The Design of a Desalination Plant in the Bay of Bengal

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

In Chennai, India, a combination of climate change and increased water pollution has led the city to a water shortage of 300 million liters per day. The implementation of a proposed RO desalination plant producing 150 MLD of potable water, will reduce the current freshwater deficiency in Chennai by 50%. The proposed design is a Reverse Osmosis (RO) Desalination plant that will take in 196 MLD of seawater from India's Bay of Bengal, and separate the water from the salt to produce two products: 150 MLD of potable water and 2.3 million kg/day of crystallized salt. This desalination plant consists of a three fold pretreatment system consisting of a strainer, ultrafiltration unit, and nanofiltration system, which removes sediment, pollutants, and unwanted metals. The treated seawater then enters the RO process where RO membranes separate the salt from the water. The resulting freshwater and brine are sent into separate post-treatment processes where they are prepared for consumers. During the final stage of production, freshwater is injected with chlorine and calcium bicarbonate in order to meet India's drinking water standards, and the brine is run through electrodialysis and a crystallizer where NaCl is entirely separated from the water and crystallized. The proposed design will have a significantly high positive impact on its community due the plant having a total recovery rate of 77.5% — a production rate over 35% higher than most saltwater desalination plants currently in production.

The overall capital cost of this project will be around \$1 billion. The yearly revenue produced from the water and food-grade salt will be around \$87MM. Using a best-case scenario of removing the crystallizer and continuing production of food grade salt, these values result in a negative profit of about \$72MM over the 20 year life of the plant, which means that overall the desalination plant is not economically favorable. However, if the plant focused solely on water

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production, there would no longer be a need for the high pumping powers to achieve a larger recovery rate of water because there could simply be more water pumped into the plant to produce 150 MLD, and then the costs associated with these pumping powers would be greatly decreased giving the plant positive economic potential. If future design teams were to take this route, salt could still be produced, but their brine streams would not be nearly as concentrated and so the brine would have to sit in salt pools for longer periods of time or require more inexpensive forms of active crystallization, requiring further research.

2. Introduction

A desalination system that can produce clean and drinkable freshwater from imported seawater could become the primary solution to solving many freshwater shortage issues all over the world. The desalination process has been part of human history for thousands of years, beginning with clay filtration systems created to trap salt from sea water (Victoria State Government, 2019). Over the years the design of desalination systems has become much more complex in order to accommodate larger populations and increased demand; nonetheless current sophisticated desalination methods continue to rely on the basic filtration and distillation concepts originally developed from ancient practices. There are approximately 16,000 operational desalination plants located across 177 different countries that collectively produce roughly 95 million m³ of freshwater per day (Victoria State Government, 2019). Many researchers have suggested that improved desalination systems could significantly aid in the reduction of global freshwater scarcity.

In areas with extreme droughts, freshwater scarcity has become a serious issue for many communities. There are currently 17 different countries, amounting to roughly 25% of the global population, that are experiencing extreme water stress, India being one of the most highly affected (Palanichamy, 2019). India's population holds approximately 1.3 billion people with over half of the population unable to regularly access freshwater (*India's water crisis: Is there a solution*, 2020). This is due to high levels of pollution in rivers and lakes throughout the country. Groundwater is currently India's main source for freshwater, however, it is not large enough to meet the country's total freshwater demand. In cities like Chennai, the effect of climate change has become detrimental. The city relies heavily on the monsoon season to provide water, but

with reduced seasonal rainfall and a rapidly growing population, relying solely on their groundwater supply is not a viable long-term solution for Chennai.

A proposed solution for Chennai's water crisis is building a desalination plant on the Bay of Bengal using reverse osmosis (RO) as the desalination technique. This desalination plant will create potable water with the goal of minimal liquid discharge to treat the concentrated brine and produce the sea salt as a coproduct. In addition to RO, there will be a pretreatment system to prepare the seawater for RO and reduce the amount of fouling on the RO membranes, as well as, a post-treatment system to prepare the clean permeate water for public consumption in compliance with US and Indian water quality regulations. This desalination design can be implemented into various water scarce communities to provide clean freshwater and salt, decrease societal stress, and significantly increase their quality of life.

3. Discussion

3.1. Pretreatment

For this desalination plant design, aspects of the conventional pretreatment process were followed to ensure proper and effective waste and seawater separation. Conventional pretreatment necessitates the use of chemicals and often includes coagulation, flocculation, sedimentation, and filtration to remove colloidal particles mechanically. However, modern seawater desalination plants have seen disadvantages of conventional pretreatment systems (Valavala et al., 2011). Common obstacles have involved difficulties removing particles smaller than 10-15 microns, fluctuations in feed water quality entering RO membranes, and coagulants influencing membrane performance. Moreover, conventional pretreatment is extremely complex, labor-intensive, and space-consuming. For these reasons, our team chose to design a pretreatment process consisting of the addition of coagulants, a strainer, an ultrafiltration system, and a nanofiltration system with a PFD of this system depicted below in Figure 3.1. Each piece of equipment requires regular maintenance and cleaning to prevent fouling. Both filtration systems have backflushing processes while the strainer contains a self cleaning function and does not require additional maintenance. A detailed description for general maintenance and backflushing is discussed in Section 3.5 Production and Maintenance.



Figure 3.1: PFD of Pretreatment Process

3.1.1. Coagulant

In order to reach a product freshwater goal of 150 MLD exiting our desalination system, the intake amount needed is 196 MLD of saltwater from the Bay of Bengal. This entering stream is combined with a coagulation chemical, where an aqueous aluminum sulfate solution is added directly to the water to neutralize negative charges on non-settleable materials such as clay and color-producing organic compounds. Based on industry standards, 100 mg of aluminum sulfate is needed per liter of water from the Bay of Bengal. Thus, using this standard, a total amount of 20 tons of aluminum sulfate are used per day. High-energy, rapid mixing is required during coagulation to spread the coagulant and enhance particle collisions appropriately. The aluminum sulfate flow rate entering the seawater will be 210 g/s.

3.1.2. Strainer

Strainers are used in the water treatment industry to provide very efficient filtering. Strainers can remove pollutants of varying sizes, ranging from microscopic organic particles to massive debris. Basket, self-cleaning, and temporary strainers are most commonly used for water treatment. Furthermore, self-cleaning strainers can successfully filter water with varying pollution levels, from freshwater to saltwater. These strainers also give a high level of filtration for greater viscosity liquids. Additionally, self-cleaning strainers are frequently used as pre-filters, eliminating particles from the water before they reach downstream equipment, where blockages or damage may occur (Ligon, 2021). For this pretreatment design, a self-cleaning filter is required to remove the colloidal particles generated by adding aluminum sulfate to saltwater.

The chosen vessel, Eaton's High Flow MCS-1500 Mechanically Cleaned Strainer, is utilized in this process due to its use in American and European desalination plants (Eaton, 2020). This strainer is made from 316 Stainless Steel and has a height of 2.6 m. Its max operating temperature and range of operating pressures are 82° C and 2-10.5 bar, respectively. Because the strainer's range of standard retention particle sizes is 0.15-1.1 mm, it will strain the large colloidal particles allowing for cleaner seawater to enter further filtration. When the Eaton's High Flow MCS-1500 Mechanically Cleaned Strainer is operating, the water rises toward the filter, preventing suspended particles from flowing through. These particles are then pushed downward and collected. When pressure lowers to a preset level, pollutants are immediately removed through the flush valve. No shutdowns are necessary since a backwashing mechanism is used to continuously clean the filter and remove dirt. Because a single Eaton strainer has a recovery rate of 98%, we require 33 strainers to filter 196 MLD of seawater. Of the 196 MLD seawater, 3.9 MLD will be removed through a waste stream. Additionally, the strainer will be operated at a pressure of 5 bar, with a final seawater exit flow rate of 192.1 MLD.

3.1.3. Ultrafiltration

Strained seawater enters an ultrafiltration unit which acts similarly to an RO membrane system where water placed at high pressures is pushed through a membrane to separate its impurities. Ultrafiltration (UF) modules were selected with the model name SFP-2660 from DOWTM which removes bacteria, viruses, and parasites from the fluid. These specific modules contain thousands of high strength, hollow fiber membranes for contaminant removal as well as colloids to protect downstream processes such as the awaiting RO system (DOW Chemical Company, 2022). This UF module will have a membrane pore diameter of 0.03 µm with a diameter, length, and width of 16.5, 186, and 25 cm respectively. The total module area is 33 m² and contains a single element recovery rate of 95%. This indicates that each module will recover almost all the water introduced into the system, with only 5% waste remaining. A feed seawater rate of 192.1 MLD will enter the UF process with a permeate flow of 182.5 MLD and a waste stream of 9.6 MLD. In order to obtain our desired 150 MLD target potable water goal, the design system will contain 196 UF modules, each at a pressure of 6.25 bar and a temperature of 25° C. These modules can withstand a pH range of 2-11, therefore the entering water will be well within this range at a pH of roughly 8.2. Although backflushing for these UF systems is necessary, the amount of water needed to complete this process is 1114.4 L per day which amounts to almost negligible numbers when compared to the larger process.

3.1.4. Nanofiltration

Proceeding after ultrafiltration, nanofiltration (NF) is used to remove any remaining viruses, most organic matter, and all salts present in the water with the exception of NaCl. In the wastewater treatment industry, microscopic dissolved pollutants and colloidal particles are routinely removed with NF membranes. Our NF membrane will be purchased from FilmTec

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Fortilife with the ID name: XC-N HP (Dupont, 2020). This is a spiral-wound element with a poly(piperazine) thin-film composite membrane inside. The pore diameter for this NF membrane is 1 nm and produces a flow rate of 31 m³/day per element. The physical dimensions for this membrane are a diameter and length of 20.1 cm and 102 cm respectively. The total membrane area per filter for the NF system is 37 m² and contains a single element recovery rate of 15%. Due to such a low individual recovery, there are 6 NF filter membranes connected in series within a single stack; a similar process to the RO pressure vessel system. With each stack containing 6 membranes, these stacks will have an individual recovery rate of 90%. A total of 5,145 NF membranes and 851 membrane stacks are needed to process the incoming water flow rate and volume. The NF system conditions will include a system temperature of 25° C, a necessary pressure of 35 bar, and an available pH range of 3-10 which falls well within the Bay of Bengal water quality pH level. For this NF system, an incoming feed stream flow rate of 182.5 MLD will enter, with a permeate flow of 164.2 MLD and a waste stream of 18.2 MLD at a pH of 8.2.

The total flow rates entering and exiting the 3 different stages in the pretreatment process are shown below in **Table 3.1**, along with the back flushing process flow rates.

Pretreatment Stages	Feed Flow Rate (MLD)	Permeate (MLD)	Waste Stream (MLD)
Strainer	196	192.1	3.92
Ultrafiltration	192.1	182.2	9.6
Nanofiltration	182.2	164.2	18.2

 Table 3.1: Inlet and outlet flow rates for pretreatment

3.2. Reverse Osmosis System

Reverse osmosis (RO) is the main process in the desalination plant in which salt and other impurities are removed from the Bay of Bengal's sea water. A total of 190 millions liters of water per day (MLD) will enter the RO system (164 MLD from pretreatment, 26 MLD recycled from electrodialysis). A process flow diagram (PFD) for the RO process is shown below in **Figure 3.2**.



Figure 3.2: Reverse Osmosis Process PFD

The stream entering the RO system is 190 MLD of treated seawater. The two stages achieve a 56% recovery and so exiting the first stage is a 106.4 MLD permeate stream and a 83.6 MLD retentate stream. In the second stage, the 83.6 MLD retentate enters as the inlet stream and the two exiting streams are 46.8 MLD of permeate and 36.8 MLD of retentate. The total permeate flow adds up to 153.2 MLD, but when maintenance in the form of backflushing is

accounted for in the RO and pretreatment systems, 149.6 MLD of permeate is actually exiting the system.

In this RO system, the membranes are organized in a two-part tapered array with the first stage containing 2975 RO membranes within 425 pressure vessels, and the second stage containing 2156 RO membranes within 308 pressure vessels. The tapered array was chosen as opposed to a series or parallel array because this structure is more cost effective. The series array would require all of the RO membranes to be in a single line which would take up more space and require significantly more membranes to process the water. The parallel array would mean having the same number of pressure vessels in stage one and two. The issue with that setup is that there is less water going into the second stage and so less membranes are required to process it, making it more costly to have a parallel array.

In order to make sure a flux occurs across the membrane, the membranes are housed in pressure vessels operating at 55 barg for the first stage and 80 barg for the second stage. These pressures were determined based on the osmotic pressure (OP) of the salt water exiting each stage which was calculated using **Equation 3.1**, where *C* is the molar concentration, *i* is van't Hoff's index with a value of 2, *R* is the ideal gas constant, *T* is the temperature of the water, and Π is the OP:

Inlet Sea Water: C = 0.56 mol/L $\Pi = 2 \ge 0.56 \ge 0.08206 \ge 298 = 27.4 \text{ atm} = 27.8 \text{ bar}$ Retentate Stage 1: C = 0.88 mol/L $\Pi = 2 \ge 0.88 \ge 0.08206 \ge 298 = 43.1 \text{ atm} = 43.7 \text{ bar}$ Retentate Stage 2: C = 1.37 mol/L $\Pi = 2 \ge 1.37 \ge 0.08206 \ge 298 = 67 \text{ atm} = 67.8 \text{ bar}$

Π

There must be a positive difference between the pressure vessel pressure and the OP exiting the system so that a flux occurs. The stream entering the RO system has a salt concentration of 33 g/L with an OP of 27.8 bar. The OP leaving the first stage is 43.7 bar, so the pressure vessel pressure of 55 bar provides a differential pressure of 11.3 bar, and will be enough to create the flux. The retentate leaving the second stage has an OP of 67.8 bar, so the pressure vessel pressure of 80 bar provides a positive differential pressure of 12.2 bar, and will also be enough to create a flux. In addition, there are two pumps placed before each RO stage that push the saltwater into the RO membranes. The hydraulic power for the first pump was calculated to be 3.1 MW however the second pump requires only 2.6 MW due to a reduced inlet volume in stream 3.

The specific RO membrane elements used in the first stage of this system are bought from the company Lenntech and have the ID name: FilmTecTM SW30XLE-440i (Dupont, 2019). The other seawater RO membrane options have lower exiting permeate flow rates which would mean the system would need more membranes and this would increase the cost and production time. As for the second stage, they will also be bought from Lenntech and have the ID name: FilmTecTM SW30HR-320 (DOW, 2022). The second stage membranes are high fouling which means they account for the increased salt concentration entering that stage and will allow for less frequent replacement of the membranes. Each membrane is spiral wound with the bulk of the area being the membrane where water will pass through. The specifications for each membrane are shown below in **Table 3.2** and **Table 3.3**. A diagram of a spiral wound membrane is also depicted in **Figure 3.3** with the membrane dimensions which are the same for both types. Overall, the entire RO process will require 5131 RO elements and 733 pressure vessels.

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Membrane Specifications	
Outer Diameter	7.9 in
Inner Diameter	1.125 in
Length	40 in
Single Element Recovery Rate	8%
Exiting Flow Rate	37,400 L/day
Active Area	440 ft ²

 Table 3.2: Specifications for the FilmTecTM SW30XLE-440i membrane

Table 3.3: Specifications for the FilmTecTM SW30HR-320 membrane

Membrane Specifications		
Outer Diameter	7.9 in	
Inner Diameter	1.125 in	
Length	40 in	
Single Element Recovery Rate	8%	
Exiting Flow Rate	23,000 L/day	
Active Area	320 ft ²	



Figure 3.3: Diagram of RO Membrane Dimensions

As mentioned for this process, there will be seven membranes placed into a pressure vessel. The number of membranes per pressure vessel was determined through literature which said the most common configuration had between 6 and 8 membranes in series per vessel (Fritzmann, 2007). A diagram of this setup is shown in **Figure 3.4** along with the pressure vessel specifications given in **Table 3.4**. The pressure vessels are bought from LennTech and there are two different types of vessels to account for the different pressures required for the two stages. The first stage will use the 8-E-1000-1M-to-8M-R6 pressure vessels (LennTech, 2017), which have a maximum operating pressure of 69 bar, while the second stage will use 8-E-1200-1M-to-8M-R6 pressure vessels (LennTech, 2017), which have a maximum operating pressure vessels (LennTech, 2017), which have a maximum operating pressure vessels (LennTech, 2017), which have a maximum operating pressure vessels (LennTech, 2017), which have a maximum operating pressure vessels (LennTech, 2017), which have a maximum operating pressure vessels (LennTech, 2017), which have a maximum operating pressure vessels (LennTech, 2017), which have a maximum operating pressure vessels (LennTech, 2017), which have a maximum operating pressure vessels (LennTech, 2017), which have a maximum operating pressure vessels (LennTech, 2017), which have a maximum operating pressure vessels (LennTech, 2017), which have a maximum operating pressure vessels (LennTech, 2017), which have a maximum operating





Figure 3.4: Pressure vessel depiction with dimensions

Volume per pressure vessel	2.1 m ³
Total volume of first stage	884.0 m ³
Total volume of second stage	640.6 m ³
Mass transfer area	50.3 in ²

 Table 3.4: Pressure Vessel Specifications

In regards to the inlet material conditions, the feed water manufacturer specifications for the FilmTecTM SW30XLE membrane suggests a pH level of 8, a membrane pressure of at least 55 bar, and a temperature of 25° C (Lenntech, 2019). The feed water used for this process is collected from the Bay of Bengal, which has an average temperature range of $23-28^{\circ}$ C year round (Balakrishna, n.d.). Taking the average sea water temperature of 25° C, an OP of 27.8 bar was calculated for the entering sea water as discussed earlier. It should be mentioned that a maximum OP of 28.1 bar was calculated to ensure that the RO design process would function with a higher OP value, however the calculations for the system used the average.

In order to conserve some amount of the energy leaving RO in the brine stream, pressure exchangers are used to transfer energy from the brine stream to the RO inlet stream to reduce the pumping power required to push the water through RO. The pressure exchangers chosen were the PX Q300 and will be bought from Energy Recovery® (PX Pressure Exchanger - Energy Recovery, 2022). The whole system will consist of 23 pressure exchangers in order to run the entire brine stream exiting RO through and to recover the maximum amount of power. The entire brine stream has a flow rate of 36.78 MLD and has a power of 151 kW. That power will be transferred to 66.24 MLD of the feed stream. The pressure exchangers recover 60% of the power from the brine stream, so 90.8 kW will be added to the portion of the RO inlet stream passing through, which will increase its pressure by 26.6 bar. The feed stream is at 29 bar coming out of nanofiltration, and the pressure exchanger will bring the pressure up to 55.6 bar. A small pump will be placed after the pressure exchanger to bring the pressure up to 56.5 bar, so that the pressure matches the pressure of the streams from electrodialysis and the stream being pumped directly after nanofiltration. These pressure exchangers will be used to pressurize the 66.24 MLD of water instead of pumps being powered by electricity. This saves costs on both equipment and electricity.

3.3. Permeate Post-Treatment

3.3.1. Remineralization

Remineralization is a post-treatment process used to add $Ca(HCO_3)_2$ to the filtered water in order to help with decrease corrosivity and aggressiveness of the water. This is done using lime or limestone in various forms, and mixing it with carbon dioxide or an acid to form the $CaCO_3$ in solution. We are using hydrated or slaked lime ($Ca(OH)_2$), which will be referred to as lime and comes in a powdered form, for our calcium source. Carbon dioxide will be used as our acid to make carbonic acid to fulfill the reaction below (Voutchkov, 2011).

$$2 \text{ CO}_2 + \text{Ca}(\text{OH})_2 \rightarrow \text{Ca}(\text{HCO}_3)_2$$

Since the maximum allowed hardness for water is 200 mg/L, and the recommended hardness for drinking water is between 80-120 mg/L, we will be aiming for a hardness of 100 mg/L, which is based on the concentration of $Ca(HCO_3)_2$ (Withers, 2005). The required dosage to reach the desired hardness is 77.36 mg/L of 95% purity lime and 87.32 mg/L carbon dioxide (Nelson & De Luca, 2021).



Figure 3.5: PFD for Lime Addition

The lime is added to the water in three stages as seen in **Figure 3.5**. The lime is added to 0.2 MLD of water at a rate of 11.6 tonnes a day to create a lime slurry. The concentration of the lime slurry at this point should be around 75 g/L. This concentration was chosen because at 100 g/L, there is a risk of the lime depositing and creating blockages, and below 50 g/L, there is a chance of carbonation (*Lime, reagent in water treatment, 2022*). The storage silo for the lime is designed to hold enough lime to last 60 days, and will have a volume of 320 m³, with a height of 12 m and a diameter of 6 m. The lime slurry tank dimensions have not been finalized, but it will be designed to have a retention time of 4 hours, resulting in a mixer volume of 25 m³. The basis for the lime slurry mixer and lime storage silo can be found in the appendix in **Table A.1**.



Figure 3.6: Diagram of Lime Saturator (*Lime, reagent in water treatment, 2022*)

The lime saturator is the second stage of lime addition where the lime slurry is diluted into limewater as shown in **Figure 3.6**. By adding 10.8 MLD of water to the lime slurry, we

create limewater that has a concentration of 1000 mg/L of Ca(OH)₂. This step is important for removing the impurities introduced by the lime. It is very hard and expensive to get pure slaked lime, so we are buying 95% purity lime, so we must remove as much of the 5% impurities as possible to keep the turbidity of the final product down. The lime saturator is a large mixing tank that consists of two major zones; the reaction zone and the settling zone. The reaction zone is a small zone in the center of the saturator that is 25-35% the diameter of the saturator (Voutchkov, 2011). This is where the permeate and lime slurry is fed into is being mixed with an impeller. It is marked by walls that have slits going down the cylinder in order to allow the mixed limewater to leave the reaction zone and enter the settling zone. In the settling zone, there are rakes moving more slowly, which will collect all the settled impurities at the bottom and force them out through the bottom as waste. The settling zone allows for these impurities to settle naturally. With lower purities, it is necessary to add flocculants to speed up the settling of impurities, but we believe that we have a high enough purity that we do not require flocculants (Withers, 2005). We produce about 12.89 mg of waste per liter of water assuming we remove all the impurities and only impurities, and the waste stream is 30% solid and 70% water, and this is about 142 kg of waste per day (Nelson & De Luca, 2021). This is an incredibly small amount of waste compared to the amount of water we produced and that leaves the tanks, that it leaves a negligible effect on our final flow rate. The lime impurities will be mixed with the rest of our waste streams, which is about 30 MLD of waste water, and discharged back into the Bay of Bengal. Due to the insignificant amount of lime versus the large amount of waste water, the pH effect should be negligible.

In order to ensure that no impurities leave the saturator in the final limewater, the lime water leaves through an overflow pipe close to the top of the tank, so that we reduce the risk of

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any settling material from entering that pipe as seen in **Figure 3.6**. The design of our lime saturator is based on a retention time of 100 minutes and results in using two tanks that both have a height of 5 m and diameter of 12 m. The liquid level will only reach 4 m as it is recommended that there is 1 m of clearance above the water level. The reaction zone will have a diameter of 3 m and height of 3 m. The design basis of the lime saturator is seen in **Table A.2**. The third stage is injecting the limewater into the mainstream of water that was not diverted to the remineralization unit, which would be about 139 MLD, to create a 150 MLD flow of remineralized water that has a hardness of 100 mg/L. This can be done with an inline mixer or just an injection stream.

3.3.2. Recarbonation

Carbon dioxide is added to the water to create carbonic acid, which will react with the dissolved lime to create $Ca(HCO_3)_2$, which hardens the water. Since we require 87.32 mg/L to achieve the designed hardness, we need a flow rate of 13.1 tons per day to be added to the water. Some processes do the carbon dioxide addition in parallel, meaning water is diverted to remineralization and a separate stream is diverted to recarbonation at the same time and injected into the rest of the permeate sequentially (Withers, 2005). Other systems will recarbonate the permeate water right before or after the lime is added to the permeate. The carbon dioxide will be stored either as liquid or gas. If stored as a liquid, it will require refrigeration for storage and a vaporizer for when it is added to the water (Voutchkov, 2011). The methods we explored for injecting carbon dioxide into the water stream include two TOMCO₂ Systems: direct CO_2 gas injection and pressurized solution feed (PSF). Their direct CO_2 gas injection involves adding the CO_2 directly into the water and using static mixers to dissolve the gas into the water. This creates a two-phase flow since not all the gas will dissolve and requires degasser towers in order to

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remove the excess gas from the water flow (*TomCO*₂ *Systems*, 2022). The direct gas injection has an estimated efficiency of 55% based on their reports. Their PSF involves splitting the water stream to create a pressurized water stream. Then, pressurized CO_2 gas will be added to the pressurized water stream and be mixed using two parallel streams of static mixers. The high pressure of the water and gas will help dissolve and react the gas into the water creating carbonic acid and this can be added directly into the main water stream. This has an estimated efficiency of 95%, and since the final mixing is a liquid-liquid interface, the mixing is fast and requires minimal extra equipment. Due to the simplicity of the PSF and the removal of process units like the degasser tower, this process will use the PSF for the carbon dioxide injection. (*TomCO*₂ *Systems*, 2022). Moreover, we will have an additional PSF to serve as a redundancy and a pump will be designed to pressurize the carrier water to the required pressure.



Figure 3.7: Picture of the Pressurized Feed System (*TomCO*₂ *Systems*, 2022)

3.3.3. Chlorination

The process of adding chlorine to drinking water to eliminate parasites, germs, and viruses is known as chlorination. Different methods are used to achieve safe chlorine levels in

drinking water. Common methods include adding chlorine (liquid or gas) either in-line or in a tank (*Chlorination*, 2022 & *Water treatment*, 2022). Because chlorine is a hazardous chemical, the United States and India have set limits on the amount of chlorine in the water to prevent health issues associated with ingesting large amounts of chlorine. The United States and India limits are 4 mg/L and 250 mg/L, respectively (Samer, 2015 & CPCB, 2022). Moreover, the wastewater treatment industry standard is to add 2 to 3 mg/L during post-treatment. We have chosen to add 2.5 mg/L. Based on the 149.9 MLD of water leaving recarbonation, our Cl₂ flow rate is 375 kg/day or 260 g/hr. Our Cl₂ gas flow rate is low enough that we can make use of a compact chlorometer as seen in Figure 3.8, which can usually handle up to a flow rate of 4 kg/hr (*Gas reagents...*, 2022). The solubility of Cl₂ in water at 25 C is at most 6.30 g/L, and due to our low target concentration and flow rate of Cl₂, there is minimal risk of two-phase flow forming with any significant amount of water being diverted to the chlorometer (Whitney & Vivian, 1941). Additionally, we will have an additional Compact Chlorometer to serve as a redundancy.





3.4. Brine Post-Treatment

In order to reduce the negative environmental impact of this desalination plant, the concentrated brine waste stream leaving RO will be further concentrated and processed into crystallized salt rather than being dumped back into the Bay of Bengal and potentially harming marine life. The post-treatment brine process is two-fold including an electrodialysis system and crystallizer shown below in **Figure 3.8**. Electrodialysis will further concentrate the brine stream and have a diluate recycle stream brought back to the RO unit. The further concentrated brine will then be brought to a crystallizer where the water will boil off and crystallized salt will be left behind as a coproduct to water for the plant.



Figure 3.8: Brine Post-Treatment PFD

3.4.1. Electrodialysis

Electrodialysis is a membrane process in which ions, rather than water, are forced through a membrane by the direct current connected to an anode and cathode on either side of the system. In this specific system, Na⁺ and Cl⁻ ions will pass through the membrane to create a concentrated salt stream and a dilute stream with significantly less ions. Each membrane has a

pore size to allow either Na+ or Cl- to pass through, so the membranes are referred to as membrane pairs because there is an equal number of both membranes in an electrodialysis system. The Na+ ions are attracted by the negative charge of the anode and rejected by the positive charge of the cathode which allows for the charge to pass through the membrane. It is the same process for the Cl- charges, but with it being attracted to the positive charge of the cathode and rejected by the negative charge of the anode. This process is displayed below in **Figure 3.9**.



Figure 3.9: Electrodialysis System (Tedesco, 2016)

A 36.8 MLD brine stream will enter electrodialysis from the RO system and will have a concentration of 80 g/L. In preparation for the crystallizer, the concentration of the concentrate stream needs to be brought to 200 g/L. Similarly, the diluate stream must be returned to the concentration of the sea water entering RO, which is 33 g/L so that it can be added back into the RO inlet stream. Following **Equations 8.6 - 8.8** in the Appendix, the volumetric flow rates of the streams exiting the process were found to be 10.5 MLD of concentrate and 26.8 MLD of diluate.

The specifications for the ED system are shown below in **Table 3.5**. The membranes will be organized in a single stack to process all of the water. The exact configuration of the membranes within the ED stack is out of scope for this project and the ED system is generalized to be easier to work with. The stack voltage is taken from McGovern's study on the cost of using electrodialysis on high salinity feeds (2014).

ED System Specifications	
Inlet Flow per Unit	37 MLD
Total Number of Membrane Pairs	435
Effective Area per Membrane	0.5 m ²
Current Density	300 A/m ²
Total Current	65,250 A
Stack Voltage	6 V
Power	391.5 kW

 Table 3.5: Specifications for the ED System

3.4.2. Crystallizer

A crystallizer is a system in which salts dissolved in water become crystallized solids. The 10.5 MLD concentrated brine stream leaving electrodialysis will be brought to the crystallizer where the salt is fully separated from the water and crystallized. The unit is designed as a heated moving bed system that will allow the water to be boiled off leaving pure NaCl behind. There were two options to heat the system: heated air or coils on the moving bed. We chose the heated air because coils on the heated bed are outside of our scope and heated air was suggested by Professor Giri. There are vents above the unit to allow for the water-saturated air to exit the system. The solid salt will be collected at the end of the conveyor belt and sent off for distribution.

An Aspen Plus model was used to determine the heat duty required to evaporate all of the water out of the 10.5 MLD brine stream. The simulation used the electrolyte wizard and flash tube block. The feed stream in the simulation ran at 25 °C and 1.97 atm, and the flash tube ran at 1 atm and 109 °C. After running the simulation, 314 MW was the minimum heat duty required to evaporate all of the water. The energy will be provided to the system using low-pressure steam to heat up the salt water and boil it off. Low-pressure steam is considerably cheaper than electricity for providing heat. Since there is some inefficiency in the heating process, we assumed the heat transfer would be 90% efficient, so we require 349 MW to evaporate the water.

3.5. Production and Maintenance

3.5.1. Overall Process

The overall production schedule for this desalination plant will run for 8,322 hours per year to make a 95% uptime rate. Through literature reviews, it was found that previous desalination plants around the world operate on this schedule, since desalination processes should be a continuous process. In order to keep this high operation rate and allow for cleaning and maintenance on the equipment, multiple redundancies will be incorporated into the system so that the entire plant will never fully shut down.

3.5.2. Pretreatment

Since pretreatment involves removing physical and chemical impurities from seawater, fouling can easily occur if regular maintenance is not performed. Backflushing processes must be done on both ultrafiltration and nanofiltration, with the exception of the self-cleaning strainers, as they will regularly maintain themselves without further required cleaning. For both processes, backflushing should be performed at least once a week, indicating that extra filtration membranes should be in place to filter the continuous seawater as specific filters are being cleaned. The amount of active modules necessary for ultrafiltration is 182, and therefore there will be 14 extra modules so there will be 14 unoperational modules at any given time while 182 are active. In total, 196 modules will be required, and need a water volume of 39.8 L per module to fully backflush the equipment. The backflushing process will last 2 minutes per singular module (and therefore 28 minutes to flush 14 modules concurrently), therefore 28 modules will be fully flushed each day, amounting to a complete ultrafiltration backflush cycle once a week. The total amount of water needed for the backflushing process in ultrafiltration is 1114.4 L per day, an almost insignificant amount compared to the rest of the seawater processes.

In regards to nanofiltration, 5,100 active NF membranes are needed in order to process the 196 inlet flow into the plant. It was determined that 7 different units, each containing 105 NF membranes will be backflushed per day, resulting in a total backflushing cycle for the NF stage of 1 week. Each membrane will be backflushed for 2 minutes, and therefore, to clean a unit of 105 membranes, it takes 210 minutes. A total of 5,145 NF membranes will need to be purchased to take into account the backflushing and rundundencies needed for this process, and each membrane will need to be flushed with 194 liters of seawater to efficiently reduce the possibility and frequency of fouling. For nanofiltration, 0.143 MLD of seawater will be needed for backflushing in a final backflushing seawater amount of 0.144 MLD.

3.5.3. Reverse Osmosis

In terms of the RO system, RO membranes need to be backflushed and cleaned multiple times a day so that significant fouling does not occur. Backflushing should be done every 30 to 90 minutes, which means each stage needs a certain number of extra membranes and pressure vessels so that there are enough membranes operating to process the amount of water required. Each pressure vessel houses 7 membranes, and so backflushing was calculated based on the number of pressure vessels rather than solely on the number of membranes. In the first stage, there are 17 extra pressure vessels so that 17 of them can be inoperable at any given time and 408 pressure vessels are always active. Similarly, the second stage has 14 extra pressure vessels so that 14 pressure vessels can be backflushed while 294 pressure vessels are still active. An automatic system will allow for these backflushing processes to occur. The amount of water required to fully backflush one pressure vessel is equal to the volume of the pressure vessel, so since the volume of one pressure vessel is 231 L, that is the volume of permeate water required

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to backflush it.

Each pressure vessel takes roughly 3 minutes to backflush. If the first stage is broken into 17 units with 25 pressure vessels per unit, one backflushing cycle to clean the 25 membranes takes 75 minutes with a total of 19.2 backflushing cycles per unit per day. As for the second stage, there are 14 units with 22 pressure vessels per unit, where one backflushing cycle to clean all 22 pressure vessels takes 66 minutes with a total of 21.8 backflushing cycles per unit per day. The amount of permeate water used to backflush the first stage is 1.88 MLD, and 1.55 MLD for the second stage. The total amount of permeate lost to backflushing is 3.43 MLD. Lastly, RO membranes should be fully replaced every 2 years.

3.5.4. Permeate Post-Treatment

Lime saturators and recarbonation systems should be cleaned every 6-12 months while the chlorination tanks should be cleaned every 6 months. The lime slurry tubes will include a valve that will allow it to be backflushed occasionally to prevent any major deposits from forming.

3.5.5. Brine Post-Treatment

In the electrodialysis unit, maintenance is performed by switching the electrode polarity and reversing the salt flux direction. Valves are used to switch which streams flow into the crystallizer and recycle streams automatically when reversing the polarity, allowing for the process to be done automatically and continuously. The crystallizer system will be cleaned once a week to prevent buildup of salt on the moving bed.

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3.6. Economics

3.6.1. Material Expenses

Process Unit	Materials	Amount Required (Mass/Day)	Expense
Pretreatment	Aluminum Sulfate, Al ₂ (SO ₄) ₃	21 tons	\$4,494.00
	Carbon Dioxide, CO ₂	13.1 tons	\$262.00
Permeate Post-Treatment	Chlorine, Cl ₂	375 kg	\$1,455.00
	Lime, Ca(OH) ₂	11.6 tons	\$1,264.40
Total Daily Expense		\$7,475.40	
Total Annual Expense			\$2,728,521.00

 Table 3.6: Material Expenses

The material expenses for the plant are exclusively found in the pretreatment and post-treatment sections where chemicals are added to the water to either remove unwanted particles in pretreatment, or to remineralize and chlorinate water in permeate post-treatment. The price per ton or kg were found online from manufacturers and the total expense was calculated based on those values. These expenses can be found above in **Table 3.6**, and the total annual expense for materials adds up to about \$2.7MM.

3.6.2. Utility Expenses

Process Unit	Annual Utility Expense
Electricity (Pumps)	\$12,022,578.09
Electricity (Electrodialysis)	\$243,229.17
Steam (Crystallizer)	\$51,936,339.11
Total Annual Expense	\$66,930,667.38
Total Annual Expense without Crystallizer	\$14,994,328.27

 Table 3.7: Annual Utility Expenses

The main utilities used in this plant are electricity and steam as seen in **Table 3.7**. For the cost of electricity, we used local prices in India and found a price of ₹5.75 per kWh, which is about \$0.08 per kWh. From Turton and their capital cost macro, we were able to get the amount of energy required to run our pumps in the system, and used an efficiency factor of 86% to get the amount of electricity that would be supplied to the pumps (Turton et al, 2009). The pumping costs here take into account the energy saved from the pressure exchangers used for RO. The electrodialysis unit runs on electricity to create the driving force for ions to go across the membrane, and since the energy is going directly into cathodes, we assumed 100% efficiency for simplicity, and got our total energy requirements based off the voltage and current the electrodialysis would be operating at. The crystallizer only requires heat, so the crystallizer could be run on steam or electricity. However, the cost of electricity is much greater than the cost of steam for generating heat. From Turton, we were able to get a price for low-pressure steam at \$4.54 per GJ, and we simply calculated how many GJ per year would be necessary to evaporate our water. This gave us the total cost for steam. Steam is the majority of our utility expenses, so

we added a scenario where solar evaporation, rather than steam, would be utilized in order to estimate the plant's worth if salt could still be produced without spending money to actively heat it, and therefore determine if that scenario would make this plant valuable.

3.6.3. Operating Costs

	Operating Costs
Materials	\$2,728,521.00
Utilities	With Crystallizer: \$59,008,512.47 Without Crystallizer: \$12,265,807.27
Labor	\$1,128,400.00
Total Operating Costs with Crystallizer	\$62,865,433.00
Total Operating Costs without Crystallizer	\$16,122,728.00

 Table 3.8: Annual Operating Expenses

The annual operating expenses for the plant are a combination of the materials, utilities, and labor that will be necessary on a yearly basis, and these values are shown above in **Table 3.8**. The materials and utility cost values were taken from **Table 3.6** and **Table 3.7**, respectively. The labor costs were calculated by finding the total number of operators necessary for a single shift using **Equation 3.2**, and then multiplying that value by 4 to account for three shift cycles per day, as well as, holidays, sick days, and vacation days. The number of operators was then multiplied by ₹350,000 (\$4550), the average salary in Chennai for a desalination plant operator in order to determine the total labor cost (*Chennai Water Desalination RO Plant Operator Salaries in India*, 2022).

$$N_{OL} = (6.29 + 31.7P^2 + 0.23N_{np})^{0.5}$$
 (Equation 3.2)

In the above 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. The number of operators for one shift, calculated based on **Equation 3.2** is 62 operators, which would mean a total of 248 operators to cover all four shifts and result in a labor cost of about \$1MM.

Additionally, the utility operating costs include two scenarios: using a crystallizer and not using a crystallizer. Removing the crystallizer reduces the annual operating costs by about \$46MM.

3.6.4. Equipment and Capital Costs

Process Unit	Equipment	Expense
Pretreatment	Strainer, Membranes, Tanks	\$119,838,495.00
Reverse Osmosis	Membranes, Pressure Vessels, Pressure Exchangers	\$65,000,000.00
Permeate Post-Treatment	Lime Addition, Carbonation, Chlorometer, Tanks	\$3,652,493.26
Brine Post-Treatment	Membranes, Furnace	\$1,878,540.00
Pumps	Pumps	\$10,585,940.00
Total Equipment Expense		\$200,955,468.26

Table 3.9:	Equipment	Expenses
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In order to build the plant, the equipment for each unit must be purchased and installed. The breakdown of the expense for each process unit is shown above in **Table 3.9**. Due to the large number of membranes and multiple expensive tanks, the pretreatment process has the most expensive equipment. Additionally, the pump equipment costs are placed as a separate line item in the table due to how specific pump costing is. There are a total of 12 pumps located throughout the plant, requiring different construction materials depending on what comes in contact with seawater, and in total costing about \$10.6MM.

Capital Costs	Expense
Land	\$509,652.00
Equipment	\$200,955,468.26
Plumbing, Insulation, Wiring, etc.	Lang Factor: 4.74
Total Initial Capital Investment	\$954,944,670.03

 Table 3.10: Capital Costs

The capital costs for the plant include purchasing the land to build on, the equipment for each process unit, and other expenses such as plumbing and insulation and the breakdown is shown above in **Table 3.10**. A total of 6 acres will be required to build this plant, which was determined based on the amount of land the Carlsbad, California desalination plant takes up (*Claude "Bud" Lewis Carlsbad Desalination Plant*, n.d.). The price per square foot was determined based on an Indian agricultural real estate site which listed an average of ₹150 per square foot for agricultural land in Chennai, India (*Agriculture Land for Sale in Chennai South - Farm Land for Sale in Chennai South*, 2022). These values resulted in a total cost for land of about \$500,000.

3.6.5. Revenue

Product	Daily Amount Produced	Sale Unit	Daily Revenue	Yearly Revenue
Water	150 MLD	₹45.00/kL	₹6,750,000.00	₹2,463,750,000.00
Food Grade Salt	2315 tons	₹5,000.00/ton	₹11,576,626.24	₹4,225,468,577.73
Total Revenue (₹)				₹6,689,218,577.73
Total Revenue (\$)				\$86,959,841.51

Table 3.11: Total Revenue (Food Grade Salt)

 Table 3.12: Total Revenue (Road Salt)

Product	Daily Amount Produced	Sale Unit	Daily Revenue	Yearly Revenue
Water	150 MLD	₹45.00/kL	₹6,750,000.00	₹2,463,750,000.00
Road Salt	2315 tons	₹2,000.00/ton	₹4,630,650.50	₹1,690,187,431.09
Total Revenue (₹4,153,937,431.09			
Total Revenue (\$)				\$54,001,186.60

The total revenue of the plant is based on the two products produced: potable water and salt, and these values and total revenue can be found above in **Table 3.11** and **Table 3.12**. The price per kL of desalinated water in India is ₹45 (Lahiry, 2017). Depending on the quality of the salt produced, the price can vary fairly significantly. For instance, the difference between food grade salt and road salt is ₹3,000/ton which results in a yearly revenue difference of about \$33MM. Based on this difference, food-grade salt would be the most profitable option and will be the product produced through brine post-treatment.

3.6.6 Profitability Analysis

Scenario	Yearly Net Cash Flow
Crystallizer Utility w/ Road Salt Prices	-\$14,057,881
Crystallizer Utility w/ Food Salt Prices	\$18,900,774
No Crystallizer Utility w/ Road Salt Prices	\$37,878,458
No Crystallizer Utility w/ Food Salt Prices	\$70,837,113

 Table 3.13: Yearly Net Cash Flow

When determining the profitability of the plant, we first looked at the yearly net cash flow. This was done by subtracting the annual operating expenses from the revenue, and we found the net cash flow for each scenario as shown in **Table 3.13**. From this, we can see that we do make money year to year in all our scenarios except our worst one, where we use steam to evaporate the water and have to sell all the salt at road salt prices. We can also see that if we were to have to sell half our salt at food price and half at road price, we would barely break even on an annual basis.

Net Present Value over 20 years with a 5% interest rate				
Scenario	Discounted Cash Flow	Net Present Value		
Crystallizer Utility w/ Road Salt Prices	-\$175,192,267	-\$1,130,136,937		
Crystallizer Utility w/ Food Salt Prices	\$235,545,423	-\$719,399,247		
No Crystallizer Utility w/ Road Salt Prices	\$472,049,315	-\$482,895,355		
No Crystallizer Utility w/ Food Salt Prices	\$882,787,005	-\$72,157,665		

 Table 3.14: Discounted Cash Flow and Net Present Value Breakdown

In order to determine the profitability of the plant over its 20 year lifetime, a discounted cash flow analysis was performed on the scenarios which are shown above in **Table 3.14**. For this analysis, we assumed that we would be operating at full capacity starting year 1 once construction finished, and we would be operating at full capacity for all 20 years with the same amount of revenue and operating expenses. The worst case scenario involves keeping the crystallizer unit and producing road salt, which results in a negative profit over 20 years of \$1.1B. The best case scenario involves removing the crystallizer unit and producing food-grade salt, which still results in a negative profit, but a significantly smaller loss at about \$72MM. From these results, we can see that all our scenarios are not profitable over a 20 year span even though we do make money year to year. For normal business investments, this may not be worth it, however the best scenario is potentially worth investigating further since the margin for profitability is quite small comparatively. This venture would also be providing a service and public necessity to the people, so profitability is not necessarily the goal for this plant.

3.7. Safety, Health, Environment, and Social Impact

3.7.1. Physical Hazards

The main physical hazards that are present in this plant are high pressure water and low pressure steam. For multiple unit operations, water will be pumped to a very high pressure, which can pose a danger if a release occurs. A release of high pressure water could lead to severe injuries and even fatality if enough water is released and directly impacts a person. Concentrated salt water at high pressure can pose an even greater threat due to the higher density and corrosivity when impacting a person, as well as damaging equipment. In order to prevent a release from occurring, regular maintenance checks on the pipes, flanges, seal, and pumps will be required to check for corroded or damaged parts. The use of copper alloy and other non-ferrous materials will be used to mitigate the corrosive effects of salt water on equipment such as pumps and pipes. Drains should also be installed around the facilities in order to mitigate the effects of flooding in case of a leak. Drains would limit the spread of the water leaking and mitigate any flooding that would occur. Other mitigating measures may include release valves into a drain or emergency tank to alleviate pressure in case of releases or a shut-off valve to stop flow to the damaged area completely. Low-pressure steam is only used to heat up the crystallizer, and is not present throughout the rest of the facility. So, the area in which steam is a hazard is limited. If steam is ever released, the heat can lead to serious burns and injuries. Mitigating measures that can be put in place would be either a shut off valve to stop the release of steam, or a relief valve into the atmosphere in case there is an unexpected build-up of pressure in the system.

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3.7.2. Chemical Hazards

The three chemical additives that are used in this plant are aluminum sulfate solution, carbon dioxide (CO_2), and chlorine (Cl_2). Aluminum sulfate solution will be stored in a tank in solution form. Under OSHA regulations, the solution is listed as category 1 for skin corrosion/irritation, serious eye damage/eye irritation, and corrosive to metals (*Aluminum Sulfate Solution*, 2020). If a loss of primary containment (LOPC) occurs, severe injury could be inflicted onto the workers and other personnel, so any personnel in the area of solution should be wearing sufficient PPE and understand the first aid procedures and responses if an LOPC occurs. Since the solution is only used for one process, the area that the solution is present is narrow. For safety measures, the material that will be used to contain the solution will be resistant to corrosion and regularly checked on for integrity. Mitigation measures such as a dike should be installed around the tank in order to contain the spread of the liquid, as well as a shut-off valve installed in case the release source is from a pipe.

 CO_2 is being stored under pressure to be injected for remineralization of the permeate water. CO_2 does not have any major chemical hazards, however a release of CO_2 could lead to widespread asphyxiation if not properly ventilated (*Carbon Dioxide*, 2022). In order to prevent a concentrated amount of CO_2 settling, the area the CO_2 is stored must be properly ventilated and not enclosed. The storage tank will most likely be outdoors, so this will not pose a major hazard. However, high level carbon dioxide sensors should still be installed, as well as emergency fans if deemed necessary in order to quickly increase ventilation in case of an LOPC. Since CO_2 will be stored under pressure, another hazard would be overpressure leading to a rupture or BLEVE. For this to take place, the storage tank would need to be heated for an extended period of time. The only flammable material that should be present in the facility is the chlorine, so the risk of this occurring is low. However, it may still be recommended to install high pressure and temperature alarms on the tank in order to prevent a BLEVE or rupture from the tank. To alleviate this pressure, a relief valve could be installed to release the pressure either into the atmosphere or into a separate storage tank.

Chlorine is used to disinfect the treated water, and is being kept under pressure as a liquid. Chlorine is exceptionally hazardous as it is extremely toxic with a low exposure limit and acts as an oxidizer, so it can worsen or create fires (Chlorine, 2022). An LOPC would result in widespread injury and potentially multiple fatalities, so having the required safeguards in place is essential for the safety of the workers and surrounding communities. For containing the chlorine tanks, the tanks can either be kept in an open enclosure, or a closed enclosure depending on the safeguards that are in place. If the tanks are kept in the open, a water curtain or neutralization solution should be implemented in order to minimize the spread of the chlorine gas and convert into solution. If the tanks are enclosed tightly by walls or another vessel, then a nitrogen purge, a flare, or a scrubber can be implemented to mitigate the spread of the chlorine. Other safeguards that must be considered are the potential requirements for containment ponds of the neutralization solution or water after they have been used in order to prevent ecological damage from the chlorine, the use of high flow check valves to stem any leaks from hoses, making sure that the procedures for switching or filling tanks is well-established and followed closely, and potentially placing the tanks underground in order to prevent above ground spread or damage. Small leaks can also lead to injury as well, so making sure chlorine detection alarms are placed around the tank and immediate area around the chlorine storage will be essential for the safety of those working around the disinfection unit. While there are other methods that can be used with less hazards, such as sodium hypochlorite, chlorine is the most common and efficient way to

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disinfect this volume of water (Bazargan, 2018). To prevent a massive leak and spread of chlorine, we are limiting our on-site inventory greatly, as well as spreading our chlorine storage among multiple tanks, so that even if one tank ruptures, the damage would be greatly reduced by the smaller volume of the tank as well as the lower inventory.

3.7.3. Environmental Concerns

If we are discharging brine waste into the Bay of Bengal. The environmental impacts are mostly caused by brine discharge, but also, to a lesser extent, by the chemicals used in the cleaning of various modules, and also thermal pollution. Because brine is denser than seawater, it tends to settle on the ocean floor, where it can harm susceptible marine species. It also contains less oxygen than seawater and may contain potentially harmful compounds added during the desalination process, such as copper or chlorine. Furthermore, because brine settles to the bottom, away from waves and currents, it may be difficult to disperse. Moreover, research shows that there are little to no changes in temperature and pH of the body of water when brine is disposed of. However, if this becomes a problem in future years, we will adhere to any regulations from local and federal governments in India.

3.7.4. Social Motivation and Impact

We were inspired to situate our facility in Chennai since India offers unparalleled chances for social investing due to the magnitude and possibility for long-term rewards, regardless of market uncertainties. Concerns that affect India include poverty, unemployment, gender inequity, and insufficient access to water. Only via such innovative solutions will India be able to continue its development trajectory and deliver inclusive growth. Millions of people will gain access to clean water as a result of the plant's construction in India. Additionally, we would recruit the locals allowing them to obtain and preserve transferable skills; their salary would ameliorate

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poverty. Moreover, we are committed to being an equal opportunity employer, pledging not discriminate against employees based on race, color, religion, sex, national origin, age, disability, or genetic information. We hope to champion women's rights, alleviating some of the gender inequality Indian women encounter. Our desalination plant will effectively bridge the enormous divide that exists between the prosperous and this underserved country.

4. Final Design



Figure 4.1: PFD of Desalination Plant

4.1. Pretreatment

Figure 4.2: PFD of Pretreatment

The entire pretreatment stage processes 196 MLD of raw seawater and distributes 164.1 MLD of treated seawater into the RO separation stage. The pretreatment portion of the desalination process contains 4 separate steps: the addition of coagulants, removal of sediment through strainers, ultrafiltration, and nanofiltration purification. A total of 20 tons per day of aluminum sulfate coagulant will be distributed into the incoming seawater, and once coagulation has occurred, 33 Eaton's High Flow MCS-1500 Mechanically Cleaned strainers will be implemented with a recovery rate of 98%. After the strainer has processed the water, 192.1 MLD enters the UF portion where 182 active UF membranes filter out particles larger than 0.03μ m. These UF membranes contain a single element recovery rate of 95% and will be purchased from DOWTM with the model name SP-2660. From this stage, 182.2 MLD of seawater will enter the nanofiltration unit, and 9.6 MLD will join the 3.92 MLD waste stream from the strainer and return to the Bay of Bengal.

During nanofiltration, impurities with a diameter larger than 1 nm are filtered out and returned to the Bay. These NF membranes will be purchased from Dupont with the model name

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XC-N HP. Due to each NF membrane containing a low recovery rate of 15%, 6 membranes will be connected within a single stack to achieve a single stack recovery rate of 90%. All membrane specifications are described in Section 3.1.4: Nanofiltration. The incoming seawater flow for this stage is 182.2 MLD with a waste stream of 18.2 MLD and a final pretreatment permeate of 164.1 MLD. Through both ultrafiltration and nanofiltration, a total of 0.144 MLD of seawater will be diverted to backflush both filtration unit stages regularly. This 0.144 MLD has already been omitted from the final pretreatment exit flow rate value of 164.1 MLD. **Table 3.1** depicts all the entering and exiting seawater flow rates for each stage in the pretreatment process.

4.2. Reverse Osmosis

Figure 4.3: PFD of Reverse Osmosis System

The reverse osmosis system processes 190 MLD of treated seawater, 164 MLD from pretreatment and 26 MLD from ED recycle stream, and consists of 5,131 membrane elements and 733 pressure vessels. The membranes are split into two stages, process unit RO101 in **Figure 4.3** processes all of the water, and process unit RO102 in **Figure 4.3** processes the retentate stream from the first stage. Both stages achieve 56% recovery of water. The first stage consists of 2975 FilmTecTM SW30XLE-440i membranes which have a permeate flow rate per membrane of 37,400 L/day. The second stage consists of 2156 high-fouling FilmTecTM SW30HR-320 membranes which have a permeate flow rate per membrane of 23,000 L/day and are built to handle the increased salt concentration going into the second stage. The membrane dimensions are shown in **Figure 3.3**, and the membrane specifications for the first and second stage membranes are shown in **Table 3.2** and **Table 3.3**, respectively. The permeate stream exits the system at 150 MLD and the retentate stream exits at 36.8 MLD.

The RO membranes are housed within pressure vessels, with 7 membranes per vessel.

These pressure vessels are necessary to ensure that the RO membranes remain at a specific pressure. The first stage pressure vessels operate at 55 barg and the second stage operates at 80 barg. Additionally, 8-E-1000-1M-to-8M-R6 is the pressure vessel we are using from LennTech for the first stage. The diameter and length of the pressure vessel are 8 in and 280 in, respectively.

The body material of the pressure vessel is glass/epoxy, and there are a total of 425 vessels necessary for stage 1. For stage two, the specific vessel we are using from LennTech is 8-E-1200-1M-to-8M-R6. The diameter, height, and body material of these vessels are the same as that of stage one. However, we only need 308 vessels for stage two due to the fewer number of RO membranes.

Furthermore, we are utilizing the PX Q300 Pressure Exchanger Device from Energy Recovery®. Pressure exchangers transfer the power from a high pressure stream to a low pressure stream. In this case, the high pressure brine stream will transfer its energy to part of the low pressure RO inlet stream, which allows us to decrease the pumping power required for the entire inlet stream. The system will consist of 23 pressure exchangers and each pressure exchanger can transfer 60% of the power to the low pressure stream.

4.3. Permeate Post-treatment

Figure 4.4: PFD of Permeate Post-Treatment

After the water is filtered through reverse osmosis, the permeate is sent to post-treatment. First, 11 MLD of permeate is diverted to remineralization. 0.2 MLD is used in the lime slurry mixer to be mixed with 11.6 tons/day of Ca(OH)₂. The lime slurry and the rest of the 10.8 MLD of permeate is mixed in a lime saturator in order to make a 1000 mg/L Ca(OH)₂ solution. The lime saturator is designed so that any impurities from the lime have fallen out of solution and have been removed. After the lime saturator, 6 MLD of the treated water will be diverted to a TOMCO₂ Systems PSF, where it will be pressurized to 60 bar and be mixed with 13.1 tons/day of CO₂ at 60 bar to create carbonic acid. The carbonic acid will then be injected into the rest of the untreated and treated permeate water stream to create 150 MLD of remineralized water. 15 MLD of the remineralized water is redirected to a seperate stream to be injected with 375 kg/day of Cl₂ using a compact chlorometer, and the chlorinated water is then injected back into the mainstream to create 150 MLD of product water, which is then sold.

4.4. Brine Post-Treatment

Figure 4.5: PFD of Brine Post-Treatment

4.4.1. Electrodialysis

The electrodialysis system will consist of 435 membrane pairs in a single stack to process the 36.8 MLD brine stream entering. The total current and stack voltage are 65,250 A and 6 V, respectively, which means a power of 391.5 kW is required for the system. The inlet stream has a salt concentration of 80 g/L which is decreased to 33 g/L in the diluate stream and increased to 200 g/L in the concentrate stream. The diluate stream will be returned to the RO system at 26.8 MLD. The concentrate stream will be brought to the crystallizer at 10.5 MLD. The remaining system specifications can be found in **Table 3.5**.

4.4.2. Crystallizer

The crystallizer system is designed as a heated moving bed with vents above the unit to release water-saturated air into the atmosphere. The 10.5 MLD brine stream from the electrodialysis unit will be the inlet stream, and 349 MW will be the heat duty provided to the system to evaporate the water off. The total 2.1 million kilograms of salt will be recovered and deposited in silos for distribution.

4.5. Pumps

The desalination plant contains 12 total pumps to help transfer water throughout the system. The pump ID names are located in **Figure 4.1**, and the pumping power for each unit was calculated with the differential pressure in order to determine the amount of power needed.

Pump Location	Flow Rate (MLD)	Initial Pressure (bar)	Goal Pressure (bar)	Final Pressure (bar)	Pumping Power (kW)	Pumping Material
P101	196.0	1	5	6.5	1249.16	Copper Alloy
P102	192.1	4	6.25	7.75	838.10	Copper Alloy
P103	182.5	5	35	36.5	6657.76	Copper Alloy
P104	97.8	29	55	56.5	3113.78	Copper Alloy
P105	66.2	55	55	56.5	21.43	Copper Alloy
P106	83.6	55	80	81.5	2566.01	Copper Alloy
P107	36.8	2	5	6.5	192.50	Copper Alloy
P108	26.0	3	55	56.5	1610.54	Copper Alloy
P109	6.0	2	60	61.5	413.33	Stainless Steel
P110	11.0	1	3	4.5	44.81	Stainless Steel
P111	11.0	2	3	4.5	32.08	Stainless Steel
P112	15.0	2	3	4.5	43.74	Stainless Steel
Total Pumping Power			16783.24			

Table 4.1: Pumping power for each pump unit

Table 4.1 shown above depicts the power for each pump as well as their construction material. 1.5 bar is added to the goal pressure to account for frictional losses to give the final pressure. All pretreatment and brine post-treatment pumps must be made from copper alloy because it will prevent corrosion from salt water. However, once the salt is separated from the water, the pump material for the permeate post-treatment will be stainless steel.

4.6. Storage Tanks

Various storage tanks are necessary for this production process in order to accommodate for redundancies and allow for unexpected maintenance on broken equipment. Within the pretreatment section, two storage tanks are used for the aluminum sulfate coagulant and pretreated saltwater. Aluminum sulfate solution is stored in 1 cylindrical steel tank with a thermoset powder coating that is capable of storing 14 days worth of solution; approximately 0.11 ML. This aluminum sulfate solution storage tank will be purchased through National Storage Tank Inc, and measures a height and diameter of 2.4 and 2.7 m respectively (National Storage Tank Inc, 2020). The tank volume is 4,000 gal which can hold 0.15 ML of solution at ambient pressure.

A secondary pretreatment storage tank is placed at the end of pretreatment in order to serve as storage in case the RO process section must be temporarily shut down for maintenance. 22 steelcore corrugated bolted steel storage tanks are capable of holding 1 full day's worth of treated saltwater, or 164 ML. These tanks will be purchased from the National Storage Tank Outlet with tank dimensions of 21.6 m in diameter and a height of 21.3 m (National Storage Tank Outlet, 2020). The tanks are manufactured with zinc galvanized steel and will contain water at ambient pressure. Each tank contains a volume of 2,000,000 gallons or the equivalent 7.6 ML of water.

The post-treatment freshwater process contains three storage tanks as well; a slaked lime storage silo, liquid chlorine storage tank, and carbon dioxide storage tanks. The lime storage silo is manufactured to contain 60 days worth of lime powder, equivalent to 696 tons of lime, and is purchased from the National Storage Tank Outlet. This tank is a factory coated bolted steel tank with a thermoset powder coating with a conical roof style (National Storage Tank Outlet, 2020).

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The diameter and height dimensions of this cylindrical tank are 5.5 and 12.2 m respectively. Its total volume is 80,000 gallons and is held at atmospheric pressure.

The liquid chlorine storage tank is built to hold 2 days worth of solution which amounts to 135,000 gallons of liquid chlorine. To accommodate this need, 8 plastic polyethylene storage tanks from National Tank Outlet will hold 16,874 gallons of liquid chlorine at a pressure of 6 bar (Plastic Mart, 2022). Each cylindrical tank has diameter and height dimensions of 4.2 and 6.15 m respectively and has a 20,000 gallon capacity, allowing for each tank to be 84% filled.

Finally, a carbon dioxide storage tank is used to store CO_2 for its injection into the incoming freshwater stream. In order to store 14 days worth of CO_2 gas, 3 70-ton large portable cylindrical storage units will be purchased from TOMCO₂ Systems. Each tank contains an outer jacket aluminum cover as well as polyurethane insulation. Each tank can hold 70 tons of CO_2 gas and contains the following dimensions: height of 4.1 m, length of 17.5 m, and a width of 2.3 m (TOMCO₂ Systems, 2022). Lastly, the tanks will be maintained at a pressure of 24 bar.

5. Future Work and Recommendations

In order to improve the design of this desalination plant in the future, we would suggest removing the crystallizer. Challenges that would be faced by removing the crystallizer include buying a significant amount of extra land for salt pools, having to dump the brine 3 months out of the year during the monsoon season, including a more in depth cleaning process for the salt once the water has evaporated since it was open to the elements, and finally paying for the negative environmental impacts that highly concentrated salt pools will have on the soil and wildlife.

A recommendation for the crystallizer would be investigating the use of a multi-effect evaporation to use heat more efficiently to evaporate the water. Also, investigating a method of crystallization to recover and process the steam into potable water. This could lead to potentially lower costs for pumping and RO costs since the RO could be run at a lower pressure since more water will be recovered from the brine.

Because creating salt was not profitable, future groups could consider disposing brine instead of producing salt. Brine can be hazardous to the marine environment due to its higher-than-normal salinity or the presence of pollutants that would not otherwise exist in the receiving water body. However, with the right precautions, brine disposal in surface water could remain a feasible option for seawater desalination plants. To reduce salinity, the brine might be mixed with ordinary seawater or municipal wastewater before disposal. Another type of disposal is sewer discharge, which involves dumping brine into a neighboring wastewater collection system. Because of the possible detrimental influence of the brine's high total dissolved solids (TDS) concentration on the receiving wastewater treatment facility, this method is only widely

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used by small-scale brackish water desalination plants. However, if the brine was treated to reduce the TDS content, this method could be applied to seawater desalination plants.

6. Conclusion

The implementation of an RO desalination plant in Chennai, India will use pretreatment, RO separation, and post-treatment methods to produce potable water for the local community. An inlet flow of 196 MLD of saltwater ultimately results in 150 MLD of freshwater which cuts half of Chennai's freshwater scarcity. Using various filtration methods, both physical and chemical impurities are eliminated, proceeding with a two stage RO process which separates salt NaCl from the water. Finally a post-treatment process is done for both brine retentate and freshwater permeate in order to meet India and US quality standards. The resulting amounts are 150 ML daily of potable freshwater and 2.3 million kg/day of crystalized salt. There are a few limitations for this project design. While this project was not profitable due to the salt co-production and the high utility costs associated with it, future design teams could still produce salt but have a less concentrated brine stream to reduce pumping powers, as well as, use passive evaporation or less expensive active evaporation means to crystallize the salt and reduce the crystallization costs. It should be noted that RO has been and continues to be used to provide potable water within and outside of India. If alternative methods were utilized for brine post-treatment, over a 20 year period, this plant design may result in a lower economic deficit than currently predicted.

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8. Appendix

8.1. Reverse Osmosis

Equation 8.1:

Pumping Power:

Stage 1: Power = $(1.13m^3/s * 2752 \text{ kPa}) = 3.1 \text{ MW}$

Stage 2: Power = $(0.97m^3/s * 2652 \text{ kPa}) = 2.6 \text{ MW}$

8.2. Permeate Post-Treatment

Equation 8.2:

Dosages for Ca(OH)₂ and CO₂ for 100 mg/L of Hardness (CaCO₃):

 $Ca(OH)_2 (mg/L) = 100 mg/L \div 100 g/mol x 74.1 g/mol \div 0.95 = 77.36 mg/L$

 $CO_2 (mg/L) = 100 mg/L \div 100.08 g/mol x 44.01 g/mol x 2.0 = 87.32 mg/L$

Equation 8.3:

Waste from impurities with 30% Solid Content:

Waste $(mg/L) = 77.36 mg/L \ge 0.05 \div 0.30 = 12.89 mg/L$

Equation 8.4:

Volume for Lime Slurry:

Retention Time = 240 min

Flow Rate = 102 L/min

Volume Size = 240 min x 102 L/min = 24480 L = 24.48 m³

Equation 8.5:

Volume for Lime Saturator:

Retention Time = 100 min

Flow Rate = 7655 L/min

Volume Size = 100 min x 7655 L/min = 765500 L = **765.5 m³**

8.3. Brine Post-Treatment

Variables:

 N_m [mol/s] = molar flux of salt across the membrane

$$\eta$$
 = current efficiency = 0.85

$$I [A] = electric current = 28200$$

$$N_{mp}$$
 = number of membrane pairs = 188

z = charge number of salt = 1 for NaCl

$$F$$
 [C/mol] = Faraday's constant = 96485

 $c_{D,in}$ [mol/m³] = molar concentration of diluate stream entering = 1370

 $c_{D,out}$ [mol/m³] = molar concentration of diluate stream leaving = 565

 $c_{C,in}$ [mol/m³] = molar concentration of concentrate stream entering = 1370

 $c_{C,out}$ [mol/m³] = molar concentration of concentrate stream entering = 3420

 Q_{D} [m³/s] = volumetric flow of diluate stream

 Q_{c} [m³/s] = volumetric flow of concentrate stream

Equation 8.6:

$$N_m = \frac{\eta I N_{mp}}{zF} = \frac{(0.85)(28200)(188)}{(1)(96485)} = 46.7 \text{ mol/s}$$

Equation 8.7:

$$c_{D,in} - c_{D,out} = + \frac{\eta I N_{mp}}{Q_D z F}$$
$$Q_D = \frac{\eta I N_{mp}}{(c_{D,in} - c_{D,out}) z F} = 0.058 \text{ m}^3/\text{s}$$

Equation 8.8:

$$c_{C,in} - c_{C,out} = -\frac{\eta I N_{mp}}{Q_C zF}$$

 $Q_C = -\frac{\eta I N_{mp}}{(c_{C,in} - c_{C,out})zF} = 0.023 \text{ m}^3/\text{s}$

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