**The Design of a Desalination Plant in New Orleans, Louisiana**

A Technical Report submitted to the Department of Chemical Engineering

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On my honor as a University Student, I have neither given nor received unauthorized aid on this assignment as defined by the Honor Guidelines for Thesis-Related Assignments

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## 1. Executive Summary

The increasing intrusion of saltwater from the Gulf of Mexico into the Mississippi River has raised significant concerns about the availability of clean water in New Orleans, Louisiana, particularly the Algiers neighborhood. To combat this emerging water crisis, the proposed solution is the establishment of a desalination plant located 60 miles south of the city in Port Fourchon. This initiative aims to provide a sustainable and reliable source of potable water, leveraging the proximity of the Gulf of Mexico.

The desalination process, designed to meet the daily water demand of approximately 54,000 residents, is anticipated to produce 10 million gallons of water daily. Key components of the process include pretreatment stages (straining, microfiltration, ultrafiltration) to remove contaminants, followed by reverse osmosis (RO) to separate salt from seawater, ensuring the production of high-quality freshwater. Post-treatment processes, including remineralization, recarbonation, and chlorination, are implemented to meet the US water quality standards.

In addition to water production, the plant's innovative design addresses the environmental challenge of brine disposal. By treating the byproduct brine to produce fertilizer and rock salt, the project not only minimizes waste but also enhances economic viability. The integration of these by-products into the market demonstrates the plant's role in promoting environmental sustainability and economic growth. The plant's production includes daily outputs of 10 million gallons of tap water, 15 million kilograms of fertilizer, and 13 million kilograms of rock salt.

The plant's financial assessment reveals substantial economic viability. The total annual revenue from tap water, fertilizer, and rock salt production is estimated at approximately $3.5 billion, with the fertilizer being the most lucrative product. The initial capital investment for the plant is significant, at over $2.6 billion, covering land, equipment, installation, and other expenses. The brine post-treatment equipment, primarily the large multi-stage evaporators, represents the bulk of the equipment cost, amounting to approximately $565 million. However, the projected internal rate of return (IRR) of 27% and a net present value (NPV) exceeding $4 billion highlights the project's potential profitability. This financial outlook, combined with the plant's ability to ensure a consistent water supply, proves its importance as a strategic investment for the future.

For future design teams, it is recommended to address the discrepancies in flow rates and operational parameters between the design and real-world plants to enhance efficiency and scalability. Optimizing the overall recovery rate of the RO unit, which is currently at 40%, is essential to improve water production capacity and plant performance. It is also recommended to further investigate alternative ways to remove water from the brine stream, in hopes of eliminating the large multi-stage evaporators that represent the bulk of equipment costs.

In summary, the desalination plant is positioned as an innovative solution to New Orleans' water crisis, with a strong emphasis on economic and environmental sustainability. Through careful planning, innovative engineering, and strategic financial management, the project presents a promising model for addressing water scarcity and supporting regional development.

## 2. Introduction

Several countries across the globe struggle to access uncontaminated, drinkable water for irrigation, sewage, and industrial means. Of those countries, most of them are considered third world; however, the United States has a water crisis of its own. According to Time Magazine as of 2023, “nearly half a million U.S. households lack complete plumbing, while many more were living in communities with unclean water.” (Nelson, 2023). After further research into America’s potential water crisis, New Orleans, Louisiana seemed to stand out because of one river: the Mississippi.

In the vibrant city of New Orleans, Louisiana, the Mississippi River has long served as the lifeblood, providing essential tap water to its residents. In recent years, saltwater from the Gulf of Mexico has crept into this freshwater river. Not only does this intrusion endanger marine life and local ecosystems located close to the river, but also threatens Louisiana's source of clean water, posing a significant challenge to the city's water supply security. An augmented sill made from sand has been formed to mitigate the intrusion from encroaching farther up the river. The US Army Corps of Engineers explains that a sand sill constructed at a proper height of the rivers’ streambed can reduce incoming saltwater flow. (USACE, n.d.). However, this solution is not efficient enough to solve the problem entirely. An article written by Tulane University’s School of Public Health and Tropical Medicine agreed with this sentiment: “That said, the sill is designed to only buy more time and not meant to altogether prevent the saltwater wedge from proceeding upriver. The low flow remains the problem and the wedge will top the sill without more flow.” (SPHTM Communications., n.d.).

Recognizing the pressing need for an alternative water source, it is being suggested to harness the vast expanse of the nearby Gulf of Mexico as a potential solution. The plan is to build a desalination plant 60 miles south of the city in Port Fourchon to take in water from the Gulf and turn it into usable tap water. The proposed idea plans to generate enough water for the neighborhood of Algiers in New Orleans, a neighborhood representing a seventh of the city's total population. Further, the proposed plant will meet the daily water demand of approximately 54,000 residents, with predicted production of 10 million gallons of water daily. This proposed desalination plant presents a forward-thinking strategy to secure a dependable water supply for Algiers while mitigating the risks associated with saltwater intrusion in the Mississippi River.

Desalination, while effective in producing freshwater, often confronts challenges in the disposal of brine, a by-product that can pose environmental concerns if not managed appropriately. Currently, the world produces almost 27 billion gallons of water per day using desalination, which leaves a similar volume of concentrated brine leftover; most of which is pumped back out to sea. (Chandler, 2019). Our paper addresses this issue by introducing an integrated process that not only resolves the brine disposal dilemma but also transforms it into valuable by-products. Specifically, we propose the utilization of the brine to produce fertilizer and rock salt, thereby minimizing waste and creating additional resources with tangible benefits.

Through the process of desalination, the salt removed from the seawater forms a brine solution. In this brine there are various divalent metal cations (Hughes, 1986). These valuable ions are lost in conventional brine disposal methods. When mixed with ammonia and phosphoric acid, the brine becomes a relatively insoluble precipitate made up of a mixture of divalent ammonium phosphates (Hughes, 1986). This mixture makes an ideal fertilizer due to the high phosphorus content, availability of nitrogen, pH adjustment capabilities, water solubility, and flexibility in improving soil fertility and promoting plant growth. Free of all divalent metals, the brine is dried and is of perfect composition and grade to be sold as rock salt. This provides a perfect alternative to typical brine disposal methods.

As we delve into the technical details of the proposed desalination plant and explore the intricacies of the by-product utilization process, this paper aims to shed light on an innovative and comprehensive solution that can serve as a model for sustainable water management in coastal cities facing similar challenges.

# 3. Discussion

## 3.1. Pretreatment

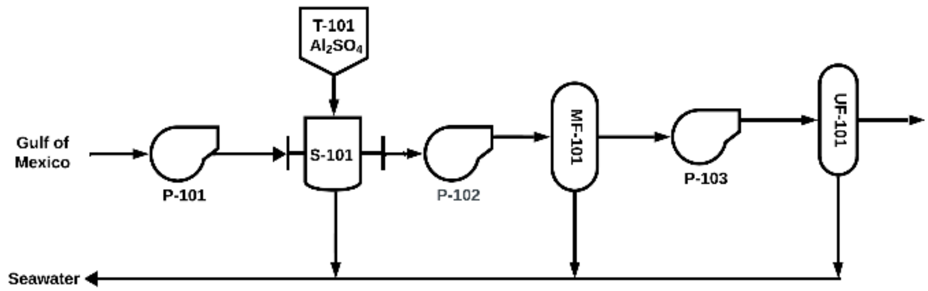



**Figure 3.1.1:** PFD for Pretreatment

### *3.1.1 Strainer*

Strainers play a crucial role in pretreatment within the desalination process, ensuring the removal of contaminants that could potentially damage equipment downstream. Commonly utilized strainers include basket strainers, duplex strainers, self-cleaning strainers, and temporary strainers. Basket strainers, often constructed from stainless steel, offer a large filter area and high contaminant-holding capacity, making them ideal for applications where ultra-fine filtration is required. Duplex strainers facilitate uninterrupted operation by enabling easy swapping of full strainers while the system remains functional, crucial for processes with minimal downtime tolerance. Self-cleaning strainers are particularly advantageous in systems unable to accommodate multiple strainers, as they employ motorized filtration to continually remove debris without necessitating system shutdowns. Temporary strainers serve as a vital first line of defense, clearing coarse debris before the treatment process begins or after prolonged shutdowns. In the intricate landscape of water treatment, selecting the appropriate strainer based on filtration needs, system configuration, and maintenance considerations is essential for optimal efficiency and equipment longevity (Ligon, 2021).

We opted to utilize the FilterSafe Coarse Filtration Leviathan 300 as the primary strainer for our desalination plant due to its capability of effectively removing larger, suspended particles, and safeguarding the integrity of downstream equipment. With a capacity to handle 2,700 cubic meters per hour and a wide filtration degree ranging from 200 to 6,000 microns, the Leviathan 300 offers versatile filtration suitable for various water qualities and contaminants. Its substantial filter screen area of 30,000 square centimeters ensures ample filtration surface within a compact footprint, optimizing space utilization in our plant. In order to handle our flow rate, we will be utilizing two of these strainers. Additionally, the Leviathan 300's design, featuring a single moving part, promises minimal maintenance requirements, ensuring continual operation without downtime during flushing cycles. By selecting the Leviathan 300, we prioritize reliability, efficiency, and cost-effectiveness in our desalination process, mitigating the risk of equipment damage and optimizing overall plant performance.

*3.1.2. Microfiltration*

Microfiltration is the second type of filtration used in this plant. It is responsible for removing microorganisms, such as bacteria and viruses, and smaller suspended particles. The mechanism involves a pressure difference between the membrane to allow liquid to pass through and separate it from unwanted particles. This is achieved by two methods, cross-flow or dead-end microfiltration. Cross-flow involves the liquid passing through along the membrane interface, allowing the liquid to filter out as a permeate, while the solids flow to the end as a concentrate. To effectively clean the membrane and reduce concentration polarization, the concentrate is circulated in a loop back to the membrane. This is normally used on high solid concentrations. Dead-end involves the liquid passing through the membrane interface perpendicularly, leading to an accumulation of solids as a filter cake, while the liquid continues flowing to the end as a permeate. A common issue with this method is that at a certain point in this process, the filter cake increases in size and prevents the flow of the solution through the membrane by reducing permeate flux. So, at regular intervals, the membrane is cleaned by backwashing or chemical or mechanical methods (Envirogen Group, 2021). This will allow a constant flow of solution and an efficient filtration process.

For this plant, we decided to utilize the Alfa Laval MFG1-8338/48 spiral membrane that operates under the cross-flow method. This membrane has been chosen due to its effective removal of small-diameter particles and wide applications in high-sanitary processes. Additionally, it offers the advantage of eliminating the frequent replacement and disposal of cartridges which is typically done in dead-end filtration. That will, in return, reduce maintenance costs as well as improve operating conditions. Its 0.1 μm pore size membrane provides an excellent recovery rate of 90% to retain a high amount of the permeate needed. The dimensions of this membrane include an outer diameter of 211 millimeters and a length of 965 millimeters to help reach the provided recovery rate. This membrane can handle a throughput of 23 cubic meters per hour, which requires more membranes operating simultaneously. To accommodate 105.5 MLD flow rate exiting the previous unit, the strainer, 147 modules will be needed. Additionally, a differential pressure of 1.5 bar is required following the strainer in pump P-102.

*3.1.3. Ultrafiltration*

Following microfiltration, ultrafiltration will be used. The Dow Ultrafiltration (UF) Module, Model SFP 2880, emerges as a standout choice for our proposed plant. These modules are made from high strength, hollow fiber membranes that have features perfect for desalination. Its 0.03 μm nominal pore diameter is used for the removal of bacteria, viruses, and particulates including colloids to protect downstream processes like reverse osmosis. Along with that, the high strength and chemical resistant polymeric hollow fibers allow for long membrane life. It is recommended for use in a wide variety of applications like treatment of seawater. The membranes within the UF ensure a 95% recovery rate, highlighting the precision with which contaminants are filtered out. This fine-tuned filtration capability is essential in safeguarding the public health of Algiers' residents by meeting stringent water quality standards. Moreover, a differential pressure of 4.5 bar is required following microfiltration in pump P-103.

The structural design of the UF Module offers a highly effective membrane area. This is crucial for maintaining consistent performance and ensuring cost-effective operations. Each module, with an 8 inch diameter and 80 inch length, contains an array of 10,000 fibers. This architecture maximizes the surface area available for water filtration, optimizing the module's performance. This module is ideal for systems with capacities greater than 50 cubic meters per hour, like the one proposed. Given the flow rate coming out of MF, 95 million liters per day, and knowing that each module can handle a flow rate about 75 cubic meters per hour, it was calculated that 56 modules will be needed.

### 3.2 Reverse Osmosis System

Reverse osmosis (RO) is the process by which mainly salt and other impurities are removed from the Gulf's sea water under high pressure. A total of 90.2 million liters of treated water per day will be fed to the RO system. A process flow diagram of the system is shown below in **Figure 3.2.1**.

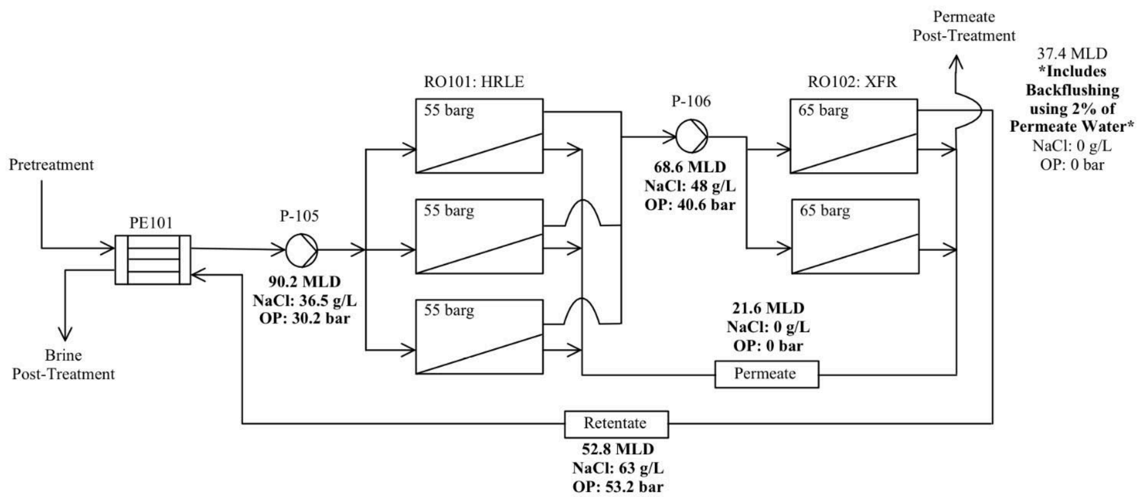



**Figure 3.2.1:** Reverse Osmosis Process Flow Diagram

Two stages will be used with a 24% recovery between each stage, therefore exiting the first stage is a 21.6 MLD permeate stream and a 68.6 MLD retentate stream. This 24% was chosen to achieve higher brine yield for fertilizer and rock salt production. Next, the second stage will be fed using the retentate stream from stage 1. The resulting streams from stage 2 are 16.5 MLD of permeate and 52.1 MLD as a retentate stream. In total, the amount of permeate produced will be 38.1 MLD. However, 2% of the total permeate stream will be utilized to account for backflushing requirements for both RO and pretreatment filtration systems. Therefore, 37.4 MLD will enter permeate post-treatment to be made into potable water.

This system will be organized as a double pass reverse osmosis (DPRO) system. DPRO offers an efficient and flexible solution for desalination, providing high-quality water with reduced energy consumption and operating costs compared to traditional methods. The first stage contains 864 RO membranes within 288 pressure vessels, while the second stage contains 588 membranes within 196 pressure vessels; these calculations can be found in Appendix B. The DPRO system will be designed as a 3:2 array. The first stage will consist of 96 pressure vessels per unit, and the second stage will contain 98 pressure vessels per unit. This type of array was chosen to ensure the people of Algiers community received 10 million gallons of water per day, while maintaining a surplus of retentate for post-brine treatment and our side products.

To ensure continuous flux of water across each membrane, the membranes are encased in pressure vessels that operate at pressures higher than the osmotic pressure. The first stage vessels operate at 55 barg and 65 barg for the second stage. The osmotic pressure (OP) of the gulf water exiting each stage was calculated using **Equation 3.2**, where *c* is the molar concentration [mol/L] of salt in the water, *i* is the van't Hoff's index, *R* is the ideal gas constant [L•bar/mol/K], *T* is the temperature of water [ºC], and *Π* is the OP [bar]. Additionally, a van't Hoff's index value of 2 was chosen based on the fact that sodium chloride dissociates into two ions in water; $Na^+$ and $Cl^-$ (Helmenstine, 2020).

$$\Pi = \mathbf{icRT} \qquad \textbf{(Equation 3.2)}$$

*Inlet Gulf Water:*

$$c = \frac{36.5\ g/L}{58.44\ g/mol\ NaCl} = 0.62\ mol/L$$

$$\Pi = 2 * 0.62 * 0.08314 * 298 = 30.7 \text{ bar}$$

*1st Stage Retentate:*

$$c = 36.5\ g/L * \left(\frac{1}{1-0.24}\right) * \frac{1}{58.44\ g/mol\ NaCl} = 0.82\ mol/L$$

$$\Pi = 2 * 0.82 * 0.08314 * 298 = 40.6 \text{ bar}$$

*2nd Stage Retentate:*

$$c = 48\ g/L * \left(\frac{1}{1-0.24}\right) * \frac{1}{58.44\ g/mol\ NaCl} = 1.08\ mol/L$$

$$\Pi = 2 * 1.08 * 0.08314 * 298 = 53.5 \text{ bar}$$

In order for flux to occur, the differential pressure between the operating pressure of the RO pressure vessels and the OP must remain positive. The stream entering RO has a salt concentration of 35.6 g/L and an osmotic pressure of 30.7 bar. The OP leaving the first stage is calculated to be 40.6 bar, therefore the first stage pressure vessels will operate at 55 barg to ensure sufficient flux is maintained across each membrane. Since the OP leaving the second stage increases to 53.5 bar, the pressure vessels in this stage will also increase to 65 barg. Additionally, one pump will be used to supply salt water prior to each RO stage. Both pumps equally use 4.2 kW of hydraulic power to maintain the water requirement for each stage; this calculation can be found in Appendix A.

The specific membrane elements chosen for the first stage of RO are supplied by the company of Dupont and have the ID name: FilmTec™ SW30HRLE-370/34i Wet (Dupont, 2024). This element offers high performance capabilities and high salt rejection while maintaining low water costs compared to other traditional membrane elements. The second stage membrane elements are also supplied by Dupont and have the ID name: FilmTec™ SW30XFR-400/34i Wet (Dupont, 2024). During the second stage, salt concentrations are higher and this element offers advantages by providing extra fouling resistance, thus reducing both the amount of maintenance required and replacement costs. Each chosen membrane element is spiral bound and composed of polyamide thin-film composite material. The specifications for each membrane are shown below in **Table 3.2.1** and **Table 3.2.2**. Additionally, a diagram of a spiral wound membrane is also depicted in **Figure 3.2.2**. This figure includes the membrane dimensions which are the same for both types of elements. Overall, the entire RO process will require 1452 RO membrane elements and 484 pressure vessels.

**Table 3.2.1 :** Specifications for the FilmTec™ SW30HRLE-370/34i Wet

| **Membrane Specifications for Stage 1** | |
|---|---|
| Inner Diameter | 1.125 in |
| Outer Diameter | 7.9 in |
| Length | 40 in |
| Single Element Recovery Rate | 8% |
| Exiting Flow Rate | 25,000 L/day |
| Active Area | 370 $ft^2$ |

**Table 3.2.2:** Specifications for the FilmTec™ SW30XFR-400/34i Wet

| **Membrane Specifications for Stage 2** | |
|---|---|
| Inner Diameter | 1.125 in |
| Outer Diameter | 7.9 in |
| Length | 40 in |
| Single Element Recovery Rate | 8% |
| Exiting Flow Rate | 28,000 L/day |
| Active Area | 400 $ft^2$ |

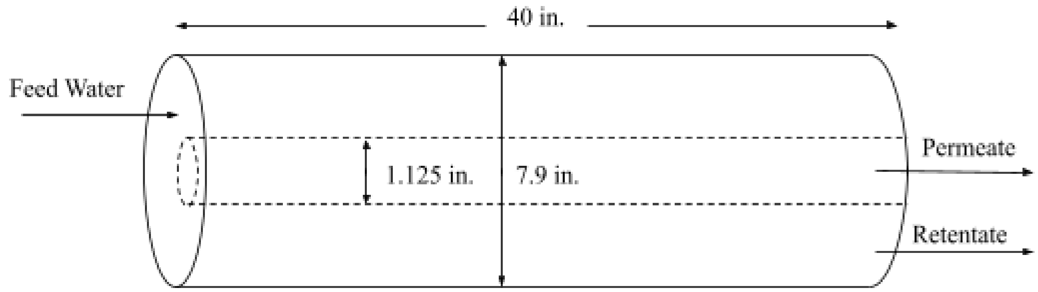



**Figure 3.2.2:** Diagram of RO Membrane Element Dimensions

It is important to note that 3 RO membrane elements will be placed into each pressure vessel. This was to ensure there was a 24% recovery of permeate water in between stages, based on a single element recovery rate of 8%. However, pressure vessels may contain anywhere between one and seven membrane elements per vessel (Lenntech, 2001). A diagram showing the interconnections of 3 membrane elements in one pressure vessel can be found in **Figure 3.2.3**. The pressure vessels are purchased from LennTech with a maximum operating pressure of 69 bar. Therefore, the same pressure vessel can be used for both stages based on the operating pressure required in stage two is only 65 barg. Both stages will utilize the BEL8-E-1000 psi. RO Pressure Vessel, with further specifications found in **Table 3.2.3**.

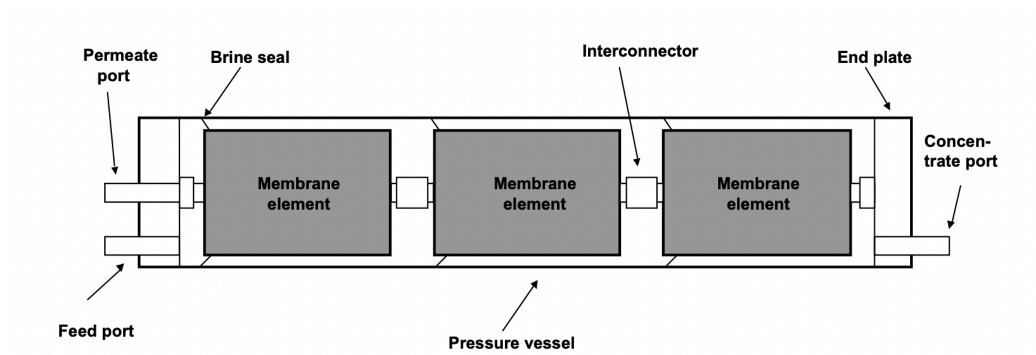



**Figure 3.2.3:** Pressure Vessel Depiction with RO Membrane Elements Inside

(ResearchGate, n.d.)

**Table 3.2.3:** Pressure Vessel Specifications

| | |
|---|---|
| Total Length | 120.6 in |
| Max. Operating Pressure | 69 bar |
| Volume per Pressure Vessel | 0.10 $m^3$ |
| Total Volume of Stage 1 | 28.8 $m^3$ |
| Total Volume of Stage 2 | 19.6 $m^3$ |
| Mass Transfer Area | 50.3 $in^2$ |

Additionally, pressure exchangers are utilized to transfer energy from the brine post-treatment stream to the inlet RO Stream; shown as PE101 and pump 105 (P-105) respectively in **Figure 3.2.1**. These pressure vessels are designed to capture energy that would otherwise be wasted in high-pressure pump flow requirements. More specifically, all of the pressure exchangers will be sourced from Energy Recovery® using the PX Q400 series (*High-Pressure PX. Energy Recovery*, 2024). Reject concentrate enters the PX alongside seawater, where hydraulic pressure transfers energy to the seawater inside a ceramic rotor. The pressurized seawater re-enters the system while any remaining concentrate exits at low pressure. The rotor's ducts function acts similar to a cartridge, enabling continuous pressure transfer, allowing it to rotate at impressive speeds up to 1,200 RPM Overall, 60% energy will be saved after implementing 12 PX Q400 series pressure exchangers prior to the RO system (*High-Pressure PX. Energy Recovery*, 2024).

## 3.3. Permeate Post-Treatment

### *3.3.1. Remineralization*

Remineralization is a post-treatment process that involves adding minerals back into the water that may have been removed or reduced during the water treatment process. When deciding which technique, it was decided to use a process called lime dosing to remineralize the water (Nelson & Luca, 2021). This process includes adding a saturated lime solution ($Ca(OH)_2$) along with carbon dioxide ($CO_2$) to form bicarbonate alkalinity in the following reaction:

$$Ca(OH)_{2\,(aq)} + 2CO_{2\,(g)} \rightarrow Ca(HCO_3)_{2\,(aq)}$$

We will be using $Ca(OH)_2$ as our saturated lime source and will be referring to it as lime for the following time.

In efforts to keep the water quality similar to that in other neighborhoods throughout the city, calculations were done to mimic typical water hardness. Using national water tables, it was found that Louisiana has relatively hard water overall. New Orleans has a water hardness of 138 mg/L, and that was what we used as our target value hardness (HydroFlow, n.d.). In order to reach this targeted hardness, the required dosage needed for lime is 113.5 mg/L at 90% purity and 121.37 mg/L for carbon dioxide (Nelson & Luca, 2021).

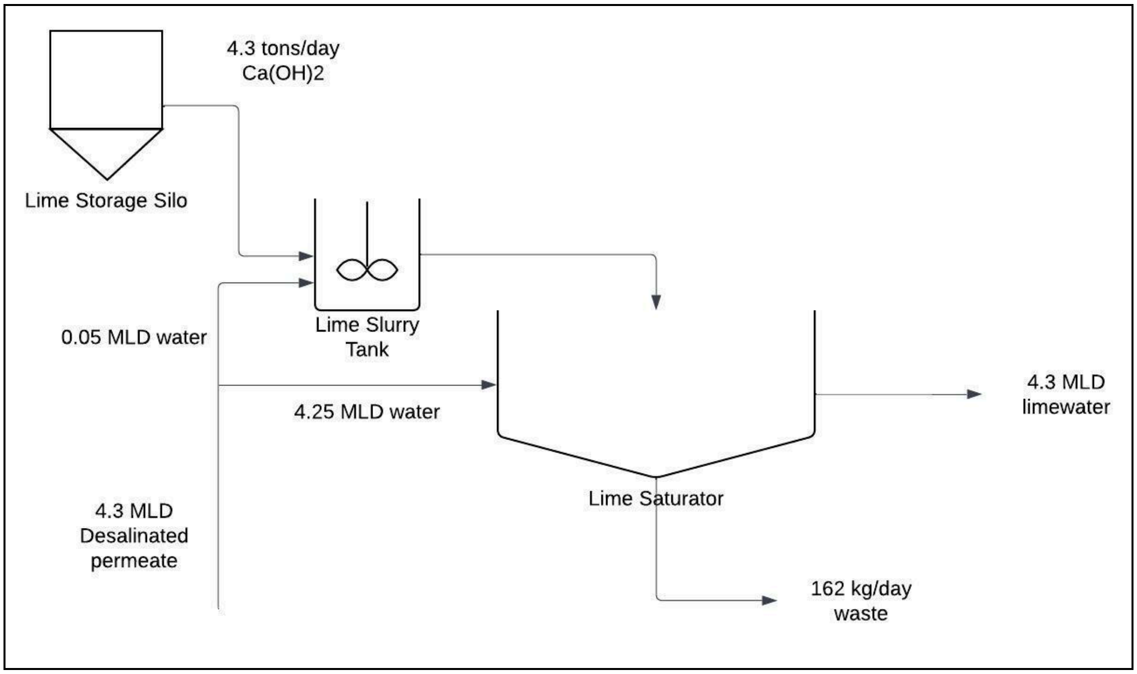



**Figure 3.3.1:** PFD for Lime Addition

The lime begins in the lime storage silo as seen in **Figure 3.3.1**. This silo was designed to hold about 60 days worth of lime. In order to hold that much lime, the volume of the silo needs to be 282 $m^3$ with a height of 10 meters and a diameter of 6 meters. From the storage silo, the solid lime will then be transported into the lime slurry tank at a rate of 4.3 tons a day via a screw conveyor. The target concentration of the lime slurry tank is about 75 g/L. This concentration was chosen because when the concentration is larger than 100 g/L there are dangers of deposits and blockages, and when it is lower than 50 g/L there is a danger of carbonation, making 75 g/L the sweet spot (BeClood.com, n.d.). The lime slurry tank was designed to have a volume of 0.76 $m^3$, a height of 1.25 meters, and a diameter of 1.27 meters. According to the Office of Surface Mining Reclamation and Enforcement, "Studies indicate that a minimum of 6 to 12 minutes of retention time is needed to achieve adequate dissolution of lime products" (OSMRE, n.d.). We designed our tank to have a retention time of 12 minutes to ensure full and proper mixing. It will have a pitch blade impeller with a diameter of 0.3 meters and an agitation rate of 70 RPM (Gaolin, 2022).

After mixing to achieve the target concentration, the lime slurry will be pumped into a lime saturator. By pumping 4.25 million liters per day of water in along with the lime slurry, the concentration of the lime saturator is managed to be kept at 1000 mg/L lime. This step is crucial to efficiently dissolve the lime into water creating a saturated lime solution, ensuring that the lime is uniformly dispersed in the water. In the design, we opted for 2 saturators, both with a volume of 37.5 $m^3$, a height of 5 meters, and a diameter of 3 meters. The retention time of the lime saturator will be 90 minutes (Zinck, n.d.). Because we are utilizing lime with a 90% purity, there are impurities that we are introducing into the system that we need to get rid of. The lime saturator has a settling zone where impurities settle naturally, leaving behind about 162 kg/day of waste. This small amount of waste is also essentially harmless to the environment and will be pumped out back into the Gulf of Mexico along with our waste streams from the pretreatment process.

The final part of this process is injecting this new lime water back into the main flow of water to hit the target hardness of 138 mg/L. The limewater will be pumped out of the saturator at a rate of 4.3 million liters a day.

*3.3.2. Recarbonation*

As part of the lime dosing process, $CO_2$ must also be added to the water. By adding carbon dioxide to the water, carbonic acid will be formed. As stated before, this combined with the remineralization process will help achieve our target water hardness. It was found that our process will require 4.6 tons of carbon dioxide per day (Nelson & Luca, 2021). To do this, we will be using the TOMCO$_2$ Systems' Pressurized Solution Feed (PSF). Conventional $CO_2$ gas control systems lack efficiency in introducing $CO_2$ into water, leading to the escape of carbon dioxide gas into the atmosphere. Their state-of-the-art technology has been designed to force $CO_2$ gas molecules into a carbonic acid solution, effectively injecting carbonic acid directly into the water flow. A pressurized stream of $CO_2$ will be added to a pressurized stream of water at 5 bar. The high pressure will help dissolve the gas resulting in liquid carbonic acid that can be added directly to the water. The process is designed to maximize the conversion of carbon dioxide gas resulting in a 95% efficiency.

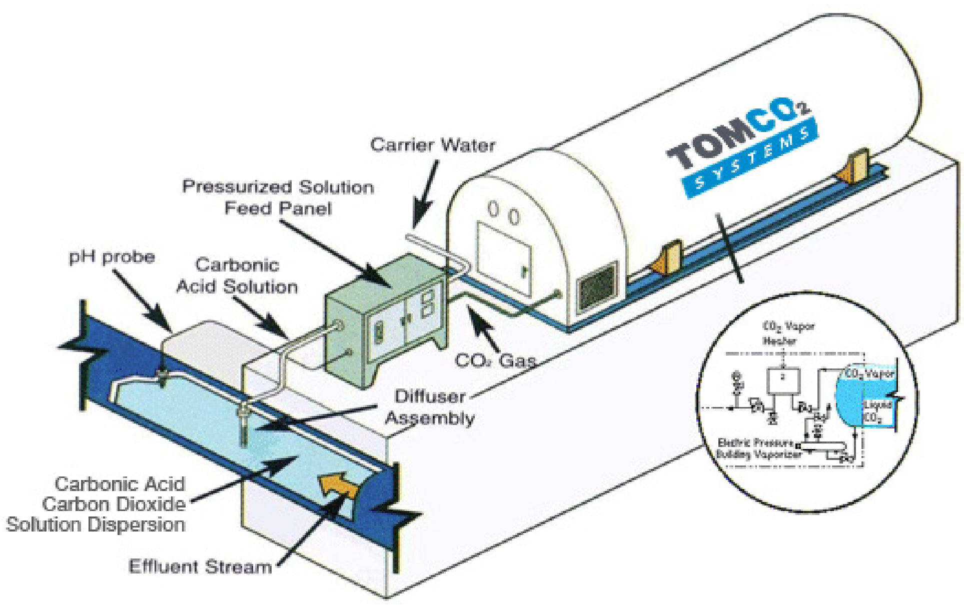



**Figure 3.3.2**. TOMCO$_2$ Systems' Pressurized Solution Feed System (TOMCO$_2$, 2018)

*3.3.3. Chlorination*

To ensure that the water is safe and potable by the time it reaches the faucet, all parasites, algae and bacteria must be removed. Following both remineralization and recarbonation, chlorination takes place. Chlorination is a widely employed method for water post treatment due to its effectiveness in destroying pathogens. By introducing chlorine ($Cl_2$) into the water, it reacts with and neutralizes microorganisms, preventing the spread of waterborne diseases. This process also aids in keeping the pipes downstream from affecting the water flowing through it.

According to water tables, New Orleans maintains a chlorine concentration of 4 parts per million (ppm) in its water. This level of chlorine serves as a critical measure to ensure effective disinfection. Chlorine concentrations are carefully regulated to balance eliminating harmful microorganisms and avoiding any adverse effects on human health. To maintain the suggested chlorine concentration, 151.6 kg of $Cl_2$ per day is needed.

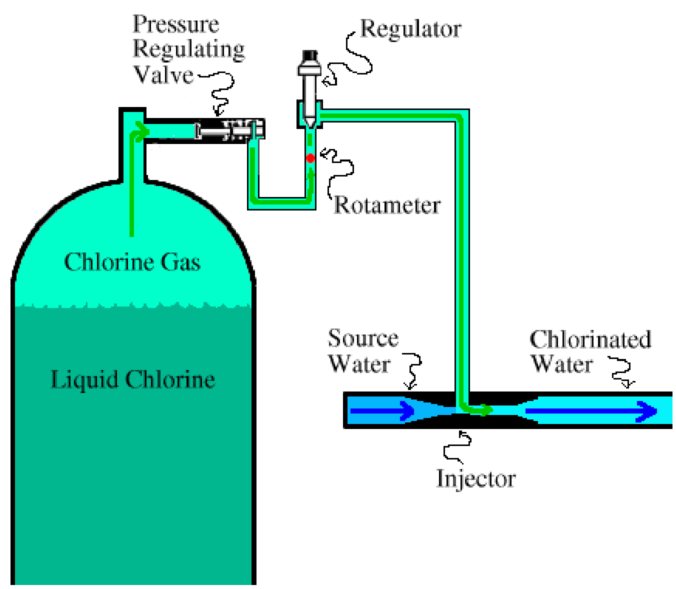



**Figure 3.3.3**. Vacuum Chlorination System (Chlorinator, 2024)

In the design, we plan to utilize a vacuum chlorinator, the most common type of chlorinator. Chlorine gas is pulled out of the cylinder via vacuum into the water. The vacuum is formed by water flowing through the injector which creates a negative head. The negative head forces open the pressure regulating valve on the cylinder, allowing chlorine gas to flow out. The chlorine flow rate is measured using a rotameter. An example of the proposed design can be seen in Figure 3.3.3. This equipment allows for a controlled and efficient introduction of chlorine into the water supply. The vacuum chlorinator ensures accurate dosing, preventing both under-chlorination, which may lead to inadequate disinfection, and over-chlorination, which could pose health risks.

### 3.4. Brine Post-Treatment

*3.4.1. Evaporation Units*

Once the solution is separated by the RO unit, a retentate flows out as a concentrated brine solution. The majority of the composition of this solution is still water, which is undesirable to produce fertilizers and rock salt. To increase the salt concentration, 3 evaporation units will vaporize the excess water in the brine, fertilizer, and rock salt streams, as shown in **Figure 4.4**. The purpose of these units is to vaporize the excess water in the brine stream by utilizing an external heat source, such as steam. The first evaporation unit (E-101) will cause the brine solution to go from 6.3 wt% salt to 15 wt% salt, the second unit (E-102) will dry out the wet fertilizer from 51.7 wt% salt to 99.5 wt% salt, and the third unit (E-103) will evaporate the rock salt solution from 15.9 wt% salt to 99.5 wt% salt. There are different types of evaporation units like plate, bare tube, falling film, and flash evaporators (Vobis LLC, n.d.). These are used in many industries such as food and beverage, pharmaceuticals, and water treatment. Desalination plants typically use the flash evaporator as it can accommodate large amounts of flow as well as reduce energy costs.

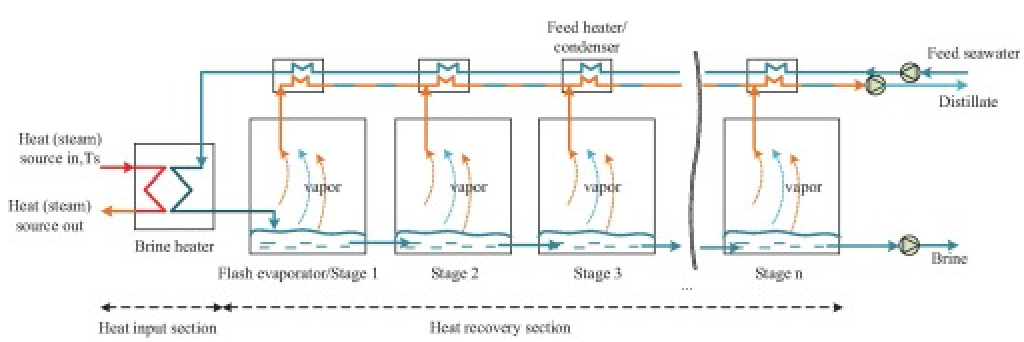



**Figure 3.4.1:** Diagram of Multi-Stage Flash Distillation Process (Rosen & Farsi, 2017)

In this plant, a Multi-Stage Flash Distillation unit will be utilized to effectively vaporize water by changing the pressure and temperature of each stage vessel, as shown in **Figure 3.4.1**. This type of evaporation unit is widely used because it only requires an external heat source, steam, in the first stage. The remaining stages will take advantage of consecutive decreases in pressure and temperature in each stage to lower the boiling point of the solution and generate a flash vaporization of water that will be extracted as water vapor and reused in other evaporation units in the process. In **Figure 3.4.1**, the diagram illustrates that the vapor created during evaporation leaves each vessel and is condensed by passing the vapor over the inlet brine stream feed, which is cooler than the vapor. This results in the production of liquid water distillate. However, for the purpose of reusing the water vapor in the other two evaporation units, an alternate approach is employed. Instead of allowing the vapor to flow over the brine stream feed and condense into liquid water, it is directed as vapor straight to the other two evaporation units to serve as steam replacements. Any excess vapor is then disposed of into the atmosphere. In this design, the brine stream feed is initially heated up by steam only in the first stage. This method offers the benefit of reducing steam costs in the last two evaporation units. By using water vapor directly as steam replacements, the process ensures efficient vaporization and minimizes energy consumption. The retentate flow from the RO unit is 52.3 MLD with a 6.3 wt% salt, and to accommodate it, 21 stages will be utilized to vaporize the water until it reaches 15 wt% salt (Daly et al., 2016). Each vessel will be operating at certain decreasing pressures and temperatures shown in **Table 3.4.1**.

**Table 3.4.1:** Stage Vessels' Operating Conditions and Wall Thickness (Daly et al., 2016)

| Stage | Temperature (°C) | Pressure (bar) | Thickness (mm) |
|---|---|---|---|
| 1 | 103.6 | 1.11 | 17.1 |
| 2 | 102.3 | 1.06 | 16.6 |
| 3 | 100.9 | 1.01 | 16.1 |
| 4 | 99.5 | 0.96 | 15.6 |
| 5 | 98.0 | 0.91 | 15.1 |
| 6 | 96.4 | 0.86 | 14.5 |
| 7 | 94.8 | 0.81 | 14.0 |
| 8 | 93.1 | 0.76 | 13.5 |
| 9 | 91.3 | 0.71 | 13.0 |
| 10 | 89.3 | 0.66 | 12.5 |
| 11 | 87.3 | 0.61 | 11.9 |
| 12 | 85.1 | 0.56 | 11.4 |
| 13 | 82.7 | 0.51 | 10.9 |
| 14 | 80.1 | 0.46 | 10.4 |
| 15 | 77.3 | 0.41 | 9.8 |
| 16 | 74.1 | 0.36 | 9.3 |
| 17 | 70.6 | 0.31 | 8.8 |
| 18 | 66.5 | 0.26 | 8.3 |
| 19 | 61.7 | 0.21 | 7.8 |
| 20 | 55.8 | 0.16 | 7.2 |
| 21 | 47.9 | 0.11 | 6.7 |

By using the energy balance in Appendix D, the heat duty required to heat up the inlet brine solution in stage 1 from 25℃ to 100℃ using 28-bar steam is around 900 MW. The high energy requirement is due to the high flow rate of brine solution needed to vaporize daily. The vessels will be constructed using carbon steel. However, high corrosion areas, such as the tube side of each heat exchanger found between each stage will use a 70/30 (Cu/Ni) alloy to prevent fouling due to high salt concentrations. The dimensions of each vessel are 3 meters in length, 18 meters in width, and 4 meters in height. Although these dimensions are the same for each vessel, the thickness of the vessel walls will differ. A constant change in operating conditions between each vessel and a high chance of corrosion causes it to change accordingly. These values are shown in **Table 3.4.1**.

Towards the end of the fertilizer and rock salt production process, 2 more evaporation units are utilized to completely vaporize the water from the streams exiting the rotary drum filter.

The evaporation unit responsible for vaporizing the wet fertilizer stream will consist of a 9-stage flash distillation to accommodate a flow rate of 28.9 MLD. This will dry the product until 99.5 wt% salt is reached in the final product. Similar to the previous unit's setup, it will have the same vessel dimensions and material of construction, as well as the material replacement for high-corrosion areas. The operating conditions and thickness for each vessel will be the first 9 rows shown in **Table 3.4.1**. By using the same energy balance, the total required heat duty to heat up the inlet brine solution from 25℃ to 100℃ in stage 1 using reused water vapor is 390 MW, hence the low number of stages. However, this method may not be practical because the fertilizer is still wet solid prior to entering the dryer. It is important to note that the pressure difference between stages is too low of a driving force to transport all of the feed from one stage to another. Ideally, a higher viscosity liquid and larger pressure gradient between stages would allow for easier transport of the fertilizer between stages. However, for this design a mechanical method is required to move the fertilizer from one stage to another. One way to accomplish this is to utilize screw feeders. A screw feeder is a mechanical mechanism that uses a screw or a helix to move solid bulk material from one place to another (Schenck Process, n.d.). Implementing screw feeders in the design of this dryer would improve the transportation process of the wet solid fertilizer from one stage to another.

For the rock salt stream, the evaporation unit will consist of only 1 stage which translates to a simple, typical evaporation process. This will accommodate a flow rate of 7.2 MLD and will vaporize water until 99.5 wt% salt is reached in the final product. The dimensions of this vessel will be the same as the previous units, with a 70/30 (Cu/Ni) alloy material of construction. The operating conditions and thickness will be the first row in **Table 3.4.1**. By using the same energy balance, the total required heat duty to heat up the inlet brine solution from 25℃ to 100℃ in stage 1 using reused water vapor from the first evaporation unit is 72 MW, hence the low number of stages.

However, our process of reusing the water vapor from the first evaporation unit to supply energy for the other two evaporation units is not considered practical. This approach aimed to reduce the cost of purchasing steam and minimizing water vapor waste and energy consumption. In hindsight, a different design may have saved significant capital and utility costs. In reality, this type of technology may be more suited for streams containing lower salt concentrations. An alternative approach could have utilized the water vapor in the first evaporation unit to recycle excess condensate to RO and post-treatment water. Additionally, purchasing steam or employing a furnace for the two downstream evaporation units would reduce complexity of the process while also increasing overall efficiency of vaporization. Overall, this process may require a more detailed analysis prior to large-scale production.

*3.4.2. Storage Tanks*

Storage tanks are crucial elements to our system as they facilitate the treatment, processing, and storage of each chemical. Firstly during the pretreatment stage, a storage tank will be utilized for the aluminum sulfate coagulant. This tank will be a high-density polyethylene vertical storage tank which is also referred to as a heavy duty vertical poly(HDPE) tank. These types of tanks are commonly used in the agricultural, industrial, manufacturing, and commercial sectors for a variety of applications. We will be purchasing this tank from the National Tank Outlet (*Aluminum Sulfate Storage Tanks: Al2(SO4)3, Al2O12S3, Alum*, n.d.). We will be purchasing a tank capable of storing 14 days worth of solution. This is about 5250 gallons, and in turn, we chose a 5300 gallon tank with a diameter of 2.6 m and a height of 4.2 m which will be held at ambient pressure.

During the post-treatment process, there will be three areas where storage tanks are utilized. These are during the remineralization, recarbonation, and chlorination processes. Remineralization involves the use of a slaked lime storage silo which will contain 60 days worth of lime powder which is about 28,000 gallons. We will purchase these storage tanks from the National Storage Tank, Inc. and they will have a 2.6 m diameter and a height of 2 m (*2,650 Gallon Poly HDPE Water Storage Tank 102"D X 80"H*, n.d.). Each tank has a volume of 2,650 gallons, so we will use 11 of them which will be held at atmospheric pressure. These tanks are made out of high-density polyethylene and held at ambient pressure.

In the recarbonation stage, we will incorporate a carbon dioxide storage tank. The system deploys 4.6 tons or about 1004 gallons of $CO_2$ per day. In order to store 14 days worth of liquid $CO_2$ which is about 14,100 gallons, there will be 10 aluminum tanks that can each hold 1,500 gallons. These tanks will be purchased from the National Tank Outlet and will have a height of 3.2 meters and a diameter of 1.6 meters and they will be held at a pressure of 24 bar (*1500 Gallon Aluminum Sulfate Storage Tank*, n.d.).

The liquid chlorine storage tank embedded into the chlorination process will be built to hold about 2 days worth which is about 55 gallons of pure liquid chlorine. We will use 1 tank that is made out of Hastelloy C. Hastelloy C is used due to its excellent corrosion resistance properties, especially in highly corrosive environments like this which involves chlorine. This tank will have a diameter of 0.3 m and a height of 1 m. This tank will be held at a pressure of 6 bar.

Furthermore, storage tanks will be used during the fertilizer production. We will store enough phosphoric acid for 4 days or about 4.3 million gallons. In turn, 2,180,000 gallon storage tanks will be used. In order to accommodate for the amount stored at a time, 4 of them will be used. These will be made out of high-density polyethylene and will have a height of 6 meters and a radius of 21 meters. These will be held at ambient pressure.

We will also store enough ammonia for 2 days or about 10.5 million gallons. We will use 4 storage tanks that are made out of high-density polyethylene that can hold 2,620,000 gallons. These tanks will have a diameter of 10 meters and a height of 33.3 meters. They will also be held at a pressure of 10 bar. However, it is important to note that HDPE tanks with these dimensions are difficult to purchase. These specific tanks are not currently on the market and likely hard to quote in real life. Therefore, prior to large-scale implementation the team may pivot to glass-lined metal tanks of similar size with a re-evaluation of initial equipment costs or seek out a third-party company willing to custom make similar HDPE tanks with these dimensions.

Storage tanks are incorporated into our plant before and after the phosphoric acid and ammonia are added to the brine. This is in order to allow for water analysis and calculate the pH. The storage tanks utilized before the chemicals are added will be purchased from the National Storage tank Inc. and have a diameter of 7.6 m and a height of 2.2 m (*25000 Gallons Galvanized Water Storage Tank*, n.d.). 22.1 MLD of brine will be entering these steel storage tanks and, in turn, there will be 4 of them before the mixing stage and during analysis. They will be held at ambient pressure. Furthermore, about 36.1 MLD of the fertilizer slurry product will be exiting the mixing stage. The storage tanks during this stage will be made out of high-density polyethylene and have a height of 2 meters and the radius would be 4 meters. These tanks will be held at ambient pressure.

*3.4.3. Static Mixer*

During the post-brine treatment process addition of phosphoric acid to the brine solution is required. To ensure this addition results in a homogeneous solution, a static inline mixer is used. A static mixer typically consists of stationary elements arranged in a specific configuration to promote mixing by inducing turbulence and interfacial contact between the components being mixed. For this design 4.1 MLD of 75% phosphoric acid will be added to the exiting stream of the first multi-stage evaporator which consists of the brine concentrate solution and water. The total volume of the mixed solution, 26.2 MLD, will be fed to a heat exchanger and a series of agitation tanks to be further sparged with 29% aqueous ammonia.

*3.4.4 Shell and Tube Heat Exchanger*

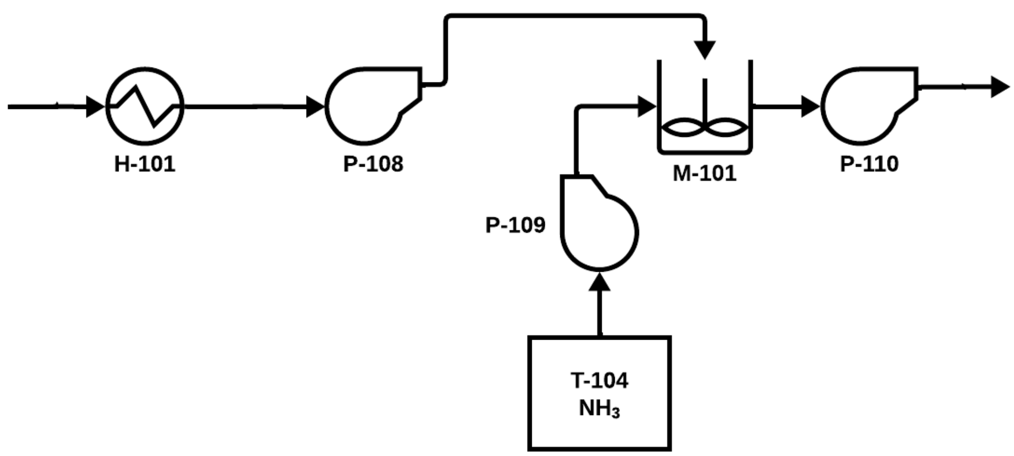



**Figure 3.4.4:** Diagram of Heat Exchanger PFD

Before the mixed solution enters the agitation tanks, as shown in **Figure 3.4.4**, it undergoes a crucial cooling process to prevent any potential runaway reactions. This cooling is achieved through the use of heat exchangers, with this design employing 6,293 $ft^2$ shell and tube heat exchangers constructed from stainless steel. Among these, 3 will operate simultaneously, while another 3 will serve as backups for emergencies or maintenance. Each heat exchanger is capable of handling up to 8.72 MLD of flow and possesses a heat duty of 4.69 MW. The total flow rate entering the heat exchangers is 26.1 MLD. This specific type of heat exchanger was selected due to its ability to accommodate high pressures, temperatures, and flow rates, along with its ease of maintenance and relatively affordable cost. It aligns perfectly with the conditions of our plant, ensuring efficient and reliable operation.

*3.4.5. Ammonia Phosphate Agitation Tank & Sparger*

Following the static mixer, a total of 8 agitation tanks in parallel will be used to facilitate the precipitation of the ammonia phosphate based fertilizer including attached divalent metals such as calcium and magnesium (Hughes, 1986). The use of 8 tanks is required in order to meet the needs of production; which in the case of the fertilizer product is 15 Mkg/day. Although one tank may be more efficient, it is difficult to quote a tank of that size. Each agitation tank will utilize a sparger that will feed vapor 29% aqueous ammonia at a flow rate of 5.4 kg/min; total of 9.9 MLD. Each tank will hold a volume of 24 $m^3$, however additional specifications can be found in **Table 3.4.5**.

**Table 3.4.5:** Agitation Tank Design Specifications

| | |
|---|---:|
| Total Volume | 24,000 L |
| Tank Height | 7.7 m |
| Tank Diameter | 2.8 m |
| Impeller Diameter | 0.7 m |
| Impeller Speed | 40 RPM |
| Material Type | Hastelloy C |

*3.4.6. Rotary Drum Filter*

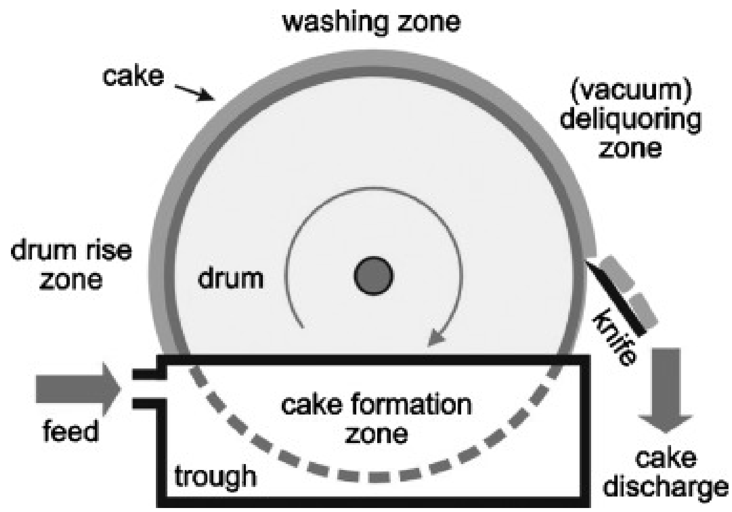



**Figure 3.4.6:** Rotary Drum Filter Diagram

Rotary drum filters serve as mechanical units that efficiently separate solid and organic matter from water. These filters operate by rotating a cylindrical drum immersed in a liquid, where solids are captured on the drum's surface and removed by a scraper or vacuum system, allowing clarified liquid to pass through, such as the one depicted in **Figure 3.4.6**. This type of filter will be used in order to separate the fertilizer product from the solution produced during the mixing stage. We will purchase this rotary drum filter from Mat-Kuling. These drum filters boast corrosion-resistant mechanisms requiring minimal maintenance and no lubrication, making them easily manageable by all staff members. With their robust structure, they offer long warranty duration. Moreover, these filters are energy-efficient, offering both intermittent and continuous backwash modes to minimize water consumption. Equipped with Smart Wash Logic Control, operators can easily select operational modes, adjust rotational speeds, and autonomously optimize washing speed and time for maximum efficiency (*Drum Filters for Aquaculture - Land Based Salmon Farms*, n.d.).

First, the filter cake, a 50/50 wt% of liquid salt water to solid product, will be scraped with a blade from the drum's surface. Next, the fertilizer cake will enter a 9-stage evaporator to further dry and solidify into our final fertilizer product. Moreover, the filtrate, also known as the rock salt solution, will enter a separate 1-stage evaporator to dry before forming our rock salt product.

Our rotary filter will have a diameter of 0.2 m and length of 0.2 m and there will be 4 of these filters. Additionally, based on a linear velocity of 16.6 cm/sec, the system requires 15.9 RPM. Assuming a 4 inch filter cake and a 5.5 bar/m pressure gradient through the depth of the cake, the pressure gradient is 0.56 bar.

### 3.5 Production and Maintenance

*3.5.1 Overall Process*

The desalination plant will run a total of 8,400 hours per year to make a 96% uptime rate. Because desalination is a continuous process, it is critical that we keep the plant running year round. By operating the desalination plant for a significant portion of the year, communities can secure a reliable and sustainable water supply. In not running the plant at 100% uptime rate and by having multiple redundancies, we can ensure time for cleaning and maintenance without having to ever completely shut down our process.

*3.5.2. Pretreatment*

Pretreatment is crucial for removing impurities, suspended solids, microorganisms, and other contaminants from the feed water to protect the desalination equipment and ensure efficient operation. Because we are working with saltwater as the feed, if regular maintenance is not done, fouling will occur. Routine maintenance must be done on all our major units: strainers, microfilters, and ultrafilters.

The strainers that will be used are self-cleaning, eliminating the need for human intervention when it comes to their maintenance. For both microfiltration and ultrafiltration, backwashing will be performed once a week. Because of this, extra membranes will be needed to account for the continuous flow of inlet water. In total we will have 160 microfiltration modules. In order to account for the total flow rate coming in, 147 active microfiltration modules will be needed, along with an additional 13 modules. These 13 additional modules will not be in operation. The process of backflushing will take 2 minutes per module. To ensure that each module is backwashed once a week, 23 modules will be fully flushed each day. For ultrafiltration, 56 modules are needed for the amount of inlet flow rate coming through. In addition to those, there will be 6 extra ultrafiltration modules. This means at any given time there will be 6 modules not in operation. Backflushing for these will also be 2 minutes per module. To ensure that each module is backwashed once a week, 9 modules will be fully flushed each day.

*3.5.3. Reverse Osmosis*

Considering the configuration of the RO system, to ensure the optimal functioning of RO membranes regular backflushing and cleaning must be done to prevent significant fouling. This necessitates scheduling backflushing activities every 30 to 90 minutes, requiring an adequate number of extra membranes and pressure vessels in each stage to maintain continuous operation. In stage 1, comprising 288 pressure vessels and 864 membrane elements, there are 10 additional pressure vessels allocated for backflushing purposes, ensuring 278 pressure vessels remain active at all times. Similarly, in stage 2, with 196 pressure vessels and 588 membrane elements, 12 additional pressure vessels are designated for backflushing, leaving 184 pressure vessels operational. An automated system facilitates these backflushing procedures seamlessly.

This process typically takes around 3 minutes per pressure vessel. In stage 1, divided into 12 units with 24 pressure vessels per unit, one backflushing cycle for 24 membranes consumes 72 minutes, totaling 20 cycles per unit per day. In stage 2, divided into 7 units with 28 pressure vessels per unit, one backflushing cycle for all 28 pressure vessels takes 84 minutes, totaling 17.1 cycles per unit per day. We will use 2% of the permeate water coming out of reverse osmosis for backwashing of RO and pretreatment, which is about 0.76 MLD. Moreover, it's imperative to replace RO membranes entirely every 2 years to maintain optimal system performance and longevity.

*3.5.4. Permeate Post-Treatment*

It is recommended to clean lime saturators and recarbonation systems every 6-12 months, and chlorination tanks should undergo cleaning every 6 months. Additionally, incorporating a valve in lime slurry tubes enables occasional backflushing to mitigate the accumulation of significant deposits.

*3.5.5. Brine Post-Treatment*

Brine Post-Treatment is a crucial aspect of the design, aimed at transforming the common by-product of desalination, brine, into valuable resources like fertilizers and rock salt. This process involves multiple unit operations, emphasizing the necessity of regular maintenance to ensure smooth operation and safety.

Priority maintenance is allocated to the evaporation units and heat exchangers. The evaporation units, comprising large vessels arranged in a series of stages, require thorough inspection for any cracks or damage to the liners of the carbon steel in each stage. This preventive measure not only prevents leaks but also helps maintain the required temperatures and pressures for optimal performance.

Similarly, the heat exchangers, utilized in both the evaporation units and before the sparging mixing tank to adjust brine solution temperatures, demand attention. Cleaning the tubes to remove scale and marine growth is essential for efficient heat transfer. Techniques like high-pressure sprayers, such as hydro lasers, are effective in dislodging buildup by applying pressure against it. Moreover, inspecting the shell and tube areas for damages and cracks and promptly replacing tubes as needed are critical steps to ensure the overall efficiency of the operation.

## 3.6 Economics

### *3.6.1. Material Expenses*

**Table 3.6.1:** Material Expenses

| Process Unit | Materials | Amount Required (Mass/Day) | USD/kg | Daily Expense |
|---|---|---|---|---|
| Pretreatment | Aluminum Sulfate | 4.2 tons | $0.27 | $1,027.73 |
| Post-Permeate Treatment | 90% Calcium Hydroxide (Lime) | 4.3 tons | $0.34 | $1,307.95 |
| | Carbon Dioxide | 4.6 tons | $0.67 | $2,795.94 |
| | Chlorine | 150 kg | $2.20 | $334.22 |
| Post-Brine Treatment | 75% Phosphoric Acid | 7,140 tons | $0.89 | $5,765,420.00 |
| | 29% aq. Ammonia | 8 tons | $0.21 | $1,517.67 |
| **Total Daily Expense** | | | | $5,772,420.00 |
| **Total Annual Expense** | | | | $2,020,341,227.86 |

The material expenses for the plant, found in **Table 3.6.1**, arise from pretreatment, post-permeate and post-brine treatments. These include Aluminum Sulfate for pretreatment, Calcium Hydroxide, Carbon Dioxide, and Chlorine for post-permeate treatment, and significant quantities of Phosphoric Acid and Ammonia for post-brine treatment. Notably, Phosphoric Acid comprises a significant portion of the total daily expense due to its large quantity requirement. Additionally, despite its low mass requirement, Carbon Dioxide significantly contributes to daily expenditure due to its relatively higher cost per kg compared to other materials listed in **Table 3.6.1**. Overall, these expenses sum up to approximately $2.0B annually.

*3.6.2. Utility Expenses*

**Table 3.6.2:** Annual Utility Expenses

| Utility | Usage (Unit/day) | Unit Price (USD/unit) | Annual Expense |
|---|---|---|---|
| Steam[(1)] | 7310 kg/day | \$0.018/kg | \$46,053.00 |
| Cooling Water[(2)] | 16131 $m^3$/day | \$0.06/$m^3$ | \$338,746.65 |
| Pumps[(3)] (Electricity) | 12435 kWh/day | \$0.076/kWh | \$8,136,791.04 |
| 21-Stage Evaporation Unit[(4)] | | | \$56,000,000.00 |
| 9-Stage Evaporation Unit[(4)] | | | \$24,000,000.00 |
| 1-Stage Evaporation Unit[(4)] | | | \$2,700,000.00 |
| **Total Utility Cost** | | | \$91,221,590.69 |

*Note.* (1) This is only the steam used in the heat exchanger/brine heater in the first stage of the 21-stage evaporation unit. (2) This is the cooling water used in the heat exchanger before the agitation tank. (3) This is the total pumping power for pumps not included in the evaporation units. (4) This includes running heat exchangers and pumps in all the evaporation units.

The annual utility expenses of the water desalination plant comprise three main components: the cost of steam, cooling water, the cost of running pumps, and the operation of the three multi-stage evaporation units, as outlined in **Table 3.6.2**. The cost of steam, priced at \$0.018 per kg, is calculated for the heat exchanger/brine heater before the 21-stage evaporation unit, making up a mere 0.05% of the total annual utility cost due to the plant's low number of heat exchangers (*Industrial Steam Cost: Industrial Utilities*, n.d.). The cost of cooling water, priced at \$0.06 per $m^3$, is also calculated for the heat exchanger before the agitation tank, refer to **Figure 3.4.4**, which only takes up about 0.37% of the total annual utility cost due to the low number of heat exchangers needed (*Cooling Water Cost: Industrial Utilities,* n.d.). The cost of running pumps, at \$0.076 per kWh based on the industrial market price in New Orleans in November 2023 (Zdanov, 2023), applies to pumps outside the evaporation units and accounts for about 9% of the total annual utility cost. These pumps, mostly operating at low pressures and benefiting from relatively low electricity prices, contribute modestly to the overall expenses. Conversely, the substantial cost of running the three multi-stage evaporation units encompasses pumping and heat exchangers utilized in the process, constituting the majority share of approximately 91% of the total annual utility cost (Daly et al., 2016). This allocation reflects the scale of these units, necessary to handle the high flow rates of water through the desalination process.

### *3.6.3. Operating Costs*

**Table 3.6.3:** Annual Operating Expenses

| | **Operating Costs** |
|---|---:|
| Materials | \$2,020,341,227.86 |
| Utilities | \$91,221,590.69 |
| Labor (including Benefits 1.3x) | \$11,104,788.50 |
| **Total Operating Costs** | \$2,122,667,607.05 |

The annual operating expenses consist of the cost of materials, utilities, labor, and benefits that will be distributed yearly. These specific values are listed in **Table 3.6.3.** Furthermore, the costs for materials and utilities are shown in **Table 3.6.1** and **Table 3.6.2,** respectively.

$$N_{OL}=(6.29 + 31.7P^2 + 0.23N_{np})^{0.5} \qquad \textbf{(Equation 3.6)}$$

Using **Equation 3.6,** the amount of operators needed for a single shift was calculated. In this equation, $N_{OL}$ represents the number of operators per shift, P represents the number of processing steps, and $N_{np}$ represents the number of nonparticulate processing steps. This was calculated to be 41 operators necessary per shift. A total number of operators for the plant was calculated to be 164 by multiplying 41 by 4 in order to account for three shift cycles per day, holidays, sick days, and vacation days. The plant will have 1 plant manager with a salary of \$138,100 per year, 11 engineers and supervisors with a salary of \$88,745 per year, and 164 lower level workers and operators with a salary of \$45,113 per year. These salaries are the average salaries for each respective role in Louisiana (*What is the Average Salary in Louisiana?*, n.d.). The sum of these salaries was multiplied by 1.3 in order to account for any benefits the operators may receive. This results in a labor cost of about \$11MM.

*3.6.4. Equipment and Capital Costs*

**Table 3.6.4:** Equipment Costs

| **Process Step** | **Equipment** | **Total Expense (USD)** |
|---|---|---|
| Pre-treatment | Strainers, Filters, Membranes, Tanks | \$73,724,311.21 |
| Reverse Osmosis | Membrane, Pressure Vessels | \$45,862,155.60 |
| Permeate Post-Treatment | Tanks, Lime Dosing, Carbonation, Direct Injector | \$6,790,825.55 |
| Brine Post-Treatment | Multi-Stage Evaporators, Rotary Drum Filter, Storage Tanks, Agitators | \$565,385,404.30 |
| Ancillary Equipment | Pumps, Heat Exchangers, etc. | \$35,270,411.42 |
| **Total Equipment Cost** | | \$727,033,108.08 |

The equipment costs for our plant can be found in **Table 3.6.4**. From the process flow diagram, we compiled a list of all major equipment needed to make our plant run. For many pieces of equipment, we were able to find exact brands and models online that worked well for our process. These sources provided exact pricing of much of the equipment. For equipment prices that were not updated for the current year, the chemical engineering plant cost index

(CEPCI) tool was used to adjust the cost. For the pieces of equipment where prices weren't easily found by manufacturers, cost curves were used to approximate the costs (Towler & Sinnott, 2012).

Brine Post-Treatment equipment accounts for the majority of our total equipment costs. The 21 stage evaporator unit alone costs $226.6M and includes pumps, heat exchangers, and vessels (Daly et al., 2016). Furthermore, all three of the multi-stage evaporators make up 82% of the brine post-treatment equipment cost. Also notice the process step labeled ancillary equipment that incorporates the costs of pumps and heat exchangers. These items were placed on their own line due to how specific costing of them is. The total equipment expense is $727MM.

**Table 3.6.5:** Capital Costs

| Capital Costs | Expense |
|---|---|
| Land | $268,000.00 |
| Equipment | $727,033,108.08 |
| Solid-Fluid Processing Plant (Installation, Piping, Controls, etc.) | Lang Factor: 3.63 |
| **Total Initial Capital Investment** | $2,639,130,182.33 |

The capital costs of our plant can be found in **Table 3.6.5** and takes into account purchasing of the land, equipment costs, installation, and other expenses. Given the size of our plant, it was determined that 10 acres of land were needed. The average price per acre of land in Louisiana is $26,800, making the total cost of land $268,000 (Morris, 2023). Additional information regarding the direct and indirect expenses can be found in **Table 3.6.5** below.

**Table 3.6.6**: Total Direct and Indirect Plant Costs

| Direct Cost | Solid-fluid Plant | Cost (USD) |
|---|---|---|
| Purchased Equip. Delivered | 100 | $616,619,201 |
| Purchased Equip. Installation | 39 | $240,481,488 |
| Instrumentation & Controls | 26 | $160,320,992 |
| Piping | 31 | $191,151,952 |
| Electrical systems | 10 | $61,661,920 |
| Buildings | 29 | $178,819,568 |
| Yard improvements | 12 | $73,994,304 |
| Service facilities | 55 | $339,140,560 |
| Total direct plant cost | 302 | $1,862,189,988 |
| Indirect Costs | | |
| Engineering and supervision | 32 | $197,318,144 |
| Construction expenses | 34 | $209,650,528 |
| Legal expenses | 4 | $24,664,768 |
| Contractor's fee | 19 | $117,157,648 |
| Contingency | 37 | $228,149,104 |
| Total indirect plant cost | 126 | $776,940,193 |
| Fixed Capital Investment | 428 | $2,639,130,182 |
| Working Capital | 75 | $263,913,018 |
| Water Transportation Pipeline | - | $300,000,000 |
| Total Capital Investment | 503 | $3,203,043,200 |

*3.6.5 Total Product Revenue*

**Table 3.6.7:** Total Product Revenue

| Product | Daily Production Amount | Sale Value (USD/unit) | Daily Revenue | Annual Revenue |
|---|---|---|---|---|
| Tap Water | 10M Gallons | $3.48/m^3$ | $130,152.00 | $45,553,200.00 |
| Fertilizer | 15 Mkg | $0.58/kg | $8,727,450.00 | $3,054,607,500.00 |
| Rock Salt | 13 Mkg | $0.08/kg | $1,003,102.10 | $351,085,735.00 |
| **Total Daily Revenue** | | | | $9,860,704.10 |
| **Total Annual Revenue** | | | | $3,451,246,435.00 |

The total revenue of our desalination plant is shown above in **Table 3.6.7** is based on the following three products: tap water, divalent metal ammonia phosphate based fertilizer, and rock salt. First, Algiers's tap water, with a daily production of 10 million gallons, yields an annual revenue of almost $46MM (Salas, 2023). Next, our rock salt produced at 13 million kg daily has a lower sale value compared to our fertilizer product, but will account for over $350MM in annual revenue. Finally, the divalent metal fertilizer is our most profitable product with a sale value of $0.58 per kg based on current trends of Diammonium Phosphate (DAP) fertilizer sold in the US (*US Diammonium phosphate spot price (Gulf)*, n.d.). It is important to highlight that this product's market value can fluctuate based on different market trends. According to Procurement Resource, "–the price trend for DAP (Diammonium Phosphate) are expected to continue behaving in a similar fluctuating manner since the dull demands and high inventories will continue to influence the market dynamics." (*DAP (diammonium phosphate) price trend and forecast*, 2022). However, it is important to note that DAP is only a reference product and the price of the actual fertilizer will vary based on market demands. This is critical to the profitability of our plant and trials of our product should be conducted prior to full scale up.

Using DAP pricing, this product contributes significantly to the profitability of our plant, and accounts for approximately 90% of our total annual revenue of $3.5B.

*3.6.6 Alternative Scenarios & Cash Flow Position*

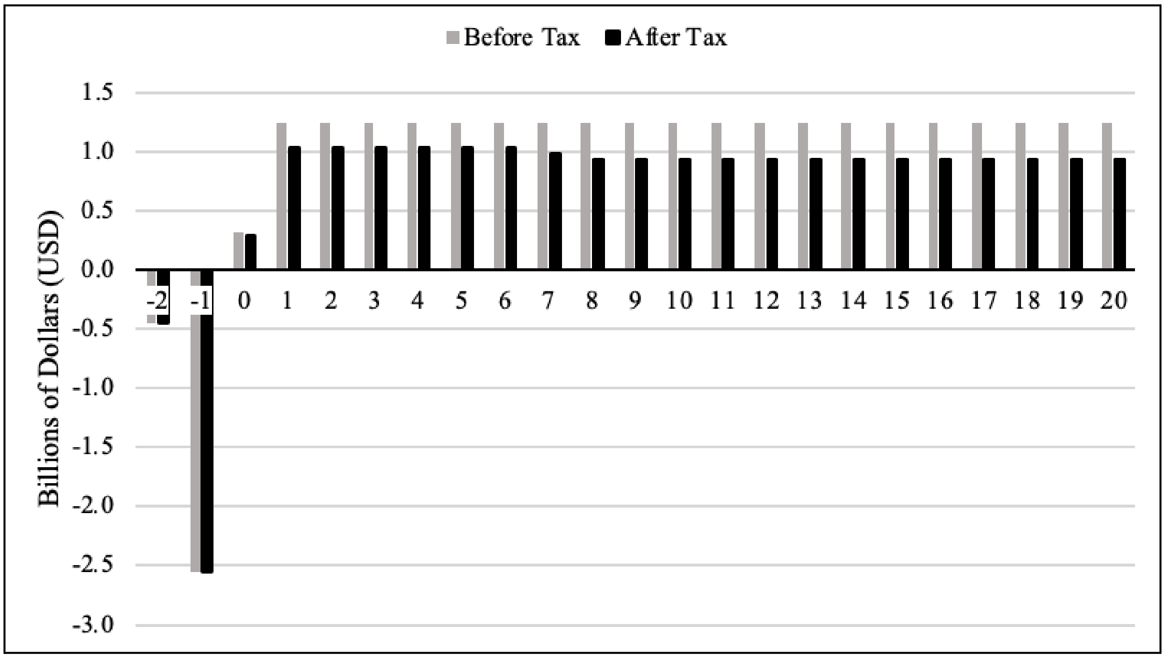



**Figure 3.6.6:** Desalination Plant's Tax Cash Flow Position

Using **Figure 3.6.6**, we were able to get a sense of what our cash flow position would look like assuming a total plant life of 22 years, including 2 years of construction and purchasing of equipment. During the first two years of construction, the plant will be in approximately $3B in debt which is equivalent to our initial capital investment. At year zero, we are able to start operations at 50% capacity and will begin profiting approximately $300MM after tax. Between year zero and year eight, we will save money on taxes assuming an 8-yr straight line depreciation on our equipment. After year eight we have a steady after tax cash flow of $930MM. The after-tax cash flow appears to be positive for all of the operational years, indicating a financially viable project post-tax obligations. This trend suggests not only a swift recovery of the initial investment but also a strong, stable financial future of the plant which can help guarantee dividends to paid out shareholders each year.

**Table 3.6.8:** Alternative Scenarios' IRR & Net Present Value After 20 Years of Operation

| Scenario | IRR | Net Present Value |
|---|---|---|
| Without Fertilizer & Rock Salt<br>*(Tap Water Transportation Included)* | 0% | -$13,418,062,988.89 |
| Without Fertilizer & Rock Salt<br>*(Tap Water Transportation Cost is $0)* | 46% | $674,457,399.41 |
| Current Proposal of Plant Design<br>*(Fertilizer, Rock Salt, and Tap Water Transportation Included)* | 27% | $4,069,697,111.54 |

*Note.* The discount rate for Net Present Value is assumed to be 10%.

In order to evaluate the profitability of the plant, we compared alternative scenarios to our current plant design. As shown in **Table 3.6.8**, both the internal rate of return (IRR) and net present value after 20 years of operation were used as metrics to determine if the scenario promoted an economically viable alternative. A fundamental consideration when calculating the net present value (NPV) of an investment is the time value of money. The principle of the time value of money suggests that a dollar received today holds greater worth than a dollar received in the future (Fernando, n.d.). After further analysis, a net present value of the plant would be negative $13.4B after 20 years if it were to operate solely as a desalination plant without producing fertilizer or rock salt. It's important to note this is only the case if a 60 mile tap water transportation pipeline is included in the capital and annual utility costs. This helps us further highlight the profitability of our side products and emphasizes their need in our plant design. However the second scenario, where the cost of tap water transportation is assumed to be zero annually, results in an IRR of 46% and a positive NPV of approximately $670MM. This indicates that without transportation costs, the investment becomes financially attractive. In summary, the current plant design, even with a smaller IRR, has an NPV of over $4.0B and should be considered by shareholders for potential investment.

### 3.7 Safety, Health, and Environment

In order to achieve a successful capstone project, our proposal on the design of a desalination plant in New Orleans, Louisiana, must require a comprehensive analysis of societal, environmental, and safety considerations. This project holds significant importance in addressing the water crisis facing the Gulf region and presents a holistic approach to sustainable water management.

From an environmental perspective, the project acknowledges the negative impacts of traditional diammonium phosphate (DAP) fertilizer production. While fertilizer production can contribute to nutrient runoff, water pollution, and eutrophication of water bodies, diammonium phosphate stands out for its significant benefits in agriculture. Ammonium phosphate, a common phosphatic fertilizer, is widely used due to its high nutrient content and favorable physical properties (Phosphate Fertilizers, n.d.). It provides essential phosphorus and nitrogen for plant nutrition, supporting the growth of crops such as wheat, barley, fruits, and vegetables.

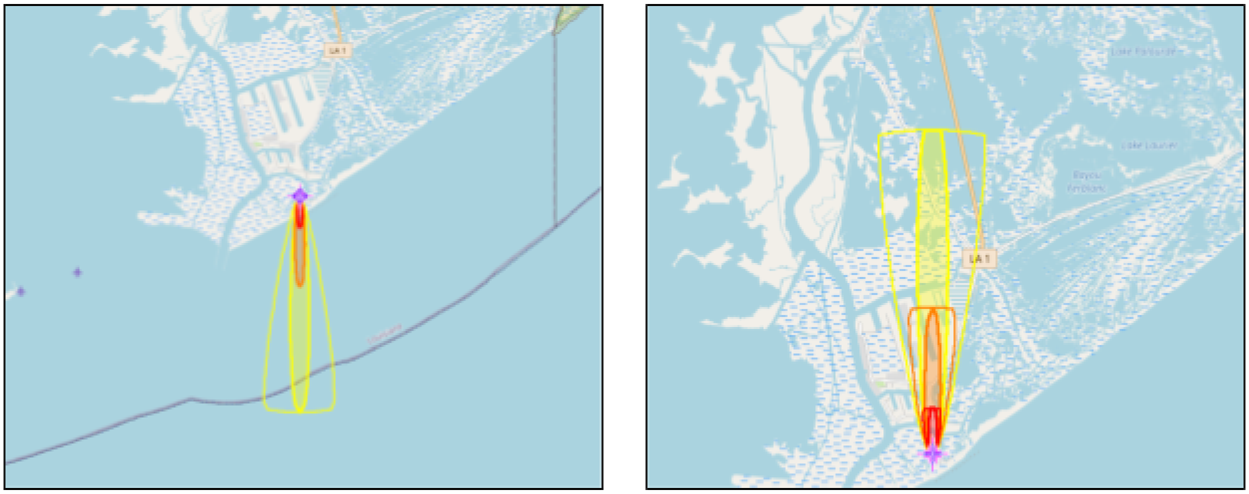
However, the production and use of diammonium phosphate based fertilizers also come with challenges. The extraction and processing of phosphate rock can have environmental consequences, including habitat destruction and water resource depletion. Moreover, the runoff of excess fertilizers into water bodies can lead to algal blooms, oxygen depletion, and ecosystem disturbances.

Regarding safety, the project acknowledges the risks associated with both chemical exposure and high pressures surrounding the reverse osmosis units. The high pressure required to operate a reverse osmosis system can be dangerous if not properly maintained and controlled. Electrical hazards from equipment in the plant, such as pumps and controls, also pose risks (Lucky, 2023).

The three main chemical hazards for the plant include the storage of 75% phosphoric acid, 29% aqueous ammonia, and chlorine gas. First, the storage of 29% aqueous ammonia is considered our most dangerous hazard due to the large amount contained on site. Ammonia is highly corrosive, toxic to aquatic environments, and is acutely toxic including respiratory illness if inhaled (CFIndustries, n.d.). To mitigate a potential loss of primary containment, a dike and water deluge system will be built around each storage container.

Additionally, 75% phosphoric acid is highly corrosive and if come into direct contact with can cause severe burns and eye damage (Univar Solutions, n.d.). Fortunately, phosphoric acid is not volatile or flammable unless mixed with incompatible materials. Similarly a dike and water deluge system will be built around each phosphoric acid storage container to mitigate any potential release scenarios.

Finally, the use of chlorine gas for water treatment, while effective in destroying harmful bacteria and viruses, can also pose significant risks to human health and the environment (admin, 2023). Exposure to chemicals used in the pretreatment process can be hazardous without proper protective equipment (Lucky, 2023). Accidental releases of chlorine gas can cause severe respiratory problems and environmental damage. It’s also important to note that although we have small inventories kept on site we will need weekly shipments of chlorine via rail to keep up with production requirements.



*Note.* The left image is modeled during the month of January, while the right image is during July.

**Figure 3.7:** ALOHA Ammonia Release Models

Next, we wanted to analyze the plant's largest hazard; loss of primary containment of an Ammonia tank. We assumed that a two foot long crack would develop towards the bottom of one of our tanks. After one hour, approximately 262,675 pounds or 51,000 gallons of ammonia was released into the surrounding environment. Additional material regarding the assumptions made for these models can be found in Appendix E. The IDLH for ammonia, also known as the concentration that is immediately dangerous to life and health, is 300 ppm. As shown in **Figure 3.7**, the red area for both models represents concentrations at or above the IDLH. The model shows during the month of January concentrations exceeding 300 ppm goes out about 1200 yards in the southerly direction. While the July model shows a more severe consequence in which the IDLH concentrations can be seen up to almost one mile away from the plant in the northern direction. Although Port Fourchon is far from populated areas, this hazard would still result in multiple fatalities on site. Thankfully, due to the sites' remote location, impact to residential areas will be minimal if this event were to occur. As a result of this study additional layers of protection including both dikes and deluge systems will be installed for each tank in order to mitigate the impact of similar releases. Moreover, routine maintenance and proper emergency response will also be implemented to ensure the safety of all individuals on site.

Despite these challenges, the societal benefits of desalination are evident. Access to clean water is fundamental to various aspects of society, including agriculture, industry, and overall economic development. Desalinating water provides drinkable water in places where it is scarce, supporting larger populations and fostering economic growth. For instance, the United Arab Emirates heavily depends on water desalination for drinking water, and the prosperity of the nation has increased with improved access to fresh water (Water Supply and Sanitation in Abu Dhabi).

In conclusion, this capstone project represents an approach that carefully considers environmental stewardship, safety measures, and societal benefits. Through innovative technologies, responsible practices, and stakeholder engagement, the project aims to address water scarcity challenges while promoting environmental integrity, public health, and social well-being in the region.

## 4. Final Design

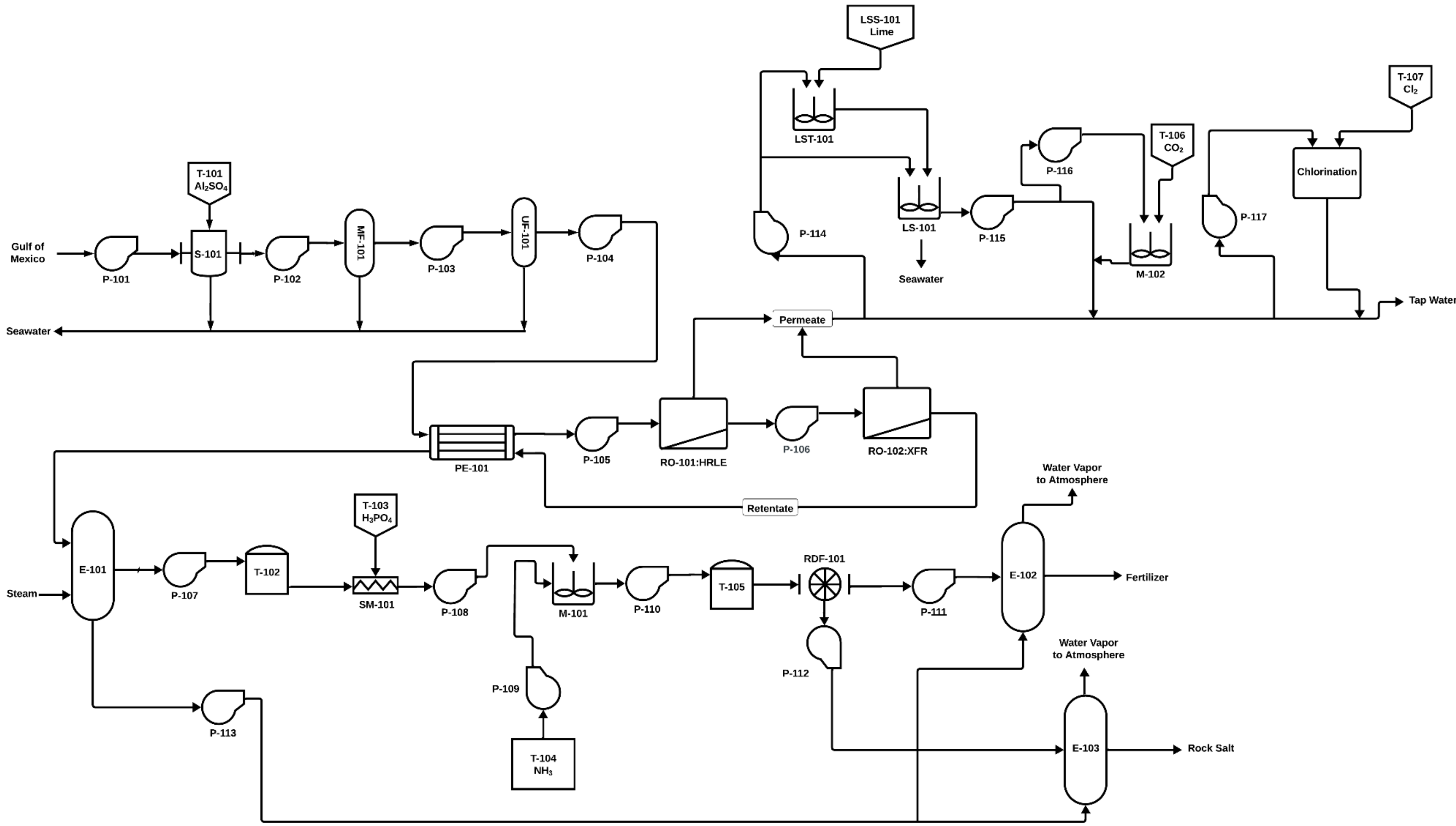



**Figure 4:** Overall PFD of Final Design

## 4.1 Pretreatment

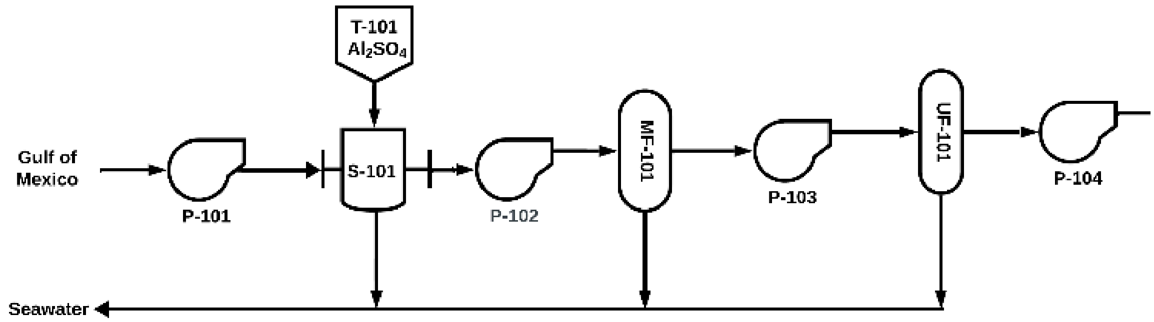



**Figure 4.1:** PFD for Pretreatment

The pretreatment process takes in 106.4 MLD of seawater from the Gulf of Mexico, where it goes through a series of filtration processes, preparing it for reverse osmosis. The pretreatment process is divided into 4 parts: coagulation, straining, microfiltration, and ultrafiltration. To the initial 106.4 MLD coming into the system 3,790 kg/day of aluminum sulfate coagulants will be added. These coagulants will promote the aggregation of suspended particles, colloids, and other impurities in water, forming larger flocs that are easier to remove through subsequent filtration processes. Following the addition of the coagulants, the seawater will then go through a strainer. With a 99.2% recovery rate, 2 *FilterSafe Coarse Filtration Leviathan 300* strainers will be used. From the strainer, 105.5 MLD of seawater will flow into microfiltration. To remove small-diameter dispersed solids at a water recovery rate of 90%, 147 *Alfa Laval MF Spiral Membranes* will be used. The 0.9 MLD of waste coming out of the strainer will be combined in a waste stream with the 10. MLD of waste coming out of the microfiltration, that will then be pumped back into the Gulf of Mexico. Ultrafiltration will then follow microfiltration. From microfiltration, 95 MLD will be pumped into the *Dow Ultrafiltration Module Model SFP 2880*. Based on the flow rate each ultrafilter will be able to handle, 56 modules will be needed. Given that ultrafiltration has a recovery rate of 95%, 90.2 MLD of seawater will continue and flow into RO, and 4.8 MLD of waste will join the waste stream and also be pumped back into the Gulf of Mexico. The careful selection of filtration equipment in the pretreatment process ensures efficient removal of contaminants, safeguarding downstream processes, and ultimately contributing to the optimal performance of the desalination plant.

## 4.2 Reverse Osmosis

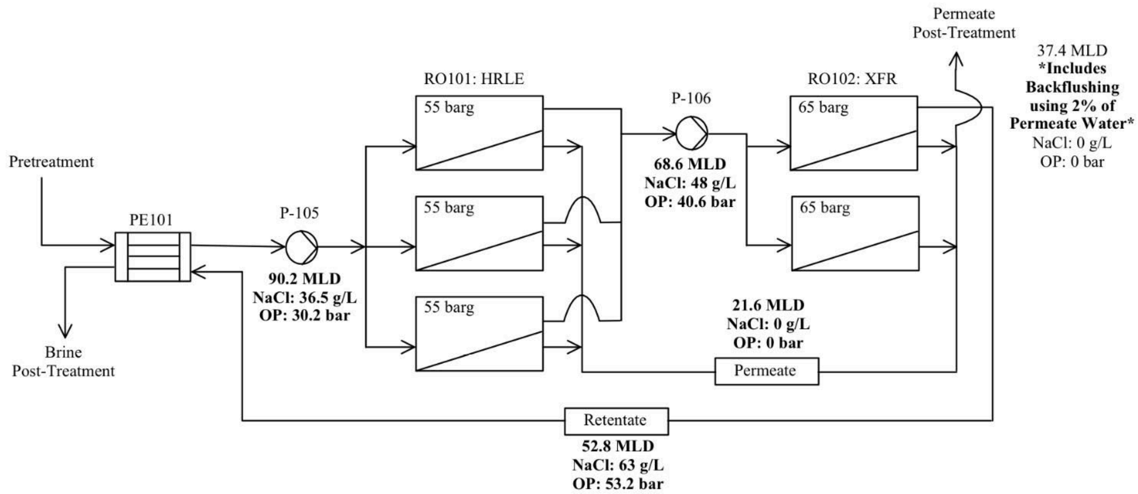



**Figure 4.2:** PFD for Reverse Osmosis System

The reverse osmosis (RO) process for desalinating Gulf of Mexico's seawater takes an inlet of 90.2 MLD and features two stages with a 24% recovery rate between them, yielding a production capacity of 37.3 MLD of treated water. This process is organized as a double pass reverse osmosis (DPRO) system, ensuring optimal efficiency and resource utilization. The initial stage comprises 864 RO membranes housed within 288 pressure vessels, while the second stage incorporates 588 RO membranes within 196 pressure vessels, designed in a 3:2 array configuration. To maintain a continuous flux of water across each membrane, operating pressures are carefully calculated based on osmotic pressures, with the first stage vessels operating at 55 barg and the second stage at 65 barg.

The chosen membrane elements, supplied by Dupont, exhibit superior performance characteristics, with the first stage employing FilmTec™ SW30HRLE-370/34i Wet membranes and the second stage utilizing FilmTec™ SW30XFR-400/34i Wet membranes. These membranes, composed of polyamide thin-film composite material, ensure high salt rejection, reduced fouling, and lower maintenance costs. In total, the RO process requires 1452 membrane elements and 484 pressure vessels, with each stage supported by pumps consuming 4.2 kW of hydraulic power to meet water requirements.

### 4.3 Permeate Post-Treatment

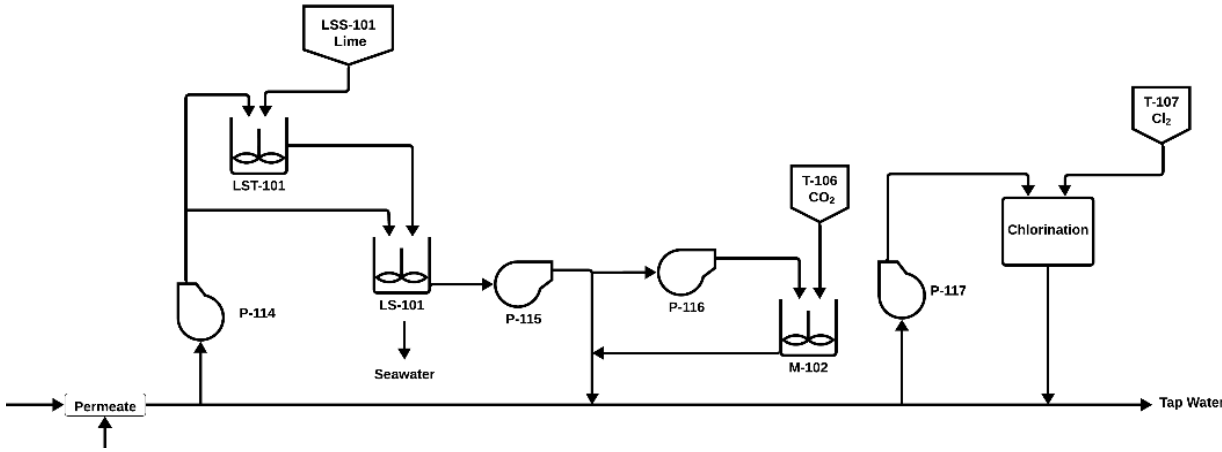



**Figure 4.3:** PFD for Permeate Post-Treatment

Following RO, 37.4 MLD of water will flow through the permeate post-treatment process. This process is made up of 3 steps: remineralization, recarbonation, and chlorination. From the main 37.4 MLD of water, there are split streams of 4.3 MLD flowing through remineralization and recarbonation and 0.05 MLD added to a lime slurry tank with 4.3 tons of calcium hydroxide per day in a lime slurry tank. The two streams will be mixed at a retention time of 12 minutes. From the lime slurry tank, the slurry flows into the lime saturator and is combined with the rest of the 4.3 MLD to create a limewater. Due to the volume, 2 lime saturators of height 5 meters and diameter of 3 meters will be needed. Because of the use of 90% pure lime, there is a waste stream coming out of the lime saturator. This waste stream of 162 kg/day can be combined with the pretreatment waste streams and pumped back into the Gulf of Mexico. From the limewater stream, a side stream is pressurized and is added to a pressurized stream of $CO_2$ to recarbonate the water. This process will utilize the *TOMCO$_2$ Pressurized Solution Feed*. With an efficiency of 95%, the process will require 4.6 tons per day of $CO_2$. The newly carbonated stream and the rest of the limewater stream will then reenter the main stream of water of 37.4 MLD.

After both remineralization and recarbonation, the water has been hardened. The third and final step that the water must go through to get to the tap is chlorination. Based on national water tables, New Orleans has 4 ppm chlorine in their water. This will be achieved using a vacuum chlorinator injecting 151.6 kg of chlorine gas per day. Direct injection was decided to be the most effective way to add chlorine gas to the water in order to kill parasites, bacteria, and algae, along with protecting the pipes as the water is pumped to the population's tap. The total stream of 37.4 MLD of potable water is what will be sold.

## 4.4 Brine Post-Treatment

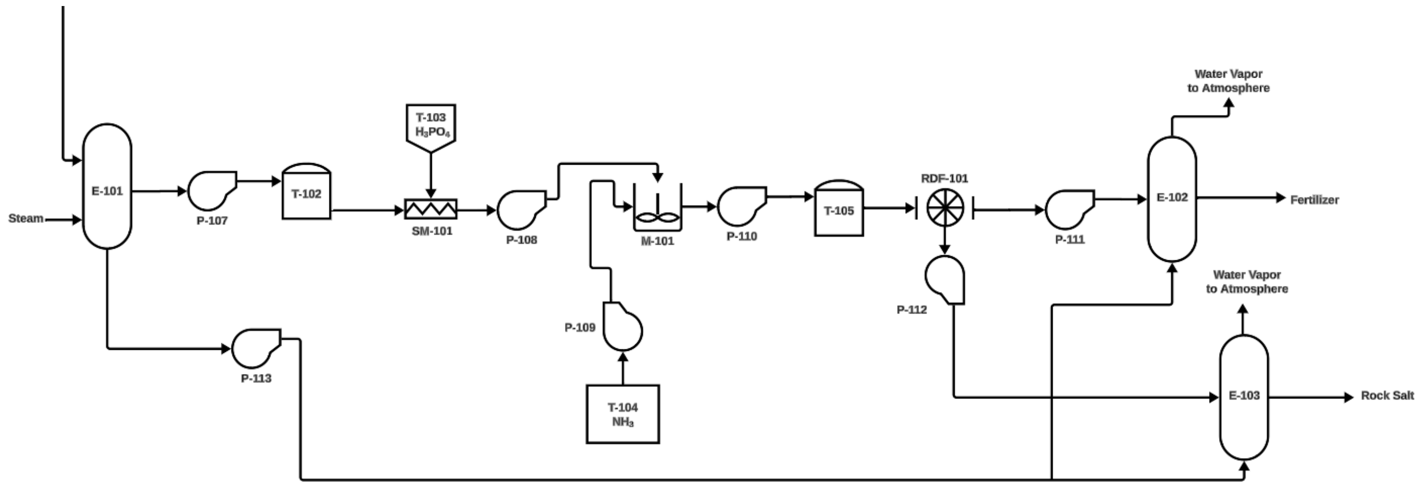



**Figure 4.4:** PDF for Brine Post-Treatment

Following RO, the retentate brine solution will flow through the brine post-treatment process. The first step in this process is vaporizing the water content of the brine solution using an evaporation unit. This evaporation unit is a Multi-Stage Flash Distillation unit that will utilize 21 stages to vaporize the water content of 52.3 MLD brine solution to reach 85 wt% salt from 94 wt% salt. The stages will operate at decreasing pressures and temperatures to conserve energy cost by utilizing the flash point. This produces 22 MLD of less diluted brine solution as well as 30 MLD of water vapor, and will require a heat duty of around 900 MW using 28-bar steam.

Following the evaporation unit, five, 25,000 L storage tanks are placed to analyze the solution for divalent metals to determine the amounts of phosphoric acid and ammonia needed to be added to the static inline mixer and agitation tank used in the process.

A static inline mixer is employed to homogeneously mix 4.1 MLD of 75% phosphoric acid with a brine solution, facilitated by stationary elements to induce turbulence. Following this, 8 agitation tanks, each holding a volume of 24 m$^3$, will be used in parallel to precipitate ammonia phosphate fertilizer, aided by spargers introducing vapor 29% aqueous ammonia at a flow rate of 5.4 kg/min (6.9 $m^3$/min), totaling 9.9 MLD.

An additional, five, 25,000 L storage tanks are placed to analyze and adjust the pH of the solution flowing out of the mixing process by using either an acid or a base. This storage tank will have the same dimensions as the previous tank. It will accommodate a flow of 36.1 MLD of solution flowing out of the mixing process.

The plant will utilize rotary drum filters, each with a diameter and length of 0.2 m, will be used to separate fertilizer from the post-mixing solution, operating at a pressure gradient of 0.56 bar and a linear velocity of 16.6 cm/sec. Four of these filters will be deployed, featuring corrosion-resistant mechanisms and Smart Wash Logic Control. They offer both intermittent and continuous backwash modes to minimize water consumption while ensuring optimal performance.

The second evaporation unit has 28.9 MLD of solid fertilizer with entrained water entering from the rotary drum filters as a filter cake with 51.7 wt% fertilizer salt. 13.9 MLD of water is vaporized and disposed of into the atmosphere as it is no longer needed, as well as 15 million kg of fertilizer per day is produced with 99.5 wt% fertilizer salt. This unit requires a heat duty of around 390 MW, which is lower than the previous unit due to the lower flow rate entering the unit, utilizing reused water vapor from the first evaporation unit.

The last evaporation unit has 7.21 MLD of rock salt solution entering from the rotary drum filters as a liquid stream with 15.1 wt% rock salt. 2.6 MLD of water is vaporized and disposed of into the atmosphere, and 13 million kg of rock salt per day is produced with 99.5 wt% rock salt. This unit requires a heat duty of 72 MW as the flow rate entering the unit is much lower than the previous units, using reused water vapor from the first evaporation unit.

## 4.5 Pumps

**Table 4.5:** Pumping Design for Each Pumping Unit

| Pump Location | Flow Rate (MLD) | Initial Pressure (bar) | Goal Pressure (bar) | Final Pressure (bar) | Pumping Power (kW) | Pump Material |
|---|---|---|---|---|---|---|
| P-101 | 106.4 | 1 | 2 | 3.02 | 248.67 | Bronze |
| P-102 | 105.5 | 2 | 2.5 | 3.52 | 185.54 | Bronze |
| P-103 | 95.0 | 2.5 | 6 | 7.02 | 496.82 | Bronze |
| P-104 | 90.2 | 6 | 35 | 36.02 | 3132.93 | Bronze |
| P-105 | 90.2 | 35 | 55 | 56.02 | 4222.40 | Bronze |
| P-106 | 68.6 | 55 | 65 | 66.02 | 4226.50 | Bronze |
| P-107 | 22.1 | 0.107 | 1 | 2.20 | 54.54 | Bronze |
| P-108 | 26.2 | 1 | 2 | 4.80 | 114.28 | Polypropylene |
| P-109 | 9.9 | 1 | 2 | 3.00 | 23.14 | Polypropylene |
| P-110 | 36.1 | 2 | 1 | 2.20 | 10.02 | Polypropylene |
| P-111 | 28.9 | 0.56 | 1.1 | 2.50 | 65.20 | Bronze |
| P-112 | 7.2 | 0.56 | 1.1 | 2.50 | 16.24 | Stainless Steel |
| P-113 | 15.1 | 0.58 | 1.1 | 2.10 | 26.90 | Stainless Steel |
| P-114 | 4.3 | 1 | 3 | 4.02 | 15.02 | Stainless Steel |
| P-115 | 4.3 | 3 | 3 | 4.02 | 5.07 | Stainless Steel |
| P-116 | 6.0 | 3 | 5 | 6.02 | 20.96 | Stainless Steel |
| P-117 | 15 | 1 | 4 | 5.02 | 69.77 | Stainless Steel |

Pumps will be utilized to adjust pressure and transport water and brine between unit operations. Different types of pumps are used in desalination processes which include positive or dynamic displacement pumps. Positive displacement pumps tend to use air or water pressure to achieve flow. This type includes diaphragm, gear, and lobe pumps. On the other hand, dynamic displacement pumps use propellers or impellers to achieve flow. Centrifugal and submersible pumps are a few examples of this category (Elitepumps, 2018).

In this plant, dynamic displacement pumps, specifically centrifugal pumps, are used. 18 pumps are used to increase the flow pressure before entering a unit operation. This will be based on the operating pressures of the unit operation, the initial pressure flowing into the pump, and accommodations for loss of pressure due to pipes, control valves, heat exchangers, and gravity head. These pressures, as well as pump ID and construction materials, are outlined in **Table 4.5**.

Bronze alloy is used as the material of construction for salty water in the pretreatment and parts brine post-treatment processes to prevent corrosion within the pump. Polypropylene is also used for pumps in the brine post-treatment part of our process due to the phosphoric acid and ammonia that enter the system. In the permeate post-treatment process, stainless steel is used due to the low chance of corrosion occurring within the pump, as well as the price of it.

### 4.6 Water Pipeline

Since the desalination plant is located in Port Fourchon, Louisiana, and the plant intends to provide tap water for the Algiers neighborhood, the water must be transported to its final destination. A 60 mile underground pipeline will be built and used to accomplish this. The pipeline will be built using high-density polyethylene (HDPE) pipes estimated to cost $5M per mile. The capital cost of building this pipeline is $300M. In addition, taking into account maintenance, energy, personnel, and regulatory expenses, the total annual operating cost to run this pipeline is about $50M per year.

## 5. Future Works and Recommendations

The design of desalination plants is a meticulous process, balancing efficiency, safety, and cost considerations. This particular design incorporates multi-stage flash distillation units, recommended for efficient water vaporization. However, with 21 stages, the design faces complexities in practical implementation. The notable disparity between the design's seawater flow rate of 106.4 MLD and the flow rate of 200 MLD used in the referred source the design was taken from (Daly et al., 2016), significantly impacts heat and mass requirements, posing challenges in scaling up. Merely increasing scale based on flow rates overlooks non-linear factors like heat requirements and equipment selection, potentially leading to inefficiencies. Additionally, the design's reliance on pressures below local atmospheric hampers overall efficiency. Optimizing these parameters and adjusting heat duty independently for each stage could vastly improve performance, economics, and production capacity. This highlights the pressing need for further research to align desalination plant designs with real-world conditions, ultimately aiming for heightened efficiency and cost-effectiveness.

Further optimization is necessary for this plant, as the amount of pure water exiting the RO unit is lower than that of a standard desalination plant. This disparity arises from our plant's unique focus on treating brine to produce fertilizer and rock salt, alongside the production of tap water. This strategy aims to enhance profitability, reduce waste, and promote environmental sustainability. Typically, a fully optimized RO unit in a desalination plant can recover about 90% of the seawater feed; however, our plant currently achieves only around 40% recovery (U.S. Department of Energy, 2013). It would be advisable for future groups to emphasize on increasing water production capacity.

## 6. Conclusion

The proposed desalination plant in Port Fourchon, Louisiana, offers a comprehensive solution to address the salt water intrusion crisis in New Orleans, and more specifically, the Algiers neighborhood. By implementing a series of integrated processes, including pretreatment, RO separation, permeate post-treatment methods as well as brine post-treatment, the plant aims to produce tap water and ensure a reliable water supply for the community, while also producing fertilizer and rock salt in order to offer a more economically viable and sustainable solution to brine disposal.

The pretreatment process involves the use of strainers, microfiltration, and ultrafiltration to remove physical impurities and particulate matter from the seawater intake of 106.4 MLD. This step maintains the efficiency and longevity of the RO membranes in the subsequent stages by removing contaminants from the seawater. The RO system takes an inlet of 90.2 MLD and features two stages with a daily production capacity of 37.3 million liters of treated water. This process is organized as a double pass reverse osmosis (DPRO) system. This system separates the salt ions from the seawater. The permeate post-treatment stage which includes remineralization, recarbonation, and chlorination further refines the freshwater quality, ensuring compliance with US water quality standards. This stream results in 37.4 MLD that will be pumped on a 60 mile underground pipeline to the Algiers Neighborhood.

In order to address the challenge of brine disposal, the byproduct brine also enters a post-treatment stage. Methods such as evaporation units and rotary drum filters are utilized to manage and treat the concentrated brine stream effectively. As a result 15 million kilograms per day exits the plant to be sold for profit. Furthermore, 13 million kilograms per day is produced which is also sold for profit.

With diversified streams from tap water, fertilizer, and rock salt, the annual revenue generated by the desalination plant is \$3.5B, which showcases its economic viability and potential sustainability. The significant revenue from this plant demonstrates the economic benefits from the innovative processes applied in this project, not only addressing water scarcity but also contributing to economic growth and environmental stewardship in the region.

## 7. Acknowledgements

This capstone project could not have been completed without the help of many of our professors and peers. First and foremost, we extend our gratitude to Geoffrey M. Geise for guiding us in researching the correct membrane elements for both pretreatment and the RO system in our process.

Additionally, we thank Ronald J. Unnerstall for offering insightful feedback on the safety considerations of our project, including recommendations on implementing additional layers of protection for our three main hazards: ammonia, chlorine, and phosphoric acid.

Finally we want to thank National Storage Tank, Inc., specifically Emily Sloan, for providing us pricing and design information which we could incorporate into our calculations and economics sections.

Thank you to everyone who has played a part in the completion of this technical report. Your contributions have been indispensable and deeply appreciated.

## 8. Appendix

*Appendix A* - RO Pumping Power Requirements:

Stage 1: 90.2 MLD = 1.04 $m^3/s$

$(1.04\ m^3/s)(40.6\ bar)(100\ kPa/bar)$ = **4.2 MW**

Stage 2: 68.6 MLD = 0.79 $m^3/s$

$(0.79\ m^3/s)(53.5\ bar)(100\ kPa/bar)$ = **4.2 MW**

*Appendix B* - RO Membrane & Pressure Vessel Requirements

Stage 1:

Variables: Inlet Water Flow Rate: 90.2 MLD

Permeate Flow Rate: (25 $m^3$)(3 membrane elements) = 75 $m^3$ = 75,000 L/day

Single Element recovery rate: 8%; Overall recovery rate for one pressure vessel: 24%

No. of Pressure Vessels: $\frac{(90.2 \times 10^6 L/day)(0.24)}{(75{,}000\ L/day)} = 288\ Pressure\ Vessels$

No. of Membranes: $288 \times 3 = 864\ Membranes$

Stage 2:

Variables: Inlet Water Flow Rate: 90.2 MLD(1 - 0.24) = 68.6 MLD

Permeate Flow Rate: (28 $m^3$)(3 membrane elements) = 84 $m^3$ = 84,000 L/day

Single Element recovery rate: 8%; Overall recovery rate for one pressure vessel: 24%

No. of Pressure Vessels (Stage 2): $\frac{(68.6 \times 10^6 L/day)(0.24)}{(84{,}000\ L/day)} = 196\ Pressure\ Vessels$

No. of Membranes (Stage 2): $196 \times 3 = 588\ Membranes$

*Appendix C* - Lime Slurry Tank Dimension Calculations

Volume for Lime Slurry Tank:

Retention Time: 12 min

Flow Rate: (51,765 L/day)(1 day/1440 min) = 35.95 L/min

Volume: (12 min)(35.95 L/min) = 431.4 L; equivalent to approximately 115 Gallon Tank; 200 Gallon Tank is used to account for excess space and bubbling.

*Appendix D* - Evaporation Units Energy Balance

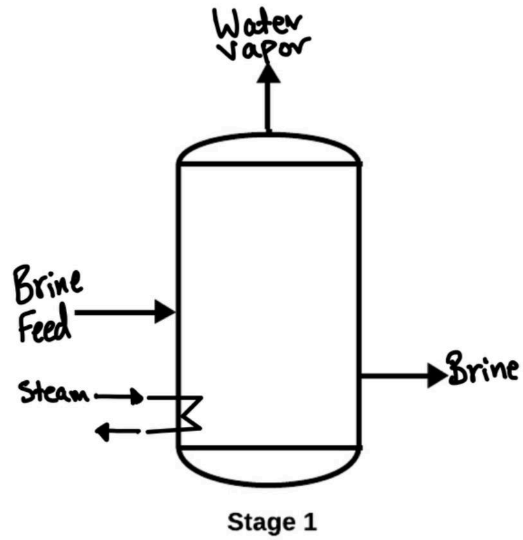



**Figure D:** Example of Stage 1 in Evaporation Unit

Specific Heat Capacity of Water: 4182 $\frac{J}{kg°C}$

Enthalpy of Vaporization of Water: 2260 $\frac{kJ}{kg}$

<u>Evaporation Unit #1</u>

Water Vapor Flow Rate (Water needed to be vaporized): (Feed Flow Rate - Brine Flow Rate) =

(52.3 $\frac{Mkg}{day}$ - 22.07 $\frac{Mkg}{day}$) = 30.23 $\frac{Mkg}{day}$

Brine Solution Temperature Change: (Final brine temperature - Initial feed temperature) =

(100°C - 25°C) = 75°C

Sensible Heat = (30.23 $\frac{Mkg}{day}$)(4182$\frac{J}{kg°C}$)(75°C)($\frac{1\ day}{86400\ s}$) = 110 MW

Heat of Vaporization = (30.23 $\frac{Mkg}{day}$)(2260 $\frac{kJ}{kg}$)($\frac{1\ day}{86400\ s}$) = 791 MW

Total Heat Duty = (110 MW) + (791 MW) = 901 MW

Evaporation Unit #2

Water Vapor Flow Rate (Water needed to be vaporized): (Feed Flow Rate - Fertilizer Flow Rate)

= (28.86 $\frac{Mkg}{day}$ - 15.00 $\frac{Mkg}{day}$) = 13.86 $\frac{Mkg}{day}$

Wet Fertilizer Heat Temperature Change: (Final product temperature - Initial feed temperature) =

(100°C - 62.5°C) = 27.5°C

Sensible Heat = (13.86 $\frac{Mkg}{day}$)(4182 $\frac{J}{kg°C}$)(27.5°C)($\frac{1\ day}{86400\ s}$) = 25 MW

Heat of Vaporization = (13.86 $\frac{Mkg}{day}$)(2260 $\frac{kJ}{kg}$)($\frac{1\ day}{86400\ s}$) = 363 MW

Total Heat Duty = (25 MW) + (363 MW) = 388 MW

Evaporation Unit #3

Water Vapor Flow Rate (Water needed to be vaporized): (Feed Flow Rate - Rock Salt Flow Rate)

= (15.6 $\frac{Mkg}{day}$ - 13.0 $\frac{Mkg}{day}$) = 2.6 $\frac{Mkg}{day}$

Rock Solution Temperature Change: (Final product temperature - Initial feed temperature) =

(100°C - 62.5°C) = 27.5°C

Sensible Heat = (2.6 $\frac{Mkg}{day}$)(4182 $\frac{J}{kg°C}$)(27.5°C)($\frac{1\ day}{86400\ s}$) = 3.5 MW

Heat of Vaporization = (2.6 $\frac{Mkg}{day}$)(2260 $\frac{kJ}{kg}$)($\frac{1\ day}{86400\ s}$) = 68.0 MW

Total Heat Duty = (3.5 MW) + (68.0 MW) = 71.5 MW

*Appendix E* - ALOHA Modeling Outputs and Assumptions

SITE DATA:
Location: PORT FOURCHON, LOUISIANA
Building Air Exchanges Per Hour: 0.91 (unsheltered single storied)
Time: **January 1, 2024 0900 hours** CDT (user specified)

CHEMICAL DATA:
Chemical Name: AMMONIA
CAS Number: 7664-41-7     Molecular Weight: 17.03 g/mol
AEGL-1 (60 min): 30 ppm     AEGL-2 (60 min): 160 ppm     AEGL-3 (60 min): 1100 ppm
IDLH: 300 ppm     LEL: 150000 ppm     UEL: 280000 ppm
Ambient Boiling Point: -28.2° F
Vapor Pressure at Ambient Temperature: greater than 1 atm
Ambient Saturation Concentration: 1,000,000 ppm or 100%

ATMOSPHERIC DATA: (MANUAL INPUT OF DATA)
Wind: 11.4 miles/hour from N at 3 meters
Ground Roughness: open country     Cloud Cover: 5 tenths
Air Temperature: 57° F     Stability Class: D
No Inversion Height     Relative Humidity: 15%

SOURCE STRENGTH:
Leak from hole in vertical cylindrical tank
Flammable chemical escaping from tank (not burning)
Tank Diameter: 10 meters     Tank Length: 33.3 meters
Tank Volume: 2,615 cubic meters     Internal Temperature: 57° F
Tank contains liquid
Chemical Mass in Tank: 1,296,454 kilograms
Tank is 80% full
Opening Length: 0.609 meters     Opening Width: 0.003 meters
Opening is 2.00 meters from tank bottom
Release Duration: ALOHA limited the duration to 1 hour
Max Average Sustained Release Rate: 4,400 pounds/min (averaged over a minute or more)
Total Amount Released: 262,675 pounds
Note: The chemical escaped as a mixture of gas and aerosol (two-phase flow).

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SITE DATA:
Location: PORT FOURCHON, LOUISIANA
Building Air Exchanges Per Hour: 0.91 (unsheltered single storied)
Time: **July 1, 2024 0900 hours** CDT (user specified)

CHEMICAL DATA:
Chemical Name: AMMONIA
CAS Number: 7664-41-7     Molecular Weight: 17.03 g/mol

AEGL-1 (60 min): 30 ppm AEGL-2 (60 min): 160 ppm AEGL-3 (60 min): 1100 ppm
IDLH: 300 ppm LEL: 150000 ppm UEL: 280000 ppm
Ambient Boiling Point: -28.2° F
Vapor Pressure at Ambient Temperature: greater than 1 atm
Ambient Saturation Concentration: 1,000,000 ppm or 100.0%

ATMOSPHERIC DATA: (MANUAL INPUT OF DATA)
Wind: 9.2 miles/hour from s at 3 meters
Ground Roughness: open country Cloud Cover: 4 tenths
Air Temperature: 84° F Stability Class: D
No Inversion Height Relative Humidity: 95%

SOURCE STRENGTH:
Leak from hole in horizontal cylindrical tank
Flammable chemical escaping from tank (not burning)
Tank Diameter: 10 meters Tank Length: 33.3 meters
Tank Volume: 2,615 cubic meters Internal Temperature: 57° F
Tank contains liquid
Chemical Mass in Tank: 1,379 tons Tank is 80% full
Opening Length: .609 meters Opening Width: .003 meters
Opening is 2.00 meters from tank bottom
Note: RAILCAR predicts a stationary cloud or 'mist pool' will form.
Model Run: traditional ALOHA tank
Release Duration: ALOHA limited the duration to 1 hour
Max Average Sustained Release Rate: 5,230 pounds/min (averaged over a minute or more)
Total Amount Released: 312,556 pounds
Note: The chemical escaped as a mixture of gas and aerosol (two-phase flow).

THREAT ZONE:
Model Run: Heavy Gas
Red: 1523 yards --- (1100 ppm = AEGL-3 [60 min])
Orange: 2.7 miles --- (160 ppm = AEGL-2 [60 min])
Yellow: greater than 6 miles --- (30 ppm = AEGL-1 [60 min])

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