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# Green RO pretreatment

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# **Green RO Pretreatment**

An Undergraduate Honors College Thesis

in the

Ralph E. Martin Department of Chemical Engineering  
College of Engineering  
University of Arkansas  
Fayetteville, AR

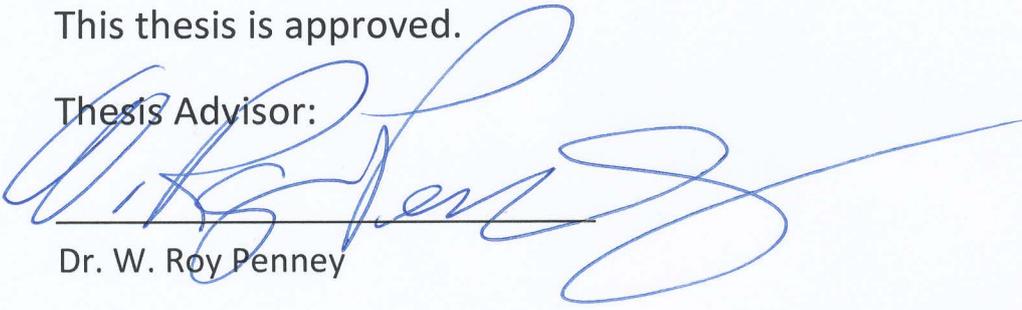
by

Elizabeth Fullerton

April 16, 2012

This thesis is approved.

Thesis Advisor:

A handwritten signature in blue ink, appearing to read 'W. Roy Penney', is written over a horizontal line. The signature is fluid and cursive, extending to the right of the line.

Dr. W. Roy Penney

Thesis Committee:

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I participated in a senior design course known as WERC: A Consortium for Environmental Education and Technology Development. I was selected as team coordinator for this year's project. My team and I selected Task 5: "Green" RO Pretreatment. As team coordinator I led biweekly meetings; managed a team of four individuals, delegating projects to each individual; and contributed in every aspect of this project.

At the start of this project my team and I researched various Reverse Osmosis (RO) pretreatment systems. I specifically researched Ultrafiltration (UF), Microfiltration (MF), Sand Filters and Bag and Cartridge filters. Each technology was compared based on its advantages and disadvantages. Initially we selected a microfiltration unit with air backflush. The air backflush feature would allow us to drastically reduce the use of hazardous chemicals and thus create a "Green" system. The department had actually acquired a manually operated air backflush MF unit from a company once known as MEMCOR a few years ago. The unit was operated for the first time to determine if it was in working order. Once this was done I helped with the design of several experiments to test the unit. As problems arose I helped to revise the system testing environment and individual experiments. I personally helped perform several experiments to determine the frequency and duration of air backflush cycles that would optimize the lifetime of the membrane. It became evident early on that chemical cleaning would be required, and following the recommendation of our advisor, a sand filter was added in front of the MF unit to reduce the amount of foulants to come in contact with the membrane.

The assembly of the apparatus was done by other members of my team; however, I was involved in the purchasing of materials for automation of the unit. I was in charge of sample collection and delivery to the Arkansas Water Quality Laboratory of the Arkansas Water Resources Center for turbidity, total dissolved solids, and electrical conductivity tests.

My primary contributions were in the completion of the economic analysis, review of government regulations, and the first draft of the paper. I wrote all portions of the first draft of the paper other than technologies considered, Sand Filter experimental results and HMF Unit experimental results; additionally all process flow diagrams were completed by other members of my team. My team members and I contributed to the paper review and rewrites. Our paper had to be audited by three professional engineers; I was in charge of soliciting the audits. For the presentation I focused on the objective of Task 5, design premises and the technology evaluated. I assisted in edits of the presentation and combine the slides completed by my other team members.

The 2012 WERC design competition was held at New Mexico State University in Las Cruces, New Mexico, over a three day period. I assisted in the presentation of our project to a group of judges, faculty and students from institutions across the country. Although our team did not place at the competition, I obtained valuable experience working on a real-world problem that could not have been learned in a traditional classroom setting.

# Green RO Pretreatment

WERC 2012

Task # 5

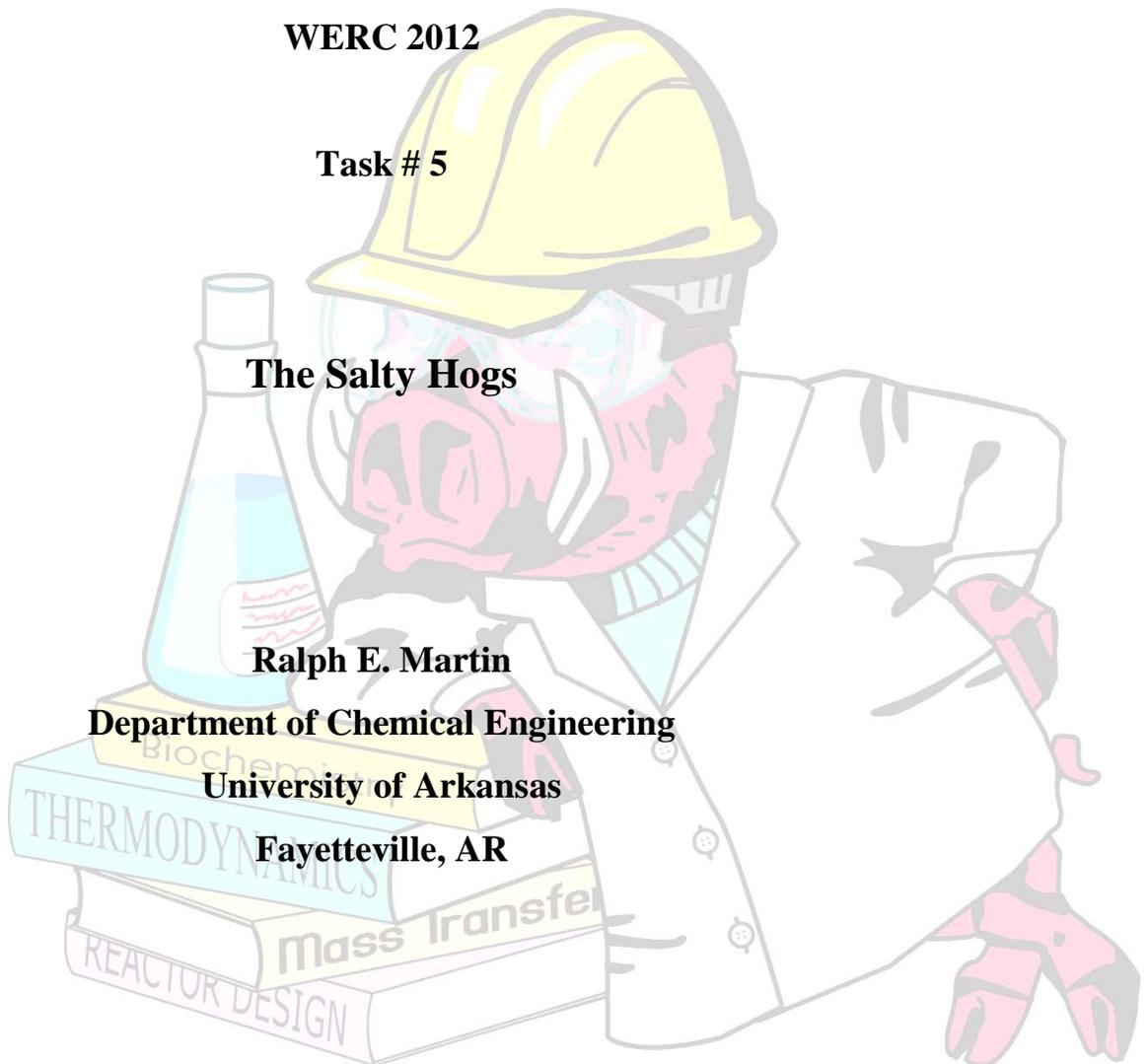
The Salty Hogs

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# **Green RO Pretreatment**

## **TASK # 5**

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## **EXECUTIVE SUMMARY**

The use of sea water desalination to produce potable water for maritime vessels is essential for eliminating storage of potable water. Reverse osmosis (RO) is the technology of choice for desalination purposes; however, raw seawater must undergo pretreatment to avoid rapid fouling and plugging of RO membranes and modules. Current pretreatment technology is effective in the open ocean but in coastal regions involves frequent replacement of filter cartridges and increased maintenance costs.

The standard pretreatment process consists of 20  $\mu\text{m}$  cartridge filters and 3  $\mu\text{m}$  cartridge filters in series. Several technologies were considered as options for improving the current pretreatment process. Multimedia filtration followed by hollow-fiber ultrafiltration was considered, but it was rejected because of the high required membrane surface area, high capital cost, and high pressure drop requirements. Mechanically-enhanced self-cleaning filters, followed by either microfiltration (MF) or ultrafiltration (UF), were also considered. The self-cleaning filters were rejected because of mechanical complexity and high initial investment. The technology selected by The Salty Hogs is a sand filter followed by a 0.1  $\mu\text{m}$  hollow-fiber microfiltration (HMF) unit. The sand filter is less expensive and less complex than a multimedia filter, and the HMF unit requires much less area and has lower capital costs than a UF unit. The sand filter/HMF combination can handle feeds with various turbidity levels and meet the task requirements to feed 21 gpm of filtered water to the RO unit, which is the flow rate required for the RO unit to produce 12,000 gallons/day of potable water with 40% recovery.

Although the project focused on oceanic applications, inland water desalination plants can integrate The Salty Hogs' technology into existing RO desalination processes. This report presents the design for a full-scale, shipboard system that incorporates sand filtration and hollow-fiber microfiltration. The shipboard unit is designed for optimal power usage and minimal space requirements.

The water provided by the recommended process is filtered to 0.1  $\mu\text{m}$ , an improvement that will lengthen the lifetime of the current spiral-wound RO membranes. The system is automated to back-wash the sand filters with water and back-flush the HMF membranes with air. Periodic chemical cleaning is required and involves the use of a sodium hydroxide wash followed by an acetic acid rinse. The spent sodium hydroxide and acetic acid solutions will be combined and neutralized before being discharged to the ocean.

The economic parameters for the system have been determined. Purchased equipment cost is \$21,000, and the total capital investment, based on factored estimates, is \$133,000. Over the projected three year lifetime of the RO membranes, the proposed process provides an incremental savings, including reduced maintenance and filter usage, of \$92,000, over the three year project lifetime when compared to the current cartridge filter system.

The WERC task statement specified a goal for energy usage of 1.2 HP, which is about 10% of the total power required for the RO operation. The current full-scale design uses ~3 HP.

## **INTRODUCTION**

The production of potable water aboard sea going ships has been a concern for some time. The U.S. Navy has ships that are offshore for six to nine months at a time, during which time they must provide freshwater for the crew. Desalination using RO membranes has been used for this purpose since 1988, when the first prototype was installed aboard the destroyer USS Fletcher.<sup>1</sup> Since then RO has become the standard method of desalination for its relatively low operating costs, independence from the ship's steam system, and reduced power requirements, chemical additives and maintenance.<sup>2</sup> RO units typically last three to five years before requiring replacement after operating at capacities varying from 2,000 gpd to 125,000 gpd aboard Navy ships. However, as specified by the current WERC Task 5, a 12,000 gpd capacity is characteristic of most commissioned vessels.

In coastal environments unprotected RO membranes tend to foul very rapidly. Consequently, a pretreatment step is required to remove particulates in order to extend the life of the RO membranes. The most common RO Navy pretreatment process consists of (1) a centrifugal separator and (2) a 20  $\mu\text{m}$  depth filter followed by (3) a 3  $\mu\text{m}$  depth filter. The RO triplex plunger feed pump takes suction from the 3  $\mu\text{m}$  filter. Pressure pulsations caused by the plunger pumps are minimized by an accumulator on the pump discharge and the RO units are installed downstream of the accumulator. A turbine is used downstream of the RO unit on the concentrated retentate stream for energy recovery. The turbine discharges retentate to the ocean.

Task 5 specifically addresses the pretreatment portion of the potable water system. The current cartridge filters last 4 to 6 weeks at sea before requiring replacement, which requires 1-6 hours of down-time. When the ship is in coastal regions, the seawater has a high level of suspended particles and a larger Silt Density Index (SDI) than the open ocean. The SDI

measures the rate of change of the permeate flux with respect to time (%/min) of a 0.45  $\mu\text{m}$  filter when subjected to a constant pressure drop of 30 psi across the filter for 15 minutes.<sup>3</sup> Contaminants include silt, bacteria, algae, sand, krill, starfish larva and mixed plankton.<sup>1</sup> RO membranes cannot handle SDI values of more than 3.0-4.0.<sup>4</sup> The Office of Naval Research is searching for a pretreatment method to meet “15 minute SDI values (ASTM D4189-07) of less than 3.0 and turbidity values of less than 1 NTU when operating in coastal regions”.<sup>2</sup> In coastal regions, cartridges may only last 10 hours, which is why alternative pretreatment methods are being investigated to decrease replacement requirements and maintenance times.

## **TASK PARAMETERS**

The purpose of this task is to develop an alternative to current RO pretreatment methods. The task parameters are:

### **Feed Composition**

1. De-chlorinated tap water.
2. 32,000 ppm Sea Salt.
3. 75 mg/L Klamath Blue Green Algae Powder.
4. 20 mg/L Orchid Pro (by Turf Pro USA).

### **Task Premises**

1. RO recovery will be 40%.
2. The well mixed feed is strained through a 500  $\mu\text{m}$  screen.
3. A modern energy recovery device will be used to minimize pump power.
4. Energy usage cannot exceed 10% of what is used for RO operation.
5. Operating space cannot exceed three times the amount currently used.
6. System must produce 30,000 GPD of filtered seawater for the RO system.
7. Volume of stored items must be less than 100  $\text{ft}^3$  (one-tenth the volume of the currently stored filter elements assuming replacement every four days).
8. No hazardous chemicals can be used when the vessel is at sea.
9. Maintenance should be less than that of the current system; currently cartridge replacements occur every four days, requiring four hours of down time.

10. Particles must be filtered down to 0.1  $\mu\text{m}$ .

## **TECHNOLOGIES CONSIDERED**

Several pretreatment methods were evaluated while considering the task requirements. The advantages and disadvantages of each method were carefully considered. The methods evaluated were: (1) Bag and cartridge filters, (2) Sand filters, (3) Mechanical self-cleaning filters, (4) UF membranes, and (5) MF membranes. Other methods, such as forward osmosis, are technically feasible but not economical.

### **Bag and Cartridge Filters**

Bag and cartridge filters have the lowest capital cost of any filter type and are easily replaced and maintained with minimal downtime. Cartridge filters are currently used and the Navy is searching for a better filtration system because of their frequent replacement and large storage space requirements. Bag filters are available for particulate removal down to 1  $\mu\text{m}$ . Thus, they could be used in series and as a pretreatment for UF or MF membranes, but this was not deemed appropriate for the current study.

### **Sand Filters**

Sand filters utilize the same phenomena by which groundwater is purified naturally. Depending on the filtration media, particles as small as 5 microns can be removed from the water.<sup>5</sup> Typically, sand filters are used and require a back-wash fluidization to remove the filtered material that is collected within the sand bed. These filters come in various forms, most notably the mono, dual, and multimedia filters. An example of an acceptable multimedia prefilter was utilized by the University of Arkansas Team that won first place for their design in WERC 2010's Task 4, the same as this year's Task 5.<sup>6</sup> The filter consisted of sand, anthracite, and garnet. The effort this year is to identify a pretreatment method that is more economical than the multimedia/UF combination utilized in 2010. The multimedia/UF combination was thoroughly investigated by a literature search and experimental testing in 2010. The work is summarized in the appropriate task report which is available to the public for review. No additional work was done on these methods for the current task.

### **Self-cleaning Filters**

Typical self-cleaning devices use a rotating suction arm to create a localized back-wash

through the filter and high-velocity shear across the filter surface to dislodge any caking or fouling layers that may form, removing them from the system through the suction arm.<sup>8</sup> This method is theoretically superior to units that only utilize periodic back-washing or chemical cleaning due to fouling over time. Self-cleaning devices typically filter to 5-10 microns although some can filter as small as 1  $\mu\text{m}$ . A large initial investment is required but is compensated for by a long operational lifetime. Self-cleaning filters can last up to 8 years, far longer than the average lifetimes of any of other filters.<sup>9</sup> In order to meet the requirements of Task 5, a self-cleaning filter would require additional filtration downstream. High initial investment and mechanical complexity eliminated self-cleaning filters from consideration.

### **Ultrafiltration**

Ultrafiltration, using polymeric or ceramic membranes, removes particles larger than 0.01  $\mu\text{m}$ . UF membranes are capable of maintaining the permeate SDI consistently between 0.7 and 3.0.<sup>4</sup> The WERC 2010 team found that UF, although filtering beyond the requirements of Task 4 in 2010, was a more economical choice than MF because the smaller pore size in the UF membrane was less susceptible to fouling and was easily cleaned by back-flushing with water. A literature search found that maintenance for a typical UF system consisted of back-washing every 30-90 minutes, along with a 1 ppm solution of sodium hypochlorite to prevent biological fouling.<sup>4</sup> Chemical cleaning cannot be avoided on these extremely fine filters because extracellular polymeric substances (EPS) are produced by algae on the surface of the membrane and can only be removed via a chemical wash.<sup>7</sup> Another disadvantage of UF membranes is a large surface area and pressure drop requirement.

### **Microfiltration**

Microfiltration removes particles larger than 0.1  $\mu\text{m}$ . For coastal water, MF membranes have been shown to initially lower the SDI to 3.8 but ultimately stabilize between 2.0 and 3.0.<sup>4</sup> Literature reveals that air, or filtrate, back-flushing and periodic chemical cleaning have been used to remove fouling from MF membranes.<sup>4</sup>

There are several configurations for operating MF or UF membranes that need to be considered. The module types include: plate-and-frame, spiral-wound, hollow-fiber, and tubular, among others, each having advantages and disadvantages depending on the system.

Plate-and-frame modules consist of flat membrane sheets separated by support plates that channel and distribute flow across the surface of the membranes. Their major disadvantage is the low surface area to volume ratio, meaning large volumes are required to achieve the necessary membrane area. Other disadvantages include the difficulty of removing and replacing the plates and the high relative cost.

Spiral-wound membrane modules consist of the same type of membranes involved in plate-and-frame modules rolled into a spiral configuration. This produces a higher surface area to volume ratio, lowering the relative cost and volume requirements. However, spiral-wound membranes cannot be back-washed without damaging the membranes. This makes cleaning the membranes a difficult process.

Tubular modules typically have tube diameters of about 0.5-5 cm.<sup>10</sup> They operate by feeding through the tube-side of the bundle, allowing permeate to travel through the tube walls to the shell-side where it exits the module. A minor disadvantage for the tubular units is a high capital cost per unit area; however, they are easily cleaned which somewhat compensates for their high relative cost.

Hollow-fiber modules are similar to tubular modules but have much smaller tube size, typically less than 1.0 mm, giving these units the highest surface area to volume ratio of all the membrane filtration units.<sup>10</sup> Feed can be supplied to either the shell or the tube-side with permeate exiting through the corresponding tube or shell-side. Unfortunately they can be difficult to clean, but the low relative cost makes them an appealing option in many cases.

## **BENCH-SCALE APPARATUS**

The bench-scale unit used in this study consists of a sand filter followed by an HMF unit. Sand filtration was used as an initial filtration of the feed mixture in order to minimize the HMF membrane's fouling due to larger particles.



**Figure 1.** Sand Filter Apparatus

### **Sand Filter**

The sand filter (see photograph in Figure 1, at left) consists of a feed tank, feed pump, and filtrate tank in addition to the sand bed, itself. The feed pump is a variable speed, 5 gpm, rotary vane pump. The sand bed was constructed from a 2.5 in. S40 (2.47 in. ID) x 60 in. long clear PVC pipe and a 20 in. layer of fracking sand. Stainless steel mesh distributors at the ends of the pipe prevent sand loss during filtration and fluidization. The filtrate tank is a 100 gallon, epoxy-coated, carbon steel tank. The sand was 100-300  $\mu\text{m}$  fracking sand.

### **HMF Unit**

The HMF unit (see photograph in Figure 2, below) consists of a 5 gpm centrifugal pump and a 0.2  $\mu\text{m}$ , polypropylene hollow-fiber bundle with roughly 10  $\text{ft}^2$  of membrane surface area. The housing for the bundle is approximately 2.5 in. ID x 7 in. long. Feed is circulated through the shell-side of the bundle by the centrifugal pump and permeate exits the tube-side. PLC-controlled solenoids operate pneumatic pistons that actuate three valves so that back-flushing of the membrane can be automated.



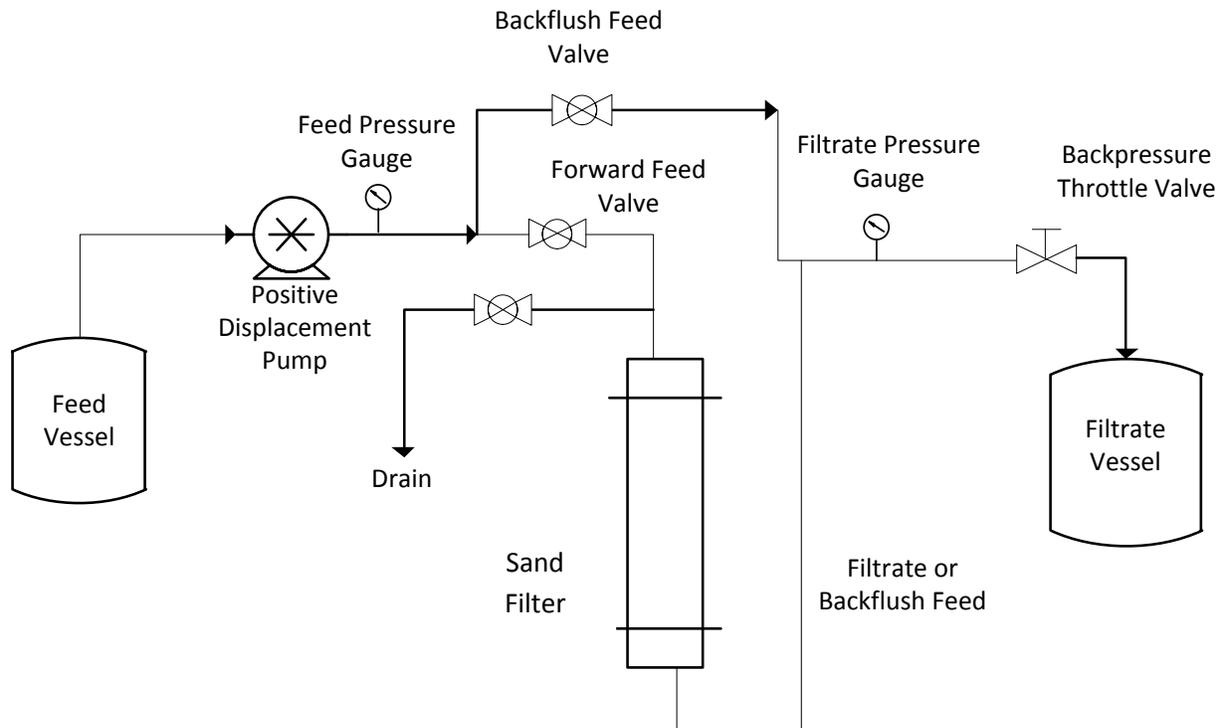
**Figure 2.** HMF Unit

## BENCH-SCALE OPERATION

Approximately 120 gallons of feed mixture were made in accordance with the previously stated WERC specifications. The sand filter and HMF unit were operated separately with the sand filter providing about 100 gallons of filtrate which was then used as feed for the HMF unit.

### Sand Filter

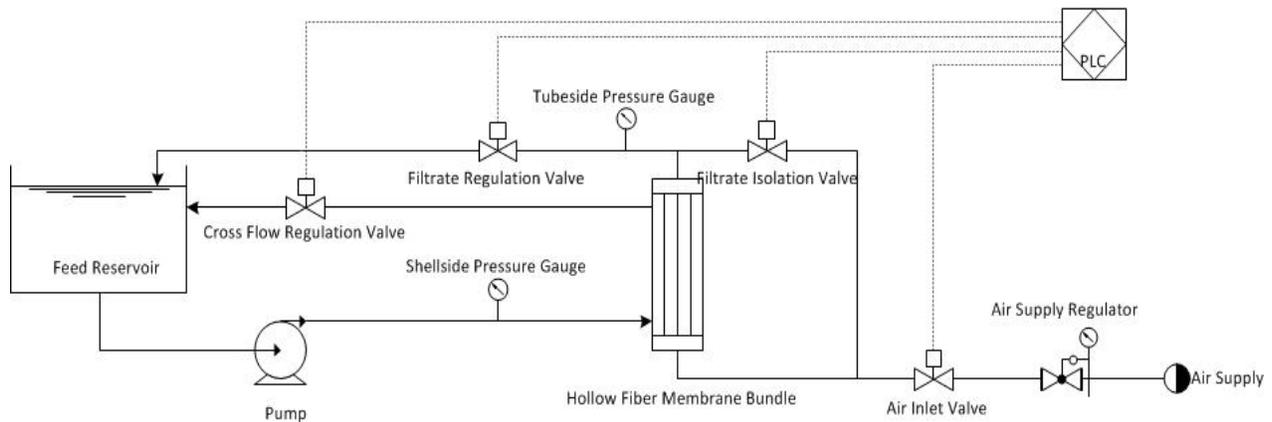
The feed pump was set to provide an initial inlet feed pressure of 20 psig to the top of the sand bed. Filtrate from the sand bed was directed to the 100 gallon storage tank. When the inlet pressure increased to 40 psig, the sand bed was back-washed with DI water. In the full-scale process, back-washing will utilize RO permeate. A description of the back-wash procedure is as follows: (1) The pump was turned off. (2) The pump discharge was directed to the bottom of the sand bed. (3) The feed was directed to a bucket of DI water. (4) The effluent from the sand bed was directed to a drain. (5) The bed was rapidly fluidized and allowed to settle. (6) Step 5 was repeated three times or until the effluent was colorless. Normal operation was resumed after back-washing, which lasted approximately 10 minutes. Figure 3, below, presents a process flow schematic for the sand filter.



**Figure 3.** Sand Filter Process Flow Schematic

## HMF Unit

Feed to the HMF unit was circulated through the shell-side of the hollow-fiber bundle by the centrifugal pump. The permeate and retentate streams were returned to the storage tank in an effort to reduce the overall amount of feed required for extended testing. The retentate, or cross flow, valve was operated at 80 degrees toward its 90-degree closed position to provide a constant 10 psig pressure drop across the membrane. Every five minutes the system was back-flushed by pulses of compressed air at 80 psig delivered to the tube-side of the fibers. The automated back-flush procedure for the HMF unit is as follows: (1) The retentate and permeate valves were closed. (2) The air supply line was opened and the pressure in the bundle housing was allowed to build to 80 psig, at which point the air supply line was closed. (3) The retentate valve was rapidly opened. (4) Steps 2 and 3 were repeated three times. Normal operation was resumed after back-flushing, which lasted approximately 40 seconds. A process flow schematic for the HMF unit is given in Figure 4, below.



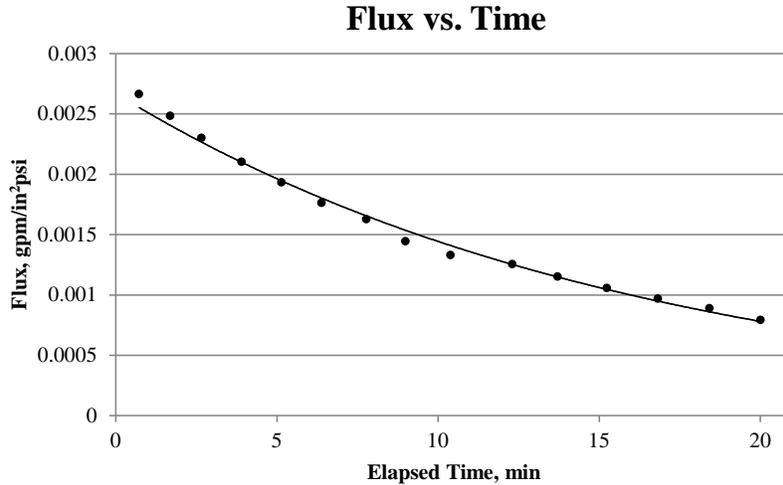
**Figure 4.** HMF Process Flow Schematic

## LABORATORY EXPERIMENTAL RESULTS

Laboratory experiments were conducted using the bench-scale apparatus described above. Before experiments were conducted, the HMF unit was chemically cleaned and back-washed as per manufacturer's instructions. Experimentation focused on three factors: sand filter performance, HMF permeate flux determination, and optimization of HMF back-flush frequency and duration to minimize fouling.

## Sand Filter

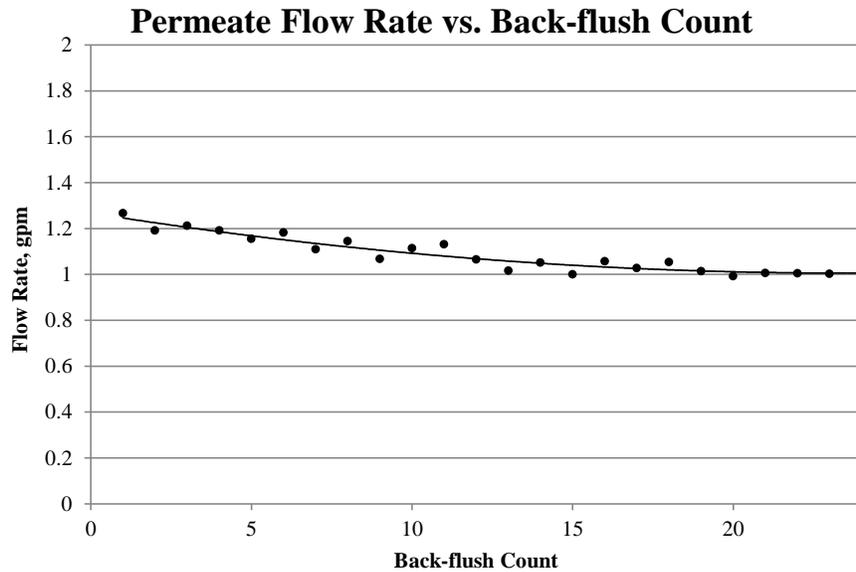
The sand filter was operated at an initial inlet pressure of 20 psig and filtrate flow rate of 0.51 gpm, a flux of  $4.7 \times 10^{-3}$  gpm/in<sup>2</sup> psi. After 20 minutes, the pressure increased to 40 psig and the filtrate flow rate had decreased to 0.19 gpm, a flux of  $9.7 \times 10^{-4}$  gpm/in<sup>2</sup> psi. The data collected during the experiment, shown below in Figure 5, indicates that the 20 minute interval between back-washes may be able to be increased.



**Figure 5.** Plot of Filtrate Flux vs. Elapsed Time

## HMF Unit

Throughout the testing period, the retentate flow rate was recorded once and the permeate flow rate was recorded multiple times between each back-flush cycle. The optimal back-flush interval was determined to be five minutes. As shown in Figure 6, this back-flush procedure resulted in permeate flow rates that asymptotically approached 1.0 gpm. It should be noted that the height of the permeate line relative to the circulation pump had a noticeable effect on the permeate flow rate. For the data reported in the figure, the permeate line was even with top of the storage tank, approximately 4 ft above the circulation pump. Measurements taken closer to the ground resulted in flow rates closer to 1.5 gpm.



**Figure 6.** Permeate Flow Rate vs. Back-flush Cycle Count  
(Each count represents five minutes of elapsed time.)

### Membrane Cleaning

During laboratory tests, the hollow fiber membrane was chemically cleaned between each experimental run. A cleaning solution of 2 wt% sodium hydroxide in water was used, to break down any organic material, followed by a rinse with 1% acetic acid, to dissolve any inorganic compounds that may have precipitated onto the fibers. The chemical cleaning procedure used is as follows: (1) Caustic solution was circulated through the unit in its normal operating mode for 10 minutes. (2) Permeate flow was stopped and caustic solution was circulated on the shell-side for 15 minutes. (3) The pump was stopped and the solution was allowed to soak for 30 minutes. (4) The pump was turned on for five minutes and then stopped. (5) Permeate flow was started and the unit was rinsed with a single pass of DI water for 10 minutes, back-flushing intermittently, discharging the sodium hydroxide solution. (6) The unit was returned to standard operating mode and circulated the 1% acetic acid solution for 20 minutes. (7) The unit was back-flushed and then the acetic acid was rinsed with a single pass of DI water for five minutes before feed was returned to the system.

Permeate flow, after the caustic cleaning, was typically restored to ~2 gpm and then to ~3 gpm after the acid rinse. These measurements were taken while holding the permeate line approximately 1 ft above the ground.

## Suspended Solids and Turbidity

The testing procedures used corresponded to the testing procedure which will be utilized at the WERC competition for Task 5. The various water quality indicators WERC will be performing include turbidity, total dissolved solids (TDS), and electrical conductivity (EC).<sup>11</sup>

The feed and MF permeate were tested for total dissolved solids (TDS) and turbidity according to EPA Methods 160.1 and 180.1 by the Arkansas Water Quality Lab.<sup>12,13</sup> Electrical conductivity was also determined by EPA Method 120.1.<sup>14</sup> The turbidity and electrical conductivity results are given in Table 1, below.

**Table 1.** Water Test Results

<b>Stream</b>	<b>Turbidity (NTU)</b>	<b>TDS (mg/L)</b>	<b>EC (<math>\mu</math>S)</b>
Feed	23	20,671.80	35,600
Permeate	1.6	20,544.70	36,300

The SDI is often used as a performance indicator for MF units.<sup>3</sup> The SDI was not performed because the procedure typically requires filtering through a 0.45  $\mu$ m filter which, because the HMF unit filters to 0.2  $\mu$ m, would result in an invalid measurement. The permeate turbidity is high compared to literature values of less than 0.2 NTU for an MF membrane.<sup>15</sup> The turbidity was tested using methods applicable primarily to fresh wastewater, but the presence of high salt concentration increased the turbidity.

## FULL-SCALE DESIGN

The equipment for the full-scale system consists of (1) a stainless steel vane feed pump, (2) a stainless steel centrifugal pump for cleaning solution circulation, (3) a train of five sand filters, (4) a bank of six HMF membranes modules, (5) three peristaltic pumps for preparation and mixing of cleaning solutions, (6) sodium hydroxide and acetic acid metering tanks, (7) a mixing tank for preparation and neutralization of the cleaning solutions, and (8) a temporary holding tank for the sodium hydroxide solution during the acetic acid rinse.

## Sand Filter

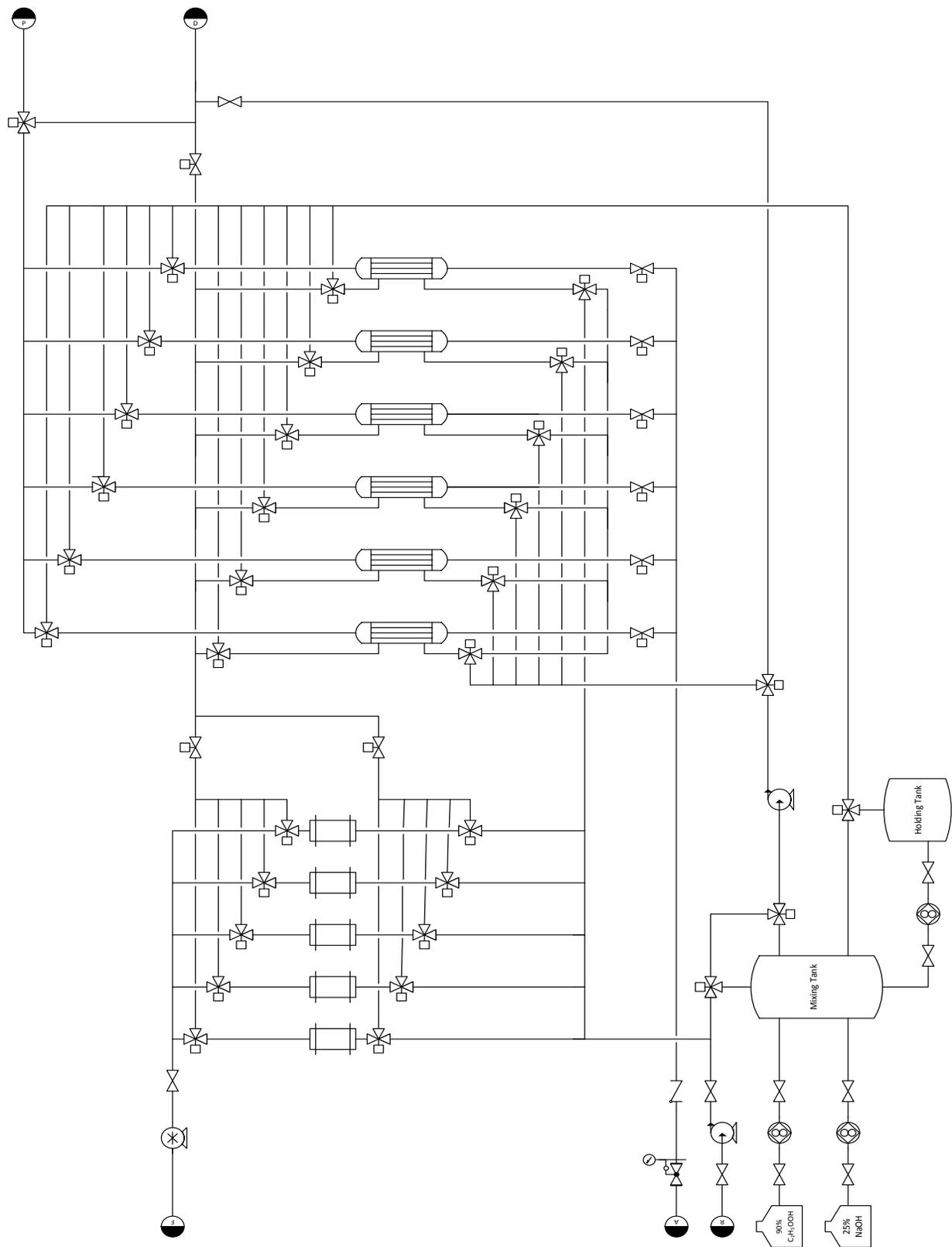
Each sand filter will be constructed from a 2 ft ID x 4 ft long pipe and will contain a 5-6 in. layer of fine sand similar to the fracking sand, 100-300  $\mu\text{m}$  particle diameter, used in the experimental apparatus. The sand filter feed pump provides a flow of 40 gpm and a maximum discharge pressure of 50 psig. The pressure drop across the sand filter train will be 30 psi when clean and 40 psi when loaded. At least four of the sand filters will be operating at all times, yielding an average pressure drop of  $\sim 8$  psi across each bed during continuous operation. The leading filter in the train is back-washed with a stream of RO permeate when its pressure drop approaches 5% of the steady state value. After back-washing, the fresh filter is positioned at the end of the train.

### **HMF Modules**

The filtrate from the sand beds enters the inlet line of the HMF circulation pump and is circulated through the shell side of the membrane modules. Each module is 2.9 in. ID x 20 in. long and recommended to operate at 8,640 gpd (6 gpm) at a 7 psi pressure drop. The HMF bank is typically operated with five modules in parallel and at 50% recovery (i.e., half of the feed permeates the membrane) with one of the modules being cleaned at any given time. When five membranes are online, permeate can be produced at a rate of 43,200 gpd (30 gpm) 144% of the flow currently delivered to the RO system. Alternatively the pumps can be throttled to reduce power requirements.

The HMF membranes require daily cleaning to remove fouling and the automated cleaning procedure is expected to be similar to that of the bench scale apparatus. The exact cleaning procedure must be determined through pilot testing. After each caustic wash, the sodium hydroxide solution is drained into the holding tank along with a small amount of RO permeate to rinse the module until the end of the acid cycle. When the acid rinse is complete, the acid solution is drained into the mixing tank and slowly mixed with the caustic solution until all of the chemical solutions have been mixed and the alkalinity has reached a minimum of pH 9. Additional acid may be required to neutralize all of the caustic solution before being discharged to the ocean. A pH meter is installed on the HMF permeate line to detect any cleaning chemicals that may enter the RO unit.

Both the sand filter train and the HMF bank must be completely automated and the required control valves and electrical components are included in the economic analysis.



**Figure 7.** Full-Scale Process Flow Schematic (F = Feed, P = Permeate, D = Discharge, A = Air, R = RO Permeate)

## **Space and Power Requirements**

The total power requirement for the system, using a 70% efficient rotary vane pump, is approximately 3 HP, which exceeds the WERC task limitation of 1.2 HP. The majority of the power is expended during the pulses in the air back-flush cycle. The power requirement could be reduced by adding more surface area through larger sand filters, more HMF units, or operating at a higher recovery in the MF units. However the first two options would require more space and investment, and the latter would increase membrane fouling. An increase in membrane fouling would then require an increased volume of chemical storage and membrane replacement cartridges, should they become permanently damaged. At higher recovery, cleaning cycles would also become more frequent and maintenance time would increase. As membrane fouling is already high, and space limitations are stringent onboard ships, the power requirement was determined to be the most acceptable constraint to exceed. The designed power usage will have a modest impact on the overall power requirement of the vessel.

Sodium hydroxide and acetic acid should be stored in fiberglass-reinforced plastic (FRP) vessels to ensure service. The sand filters are also constructed from either FRP or thin wall stainless steel to combat the corrosive nature of seawater. Wetted materials for pumps will all be compatible with handling seawater; FRP and 316L stainless steel are acceptable.

## **ECONOMIC ANALYSIS**

Table 2, below, summarizes the economic analysis for the shipboard system. The total capital investment for the full scale system is estimated based on the total purchased equipment cost.<sup>16</sup> The total purchased equipment cost is \$21,000, including spare feed and cleaning pumps. The direct cost incorporates purchased equipment delivery and installation, instrumentation and controls, piping, electrical and automation systems. Table 2 also indicates the estimated indirect costs associated with system installation. The total capital investment of The Salty Hogs' process is \$133,000.

An operating cost analysis comparing the current RO system to the proposed treated system was performed using a basis of 30,000 gpd (20 gpm) feed to the RO and the results are given in Table 3, below. From this feed, the RO unit produces 12,000 gpd of potable water. The cost analysis was performed for a three year project lifetime, the "lifetime of a reverse osmosis element under blue water feed conditions".<sup>6</sup> Currently, the pretreatment system for the RO unit

requires four hours of labor every four days to replace filter cartridges, amounting to 365 worker-hours/year. In coastal regions, the lifespan of the RO membrane was assumed to be relatively short, approximately six months. In the proposed process, the higher quality of the RO feed stream will allow for longer RO membrane service times. The HMF membrane elements are estimated to be replaced every year for harsh feed compositions.

**Table 2: Process Capital Costs<sup>15</sup>**

	<b>Basis</b>	<b>Cost</b>
<b>Equipment Costs</b>		
Feed Pump (2)	Manufacturer	\$2,000
Peristaltic Pump (3)	Manufacturer	\$1,500
Sand Filters (5)	Manufacturer	\$10,000
Hollow Fiber MF Membranes (6)	Manufacturer	\$2,000
Sodium Hydroxide Storage Tank	Manufacturer	\$2,000
Acetic Acid Storage Tank	Manufacturer	\$2,000
Holding Tank	Manufacturer	\$500
Mixing Tank	Manufacturer	\$1,000
<b>Total Equipment Costs</b>		<b>\$21,000</b>
<b>Direct Costs</b>		
Purchased Equipment Cost		\$21,000
Purchased Equipment Delivery	10% of Purchased Equipment Cost	\$2,100
Purchased Equipment Installation	47% of Purchased Equipment Cost	\$9,900
Instrumentation and Controls	36% of Purchased Equipment Cost	\$7,600
Piping	75% of Purchased Equipment Cost	\$15,800
Electrical Plus Automation Systems	100% of Purchased Equipment Cost	\$21,000
<b>Total Direct Plant Costs</b>		<b>\$77,400</b>
<b>Indirect Costs</b>		
Engineering and Supervision	100% of Purchased Equipment Cost	\$21,000
Construction Expenses	41% of Purchased Equipment Cost	\$8,600
Legal Expenses	4% of Purchased Equipment Cost	\$800
Contractor's Fee	22% of Purchased Equipment Cost	\$4,600
Contingency	40% of Purchased Equipment Cost	\$8,400
<b>Total Indirect Plant Cost</b>		<b>\$43,400</b>
<b>Fixed Capital Investment</b>	<b>Sum of Direct and Indirect Costs</b>	<b>\$121,000</b>
<b>Working Capital</b>	<b>10% of Fixed Capital Investment</b>	<b>\$12,100</b>
<b>Total Capital Investment</b>	<b>Sum of Fixed and Working Costs</b>	<b>\$133,100</b>

**Table 3. Three Year Operating Cost Comparison**

	<b>Basis</b>	<b>Cost</b>
Hollow Fiber MF Membranes	Manufacturer	\$6,000
Sodium Hydroxide (25%)	Manufacturer	\$500
Acetic Acid (Glacial)	Manufacturer	\$500
Power Required for Compressed Air	\$0.135/kWh	\$1,500
Power Requirements: Pump/Controls	\$0.135/kWh	\$3,000
Replacement Media		Nil
Reverse Osmosis Membranes	Manufacturer	\$6,000
Operating Labor	183 hour/year at \$25/hour	\$13,725
<b>Total Costs per 3 Years</b>		<b>\$31,000</b>

	<b>Basis</b>	<b>Cost</b>
20 Micron Cartridge Filter	Manufacturer	\$29,700
5 Micron Cartridge Filter	Manufacturer	\$29,700
Reverse Osmosis Membranes	Manufacturer	\$36,000
Operating Labor	365 hours/year at \$25/hour	\$27,600
<b>Total Costs per 3 Years</b>		<b>\$123,000</b>

The overall maintenance of the system is low because of automation and decreased component replacement frequency. The cost of labor was calculated based on man-hours required for system operation. Other power cost differences between the two systems are negligible. In total, The Salty Hogs’ pretreatment design will save the U.S. government \$92,000 in operating costs over a three year period. If RO membranes must be replaced more than once per year in coastal waters with the current system, the cost benefits of The Salty Hogs’ design will be even greater.

### **INLAND DESALINATION APPLICATION**

Desalination units, though primarily thought of for their usage in coastal areas, can also be used inland in brackish water applications, providing fresh water to regions that lack adequate groundwater. A major concern with any desalination unit used in inland applications is the production of a concentrated salt stream from the RO facility. In brackish water RO systems, the retentate stream is 15% of the feed flow rates, while in seawater application it is 50%. Currently, the retentate can be disposed of by “surface water discharge, sewer discharge, deep well injection, evaporation ponds, infiltration basins and irrigation”.<sup>1</sup> The two most common methods, surface water discharge and sewer discharge, cannot be used in inland applications because it will lead to contamination of the fresh groundwater. Instead of disposing of the

concentrated solution, recycling the concentrated brine for industrial processes or use in energy generation from solar ponds has been considered.<sup>17</sup>

## **REGULATIONS**

### **Environmental Considerations**

The Clean Water Act (CWA) was enacted in 1977 by the Oceans and Coastal Protection Division of the Environmental Protection Agency (EPA). The CWA established a basis for the regulation of pollutant discharge into the waters of the United States and quality standards for ground water. Under the National Pollutant Discharge Elimination System, vessel discharges incidental to the normal operation of a vessel into the waters of the United States are required to submit a notice of intent if the vessel is over 300 gross tons. A Vessel General Permit (VGP) is required for such vessels and is applicable to 26 discharge streams, which includes a concentrated sea water stream resulting from a desalination process. According to the VGP guidelines, brine from any reject water “shall not contain or come in contact with machinery or industrial equipment (other than that necessary for the production of potable water), toxic or hazardous materials, or wastes”.<sup>18</sup> Additional state regulations may apply. For example, Hawaii bans the formation of objectionable sludge or bottom deposits. Although vessels of the Armed forces are currently exempt from obtaining the VGP, our system will abide by the regulations declared in the CWA.

In 1996, the National Defense Authorization Act amended section 312(n) of the CWA. Through this amendment the EPA and the Department of Defense (DOD) were required to develop Uniform National Discharge Standards (UNDS) for any discharges incidental to the normal operation of a vessel of the Armed Forces. UNDS is in the process of being implemented under three phases. The DOD is currently in the second phase. In the first phase, it was determined that, of the 39 discharge streams, only 25 will require control. Once the third phase is complete, all discharges of proposed systems must meet the UNDS regulations. Under Phase I, it was found that discharge from the RO system has the potential to cause adverse environmental effects. These adverse effects were due to high metal concentration in the discharge above the water quality criteria. In the designed process, consisting of a 0.1 µm hollow fiber membrane, the metal concentration in the system discharge will be unchanged because they will permeate the MF membrane.

## **Chemical Considerations**

Excess sodium hydroxide from the chemical cleaning cycle needs to be completely neutralized before being discharged to the sea. The acetic acid from the second part of the cleaning cycle will be used to neutralize the sodium hydroxide; the sodium acetate produced is environmentally benign. Sodium hydroxide, although commercially available in a 50 wt% solution, must be diluted to 25 wt% to prevent freezing.<sup>19</sup> Glacial acetic acid cannot be used due to similar problems as the sodium hydroxide and must be diluted to at least 90 wt% to prevent crystallization.<sup>20</sup> The concentrated solutions allow the storage volume to be relatively small. For six months at sea, 0.2 tons of glacial acetic acid and 0.42 tons of 25 wt% sodium hydroxide are needed. Before the cleaning solution can be safely discharged, it must be titrated to a maximum pH of 9. This is done in an effort to protect the oceanic environment from harmful chemicals.

## **Worker Safety**

While prevention and training are the primary safety measures for the proposed system, additional precautions should be considered. Care should be taken while maintaining the system as it has been designed for operating pressures of about 80 psig. Workers should be informed of all electrical dangers associated with the system as it is fully automated. The cleaning agents, 25 wt% sodium hydroxide, a strong base and reducing agent, and 90 wt% acetic acid, an oxidizer, can be hazardous to human health at the operating concentrations, so proper Personal Protective Equipment must be worn during chemical handling. These include (1) chemical gloves, (2) eye protection, (3) chemical protective suit, (4) chemical resistant boots (with pant legs outside the boots), (5) transparent face shield, and (6) hard hat, all conforming to OSHA standards. All personnel must be trained on the hazards associated with each chemical as per OSHA standards. Material Safety Data Sheets for each chemical must be aboard the vessel. In order to prevent injury to personnel and damage to equipment, the lock-out and tag-out programs should be implemented for all equipment, components, and systems (mandatory aboard USN ships). MIL-STD-882 provides protocol for implementing a safety program for the life of the system if used on military vessels.

## **Additional Regulations**

Any pretreatment designs considered for installation aboard sea vessels are required to conform to standards unique to shipboard operation. Design considerations include being electromagnetically compatible with surrounding equipment and able to withstand certain vibrations and shock loading. These mechanical requirements for the installed system are categorized in MIL-S-901D, MIL-STD-167-1A, and MIL-STD-461. Additionally, the system must not exceed the noise requirements (84 dB TWA, OPNAVINST 5090) and be constructed from materials that, in the case of a fire, will not create toxic fumes in a confined space (MIL-STD-2031).<sup>1</sup> Militarization of the pretreatment process involves removal or enclosure of all plastic parts and the removal of pipe-threaded parts.<sup>1</sup> The full-scale system is easily applicable and adaptable to such standards.

## **CONCLUSIONS AND RECOMMENDATIONS**

1. The Salty Hogs team has determined that sand filtration/HMF is the most economically viable green pretreatment method for seawater RO. The process removes particles to 0.1  $\mu\text{m}$  to minimize the fouling potential of the RO feed stream and prolong the life of downstream RO membranes.
2. The combination of fluidized sand filters and air-back-flushed HMF membranes eliminates the need for replacement cartridges, decreases maintenance costs, and minimizes unit down time.
3. Automated back-washing and cleaning protocols allow operation of the sand filters for at least a year and maintain high permeation rates through the HMF membranes. The performance of the sand filter does not deteriorate over time and the experimental data has shown that the HMF unit reaches an asymptotic flux after several hours.
4. The unit requires approximately 250  $\text{ft}^3$  for operation and 40  $\text{ft}^3$  of storage space for six months at sea, which is below the Task 5 specifications of 300  $\text{ft}^3$  and 100  $\text{ft}^3$ , respectively.
5. The total capital investment for The Salty Hogs' process is \$133,000. The operating cost per three-year period is \$31,000, a savings of \$92,000 over the existing system during the same time period.

6. It is recommended that long term (90 days) pilot plant testing be conducted before the implementation of this process onboard ships. Tests will confirm projected service times and can address the optimization of back-wash and cleaning cycles. The use of actual seawater, instead of a surrogate, will more closely approximate the conditions encountered at sea.
7. Before constructing any inland desalination units, pilot plant testing should be conducted to fine tune the process to the specific feed conditions in that area. Altering the back-washing schedule for less turbid feeds may lead to significantly increased permeate production through reduced downtime.
8. The Salty Hogs' system will have a peak power requirement of approximately 3 HP.

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## **APPENDIX A - Audits**

Date: March 12, 2012

The Salty Hogs

Department of Chemical Engineering

University of Arkansas

Subject: Review and comments on WERC 2012 Paper

Dear Salty Hogs Team,

I have reviewed your paper submitted to the WERC 2012 competition entitled Green RO Pretreatment. My understanding of the project rules is that you needed to have the paper reviewed by a professional in the water field. Additionally you requested specific comments with regards to the legal issues and regulations the proposed pretreatment design may have to conform with.

Comments on Regulations Section-Legal Issues

### 1. Environmental Considerations

I would recommend listing the specific metal and estimated concentrations from the RO system which may be in violation of the UNDS standards. The metal concentration in the discharge would probably be due more to corrosion of the piping systems from the seawater. Test experience and lab results from our military water purification systems in Port Hueneme shows leaching from the copper nickel piping the primary source of exceeding the environmental limits. The copper discharge limits into seawater are very low due to the impact on marine life and this is typically the primary metal which causes environmental discharge issues. I would also recommend looking at the wetted materials in the proposed design and try to estimate the potential discharges concentrations. This could be estimated by assuming a corrosion or degradation rate of the material.

### 2. Chemical Considerations

Your paper indicated the laboratory hollow fiber membrane was cleaned with a solution of sodium hydroxide and flushed with DI water. I would agree with the statements regarding neutralizing the NaOH and adjusting the pH level below 9. Circulation of the cleaning solution through the membranes would improve cleaning effectiveness versus a static soak and flush.

However my experience with cleaning hollow fiber microfilters and ultrafilter membranes fouled by seawater is somewhat different. A two step process involving a low pH cleaner (typically citric acid) followed by a high pH cleaner (chlorine, detergent ) is typically used. Sodium hydroxide is sometimes used to increase the pH level of the cleaning solution to around 12 which is shown to be effective against removing biofouling deposits. The low pH solution can be effective at removing foulants caused by metals such as iron. The cleaning solutions are typically mixed using RO product water and the temperature of the solution is usually between 90 to 100 F. A complete drain and flush of the tanks and piping with RO product water would be conducted between the low and high pH cleaners to avoid unintended chemical reactions. If seawater is used to prepare the cleaning solutions then it is important to determine any potential interactions with the chemical solutions prior to proceeding. The cleaning effectiveness is directly proportional to the Ct (concentration of chemical x time of exposure) value. For the proposed shipboard application I would recommend the cleaning solutions be directed to waste storage tanks that can be disposed of upon return to shore.

Each membrane manufacturer typically develops their own specific cleaning protocol with the acceptable chemicals (citric acid, bleach, detergent), chemical concentrations (percent weight or percent volume, X mg/L etc) , exposure time (30 min), temperature limits (don't exceed 105 F) and recirculation flow rates to warranty their membranes.

The material of construction used in the hollow fiber membrane is what determines the chemical compatibility. For example chlorine bleach has been shown to be a very effective cleaner in PVDF membranes while causing degradation in polypropylene membranes. Additionally the implementation of a chemical cleaning would be the result of a maximum transmembrane pressure (TMP) being reached. For example the operators could be instructed to conduct a chemical cleaning once the TMP reached 20 psi.

### 3. Worker Safety

I would recommend having proper lab equipment for measuring and dispensing the chemical solutions. This could include an accurate scale for dry chemicals or volumetric flask or pipet for liquid solutions. Discharge of the chemical solutions should be through dedicated piping to avoid inadvertent discharge to sea. This could be done with automated or manually activated valves.

#### General Comments on Paper

- 1) The opening sentence of the executive summary is cumbersome. I would recommend rewriting and avoid repeating terms.
- 2) I would provide the reader with more specific descriptions of what each task was and what your group is trying to achieve. You mention task 5 but it may be helpful to understand the overall context of the project. Also define what WERC stands for.
- 3) Under the discussion on self cleaning filters you may want to consider the buildup over time of marine organisms such as barnacles on the filter housing and cleaning mechanisms. These have to be manually scrapped off. Also shells, seaweed, and other debris from the ocean can be trapped by these filters requiring manual intervention to clean.
- 4) I would recommend listing the design loading rate on your laboratory sand filter. Based on the dimensions listed of 4 inch diameter and flowrate of 0.5 gpm I calculated a media filter loading rate of about 6.25 gpm/ft<sup>2</sup> which is fairly low. By comparison the loading rate on the media filter used on military 600 GPH ROWPU system is about 7.6 gpm/ft<sup>2</sup>. So your design is about 20% lower which should translate into longer run times between backwashes.
- 5) Your paper indicates the sand filter was backwashed when the pressure drop was between 35 to 45 psig. This may be too high in certain cases since multimedia filters are usually backwashed when the pressure drop reaches 20 psi.
- 6) Under the laboratory experimental results section you indicated the best time interval for backflushes was every 5 minutes for 100 seconds. My experience with microfilter and ultrafilter membranes is the minimum backwash time is 15 minutes. Your approach would probably require an intermediate holding tank for the MF/UF permeate while the system backwashed.
- 7) I would suggest listing the manufacturer, material construction, flow path (out/in or in/out), dimensions, housing requirements, pore size, fiber diameters of the hollow fiber UF membrane used in the experiments and proposed full scale design.

## Background and Experience

I have worked as a mechanical engineer for the NAVFAC Engineering Service Center (NAVFAC ESC) since 1991 working primarily on desalination and water purification issues. Over the past 10 years I have supported the Office of Naval Research (ONR) and Naval Sea Systems Command (NAVSEA) on various research efforts focused on improving shipboard desalination. Additionally I serve as a subject matter expert on water purification issues for the Navy and Marine Corps activities.

I currently manage and oversee testing operations at the Navy's Seawater Desalination Test Facility in Port Hueneme, CA where several pretreatment, reverse osmosis and system tests have been conducted for the military and private sector. I have B.S. degree in Mechanical Engineering from Cal Poly State University, San Luis Obispo and a professional engineer license from the State of California.

If there are any further questions or comments regarding my evaluation I can be reached at 805-982-6640 or email: [William.varnava@navy.mil](mailto:William.varnava@navy.mil).

Sincerely yours,

Bill Varnava, PE

NAVFAC ESC

Code EX32

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**APTwater, Inc.**

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DATE: 12 March 2012

TO: Elizabeth Fullerton

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FROM: Erick Neuman

MESSAGE:

**RE: Safety Review – Salty Hogs WERC 2012 Project Green RO Pretreatment**

Thank you for the opportunity to review your project in reference to the safety of operating and maintaining the full scale process. I provide to you my comments in the following paragraphs.

**Overview**

It is encouraging to follow your efforts to develop a green RO pretreatment system. Your considerations of other technologies and bench scale program were impressive as they related to your project objectives.

In implementing and maintaining a new pretreatment system three aspects are important

to establish and preserve safe working conditions. They are: design/equipment selection, operator interfaces, and training/written procedures.

I preface all of my comments that they are provided to add additional guidance to your work. I recognize a lot of great work has gone into this design project. I do not want my comments to detract from all the positive aspects of the work that has been done. These comments are meant to supplement the effort to ensure the key aspects are addressed. There are not a lot of comments that can be provided at this stage and I do not want to seem that they are overly critical.

### **Design/Equipment Selection**

Based on the full-scale process flow diagram (Figure 7), there is limited detailed information on the type of equipment that will be utilized for the marine service. Materials of construction are properly considered and I suggest further work can be completed on the types of controls and process equipment. It is important that the equipment can withstand the requirements of a marine environment. Failures of equipment at sea can result in operators taking shortcuts to maintain the system in an operating condition. These shortcuts can result in damage to the system and possible injury.

Consideration should be given to the need for and location of pressure relief devices. There are several vessels that could be pressurized (not under normal operation).

Protection should be added to contain the possible hazards related to the use of sulfuric acid and sodium hydroxide. Operator handling of these chemicals must be minimized. Consideration should be made to make sure that the raw chemicals do not mix or contact each other. These types of details are not normally presented at the process flow diagram level. It is important to make sure that the raw chemicals do not come in contact with any process equipment downstream of the mixing tank.

### **Operator Interfaces**

The project has correctly presented the need for automated processes and PLC control. Sulfuric acid and sodium hydroxide are dangerous chemicals that should only be handled with the proper safety procedures. The project correctly presents the need for proper PPE when handling these chemicals. Additional care must be given to the equipment and piping utilized to contain and transport these chemicals. Any failures will risk injury to the operator and other equipment that comes in contact with these chemicals.

Automation must limit the possibility of operator error. Again at the level of detail of the process flow diagram it is not normal to describe the automation philosophy but must be included in the more detailed tasks.

### **Training/Written Procedures**

It should be recognized that naval personnel have limited background in the operation and maintenance of these types of processes. In addition, while at sea there is limited opportunity for the support by the manufacturer. The operators must be training to a sufficient level that they can effectively troubleshoot possible situations.

This training can be provided through operating guidance (manuals and procedures) and effective training programs. Again at the level of process flow diagrams and process descriptions, this level of detail is not expected but must be considered as the project details are developed.

I hope these comments are helpful. Best regards.

Audit from Paul Gallagher of Siemens Industry, Inc. received March 14, 2012.  
Email: paul.m.gallagher@siemens.com  
Phone: (978) 614-7300

My comments are focused on the economic analysis pp19-20 of the report and on other parts of the document that are closely connected to process economics.

### **Economic Analysis:**

While working with a novel design concept for which cost data are largely absent in the published literature, the team has done a good job of preparing budgetary cost estimates for both capital costs and operating costs.

The team used escalation factors to derive complete installed costs from what appear to be budgetary estimates of proprietary equipment costs provided by equipment vendors. This is a valid approach for initial budget costs for a variety of project scenarios. I am concerned that since equipment of the scale being considered is normally sold as fully built up skids, the value provided by the equipment supplier has been underestimated while that supplied by the installing contractor has been overestimated. Specific costs incurred to render equipment suitable for the naval ship environment do not seem to have been considered. Those costs can be a substantial fraction of the system cost. I did not see a direct comparison to the capital cost of the currently installed solution. One imagines that the systems now in use would have lower capital cost than the new system but that this difference would be more than compensated for in operating cost savings over the lifecycle of the systems.

The operating costs appear to have been investigated in considerable detail. The analysis of the proposed system contains membrane replacement, cleaning chemical, energy, replacement media and operating labor costs. I am a little surprised that the operating costs are considered for only a three year period. I imagine that the proposed equipment would have a service life of ten years or more, and then a full lifecycle cost analysis covering initial capital cost plus ten years of operating costs would have been instructive. The updated version of Table 4 shows three year operating cost saving of \$92k for the proposed system (\$123k less \$31k). It is unclear why the text below the table refers to a three year saving of \$14k.

If time permits, I would recommend presenting the economic analysis on a lifecycle cost basis. That would require an estimate of the cartridge filter assembly installed cost and also a discounted cash flow analysis of operating costs over the equipment lifetime (say 10 years). It is likely that such an analysis would demonstrate a significant benefit for the proposed solution.

**Other Technical Comments:** The team demonstrates considerable knowledge about the unit operations involved in the desalination system. There are a few items that I would ask the team to reconsider. The report states that SDI of microfiltration filtrate was not measured in the feasibility tests since water filtered to 0.1micron could not foul a 0.45 micron filter. In general, this is not a good assumption since dissolved or fine colloidal organic material can foul a 0.45

micron filter and since the 0.1 micron filter must be shown to be integral (this cannot be assumed). The team also discounts a higher than expected filtrate turbidity as being related to the type of turbidimeter used. Again, I would advise caution in reaching this conclusion since several other explanations are possible, including leakage across the microfilter, failure of seals in the MF system, precipitation in the filtrate, contamination of sample vessels.

The team has considered several of the applicable safety regulations and has identified some important hazards related to the proposed equipment. In particular, risks associated with high pressures and with chemical hazards associated with sodium hydroxide and sulfuric acid are discussed in some detail. The team could further develop this analysis for example by discussing how equipment design features could be included (automatic Clean-in-Place is one example) that would greatly reduce the need for chemical handling. The compressed air backwash raises some specific concerns since certain plastics such as PVC and CPVC are prone to shattering. It would be appropriate to mention that such materials should be avoided for any lines that may become filled with compressed air.

In summary, I am impressed with the level of detailed understanding displayed by the team, particularly when taking into account the limited availability of directly relevant published information. Their design concept is very likely a significant improvement over the currently deployed system design with the bulk of this advantage lying in lower operating costs. The team has identified that their system's advantage is low operating cost and they have analysed this in some detail. While the presentation of the operating cost analysis could be refined, the essential point is clear. The team has done a good job of identifying safety hazards related to their design and of proposing mitigation measures for those hazards.