 0.00000 KG/HR
UTILITIES CO2E PRODUCTION 0.00000 KG/HR
TOTAL CO2E PRODUCTION 0.00000 KG/HR

*** INPUT DATA ***
OUTLET PRESSURE BAR 4.00000
DRIVER EFFICIENCY 1.00000

FLASH SPECIFICATIONS:
LIQUID PHASE CALCULATION
NO FLASH PERFORMED
MAXIMUM NUMBER OF ITERATIONS 30
TOLERANCE 0.000100000

*** RESULTS ***
VOLUMETRIC FLOW RATE CUM/HR 167.051
PRESSURE CHANGE BAR 3.50000
NPSH AVAILABLE METER 0.0
FLUID POWER KW 16.2411
BRAKE POWER KW 21.7555
ELECTRICITY KW 21.7555
PUMP EFFICIENCY USED 0.74653
NET WORK REQUIRED KW 21.7555
HEAD DEVELOPED METER 28.7921

BLOCK: BRIPUMP3 MODEL: PUMP

INLET STREAM: HTBRIN10
OUTLET STREAM: HTBRIN11
PROPERTY OPTION SET: ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
HENRY-COMPS ID: GLOBAL
CHEMISTRY ID: GLOBAL - TRUE SPECIES

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***********************************************************************
*                                                                     *
*    AT LEAST ONE OF THE INLET OR OUTLET STREAMS                      *
*    IS NOT IN CHARGE BALANCE                                         *
*                                                                     *
***********************************************************************
```

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(KMOL/HR)	12371.7	12371.7	-0.366043E-06
MASS(KG/HR)	304441.	304441.	0.00000
ENTHALPY(GCAL/HR)	-845.090	-845.066	-0.288685E-04

*** CO2 EQUIVALENT SUMMARY ***
FEED STREAMS CO2E 0.257253E-07 KG/HR
PRODUCT STREAMS CO2E 0.257258E-07 KG/HR
NET STREAMS CO2E PRODUCTION 0.597103E-12 KG/HR
UTILITIES CO2E PRODUCTION 0.00000 KG/HR
TOTAL CO2E PRODUCTION 0.597103E-12 KG/HR

```
        *** INPUT DATA ***
 OUTLET PRESSURE  BAR                        4.00000
 DRIVER EFFICIENCY                           1.00000

 FLASH SPECIFICATIONS:
 LIQUID PHASE CALCULATION
 NO FLASH PERFORMED
 MAXIMUM NUMBER OF ITERATIONS                   30
 TOLERANCE                                 0.000100000

        *** RESULTS ***
 VOLUMETRIC FLOW RATE  CUM/HR                 224.145
 PRESSURE CHANGE  BAR                        3.50000
 NPSH AVAILABLE   METER                        0.0
 FLUID POWER  KW                            21.7918
 BRAKE POWER  KW                            28.3174
 ELECTRICITY  KW                            28.3174
 PUMP EFFICIENCY USED                        0.76956
 NET WORK REQUIRED  KW                       28.3174
 HEAD DEVELOPED METER                        26.2768

BLOCK:  CRYST    MODEL: FLASH2
------------------------------
 INLET STREAM:          HTBRINE8
 OUTLET VAPOR STREAM:   HTVAP2
 OUTLET LIQUID STREAM:  HTBRIN10
 PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:        GLOBAL
 CHEMISTRY ID:          GLOBAL   - TRUE SPECIES


  ***********************************************************************
  *                                                                     *
  *    AT LEAST ONE OF THE INLET OR OUTLET STREAMS                      *
  *    IS NOT IN CHARGE BALANCE                                         *
  *                                                                     *
  ***********************************************************************


       *** MASS AND ENERGY BALANCE ***
                         IN          OUT       RELATIVE DIFF.
 TOTAL BALANCE
   MOLE(KMOL/HR )       13651.1      13589.6      0.450248E-02
   MASS(KG/HR   )       326384.      326384.      0.665212E-13
   ENTHALPY(GCAL/HR )   -914.232     -914.861     0.687176E-03

       *** CO2 EQUIVALENT SUMMARY ***
 FEED STREAMS CO2E            0.231054E-04  KG/HR
 PRODUCT STREAMS CO2E         0.231054E-04  KG/HR
 NET STREAMS CO2E PRODUCTION  0.00000       KG/HR
 UTILITIES CO2E PRODUCTION    0.00000       KG/HR
 TOTAL CO2E PRODUCTION        0.00000       KG/HR

        *** INPUT DATA ***
 TWO    PHASE  PV  FLASH
 SPECIFIED PRESSURE    BAR                   0.50000
 VAPOR FRACTION                              0.100000
 MAXIMUM NO. ITERATIONS                        30
 CONVERGENCE TOLERANCE                         0.000100000

        *** RESULTS ***
 OUTLET TEMPERATURE    C                     89.814
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```
OUTLET PRESSURE       BAR                            0.50000
HEAT DUTY             GCAL/HR                       -0.62867
VAPOR FRACTION                                     0.10000


V-L PHASE EQUILIBRIUM :

  COMP        F(I)         X(I)         Y(I)         K(I)
  H2O         0.82253      0.81204      1.0000       1.2315
  NA+         0.64787E-01  0.67105E-01  0.0000       0.0000
  CL-         0.96707E-01  0.10293      0.0000       0.0000
  CO2         0.42674E-10  0.53324E-13  0.43056E-09  8074.5
  O2          0.65845E-13  0.14079E-17  0.66509E-12  0.47239E+06
  MG++        0.15973E-01  0.17927E-01  0.0000       0.0000

BLOCK:  DEAERATE MODEL: FLASH2
------------------------------
 INLET STREAMS:        HTBRINE1    VAPREC
 OUTLET VAPOR STREAM:   AERATE
 OUTLET LIQUID STREAM:  AERBRINE
 PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:       GLOBAL
 CHEMISTRY ID:         GLOBAL   - TRUE SPECIES

             ***  MASS AND ENERGY BALANCE  ***
                       IN           OUT       RELATIVE DIFF.
 TOTAL BALANCE
   MOLE(KMOL/HR )       8089.01       8089.01      -0.337308E-15
   MASS(KG/HR   )       160087.       160087.      -0.109080E-14
   ENTHALPY(GCAL/HR )   -519.355      -515.198     -0.800381E-02

             ***  CO2 EQUIVALENT SUMMARY ***
 FEED STREAMS CO2E            311.862      KG/HR
 PRODUCT STREAMS CO2E         311.862      KG/HR
 NET STREAMS CO2E PRODUCTION   0.00000      KG/HR
 UTILITIES CO2E PRODUCTION     0.00000      KG/HR
 TOTAL CO2E PRODUCTION         0.00000      KG/HR

               ***  INPUT DATA  ***
 TWO    PHASE  PV  FLASH
 PRESSURE DROP         BAR                         0.0
 VAPOR FRACTION                                  0.100000
 MAXIMUM NO. ITERATIONS                              30
 CONVERGENCE TOLERANCE                              0.000100000

               ***  RESULTS  ***
 OUTLET TEMPERATURE    C                            87.221
 OUTLET PRESSURE       BAR                          0.50000
 HEAT DUTY             GCAL/HR                      4.1568
 VAPOR FRACTION                                   0.10000


 V-L PHASE EQUILIBRIUM :

  COMP        F(I)         X(I)         Y(I)         K(I)
  H2O         0.84711      0.83243      0.97920      1.1763
  NA+         0.66295E-01  0.73661E-01  0.0000       0.0000
  CL-         0.78442E-01  0.87158E-01  0.0000       0.0000
  CO2         0.87603E-03  0.59299E-06  0.87549E-02  14764.
  O2          0.12048E-02  0.30554E-07  0.12048E-01  0.39430E+06
```

MG++	0.60736E-02	0.67485E-02	0.0000	0.0000

BLOCK: EVAP MODEL: FLASH2

INLET STREAMS: HTBRINE4 PUMPBRI2
OUTLET VAPOR STREAM: VAP1
OUTLET LIQUID STREAM: HTBRINE2
PROPERTY OPTION SET: ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
HENRY-COMPS ID: GLOBAL
CHEMISTRY ID: GLOBAL - TRUE SPECIES

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(KMOL/HR)	11935.4	11904.9	0.255716E-02
MASS(KG/HR)	248731.	248731.	-0.929053E-13
ENTHALPY(GCAL/HR)	-766.306	-759.594	-0.875956E-02

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.190019	KG/HR
PRODUCT STREAMS CO2E	0.190019	KG/HR
NET STREAMS CO2E PRODUCTION	0.00000	KG/HR
UTILITIES CO2E PRODUCTION	0.00000	KG/HR
TOTAL CO2E PRODUCTION	0.00000	KG/HR

*** INPUT DATA ***

TWO PHASE PV FLASH

SPECIFIED PRESSURE	BAR	0.50000
VAPOR FRACTION		0.20000
MAXIMUM NO. ITERATIONS		30
CONVERGENCE TOLERANCE		0.000100000

*** RESULTS ***

OUTLET TEMPERATURE	C	89.328
OUTLET PRESSURE	BAR	0.50000
HEAT DUTY	GCAL/HR	6.7125
VAPOR FRACTION		0.20000

V-L PHASE EQUILIBRIUM :

COMP	F(I)	X(I)	Y(I)	K(I)
H2O	0.84386	0.81039	1.0000	1.2340
NA+	0.65391E-01	0.78870E-01	0.0000	0.0000
CL-	0.82298E-01	0.10012	0.0000	0.0000
CO2	0.37148E-06	0.14177E-09	0.18666E-05	13167.
O2	0.19138E-07	0.21895E-12	0.96193E-07	0.43934E+06
MG++	0.84539E-02	0.10623E-01	0.0000	0.0000

BLOCK: EVAPUMP MODEL: PUMP

INLET STREAM: AERBRINE
OUTLET STREAM: PUMPBRI2
PROPERTY OPTION SET: ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
HENRY-COMPS ID: GLOBAL
CHEMISTRY ID: GLOBAL - TRUE SPECIES

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(KMOL/HR)	7280.11	7280.11	0.00000

MASS(KG/HR) 145194. 145194. -0.400895E-15
ENTHALPY(GCAL/HR) -469.151 -469.147 -0.892457E-05

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.189990	KG/HR
PRODUCT STREAMS CO2E	0.189990	KG/HR
NET STREAMS CO2E PRODUCTION	0.00000	KG/HR
UTILITIES CO2E PRODUCTION	0.00000	KG/HR
TOTAL CO2E PRODUCTION	0.00000	KG/HR

*** INPUT DATA ***

PRESSURE CHANGE BAR	1.00000
DRIVER EFFICIENCY	1.00000

FLASH SPECIFICATIONS:
LIQUID PHASE CALCULATION
NO FLASH PERFORMED

MAXIMUM NUMBER OF ITERATIONS	30
TOLERANCE	0.000100000

*** RESULTS ***

VOLUMETRIC FLOW RATE CUM/HR	125.445
PRESSURE CHANGE BAR	1.00000
NPSH AVAILABLE METER	0.0
FLUID POWER KW	3.48458
BRAKE POWER KW	4.82564
ELECTRICITY KW	4.82564
PUMP EFFICIENCY USED	0.72210
NET WORK REQUIRED KW	4.82564
HEAD DEVELOPED METER	8.81016

BLOCK: HEATIN MODEL: HEATER

INLET STREAM: BRINEIN
OUTLET STREAM: HTBRINE1
OUTLET HEAT STREAM: HEAT1
PROPERTY OPTION SET: ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
HENRY-COMPS ID: GLOBAL
CHEMISTRY ID: GLOBAL - TRUE SPECIES

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(KMOL/HR)	7857.77	7857.77	0.00000
MASS(KG/HR)	155921.	155921.	-0.186657E-15
ENTHALPY(GCAL/HR)	-514.739	-514.739	0.220863E-15

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	311.843	KG/HR
PRODUCT STREAMS CO2E	311.843	KG/HR
NET STREAMS CO2E PRODUCTION	0.00000	KG/HR
UTILITIES CO2E PRODUCTION	0.00000	KG/HR
TOTAL CO2E PRODUCTION	0.00000	KG/HR

*** INPUT DATA ***

TWO PHASE TV FLASH

SPECIFIED TEMPERATURE	C	94.0000
VAPOR FRACTION		0.0100000
MAXIMUM NO. ITERATIONS		30
CONVERGENCE TOLERANCE		0.000100000

*** RESULTS ***

OUTLET TEMPERATURE	C	94.000
OUTLET PRESSURE	BAR	0.82500
HEAT DUTY	GCAL/HR	8.6309
OUTLET VAPOR FRACTION		0.10000E-01

V-L PHASE EQUILIBRIUM :

COMP	F(I)	X(I)	Y(I)	K(I)
H2O	0.84261	0.84317	0.78683	0.93318
NA+	0.68245E-01	0.68935E-01	0.0000	0.0000
CL-	0.80750E-01	0.81566E-01	0.0000	0.0000
CO2	0.90175E-03	0.98610E-05	0.89199E-01	9045.6
O2	0.12402E-02	0.53471E-06	0.12397	0.23184E+06
MG++	0.62523E-02	0.63155E-02	0.0000	0.0000

BLOCK: HEATOUT MODEL: HEATER

INLET STREAM: COOLVAP
INLET HEAT STREAM: HEAT1
OUTLET STREAM: DIST1
PROPERTY OPTION SET: ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
HENRY-COMPS ID: GLOBAL
CHEMISTRY ID: GLOBAL - TRUE SPECIES

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(KMOL/HR)	2081.14	2081.14	0.00000
MASS(KG/HR)	37492.4	37492.4	0.00000
ENTHALPY(GCAL/HR)	-138.889	-138.889	0.140797E-08

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.170965	KG/HR
PRODUCT STREAMS CO2E	0.170965	KG/HR
NET STREAMS CO2E PRODUCTION	0.00000	KG/HR
UTILITIES CO2E PRODUCTION	0.00000	KG/HR
TOTAL CO2E PRODUCTION	0.00000	KG/HR

*** INPUT DATA ***

TWO PHASE PQ FLASH

SPECIFIED PRESSURE	BAR	2.00000
DUTY FROM INLET HEAT STREAM(S)	GCAL/HR	-8.63093
MAXIMUM NO. ITERATIONS		30
CONVERGENCE TOLERANCE		0.000100000

*** RESULTS ***

OUTLET TEMPERATURE	C	112.24
OUTLET PRESSURE	BAR	2.0000
OUTLET VAPOR FRACTION		0.0000

V-L PHASE EQUILIBRIUM :

COMP	F(I)	X(I)	Y(I)	K(I)
H2O	1.0000	1.0000	0.98848	0.77396

```
   CO2           0.18666E-05    0.18666E-05    0.71144E-02     2984.3
   O2            0.96193E-07    0.96193E-07    0.44063E-02     35867.

BLOCK:  MIX1     MODEL: MIXER
-----------------------------
 INLET STREAMS:         HTBRINE5    HTBRIN11
 OUTLET STREAM:         HTBRINE6
 PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:        GLOBAL
 CHEMISTRY ID:          GLOBAL   - TRUE SPECIES


   ***********************************************************************
   *                                                                     *
   *    AT LEAST ONE OF THE INLET OR OUTLET STREAMS                      *
   *    IS NOT IN CHARGE BALANCE                                         *
   *                                                                     *
   ***********************************************************************


              ***  MASS AND ENERGY BALANCE  ***
                         IN            OUT       RELATIVE DIFF.
  TOTAL BALANCE
    MOLE(KMOL/HR )           17167.9        17167.3       0.335501E-04
    MASS(KG/HR   )           407978.        407978.       0.142674E-15
    ENTHALPY(GCAL/HR )      -1158.62       -1158.62      -0.488880E-07

              ***  CO2 EQUIVALENT SUMMARY ***
  FEED STREAMS CO2E            0.288818E-04  KG/HR
  PRODUCT STREAMS CO2E         0.288818E-04  KG/HR
  NET STREAMS CO2E PRODUCTION -0.597103E-12  KG/HR
  UTILITIES CO2E PRODUCTION    0.00000       KG/HR
  TOTAL CO2E PRODUCTION       -0.597103E-12  KG/HR

                 ***  INPUT DATA  ***
 TWO    PHASE     FLASH
 MAXIMUM NO. ITERATIONS                            30
 CONVERGENCE TOLERANCE                               0.000100000
 OUTLET PRESSURE:  MINIMUM OF INLET STREAM PRESSURES

BLOCK:  RECSPLIT MODEL: FSPLIT
------------------------------
 INLET STREAM:          VAP1
 OUTLET STREAMS:        VAP2       VAPREC
 PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:        GLOBAL
 CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

              ***  MASS AND ENERGY BALANCE  ***
                         IN            OUT       RELATIVE DIFF.
  TOTAL BALANCE
    MOLE(KMOL/HR )           2312.38        2312.38      -0.882746E-06
    MASS(KG/HR   )           41658.2        41658.3      -0.882742E-06
    ENTHALPY(GCAL/HR )      -132.469       -132.470       0.882825E-06

              ***  CO2 EQUIVALENT SUMMARY ***
  FEED STREAMS CO2E            0.189961      KG/HR
  PRODUCT STREAMS CO2E         0.189961      KG/HR
  NET STREAMS CO2E PRODUCTION -0.100899E-06  KG/HR
  UTILITIES CO2E PRODUCTION    0.00000       KG/HR
  TOTAL CO2E PRODUCTION       -0.100899E-06  KG/HR
```

```
              ***  INPUT DATA  ***

 FRACTION OF FLOW              STRM=VAPREC   FRAC=        0.100000

              ***  RESULTS  ***

 STREAM= VAP2          SPLIT=        0.90000     KEY=  0    STREAM-ORDER=   2
         VAPREC                  0.100000         0                 1

BLOCK:  SPLIT    MODEL: FSPLIT
------------------------------
 INLET STREAM:         HTBRINE9
 OUTLET STREAMS:       HTBRINE5    HTBRINE3
 PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:       GLOBAL
 CHEMISTRY ID:         GLOBAL   - TRUE SPECIES

              ***  MASS AND ENERGY BALANCE  ***
                        IN            OUT       RELATIVE DIFF.
  TOTAL BALANCE
    MOLE(KMOL/HR )           9592.48        9592.48       0.489785E-06
    MASS(KG/HR   )         207073.        207073.      -0.140549E-15
    ENTHALPY(GCAL/HR )       -627.105       -627.105      -0.699310E-07

              ***  CO2 EQUIVALENT SUMMARY ***
  FEED STREAMS CO2E           0.577121E-04  KG/HR
  PRODUCT STREAMS CO2E        0.577121E-04  KG/HR
  NET STREAMS CO2E PRODUCTION   0.00000      KG/HR
  UTILITIES CO2E PRODUCTION     0.00000      KG/HR
  TOTAL CO2E PRODUCTION         0.00000      KG/HR

              ***  INPUT DATA  ***

 FRACTION OF FLOW              STRM=HTBRINE3 FRAC=        0.50000

              ***  RESULTS  ***

 STREAM= HTBRINE5      SPLIT=        0.50000     KEY=  0    STREAM-ORDER=   2
         HTBRINE3                0.50000          0                 1

BLOCK:  SPLIT2   MODEL: FSPLIT
------------------------------
 INLET STREAM:         HTBRINE6
 OUTLET STREAMS:       HTBRINE7    WASTE
 PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:       GLOBAL
 CHEMISTRY ID:         GLOBAL   - TRUE SPECIES


   ********************************************************************
   *                                                                  *
   *    AT LEAST ONE OF THE INLET OR OUTLET STREAMS                   *
   *    IS NOT IN CHARGE BALANCE                                      *
   *                                                                  *
   ********************************************************************


              ***  MASS AND ENERGY BALANCE  ***
                        IN            OUT       RELATIVE DIFF.
  TOTAL BALANCE
    MOLE(KMOL/HR )          17167.3        17167.3      -0.211913E-15
    MASS(KG/HR   )         407978.        407978.      -0.142674E-15
```

```
    ENTHALPY(GCAL/HR )      -1158.62       -1158.62       0.196245E-15

              ***  CO2 EQUIVALENT SUMMARY ***
  FEED STREAMS CO2E           0.288818E-04  KG/HR
  PRODUCT STREAMS CO2E        0.288818E-04  KG/HR
  NET STREAMS CO2E PRODUCTION   0.00000      KG/HR
  UTILITIES CO2E PRODUCTION     0.00000      KG/HR
  TOTAL CO2E PRODUCTION         0.00000      KG/HR

                 ***  INPUT DATA  ***

 FRACTION OF FLOW             STRM=WASTE    FRAC=        0.20000

                  ***  RESULTS  ***

 STREAM= HTBRINE7      SPLIT=         0.80000     KEY=  0    STREAM-ORDER=   2
      WASTE                   0.20000         0               1

BLOCK:  VAPCOMP  MODEL: COMPR
-----------------------------
 INLET STREAM:          VAP2
 OUTLET STREAM:         COMPVAP
 PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:        GLOBAL
 CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

              ***  MASS AND ENERGY BALANCE  ***
                         IN            OUT       RELATIVE DIFF.
  TOTAL BALANCE
    MOLE(KMOL/HR )          2081.14        2081.14       0.218509E-15
    MASS(KG/HR   )        37492.4        37492.4       0.194065E-15
    ENTHALPY(GCAL/HR )      -119.223       -113.863      -0.449569E-01

              ***  CO2 EQUIVALENT SUMMARY ***
  FEED STREAMS CO2E           0.170965      KG/HR
  PRODUCT STREAMS CO2E        0.170965      KG/HR
  NET STREAMS CO2E PRODUCTION   0.00000      KG/HR
  UTILITIES CO2E PRODUCTION     0.00000      KG/HR
  TOTAL CO2E PRODUCTION         0.00000      KG/HR

                  ***  INPUT DATA  ***

 ISENTROPIC CENTRIFUGAL COMPRESSOR
  OUTLET PRESSURE  BAR                          4.00000
  ISENTROPIC EFFICIENCY                         0.75000
  MECHANICAL EFFICIENCY                         0.75000

                  ***  RESULTS  ***

  INDICATED  HORSEPOWER REQUIREMENT  KW              6,233.53
  BRAKE      HORSEPOWER REQUIREMENT  KW              8,311.38
  NET WORK REQUIRED                 KW               8,311.38
  POWER LOSSES                      KW               2,077.84
  ISENTROPIC HORSEPOWER REQUIREMENT  KW              4,675.15
  CALCULATED OUTLET TEMP  C                          395.285
  ISENTROPIC TEMPERATURE  C                          322.252
  EFFICIENCY (POLYTR/ISENTR) USED                    0.75000
  OUTLET VAPOR FRACTION                              1.00000
  HEAD DEVELOPED,      METER                         45,775.6
  MECHANICAL EFFICIENCY USED                         0.75000
  INLET HEAT CAPACITY RATIO                          1.32710
  INLET VOLUMETRIC FLOW RATE , CUM/HR                124,965.
```

```
  OUTLET VOLUMETRIC FLOW RATE, CUM/HR              28,751.8
  INLET  COMPRESSIBILITY FACTOR                    0.99621
  OUTLET COMPRESSIBILITY FACTOR                    0.99435
  AV. ISENT. VOL. EXPONENT                       1.31006
  AV. ISENT. TEMP EXPONENT                       1.31347
  AV. ACTUAL VOL. EXPONENT                        1.41523
  AV. ACTUAL TEMP EXPONENT                        1.41703

BLOCK:  VAPCOMP2 MODEL: COMPR
----------------------------
 INLET STREAM:         HTVAP2
 OUTLET STREAM:        COMPVAP2
 PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
 HENRY-COMPS ID:       GLOBAL
 CHEMISTRY ID:         GLOBAL   - TRUE SPECIES

              ***  MASS AND ENERGY BALANCE  ***
                     IN            OUT       RELATIVE DIFF.
  TOTAL BALANCE
    MOLE(KMOL/HR )          1217.99        1217.99        0.00000
    MASS(KG/HR   )          21942.4        21942.4       0.165797E-15
    ENTHALPY(GCAL/HR )      -69.7702       -68.3032      -0.210263E-01

              ***  CO2 EQUIVALENT SUMMARY ***
  FEED STREAMS CO2E           0.230797E-04  KG/HR
  PRODUCT STREAMS CO2E        0.230797E-04  KG/HR
  NET STREAMS CO2E PRODUCTION   0.00000      KG/HR
  UTILITIES CO2E PRODUCTION     0.00000      KG/HR
  TOTAL CO2E PRODUCTION         0.00000      KG/HR

                  ***  INPUT DATA  ***

 ISENTROPIC CENTRIFUGAL COMPRESSOR
  PRESSURE CHANGE  BAR                           1.00000
  ISENTROPIC EFFICIENCY                          0.75000
  MECHANICAL EFFICIENCY                            0.75000

                  ***  RESULTS  ***

  INDICATED  HORSEPOWER REQUIREMENT  KW             1,706.13
  BRAKE      HORSEPOWER REQUIREMENT  KW             2,274.84
  NET WORK REQUIRED                 KW              2,274.84
  POWER LOSSES                      KW              568.709
  ISENTROPIC HORSEPOWER REQUIREMENT  KW             1,279.60
  CALCULATED OUTLET PRES  BAR                      1.50000
  CALCULATED OUTLET TEMP  C                        235.993
  ISENTROPIC TEMPERATURE  C                        200.308
  EFFICIENCY (POLYTR/ISENTR) USED                   0.75000
  OUTLET VAPOR FRACTION                            1.00000
  HEAD DEVELOPED,       METER                      21,407.7
  MECHANICAL EFFICIENCY USED                        0.75000
  INLET HEAT CAPACITY RATIO                        1.32704
  INLET VOLUMETRIC FLOW RATE , CUM/HR              73,235.2
  OUTLET VOLUMETRIC FLOW RATE, CUM/HR              34,216.2
  INLET  COMPRESSIBILITY FACTOR                    0.99622
  OUTLET COMPRESSIBILITY FACTOR                    0.99543
  AV. ISENT. VOL. EXPONENT                       1.31624
  AV. ISENT. TEMP EXPONENT                       1.31909
  AV. ACTUAL VOL. EXPONENT                        1.44369
  AV. ACTUAL TEMP EXPONENT                        1.44519

BLOCK:  VAPHEAT  MODEL: HEATER
```

INLET STREAM: COMPVAP
OUTLET STREAM: COOLVAP
OUTLET HEAT STREAM: HEAT2
PROPERTY OPTION SET: ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
HENRY-COMPS ID: GLOBAL
CHEMISTRY ID: GLOBAL - TRUE SPECIES

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(KMOL/HR)	2081.14	2081.14	0.00000
MASS(KG/HR)	37492.4	37492.4	0.00000
ENTHALPY(GCAL/HR)	-113.863	-113.863	-0.124807E-15

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.170965	KG/HR
PRODUCT STREAMS CO2E	0.170965	KG/HR
NET STREAMS CO2E PRODUCTION	0.00000	KG/HR
UTILITIES CO2E PRODUCTION	0.00000	KG/HR
TOTAL CO2E PRODUCTION	0.00000	KG/HR

*** INPUT DATA ***

TWO PHASE PV FLASH

SPECIFIED PRESSURE	BAR	3.00000
VAPOR FRACTION		0.40000
MAXIMUM NO. ITERATIONS		30
CONVERGENCE TOLERANCE		0.000100000

*** RESULTS ***

OUTLET TEMPERATURE	C	133.58
OUTLET PRESSURE	BAR	3.0000
HEAT DUTY	GCAL/HR	-16.395
OUTLET VAPOR FRACTION		0.40000

V-L PHASE EQUILIBRIUM :

COMP	F(I)	X(I)	Y(I)	K(I)
H2O	1.0000	1.0000	1.0000	1.0000
CO2	0.18666E-05	0.21280E-08	0.46634E-05	2191.4
O2	0.96193E-07	0.96640E-11	0.24047E-06	24883.

BLOCK: VAPHEAT2 MODEL: HEATER

INLET STREAM: COMPVAP2
OUTLET STREAM: DIST2
OUTLET HEAT STREAM: HEAT3
PROPERTY OPTION SET: ELECNRTL ELECTROLYTE NRTL / REDLICH-KWONG
HENRY-COMPS ID: GLOBAL
CHEMISTRY ID: GLOBAL - TRUE SPECIES

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(KMOL/HR)	1217.99	1217.99	0.00000
MASS(KG/HR)	21942.4	21942.4	0.00000
ENTHALPY(GCAL/HR)	-68.3032	-68.3032	0.00000

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.230797E-04	KG/HR
PRODUCT STREAMS CO2E	0.230797E-04	KG/HR
NET STREAMS CO2E PRODUCTION	0.00000	KG/HR
UTILITIES CO2E PRODUCTION	0.00000	KG/HR
TOTAL CO2E PRODUCTION	0.00000	KG/HR

*** INPUT DATA ***

TWO PHASE PV FLASH

SPECIFIED PRESSURE	BAR	1.00000
VAPOR FRACTION		0.050000
MAXIMUM NO. ITERATIONS		30
CONVERGENCE TOLERANCE		0.000100000

*** RESULTS ***

OUTLET TEMPERATURE	C	99.629
OUTLET PRESSURE	BAR	1.0000
HEAT DUTY	GCAL/HR	-12.667
OUTLET VAPOR FRACTION		0.50000E-01

V-L PHASE EQUILIBRIUM :

COMP	F(I)	X(I)	Y(I)	K(I)
H2O	1.0000	1.0000	1.0000	1.0000
CO2	0.43056E-09	0.15724E-11	0.85814E-08	5457.4
O2	0.66509E-12	0.18963E-15	0.13298E-10	70127.

<u>Heat Exchanger Design - GE Brine Concentrator</u>

Heat Exchanger HEAT1

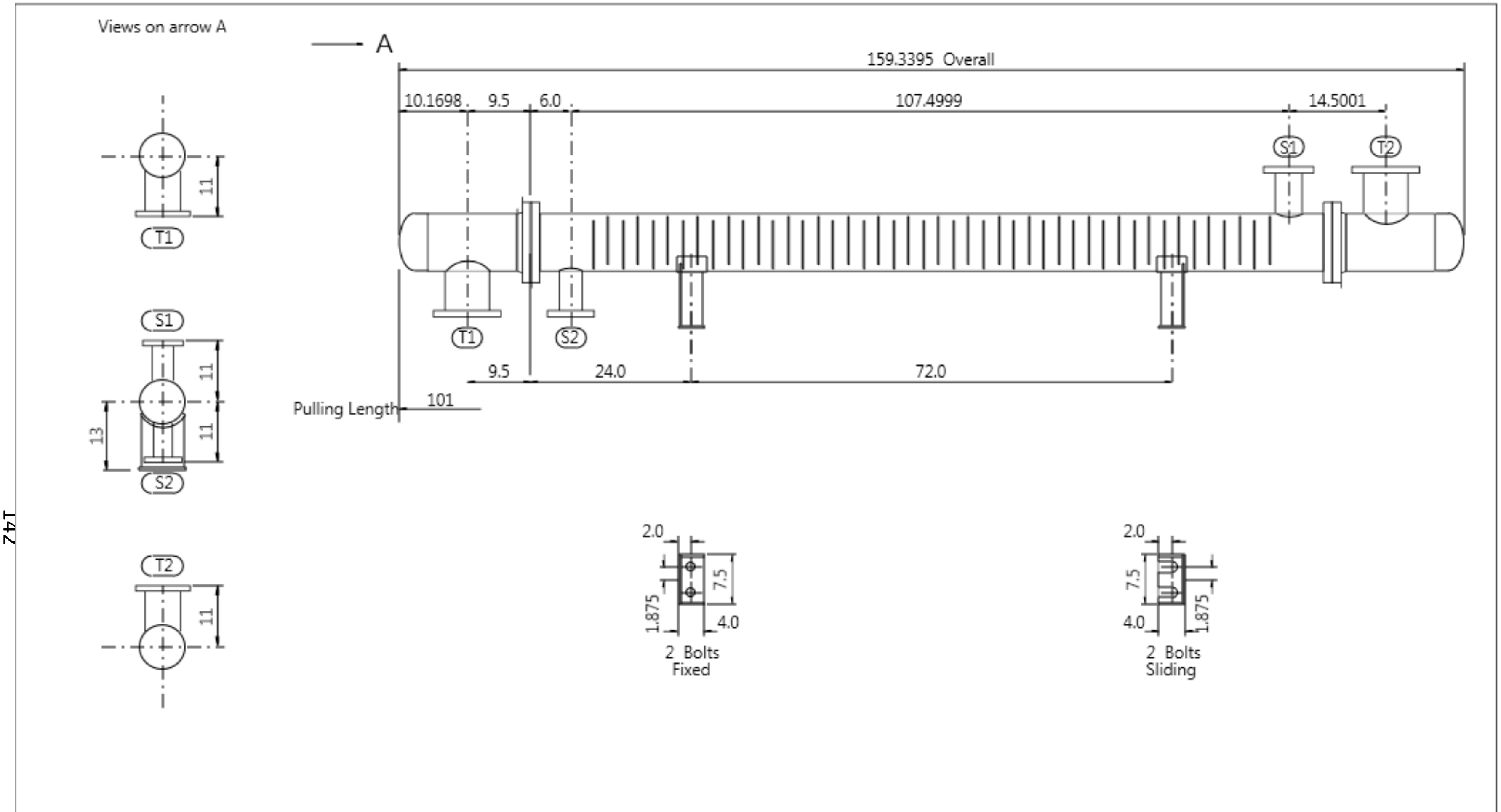
Heat Transfer Parameters			
Total heat load	Mkcal/	5.975	
Eff. MTD/ 1 pass MTD	°F 264.13	/	264.13
Actual/Reqd area ratio - fouled/clean	1	/	1

Coef./Resist.	BTU/(h-ft²-F)	ft²-h-F/BTU	%
Overall fouled	999.84	0.001	
Overall clean	999.84	0.001	
Tube side film	3185.8	0.0003	31.38
Tube side fouling		0	0
Tube wall	3033.5	0.0003	32.96
Outside fouling		0	0
Outside film	2804.11	0.0004	35.66

Overall Coefficient / Resistance Summary			**Clean**	**Dirty**	**Max Dirty**
Area required		ft²	89.8	89.8	90
Area ratio: actual/required			1	1	1
Overall coefficient		BTU/(h-ft²-F)	999.84	999.84	997.68
Overall resistance		ft²-h-F/BTU	0.001	0.001	0.001
Shell side fouling		ft²-h-F/BTU	0.0	0	0
Tube side fouling			0.0	0	0
Resistance Distribution	BTU/(h-ft²-F)	ft²-h-F/BTU	%	%	%
Shell side film	2804.11	0.0004	35.66	35.66	35.58
Shell side fouling		0		0	0.11
Tube wall	3033.5	0.0003	32.96	32.96	32.89
Tube side fouling *		0		0	0.11
Tube side film *	3185.8	0.0003	31.38	31.38	31.32

* Based on outside surface - Area ratio: Ao/Ai = 1.15

1	Company: Desalination Team 12					
2	Location:					
3	Service of Unit:	Our Reference:				
4	Item No.:	Your Reference:				
5	Date:	Rev No.:	Job No.:			
6	Size : 8.071 - 120 in	Type: BEM	Horizontal	Connected in: 1 parallel	1 series	
7	Surf/unit(eff.) 90 ft²	Shells/unit 1		Surf/shell(eff.) 90 ft²		
8	**PERFORMANCE OF ONE UNIT**					
9	Fluid allocation		Shell Side		Tube Side	
10	Fluid name		PUMPVAP		BRINEIN	
11	Fluid quantity, Total	lb/h	50641		235042	
12	Vapor (In/Out)	lb/h	0	0	0	0
13	Liquid	lb/h	50641	50641	235042	235042
14	Noncondensable	lb/h	0	0	0	0
15						
16	Temperature (In/Out)	°F	659.45	207.32	77	197.86
17	Dew / Bubble point	°F				
18	Density Vapor/Liquid	lb/ft³	/ 43.929	/ 59.987	/ 67.496	/ 65.518
19	Viscosity	cp	/ 0.0692	/ 0.29	/ 1.3349	/ 0.4498
20	Molecular wt, Vap					
21	Molecular wt, NC					
22	Specific heat	BTU/(lb-F)	/ 1.1602	/ 1.0017	/ 0.8347	/ 0.8354
23	Thermal conductivity	BTU/(ft-h-F)	/ 0.296	/ 0.39	/ 1.16	/ 0.996
24	Latent heat	BTU/lb				
25	Pressure (abs)	psi	304.58	279.98	14.5	11.55
26	Velocity (Mean/Max)	ft/s	7.07 / 12.71		9.01 / 9.14	
27	Pressure drop, allow./calc.	psi	159.54	24.6	3	2.95
28	Fouling resistance (min)	ft²-h-F/BTU	0		0	0 Ao based
29	Heat exchanged 23710710	BTU/h		MTD (corrected) 264.13		°F
30	Transfer rate, Service 997.68		Dirty 999.84	Clean 999.84		BTU/(h-ft²-F)
31	**CONSTRUCTION OF ONE SHELL**				**Sketch**	
32			Shell Side	Tube Side		
33	Design/Vacuum/test pressure	psi	340 / /	50 / /		
34	Design temperature	°F	480	270		
35	Number passes per shell		1	1		
36	Corrosion allowance	in	0.125	0		
37	Connections	In in	1 3.5 / -	1 6 / -		
38	Size/Rating	Out	1 3 / -	1 6 / -		
39	Nominal	Intermediate	/ -	/ -		
40	Tube No. 47 OD 0.75 Tks Average 0.049	in Length 120	in Pitch 0.9375 in			
41	Tube type Plain	#/in Material Monel	Tube pattern 30			
42	Shell Carbon Steel ID 8.071 OD 8.625 in	Shell cover -				
43	Channel or bonnet Carbon Steel	Channel cover -				
44	Tubesheet-stationary Carbon Steel -	Tubesheet-floating -				
45	Floating head cover -	Impingement protection None				
46	Baffle-cross Carbon Steel Type Single segmental Cut(%d) 13.05	H Spacing: c/c 2.25 in				
47	Baffle-long - Seal Type	Inlet 7.875 in				
48	Supports-tube U-bend 0	Type				
49	Bypass seal	Tube-tubesheet joint Expanded only (2 grooves)(App.A 'i')				
50	Expansion joint -	Type None				
51	RhoV2-Inlet nozzle 956	Bundle entrance 648	Bundle exit 475	lb/(ft-s²)		
52	Gaskets - Shell side -	Tube side	Flat Metal Jacket Fibe			



Nozzle Data				
Ref	OD	Wall	Standard	Notes
S1	4.0"	0.226"	150 ANSI Slip on	
S2	3.5"	0.216"	150 ANSI Slip on	
T1	6.625"	0.28"	150 ANSI Slip on	
T2	6.625"	0.28"	150 ANSI Slip on	

Design Data	Units	Shell	Channel
Design Pressure	psi	340	50
Design Temperature	F	480	270
Full Vacuum		0	0
Corrosion Allowance	in	0.125	0
Test Pressure	psi		
Number of Passes		1	1
Radiography		0	0
PWHT		0	0
Internal Volume	ft^3	3.3867	1.3781

Weight Summary		
Empty	Flooded	Bundle
918 lb	1150 lb	312 lb

Company: Desalination Team 12	
Location:	
Service of Unit:	Our Reference:
Item No.:	Your Reference:
Date: Rev No.:	Job No.:
Design Codes	
0	
TEMA 0	
Customer Specifications	

Aspen Shell & Tube Exchanger

Setting Plan

BEM 8 - 120

Drawing Number

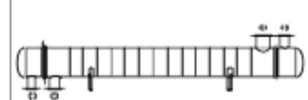
Revision	Date	Dwg.	Chk.	App.
	4/3/2016			

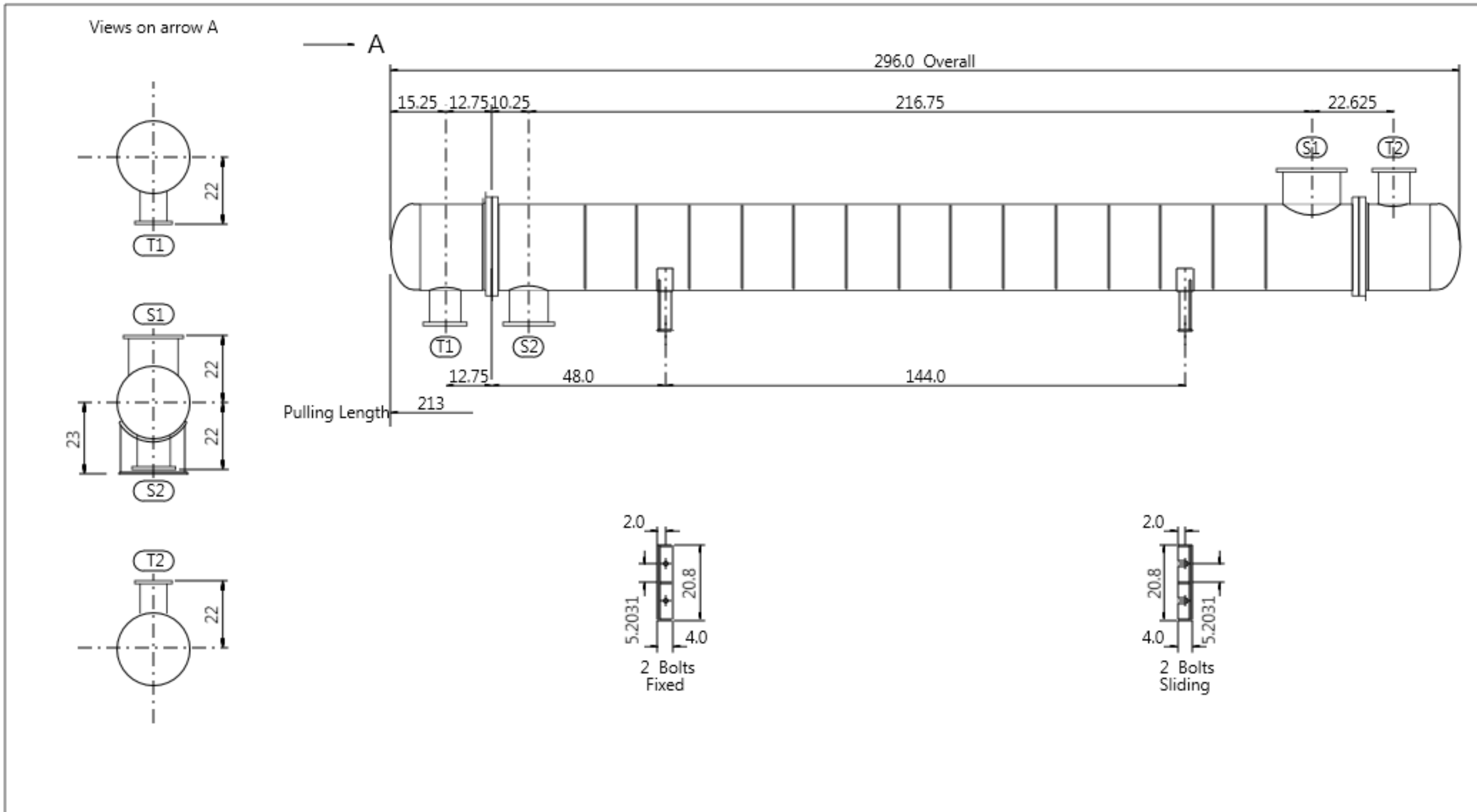
Heat Exchanger HEAT2

Heat Transfer Parameters				
Total heat load		Mkcal/	9	
Eff. MTD/ 1 pass MTD		°F	43.41 /	48.95
Actual/Reqd area ratio - fouled/clean			1.04 /	1.04
Coef./Resist.	BTU/(h-ft²-F)	ft²-h-F/BTU	%	
Overall fouled	554.7	0.0018		
Overall clean	554.7	0.0018		
Tube side film	1901.09	0.0005	29.18	
Tube side fouling		0	0	
Tube wall	3283.57	0.0003	16.89	
Outside fouling		0	0	
Outside film	1028.57	0.001	53.93	

Overall Coefficient / Resistance Summary			**Clean**	**Dirty**	**Max Dirty**
Area required		ft^2	1483.3	1483.3	1537.7
Area ratio: actual/required			1.04	1.04	1
Overall coefficient		BTU/(h-ft²-F)	554.7	554.7	535.07
Overall resistance		ft²-h-F/BTU	0.0018	0.0018	0.0019
Shell side fouling		ft²-h-F/BTU	0.0	0	0
Tube side fouling			0.0	0	0
Resistance Distribution	BTU/(h-ft²-F)	ft²-h-F/BTU	%	%	%
Shell side film	1028.57	0.001	53.93	53.93	52.02
Shell side fouling		0		0	1.77
Tube wall	3283.57	0.0003	16.89	16.89	16.3
Tube side fouling *		0		0	1.77
Tube side film *	1901.09	0.0005	29.18	29.18	28.15

1	Company:							
2	Location:							
3	Service of Unit:	Our Reference:						
4	Item No.:	Your Reference:						
5	Date:	Rev No.:	Job No.:					
6	Size : 23 - 240	in	Type: BEM	Horizontal	Connected in:	1 parallel	1 series	
7	Surf/unit(eff.)	1537.7	ft^2	Shells/unit 1	Surf/shell(eff.)	1537.7	ft^2	
8	**PERFORMANCE OF ONE UNIT**							
9	Fluid allocation		Shell Side		Tube Side			
10	Fluid name		COMPVAP		HTBRINE3			
11	Fluid quantity, Total	lb/h	45931		566079			
12	Vapor (In/Out)	lb/h	45931	18128	0	0		
13	Liquid	lb/h	0	27803	566079	566079		
14	Noncondensable	lb/h	0	0	0	0		
15								
16	Temperature (In/Out)	°F	589.99	276.31	192.2	293		
17	Dew / Bubble point	°F	268.63	268.43				
18	Density Vapor/Liquid	lb/ft^3	0.097 /	0.147 / 56.67	/ 71.032	/ 67.827		
19	Viscosity	cp	0.0208 /	0.0199 / 0.1742	/ 0.7723	/ 0.4835		
20	Molecular wt, Vap		18.01	18.01				
21	Molecular wt, NC							
22	Specific heat	BTU/(lb-F)	0.5899 /	0.5927 / 1.011	/ 0.6166	/ 0.6312		
23	Thermal conductivity	BTU/(ft-h-F)	0.027 /	0.021 / 0.405	/ 4.756	/ 3.517		
24	Latent heat	BTU/lb	816	816				
25	Pressure (abs)	psi	58.02	47.76	58.02	57.06		
26	Velocity (Mean/Max)	ft/s	70.34 / 211.44		2.45 / 2.51			
27	Pressure drop, allow./calc.	psi	14.5	10.25	14.5	0.95		
28	Fouling resistance (min)	ft^2-h-F/BTU	0		0	0 Ao based		
29	Heat exchanged	9	Mkcal/h		MTD (corrected)	43.41	°F	
30	Transfer rate, Service	535.07	Dirty	554.7	Clean	554.7	BTU/(h-ft^2-F)	
31	**CONSTRUCTION OF ONE SHELL**				**Sketch**			
32			Shell Side	Tube Side				
33	Design/Vacuum/test pressure	psi	70 / /	70 / /				
34	Design temperature	°F	810	360				
35	Number passes per shell		1	1				
36	Corrosion allowance	in	0.125	0				
37	Connections	In in	1 16 / -	1 8 / -				
38	Size/Rating	Out	1 10 / -	1 8 / -				
39	Nominal	Intermediate	/ -	/ -				
40	Tube No. 398	OD 0.75	Tks Average 0.049	in	Length 240	in	Pitch 0.9375	in
41	Tube type Plain	#/in	Material	Monel	Tube pattern	30		
42	Shell Carbon Steel	ID 23.25	OD 24	in	Shell cover	-		
43	Channel or bonnet	Carbon Steel			Channel cover	-		
44	Tubesheet-stationary	Carbon Steel	-		Tubesheet-floating	-		
45	Floating head cover	-			Impingement protection	None		
46	Baffle-cross Carbon Steel	Type	Single segmental	Cut(%d)	39.52	V Spacing: c/c	14.5	in
47	Baffle-long -		Seal Type		Inlet	23.8125	in	
48	Supports-tube	U-bend	0		Type			
49	Bypass seal		Tube-tubesheet joint		Expanded only (2 grooves)(App.A 'i')			
50	Expansion joint	-	Type	None				
51	RhoV2-Inlet nozzle	1045	Bundle entrance	2138	Bundle exit	831	lb/(ft-s^2)	
52	Gaskets - Shell side	-	Tube side		Flat Metal Jacket Fibe			





Nozzle Data				
Ref	OD	Wall	Standard	Notes
S1	16.0"	0.375"	150 ANSI Slip on	
S2	10.75"	0.365"	150 ANSI Slip on	
T1	8.625"	0.322"	150 ANSI Slip on	
T2	8.625"	0.322"	150 ANSI Slip on	

Design Data	Units	Shell	Channel
Design Pressure	psi	70	70
Design Temperature	F	810	360
Full Vacuum		0	0
Corrosion Allowance	in	0.125	0
Test Pressure	psi		
Number of Passes		1	1
Radiography		0	0
PWHT		0	0
Internal Volume	ft^3	59.3627	13.216

Weight Summary		
Empty	Flooded	Bundle
6843 lb	10581 lb	3960 lb

Company:
Location:
Service of Unit: Our Reference:
Item No.: Your Reference:
Date: Rev No.: Job No.:

Design Codes
0
TEMA 0

Customer Specifications

Aspen Shell & Tube Exchanger

Setting Plan

BEM 23 - 240

Drawing Number

Revision	Date	Dwg.	Chk.	App.
	4/3/2016			

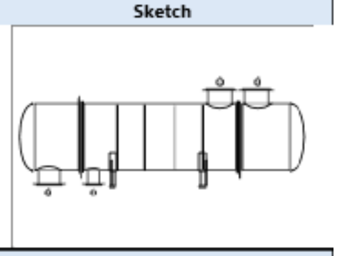
Heat Exchanger HEAT3

Heat Transfer Parameters				
Total heat load		Mkcal/	24	
Eff. MTD/ 1 pass MTD		°F	25.54 /	27.47
Actual/Reqd area ratio - fouled/clean			1.15 /	1.15
Coef./Resist.	BTU/(h-ft^2-F)	ft^2-h-F/BTU		%
Overall fouled	952.11	0.0011		
Overall clean	952.11	0.0011		
Tube side film	3032.73	0.0003		31.39
Tube side fouling		0		0
Tube wall	3204.39	0.0003		29.71
Outside fouling		0		0
Outside film	2448	0.0004		38.89

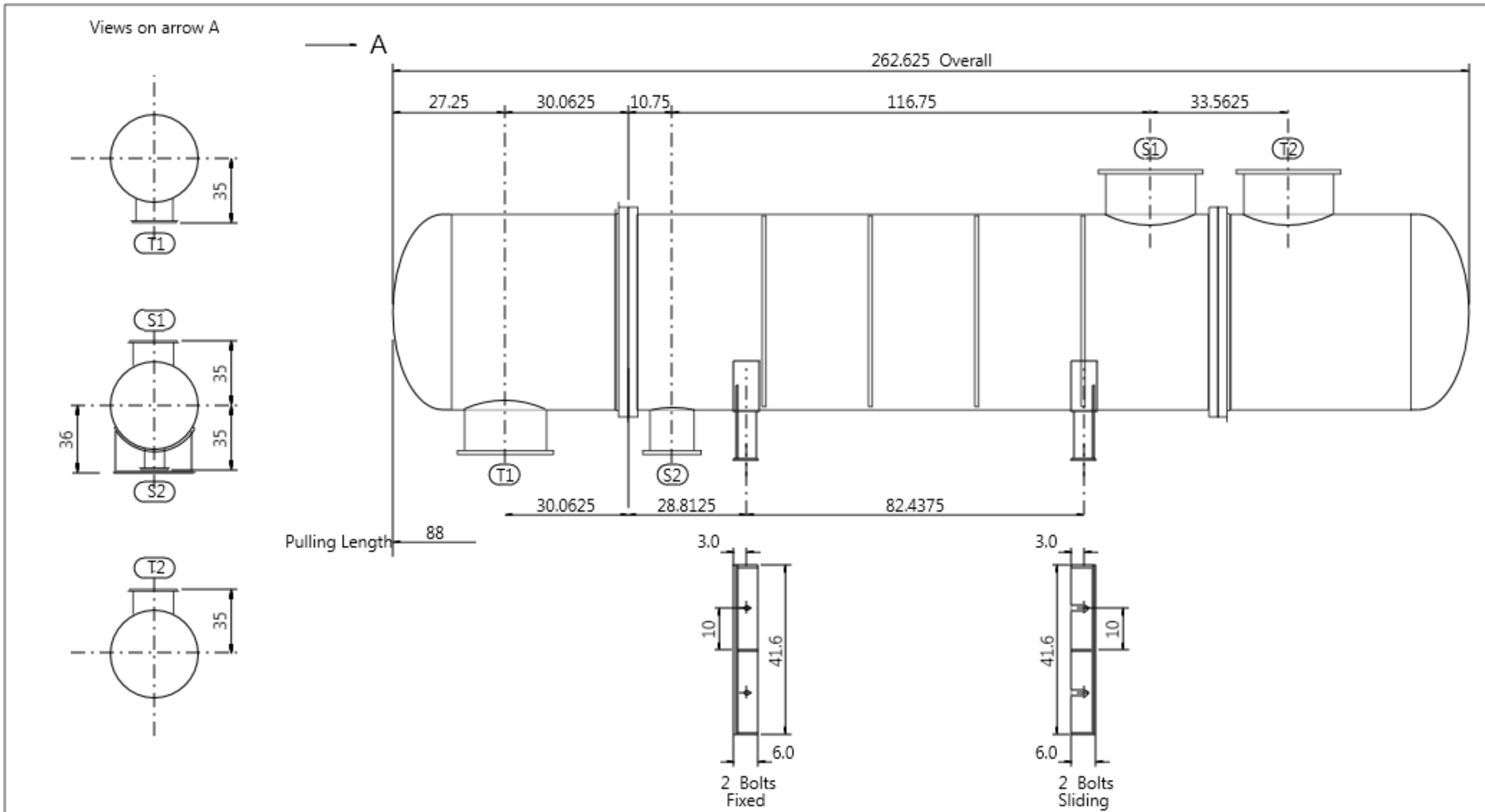
Overall Coefficient / Resistance Summary			**Clean**	**Dirty**	**Max Dirty**
Area required		ft^2	3916.4	3916.4	4484.6
Area ratio: actual/required			1.15	1.15	1
Overall coefficient		BTU/(h-ft^2-F)	952.11	952.11	831.49
Overall resistance		ft^2-h-F/BTU	0.0011	0.0011	0.0012
Shell side fouling		ft^2-h-F/BTU	0.0	0	0.0001
Tube side fouling			0.0	0	0.0001
Resistance Distribution	BTU/(h-ft^2-F)	ft^2-h-F/BTU	%	%	%
Shell side film	2448	0.0004	38.89	38.89	33.97
Shell side fouling		0		0	6.33
Tube wall	3204.39	0.0003	29.71	29.71	25.95
Tube side fouling *		0		0	6.33
Tube side film *	3032.73	0.0003	31.39	31.39	27.42

1	Company:						
2	Location:						
3	Service of Unit:	Our Reference:					
4	Item No.:	Your Reference:					
5	Date:	Rev No.:	Job No.:				
6	Size : 47 - 144 in	Type: BEM	Horizontal		Connected in: 1 parallel	1 series	
7	Surf/unit(eff.) 4484.6 ft²	Shells/unit 1			Surf/shell(eff.) 4484.6 ft²		
8	**PERFORMANCE OF ONE UNIT**						
9	Fluid allocation		Shell Side		Tube Side		
10	Fluid name		COMPVAP2		HTBRINE7		
11	Fluid quantity, Total	lb/h	90169		4030465		
12	Vapor (In/Out)	lb/h	90169	0	0	0	
13	Liquid	lb/h	0	90169	4030465	4030465	
14	Noncondensable	lb/h	0	0	0	0	
15							
16	Temperature (In/Out)	°F	246.97	208.15	194	228.2	
17	Dew / Bubble point	°F	203.88	203.88			
18	Density Vapor/Liquid	lb/ft³	0.08 /	/ 59.971	/ 67.469	/ 65.831	
19	Viscosity	cp	0.013 /	/ 0.2887	/ 0.5394	/ 0.4586	
20	Molecular wt, Vap		18.01				
21	Molecular wt, NC						
22	Specific heat	BTU/(lb-F)	1.1632 /	/ 1.0017	/ 0.6863	/ 0.6963	
23	Thermal conductivity	BTU/(ft-h-F)	0.018 /	/ 0.391	/ 4.699	/ 3.934	
24	Latent heat	BTU/lb	816	816			
25	Pressure (abs)	psi	29.01	26.18	58.02	56.77	
26	Velocity (Mean/Max)	ft/s	57.38 / 145.81		3.68 / 3.72		
27	Pressure drop, allow./calc.	psi	14.5	2.83	43.51	1.25	
28	Fouling resistance (min)	ft²-h-F/BTU	0		0	0	Ao based
29	Heat exchanged 24	Mkcal/h			MTD (corrected) 25.54	°F	
30	Transfer rate, Service 831.49		Dirty 952.11		Clean 952.11	BTU/(h-ft²-F)	

31	**CONSTRUCTION OF ONE SHELL**										**Sketch**
32				Shell Side				Tube Side			
33	Design/Vacuum/test pressure		psi	50 / /				70 / /			
34	Design temperature		°F	520				300			
35	Number passes per shell			1				1			
36	Corrosion allowance		in	0.125				0			
37	Connections	In	in	1	22	/	-	1	20	/	-
38	Size/Rating	Out		1	10	/	-	1	22	/	-
39	Nominal	Intermediate				/	-			/	-



40	Tube No. 1970	OD 0.75	Tks Average 0.049	in Length 144	in Pitch 0.9375	in
41	Tube type Plain		#/in Material Monel		Tube pattern 30	
42	Shell Carbon Steel	ID 47	OD 48	in	Shell cover -	
43	Channel or bonnet	Carbon Steel			Channel cover -	
44	Tubesheet-stationary	Carbon Steel	-		Tubesheet-floating -	
45	Floating head cover -				Impingement protection None	
46	Baffle-cross Carbon Steel	Type Single segmental	Cut(%d) 39.64		V‹ Spacing: c/c 26	in
47	Baffle-long -		Seal Type		Inlet 30.5625	in
48	Supports-tube	U-bend 0			Type	
49	Bypass seal		Tube-tubesheet joint		Expanded only (2 grooves)(App.A 'i')	
50	Expansion joint	-	Type None			
51	RhoV2-Inlet nozzle 1295		Bundle entrance 2217		Bundle exit 8	lb/(ft-s²)
52	Gaskets - Shell side -		Tube side		Flat Metal Jacket Fibe	



Nozzle Data				
Ref	OD	Wall	Standard	Notes
S1	22.0"	0.375"	150 ANSI Slip on	
S2	10.75"	0.365"	150 ANSI Slip on	
T1	20.0"	0.375"	150 ANSI Slip on	
T2	22.0"	0.375"	150 ANSI Slip on	

Design Data	Units	Shell	Channel
Design Pressure	psi	50	70
Design Temperature	F	520	300
Full Vacuum		0	0
Corrosion Allowance	in	0.125	0
Test Pressure	psi		
Number of Passes		1	1
Radiography		0	0
PWHT		0	0
Internal Volume	ft³	141.5869	115.0999

Weight Summary		
Empty	Flooded	Bundle
19065 lb	31911 lb	11912 lb

Company:
Location:
Service of Unit: Our Reference:
Item No.: Your Reference:
Date: Rev No.: Job No.:

Design Codes
0
TEMA 0

Customer Specifications

Aspen Shell & Tube Exchanger

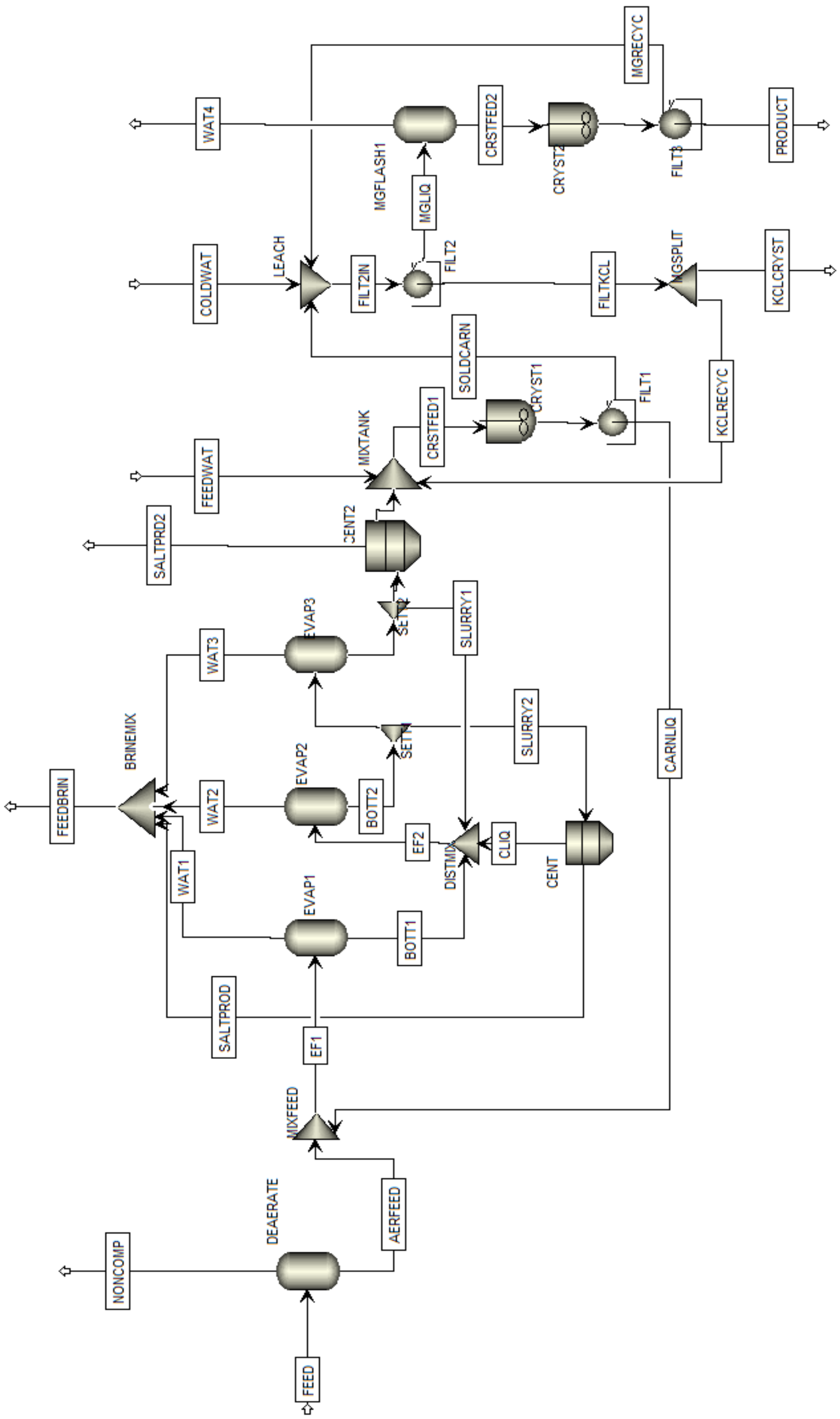
Setting Plan

BEM 47 - 144

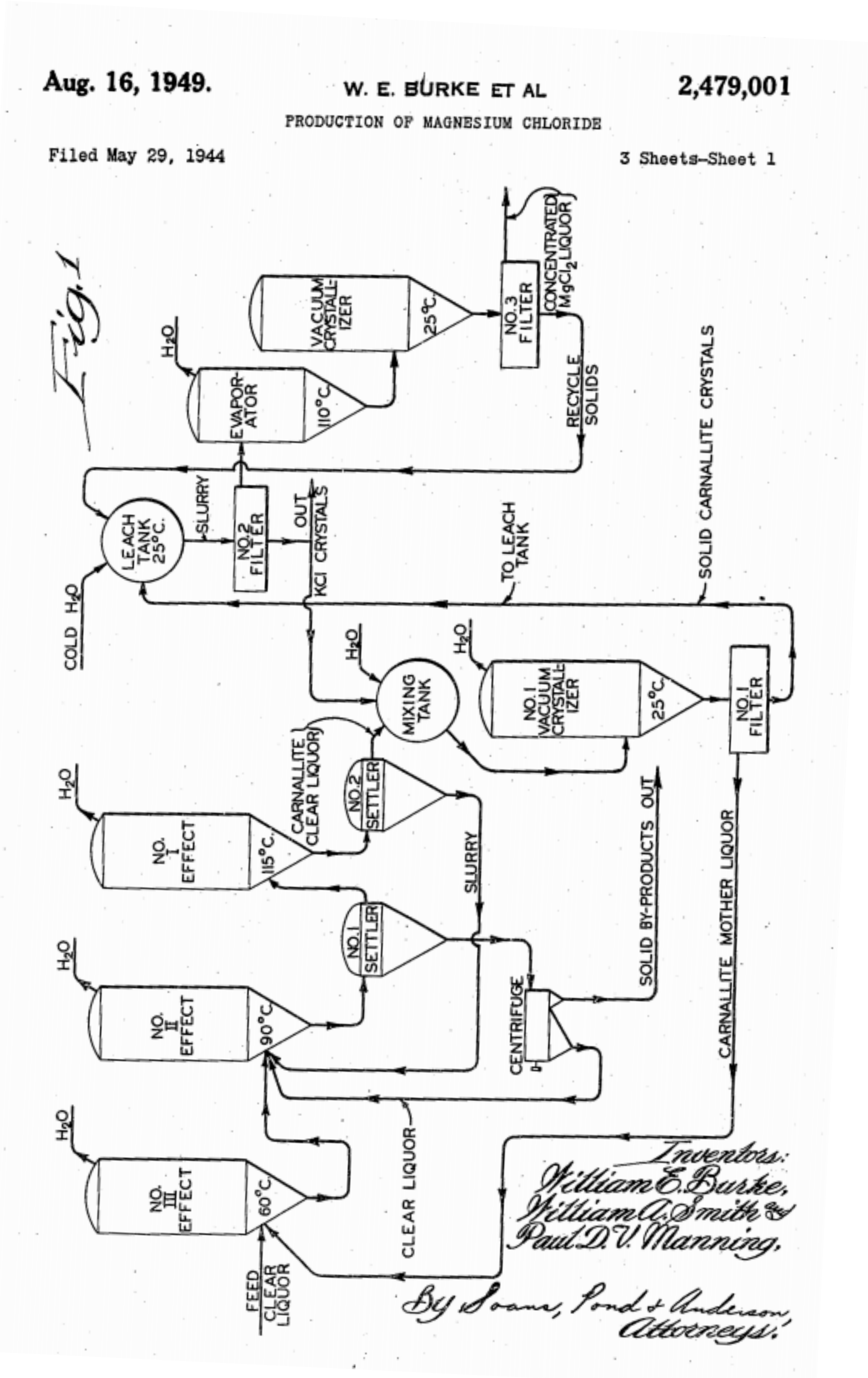
Drawing Number

Revision	Date	Dwg.	Chk.	App.
	4/3/2016			

$MgCl_2$ Separation Unit Flowsheet

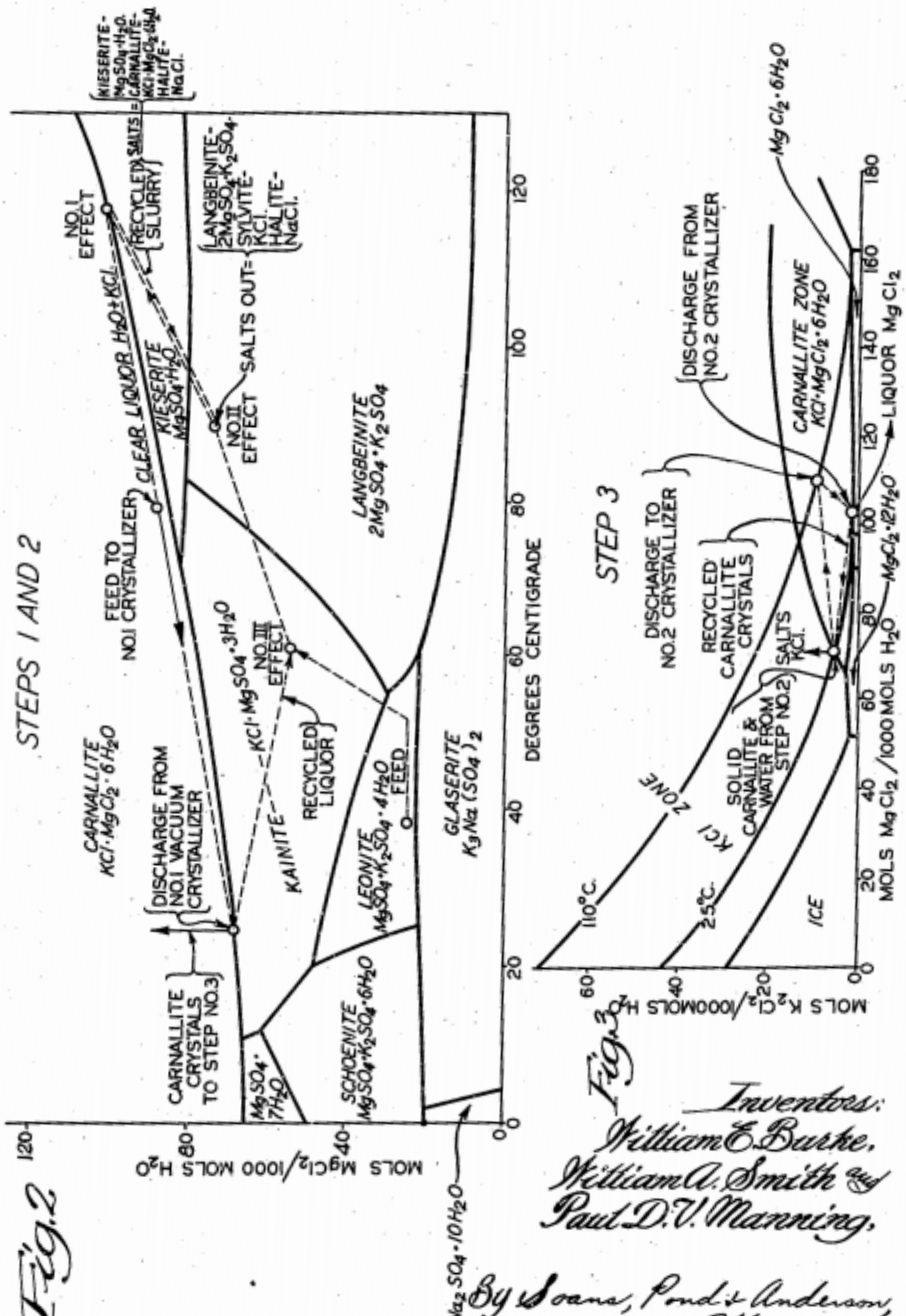


Appendix C. International Minerals & Chemicals Corporation Patent



PRODUCTION OF MAGNESIUM CHLORIDE

Filed May 29, 1944 3 Sheets-Sheet 2



Inventors:
William E. Burke,
William A. Smith and
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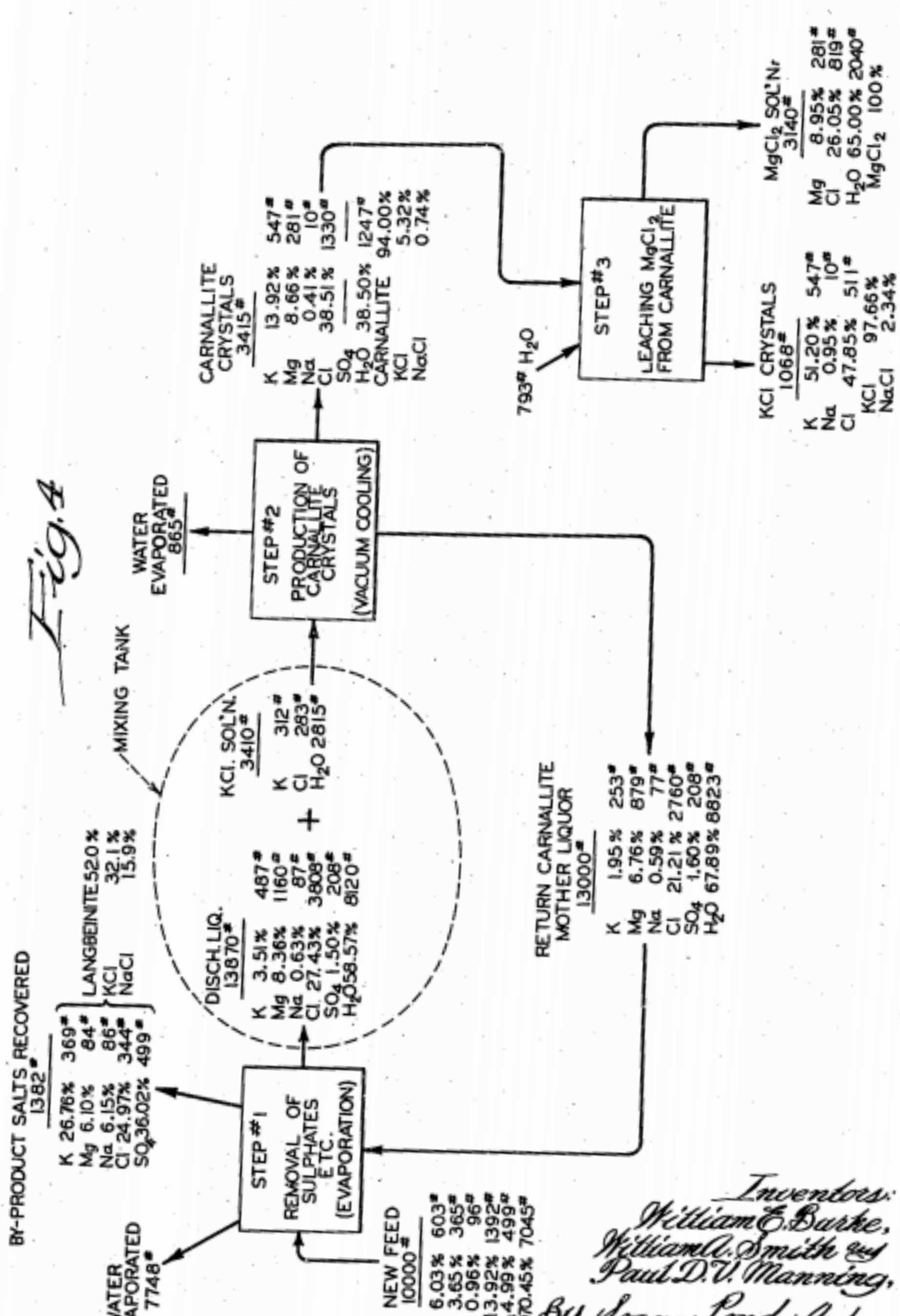
Attorneys.

Aug. 16, 1949. W. E. BURKE ET AL 2,479,001

PRODUCTION OF MAGNESIUM CHLORIDE

Filed May 29, 1944 3 Sheets-Sheet 3



Inventors:
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William A. Smith and
Paul D. V. Manning,
By Soans, Pond & Anderson,
Attorneys.

Patented Aug. 16, 1949

2,479,001

UNITED STATES PATENT OFFICE

2,479,001

PRODUCTION OF MAGNESIUM CHLORIDE

William E. Burke and William A. Smith, Carlsbad, N. Mex., and Paul D. V. Manning, Glencoe, Ill., assignors to International Minerals & Chemical Corporation, a corporation of New York

Application May 29, 1944, Serial No. 537,764

3 Claims. (Cl. 23—91)

1

Our invention relates to improvements in the production of magnesium chloride from an aqueous solution containing not only magnesium and chlorine ions, but also undesired substances such as potassium, sodium and sulphate ions.

For certain purposes, for example when used as a feed for electrolytic cells employed in the production of magnesium metal, the magnesium chloride should be of high purity, and particularly it should contain the minimum amount of sulphate, because sulphate is very undesirable from the standpoint of satisfactory operation of the electrolytic cells used in the production of magnesium metal. However, certain minerals containing magnesium frequently also contain sulphates in addition to other substances. For instance, the magnesium-containing minerals such as are found in the Carlsbad, New Mexico, or other similar mineral deposits contain magnesium principally in the form of Langbeinite ($2MgSO_4.K_2SO_4$), Kainite ($KCl.MgSO_4.3H_2O$), or Kieserite ($MgSO_4.H_2O$).

Also, as is usual in the case of minerals obtained from such deposits, other undesirable materials are present in the ores, for example Halite (NaCl) or Sylvite (KCl). In the commercial beneficiating processes in which various products, for example potash and magnesium compounds, are obtained as the principal products or as the by-products of a concentrating, refining or chemical process, it is not possible to obtain or recover substances which are 100% pure.

For example, a typical by-product liquor or brine which is produced in the Carlsbad, New Mexico, area may contain the following substances:

	Per cent
K	6.03
Mg	3.65
Na	0.96
Cl	13.92
SO_4	4.99
H_2O	70.45

This invention will be described as applied to the production of a relatively pure solution of $MgCl_2$ from a liquor containing the above-mentioned substances.

In general, it may be stated that the invention contains a number of features some of which may be independently performed, while other features of the invention should be practiced in combination with each other. The principal process steps which are useful in producing a relatively pure concentrated magnesium chloride solution from

2

a brine or liquor of the general character indicated may be said to consist of three main steps or stages.

The first of these stages consists in the elimination of undesirable substances, such as sodium or sulphates in the solid phase, or at least enough of these substances so that they will not be carried forward in the brine to the subsequent steps of the process in a sufficient amount to interfere with the production of an end product having the desired degree of purity.

The second stage of the process may be said to consist in the production of substantially pure crystals of Carnallite ($KCl.MgCl_2.6H_2O$) from the liquor from which the objectionable substances have been largely removed in the first stage of the process.

The third stage of the process consists in forming the magnesium chloride solution from the Carnallite crystals and water, preferably by a relatively simple leaching operation, and the subsequent separation of the magnesium chloride in the form of a relatively pure concentrated solution, as an end product of the process.

As an incident to the practice of the entire process, various by-products are formed, which by-products may be returned to the system or may be otherwise disposed of, it being understood that the primary object of the present process is the production of as large a percentage of pure magnesium chloride solution as possible from the feed liquor.

In order to assist in describing and understanding the various features of the complete process, we have prepared certain drawings, which accompany this application.

In said drawings,

Fig. 1 is a flow sheet of the entire process;

Fig. 2 is a phase diagram useful in understanding the first two steps of the process as above described;

Fig. 3 is a phase diagram pertaining to the third stage of the process as above described; and

Fig. 4 is a chart showing the composition of the flow at various points in the system, and the amounts and percentages of materials at such points.

In the first stage of the process, the feed liquor is preferably first concentrated by evaporation at a relatively low temperature so as to eliminate a substantial amount of water. Then the temperature is raised, and the liquor is further concentrated by evaporation. At this point in the process, there are formed crystals of materials such as Langbeinite, also chlorides of potassium and sodium, the percentage of these other substances depending, of course, upon the composition of the feed liquor. Concentration by evaporation is then continued at a higher temperature, and at this zone of higher temperature and concentration, crystals of other materials come out of solution, for example Kieserite and some Carnallite, together with crystals of other materials, and usually a little Halite.

The crystals formed in the final concentration of the original liquor could be discarded, but if, as in the present process, it is desirable to save as much magnesium as possible, the discarding of the crystals at this stage would result in discarding a molecule of magnesium with each molecule of sulphate, in view of the fact that most of the sulphate crystallizes out as Kieserite ($MgSO_4.H_2O$). Therefore, preferably the crystals formed in the final concentration are settled to a dense slurry and returned to a prior stage of the process, where the Kieserite is converted into Langbeinite under favorable conditions, and, when these salts formed in the prior stage are separated and discarded, one-third of the magnesium will be retained in solution as compared with the elimination of the sulphate in the form of Kieserite.

The strong, hot liquor resulting from the final concentration and the removal of the Kieserite in the form of a slurry, is filtered to remove all of the suspended solids. A solution of KCl is then added to the clear liquor and the total solution thus formed is cooled, preferably in a vacuum cooler in which some water is evaporated during the cooling cycle. The purpose of adding the KCl solution is to supply a sufficient amount of potassium to assure the maximum production of Carnallite crystals and to supply a sufficient amount of water to retain the sulphates in solution, it being understood that, in this type of process, it is not always possible to obtain complete elimination of undesirable materials by a simple, single crystallization stage.

After the liquor has been diluted as stated, it is cooled and concentrated under vacuum as a result of which relatively pure crystals of solid Carnallite are formed. After this cooling has proceeded to the desired point, the liquor is recycled back to the system, preferably being combined with the feed liquor. The solid Carnallite crystals which then contain the desired magnesium which is to appear in the end product constitute the feed for the final stage of the process.

In the third or final stage of the process, the Carnallite crystals formed in the prior or second stage of the process are leached with pure water, at a relatively low temperature, so that there are formed crystals of potassium chloride in a magnesium chloride solution. These crystals of potassium chloride are separated from the liquor and may be used in part in a prior stage of the process or may be otherwise disposed of.

Preferably, after the potassium chloride crystals have been separated from the solution, the latter is heated and concentrated, during which stage there may be formed crystals of Carnallite and possibly small quantities of other undesired substances, possibly also small percentages of magnesium chloride, according to conditions.

The hot, concentrated liquor, together with any crystals which may have formed, is cooled, preferably in a vacuum crystallizer, so as to separate as completely as possible the KCl in the form of Carnallite. The crystals are then removed by filtration from the strong liquor, which is the end product of the process.

If it is desired to use magnesium chloride as a raw material for electrolytic cells in the production of magnesium metal, it will of course be obvious that it is necessary to convert the solution into solid form by any appropriate dehydration method, for example, by spray drying or otherwise.

Specific example

The particular example of the complete process referred to in the drawings is based upon the treatment of feed liquor having substantially the same composition as the Carlsbad by-product liquor above referred to, which is delivered to the system at the rate of 10,000 lbs. per hour. The quantities of materials shown in Fig. 4 of the drawings are those quantities of those materials which are used, produced or disposed of during each hour of operation of the system when supplied with feed liquor at the aforesaid rate.

The raw or feed liquor is first introduced into the low temperature, or No. 3 effect, of a triple effect evaporator, the liquor being held at a temperature of about 60° C. while being evaporated. In this effect 2034 lbs. of water are evaporated per hour.

The liquor partially concentrated at the low temperature of the No. 3 effect is then conducted into the No. 2 effect, where a liquid temperature of about 90° C. is maintained. At this point in the operation, water is evaporated at the rate of about 2366 lbs. per hour of operation.

From the No. 2 effect of the evaporator, (see Fig. 2), the liquor, which at that time contains crystals of Langbeinite ($2MgSO_4.K_2SO_4$), Sylvite (KCl), and Halite (NaCl), is delivered into the No. 1 settling tank, shown in Fig. 1, and the clear liquor from the No. 1 settler is then introduced into the high temperature of the No. 1 effect of the evaporator, the liquor being maintained at a temperature of about 125° C. At this point in the system 3348 lbs. of water per hour are removed.

From the No. 1 effect, the liquor which, as shown in Fig. 2 of the drawings, at that time contains crystals of Kieserite ($MgSO_4.H_2O$), Carnallite ($KCl.MgCl_2.6H_2O$), and Halite (NaCl), is delivered to the No. 2 settling tank.

The phase diagram, Fig. 2, shows the crystalline compounds which are formed in the complex brine under various temperature conditions and under various concentrations of magnesium ion, measured as magnesium chloride. It is clearly evident from the phase diagram that Langbeinite ($2MgSO_4.K_2SO_4$) will crystallize out of the solution and Kieserite ($MgSO_4.H_2O$) will remain in solution when the temperature of the solution is over about 60° C. and the concentration of magnesium ion, measured as magnesium chloride, is less than about 80 mols of magnesium chloride per 1000 mols of water. When the concentration of magnesium ion increases to a value where there are over about 80 mols of magnesium chloride to 1000 mols of water and the temperature of the solution is over about 70° C., Kieserite ($MgSO_4.H_2O$) will crystallize out of the solution. The first mentioned conditions, Langbeinite crystallizing out and Kieserite remaining in solution, occur in the No. 2 effect and the second mentioned conditions, wherein Kieserite and some Carnallite crystallizes out, are maintained in the No. 1 effect. Thus, crystals of Kieserite which may be recycled into the No. 2 effect from the No.

1 effect will be dissolved in the liquor in the No. 2 effect.

From the No. 2 settling tank the clear liquor, most of the non-aqueous constituents of which may be represented by the formula for Carnallite, is then delivered to a mixing tank, (see Fig. 1).

At this point in the process there is added to the clear Carnallite liquor a quantity of potassium chloride and a considerable volume of cold water. The amounts of the various substances which are put into the mixing tank are as shown in the circle on Fig. 4. By the time thorough mixing of the clear liquor and the KCl solution has been effected, the temperature of the liquor has been reduced to about 80° C. From the mixing tank the liquor is discharged into the No. 1 vacuum crystallizer where the liquor is cooled by evaporation to a temperature of about 25° C. At this point, water is evaporated at the rate of about 865 lbs. per hour, (see Fig. 4).

The slurry formed in the No. 1 vacuum crystallizer is then delivered to the No. 1 filter where the crystals, which are almost pure Carnallite, are separated, and are fed to the next stage of the process.

These solid Carnallite crystals are charged into the leach tank which is at the same temperature, 25° C., as the No. 1 vacuum crystallizer previously referred to, and at this point cold water at the rate of about 793 lbs. per hour is introduced. When the Carnallite crystals dissolve in the cold water the $MgCl_2$ part of the Carnallite formula stays in solution but the KCl part immediately recrystallizes, so that the liquor in the leach tank is in the form of a slurry which contains substantially pure crystals of KCl in a substantially pure solution of $MgCl_2$. This slurry is delivered to the No. 2 filter which separates the clear liquor from the crystals. The clear $MgCl_2$ liquor is then delivered to an Ozark evaporator where it is heated to a temperature of about 110° C. and further concentrated. The concentrated liquor flowing from the Ozark evaporator together with any salts which may have come out is then delivered to the No. 2 vacuum crystallizer where it is then cooled to a temperature of about 25° C. Any crystals which have been formed are then removed by No. 3 filter so as to produce clear liquor which is a virtually pure solution of $MgCl_2$, the desired end-product of the process.

In order to obtain as high a yield as possible in respect of magnesium recovery, it may be and usually is advantageous to re-cycle certain of the side products resulting from the various treatment stages which have been described. For example, in the third or leaching stage of the process, the solid crystals separated from the concentrated end-product in filter No. 3, which consist largely of Carnallite, are returned to the leach tank.

Part of the potassium chloride crystals which constitute the cake from the No. 3 filter at this third stage are used to supply the potassium chloride which is added to the mixing tank in the previous Carnallite or second stage of the process.

Also, it is advantageous to save the clear Carnellite mother liquor which flows from the No. 1 filter in the Carnellite stage, this mother liquor being added to the raw feed liquor which flows into the first evaporator (the No. 3 effect).

Similarly, the slurry constituting the underflow from the No. 2 settler in the Carnellite stage, is returned to the intermediate evaporating stage in the No. 2 effect. As previously stated, the purpose of this is to avoid the discarding of Mg. in the form of a simple sulphate of magnesium.

The underflow from the No. 1 settler is conducted to a centrifuge which separates the solids from the clear liquor and the clear liquor is also re-cyled back into the No. 2 effect. The solids coming from the centrifuge, the composition and quantities of which are shown in the drawings as by-product salts recovered, (see Fig. 4), pass out of the system at this point. These by-product salts may be sold as such or may be beneficiated to produce more valuable substances. The other valuable by-product of the combined operation consists of the unused part of the potassium chloride which is produced as a cake from the No. 2 filter which handles the slurry from the leach tank.

We claim:

1. The improvement in the art of removing undesired sulphate ions from an aqueous solution containing magnesium, potassium, chloride, and sulphate ions which comprises: evaporating water from the solution under conditions such that the solution passes through a phase in which Langbeinite will crystallize out of solution and Kieserite will remain in solution, said evaporation being carried on at a temperature of at least about 60° C. and at a concentration of magnesium ions, measured as magnesium chloride, such that there are less than about 80 mols of magnesium chloride per 1000 mols of water; and separating Langbeinite crystals from the solution while maintaining the solution at the aforesaid conditions.

2. The improvement in the art of removing undesired sulphate ions from an aqueous solution containing magnesium, potassium, chloride, and sulphate ions which comprises: first, evaporating water from the solution under conditions such that the solution passes through a phase in which Langbeinite will crystallize out of solution and Kieserite will remain in solution, said evaporation being carried on at a temperature of at least about 60° C. and at a concentration of magnesium ions, measured as magnesium chloride, such that there are less than about 80 mols of magnesium chloride per 1000 mols of water, separating Langbeinite crystals from the solution while maintaining said solution at the aforesaid conditions; evaporating water from the resultant liquor until the concentration of magnesium ion, measured as magnesium chloride, is over about 80 mols of magnesium chloride per 1000 mols of water, maintaining the temperature of the liquor over about 70° C. and separating the Kieserite crystals, which are formed, from the solution; and finally recycling the separated Kieserite crystals to the first evaporation step so that the Kieserite crystals become dissolved and Langbeinite crystals are formed, thereby effecting the removal of sulphate ions from the solution as Langbeinite.

3. The method of recovering sulphate free magnesium chloride from an aqueous solution containing magnesium, potassium, chloride, and sulphate ions which includes the steps of, first, evaporating water from the solution under conditions such that the solution passes through a phase in which Langbeinite will crystallize out of solution and Kieserite will remain in solution, said evaporation being carried out at a temperature over about 60° C. and a concentration

of magnesium ions, measured as magnesium chloride, such that there are less than about 80 mols of magnesium chloride per 1000 mols of water, separating Langbeinite crystals from the solution while maintaining said solution at the aforesaid conditions; evaporating water from the resultant liquor till the concentration of magnesium ion, measured as magnesium chloride, is over about 80 mols of magnesium chloride per 1000 mols of water, maintaining the temperature of the liquor over about 70° C. and separating the Kieserite crystals which are formed from the solution; recycling the separated Kieserite crystals into the zone of the first evaporation so that the Kieserite crystals become dissolved and Langbeinite crystals are formed; mixing a potassium chloride solution with the liquor from which the Kieserite has been removed, evaporating water from the resulting solution at a temperature of about 25° C., whereby Carnallite crystals are formed; removing the Carnallite crystals from the liquor; recycling the Carnallite liquor to the first evaporation step; leaching said Carnallite crystals with

cold water to dissolve magnesium chloride from the Carnallite; and evaporating the water from the magnesium chloride solution to form sulphate free magnesium chloride crystals.

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WILLIAM A. SMITH.
PAUL D. V. MANNING.

REFERENCES CITED

The following references are of record in the file of this patent:

UNITED STATES PATENTS

Number	Name	Date
1,304,097	Reeve	May 20, 1919
1,305,566	Reeve	June 3, 1919

OTHER REFERENCES

Seidell, Solubilities of Inorganic and Organic Compounds, vol. 1, page 641 (1919), and vol. 2, page 1283 (1928) pub. by D. Van Nostrand Co., N. Y.

Appendix D. $MgCl_2$ Excel Material Balance

Manual Material Balance Excel Spreadsheet

Material Balances (kmol/hr)

	FEED	AERFED	NONCOMP	MIXFEED	EF1	BOTT1	PUMPBOTT1	WAT1	EF2
Temp (c)	25	50	50	44	44	60	60	60	63
Pressure (bar)	1	0.15	0.15	0.15	2	0.15	2	0.15	1
Solid Fraction	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.8
Sat. Pt (mol Nacl/mol H2O)	11.08	11.27	11.27	11.22	11.22	11.38	11.38	11.38	11.42
Sat. Pt. (mol KCl/ molH2O)	8.54	10.35	10.35	9.91	9.91	11.07	11.07	11.07	11.29
St. Pt. (molMgCl2/molH2O)	10.41	11.16	11.16	10.94	10.94	11.56	11.56	11.56	11.70
H2O	8000.0	7200.0	800.0	8202.9	8202.9	6562.3	6562.3	1640.6	11073.8
NA+	530.6	530.6	0.0	545.1	545.1	545.1	545.1	0.0	1264.8
NACL(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	115.1
CL-	530.6	530.6	0.0	530.6	530.6	530.6	530.6	0.0	1227.7
CO2	3.5	0.0	3.5	0.0	0.0	0.0	0.0	0.0	0.0
O2	4.8	0.0	4.8	0.0	0.0	0.0	0.0	0.0	0.0
MG++	80.0	80.0	0.0	102.6	102.6	102.6	102.6	0.0	759.7
MGCL2(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
K+	25.0	25.0	0.0	39.2	39.2	39.2	39.2	0.0	290.5
KCL(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
KMgCl3*6H2O	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
%mol Na	*6.63*	*7.37*	*0.00*	*6.65*	*6.65*	*8.31*	*8.31*	*0.00*	*11.42*
%mol Mg2+	1.00	1.11	0.00	1.25	1.25	1.56	1.56	0.00	6.86
%mol K	0.31	0.35	0.00	0.48	0.48	0.60	0.60	0.00	2.62
%mol Cl-	6.63	7.37	0.00	6.47	6.47	8.09	8.09	0.00	11.09
Total	9174.6	8366.2	808.4	9420.4	9420.4	7779.8	7779.8	1640.6	14731.6

Material Balances (kmol/

	BOTT2	PUMPBOTT2	WAT2	EF3*	BOTT3	PUMPBOTT3	WAT3	SLURRY1	SLURRY2
Temp (c)	90	90	90	90	115	115	115	115	90
Pressure (bar)	0.5	2	0.5	1.5	1.15	2	1.15	1.5	1.5
Solid Fraction	4.2	4.2	0.0	4.2	10.6	10.6	0.0	10.6	4.2
Sat. Pt (mol Nacl/mol H2O)	11.84	11.84	11.84	11.84	12.35	12.35	12.35	12.35	11.84
Sat. Pt. (mol KCl/ molH2O)	13.24	13.24	13.24	13.24	15.05	15.05	15.05	15.05	13.24
St. Pt. (molMgCl2/molH2O)	13.14	13.14	13.14	13.14	14.86	14.86	14.86	14.86	13.14
H2O	7751.6	7751.6	3322.1	5426.2	3255.7	3255.7	2170.5	2767.3	2325.5
NA+	917.6	917.6	0.0	642.3	402.1	402.1	0.0	341.8	275.3
NACL(S)	462.3	462.3	0.0	323.6	563.8	563.8	0.0	479.2	138.7
CL-	880.4	880.4	0.0	616.3	376.1	376.1	0.0	319.7	264.1
CO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MG++	759.7	759.7	0.0	531.8	531.8	531.8	0.0	452.0	227.9
MGCL2(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
K+	290.5	290.5	0.0	203.4	203.4	203.4	0.0	172.9	87.2
KCL(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
KMgCl3*6H2O	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
%mol Na	*11.84*	*11.84*	*0.00*	*11.84*	*12.35*	*12.35*	*0.00*	*12.35*	*11.84*
%mol Mg2+	9.80	9.80	0.00	9.80	16.34	16.34	0.00	16.34	9.80
%mol K	3.75	3.75	0.00	3.75	6.25	6.25	0.00	6.25	3.75
%mol Cl-	11.36	11.36	0.00	11.36	11.55	11.55	0.00	11.55	11.36
Total	11062.2	11062.2	3322.1	7743.6	5332.9	5332.9	2170.5	4533.0	3318.7

Material Balances (kmol/

	CLIQ***	SALTPROD	CARNAL	SALTPRD2	CARNAL2**	FEEDWAT	CRSTFED1	FILT1IN	FILTPUMP1	CARNLIQ
Temp (c)	25	25	100	25	25	25	35	25	25	25
Pressure (bar)	1	1	1.5	1	1	1	1	0.5	3	2
Solid Fraction	0.0	31.0	10.6	35.6	8.5	0.0	2.9	0.0	0.0	0.0
Sat. Pt (mol Nacl/mol H2O)	11.08	11.08	12.03	11.08	11.08	11.08	11.15	11.08	11.08	11.08
Sat. Pt. (mol KCl/ molH2O)	8.54	8.54	13.97	8.54	8.54	8.54	9.26	8.54	8.54	8.54
St. Pt. (molMgCl2/molH2O)	10.41	10.41	13.78	10.41	10.41	10.41	10.66	10.41	10.41	10.41
H2O	1744.1	581.4	488.4	146.5	341.8	872.9	1238.2	1202.8	1202.8	1002.9
NA+	13.8	64.4	60.3	16.2	14.5	0.0	16.4	16.4	16.4	14.5
NACL(S)	0.0	328.8	84.6	114.2	0.0	0.0	0.0	0.0	0.0	0.0
CL-	13.2	53.8	56.4	38.1	14.1	0.0	17.7	0.0	0.0	0.0
CO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MG++	205.1	22.8	79.8	4.0	35.6	0.0	35.6	29.7	29.7	22.6
MGCL2(S)	0.0	0.0	0.0	0.0	40.2	0.0	40.2	0.0	0.0	0.0
K+	78.4	8.7	30.5	1.5	29.0	0.0	50.8	44.9	44.9	14.2
KCL(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
KMgCl3*6H2O	0.0	0.0	0.0	0.0	0.0	0.0	0.0	5.9	5.9	0.0
%mol Na	*0.79*	*11.08*	*12.35*	*11.08*	*4.24*	*0.00*	*1.32*	*1.36*	*1.36*	*1.44*
%mol Mg2+	11.76	3.92	16.34	2.72	10.41	0.00	2.87	2.47	2.47	2.25
%mol K	4.50	1.50	6.25	1.04	8.48	0.00	4.10	3.73	3.73	1.42
%mol Cl-	0.76	9.26	11.55	25.98	4.12	0.00	1.43	0.00	0.00	0.00
Total	2054.7	1060.0	799.9	320.5	475.2	872.9	1398.8	1299.7	1299.7	1054.2

Material Balances (kmol/

	CARNLIQ	SOLDCARN	COLDWAT	FILT2IN	FILTPUMP2	FILTKCL	KCLRECYC	KCLCRYST	MGLIQ
Temp (c)	25	25	5	15	15	25	25	25	25
Pressure (bar)	2	2	1	0.5	3	2	2	2	2
Solid Fraction	0.0	5.5	0.0	0.0	0.0	39.1	39.1	28.1	0.0
Sat. Pt (mol Nacl/mol H2O)	11.08	11.08	11.02	11.04	11.04	11.08	11.08	11.08	11.08
Sat. Pt. (mol KCl/ molH2O)	8.54	8.54	7.09	7.81	7.81	8.54	8.54	8.54	8.54
St. Pt. (molMgCl2/molH2O)	10.41	10.41	10.08	10.21	10.21	10.41	10.41	10.41	10.41
H2O	1002.9	199.9	875.4	1230.2	1230.2	24.6	23.4	1.2	1205.6
NA+	14.5	1.9	0.0	1.9	1.9	1.9	1.9	0.0	0.0
NACL(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CL-	0.0	2.8	0.0	22.6	22.6	3.9	3.6	1.3	18.8
CO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MG++	22.6	7.1	0.0	13.7	13.7	0.0	0.0	0.0	13.7
MGCL2(S)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
K+	14.2	17.1	0.0	23.0	23.0	2.1	2.0	0.1	0.0
KCL(S)	0.0	13.6	0.0	0.0	0.0	20.9	19.8	1.0	0.0
KMgCl3*6H2O	0.0	5.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0
%mol Na	*1.44*	*0.94*	*0.00*	*0.15*	*0.15*	*7.65*	*8.06*	*0.00*	*0.00*
%mol Mg2+	2.25	3.56	0.00	1.11	1.11	0.00	0.00	0.00	1.14
%mol K	1.42	8.54	0.00	1.87	1.87	8.54	8.54	8.54	0.00
%mol Cl-	0.00	1.42	0.00	1.84	1.84	15.73	15.30	108.62	1.56
Total	1054.2	248.3	875.4	1291.3	1291.3	53.3	50.6	3.7	1238.0

Material Balances (kmol/

	CRSTFED2	WAT4	FILT3IN	FILTPUMP3	MGREYC*****	PRODUCT
Temp (c)	110	110	25	25	25	25
Pressure (bar)	1.424	1.424	0.5	3	2	2
Solid Fraction	0.0	0.0	2.0	2.0	0.0	91.5
Sat. Pt (mol Nacl/mol H2O)	12.24	12.24	11.08	11.08	11.08	11.08
Sat. Pt. (mol KCl/ molH2O)	14.69	14.69	8.54	8.54	8.54	8.54
St. Pt. (molMgCl2/molH2O)	14.49	14.49	10.41	10.41	10.41	10.41
H2O	120.6	1085.0	120.6	120.6	119.5	1.1
NA+	0.0	0.0	0.0	0.0	0.0	0.0
NACL(S)	0.0	0.0	0.0	0.0	0.0	0.0
CL-	18.8	0.0	21.1	21.1	2.1	0.0
CO2	0.0	0.0	0.0	0.0	0.0	0.0
O2	0.0	0.0	0.0	0.0	0.0	0.0
MG++	13.7	0.0	12.5	12.5	0.7	0.1
MGCL2(S)	0.0	0.0	1.2	1.2	0.0	12.9
K+	0.0	0.0	0.0	0.0	0.0	0.0
KCL(S)	0.0	0.0	0.0	0.0	0.0	0.0
KMgCl3*6H2O	0.0	0.0	0.0	0.0	0.0	0.0
%mol Na	*0.00*	*0.00*	*0.00*	*0.00*	*0.00*	*0.00*
%mol Mg2+	*11.37*	*0.00*	*10.41*	*10.41*	*0.57*	*10.41*
%mol K	*0.00*	*0.00*	*0.00*	*0.00*	*0.00*	*0.00*
%mol Cl-	*15.56*	*0.00*	*17.47*	*17.47*	*1.76*	*0.00*
Total	153.0	1085.0	155.3	155.3	122.3	14.1

*split fraction sett1	0.7	EF3
**split fraction sett2	0.85	SLURRY1
***%eff CENTRIFUGE1	0.05	NaCl Kept
***% K+ retention CENT1	0.9	K+ Kept
***% Mg2+ retention CENT2	0.9	Mg2+ Kept
****Split Fraction MGSPLIT	0.05	KCL CRYSTALS
*****%WAT RECYCLED	0.991	WATER IN RECYC
*****% Mg RECYCLED	0.05	MG in RECYCL
Vapor Fractions		
EVAP1	0.2	
EVAP2	0.3	
EVAP3	0.4	
DEAERATOR		
Vap Frac	0.1	
CENTRIFUGE 2		
Fraction NaCl Lost	0.9	
Fraction K+ Lost	0.05	
Fraction Mg2+ Lost	0.05	

Sample Code

Determining Cl^- composition in stream SOLDCARN ($MgCl_2$ Process Flowsheet)

```
=IF(AND(AE12=0, AE21=0),AC13-AD13,IF(AND(AE12=0, AE21<>0), AC13-AD13 - (AD20-AE20),
IF(AND(AE21 =0, AE12<>0), AC13-AD13-(0.115*AC11-AE6/100*AE9), IF(AND(AE12<>0, AE21<>0), AC13-
AD13 - (  (AD20-AE20) - (0.115*AC11-AE6/100*AE9))))))
```

The above sample code was used in the manually calculated Excel material balances to determine the Cl^- molar composition in stream SOLDCARN. As shown, four sets of conditional nested loops were used; in this particular case, the Cl^- molar flow was dependent on the molar flow of both $MgCl_{2(s)}$ and $NaCl_{(s)}$ present in solution, which further depend on the stream temperature and their respective saturation points. Conditional statements similar to this were utilized throughout the Excel based materials balance file.

Appendix E. MSDS Brine Solution

NOVA Chemicals®

Material Safety Data Sheet

Material Name: Brine Solution **MSDS ID: NOVA-0087**

Section 1 - Product and Company Identification

Synonyms: Salt water, Brine recycle stream, Sodium chloride solution
Chemical Name: Brine solution
Chemical Family: Mixture
Material Use: Operation of underground storage caverns and for salt manufacturing
Chemical Formula: Na^{+} (aq) Cl^{-} (aq); sodium chloride in solution

NOVA Chemicals
P.O. Box 2518, Station M
Calgary, Alberta, Canada T2P 5C6

EMERGENCY Telephone Numbers:
1-800-561-6682, 1-403-314-8767 (NOVA Chemicals) (24 hours)
1-613-996-6666 (Canutec-Canada) (24 hours)

Product Information: 1-412-490-4063
MSDS Information Email: msdsemail@novachem.com

Section 2 - Hazards Identification

NFPA Ratings: Health: 0 **Fire:** 0 **Reactivity:** 0
Hazard Scale: 0 = Minimal 1 = Slight 2 = Moderate 3 = Serious 4 = Severe

Emergency Overview
CAUTION! Product is a clear to cloudy white liquid with no odour. This product may be irritating to the eyes, skin, and respiratory system.

Potential Health Effects: Eye
This product may cause eye irritation. Symptoms may include itching, reddening, excess tearing and swelling.

Potential Health Effects: Skin
This product may cause drying, irritation and possible dermatitis.

Potential Health Effects: Ingestion
Ingestion of very large quantities may cause nausea, vomiting, dehydration, diarrhoea, oedema, and possible death. Prolonged over consumption may result in high blood pressure and heart problems.

Potential Health Effects: Inhalation
This product may cause irritation to the respiratory system.

Section 3 - Composition/Information on Ingredients

CAS No.	Component	Percent by Wt.
7732-18-5	Water	74-87
7647-14-5	Sodium chloride	13-26

Additional Information
This material is a controlled product under Canadian WHMIS regulations.
This material is not regulated as dangerous goods for transportation.

See Section 8 for applicable exposure limits. See Section 11 for applicable toxicity data.

Section 4 - First Aid Measures

First Aid: Eyes
Remove contact lenses, if it can be done safely. Immediately flush eyes with water for at least 15 minutes, while holding eyelids open. Seek medical attention if symptoms develop or persist.

First Aid: Skin
Remove contaminated clothing and shoes. Wash skin immediately with soap and water. Seek medical attention if symptoms develop or persist.

First Aid: Inhalation
Move affected individual to non-contaminated air. Loosen tight clothing such as a collar, tie, belt or waistband to facilitate breathing. Seek immediate medical attention if the individual is not breathing, is unconscious or if any other symptoms persist.

First Aid: Ingestion

DO NOT INDUCE VOMITING. Loosen tight clothing such as a collar, tie, belt or waistband. Seek immediate medical attention.

First Aid: Notes to Physician

Treat symptomatically. Treatment for overexposure should be directed at controlling the symptoms and clinical condition of the patient. Unless symptoms reappear, no further treatment is required. For more detailed medical emergency support information call 1-800-561-6682 or 1-403-314-8767 (24 hours, NOVA Chemicals Emergency Response).

Section 5 - Fire Fighting Measures

See Section 9: Physical Properties for flammability limits, flash point and auto-ignition information.

General Fire Hazards

Not a fire hazard. Does not burn.

Explosion Hazards

Not an explosion hazard.

Hazardous Combustion Products

None. Does not burn.

Extinguishing Media

Does not burn. Use extinguishing media suitable to surrounding fire conditions; e.g. dry chemical, foam, carbon dioxide, water fog or water spray.

Fire Fighting Equipment/Instructions

Firefighters should wear personal protective equipment suitable for the fire conditions and the materials burning.

Section 6 - Accidental Release Measures

Evacuation Procedures

Isolate area. Keep unnecessary personnel away.

Small Spills

Stop or reduce discharge if safe to do so. Prevent entry into water intakes and waterways. Remove liquid material with approved pumps or vacuum equipment.

Large Spills

Stop or reduce leak. Isolate, contain, and attempt to recover. Prevent entry into water intakes and waterways. Remove liquid material with approved pumps or vacuum equipment. Spill area may be washed down with water, with wash waters collected for testing and proper disposal.

Special Procedures

Contact local police/emergency services and appropriate emergency telephone numbers provided in Section 1. Ensure that statutory and regulatory reporting requirements in the applicable jurisdiction are met. Wear appropriate protective equipment and clothing during cleanup. Individuals without appropriate protective equipment should be excluded from area of spill until cleanup has been completed.

See Section 8 for recommended Personal Protective Equipment and see Section 13 for waste disposal considerations.

Section 7 - Handling and Storage

Handling Procedures

Material is slowly corrosive to metal. Handle in properly designed and approved equipment systems. Periodically inspect pipelines and other equipment for integrity and corrosion. Do not ingest or inhale. If ingested, seek medical advice immediately. Avoid contact with skin and eyes. Keep away from incompatible materials. After handling, always wash hands thoroughly with soap and water.

Storage Procedures

Storage area should be clearly identified, well-illuminated, clear of obstruction and accessible only to trained and authorized personnel. Adequate security must be provided so that unauthorized personnel do not have access to the product. Storage ponds and tank areas should be periodically inspected and kept separate from fresh water supply or outlets.

See Section 8: Exposure Controls/Personal Protection for appropriate Personal Protective Equipment. See Section 10 for information on Incompatibilities.

Section 8 - Exposure Controls / Personal Protection

Exposure Guidelines

A: General Product Information

Keep formation of airborne dusts or mists to a minimum. Ensure that eyewash stations and safety showers are in close proximity to the work locations.

B: Component Exposure Limits

ACGIH, NIOSH, Alberta and Ontario have not developed exposure limits for any of this product's components. Other exposure limits may apply, check with proper authorities.

ENGINEERING CONTROLS

Provide adequate ventilation to maintain worker exposure below levels that are irritating to the eyes or skin. Administrative (procedure) controls and use of personal protective equipment may also be required.

PERSONAL PROTECTIVE EQUIPMENT

Personal Protective Equipment: Eyes/Face

Chemical goggles are recommended. If splashing is possible use chemical goggles and a full-face shield. Carefully rinse off contaminated goggles before removing.

Personal Protective Equipment: Skin/Hands/Feet

Use chemically resistant gloves when handling product. Wear chemical-resistant safety footwear with good traction to prevent slipping. Work clothing that sufficiently prevents skin contact should be worn, such as coveralls and/or long sleeves and pants. If splashing or contact with liquid material is possible, consider the need for an impervious overcoat.

Personal Protective Equipment: Respiratory

If engineering controls and ventilation are not sufficient to prevent buildup of aerosols or vapours, appropriate NIOSH approved respiratory protection should be used.

Personal Protective Equipment: General

Personal protective equipment (PPE) should not be considered a long-term solution to exposure control. Employer programs to properly select, fit, maintain, and train employees to use equipment must accompany PPE. Consult a competent industrial hygiene resource, the PPE manufacturer's recommendation, and/or applicable regulations to determine hazard potential and ensure adequate protection.

Section 9 - Physical & Chemical Properties

Physical State and Appearance:	Clear/Cloudy Liquid	**Colour:**	Clear to white
Odour	Odourless	**pH:**	Range: 6.5 to 8.5
Vapour Pressure:	Not applicable	**Vapour Density at 0°C (Air=1):**	Not applicable
Boiling Point:	>100°C	**Freezing Point:**	-10°C
Solubility (H2O):	Miscible (water-based solution)	**Specific Gravity (Water=1):**	1.2 at 15°C
Auto Ignition:	Not applicable	**Flash Point:**	Not applicable
Flash Point Method:	Not applicable	**Upper Flammable Limit (UFL):**	Not applicable
Lower Flammable Limit (LFL):	Not applicable	**Flammability Classification:**	Non-flammable

Section 10 - Stability & Reactivity Information

Chemical Stability

This product is a stable material.

Chemical Stability: Conditions to Avoid

None identified.

Incompatibility

In presence of air, liquid contact or mists will slowly corrode most metals.

Possibility of Hazardous Reactions or Hazardous Polymerization

Hazardous polymerization will not occur.

Corrosivity

Corrosive to most metals upon prolonged contact.

Hazardous Decomposition

None identified. Does not burn.

Section 11 - Toxicological Information

A: Acute Toxicity - General Product Information

This product has not been tested.

B: Acute Toxicity - LD50/LC50

Water (7732-18-5)

Oral LD50 Rat: >90 mL/kg

Sodium chloride (7647-14-5)

Inhalation LC50 Rat: >42 g/m3/1H; Oral LD50 Rat: 3 g/kg; Dermal LD50 Rabbit: >10 g/kg

C: Chronic Toxicity - General Product Information

This product has not been tested.

D. Chronic Toxicity - Carcinogenic Effects

None of this product's components are listed by ACGIH, EPA, IARC, NIOSH, NTP, or OSHA as a carcinogen.

Section 12 - Ecological Information

Ecotoxicity

A: General Product Information

This product has not been tested. A concentrated brine solution (~26% sodium chloride) will dehydrate animal and vegetative species. Sodium chloride is practically non-toxic to aquatic organisms.

B: Component Analysis - Ecotoxicity – Aquatic/Terrestrial Toxicity

Sodium chloride (7647-14-5)

Test and Species	Results and Conditions
96 Hr LC50 Lepomis macrochirus	5560-6080 mg/L [flow-through]
96 Hr LC50 Lepomis macrochirus	12,946 mg/L [static]
96 Hr LC50 Pimephales promelas	6020-7070 mg/L [static]
96 Hr LC50 Pimephales promelas	7050 mg/L [semi-static]
96 Hr LC50 Pimephales promelas	6420-6700 mg/L [static]
96 Hr LC50 Oncorhynchus mykiss	4747-7824 mg/L [flow-through]
48 Hr EC50 Daphnia magna	1000 mg/L
48 Hr EC50 Daphnia magna	340.7 - 469.2 mg/L [static]

Environmental Fate/Mobility

This product has not been tested. Brine does not partition to air. When spilled into a body of water, the brine will disperse in and mix with the water. A large brine spill into a body of water could result in stratification with the water floating on top of the brine. Eventually the two will mix. When spilled onto soil, brine will behave similar to spilled water. Sodium chloride may leach from soil into groundwater.

Persistence/Degradability

This product has not been tested. Brine (sodium chloride) is not biodegradable.

Bioaccumulation/Accumulation

This product has not been tested.

Section 13 - Disposal Considerations

This product may meet the definition of a hazardous waste according to Canadian regulations. The use, mixing or processing of this product may alter its properties or hazards. Contact federal, provincial and local authorities in order to generate or ship a waste material associated with this product to ensure materials are handled appropriately and meet all criteria for disposal of hazardous waste.

See Section 7: Handling and Storage and Section 8: Exposure Controls/Personal Protection for additional information that may be applicable for safe handling and the protection of employees.

Section 14 - Transportation Information

Canadian TDG Information

Shipping Name: NOT REGULATED as a Dangerous Good for Transportation.

Section 15 - Regulatory Information

Canadian Regulations - Federal and Provincial

Canadian Environmental Protection Act (CEPA): This product is a mixture of naturally-occurring substances. All components are on the Domestic Substances List (DSL), and are acceptable for use under the provisions of CEPA.

Ingredient Disclosure List (IDL)

No components are listed under the Canadian Hazardous Products Act - Ingredient Disclosure List (IDL).

WHMIS Classification

Workplace Hazardous Materials Information System (WHMIS): This product has been classified in accordance with the hazard criteria of the CPR (Controlled Products Regulations) and the MSDS contains all the information required by the CPR.

WHMIS CLASS D2B: Toxic (Skin/eye irritant)

Other Regulations

Ongoing occupational hygiene, medical surveillance programs, site emission or spill reporting may be required by federal or provincial regulations. Check for applicable regulations.

For additional regulatory information, please contact your NOVA Chemicals' representative or NOVA Chemicals' Product Integrity group.

Section 16 - Other Information

Label Information

CAUTION! Product is a clear to cloudy white liquid with no odour. This product may be irritating to the eyes, skin, and respiratory system.
FIRST AID:
SKIN: Remove contaminated clothing and shoes. Wash immediately with soap and water. Seek medical attention if symptoms develop or persist.
EYES: Remove contact lenses, if it can be done safely. Immediately flush eyes with water for at least 15 minutes, while holding eyelids open. Seek medical if symptoms develop or persist.
INHALATION: Move affected individual to non-contaminated air. Loosen tight clothing such as a collar, tie, belt or waistband to facilitate breathing. Seek immediate medical attention if the individual is not breathing, is unconscious or if any other symptoms persist.
INGESTION: DO NOT INDUCE VOMITING. Loosen tight clothing such as a collar, tie, belt or waistband. Seek immediate medical attention.
IN CASE OF LARGE SPILL: Stop or reduce leak. Isolate, contain, and attempt to recover. Prevent entry into water intakes and waterways. Remove liquid material with approved pumps or vacuum equipment. Spill area may be washed down with water, with wash waters collected for testing and proper disposal.

References

Available on request

Key/Legend

ACGIH = American Conference of Governmental Industrial Hygienists; CAS = Chemical Abstracts Service; CEPA = Canadian Environmental Protection Act; CPR = Controlled Products Regulations; DSL = Domestic Substances List; EC50 = Effective Concentration 50%; EPA = Environmental Protection Agency; IARC = International Agency for Research on Cancer; IDL = Ingredient Disclosure List; IDLH = Immediately Dangerous to Life or Health; Kow = Octanol/water partition coefficient; LC50 = Lethal Concentration 50%; LD50 = Lethal Dose 50%; LEL = Lower Explosive Limit; LFL = Lower Flammable Limit; MSDS = Material Safety Data Sheet; NDSL = Non-Domestic Substances List; NFPA = National Fire Protection Association; NIOSH = National Institute for Occupational Safety and Health; NTP = National Toxicology Program; OEL = Occupational Exposure Limit; OSHA = Occupational Safety and Health Administration; PNOC = Particulates Not Otherwise Classified; PPE = Personal Protective Equipment; SCBA = Self Contained Breathing Apparatus; STEL = Short Term Exposure Limit; TDG = Transportation of Dangerous Goods; TLV = Threshold Limit Value; TWA = Time Weighted Average; UEL = Upper Explosive Limit; UFL = Upper Flammable Limit; WHMIS = Workplace Hazardous Materials Information Systems

MSDS Prepared By: NOVA Chemicals
MSDS Information Phone Number: 1-412-490-4063

[33]

Other Information

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This is the end of MSDS # NOVA-0087.