Nueces Desalination Center: Production of Drinking Water by Multi-Stage Flash Distillation

From processdesign

Team F: Rankine 672 Final Report

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Executive Summary

Desalination of seawater to drinking water is a technology of ever-increasing importance given trends of increasingly widespread water shortages around the world. Even in the United States, water shortages are becoming increasingly common in dry, highly-populated such as California and certain areas of the Gulf Coast. Our design team, Rankine 672, was tasked with evaluating the need for and designing a desalination plant to provide purified water in a place of need. Investigation revealed that the city of Corpus Christi, Texas, is a strategic location for development of a desalination facility, as surrounding region is in need of new drinking water sources and the city has expressed serious interest in contracting the development of a large-scale desalination plant. Our proposed solution is a multi-stage flash distillation desalination facility capable of producing 20,000,000 gallons (75,708 m^3) of purified drinking water for the city and surrounding region per operating day.

In this design, drinking water is produced from seawater feedstock through a 21-stage flash distillation process. By strategically locating the facility next to an existing natural gas power plant, the Nueces Bay Energy Center, pre-heated feed water can be taken directly from the outbound cooling systems of the power plant, allowing energy savings for our desalination plant. The process is designed for 16 hour-per-day operation, requires a seawater feed of 8250 m^3/hr, and produces 4758 m^3/hr of purified drinking water alongside a waste stream of 3485 m^3/hr of concentrated brine. The waste stream is isolated from a recycle stream through a 9.0% purge; this results in an overall product yield of 57.7% on a volume basis. The process requires an additional pretreatment stream of 16.5 kg/hr Belgard EV 2030, which serves as an antiscalant, as well as 9.5 kg/hr chlorine, 352.1 kg/hr lime, and 418.7 kg/hr CO_2 for post-treatment purposes.

The overall capital cost of the plant is expected to be $345MM, with a total operating cost of $103MM/yr, $59MM/yr of which comes from utilities. Due to the low price at which the product water can be sold, the 30-year net present value of the plant, assuming a 6% discount rate, is $-1,035MM. It is thus imperative this plant be constructed with public funds because it cannot return a profit to any private investor. The growing need for water however could justify the cost, because it is important that the residents of the greater Corpus Christi area have access to drinkable water.

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Introduction

Background

Clean drinking water is a basic human need that is becoming increasingly difficult to meet. Currently, over 1 billion people live in places where water is considered scarce[1]. Though often thought of as a problem exclusive to third-world countries, the issue is becoming more prevalent in the US as well, with water managers in 40 of 50 states anticipating water shortage conditions in some portion of their state in the next 10 years. These shortages are due to increasing demands on constant or diminishing freshwater resources, therefore, to correct the issue, a new supply must be found[2]. Desalination of seawater is therefore important as a new source for clean drinking water and prevent water shortages.

Project Definition

The overall goal of the project is to investigate the feasibility of a desalination plant in order to meet the needs of growing water demand. The design needs to take health, safety, and environmental concerns into account, while optimizing costs.

Design Basis

Site Location and Conditions Corpus Christi, TX was chosen as the location for the proposed desalination plant. The city is currently experiencing “Moderate Shortage Conditions”. This issue additionally comes at a time of growing demand with population growth anticipated due to new industry and Eagle Ford Group crude oil production[3].

In June 2014, the city announced the Corpus Christi Desalination Demonstration Project[4] which seeks to create a demonstration desalination plant capable of producing 20,000 gallons of water per day, with plans to later build a full scale facility generating 20,000,000 gallons of water per day. The desalination demonstration project indicates a clear need for implementation of a desalination plant of this size to accommodate growth in the Corpus Christi area.

Locating our plant in Corpus Christi will allow us to take advantage of the benefits of cogeneration through a partnership with the Nueces Bay Energy Center. By utilizing their outlet cooling water as a large portion of our feed for the desalination process, we can raise the temperature of our seawater feed, saving on energy costs in the multi stage flash distillation. The proposed location for the desalination plant is immediately west of the Nueces Bay Energy Center. See Appendix A for a map depicting the proposed location for the desalination plant along the coast of Nueces Bay.

Feed and Product Definitions

The proposed desalination facility’s feed and product streams are defined as shown in Table 1. Rationale for these definitions is also described below. The values reported in Table 1 are final derived numbers from HYSYS simulation modeling.

Table 1: Feed and product stream definitions

	Salt Concentration	Temperature	Amount
Feed Stream	30,500 ppm	45°C	8250 m^3/hr
Product Stream	~ 0 ppm	35°C	4758 m^3/hr
Byproduct Stream	69,500 ppm	35°C	3485 m^3/hr

The process feed will be seawater from Nueces Bay, preheated through the cooling water system of the nearby Nueces Bay Energy Center. Based on calculations developed from NREL (National Renewable Energy Laboratory) data, it was estimated that the Nueces Bay Energy Center uses approximately 250 million gallons of seawater per day for process and cooling needs. Therefore, the 20 million gallon per day requirement for the desalination facility can easily be met through the power plant’s cooling water[5]. Note that we are assuming the process will operate 67% of the time (16 hours per day).

Based on data acquired from two salinity-measuring stations in Nueces Bay, the average salinity of local seawater is 30.5 practical salinity units (psu)[6], thus setting the concentration of the process feed stream. This translates to a feed stream concentration of 30,500 ppm (parts per million)[7]. The composition of the salt attributed to this seawater salinity is shown in Table 6 in Appendix B.

No data on the composition of seawater around the Corpus Christi area was available, but the composition of seawater does not change significantly in different locations. The given composition is considered accurate for our feed stream[8]. Note that seawater salt is primarily sodium chloride - total dissolved solids are 85.7% sodium chloride by mass. As such, the dissolved solids in the feed stream is assumed in this project to be pure NaCl, making the feed composition 96.95% water and 3.05% NaCl by mass. A control system will likely be put in place to ensure feed concentration remains at this level, accounting for natural fluctuations in local seawater salinity.

The largest byproduct from the process will be a brine slurry with high salt concentrations. An upper corrosion limit of 700 ppm was set for recirculated and purged brine. This undesired byproduct will be sent to a nearby Class I non-hazardous injection well facility operated by GNI Group, which is approximately 12 miles away from the proposed desalination plant location as shown in the figure in Appendix A[9][10]. Through the incorporation of a Class I injection well, the brine will not have to be treated or diluted before injection. [11][12]

There are no regulations regarding the temperature of city drinking water, so the product stream temperature has no requirements based on regulations. Outlet temperatures of 35^{o}C are acceptable for input to the municipal water supply system. Since city water tends to be stored in non-temperature controlled tanks, the temperature will naturally cool to ambient temperature.[13]

Pre-Treatment and Post-Treatment

For pretreatment, Belgard EV 2030 is recommended to be added at a concentration of 1.8-2.0 mg/L for a system with top brine temperature approximately 110 °C[14]. Since our process is designed to operate at a slightly higher top brine temperature, 120 °C, we will design for the upper bound concentration: 2.0 mg/L of antiscalant. With a feed flow rate of 8250 m^3/hr, the process therefore requires 16.5 kg/hr of Belgard EV 2030.

For disinfection during post-treatment, chlorine addition typically requires a dose of roughly 2.0 mg/L as well[15]. With a pure water distillate flow rate of 4758 m^3/hr, the process requires 9.5 kg/hr of chlorine as a 5%-15% sodium hypochlorite solution. Approximately 74 mg/L of lime (calcium hydroxide) and 88mg/L of CO_2 should be added to the desalinated water to help prevent corrosion and increase alkalinity within acceptable ranges[16]. With a product stream flow rate of 4758 m3/hr, these concentrations correspond to 352.1 kg/hr of lime and 418.7 kg/hr of CO_2. Both of these can be purchased, and it is likely that the CO_2 can be procured from the neighboring natural gas power plant.

Choice of Desalination Technology

The two major technologies commonly implemented for desalination include reverse osmosis (RO), and multi-stage flash distillation (MSFD). RO utilizes high pressures to force water through a semipermeable membrane that allows for purified water to pass through, but stops organic and other wastes. Often, the pressures can reach 55 - 68 bar, which requires large pumps to maintain pressure, and introduces safety concerns to the process[17]. While RO can be less energy intensive than MSFD, and the membrane can also remove contaminants, it is susceptible to feedwater quality, membrane fouling, and requires more extensive pretreatment and posttreatment than MSFD. The cost of RO is directly proportional to the feed salinity, and so it is less expensive when desalinating low salinity water.

Multi-stage flash distillation is best selected for areas that have high salinity or low energy costs. The pretreatment process for MSFD is much easier when compared to RO because pretreatment often consists of simply adding anti-corrosives and antiscalants to the feed stream. The process works by heating water in vessels at different pressures to change the boiling point of the water and evaporate it as it passes through the system. After evaporation, the water condenses, leaving pure water as the effluent. MSFD costs are now roughly $1.00/m3, where this cost is mostly independent of salinity, but dependent on energy costs.[18] In both RO and MSFD a brine is produced which then must be appropriately treated and/or disposed of.

Multi-stage flash distillation was selected for this process, mainly due to the economic advantages offered by the selected location of Corpus Christi, TX. MSFD is also a well understood process that can be more easily modeled by chemical engineering software.

Technical Approach

Quantitative modeling of the process and all included optimization was conducted using the simulation software Aspen HYSYS. A variety of assumptions were developed to set up and define specific components of the process simulation itself. Within Aspen HYSYS, the NRTL Electrolyte property package was selected to model the behavior of salt water. Each stage of the MSF process is modeled as an individual flash vessel, with individual shell & tube heat exchangers used to model the distillate condensers present in each flash stage in the MSF system.

The HYSYS simulation model only considers the flash portion of the overall process, allowing establishment of mass and energy balances for the system. Pretreatment and post-treatment were excluded for the HYSYS model, as these steps are simple additions or removals of small concentrations of chemical components. Although the simulation models heat exchangers outside of each flash stage, in the physical design of the plant these heat exchangers will be contained within each flash stage.

One portion of the process is not modeled, which is the inlet seawater entering the heat rejection stages. As can be seen from the full process flow diagram (Appendix C), the inlet seawater in the heat rejection stage is in high excess to provide maximum cooling capacity; then a significant portion is purged to bring the stream flow rate down to what is needed for the feed stream. It is assumed that the flow rate of cooling water for those three stages is in such high excess that the cooling water does not heat up. Therefore, in the process simulation models, the inlet stream is the feed stream, which enters directly into the heat recovery portion of the process.

See Appendix D for a screenshot overview of the simulation environment model.

Process Overview

Process Flowsheet

The following diagram, Figure 1, shows a basic flowsheet for the process. Process alternatives considered for each stage are discussed below. Please see Appendix C for the full process flowsheet with finalized design decisions.

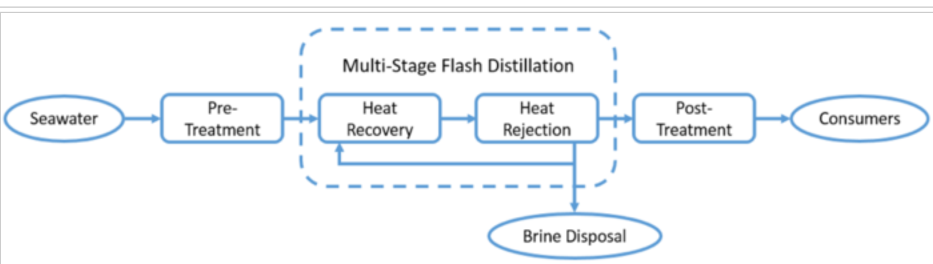

Figure 1: Basic block flowsheet of the multi-stage flash desalination process.

Major Features

As seen above, there are three primary stages of the desalination process: Pre-treatment, the multi-stage flash distillation, and post-treatment. Seawater enters the pre-treatment phase first to make the feed suitable for the rest of the process by filtering out large particles that could accumulate on pipes, deaerating the water to limit corrosion, and adding an agent to combat fouling. The water then enters the distillation process, which is broken up into two sections: heat recovery and heat rejection.

In the heat recovery stage, there are 18 flash vessels, each of which separates steam from the salinated liquid water. Each flash vessel has saltwater as its inlet and two outlets: a vapor outlet (steam) that condenses into the final product of purified water and a liquid outlet composed of water with more dissolved solids than the inlet, which is the inlet stream for the next flash vessel. The pipe containing the feed is first mixed with recycled brine, then goes through the top sections of each of the flash vessels, starting with V-118 and ending up at V-101. In this way, the feed stream pipe first serves as a heat exchanger, heating the pre-treated seawater while simultaneously condensing the steam into the final product. The feed then enters a final heat exchanger where it is brought up to 103.6 °C before entering V-101 to be flashed.

After exiting V-118, the final flash vessel in the heat recovery portion, the water enters the heat rejection portion, which consists of three flash vessels (V-201, V-202, V-203). The flash vessels of this section operate in the same manner as in the heat recovery section, with the exception that condensation occurs using excess cooling water instead of pre-treated seawater. This technique allows excess heat to be rejected into an excess water stream rather than the feed.

The purified product stream finally enters post-treatment processing in which water is re-aerated and remineralized, disinfected, and pH is adjusted before being delivered to the city. The concentrated brine leftover at the end of the flash cycle is recycled into the feed, although a purge is required to continuously remove some salt from the system. This purged brine is disposed of by use of a Class I injection well, per regional regulations.

See Appendix E to find detailed stream tables of stream specifications.

Process Alternatives

Pretreatment

Multi-stage flash distillation requires significantly less pretreatment than reverse osmosis desalination, but helpful pretreatment techniques still exist. One very inexpensive method of pretreatment is screening/filtration, which is used to filter out large contaminants from intake seawater. Using more complicated membrane technology to accomplish large particle filtration is a possible alternative, but doing so would be much more expensive than using a standard screen or filter, so the latter choice has been selected for this process.[19]

The most critical type of fouling in multi-stage flash distillation is scaling - therefore, the most critical step in the pretreatment process will be the inclusion of an anti-scalant. Alkaline scaling can be prevented by acid dosing or polyphosphate addition; however, acid dosing can result in increased corrosion and polyphosphates can lead to non-alkaline calcium phosphate fouling, so these additives will not be considered[20][21]. As such, a polymer-based additive is recommended for the anti-scalant for this process. The Belgard EV series is designed for use in high-temperature seawater MSF; of these, Belgard EV 2030 is the most neutral in pH[22][14][23][24]. As such, we have selected Belgard EV 2030 for use as an anti-scalant in our process.

Corrosion control is another significant method of pretreatment in multi-stage flash distillation plants. Using a piping material that has limited susceptibility to corrosion is typically not economically viable for MSF desalination plants due their large size, so using carbon steel for most of the piping is generally used for its economic benefits despite its susceptibility to corrosion[20]. Limiting oxygen concentration is very feasible through the use of deaerators, which can remove oxygen as well as other dissolved gases from the seawater stream. Often, an external deaeration system is used[20][25]. Given these considerations, using an external deaerator combined with a less-corrosive antiscalant like Belgard EV 2030 is much more practical for the process than using anti-corrosive piping.

Excessive foam can also be an issue in some MSF distillation facilities, so antifoaming agents are sometimes employed in the pretreatment process as well. However, many plants are operated without foaming issues, and therefore do not need anti-foaming pretreatment at all. Selection of an anti-foaming agent must be carefully weighed with selected of anti-scalant to avoid issues with additive interactions [26]. At this stage in the design process, an anti-foamer will not be included due to the risk of decreasing scalant efficacy, and the likelihood that foam will not be an issue in the desalination plant.

Multistage Flash Distillation

The main objective of the design project is desalinate water. This is done with the main process of multi stage flash distillation (MSF). MSF can be developed using a numerous number of different process configurations, such as Once-Through, Simple Mixer Brine Circulation, and Multi-Stage Heat Rejection Brine Circulation processes.

The most desired process for the project is the Three Stage Heat Rejection Brine Circulation as it builds on benefits of the other simpler technologies. By combining principles of heat rejection and heat recovery, the required heat transfer area of the process is greatly reduced as compared to once through processes or simple brine recycle processes. It has the highest performance ratio with the lowest specific feed flow rates, therefore greatly lowering the operating costs of the process. It is also known as the industry standard for MSF[26]. In addition, since our brine disposal technique involves the pumping of the brine into a Class I well, the high brine outlet salinity is not a drawback for our process since post treatment or dilution of the brine is not necessary.

Post Treatment

After processing the water via multi-stage flash, the water needs to be post-treated to make the water taste as consumers would expect, since distilled water is "flat" and unpleasant, as well as to protect the consumers and the piping through which the water will be passing[27]. There are multiple alternatives to achieve these objectives of remineralization, aeration, corrosion control, and disinfection.

Aerating and introducing some dissolved solids back into the product water are effective ways to create a quality product that is not flat, and contains flavoring that is expected of the water when delivered to the consumer. Aeration will be accomplished via an inline spray aerator which will take up less space than cascade aeration.[28] Feed blending will be used to remineralize the water as opposed to a remineralization filter because it does not require additional equipment purchases, and can easily handle the large amounts of throughput required for the process.[27]

Corrosion and pH can be controlled by adding lime (calcium hydroxide) and carbon dioxide to the water. To do so, about 74 mg/L of lime and 88mg/L of CO_2 should be added to the desalinated water to help prevent corrosion and increase alkalinity within acceptable ranges. Sodium hydroxide, soda ash, and sodium bicarbonate are alternatives to altering the alkalinity and pH of the product water, but lime and CO_2 is an industry standard [16].

Disinfection needs to be done to prevent microbial growth in the desalinated water and ensure the safety of the general public. Chlorine addition via chlorine gas or sodium hypochlorite is the most common and only requires a dose of 1.5 to 2.5 mg/liter. [15] Other alternatives include ozonation, UV light, and chloramination. Chlorine, as a 5% - 15% sodium hypochlorite solution, is sufficient for our process and relatively inexpensive, and as such will be utilized in the process.

Transport of Brine to Well

Three major possibilities exist for transporting the brine from our desalination plant to the injection well for disposal: rail, trucking, and a pipeline. A pipeline was chosen as the most feasible and cost-effective approach.

A rail line is not available at this time, and with the cost to construct a rail line at approximately $43 million per mile[29], the enormous capital cost makes rail transport infeasible for this site. For large volume operations, pipelines are cheaper than trucking. In a study of wastewater sludge produced from wastewater treatment facilities, it was found that it was more economical for facilities producing over 1200 m^3 per day to utilize pipelines rather than trucking.[30] Our facility is producing nearly 3 times this volume at 3485 m^3 per day of brine. Furthermore, brine is a less viscous fluid than wastewater sludge, decreasing pumping costs in the pipeline and further amplifying the benefit of choosing a pipeline over an injection well.

Brine Storage

Due to the choice of a pipeline, rather than trucking, as the method of transport of brine to the injection well, brine storage is unnecessary, so storage solutions such as holding tanks were not investigated. One method of storage compatible with a pipeline is to pump the brine through an evaporation pond. This solution is attractive as it would allow the brine to concentrate, saving some costs of pumping, however, the considerable land usage of evaporation ponds leads them to only be effective for smaller plants. The largest municipal plant utilizing this method produces only 5.7 million liters per day[31], whereas our desalination plant is proposed to produce 75.7 million liters per day. Therefore, evaporation ponds have been eliminated from consideration.

Major Specifications

Material of Construction Selection

While saltwater and desalinated water in general can be corrosive, there are best practices within the desalination industry for material selection. With this in mind, best practice shows carbon steel lifetime can be lengthened by monitoring the oxygen concentration in the water, specifying the paint used to coat any metal, and more expensive materials can be used in areas of extremely high corrosion.[20]. By following these best practices, it is possible to save significant amounts of money through the use of carbon steel in most areas of the plant, with specific high-corrosion areas being reinforced with, or replaced with, more corrosion resistant materials, such as Cu/Ni 70/30 alloy[26].

Heat Exchanger Sizing

See Appendix F for heat exchanger sizing data. Heat exchanger sizes appear reasonable for placement in the flash vessels.

Flash Vessel Sizing

The dimensions for the flash drums were obtained through literature values. The flash drums were determined to be 18x4x3 m in width, height, and length.[26]

Wall thickness was determined using the following conventions: Design pressure is 10% more than the operating pressure, design temperature is 50°F (27.8°C) higher than the operating temperature, a corrosion allowance of 3-mm and an effectiveness of 85% was derived from the assumption of a double-welded butt joint that was spot checked. See Appendix G for wall thicknesses calculated for each pressure vessel.

The sizing of the vessels to be constructed will be determined after talks with a manufacturer. Due to economies of scale, it may be more cost-effective to buy 21 pressure vessels with the dimensions mentioned above and a wall thickness of 17.1-mm, the highest wall thickness required in the system.

Pipe Losses and Pressure Drop

HYSYS was used to estimate the diameter of a pipe given a specified pressure drop per 100m of pipe. It was determined that for a pressure drop of 0.1 kPa/100m of pipe, a 3.05m diameter pipe is sufficient, which is more than manageable for manufacturers. This pressure drop is mostly insignificant and will be taken care of with the safety factors included in the pumps. The significant pressure drops in our system is accounted for via losses in the flash vessels and heat exchangers, which drive the flash process. Therefore overall pipe losses are insignificant and are built into intentional design losses.

Economic Analysis

Economic Assumptions

The sale of purified water yields a revenue stream of $49.5 million per year assuming sale to commercial entities within the city limits of Corpus Christi at $1.7915 per thousand liters.[32] The primary variable costs are utilities and the brine disposal, with minimal raw materials costs due to the primary feedstock, seawater, being free, and minimal amounts of pre- and post-treatment chemicals being used. The cost of brine disposal used in this analysis was high at $2.097 per thousand liters[33], an estimate for brine disposal utilizing trucking as the mode of transport. As has been shown through previous analyses, a lower cost can be achieved using a direct pipeline, however, no good estimates for this cost could be found.

The NPV was calculated using a 6% cost of capital, the approximate interest rate for a municipal bond.[34] It was assumed that the plant can be built within two years, with production at half-scale in its first year. Additionally, one month of variable costs, $10.6 million, was set aside for working capital. See Appendix H for a full table of assumptions. A 30 year NPV calculation was chosen as the decision basis since 30 years is a conservative estimate for the lifetime of a desalination plant [35]. See below for estimates of NPV for the lifetime of the plant under various assumptions and design criteria.

Optimization and Sensitivity Analysis

Looking at the multi-stage flash process, there are a number of opportunities to optimize the process. The major process variables that were investigated were the inlet water temperature, the operating pressure drop per vessel, and the number of stages over which the process takes place. Considering these optimizations, our final process consists of 21 flash vessels (18 heat recovery and 3 heat rejection stages), each with a pressure drop of 5 kPa. See Appendix D for a screenshot of the HYSYS simulation used to model the optimized process.

Operating Pressure and Pressure Drop

Initially our process was designed to pump the inlet seawater and brine to a pressure of 2 bar and allow the pressure drop in each flash stage to end our process downstream at 1 bar to prevent complications associated with designing a system under vacuum. While this provided pumping advantages, the energy in the system was significantly higher due to vapor pressure considerations. Due to this fact, the required heat exchange area for certain exchangers proved to be infeasible (on the order of 80,000m^2) due to the enormous capital and operating costs that would be associated with such a large heat exchanger.

The process was optimized to operate under vacuum, requiring additional pumping requirements but much lower required heat exchange area, and therefore lower utility costs. This optimization was done by specifying the pressure of the final flash vessel to 5.7kPa, so that the temperature of the water vapor (~30^oC) is within approximately 5^oC of the cooling water (35^oC) so that condensation can occur within a reasonable approach. This also allows our system to optimally extract pure water from the brine running through the system and reduce energy requirements for cooling the product or the recycle. Our system redesign and optimization reduced the energy in the system and reduce the same heat exchanger area requirements by more than 3 orders of magnitude to approximately 800m^2.

Once the process was optimized to operate in a vacuum, the pressure drop per flash stage regulated through values was then optimized. Three pressure drops per stages, 3kPa, 5kPa, and 7kPa, were tested and then an economic analysis was conducted to find the optimal value for the pressure drop. These changes had impacts on both operating and capital costs as they would change initial pumping requirements, the amount of water flowing through the process and recycle (and therefore the flash vessel sizes, heat exchanger sizing), and the efficiency of each stage.

The various pressure drops (3 kPa, 5 kPa, and 7 kPa) were tested on a system nominally at 24 stages under the assumption that the results would be generalizable to systems with a different number of stages. With increasing pressure drop, we found decreased capital cost but increased utility cost. As can be seen from Table 2 below, a pressure drop of 5 kPa proved to be the best option based on 30 year NPV, with a loss of $838.8 million, a savings of $65.5 million compared to the next-best option with a pressure drop of 7 kPa.

Table 2: Comparison of profitability for our MSF desalination process with 24 stages with 3 different pressure drops per stage

Number of Stages	Pressure Drop (kPa)	Capital Cost ($MM)	Utility Cost ($MM/yr)	30-Year NPV ($MM)	NPV Ratio (over 5kPa)
24	3	362.3	39.1	-905.8	1.08
24	5	255.1	53.6	-838.8	1.00
24	7	209.5	72.2	-904.3	1.08

Number of Flash Stages

The number of stages is an important area for optimization. An increased number of stages yields better recovery of purified water, however, requires higher capital cost from the additional pressure vessel. Furthermore, with an increased amount of purified water comes increasingly concentrated brine. This must be taken into account since any brine concentration in excess of about 7% presents a more significant corrosion risk[26], so increased maintenance costs for the piping or higher capital costs for more corrosion-resistant materials could be necessary for brine concentrations exceeding 7%. Therefore, to recover the same amount of pure water without exceeding corrosion limitations, increasing the number of stages will reduce the overall size of each stage.

Optimizing based on NPV, it was found that 21 stages is optimal, balancing the increased profits from the additional purified water additional costs of higher capital cost. See Table 3 below to compare the 30-year NPV of the process across various numbers of flash vessels.

Table 3: Comparison of profitability for our MSF desalination process with a 5 kPa pressure drop, 5000 m^3/hr feed flow rate, with varying number of stages.

Number of Stages	Pressure Drop (kPa)	Capital Cost ($MM)	Utility Cost ($MM/yr)	30-Year NPV ($MM)	NPV Ratio (over 21 Stages)
17	5	213.9	58.7	-813.8	1.016
18	5	218	57.6	-809.9	1.011
19	5	224.3	58	-806.1	1.007
20	5	229.8	57.9	-813.6	1.016
21	5	226.6	55.9	-800.7	1
22	5	254.7	54.6	-849.5	1.061
23	5	249.3	54.1	-833.1	1.04
24	5	255.1	53.6	-838.8	1.048
25	5	254.9	52.7	-837.4	1.046

Inlet Seawater Temperature

By pairing our desalination plant with a natural gas power plant and utilizing its cooling water we forgo initial heating of our inlet stream. Normally, water pumped from the ocean would be approximately 25^oC and by capturing waste heat from the power plant we can reduce utilities with our water inlet of 45^oC. Through an economic comparison between two plants nominally at 21 flash stages, the plant with an inlet temperature at 25^oC has a capital cost of $245MM and a utility cost of $77MM/yr and the same facility with an inlet water temperature at 45^oC has a capital cost of only $226MM and a utility cost of $56MM/yr. As shown in Table 4 below, this $19MM savings in capital costs and $21MM/yr saving in utilities is largely due to reducing the size of the heat exchangers and lowering the steam required for heating duty on the inlet seawater respectively. Therefore our process is optimized by recovering the waste heat water from the gas fired power plant.

Table 4: Seawater Inlet Temperature Process Cost Optimization

Seawater Temperature (°C)	Capital Cost ($MM)	Utility Cost ($MM/yr)
25	245	77
45	226	56
Savings	56	19

Overall Process Economics

After optimization at fixed flow rates, the process was scaled to meet the the 67% operating percentage and final product definitions as shown in Table 1, which resulted in finalized economics shown in Table 5 below. For a table of the installed costs of each process vessel, see Appendix I. For the finalized NPV calculations, see Appendix J.

Table 5: Summarized Economics for 21-stage MSFD Process

Economic Parameter	Cost (USD)
Capital Cost	345 MM
Total Operating Cost per Year	103 MM
Utility Cost per Year	59 MM
30-Year NPV (6% Discount)	-1035 MM

Conclusions and Recommendations

Project Viability

Table 5 above demonstrates that the process has reasonable operating and capital costs, but the 30-year NPV is negative by -1035 $MM due to the extremely low sell price capability of desalinated water as a product. Therefore, this plant would have to heavily be supported by the City of Corpus Christi to run as the product revenue will not make it self-sufficient. The acquisition of government subsidies are key operating requirements for all desalination facilities in the United States.

Additional Concerns

Large volumes of water, steam, and energy are needed to ensure the project runs as specified. While most of the water can easily be extracted from the ocean with little to consequence, the fuel used to generate the steam and electricity can have a significant environmental impact. As the plant is located next to a natural gas power plant, this fuel is better than coal with respect to environmental concerns, and utilizing energy from the power plant to preheat the feed stream for the desalination plant helps lower environmental impacts.

The large amounts of high pressure steam being used represents one of the the largest safety concern in the process, because a high pressure steam release could easily cause severe burns or even death. Proper precautions and control measures must be in place to prevent accidental release. Otherwise, the process does not utilize any known toxic chemicals and is relatively safe from a chemical standpoint. The flash vessels mostly operate below atmospheric pressure, so they must be constructed as vessels that can withstand vacuum without failing.

Appendix A: Proposed Location of Desalination Plant

Pictured below is the proposed location for the desalination plant, located on Nueces Bay in Corpus Christi, adjacent to the Nueces Bay Energy Center. The plant's proposed location is shown in the area outlined by the small circle in the upper right of the figure.

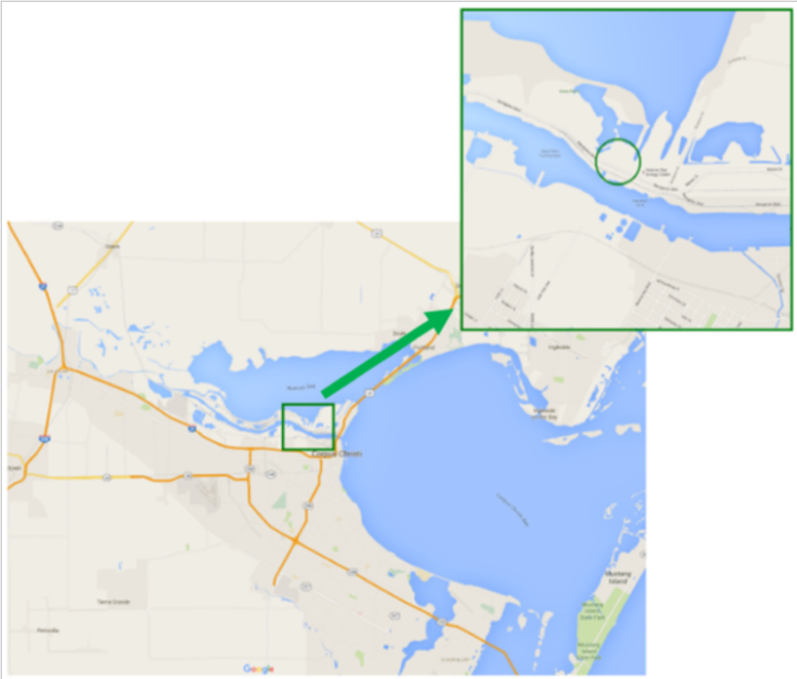

Figure A.1: Proposed mapped location of Nueces Bay Desalination Facility.

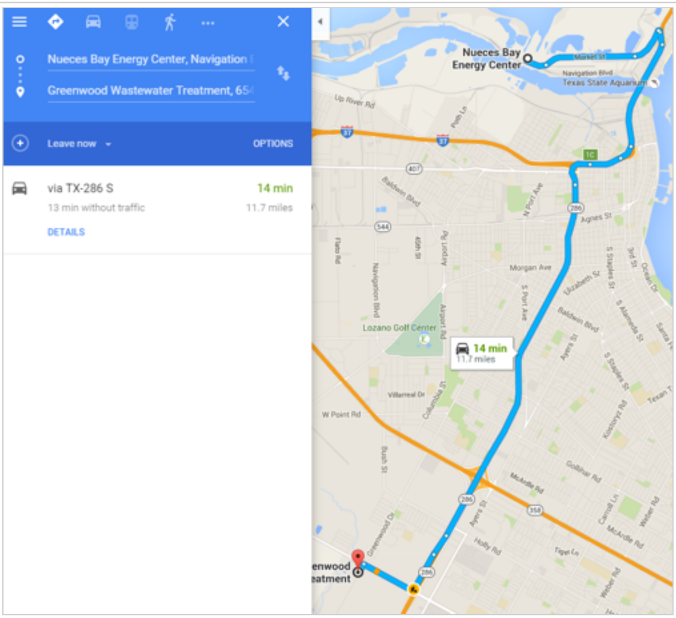

Figure A.2: Mapped distance from proposed location to Class I Injection Well

Appendix B: Feed and Product Definitions - Salinity

Table 6: Average Composition of Total Dissolved Solids in Seawater[36][37]

Compound	Mass Composition (kPa)
Chlorine	55.03%
Sodium	30.64%
Sulfate	7.70%
Magnesium	3.66%
Calcium	1.17%
Potassium	1.12%
Bicarbonate	0.36%
Bromine	0.19%
Other	0.15%

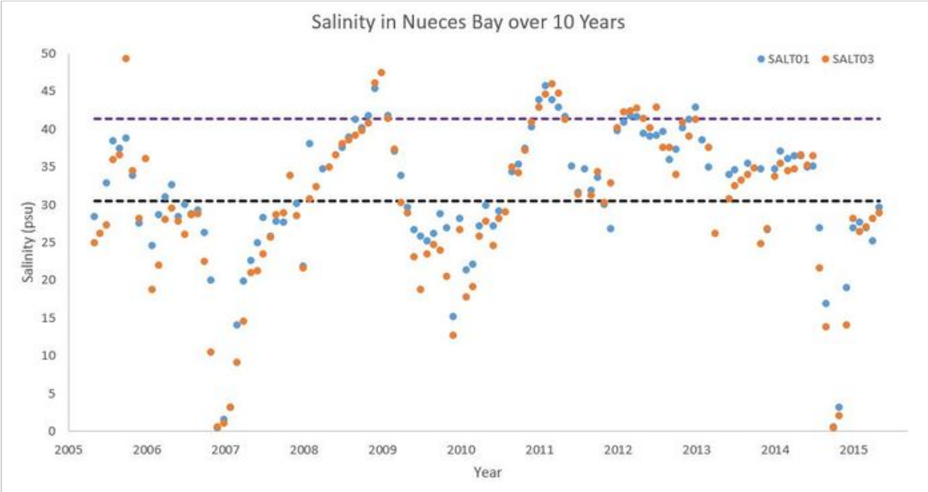

Figure C.1: Salinity in Nueces Bay over the past 10 years[7].

Appendix C: Process Flow Diagram

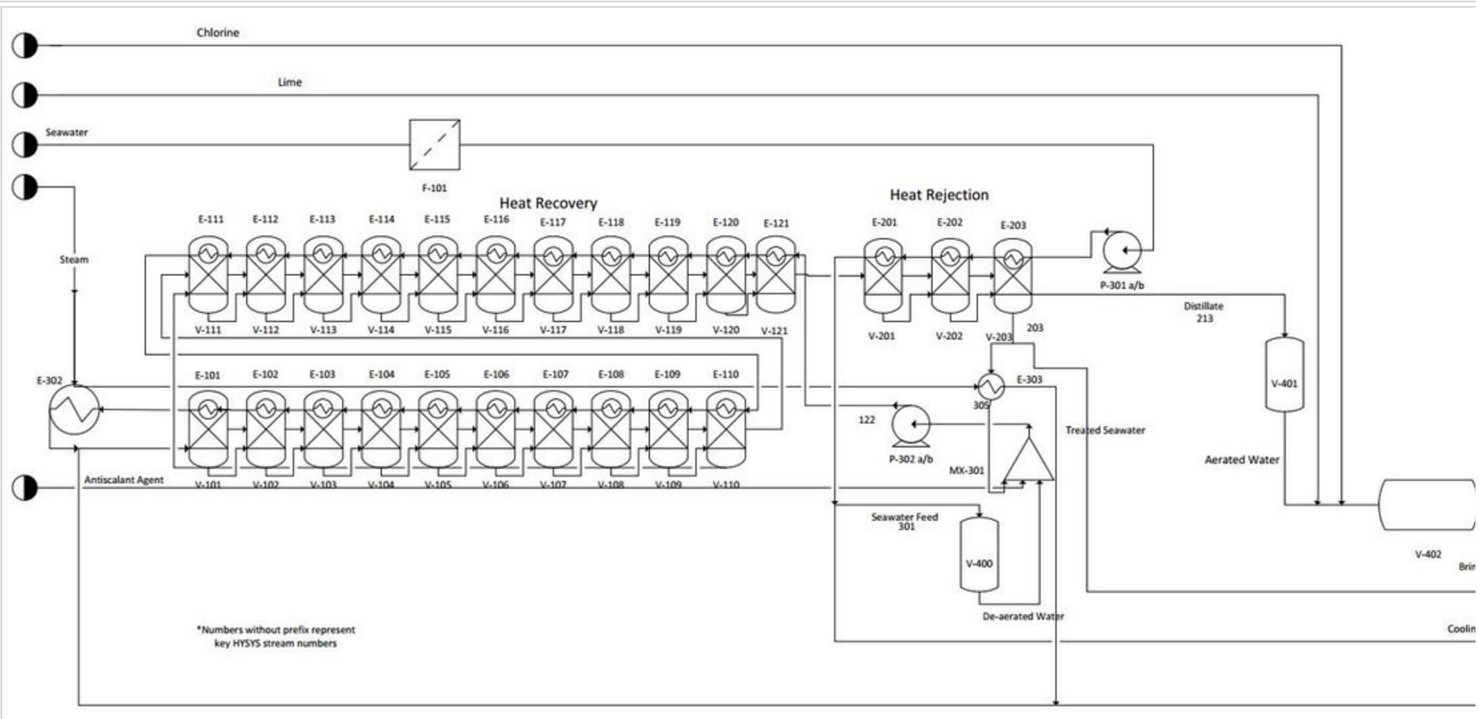

Figure C.1: Process Flow Diagram

Appendix D: HYSYS Simulation Model

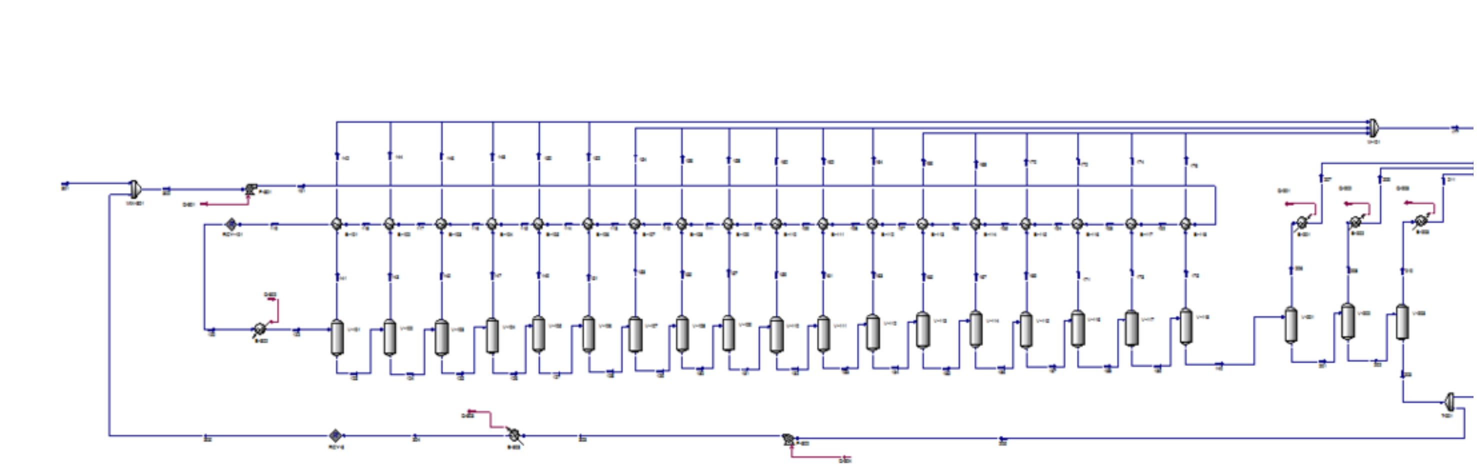

Figure D.1: Screenshot of HYSYS Model Simulation.

Appendix E: Stream Tables

A summary of some values is shown below:

Table 7: Key process material streams

	Stream	Molar Flow (10^5 kgmole/h)	Mass Flow (10^5 kg/h)
Inlet	Feed Water	454	83.6
Outlet	Brine Reject	1.90	36.0
Product	Pure Water	2.64	47.5

Table 8: Key process energy streams

Stream	Heat Flow (kJ/h)
Feed Water	1.30E+11
Q-301	5.73E+06
Q-302	3.09E+09
Q-304	4.68E+06
Q-303	-1.26E+09
Brine Reject	-5.47E+10
Pure Water	-7.46E+10

(Negative value denotes energy leaving process)

Appendix F: Heat Exchanger Sizing

Table 9: Utility Temperatures and Heat Transfer Coefficient[38]

	HP Steam	Cooling Water
Temperature (°C)	250	35
U (W/m^2K)	4000	1500

Table 10: Internal Flash Heat Exchanger Sizing, condensing steam.

Exchanger	UA (kJ/°C-h)	U (kJ/hr/m^2°C)	A (m^2)
E-101	1.47E+07	14400	1019
E-102	1.53E+07	14400	1059
E-103	1.59E+07	14400	1103
E-104	1.66E+07	14400	1153
E-105	1.74E+07	14400	1209
E-106	1.83E+07	14400	1272
E-107	1.93E+07	14400	1342
E-108	2.05E+07	14400	1424
E-109	2.19E+07	14400	1518
E-110	2.35E+07	14400	1629
E-111	2.54E+07	14400	1761
E-112	2.77E+07	14400	1921
E-113	3.05E+07	14400	2119
E-114	3.42E+07	14400	2372
E-115	3.90E+07	14400	2706
E-116	4.56E+07	14400	3167
E-117	5.55E+07	14400	3856
E-118	7.20E+07	14400	5000

Table 11: Heat Exchanger Sizing for Process Heaters and Coolers

Exchanger	Q (kJ/h)	U (J/s/m^{2}K)	U (kJ/hr/m^{2}K)	LMTD	A (m^2)
E-201	9.67E+08	1500	5400	16.03	11165
E-202	1.26E+09	1500	5400	7.57	30921
E-203	1.88E+09	1500	5400	0.45	77366
E-302	3.09E+09	4000	14400	155	1385
E-303	1.26E+09	4000	14400	209.5	418

The high heat transfer area in the final cooler of the heat rejection stage is due to the close approach temperature. Other cooling / pressurization techniques will be used to reduce the area.

Appendix G: Pressure Vessel Sizing

Table 12: Operating temperature and pressure and corresponding thickness for flash vessels

Unit	Temperature (°C)	Pressure (kPa)	Thickness (mm)
V-101	103.6	110.7	17.14
V-102	102.3	105.7	16.62
V-103	100.9	100.7	16.10
V-104	99.5	95.7	15.58
V-105	98.0	90.7	15.06
V-106	96.4	85.7	14.53
V-107	94.8	80.7	14.01
V-108	93.1	75.7	13.49
V-109	91.3	70.7	12.97
V-110	89.3	65.7	12.45
V-111	87.3	60.7	11.93
V-112	85.1	55.7	11.41
V-113	82.7	50.7	10.89
V-114	80.1	45.7	10.36
V-115	77.3	40.7	9.84
V-116	74.1	35.7	9.32
V-117	70.6	30.7	8.80
V-118	66.5	25.7	8.28
V-201	61.7	20.7	7.76
V-202	55.8	15.7	7.24
V-203	47.9	10.7	6.72

Appendix H: Expanded Economic Assumptions

Table 13: Product and raw material market prices

Parameter	Value/Unit	Unit
Sale price of water[32]	$0.0021	liter
Brine disposal[33]	$0.0016	barrel
Belgard EV 2030[39]	$1.99	kilogram
Sodium Hypochlorite 12%[40]	$4.54	kilogram
Lime[41]	$0.09	kilogram
CO_2[42]	$0.04	kilogram

Additional Economic Assumptions

- 6% cost of capital
- Construction of plant in 2 years, 50% of full-scale production in 3rd year
- 35% tax rate with 5 year MACRS depreciation
- 5 operators per shift with 3 shifts
- $50,000 salary per operator
- Supervision is 25% operating labor
- Direct overhead is 45% of labor & supervision
- Maintenance is 3% of ISBL investment per year
- Plant overhead is 65% of labor and maintenance
- Screen price negligible (~$2,000)

Appendix I: Equipment Cost Summary

Table 14: Installed costs of all major process equipment, in USD.

Unit	Equipment Cost (USD)	Installed Cost (USD)	Equipment Weight (LBS)	Total Installed Weight (LBS)
E-101	1033900	1716400	403500	568790
E-102	1065700	1748700	416500	581960
E-103	1131500	1835900	439800	609338
E-104	1206900	1913200	469800	639902
E-105	1238900	1945800	483000	653292
E-106	1315000	2023500	515400	686222
E-107	1058900	1741900	413000	578460
E-108	1447500	2458400	564900	796324
E-109	1526200	2540200	598500	830924
E-110	1656600	2698900	646400	883959
E-111	1833500	2900600	714600	953808
E-112	1974700	3046000	774000	1014446
E-113	2186200	3647000	856000	1154275
E-114	2483100	4018400	968400	1278580
E-115	2817800	4433900	1103700	1431085
E-116	3291100	5778100	1290000	1731000
E-117	4062000	6880400	1588400	2081273
E-118	5361900	10603700	2097500	3054252
E-201	1682100	2649700	633600	835319
E-202	3265400	5677900	1233000	1643496
P-301	8263300	10579300	593100	1016878
P-302	5435000	7740400	389400	811881
V-101	43800	221900	11900	40680
V-102	2070900	5175400	1051900	1717877
V-103	1759900	4866600	804500	1476730
V-104	2976300	6117800	1625400	2292006
V-105	32200	201500	8300	35146
V-106	3480300	6698800	1863900	2546114
V-107	3538500	6800500	1918700	2612432
V-108	3673200	6973400	1985400	2687585
V-109	4178000	7524400	2228400	2938816
V-110	3690700	6978300	1994800	2693811
V-111	3711900	6991200	1986700	2683352
V-112	3541500	6754700	1891600	2562046
V-113	3288600	6463300	1782100	2444134
V-114	3039500	6176100	1651800	2306017
V-115	2961900	5572100	1614900	2161959
V-117	2831600	5403700	1549900	2087670
V-118	2800000	5359400	1531200	2065331
V-201	2807200	5147700	1535700	2002614
V-202	2921700	5295200	1587700	2063167
V-203	6322000	9725700	3364500	4063966

Appendix J: Economic NPV Analysis

Company Name
Address

ECONOMIC ANALYSIS

Form XXXXX-YY-ZZ

Project Name								
Project Number						Sheet	1	
REV	DATE	BY	APVD	REV	DATE	BY	APVD	

Owner's Name	Capital Cost Basis Year	2006		
Plant Location	Units	English	Metric	
Case Description	On Stream	8,000 hr/yr	333.33 day/yr	

REVENUES AND PRODUCTION COSTS

	$MM/yr
Main product revenue	49.5
Byproduct revenue	-43.7
Raw materials cost	0.5
Utilities cost	$72.24
Consumables cost	0.0
VCOP	116.5
Salary and overheads	
Maintenance	
Interest	
Royalties	
FCOP	21.2

CAPITAL COSTS

	$MM
ISBL Capital Cost	$209.47
OSBL Capital Cost	83.8
Engineering Costs	0.0
Contingency	0.0
Total Fixed Capital Cost	293.3
Working Capital	11.5

CONSTRUCTION SCHEDULE

Year	% FC	% WC	% FCOP	% VCOP
1	50.00%	0.00%	0.00%	0.00%
2	50.00%	100.00%	0.00%	0.00%
3	0.00%	0.00%	50.00%	50.00%
4	0.00%	0.00%	100.00%	100.00%
5	0.00%	0.00%	100.00%	100.00%
6	0.00%	0.00%	100.00%	100.00%
7+	0.00%	0.00%	100.00%	100.00%

ECONOMIC ASSUMPTIONS

Cost of equity		Debt ratio	Tax rate	0.35	
Cost of debt			Depreciation method	MACRS	
Cost of capital	0.12		Depreciation period	5	years

CASH FLOW ANALYSIS

All figures in $MM unless indicated

Project year	Cap Ex	Revenue	CCOP	Gr. Profit	Deprcn	Taxbl Inc	Tax Paid	Cash Flow	PV of CF	NPV
1	146.6	0.0	0.0	0.0	0.0	0.0	0.0	-146.6	-130.9	-130.9
2	158.1	0.0	0.0	0.0	0.0	0.0	0.0	-158.1	-126.0	-257.0
3	0.0	24.8	68.8	-44.1	41.9	-86.0	0.0	-44.1	-31.4	-288.3
4	0.0	49.5	137.6	-88.1	67.0	-155.2	-30.1	-58.0	-36.9	-325.2
5	0.0	49.5	137.6	-88.1	40.2	-128.3	-54.3	-33.8	-19.2	-344.4
6	0.0	49.5	137.6	-88.1	24.1	-112.3	-44.9	-43.2	-21.9	-366.3
7	0.0	49.5	137.6	-88.1	24.1	-112.3	-39.3	-48.8	-22.1	-388.4
8	0.0	49.5	137.6	-88.1	12.1	-100.2	-39.3	-48.8	-19.7	-408.1
9	0.0	49.5	137.6	-88.1	0.0	-88.1	-35.1	-53.1	-19.1	-427.2
10	0.0	49.5	137.6	-88.1	0.0	-88.1	-30.8	-57.3	-18.4	-445.7
11	0.0	49.5	137.6	-88.1	0.0	-88.1	-30.8	-57.3	-16.5	-462.1
12	0.0	49.5	137.6	-88.1	0.0	-88.1	-30.8	-57.3	-14.7	-476.8
13	0.0	49.5	137.6	-88.1	0.0	-88.1	-30.8	-57.3	-13.1	-490.0
14	0.0	49.5	137.6	-88.1	0.0	-88.1	-30.8	-57.3	-11.7	-501.7
15	0.0	49.5	137.6	-88.1	0.0	-88.1	-30.8	-57.3	-10.5	-512.2
16	0.0	49.5	137.6	-88.1	0.0	-88.1	-30.8	-57.3	-9.3	-521.5
17	0.0	49.5	137.6	-88.1	0.0	-88.1	-30.8	-57.3	-8.3	-529.8
18	0.0	49.5	137.6	-88.1	0.0	-88.1	-30.8	-57.3	-7.4	-537.3
19	0.0	49.5	137.6	-88.1	0.0	-88.1	-30.8	-57.3	-6.7	-543.9
20	-11.5	49.5	137.6	-88.1	0.0	-88.1	-30.8	-45.8	-4.7	-548.7
21	0.0	49.5	137.6	-88.1	1.0	-89.1	-30.8	-57.3	-5.3	-554.0
22	0.0	49.5	137.6	-88.1	2.0	-90.1	-31.2	-56.9	-4.7	-558.7
23	0.0	49.5	137.6	-88.1	3.0	-91.1	-31.5	-56.6	-4.2	-562.9
24	0.0	49.5	137.6	-88.1	4.0	-92.1	-31.9	-56.2	-3.7	-566.6
25	0.0	49.5	137.6	-88.1	5.0	-93.1	-32.2	-55.9	-3.3	-569.9
26	0.0	49.5	137.6	-88.1	6.0	-94.1	-32.6	-55.5	-2.9	-572.8
27	0.0	49.5	137.6	-88.1	7.0	-95.1	-32.9	-55.2	-2.6	-575.4
28	0.0	49.5	137.6	-88.1	8.0	-96.1	-33.3	-54.8	-2.3	-577.7
29	0.0	49.5	137.6	-88.1	9.0	-97.1	-33.6	-54.5	-2.0	-579.7
30	0.0	49.5	137.6	-88.1	10.0	-98.1	-34.0	-54.1	-1.8	-581.5

ECONOMIC ANALYSIS

Average cash flow	-53.2	$MM/yr	NPV	10 years	-445.7	$MM	IRR	10 years	#NUM!
Simple pay-back period	-5.7279122	yrs		15 years	-512.2	$MM		15 years	#NUM!
Return on investment (10 yrs)	-28.56%			20 years	-548.7	$MM		20 years	#NUM!
Return on investment (15 yrs)	-28.68%		NPV to yr	1	-130.9	$MM			

NOTES

1.
2.
3.

Cap Cost ($MM)	Utility Cost ($MM)	Brine/ Product
$209.47	$72.24	72.47%

10 yr NPV	20 yr NPV	30 yr NPV
-445.7	-548.7	-581.5

Figure J.1: Screenshot of process economic analysis.

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