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## Multimedia and Ultrafiltration for Reverse Osmosis Pretreatment Aboard Naval Vessels

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# MULTIMEDIA AND ULTRAFILTRATION FOR REVERSE OSMOSIS PRETREATMENT ABOARD NAVAL VESSELS

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## Abstract

*The US Navy is interested in improving the pretreatment for shipboard Reverse Osmosis potable water systems. To investigate this problem, the Navy prepared a Task for the 2010 WERC (<http://www.werc.net>) competition. The research described in this paper was performed to compete in this WERC Task and ultimately received a 1st place award. Several technologies were considered as options for improving the current pretreatment process. Multimedia filtration followed by ultrafiltration was chosen as the most economical solution. This paper presents the final design for a full-scale shipboard system that incorporates multimedia filtration and ultrafiltration yet requires minimal space and optimal power usage. Although the work focuses on oceanic applications, inland water desalination plants can economically use the technology. The proposed process provides an annual incremental savings, including reduced maintenance and filter usage, of \$17,000 over the current process, resulting in a payback period of 14 years. However, as savings are highly dependent on the replacement frequency of all components, ships operating in coastal waters may realize greater savings of up to \$47,000 annually, thus reducing the payback period to 5 years.*

## Introduction

Production of fresh water at sea poses a significant problem for oceangoing vessels. In the past 30 years, reverse osmosis (RO) has replaced distillation as the preferred means of large-scale potable water production on ships. Since 1988, RO desalination systems have been used by the United States Navy for the production of potable water. The Navy Standard Reverse Osmosis (NSRO) plant is the standard for these systems, and by 2009, 268 NSRO plants had been installed aboard US Navy ships.<sup>1</sup> In NSRO plants, seawater first enters a centrifugal separator (hydroclone style) which removes approximately 95% of particles over 50  $\mu\text{m}$ .<sup>2</sup> Smaller particulates are removed by a series of cartridge filters (20  $\mu\text{m}$  followed by 3  $\mu\text{m}$ ), and the pretreated water is then desalinated using spiral-wound RO membranes. In open waters, the cartridge filters can last 4-6 weeks before requiring replacement, and RO elements last for 3-5 years.<sup>1</sup> However, in littoral and coastal waters, which are characterized by increased levels of suspended solids, RO membranes can fail in just a few months, and cartridge filters often last less than 10 hours. Therefore, advanced pretreatment technology is needed to reduce maintenance time and decrease storage area requirements while also providing a higher quality feed to the RO unit. High-quality feed water is commonly defined as having a silt density index (SDI) of less than 3 and a turbidity of

less than 1 NTU (nephelometric turbidity unit).<sup>2</sup>

To encourage improvements in the pretreatment of shipboard Reverse Osmosis potable water systems, the U.S. Navy prepared the Task for the 2010 WERC (<http://www.werc.net>) competition. The problem statement in this task was, "Develop and demonstrate an alternative to disposable filters or an improved disposable filter design that can last at least four months filtering somewhat turbid feed water. To increase the life of the RO membranes, ideally your proposed process will address particle size down to 0.1 microns as there are suspended solids that do pass through the current cartridge filters and foul the RO membranes." This paper describes the research of a team of chemical engineering students who responded to the 2010 WERC Task. The team's design premises for the WERC Task were as follows:

1. All particulates above 0.1 microns are to be removed.
  2. Energy usage must be less than 10% of that used for the RO system.
- Note: The calculated RO system energy requirement is 12 HP, with an energy recovery device that is 50% efficient (from ROPRO software with RO recovery of 40%).
3. 30,000 gal/day of filtered seawater is to be supplied to the RO system.
  4. The operating unit should occupy no more than 300 ft<sup>3</sup> (3 times current cartridge system).
  5. The system should be green (i.e., chemical discharge is minimized).
  6. Maintenance time must be less than that for the current system (4 hours every 4 days).
  7. Storage space for a 9-month time frame must be less than 100 ft<sup>3</sup> (1/10 of current system).
  8. The system must be designed to handle a feed of the following composition:
    - A. 32,000 ppm of sea salt
    - B. 75 mg/L of Klamath Blue-Green Algae powder
    - C. 20 mg/L of Orchid Pro (by Turf Pro USA), an orchid fertilizer liquid product
    - D. Dechlorinated tap water

A survey of the literature was conducted to identify many variations of filtering technologies which could be used to remove particles down to 0.1  $\mu\text{m}$ . However, like RO membranes, most fil-

ters are prone to fouling under turbid feed conditions. Prefiltration for these modules is required because subjecting a membrane to biological colloids increases the rate at which the transmembrane pressure increases.<sup>3</sup> Therefore, applicable filtration technologies were broken into two categories: (1) prefiltration and (2) micro- or ultrafiltration.

Prefiltration, for the purposes of this report, refers to technologies which remove particles down to 1  $\mu\text{m}$ . Technologies considered to replace the current cartridge depth filters included centrifugal separators as well as coagulation, electrocoagulation, and media filters.

Hydroclones, such as the centrifugal separators currently in use by the US Navy, can separate 50  $\mu\text{m}$  particles with a specific gravity of 1.5 or greater. Hydroclones were rejected because they can only remove particles down to 50  $\mu\text{m}$  and increased solids concentrations in ultrafiltration membrane feeds lead to a decrease in permeate flux.<sup>3</sup>

Chemical coagulation, or flocculation, is the dominant pretreatment for municipal water systems. Chemicals such as ferric chloride and ferric sulfate are added to the feed water to eliminate charges causing mutual repulsion between small particles. Moderate agitation then causes particles to flocculate. The flocs are separated by gravity or filtration. Chemical coagulation was rejected for the task solution due to disposal considerations for the sludge resulting from separating the flocs.

Electrocoagulation, where an electric potential applied between two electrodes causes floccing, has some advantages over traditional methods. Chemicals are not required; thus, their disposal and storage are not issues. Testing has shown that this method has better removal of some species than chemical coagulation.<sup>4</sup> The salt concentration in seawater reduces the voltage requirement; however, energy requirements for electrocoagulation will far exceed those allowed by the premises of this task.

Single-media filters consist of a thin bed of granular material, commonly sand or diatomaceous earth. The filtered particles are trapped in the first few inches of the bed, and water and smaller particles pass through the bed. As the bed is loaded with filtered solids, the bed pressure drop will increase to an uneconomical level, requiring backwashing for regeneration. Single-media filters perform well for a narrow particle size distribution but are not efficient for this system, which involves contaminants ranging in size from 0.1 to 500  $\mu\text{m}$ .

Multimedia filters use layers of different media to trap particles in all active layers of the bed. The media are arranged with the largest and least dense particles on top and the smallest, most dense particles on bottom, supported by an inactive layer of larger gravel. This arrangement allows for solids loading throughout the bed. After backwashing, the variation in media densities and sizes results in the bed's settling into its original layers. Additionally, the high total surface area of multimedia filters removes a greater number of particles and produces a higher quality filtrate.<sup>6</sup> Multimedia filters do not function by simply capturing particles in the voids between media particles but instead attract particulates to the surfaces of media particles by diffusion, sedimentation, and interception.<sup>5</sup> Depending on the media used, multimedia filters

will remove particulates down to 5-10  $\mu\text{m}$ . Typical flux rates for multimedia filters are between 5 and 14  $\text{gpm}/\text{ft}^2$  of the bed cross-sectional area.<sup>6</sup>

Thus, multimedia filtration was selected as the prefiltration step because it is the most economical when considering capital, energy, maintenance, and space requirements. Its high throughput and long lifetime are also advantages. Through backwashing, the performance of multimedia filters can be recovered completely after an increase in trans-filter pressure during a filtration cycle. Additionally, the media has a lifetime of several years, and replacement storage onboard is unnecessary. The media selected were anthracite coal, sand, and garnet, supported by a bed of gravel. This filter will remove particulates down to 10-20 $\mu\text{m}$ .<sup>7</sup>

In order to protect the RO membrane and filter to the 0.1  $\mu\text{m}$  level prescribed in the Task problem statement, microfiltration or ultrafiltration was needed. This was accomplished by using membranes constructed of either polymeric or ceramic materials. Advantages of ceramic membranes include the ability to withstand high temperatures and pressures and all pH values. However, desalination operations operate at low temperatures and pressures and almost neutral pH values, precluding the need for thermally and chemically robust ceramics. Moreover, there are several significant disadvantages of ceramic membranes with respect to shipboard water treatment. Although the flux rate through ceramic membranes is high, about 100 GFD ( $\text{gal}/\text{ft}^2/\text{day}$ ), the footprint needed for the surface area is far too large to include onboard a ship as the channel diameters are limited to about 1 cm. Ceramic membranes are heavy and expensive, though the high initial cost is compensated for by a long lifespan.<sup>8</sup>

Membranes are also made from a variety of natural and synthetic polymers formed into several different shapes, including flat sheets, tubes, and hollow fibers, depending on the properties of the polymer. These are then grouped into modules: the fibers and tubes are connected in bundles, and the flat sheets are wrapped into a spiral or left flat.

Flat sheets of membranes may be arranged into a plate and frame module in which the membranes are supported by plates which channel the feed, permeate, and concentrate. Flat sheets can also be used in a stacked disk design. The flow path takes the feed stream along the flat membranes. Flat membranes were rejected based on the system volume requirement.

Spiral-wound membranes offer a relatively large surface area for the volume required. This is the configuration of the RO membranes currently used aboard naval vessels. Backflushing of spiral-wound modules is not an option as the pressure would separate the layers of the membrane and destroy it.

Tubular modules, with tube diameters in the range of 5-15 mm, are fed on the tube side, allowing permeate to pass through the membrane forming the walls of the tubes. Hollow fibers function in the same manner but have smaller inner diameters, from 0.5-1.2 mm, and thus a much smaller volume for a given membrane surface area. Feed flow can be on either the tube or the shell side, which allows the modules to be backflushed to remove foulants. Tube-side flow is most advantageous as it allows a greater portion of the flow entering the unit to flow uniformly over the

membrane surface.

Of the various configurations of UF membrane modules, hollow fiber units were selected because they can be backflushed and provide high flux rates and a large membrane surface area per unit module volume. Although microfiltration is capable of filtering to  $0.1\ \mu\text{m}$ , an ultrafiltration (UF) module which filters to a 50 kDa MWCO (molecular weight cutoff) (equivalent to  $\sim 0.005\ \mu\text{m}$ ) was selected instead. The smaller pores will be less susceptible to plugging, making cleaning easier and increasing membrane life.<sup>9</sup> This enhanced filtration will also further protect the downstream RO membrane from fouling and increase its service time.

Finally, several naval and recreational vessels utilize RO units with an open-channel design between stacked discs of membranes instead of spiral-wound modules. This design minimizes pretreatment requirements by decreasing the tendency of particles to become trapped on the surface of the RO membranes. Typical pretreatment for these modules requires only cartridge or sand filters to reduce the SDI (SDI 5) of RO feed water to 15-20. Manufacturers report membrane service times of up to 5 years.<sup>10</sup> Open-channel modules appear attractive for extending RO membrane run times and should be considered for new or replacement RO units; however, the existing RO units need improved feed filtration for longer service times and improved performance.

### Bench Scale Apparatus

The multimedia filtration system consists of a feed tank, feed pump, backwashable multimedia filter, and filtrate tank. A photograph of the multimedia filter experimental unit is presented as Figure 1. The flow schematic for the multimedia experimental apparatus is presented as Figure 2. Two multimedia filters were tested. The small filter was constructed of clear 2.445" ID x 5.3' long PVC pipe, and the large filter was constructed of clear 5.625" ID x 4.5' long Plexiglas pipe. Each multimedia filter consists of four layers of media of differing sizes and densities. From the top down, the bed is composed of a 12" layer of anthracite coal (sp. gr. 1.6;  $d_p = 0.6\text{-}2\ \text{mm}$ ), a 10" layer of sand (sp. gr. 2.5;  $d_p = 0.6\text{-}1\ \text{mm}$ ), and an 8" layer of garnet (sp. gr. 4;  $d_p = 0.32\text{-}0.39\ \text{mm}$ ). The lowest bed, which provides support to the upper layers, is a 9" layer of graded pea gravel (sp. gr. 2.65;  $d_p = 2\text{-}5\ \text{mm}$ ). Filtrate quality is a function of feed quality, retention time in the filter, media characteristics, and bed depth; therefore, the units were designed with the same depths and types of media, and results were compared only at the same fluxes.

The feed, mixed to WERC specifications (32,000 ppm of sea salt, 75 mg/L of Klamath Blue-Green Algae powder, 20 mg/L of Orchid Pro), was fed to the multimedia filter using a variable speed rotary vane pump which delivered 0.08-1.5 gpm at a feed pressure of 25-35 psig. Feed entered at the top of the filter, and increasingly smaller particles were trapped as the feed traveled through the filter, giving effective solids loading throughout the bed. The filter effluent was collected in a 5-gallon pail. During backwash, flow direction was reversed. To start the backwash cycle, the pump was stopped, and the feed hose was moved to the filtrate tank. Flow of filtered water was distributed by the gravel underbed, and the filtration media were fluidized, washing out accumulated sediment. Backwash liquid was drained to a waste tank. Following backwash, normal forward operation was resumed.

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Figure 1. Photograph of Multimedia Filter

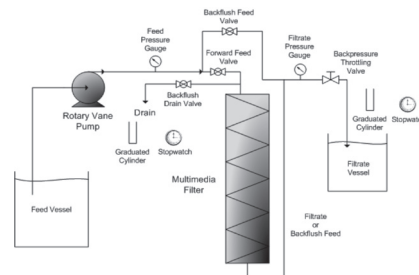


Figure 2. Multimedia Flow Schematic

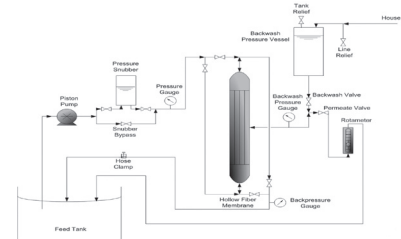


Figure 3. Hollow Fiber Membrane Flow Schematic

The effluent from the multimedia filter was sent to the Ultrafiltration Unit, which was a 1 ft<sup>2</sup> hollow fiber UF membrane (Koch HF, 1018-1.0-43-PM50) with a 50 kDa MWCO, as shown in Figure 3. The purpose of this HF unit was to filter particles larger than  $0.01\ \mu\text{m}$  from the feed stream by forcing a permeate stream through the tube wall and into the shell side of the module. Permeate exited the module shell-side and flowed through a rotameter to a permeate receiver. The exiting tube-side stream, commonly identified as the concentrate, was recycled back to the feed tank to minimize the amount of feed required for extended tests. A pressure vessel filled with permeate from previous tests was kept at 20 psig using compressed air. During backflushing, a stream of pressurized permeate was delivered to the shell side of the fibers.

A variable speed piston pump delivered feed to the membrane at 25 psig. The feed pump speed and concentrate flow rate were manually controlled to maintain this feed pressure, which fixed the permeate rate for a given level of fouling. The tube-side pressure drop was about 5 psig. The membrane fibers were cleaned by backflushing and reversing the tube-side flow direction. Backflushing and flow reversal occurred every 2-3 minutes to maintain an economically high permeation rate.

The multimedia and HF units were combined into a single automated bench scale apparatus for final experimentation and for the competition. A photograph of this apparatus is displayed in Figure 4. This setup is similar to that of the full-scale design,



where multimedia filtrate flows directly into the HF unit.

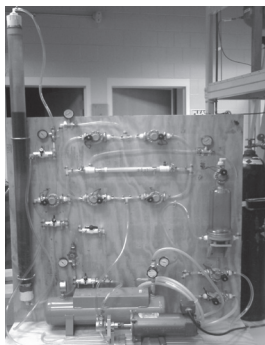


Figure 4. The Entire Experimental Apparatus

### Laboratory Experimentation

Laboratory experiments focused on two fronts: (1) determination of the pressure drop, solids loading characteristics, and filtration performance of the multimedia filter; (2) determination of the permeate flux and required backflushing frequency of the HF unit. Conditions were varied to determine the best operating parameters for the pretreatment system.

#### Multimedia Filter Testing

Multimedia filtration is used as an initial filtration step to minimize UF membrane exposure to foulants. Therefore, successful experiments were those which produced a filtrate that could be processed by the HF system without causing an unreasonable decrease in membrane flux. The fouling potential of the multimedia filtrate was determined by monitoring UF membrane permeation rates over time and by spectroscopic estimation of filtrate quality. Experiments were conducted using both multimedia filters (2.445" ID & 5.625" ID) to determine the filtration rates achievable and the backwash schedules (i.e., the filter solids loading before backwash).

The multimedia filter flux was found by determining flow rates and inlet and outlet pressures for both filters. Flow rate was found to be a linear function of pressure drop across the packed bed, indicating that laminar flow occurred in the bed. Discharge pressure was controlled while the desired feed rate was still obtained. Initial experiments indicated that filtrate quality, as measured by spectroscopic transmittance, was maximized at a flux of 5 gpm/ft<sup>2</sup> of bed area. The filtrate produced using a higher flux of 7 gpm/ft<sup>2</sup> resulted in an additional 19% decrease in membrane flux over 2 hours of operation as compared to the filtrate produced at a flux of 5 gpm/ft<sup>2</sup>. Flow rate was kept nearly constant at 5 gpm/ft<sup>2</sup>, with a clean filter pressure drop of 4 psig, during subsequent experiments and for design purposes.

The backwash cycle was initiated when the pressure drop across the filter increased by about 10 psig from the clean pressure drop of 4 – 7.5 psig. Figure 5 presents data of the multimedia filter pressure drop over time during filtering of the WERC-specified feed. The multimedia filter was backwashed after 7 hours of continuous operation. Backwashing was performed at a flow rate of 15 gpm/ft<sup>2</sup> of filter bed area to fully fluidize the bed. Following backwashing, as indicated by Figure 5, the pressure drop across the

filter returned to its original value. Spectroscopic testing showed that filtrate quality remained consistent before and after backwashing.

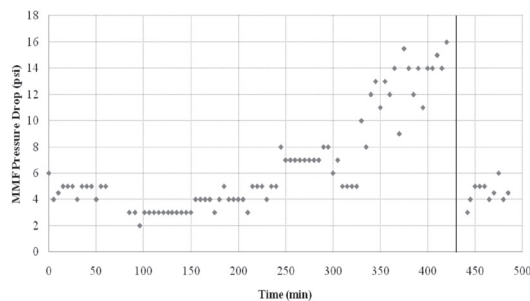


Figure 5. Multimedia Filter Pressure Drop With Backwash Results

#### Hollow Fiber Membrane Testing

Testing of the hollow fiber membrane was driven by the need to determine operating and cleaning procedures that would allow effective, economical, and long-term use of the membrane, with minimal maintenance time and required replacements. Additionally, chemical cleaning was minimized to design a green system with little onboard chemical storage. Several feed solutions were tested to determine the robustness of the system.

Tests were performed to determine permeation rates and backflush requirements. During these tests, permeate and concentrate flow rates were monitored, as were feed pressure and backpressure. System design and operational procedures were varied to maximize membrane life. A flow-reversal system, requiring only the closing and opening of a few valves, was installed to remove sediment deposits on the membrane surface. Reversing flow direction through the tubes during filtration minimizes the creation of static boundary layers inside the membrane fibers, which readily trap foulants. Additionally, permeate was used to backflush the membrane, removing foulants from the pores of the UF membrane. The membrane was backflushed at predetermined intervals by stopping the feed pump and forcing the permeate to flow backwards through the membrane. Backflushing significantly reduced the rate of permeate flow decrease and resulted in a higher steady-state permeate flow.

Backflush timing and duration were varied extensively during module testing. Testing showed that reducing the time between backflushes yielded greater improvement of steady-state flux than did increasing backflush time, which had only marginal short-term benefits. For the specified feed (after multimedia filtration), stable operation at 22 GFD was achieved after 4 hours of operation with alternating 2 minutes of forward operation with a 30 second backflush cycle, as shown in Figure 6.

A second, more concentrated, feed solution containing 20 mg/L of solids from the Orchid Pro was also tested. Under the same operating parameters, the permeate flow decreased at a faster rate, reaching 30 GFD in only 1 hour.

#### Membrane Cleaning

During laboratory testing, the membrane was chemically cleaned after each run. The cleaning procedure entailed circulating a 120 °F, 2 L 200 ppm solution of bleach or 1 wt% solution of

sodium hydroxide through the membrane for 15 minutes. This was followed by two separate 2 L washes with deionized (DI) water, each of which circulated through the membrane for 15 minutes. This ensured that the bleach was removed from the system. After cleaning, the flux through the membrane was measured with DI water to verify cleaning. This chemical cleaning restored the DI water flux through the membrane to about 150 GFD each of the 10 times it was cleaned. Cleaning with a room temperature 200 ppm bleach solution did not restore membrane flux to the same level; thus, it was essential to heat the cleaning solution to 120 °F.

The feed, multimedia filtrate, and HF permeate were tested for total suspended solids (TSS) and turbidity according to EPA Methods 160.2<sup>11</sup> and 180.1<sup>12</sup> by the Arkansas Water Quality Laboratory of the University of Arkansas at Fayetteville. Results are presented in Table 1. TSS was chosen instead of SDI because the procedure for SDI requires filtering through a 0.45 µm filter disc,

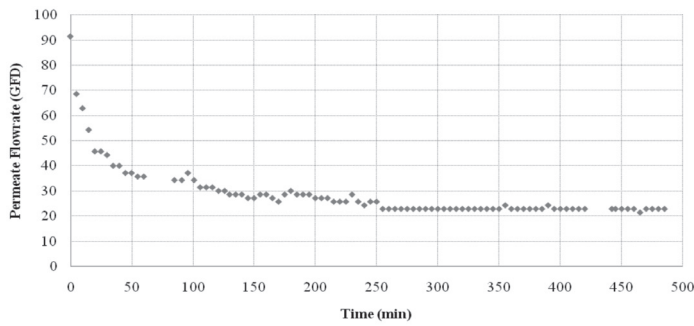


Figure 6. Hollow Fiber Permeate Fouling and Backflush Results

which would result in a measurement of zero as the UF membrane filters to a much smaller level. The multimedia filter reduced the suspended solids by nearly 80% and the turbidity by 70%. The permeate from the HF unit had a higher than expected TSS; however, due to the salt in the sample, it is likely that the results were high because careful experimental procedures are required to wash the filter paper free of salt. Additionally, contamination in the permeate backflush tank was detected after this testing was completed. The permeate turbidity was 1.1 NTUs for this sample, much higher than expected from a UF membrane (< 0.2 NTU). This result may be attributed to the contamination described above and to the use of a bench-top turbidity meter, which does not give very accurate results; in industry, flow-through turbidity meters are standard.

### Full-Scale Design

The equipment for the full-scale system consists of (1) a rotary vane feed pump, (2) a bank of 5 backwashable multimedia filters, (3) an array of 4 hollow fiber UF membranes, (4) a membrane cleaning pump, (5) a bleach metering tank, and (6) an inline heater. The pretreatment system is integrated with the RO waste streams to minimize power and water requirements. The full process flow diagram in the Appendix shows the complete system, with multiple units in parallel. The stream attributes, including all flows and pressures, are given in Table 2 for a single-unit.

Seawater is pumped to the multimedia filters at a rate of 106,000 GPD with a rotary vane pump. The raw seawater feed is pressurized to 45 psig by the feed pump before entering a distri-

bution manifold. The manifold delivers water to 5 multimedia filters, each 24" in diameter and 72" in height. Each multimedia filter is designed to process 15 gpm with a pressure drop of 5-15 psig across the packed bed. In typical operation, 5 of the filters are in operation at any one time to deliver 75 gpm of filtered water. During backwashing (which occurs for about 50 minutes over 8 hours), only 4 will be in forward operation. Any or all of the multimedia filters may be used at any one time.

Filtrate from the multimedia filters, pressurized to at least 25 psig, enters a distribution manifold leading to 4 UF membrane elements housed in a single bank array. The membranes are 72" long by 10.75" in diameter. Each cartridge contains 871 ft<sup>2</sup> of

Table 1. TSS and Turbidity Results

	TSS	Turbidity
	(mg/L)	(NTUs)
Feed	197.4	45.7
Filtrate	43	13.5
Permeate	15.5	1.1

polysulfone hollow fibers with a 100 kDa nominal MWCO. The inlet manifold is configured to allow flow reversal on the fiber side of the membranes, minimizing concentration polarization zones, which tend to cause fouling. Membranes are operated at 33% recovery (i.e., 1/3 of the feed permeates the membrane) to ensure high shear flow inside the fibers. In normal operation, two membrane modules in parallel will be online. The other two will either be in cleaning mode or in standby mode. When the feed solids concentration is high, all membranes may be operated. Permeate is produced at a rate of 35,300 GPD, and 30,000 GPD are delivered

Table 2. Stream Attributes Table for Simplified Flow Diagram

No.	Stream ID Name	Pressure psig	Flow Rate GPD	Component Flowrates				
				Water lb/day	Salt lb/day	Algae lb/day	Humus lb/day	Bleach lb/day
1	Seawater Feed	0	105750	871129	28155	66.4	0.90	0.00
2	Seawater Feed	45	105750	871129	28155	66.4	0.90	0.00
3	MMF Filtrate	40	105750	871129	28155	33.2	0.86	0.00
4	HF Feed (Forward Op.)	40	105750	871129	28155	33.2	0.86	0.00
5	HF Feed (Reverse Op.)	40	105750	871129	28155	33.2	0.86	0.00
6	HF Concentrate (Reverse Op.)	35	70500	580752	18770	32.9	0.85	0.00
7	HF Concentrate	35	70500	580752	18770	32.9	0.85	0.00
7a	HF Backflush Discharge	0	23190	190045	9361	0.3	0.01	0.00
8	HF Discharge	35	63750	525149	16973	29.7	0.77	0.00
8a	HF Backflush Discharge	0	23190	190045	9361	0.3	0.01	0.00
9	MMF Backwash	35	6750	55604	1797	3.1	0.08	0.00
10	HF Permeate	0	35250	290376	9385	0.0	0.00	0.00
10a	HF Backflush	20	23190	190045	9361	0.0	0.00	0.00
11	HF Permeate	0	35250	290376	9385	0.0	0.00	0.00
12	Recycle to Backflush	0	5250	43248	1398	0.0	0.00	0.00
13	RO Feed	0	30000	247129	7987	0.0	0.00	0.00
14	RO Brine Discharge	20	18000	147289	7987	0.0	0.00	0.00
15	Backflush	20	23250	190536	9385	0.0	0.00	0.00
16	Backflush	20	23190	190045	9361	0.0	0.00	0.00
17	Purge to Cleaning System	20	60	492	24	0.0	0.00	0.00
18	Cleaning System Recirc	35	60.08	492	24	0.0	0.00	0.11
19	Bleach (1.5%)	0	0.08	0.01	0	0.0	0.00	0.11
20	Cleaning Solution	35	60.08	492	24	0.0	0.00	0.11
21	Cleaning Solution	35	60.08	492	24	0.0	0.00	0.11
22	Cleaning Solution Discharge	35	60.08	492	24	0.0	0.00	0.11
23	Cleaning Solution Recirculation	35	60.08	492	24	0.0	0.00	0.11
24	Cleaning Solution Discharge	35	60.08	492	24	0.0	0.00	0.00
25	MMF Backwash Discharge	5	6750	55604	1797	36.3	0.13	0.00
26	Filter Discharge	20	70500	580752	18770	66.1	0.89	0.00
26a	HF Backflush Discharge	0	23190	190045	9361	0.3	0.01	0.00
27	Total Discharge to Ocean	20	70560	581244	18794	66.1	0.89	0.00

to the RO system. The tube-side pressure drop is about 5 psig. Concentrate from the HF array flows at a rate of 70,500 GPD and is used for backwashing the multimedia filters or is discharged to the sea.

Individual multimedia filters require 35 gpm during a 10-minute backwash. Backwashed sediments are discharged to the ocean (see Stream Attributes, Table 2, for average rates). Based on laboratory experiments, each filter should be backwashed every 8 hours. During backwash, the filter unit is isolated by proper positioning of valves, and the bed is fluidized.

The HF system is also equipped with a backflushing system. In the current NSRO system, the high-pressure RO reject passes through an energy recovery device with an efficiency of ~50%, undergoing a large decrease in pressure. The RO reject stream will be pressure regulated to 20 psig prior to its entering the shell side of the UF membrane. Membrane elements are backflushed every 2 minutes for a duration of 30 seconds, one element at a time. Following backflushing, the direction of flow through the membrane is reversed. Backflush fluid is discharged to the ocean.

The UF membranes require daily cleaning to remove fouling; thus, an automated membrane cleaning system is included. A cleaning cycle consists of circulation with a 120 °F, 200 ppm sodium hypochlorite solution (prepared by diluting 15% bleach), followed by two wash cycles using RO reject. One membrane is cleaned at a time. To initiate a wash cycle, feed to the membrane is terminated, and the permeate line is closed. The positive displacement cleaning pump draws brine from the RO system to fill the tube side of the fibers. Bleach is then metered into the stream from the bleach storage tank, and the stream is heated to 120 °F by an inline electric heater. This solution circulates through the membrane for 15 minutes. At the end of this time, 150 g of solid sodium thiosulfate ( $\text{Na}_2\text{S}_2\text{O}_3$ ) is introduced manually into the cleaning solution between the bleach tank and pump intake through a inline venturi. The sodium thiosulfate has a solubility of 79 g/100 mL of water (at 4 °C), which means that it should readily dissolve. Sodium thiosulfate reacts with bleach to produce sodium chloride, sodium sulfate, and water, all of which may be safely discharged directly to the ocean. The cleaning solution will circulate a further 6 minutes to allow the bleach to be neutralized. Then the solution will be discharged into the ocean. Next, two separate washes are performed with unheated brine from the RO system, each circulating for 15 minutes before discharging to the ocean. A hypochlorite monitoring device, installed in the permeate line, is necessary to ensure that no hypochlorite enters the RO membrane as RO membranes have no tolerance for free chlorine.

The design of the prefiltration system meets all space requirements. The process itself occupies approximately 250 ft<sup>3</sup>, which is only 2.5 times larger than the NSRO pretreatment system but filters solids down to 0.1 μm. Moreover, the design requires only 25 ft<sup>3</sup> for storage of spare HF membranes, sodium hypochlorite, and sodium thiosulfate over a 9-month time frame, which is only 2.5% of the volume used to store filter cartridges in the NSRO system.

The total power requirement for the system, using a 70% efficient rotary vane pump, is approximately 2.6 hp, which exceeds the WERC task limitation of about 1.2 hp. The task-specified

power requirement could have been met by adding more surface area through larger multimedia filters and more HF units or by operating at a higher recovery in the HF units. However, the first solution would exceed the operational space limitation, and the second adjustment would increase membrane fouling, requiring more storage of chemicals and membranes and a higher operating cost for replacements or more frequent cleaning cycles, thus increasing maintenance time and process footprint by necessitating more HF units. 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 typical Naval vessel, which is in excess of 15,000 hp.

Seawater is an electrolyte and can cause galvanic corrosion and crevice corrosion, so the materials used for the process must be selected with care.<sup>13</sup> A copper alloy, 70-30 Cu-Ni, is almost insensitive to stress corrosion, retains its strength well at moderately elevated temperatures, has good cold- and hot-working properties, and is readily weldable.<sup>14</sup> Already incorporated into desalination plants, this alloy provides a viable option for the wetted materials used by the process. However, 254 SMO stainless steel, developed for seawater usage, is also a good choice and is recommended for those portions of the process which come in contact with the bleach cleaning solution. Bleach should be stored in a fiberglass reinforced plastic (FRP) vessel to ensure service life in a highly corrosive environment. The multimedia filters are also constructed from FRP 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

The total capital investment for the system was determined by a factored estimate based on the purchased equipment cost.<sup>15</sup> In the interests of space, the detailed economic analysis is not provided here. Spare feed and cleaning pumps are required. The total equipment cost is \$47,300. Direct costs include equipment purchases, equipment delivery, equipment installation, instrumentation and controls, piping, and electrical systems. Indirect costs incurred in system installation were also estimated. The total capital investment for the process is \$230,000.

An incremental economic analysis comparing annual operating costs of the current reverse osmosis system and the proposed treatment system was performed on the basis of 12,000 GPD potable water production, as shown in Table 3. In this analysis, only those costs which were significantly different between the systems were considered. With the current system, 20 μm and 3 μm cartridge filters are replaced at least every 4 days, requiring 4 hours of maintenance each time, for an average of 365 hours/year of labor. Additionally, the lifetime of the reverse osmosis membranes in littoral waters is short, and it is assumed that elements are replaced every 6 months. In the proposed process, the higher quality of the RO feed stream will allow for longer RO membrane service times, and membranes should not have to be replaced over a 3-year period. Due to the harsh nature of the feed water and strict membrane backflushing and cleaning schedule, the hollow fiber



membrane modules will last at least 18 months. Media for multimedia filters should not require replacement more than once every 5 years, but the cost of replacement media for one of the five filters was included in the comparative economic analysis for completeness.

Maintenance requirements for the system are low because it is automated and requires infrequent component replacement. Monitoring of the system will be comparable to that of the existing system; however, operators will need only 30 minutes per day to add sodium thiosulfate following a cleaning cycle, compared to the 4 hours every 4 days for cartridge filter replacement. Operating labor costs are based on man-hours required for system operation, not on the number of operators. The chemicals and cleaning solution heating needed in the new design are considered. Other power cost differences between the two systems are negligible.

The design will save \$17,000 in operating costs over one year, yielding a payback period of 14 years. However, as savings are highly dependent on replacement frequency of all components, ships operating in coastal waters may realize greater savings of up to \$47,000 annually, reducing the payback period to 5 years.

**Table 3.** Annual Operating Cost Comparison

	Basis	Savings
RO Replacements	Manufacturer	\$10,000
Cartridge Filter Replacements	Manufacturer	\$20,000
Operating Labor	183 hour/year at \$25/hour	\$5,000
UF Replacements	Manufacturer	-\$15,000
Bleach (15%)	Manufacturer	-\$280
Sodium Thiosulfate	Manufacturer	-\$420
Replacement Media	Manufacturer	-\$1,600
Heating of Cleaning Solution	\$0.115/kWh	-\$1,200
<b>Annual Operational Savings</b>		<b>\$17,000</b>

### Inland Desalination Crossover

The application of multimedia filters coupled with UF membranes extends beyond shipboard systems to inland desalination plants. *“Compared to conventional chemical and physical based water treatment, ultrafiltration membrane systems require significantly less space and often have lower labor, chemical, and waste disposal costs”*.<sup>16</sup> Aside from the space requirements, differences exist in the feed water sources. The proposed system is designed for variable seawater quality in littoral regions, so inland sources of brackish water will not pose a problem for the membrane. Concerns with the direct application into inland plants lie in the production of the concentrate stream from the membrane. Common disposal methods include direct surface water discharge and sewer discharge, which are viable in coastal regions but not inland due to freshwater contamination.<sup>17</sup> The inland plant must address this issue with the RO reject, and the HF concentrate should be treated in a similar manner. Because of disposal problems, the proposed system should operate at higher water recoveries, minimizing waste from the UF membrane. HF systems have achieved water recoveries of 90%.<sup>16</sup> Higher recoveries for the proposed pretreatment process can be expected because brackish water instead of turbid seawater is used for inland desalination. *“Feed water properties, product water quality requirements, site conditions and specific project economics vary and must be carefully considered before using this process”*.<sup>16</sup> The proposed system is an attractive alternative to current physicochemical pretreatment of RO systems where “plant space is limited” and there is “high turbidity and

*variable feed water”*.<sup>16</sup>

### Regulations

Currently, the Oceans and Coastal Protection Division of the Environmental Protection Agency (EPA) has the task of preventing marine pollution. 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. The Vessel General Permit (VGP) is applicable to 26 specific discharge streams, including concentrated seawater produced as a byproduct of the processes used to produce freshwater from seawater. Following the VGP, brine from the 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> Some states, e.g., Hawaii, have additional regulations which prohibit the formation of objectionable sludge or bottom deposits. The discharges from the proposed pretreatment design meet these regulations.

The National Defense Authorization Act of 1996 amended Section 312 of the Clean Water Act to require the EPA and the Department of Defense (DOD) to develop Uniform National Discharge Standards (UNDS) for discharges incidental to the normal operation of a vessel of the Armed Forces. Although only in the second phase of three phases, the discharges of the proposed system will eventually be subject to the mandates of the UNDS. Phase I identified RO discharge brine as having the potential to cause adverse environmental effects because significant amounts of metals are discharged at concentrations above the water quality criteria. In this process, which employs multimedia filters followed by UF membranes, the concentration of metals in the discharges from the system will not be changed because they do permeate the UF membrane. The use of the RO reject for backflushing redirects an existing stream that may contain metal concentrations exceeding water quality standards into the system, but the design does not alter this concentration. The thermal effects of discharges were determined to be negligible.

### Chemical Considerations

Residual chlorine from the cleaning cycle needs to be completely eliminated before it can enter the RO membrane. Sodium hypochlorite solutions are frequently used aboard vessels to prevent biofouling, and the generated waste is continuously discharged when seawater cooling systems are in operation.<sup>19</sup> *“There are no Federal water quality criteria for chlorine. The most stringent state water quality criterion is 7.5 µg/L”*.<sup>19</sup> The 200 ppm solution is 0.2 g/L, which is an unacceptable direct discharge into the water; additionally, dilution with other system discharges does not meet the state criteria either. Abiding by this law and protecting the sea environment requires dechlorination before the cleaning solution discharges into the ocean. Neutralization with sodium thiosulfate produces sodium chloride, sodium sulfate, and water, all of which can be discharged in accordance with the strictest state regulations.

### Worker Safety

The primary safety measures for the proposed system are



prevention and training. The coupled filtration devices are operated at 45 psig, so care should be taken when maintaining the system. Since the system is automated, workers need to be aware of the electrical dangers involved. The cleaning solutions may be hazardous to human health, and under OSHA standards, personnel must be trained on the hazards of each chemical. Materials Safety Data Sheets also need to be onboard and available for each of the prescribed chemicals. While cleaning the UF membrane, workers should wear appropriate Personal Protective Equipment, including chemical gloves and eye protection, all conforming to OSHA standards. In order to prevent injury to personnel and damage to equipment, the tag-out program should be implemented for all equipment, components, and systems (mandatory aboard USN ships). MIL-STD-882 provides the protocol for implementing a safety program for the life of the system if used on military vessels.

#### *Additional Regulations*

Installation and use of the proposed pretreatment design aboard sea vessels requires conforming to standards unique to shipboard operations. The system needs to be able to withstand shock loadings and vibrations and also needs to be electromagnetically compatible with surrounding equipment. For use aboard military ships, these requirements are categorized in MIL-S-901D, MIL-STD-167-1A, and MIL-STD-461. The process also needs to meet all ship noise requirements (84 dB TWA, OPNAVINST 5090) and be made from materials that will not create toxic fumes during fires in enclosed spaces (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. Multimedia filtration followed by ultrafiltration is the most viable green pretreatment method for seawater reverse osmosis. The process removes particles down to 0.1  $\mu\text{m}$  to minimize the fouling potential of the RO feed stream and to prolong the life of downstream RO membranes.
2. The combination of backwashable multimedia filters with backflushable UF membranes minimizes the need for replacement elements, maintenance and operating time, and chemical additives.
3. Automated backwashing, backflushing, and cleaning protocols allow operation of the multimedia filters for several years and maintain high permeation rates through the UF membranes. The performance of both units returns to original levels following backwashing of the multimedia filter and chemical cleaning of the UF unit.
4. The operating unit requires 250 ft<sup>3</sup> of process space and 25 ft<sup>3</sup> of storage space for 9 months, which is below the task specifications of 300 ft<sup>3</sup> and 100 ft<sup>3</sup>.
5. The total capital investment for the designed process is \$230,000. The annual operational savings is \$17,000. The payback period is 14 years. However, as savings are highly dependent on replacement frequency of all

components, ships operating in coastal waters may realize greater savings of up to \$47,000 annually, reducing the payback period to 5 years.

6. It is recommended that long-term (90-day) pilot plant testing be conducted. Tests will confirm projected membrane service times and can address increasing the recovery rate in the HF units. The use of actual coastal seawater, instead of a surrogate mixture, is recommended to more closely approximate 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 backwashing and backflushing schedule for less turbid feeds can significantly increase the amount of permeate produced. Membrane recovery should also be increased to minimize discharges.

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**Mentor Comments:** Professor Roy Penney explains the work of the team of seniors who competed successfully in a US Navy sponsored research competition. Because Inquiry typically publishes single author manuscripts, Angela Mehner was chosen to represent the group.

*A team of 5 Chemical Engineering Seniors- Bartels, Mansell, McKnight, Mehner and Paul (the Seahogs)—competed in Task 4 of the 2010 WERC (<http://www.werc.net>) competition. The task was sponsored by the US Navy and was titled, "Green RO Pretreatment" and its objective was described on the WERC website as "Develop and demonstrate an alternative to disposable filters or an improved disposable filter design that can last at least four months filtering somewhat turbid feed water". The "disposable filters" protect shipboard potable water Reverse Osmosis Systems (RO) from foulants present in raw seawater. The Seahogs team started work on this Task January 4, 2010 and finished the competition as New Mexico State University on April 4-8, 2010. The team each received 3 credit hours for CHEG 4443, Senior Chemical Engineering Design II. The US Navy is most interested in this work because shipboard RO systems have the following problems when ships operate in coastal waters: (1) the existing prefilters*

*must be changed frequently and (2) the existing depth filters do not have the solids removal capability to adequately protect the expensive RO modules from damage by small particulates, particularly algae. Replacing depth filters and RO modules is expensive and labor intensive. The Seahogs (working independently, except for fabrication help) developed an elegant, ingenious solution to the problem, which totally eliminates the replacement of depth filters and protects the RO modules from particulates down to 0.05 $\mu$ m in size. The Seahogs design prevents any algae from reaching the RO modules. The system is totally automated and requires little operator attention, except for occasionally loading a weak bleach cleaning solution and adding sodium thiosulfate to neutralize the bleach cleaning solution. The excellent performance of the Seahogs team was rewarded by a panel of WERC judges - which included US Navy personnel - with a 1st place award for Task 4 of the 2010 WERC competition.*

**Appendix:** Full-Scale Unit Process Flow Diagram

