# FINAL REPORT

A Fully Integrated Membrane Bioreactor System for Wastewater Treatment in Remote Applications

## SERDP Project ER-2237



### OCTOBER 2018

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## **REPORT DOCUMENTATION PAGE**

Form Approved OMB No. 0704-0188

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1. REPORT DA	TE (DD-MM-YYYY	) <b>2. REPOR</b>	ΓΤΥΡΕ			3. DATES COVERED (From - To)
10/09/2018		SERDP F	inal Report			7/6/2012 - 7/6/2016
4. TITLE AND S	SUBTITLE	L			5a. CC	NTRACT NUMBER
A Fully Integr	ated Membran	e Bioreactor S	System for Wastewa	ater Treatme	nt 12-C-	0068
in Remote Applications				5b. GF	5b. GRANT NUMBER	
					5c. PR	OGRAM ELEMENT NUMBER
6. AUTHOR(S)					5d. PR	OJECT NUMBER
Amy E. Child	ress, University	of Southern	California		ER-22	037
5	, ,					
Andrea Achill	i, University of	Arizona			Je. TA	SK NUMBER
Sage R. Hiibe	el and Eric A. M	larchand, Uni	iversity of Nevada,	Reno		
Chanwoo Pa	rk, University o	f Missouri, Co	olumbia		5f. WC	ORK UNIT NUMBER
7. PERFORMIN	G ORGANIZATIO	N NAME(S) AND	DADDRESS(ES)			8. PERFORMING ORGANIZATION
University of	Southern Califo	ornia				REPORT NUMBER
Department of	of Civil and Env	ironmental Er	ngineering			ER-2237
Los Angeles,	CA 90089-253	1				
9. SPONSORIN	G/MONITORING	AGENCY NAME	(S) AND ADDRESS(ES)	)		10. SPONSOR/MONITOR'S ACRONYM(S)
Strategic Env	rironmental Res	search and De	evelopment Prograi	n		SERDP
4800 Mark C	enter Drive, Su	ite 17D03				
Alexandria, VA 22350-3605				11. SPONSOR/MONITOR'S REPORT		
		VETATEMENT				LI(-2237
		t for public re	lease: distribution (	unlimited		
DISTRIBUTI	JN A. Approve			minniteu.		
13. SUPPLEME	NTARY NOTES					
14. ABSTRACT	-					
The primary of	objective of the	research was	s to identify the ove	rall effectiven	ess of the i	ndividual system components during
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treatment sys	stem. As part of	this study, th	e operating parame	eters necessa	ary for highl	y efficient wastewater treatment and
water reuse v	vere determine	d. The systen	n was developed wi	ith online ana	lytical and o	control equipment to analyze system
performance	real-time. Ultim	lately, an inte	grated laboratory-s	cale system v	was tested i	using simulated high-strength FOB
wastewater.						
15. SUBJECT 1	ERMS					
Integrated Me	embrane Biorea	actor System,	Wastewater Treatr	nent, Remote	e Application	ns
16. SECURITY	CLASSIFICATION	OF:	17. LIMITATION OF	18. NUMBER	19a. NAME	OF RESPONSIBLE PERSON
a. REPORT	b. ABSTRACT	c. THIS PAGE	ABSTRACT	OF PAGES	Amy E. Ch	ildress
					19b. TELEP	HONE NUMBER (Include area code)
UNCLASS	UNCLASS	UNCLASS	UNCLASS	144	213-740-6	304
5	5	552,000				

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## List of acronyms

AFM	atomic force microscopy
AGMD	air-gap MD
AL-DS	active layer facing draw solution
AL-FS	active layer facing feed solution
BIO	biological reactor
CFD	computational fluid dynamics
COD	chemical oxygen demand
СР	concentration polarization
CTA	cellulose triacetate
DI	deionized water
DO	dissolved oxygen
DCMD	direct contact MD
EDS	elemental dispersive spectroscopy
EPS	extracellular polymeric substance
FO	forward osmosis
FOBs	forward operating bases
HRT	hydraulic residence time
HTI	Hydration Technologies, Inc.
HX	heat exchanger
LEP	liquid entry pressure
LMTD	log mean temperature difference
MBRs	membrane bioreactors
MD	membrane distillation
MF	microfiltration
MLSS	mixed liquor suspended solids
MVPC	membrane vapor permeability coefficient
OMBR	osmotic membrane bioreactor
ORP	oxidation-reduction potential

PP	polypropylene
PTFE	polytetrafluoroethelyne
RAS	return activated sludge
RO	reverse osmosis
RSF	reverse salt flux
SAAR	sequential anoxic/anaerobic reactor
SBR	membrane sequencing batch reactor
SEM	scanning electron microscopy
SEM-EDS	energy-dispersive x-ray spectroscopy
SRSF	specific reverse salt flux
SRT	solids retention time
SS	sidestream
SUB	submerged without aeration
SUB*	submerged with aeration
TMP	transmembrane pressure
TMWRF	Truckee Meadows Water Reclamation Facility
TOC	total organic carbon
TSS	total suspended solids
UF	ultrafiltration
UV	ultraviolet

### **Executive summary**

Introduction:

An osmotic membrane bioreactor (OMBR) system with membrane distillation (MD) reconcentration was developed for potable reuse of wastewater at forward operating bases (FOBs). The three major processes of this system (biological reactor (BIO), forward osmosis (FO), and MD) were systematically investigated at bench scale before laboratory-scale versions were designed, fabricated, and integrated into a single system. The modular system was designed to operate using low-grade ("waste") heat existing at FOBs. The abundance of low-grade heat produced by power generators at FOBs make the OMBR-MD system ideal for such locations.

Membrane bioreactors (MBRs) have proven to be effective and beneficial for wastewater reclamation due to their ability to produce high quality effluent with a small footprint [6]. In a conventional MBR, microporous (microfiltration (MF) or ultrafiltration (UF)) membranes) are immersed in a bioreactor and water is filtered through the membranes using vacuum; suspended solids are retained in the system and high levels of treatment can be achieved [7, 8]. MBRs replace the two stages of the conventional activated sludge process with a single, integrated process that has a much smaller footprint. MBR effluent may be suitable for use as irrigation water, process water, or potable water. For potable reuse, advanced treatment (e.g., reverse osmosis (RO) and chemical oxidation) is necessary after the MBR [9].

The advantages of MBRs over conventional treatment have been thoroughly reviewed [6] and include consistently high product quality, ease of operation/automation, reduced footprint, reduced sludge production due to a high biomass concentration in the bioreactor, and complete suspended solids removal from the effluent [6]. The main problem associated with MBRs is membrane fouling. Membrane fouling can occur in the MBR itself, and also in the downstream RO system [10]. Specifically, high concentrations of dissolved organic compounds in the MBR effluent can cause severe fouling of RO membranes.

Membrane fouling lowers productivity, increases energy requirements, increases frequency of membrane cleaning and replacement, and may result in deterioration of treated water quality [11]. A novel MBR system that utilizes a submerged FO membrane in the bioreactor and a membrane distillation (MD) membrane in a separate reconcentration process was investigated. Prior to describing the integrated system, the basic FO and MD processes are first outlined.

Osmosis is the net diffusive transport of water through a selectively permeable membrane from a solution of low solute concentration (low osmotic pressure) to a solution of high solute concentration (high osmotic pressure). In osmosis, the membrane allows passage of water, but rejects almost all solute molecules and ions. FO is an engineered application of osmosis [12, 13].

In FO, impaired water is in contact with the dense side of a semipermeable membrane and a highly concentrated draw solution is in contact with the support side of the membrane. The draw solution is typically an aqueous solution of a low molecular weight salt [14]. In the FO process, relatively pure water is transported from the impaired water into the draw solution; the impaired water becomes concentrated and the draw solution becomes diluted. A desalination process (e.g., RO or distillation) can be used to reconcentrate the draw solution and simultaneously produce high-quality product water (Figure 1). Thus, in most water treatment applications, FO is not the ultimate process but rather a high-level pretreatment process before an ultimate reconcentration/desalination process [14].



#### Figure

Figure 1. Forward osmosis (FO) system schematic.

FO has been evaluated for seawater and brackish water desalination [12, 13, 15], wastewater concentration and reclamation [16-19], and food concentration [20-22]. FO can also be employed in conjunction with biological processes for wastewater reuse in osmotic membrane bioreactors (OMBRs) [23, 24]. The main advantage of FO is that it operates at very low hydraulic pressure, it has high rejection of a broad range of contaminants [23, 25], and may have lower fouling propensity and/or fouling that is more reversible than in RO processes [18, 26].

MD is a non-pressure-driven membrane separation process that has the potential to become a viable tool for water desalination and brine (or draw solution) concentration. MD involves the transport of mass and heat through a hydrophobic, microporous membrane. In direct contact MD (DCMD) (Figure 2), a feed solution at elevated temperature is in contact with one side of the membrane and a permeate solution at lower temperature is in direct contact with the opposite side of the membrane; it is mainly the temperature difference between the liquids that induces the vapor pressure gradient for mass transfer. DCMD is one of the simplest configurations of MD; it requires only a membrane module and two low-pressure pumps to move the liquids over the membrane. The simplicity of this configuration makes it highly suitable for implementation in large-scale applications.

Compared to conventional distillation methods, MD requires only small temperature differences – temperature differences achievable through the use of low-grade heat/renewable energy. Compared to RO, the driving force in MD is essentially not reduced by osmotic pressure and thus, MD operates at hydrostatic pressure and can be used to treat high salinity solutions or provide enhanced recovery through brine desalination. Also compared to RO, MD produces an even higher quality distillate; it does not allow the passage of small non-volatile compounds and provides high removal of emerging contaminants. Specifically, MD has been shown to reject greater than 97% of low molecular weight organics [27]. For these reasons, MD may have substantial energy, recovery, maintenance, and water quality advantages over conventional methods.



Figure 2. DCMD: feed and permeate solutions are both in direct contact with the membrane.

One very important aspect of the MD process is the membrane cannot become hydrophilized. Hydrophilization can occur due to fouling of the MD membrane if proper pretreatment is not used. FO provides outstanding pretreatment of feed streams to MD membranes because it provides high rejection of dissolved species [18, 23].

In an OMBR system (Figure 3), wastewater is fed into a reactor which is aerated as needed to supply oxygen for the biomass and to minimally scour the membrane. Through osmosis, water diffuses from the bioreactor, across a semipermeable membrane, and into the draw solution. The FO membrane acts as a barrier to solute transport and provides high rejection of contaminants in the wastewater stream [16, 19, 25]. The diluted draw solution is sent to a reconcentration process (e.g., RO or MD), which reconcentrates the draw solution and generates a high-quality product water. Thus, FO serves as a high-level pretreatment step before the RO or MD reconcentration. Compared to the MF or UF process in a conventional MBR, the FO process in the OMBR offers the advantages of much higher rejection (semi-permeable membrane versus microporous membrane) at lower hydraulic pressure. The FO membrane inside the bioreactor also has much lower fouling propensity than the MF/UF membranes, and thus, requires minimal scouring and much less frequent backwashing. When comparing an OMBR system (OMBR followed by RO or MD reconcentration) with a conventional MBR-followed-by-RO system, the high rejection of the FO membrane results in RO influent with lower fouling propensity and ultimately produces higher quality product water. Furthermore, the FO-followed-by-RO or -MD reconcentration treatment scheme represents a tight dual barrier system. This is particularly important when pharmaceutical or personal care products are present.



Figure 3. OMBR system.

During the initial stages of development, bench-scale systems were used to demonstrate the feasibility of OMBR with MD reconcentration for wastewater treatment and reuse and to evaluate process performance and water quality with each process. Alternating aeration on/off cycles, or aeration cycling, was used to accomplish biological carbon and nitrogen removal in a single reactor with the OMBR. Bench-scale experiments were conducted to determine the effects of FO membrane module configuration (submerged in the bioreactor or as a sidestream process) on OMBR performance. Bench-scale results were used to guide development of the pilot-scale system and performance of a submerged FO membrane module and a spiral-wound membrane module were evaluated at pilot-scale. Process metrics were established to evaluate membrane performance and wastewater treatment; membranes and membrane modules were evaluated for water flux, reverse salt flux, and solute rejection and wastewater treatment was evaluated for contaminant (carbon and nitrogen) removal. Results from long-term testing using a high-strength wastewater showed essentially 100% removal of COD and 96.2% removal of  $NH_4^+$ -N could be achieved with aeration cycling in a single reactor. The research results from this SERDP project have been published in several peer-reviewed journal articles [1-5], with more in preparation.

#### Objectives

The main objective of this research was to design, fabricate, and test a pilot-scale OMBR-MD system consisting of BIO, FO, and MD processes and driven by on-site waste heat. As part of this, the operating parameters necessary for highly efficient wastewater treatment and water reuse were determined. The biologically-catalyzed processes of the OMBR provide a combined treatment process for organic matter degradation and nutrient removal. When coupled to the osmotically driven mass transport mechanisms of FO, the system offers a low-fouling alternative to traditional MBR applications while also decreasing the electrical energy demands required for the process. The FO membranes also serve as a selective barrier system capable of rejecting recalcitrant trace organic compounds to prevent passage into the draw solution stream, thereby decreasing fouling in the downstream MD process.

The steps toward achieving the goals of the proposed research were to systematically investigate the three major processes of the OMBR system (BIO, FO, and MD) and at the same time, determine the energy requirements for each process. Subsequently, the processes were integrated into a single system through the design and construction of a modular pilot-scale OMBR system that operates using renewable energy or low-grade ("waste") heat existing at FOBs. At each stage of the research, process performance was evaluated based on specific indicators (e.g., water quality and quantity, energy and resource consumption, and frequency of hands-on modification to the system).

To carry out the objectives, several tasks were performed including fundamental studies aimed at researching specific issues for each of the processes (BIO, FO, and MD) and the heat exchangers. With regard to each process, investigation of the critical phenomena enabled answering of fundamental questions about microbial ecology, FO membrane

fouling, and MD membrane flooding that are essential for scale-up and integration of these technologies. The research team collaborated with Hydration Technologies, Inc. (HTI) (now Fluid Technology Solutions, Inc.), GE Water & Process Technologies, and Parker Performance Materials to acquire the most appropriate FO and MD membranes to characterize and test. Next, process models and controls for each of the subsystems were developed. Finally, OMBR and MD processes were integrated and a system-level model was developed and tested for water quality and membrane performance.

#### **Technology Approach**

To design, fabricate, and test a pilot-scale OMBR-MD system consisting of a BIO, FO, and MD process and driven by on-site waste heat several tasks were performed, as outlined below. The project started by investigating fundamental aspects of the main components of the OMBR-MD system in Task 1: bioreactor design, instrumentation, and control, Task 2: bench-scale FO membrane fouling, Task 3: bench-scale MD process modeling, and Task 4: heat exchanger design and CFD modeling for MD. then it progressed toward the assembly of the integrated pilot-scale system in Task 5: lab-scale BIO+FO sub-system testing, Task 6: design of lab-scale MD + heat exchanger sub-system, Task 7: lab-scale FO-BIO and MD-heat exchanger sub-system assembly, Task 8: lab-scale OMBR system testing, and Task 9: OMBR system scale-up testing.

#### **Results and Discussion**

#### Development of an Oxic-Anoxic Bioreactor for Treating FOB Wastewater

A simultaneous nitrification-denitrification process was tested and optimized. The bioreactor used for these experiments was inoculated with a combination of activated sludge, bacteria from a nitrifying trickling filter biomass, and microbes from a denitrifying system at a regional wastewater treatment facility. Bench-scale experiments were conducted in a pair of bioreactors specially designed to accommodate a flat sheet membrane. The two process waste streams were tested in parallel using the two systems: (i) graywater influent [28] and (ii) combined gray+black wastewater (based on typical domestic wastewater composition). For purposes of this subtask, the unit mimicked a traditional MBR system in that it had an MF membrane solely for the purpose of retaining biomass. In subsequent research tasks, the biological system was combined with the FO system, which served to retain more than biomass. Since the FO system in subsequent tasks contains a semi-permeable membrane that is similar to a RO membrane, both suspended and dissolved constituents can be retained within the system. In the current task, which utilizes an MF membrane (pore opening size ~0.1-0.2  $\mu$ m), only suspended constituents larger than this size were retained. The bioreactor was operated in alternating aerobic/anoxic reaction modes to allow for carbon and nitrogen removal. The following general, non-balanced chemical reactions (Eqs. 1-3) describe the major reactions occurring in each mode. Operation in this fashion allowed simultaneous urea hydrolysis (Eq. 1), nitrification (Eq. 2), and denitrification (Eq. 3), which ultimately removed dissolved contaminants.

Aerobic & Anoxic:	$Urea + Microorganisms \rightarrow NH_3 + CO_2$	(Equation 1)
Aerobic only:	$NH3 + O_2 \rightarrow NO_2^{-}/-NO_3^{-} + H_2O$	(Equation 2)
Anoxic only:	Organic Carbon + $NO_2^{-}/NO_3^{-} \rightarrow N_2 + CO_2 + H_2O$	(Equation 3)

Biochemical experimental variables monitored during the experiments included DOC, ammonia-nitrogen ( $NH_3$ -N), nitrite-nitrogen ( $NO_2$ <sup>-</sup>-N), nitrate-nitrogen ( $NO_3$ <sup>-</sup>-N), and biomass (measured as volatile suspended solids, VSS). Chemical parameters were measured in the feed stream, within the bioreactor during reaction cycles, and in the final effluent stream using accepted analytical methods.

Microbial populations and ecology (see below) were monitored at reasonable long-term intervals (i.e., 15-30 days). To be effective at achieving complete treatment, cycle duration and switching between aerobic and anoxic conditions were optimized. While these processes were relatively easy to optimize individually, it was technically challenging to engineer these processes such that they occurred in a single system and consumed a minimum of resources

(specifically, oxygen and exogenous resources, such as energy for mixing, internal recirculation, and membrane cleaning).

Shortcut denitrification is a process where the nitrogen in ammonia is only oxidized to nitrite (NO<sub>2</sub><sup>-</sup>) which is subsequently reduced to nitrogen gas during denitrification. This process relies on less oxygen (and consequently less energy) than traditional nitrification-denitrification sequences. As such, a major thrust was to perform fundamental research on a single-sludge process and study the microbiological community shifts that occurred throughout the system. The central hypothesis that was addressed during this research phase was: a balanced nitrification-denitrification treatment process can be achieved with carefully tailored oxic and anoxic reaction times based on chemical signatures in the bioreactor. Thus, a major component of this task was to identify proper chemical cues (including water quality and biochemical indicators) that were required to switch between operational modes. Specifically, oxidation-reduction potential (ORP), dissolved oxygen (DO), and pH were continuously analyzed along with water quality parameters (i.e., ammonia, nitrite, nitrate, total organic carbon (TOC), and chemical oxygen demand (COD)) to determine appropriate switching times between aerobic and anaerobic reaction sequences.

Simulated FOB wastewater was developed to challenge the microbial reactor system and enable researchers to rigorously identify the switching requirements for achieving maximum nutrient removal. The key FOB wastewater constituents that were mimicked in the laboratory study were organic carbon load (measured as total organic carbon and COD), nitrogen speciation and respective constituent concentrations (urea and ammonia), typical nutrients and micronutrients (phosphorus, iron, sulfur), and suspended solid concentrations (measured as total suspended solids (TSS) and VSS).

#### Development of an Integrated Sensor-Based Control System for the Bioreactor

A control and operating program for the biological process of the OMBR system was developed; specifically, the program allowed biological nutrient removal in a single reactor. The program relied on DO, oxidation-reduction potential, pH, and time to switch between oxic and anoxic phases to achieve nitrification and denitrification, respectively. This subtask was built upon data collected in the previous task, which describe changes in biochemical reaction during wastewater treatment.

A LabVIEW-controlled system was added to the bioreactor system used during the previous task and a computercontrolled system was developed to allow the aeration system to automatically turn on and off as a function of bioreactor probe outputs. Briefly, experimental data provide a suite of ORP, DO, and pH performance curves that were correlated with biochemical environment (specifically dissolved nitrogen speciation and organic carbon concentrations). Based on these data, real time switching criteria were incorporated into the LabVIEW-controlled system to control and maintain proper air delivery to the bioreactor.

#### FO Membrane Fouling, Process Modeling, and Control

FO uses osmotic pressure differential across a semipermeable membrane, rather than hydraulic pressure differential (as in RO), as the driving force for water transport through the membrane. The FO process results in concentration of a feed stream and dilution of an osmotic agent (i.e., a draw solution).

There is evidence that in some cases, fouling occurs to a lesser extent in FO than in RO [18]. In other cases, it has been found that fouling occurs at similar rates in FO and RO; however, the fouling is more reversible in FO – possibly due to the absence of hydraulic pressure in the FO feed stream [27]. Although there is indication in the literature that the presence of salt at the membrane surface may play a role in membrane cleaning [29], the basis for the more reversible fouling has not been proven. This was explored to provide critical insight into FO transport phenomena and fouling reduction. FO membranes were used to conduct bench-scale fouling and cleaning experiments. A range of foulants (organic, inorganic, colloidal, and biological) was used to assess the ability to remove these foulants using rinsing and osmotic backwashing. During osmotic backwashing, water passed from the support side of the membrane to the active side, thereby reversing the direction of flow through the FO membrane and likely removing foulants. Foulant removal was determined by hydraulic testing and scanning electron microscopy (SEM) imaging.

#### MD Membrane Hydrophilization, Process Modeling, and Control

Maintaining membrane hydrophobicity is absolutely essential for assuring that the pores of MD membranes do not become wet. If hydrophobicity is lost, the pores can wet with water and salts and other contaminants can diffuse into the distillate [30]. Pore wetting is considered irreversible, in other words, after the pressure that caused pore wetting is removed, the pores remain wet. In order to restore permeate quality, removal of fluids and drying is required. Pore wetting results in a decline in distillate quality; downtime and prevention of this phenomenon is absolutely critical for successful MD scale-up.

The interrelationship between hydrophobicity and pore size distribution is described in terms of the liquid entry pressure (LEP). LEP is given by the difference between liquid pressure and vapor pressure:

$$P_{l} - P_{V} \le -\frac{2\gamma_{L}\cos\theta}{r} = LEP$$

#### (Equation 4)

where  $P_1$  is the pressure of the liquid in the channels of the MD module,  $P_v$  is the pressure of the vapor inside the membrane pores,  $\gamma$  is liquid surface tension,  $\theta$  is contact angle, and r is pore radius (for non-cylindrical pores, a geometry correction factor needs to be introduced). Theoretically, the membrane pores remain dry as long as the LEP is not exceeded.

#### Energy Considerations for MD

In this task, the MD system was analyzed using commercial computational fluid dynamics (CFD) tools to determine energy and flow requirements and to optimize MD system design. The thermal and hydraulic characteristics of the MD system and its heat exchanger system were numerically analyzed using ANSYS Fluent, a commercial CFD package. The operational conditions and dimensions based on the small flat-sheet MD system under testing were considered for the CFD analysis to calculate energy and flow requirements of the MD system.

#### Testing of the FO + MD Sub-System

Integration of the FO and MD processes centered on selection of an appropriate draw solution. The draw solution for the FO process was the feed solution for the MD process; the ideal draw solution was one that optimized flux in FO and was easily reconcentrated by MD. It was also one that did not reduce MD membrane hydrophobicity after exposure to high temperatures over longer time periods. After screening the draw solutions, bench-scale FO experiments were carried out to test the water flux produced by each draw solution and quantify the ability to reconcentrate the selected draw solutions.

#### **Implications for Future Research and Benefits**

An integrated OMBR-MD system was designed, fabricated, and tested at the pilot-scale for wastewater treatment and production of high-quality reuse water. There were very clear synergies in using MD to reconcentrate the draw solution used in the OMBR process, in using low-grade heat/renewable energy to drive the MD process, and in recovering heat from the distillate solution to reduce the overall heat consumption of the system. The abundance of low-grade heat produced by diesel generators at FOBs make the OMBR-MD system an ideal system for such locations. During long-term OMBR operation, carbon and nitrogen removal was achieved in a single-reactor by alternating between aerobic and anoxic bioreactor conditions. Results from long-term testing using a high-strength wastewater showed 98.4% removal of COD and 90.2% removal of NH4+-N was achieved.

Physical design aspects and process controls for system integration as well as procedures for wastewater treatment with the single-stage OMBR were established. Several areas for improvement were also identified throughout the development and testing phases of this system. Namely, passage of some dissolved, low molecular-weight contaminants into the closed-loop FO-MD solution led to MD membrane fouling and reduced MD water flux, and over time, the ammonium concentration in the distillate solution tended to increase. A suggested improved design will draw upon this and other lessons learned during the current project. Ammonium passage into the distillate solution is mitigated by improved biological ammonium removal (i.e., nitrification/denitrification) in the OMBR. To accomplish this, it is suggested to configure the OMBR with a dual-chamber tank with separate anoxic and oxic zones. Aeration in the oxic zone would be optimized for nitrification and membrane cleaning; the anoxic zone would be optimized for denitrification. The revised OBMR-MD system operates in a semi-batch configuration

defined by two separate process loops. One loop operates the OMBR and the other operates multiple MD modules. Each loop operates individually until two triggering conditions are met; the draw solution reaches a target dilution in the OMBR loop and the MD feed solution reaches a target concentration in the MD loop. Once the triggering conditions are met, the solutions in each loop are exchanged; the diluted FO draw solution is then treated by MD and the concentrated MD feed solution is then diluted by FO. This semi-batch process scheme allows for independent and "continuous" operation of the FO and MD loops, reducing the complexity of the necessary control mechanisms to maintain steady-state operation of a dynamic, continuous process and thus implementing a simplified operation that maximizes the performance of each process. Once the tank switch occurs, the diluted draw solution volume will pass to the MD loop and undergo further treatment through a fine screen, a cartridge filter, a carbon filter, and an ultraviolet (UV) reactor to reduce the possibility of membrane fouling, membrane damage, or passage through the MD membrane by small organic compounds or disinfection residual.

The suggested revised system will be programmed to periodically backwash the FO membranes to extend membrane life and improve performance. The backwash cycle will utilize a portion of the distillate stored in a standalone tank. When FO performance degrades, or after a predetermined time period has elapsed, the system will cease normal operation and change to the necessary valve configuration to route the distillate to the FO module. Backwashing in a counter-flow regime will reverse the impact of fouling and scaling that occurs over time on the FO membrane. The used backwash solution will be further treated with MD to maximize water recovery. Regarding the MD subsystem and based on testing subsequent to development of the current OMBR-MD system, it is suggested to considered use of an FO module in the revised system instead of the DCMD module that is used in the current system. In AGMD, heat recovery can be integrated into the module design.

### **Project Overview**





#### Introduction and Objectives

The availability of high-quality water at military installations is a major factor that may affect DOD missions abroad. Efficient on-site recycling of water would serve to provide additional security and ensure that armed forces personnel can focus on the mission at hand as opposed to securing reliable water resources to support base activites. To address on-site water sustainability, an osmotic membrane bioreactor (OMBR) system with membrane distillation (MD) reconcentration was developed for potable reuse of wastewater at forward operating bases (FOBs). Multiple technologies were combined to develop the treatment process, while minimizing system size and operating inputs. Specifically, a biological system was incorporated to allow for microbial degradation of organic carbon and nutrients in wastewater. A membrane process - forward osmosis (FO) - was included to provide a physical biomass separation barrier and allow for contaminant rejection and low-pressure water passage from the bioreactor into a saline draw solution. In order to re-concentrate the draw solution and produce high-quality product water, another membranebased technology (membrane distillation) was incorporated into the system. MD is an attractive option for FOBs since it utilizes waste heat instead of electricity as used in reverse osmosis (RO) applications to produce product water. The FO membranes also serve as a selective barrier system capable of rejecting recalcitrant trace organic compounds to prevent passage into the draw solution stream, thereby decreasing fouling in the downstream MD process. The combined system offers an attractive approach to producing high-quality reuse water while utilizing waste heat to drive membrane distillation.

The primary objective of the research was to identify the overall effectiveness of the individual system components during bench-scale laboratory experiments and then to combine the components to assess performance of the integrated treatment system. As part of this study, the operating parameters necessary for highly efficient wastewater treatment and water reuse were determined. The system was developed with online analytical and control equipment to analyze system performance real-time. Ultimately, an integrated laboratory-scale system was tested using simulated

high-strength FOB wastewater. System performance was assessed by monitoring water quality, chemical usage, energy requirements, and maintenance/cleaning requirements for individual sub-systems.

#### Technical Approach

An important consideration when developing water treatment and reuse systems is providing sufficient redundancy and multiple treatment barriers to remove target chemicals or biological agents. The OMBR system developed and tested contains multiple components meet this goal. The three major processes of this system (biological reactor (BIO), FO, and membrane distillation) were systematically investigated at the bench scale before laboratory-scale versions were designed, fabricated, and integrated into a single system. The modular system was designed to operate using low-grade ("waste") heat existing at FOBs. Membrane distillation can operate at ambient temperatures with a temperature difference of 20-30 degrees Celsius. The abundance of low-grade heat produced by power generators at FOBs make the OMBR-MD system ideal for such locations.

To carry out the objectives, several tasks were performed including fundamental studies aimed at researching specific challenges for each of the processes (BIO, FO, and MD) and the heat exchangers. With regard to each subsystem, investigation of the critical phenomena enabled answering of fundamental questions about microbial ecology, FO membrane fouling, and MD membrane flooding that are essential for scale-up and integration of these technologies. The research team collaborated with Hydration Technologies, Inc. (HTI, now Fluid Technology Solutions, Inc.), GE Water & Process Technologies (now Suez Water Technologies & Solutions), and Parker Performance Materials to acquire the most appropriate FO and MD membranes to characterize and test. Following research at the bench-scale, process models and controls for each of the subsystems were developed and integrated within the pilot-scale system. Finally, OMBR and MD processes were integrated, and a system-level model was developed and tested for water quality and membrane performance.

This report includes a discussion on process development, physical coupling of the BIO, FO, and MD processes into a system designed for continuous operation, and development of system controls. At this time, this is the first OMBR-MD system to be tested at pilot scale.

#### Results

During the initial stages of development, bench-scale systems were used to demonstrate the feasibility of OMBR with MD reconcentration for wastewater treatment and reuse and to evaluate process performance and water quality with each process. Alternating aeration on/off cycles, or aeration cycling, was used to accomplish biological carbon and nitrogen removal in a single reactor with the OMBR. Bench-scale experiments were conducted to determine the effects of FO membrane module configuration (submerged in the bioreactor or as a sidestream process) on OMBR performance. Bench-scale results were used to guide development of the pilot-scale system and performance of a submerged FO membrane module and a spiral-wound membrane module were evaluated at pilot-scale. Process metrics were established to evaluate membrane performance and wastewater treatment; membranes and membrane modules were evaluated for water flux, reverse salt flux, and solute rejection and wastewater treatment was evaluated for contaminant (carbon and nitrogen) removal.

An automated dosing and transfer system was developed to couple the pilot-scale OMBR and MD processes, to maintain constant FO draw solution concentration, and to prevent heat from being transferred to the bioreactor. Bioreactor salinity, which increased due to reverse salt flux, was also monitored and controlled. Thus, evaluation of mass and heat transfer played a critical role in design considerations for maintaining stable biological treatment performance. The system was equipped with electronic sensors connected to a control system for automated operation, process control, and data collection. System metrics were established to evaluate integrated system performance. Results from long-term testing using a high-strength wastewater showed essentially 100% removal of chemical oxygen demand (COD) and 96.2% removal of NH4<sup>+</sup>-N could be achieved with aeration cycling in a single reactor. The research results from this SERDP project have been presented at numerous scientific conferences and published in several peer-reviewed journal articles.

#### Benefits

An integrated OMBR-MD system was designed, fabricated, and tested at the pilot-scale for wastewater treatment and production of high-quality reuse water. There were very clear synergies in using MD to reconcentrate the draw solution used in the OMBR process, in using low-grade heat/renewable energy to drive the MD process, and in recovering heat from the distillate solution to reduce the overall heat consumption of the system. The abundance of low-grade heat produced by diesel generators at FOBs make the OMBR-MD system an ideal choice for such locations. Physical design aspects and process controls for system integration as well as procedures for wastewater treatment with the single-stage OMBR were established. Several areas for improvement were also identified throughout the development and testing phases of this system.

Overall, the OMBR-MD system exhibited very good COD and nutrient removal, making it suitable for water treatment and reuse applications at forward operating bases or other locations with an abundance of low-grade heat. Since many of the technologies incorporated are still early in the development and testing phase, additional research should be performed on the integrated system prior to deployment at FOBs. Specifically, different configurations of the FO-BIO components should be evaluated at the pilot-scale to maximize both processes. The biological process would be enhanced by designing separate stages to allow for aerobic and anoxic treatment without the need to alternate the redox environment in a single tank. This could be achieved by designing a baffled system with internal recirculation and mixing. Two FO membrane configurations (submerged flat sheet and side-stream spiral wound modules) were tested during this research. Utilizing a spiral-wound FO membrane unit in a side-stream configuration resulted in enhanced performance and allowed both systems to be independently optimized. Regarding the MD subsystem and based on testing subsequent to development of the current OMBR-MD system, it is suggested to consider the use of an air-gap MD (AGMD) module in the revised system instead of the direct-contact membrane distillation (DCMD) module that was used during this research. In AGMD, heat recovery can be integrated into the module design, thereby increasing the net water production and minimizing external heat exchanger requirements.

### **Response to In Progress Review Comments**

#### **Operator Training**

Two documents were prepared for use as operator training tools; the instrumentation and controls document and the operating manual. The instrumentation and controls document includes process descriptions of the BIO, FO, and MD subsystems and it details the physical locations of system hardware such as pumps, solenoids, and measurement devices. The instrumentation and controls document also has a supplemental wiring diagram detailing the connections between these devices and the automated LabVIEW control system. The operating manual describes system operation and maintenance procedures and it is intended to be a user guide for the specific system developed for SERDP. These documents were prepared to assist person-to-person operator training. Although a water treatment plant operator license would not be required, the system operator should have moderate knowledge of waste water treatment and membrane processes to successfully operate the system.

#### System Robustness

The biological and chemical environment within the bioreactor led to FO membrane fouling and subsequent flux reduction. To maintain flux, the OMBR subsystem was programmed to automatically initiate osmotic backwashing cycles. As feed water quality varied, the amount of aeration required by the biological system also varied. Biological wastewater treatment was maintained throughout variations in feed water quality (influent wastewater loading) by controlling aeration cycles based on real-time dissolved oxygen (DO) concentrations. For example, when the wastewater became more concentrated, the biological oxygen demand increased and aeration was continued until a DO set point was reached. In the future, an equalization tank may be installed to reduce variations in feed water quantity. While phosphorus removal was not a key objective in this study, the primary mechanisms for phosphorus removal were through incorporation in biological solids and solids wasting.

### **Objective**

The main objective of this research was to design, fabricate, and test a pilot-scale OMBR-MD system consisting of BIO, FO, and MD processes and driven by on-site waste heat. As part of this, the operating parameters necessary for highly efficient wastewater treatment and water reuse were determined. The biologically-catalyzed processes of the OMBR provide a combined treatment process for organic matter degradation and nutrient removal. When coupled to the osmotically driven mass transport mechanisms of FO, the system offers a low-fouling alternative to traditional MBR applications while also decreasing the electrical energy demands required for the process. The FO membranes also serve as a selective barrier system capable of rejecting recalcitrant trace organic compounds to prevent passage into the draw solution stream, thereby decreasing fouling in the downstream MD process.

The steps toward achieving the goals of the proposed research were to systematically investigate the three major processes of the OMBR system (biological (BIO), FO, and MD) and at the same time, determine the energy requirements for each process. Subsequently, the processes were integrated into a single system through the design and construction of a modular pilot-scale OMBR system that operates using renewable energy or low-grade ("waste") heat existing at FOBs. At each stage of the research, process performance was evaluated based on specific indicators (e.g., water quality and quantity, energy and resource consumption, and frequency of hands-on modification to the system).

To carry out the objectives, several tasks were performed including fundamental studies aimed at researching specific issues for each of the processes (BIO, FO, and MD) and the heat exchangers. With regard to each process, investigation of the critical phenomena enabled answering of fundamental questions about microbial ecology, FO membrane fouling, and MD membrane flooding that are essential for scale-up and integration of these technologies. The research team collaborated with Hydration Technologies, Inc. (HTI) (now Fluid Technology Solutions, Inc.), GE Water & Process Technologies, and Parker Performance Materials to acquire the most appropriate FO and MD membranes to characterize and test. Next, process models and controls for each of the subsystems were developed. Finally, OMBR and MD processes were integrated and a system-level model was developed and tested for water quality and membrane performance.

### Background

Membrane bioreactors (MBRs) have proven to be effective and beneficial for wastewater reclamation due to their ability to produce high quality effluent with a small footprint [6]. In a conventional MBR, microporous (microfiltration (MF) or ultrafiltration (UF)) membranes) are immersed in a bioreactor and water is filtered through the membranes using vacuum; suspended solids are retained in the system and high levels of treatment can be achieved [7, 8]. MBRs replace the two stages of the conventional activated sludge process with a single, integrated process that has a much smaller footprint. MBR effluent may be suitable for use as irrigation water, process water, or potable water. For potable reuse, advanced treatment (e.g., RO and chemical oxidation) is necessary after the MBR [9].

The advantages of MBRs over conventional treatment have been thoroughly reviewed [6] and include consistently high product quality, ease of operation/automation, reduced footprint, reduced sludge production due to a high biomass concentration in the bioreactor, and complete suspended solids removal from the effluent [6]. The main problem associated with MBRs is membrane fouling. Membrane fouling can occur in the MBR itself, and also in the downstream RO system [10]. Specifically, high concentrations of dissolved organic compounds in the MBR effluent can cause severe fouling of RO membranes.

A membrane sequencing batch reactor (SBR) was used for investigating the removal effectiveness of nutrients like nitrogen, phosphorus as well as some organic compounds [31-34]. The main feature of the sequential batch reactor technology is that both biological oxidation and sludge separation can be carried out in the same tank. But inadequate settling of sludge in SBR process can result in poor clarification and turbid effluent. To overcome this disadvantage, a membrane–coupled sequencing batch reactor was developed. In this system sludge separation is performed through membrane filtration which leads to complete biomass retention and high suspended solid concentration in the mixed liquor. Hollow fiber ultrafiltration membranes were used with an average pore size ranging between 0.2 and 0.4  $\mu$ m. Operating or working volume of the reactors ranged between 2 to 20 L. High nitrogen removal efficiencies were observed for the membrane sequencing batch reactors ranging between 87-96% [31, 33, 34]. This reactor was also effective in the removal of 35-90% of the phosphorus [31, 33] and about 99% of the TOC [33]. Arrojo et al. [32] observed removal of more than more than 95% of the organic matter, suspended solids, and coliform bacteria using membrane sequencing batch reactor.

Membrane bioreactor systems consisting of a sequential anoxic/anaerobic reactor (SAAR) were used to investigate the effect of solids retention time (SRT) and organic loading on membrane biofouling [35, 36]. The system provides alternating anoxic/anaerobic conditions with a membrane module immersed in the mixed liquor. The volume of the SAAR and aerobic reactor was 4 L and 6 L, respectively, for the studies conducted by Ahmed et al. [35] and Ahmed [36]. The membrane used for this study had an effective filtration area of 0.1 m<sup>2</sup> with a nominal pore size of 0.25  $\mu$ m. Sequential anoxic/anaerobic effectively removed 74% of the nitrogen and 89% of the phosphorus in the wastewater [36].

Li et al. [37] and Fu et al. [38] developed a hybrid/modified anoxic-oxic membrane bioreactor for investigating the microbial community and biomass characteristics associated with membrane fouling and nutrient removal. In hybrid anoxic–aerobic system, the activated sludge was alternated between aerobic and anoxic conditions using peristaltic pump. In the anoxic/oxic process, organic matter is mainly consumed in the anoxic stage which results in lower competition between the nitrifiers and hererotrophs in the following aerobic/oxic stage. In these studies, hydrophilic membranes were used with mean pore size of 0.1-0.22  $\mu$ m and effective filtration area of 0.1 m<sup>2</sup>. More than 90% of the nitrogen, phosphorus and carbon was removed by the modified anoxic/oxic membrane bioreactor [38]. Microbial community shift during the start-up period was observed that results severe membrane fouling [37].

Some researchers used other types of membrane bioreactors [39-43] for investigating the microbial community development, fouling tendencies, and nutrient removal with various organic loading conditions. Bing et al. [42] and Zhang et al. [40] investigated the impact of organic loading on membrane fouling and observed that membrane fouling tendency of biomass increases with increasing organic loading [42] and variable organic loading is more serious for

fouling than constant loading [40]. Bing et al. [38] also observed that membrane fouling propensity was strongly correlated with extracellular polymeric substance (EPS) concentrations.

Based on this literature review, the following typical MBR reactor characteristics have been gleaned from the literature:

- (a) Working volume (bench-scale units only): 2-30 L (typical range of 3-12 L)
- (b) Operational flux (for MF or UF membranes): 2.5-40 L/m<sup>2</sup>-hr (typical range of 20-30 L/m<sup>2</sup>-hr )
- (c) Effective membrane area:  $0.05-0.1 \text{ m}^2$  (majority of studies had A= $0.1 \text{ m}^2$ )
- (d) Comparison of oxic/anoxic (O/A) volumes: 3:2-2:1

A literature review was also conducted to establish the desired wastewater characteristics for the system feed. Based on this review, the following general characteristics for the feed wastewater have been established:

- (a) COD: range of 180-3,000 mg/L with anticipated value of 500 mg/L
- (b) Organic carbon constituents: glucose (250-1,000 mg/L), peptone (50-200 mg/L), acetate (250-1,000 mg/L)
- (c) Nutrient concentrations: total nitrogen (21-275 mg N/L) and total phosphorus (0.5-60 mg N/L)

Membrane fouling lowers productivity, increases energy requirements, increases frequency of membrane cleaning and replacement, and may result in deterioration of treated water quality [11]. A novel MBR system that utilizes a submerged FO membrane in the bioreactor and a membrane distillation (MD) membrane in a separate reconcentration process was investigated. Prior to describing the integrated system, the basic FO and MD processes are first outlined.

#### Forward Osmosis

Osmosis is the net diffusive transport of water through a selectively permeable membrane from a solution of low solute concentration (low osmotic pressure) to a solution of high solute concentration (high osmotic pressure). In osmosis, the membrane allows passage of water, but rejects almost all solute molecules and ions. FO is an engineered application of osmosis [12, 13].

In FO, impaired water is in contact with the dense side of a semipermeable membrane and a highly concentrated draw solution is in contact with the support side of the membrane. The draw solution is typically an aqueous solution of a low molecular weight salt [14]. In the FO process, relatively pure water is transported from the impaired water into the draw solution; the impaired water becomes concentrated and the draw solution becomes diluted. A desalination process (e.g., RO or distillation) can be used to reconcentrate the draw solution and simultaneously produce high-quality product water (Figure 2). Thus, in most water treatment applications, FO is not the ultimate process but rather a high-level pretreatment process before an ultimate reconcentration/desalination process [14].



Figure 2. Forward osmosis (FO) system schematic.

FO has been evaluated for seawater and brackish water desalination [12, 13, 15], wastewater concentration and reclamation [16-19], and food concentration [20-22]. FO can also be employed in conjunction with biological processes for wastewater reuse in osmotic membrane bioreactors (OMBRs) [23, 24]. The main advantage of FO is that it operates at very low hydraulic pressure, it has high rejection of a broad range of contaminants [23, 25], and may have lower fouling propensity and/or fouling that is more reversible than in RO processes [18, 26].

Osmotic backwashing was first suggested as a method for cleaning RO membranes by Spiegler and Macleish [44] in the early 1980's, and has more recently been suggested as a promising technique for cleaning FO membranes [18, 26, 45-51]. Osmotic backwashing utilizes the chemical potential gradient, between two streams of differing osmotic pressure to drive water transport across a semi-permeable membrane in the opposite direction of water transport during regular operation (Figure 3). Deionized (DI) water (high chemical potential) is used opposite the feed solution (moderate chemical potential) to dilute and dislodge foulants that are then removed from the membrane surface by the feed flow.

It has been suggested that the mechanics of FO membrane fouling are similar to, but not the same as those observed in RO membrane processes [52]. Whereas RO uses an applied hydraulic pressure to oppose and exceed the osmotic pressure of the feed solution (Figure 3.a), FO utilizes the osmotic pressure differential between the feed and draw solution as the sole driving force to transport water across a semi-permeable membrane (Figure 3.b). Because of this, FO has the potential to reduce energy consumption and increase water recovery in many applications. In their seminal review in 2006, Cath et al. [52] described the current state of the physical principles and applications of FO. Generally speaking, membrane fouling occurs due to either external/surface fouling or internal pore blocking, or a combination of both [45, 51, 52]. The foulants can be organic, inorganic, colloidal, or biological in nature for both internal and external fouling, although biological fouling is typically considered a surface phenomenon. The fouling in pressuredriven RO membranes is dominated by cake formation on the surface of the membrane. Because FO membranes are exposed to lower hydraulic pressure (only that required for water circulation, not as applied pressure), the external fouling is thought to be milder and the internal pore blocking to play a more significant role [45, 51-53]. Traditional chemical treatment methods typically target removal of the cake layer at the membrane surface only and do not remove foulants from the membrane pores, thus reducing their efficiency for use with fouled FO membranes. Because water is transported *through* the membranes from the permeate to the feed side during osmotic backwashing, the foulants within the pores are directly impacted. Osmotic backwashing also acts on the surface foulants via dilution, thus theoretically making it an effective cleaning tool for FO membranes.



**Figure 3.** Schematic representation of (a) normal reverse osmosis operation (water flow from feed side to permeate side) with applied pressure, (b) normal forward osmosis operation (water flow from feed side to draw side (draw solution)), and (c) osmotic backwashing (water flow from draw side (deionized water) to feed side).

There are two main methods for osmotic backwashing in RO systems: (1) under no or low pressure, and (2) with maintained operating pressure. In the first case, the applied pressure (Figure 3.a) is removed or reduced beyond the osmotic pressure difference between the permeate and feed, resulting in osmotic backwashing (Figure 3.c). Sagiv and Semiat [51] experimentally confirmed that the presence of operating pressure during the backwash procedure decreased the backwash water flux due to a reduction in the driving force. Their data also indicates that the backwash process has two distinct phases, with an initial rapid backwash flux followed by a continuously declining flux that approaches zero. The first stage was suggested to be a dilution of the concentration polarization (CP) layer, while the second stage was a dilution of the bulk phase. A simple mass balance-based model failed to fit the first stage, but was able to adequately predict the second stage of osmotic backwashing. Further work by the Semiat group [54-57] further confirmed that removing the operational pressure and reducing the chemical potential (as salinity) of the feed side increases backwash flux and membrane cleaning effectiveness. The second method of osmotic backwashing in RO systems involves injecting a high salinity solution into the feed solution for a short time, allowing for on-line operation. Liberman and Liberman [58] report that an osmotic pressure of up to 194 bar could be achieved through the injection of up to 25% NaCl solution. However, the flow rate, concentration, and duration of the high salinity solution must be well calculated with regard to the feed water chemistry and the RO pumping systems [58, 59]. A recent review by Qin et al. [59] of the osmotic backwashing of RO membranes concluded that osmotic backwashing offers a "novel environmental friendly on-line way to control fouling" by avoiding the use of aggressive chemical cleaning agents while still providing "strong and intensive cleaning effects".

Since the Cath et al. review [52], very little research effort has been devoted to evaluating the effectiveness of osmotic backwashing with FO systems. Holloway et al. [18] used osmotic backwashing as a cleaning method for a flat-sheet cellulose triacetate (CTA) FO membrane treating wastewater centrate. During osmotic backwashing, the draw solution was replaced with DI water and the feed solution was replaced with 50 g/L NaCl for 10 or 20 min. Both osmotic backwashing cycles restored the flux, with the longer cycle resulting in full flux recovery. The group also found that the fouling of the osmotic backwashed membrane was slower than was observed with the virgin membrane. It was suggested that the osmotic backwashing process adequately removed foulants from the membrane pores, but left a partial cake layer on the membrane surface that acted as a secondary membrane to reduce pore plugging in subsequent testing and resulted in a slower flux decline after osmotic backwashing. Achilli et al. [45] also utilized osmotic backwashing as a cleaning mechanism for a flat-sheet CTA membrane, this time in a novel osmotic membrane bioreactor consisting of the FO membrane inside a bioreactor. Osmotic backwashing was performed by replacing the draw solution with DI water and replacing the feed solution (synthetic waste water) with 5 g/L NaCl and circulating both sides for 1 hr at 1.5 L/min. The process was repeated three times, and flux was returned to ~90% of the original flux (from an ~20% decrease between each cleaning) through osmotic backwashing. In addition to flux recovery, no obvious effect of osmotic backwashing was observed in the reverse salt transport over the course of the experiment. Martinetti et al. [46] compared the effectiveness of osmotic backwashing to chemical cleaning of a flat-sheet CTA FO membrane. The chemical cleaning consisted of an initial feed side flush with DI water (4 L) to remove loose deposits, followed by treatment with 2 L of cleaning solution (0.029 M Na<sub>2</sub>EDTA and 0.058 M NaOH) for 30 min at full flow and then an additional 1 hr at low flow, and then a final DI rinse (8 L). The osmotic backwashing cleaning consisted of replacing the draw solution with DI water and replacing the feed solution with 100 g/L NaCl and circulating both solutions for 20 min. Flux data after cleaning indicated that osmotic backwashing was equally as effective in removing scale deposited on the membrane as was chemical cleaning, as the initial flux was equivalent after cleaning for both methods. However, flux decline occurred more rapidly after the osmotic backwashing cleaning, and some scale particles were observed to be flushed through the system after the cleaning process, suggesting that the osmotic backwashing effectively removed the scale from the membrane surface but did not remove the scale from the system or degrade the scalant. These results indicate the importance of flushed from the membrane surface is removed prior to re-starting normal operation.

More recently, Hickenbottom et al. [47] used FO to reclaim water from drilling and fracturing wastewater from oil and gas operations. In this work, hydraulic flushing with DI water on the feed side was compared to osmotic backwashing, where the draw solution was replaced with DI water and the feed solution was unchanged. Water flux was restored with both methods, however osmotic backwashing was more efficient at breaking up the foulant as determined by a shorter recirculation time (~30%) required to remove the visible fouling layer. Finally, Lay et al. [48] used osmotic backwashing as a cleaning method for comparing performance of two FO membranes: a novel thin film composite (TFC) hollow fiber membrane and a flat-sheet CTA membrane. Osmotic backwashing was performed by replacing the draw solution with DI water and circulating for 1 hr, and was found to be more effective with the more severely fouled membranes. Flux was restored to 79% of the unfouled flux (from 56%) with the TFC membranes, and to 87% (from 82%) with the CTA membranes using osmotic backwashing. The addition of aeration via bubbling during osmotic backwashing was also found to improve foulant removal, presumably by providing additional shear force across the membrane surface.

Several groups have used model species to better understand the underlying fouling/cleaning mechanisms associated with FO membranes. Arkhangelskym et al. [49] examined several cleaning techniques for inorganic scaling (CaSO<sub>4</sub>) of a flat-sheet CTA membrane. Surface flushing and osmotic backwashing procedures were performed for 1 hr with DI water and a cross-flow velocity of 48 cm/s. Hydraulic backwashing was also evaluated with 1 bar overpressure and a flowrate of 450 mL/min. Flux recovery was not recovered using surface flushing or osmotic backwashing when the membrane was fouled with CaSO<sub>4</sub>. Hydraulic backwashing did result in 75% flux recovery, however, the flux declined rapidly after cleaning with the inorganic foulant. This work suggests that the type of foulant plays a significant role in the reversibility of fouling, especially when using non-pressure driven cleaning mechanisms such as osmotic backwashing or flushing.

The most thorough evaluation of fouling mechanisms and osmotic backwashing removal was performed by Kim et al. [50]. They evaluated both alginate (200 mg/L) and humic acids (200 mg/L) as model organic foulants and colloidal silica particles (2 g/L) with different sizes (20 and 100 nm) as model inorganic particulate foulants. The group used 0.6 M NaCl as the feed solution to simulate sea water with a flat-sheet CTA membrane in two membrane orientations: active layer facing feed solution (AL-FS) and active layer facing draw solution (AL-DS, also known as pressureretarded osmosis mode [60]). The draw solution strength was adjusted so that the initial flux was the same for both orientations, and osmotic backwashing was performed by replacing the draw solution with DI water for 30 min. With the alginate, ~85% flux recovery was achieved with both membrane orientations, which agreed well with the flushing results of Mi and Elimelech [26]. In the AL-FS orientation, negligible fouling occurred with the humic acid (>95% flux after fouling), however significant fouling was observed in the AL-DS orientation. It was suggested that the hydrophobic nature of humic acid enhanced the chemical interaction with the FO membrane surface, thereby increasing the humic acid attachment to the inside pores of the porous support material that was directly in contact with the feed (AL-DS mode). Osmotic backwashing of the HA-fouled membrane in the AL-DS orientation also had minimal flux recovery (~71% from ~69%), which was much lower than what was observed with the alginate. When the FO membrane was fouled with the smaller (20 nm) silica particles, flux decline was similar to that observed with humic acid fouling: only ~5% flux decline was observed in the fouled membrane in the AL-FS orientation, but flux dropped to ~70% in the AL-DS orientation. Osmotic backwashing resulted in a slightly higher flux recovery with the small silica particles (~79%). With the larger particles, <70% flux was observed in the fouled membranes with both orientations, however the osmotic backwashing was more effective at restoring flux with these larger particles

compared to the smaller particles; flux was restored to ~84% in AL-FS and ~80% in AL-DS. Generally, less fouling was observed for the AL-FS mode, which has been observed by others [53, 60] and has been suggested is due to the ineffectiveness of the shear force exerted by cross-flow to remove the foulants entrapped in the porous support layer of the membrane in the AL-DS orientation.

The results of the literature review confirm that osmotic backwashing efficiency is highly dependent on the nature of the foulant. FO membranes fouled with two different types of organic material (alginate vs. humic acid) and two different sized colloidal particles were all affected differently by osmotic backwashing. To date, the effectiveness of osmotic backwashing for removal of inorganic scale and biological fouling have not been explored: both of these foulant types will likely to be an issue in the OMBR treating waste water. Thus, a more systematic study of is necessary to validate the use of osmotic backwashing as a cleaning mechanism for FO membranes.

#### Membrane Distillation

MD is a non-pressure-driven membrane separation process that has the potential to become a viable tool for water desalination and brine (or draw solution) concentration. MD involves the transport of mass and heat through a hydrophobic, microporous membrane. In direct contact MD (DCMD) (Figure 4), a feed solution at elevated temperature is in contact with one side of the membrane and a permeate solution at lower temperature is in direct contact with the opposite side of the membrane; it is mainly the temperature difference between the liquids that induces the vapor pressure gradient for mass transfer. DCMD is one of the simplest configurations of MD; it requires only a membrane module and two low-pressure pumps to move the liquids over the membrane. The simplicity of this configuration makes it highly suitable for implementation in large-scale applications.

Compared to conventional distillation methods, MD requires only small temperature differences – temperature differences achievable through the use of low-grade heat/renewable energy. Compared to RO, the driving force in MD is essentially not reduced by osmotic pressure and thus, MD operates at hydrostatic pressure and can be used to treat high salinity solutions or provide enhanced recovery through brine desalination. Also compared to RO, MD produces an even higher quality distillate; it does not allow the passage of small non-volatile compounds and provides high removal of emerging contaminants. Specifically, MD has been shown to reject greater than 97% of low molecular weight organics [27]. For these reasons, MD may have substantial energy, recovery, maintenance, and water quality advantages over conventional methods.



Figure 4. DCMD: feed and permeate solutions are both in direct contact with the membrane.

One way to improve energy efficiency is to include a heat exchanger for recovery of heat from the distillate to the feed ("heat recovery exchanger"). As was demonstrated earlier, membrane modules must be used in series to allow for a heat recovery exchanger to be used. The use of membrane modules in series instead of in parallel for the same

module velocity also results in higher recovery and a higher GOR, defined as the ratio of the kg of product water to the kg of input steam. GOR is evaluated a number of different ways, typically either as a mass or energy ratio, but the energy ratio definition (Eq. 1 without heat recovery exchanger or Eq. 2 with heat recovery exchanger) has been selected because it accounts for the thermal operating conditions. The mass ratio definition assumes full use of the total heat content of the input steam and does not account for heat transferred to the product water [62]. A higher GOR typically indicates a more energy efficient system with lower operating costs but larger capital costs [62].

$$GOR = \frac{\Delta h_{lat} \dot{m}_{product}}{\dot{m}_{feed} C_p (T_{feed,in} - T_{feed,out})}$$
Equation 1)

$$GOR = \frac{\Delta h_{lat} \dot{m}_{product}}{\dot{m}_{feed} C_p (T_{feed,in} - T_{HX,out})}$$
(Equation 2)

One very important aspect of the MD process is the membrane cannot become hydrophilized. Hydrophilization can occur due to fouling of the MD membrane if proper pretreatment is not used. fo provides outstanding pretreatment of feed streams to MD membranes because it provides high rejection of dissolved species [18, 23].

Maintaining membrane hydrophobicity is absolutely essential for assuring that the pores of MD membranes do not become flooded. If hydrophobicity is lost, the pores will become flooded with water and salts and other contaminants will diffuse into the distillate. Flooding is irreversible, in other words, after the pressure that caused flooding is removed, the pores remain wet. In order to restore permeate quality, removal of fluids and drying is required. Pore flooding results in a decline in distillate quality and downtime and prevention of this phenomena is absolutely critical for successful MD scale-up. This task provided insight on membrane characteristics and operating conditions that minimize pore flooding. Results from this task provided the information necessary to determine the optimal MD membrane and operating conditions to be used in the OMBR.

Membrane characteristics data were collected and analyzed from a previous study performed at UNR. The UNR study characterized a number of MD membranes prototypes, of different materials and structure, provided by General Electric (GE) Water & Process Technologies, Donaldson Filtration Solutions, and GORE (Table 1). The membrane structures are either single layer or composite, with active layers consisting of either polytetrafluoroethelyne (PTFE) or polypropylene (PP). The reported support layer material for all membranes is PP, though the support layer material was not specified for all membranes. The support layer structure for the membranes is either scrim or non-woven, though the support layer structure was not specified for many of the membranes. The susceptibility of a membrane to hydrophilization is an important factor in its long-term use. MD requires separation of water on both sides of the membrane to create a vapor pressure difference that causes water vapor to transfer through the membrane pores. When hydrophilicity increases to the point of pore flooding, salts and other contaminants will diffuse into the distillate. Membranes with high hydrophobicity are therefore desirable. Parameters describing the hydrophobicity of the membranes have been collected to aid in the selection of the optimal MD membrane for use in the OMBR system (Table 2).

**Table 1.** Properties of membranes as provided by manufacturer.

Membrane	Structure	Active material	layer	Support material	layer	Support structure	layer
AX	Single-layer	PTFE					
TX	Single-layer	PTFE					
PP22	Single-layer	PP					
PTFE45	Composite	PTFE		PP		Scrim	
QM022	Single-layer	PTFE					
QP909	Composite	PTFE		PP		Non-woven	
QL236	Composite	PTFE		PP		Non-woven	

QL822	Composite	PTFE	PP	Non-woven
QL211100	Composite	PTFE	PP	Non-woven
Gore 1	Single-layer	PTFE		
Gore 2	Composite	PTFE	PP	Scrim
Gore 3(NW)	Composite	PTFE	PP	Non-woven

The membranes have been characterized in terms of their average pore size, thickness, porosity, tortuosity, contact angle, and liquid entry pressure (LEP) (Table 2). The average pore size varies from 0.21 to 0.91  $\mu$ m. The membrane thickness ranges from 39+/-5 to 289+/- 10  $\mu$ m. Porosity of the membranes varies from a minimum value of 46.6+/-8.5 to a maximum value of 90.3+/-3.2 %. Tortuosity, a measure of a membrane's effective diffusivity, ranges from 1.33+/-0.19 to 5.05+/-0.97. Contact angle, the angle between a drop of water and the membrane surface to which it is attached, varies from 106+/-6 to 131+/-3°. LEP, the pressure that must be overcome for pore flooding to occur, ranges from 95 to 430 kPa. The characterization data will be used in determining the optimal membrane for use in the OMBR system.

Membrane	Average pore size (µm)	Thickness (µm)	Porosity (%)	Tortuosity	Contact angle (°)	LEP (kPa)
AX	0.39*	119±11	84.5±7.8	1.58±0.20	$130\pm5$	430
ТХ	0.27*	111±4	84.1±3.2	$1.60\pm0.26$	$123 \pm 2$	358
PP22	0.32	205±4	82.6±2.2	1.67±0.26	$123 \pm 2$	375
PTFE45	0.46	108±15	56.1±8.8	$3.68 \pm 0.28$	$120\pm5$	415
QM022	0.23	67±17	80.1±18	$1.79\pm0.55$	$131 \pm 3$	385
QP909	0.74*	203±8			$118\pm5$	190
QL236	0.44*	172±6	80.9±6.7	$1.75\pm0.18$	$122 \pm 5$	330
QL822	0.21	202±9	46.6±8.5	$5.05 \pm 0.97$	$106 \pm 6$	290
QL211100	0.91*	289±10	90.3±3.2	1.33±0.19	$126 \pm 1$	95
Gore 1	0.22	39±5	53.4±17	4.02±0.31	$125\pm2$	380
Gore 2	0.23	234±11	77.4±9.8	1.83±0.24	$125 \pm 2$	395
Gore 3(NW)	0.25	183±15	78±6.8	2.03±0.22	$125 \pm 2$	380

Table 2. Hydrophobicity of MD membranes.

\* further verifications are required.

#### The Integrated OMBR System

In an OMBR system (Figure 5), wastewater is fed into a reactor which is aerated as needed to supply oxygen for the biomass and to minimally scour the membrane. Through osmosis, water diffuses from the bioreactor, across a semipermeable membrane, and into the draw solution. The FO membrane acts as a barrier to solute transport and provides high rejection of contaminants in the wastewater stream [16, 19, 25]. The diluted draw solution is sent to a reconcentration process (e.g., RO or MD), which reconcentrates the draw solution and generates a high-quality product water. Thus, FO serves as a high-level pretreatment step before the RO or MD reconcentration. Compared to the MF or UF process in a conventional MBR, the FO process in the OMBR offers the advantages of much higher rejection
(semi-permeable membrane versus microporous membrane) at lower hydraulic pressure. The FO membrane inside the bioreactor also has much lower fouling propensity than the MF/UF membranes, and thus, requires minimal scouring and much less frequent backwashing. When comparing an OMBR system (OMBR followed by RO or MD reconcentration) with a conventional MBR-followed-by-RO system, the high rejection of the FO membrane results in RO influent with lower fouling propensity and ultimately produces higher quality product water. Furthermore, the FO-followed-by-RO or -MD reconcentration treatment scheme represents a tight dual barrier system. This is particularly important when pharmaceutical or personal care products are present.



Figure 5. OMBR system.

The main objective of the project was to develop a highly efficient wastewater treatment and water reuse system. To achieve this goal, the three major processes of the OMBR system (BIO, FO, and MD) were systematically investigated. The energy requirements for each process were studied in parallel. Subsequently, laboratory-scale versions were designed, fabricated, and integrated into a single system. The modular system was designed to operate using low-grade ("waste") heat existing at FOBs. The abundance of low-grade heat produced by power generators at FOBs make the OMBR-MD system ideal for such locations.

#### Development of an Oxic-Anoxic Bioreactor for Treating FOB Wastewater

A simultaneous nitrification-denitrification process was tested and optimized. The bioreactor used for these experiments was inoculated with a combination of activated sludge, bacteria from a nitrifying trickling filter biomass, and microbes from a denitrifying system at a regional wastewater treatment facility. Bench-scale experiments were conducted in a pair of bioreactors specially designed to accommodate a flat sheet membrane. The two process waste streams were tested in parallel using the two systems: (i) graywater influent [28] and (ii) combined gray+black wastewater (based on typical domestic wastewater composition). For purposes of this subtask, the unit mimicked a traditional MBR system in that it had an MF membrane solely for the purpose of retaining biomass. In subsequent research tasks, the biological system was combined with the FO system, which served to retain more than biomass. Since the FO system in subsequent tasks contains a semi-permeable membrane that is similar to a RO membrane, both suspended and dissolved constituents can be retained within the system. In the current task, which utilizes an MF membrane (pore opening size ~0.1-0.2  $\mu$ m), only suspended constituents larger than this size were retained. The bioreactor was operated in alternating aerobic/anoxic reaction modes to allow for carbon and nitrogen removal. The following general, non-balanced chemical reactions (Eqs. 3-5) describe the major reactions occurring in each mode. Operation in this fashion allowed simultaneous urea hydrolysis (Eq. 3), nitrification (Eq. 4), and denitrification (Eq. 5), which ultimately removed dissolved contaminants.

Aerobic & Anoxic: 
$$Urea + Microorganisms \rightarrow NH_3 + CO_2$$
 (Equation 3)

Aerobic only:	$NH3 + O_2 \rightarrow NO_2^{-}/-/NO_3^{-} + H_2O$	(Equation 4)
Anoxic only:	<i>Organic Carbon</i> + $NO_2^{-}/NO_3^{-} \rightarrow N_2 + CO_2 + H_2O$	(Equation 5)

Biochemical experimental variables monitored during the experiments included DOC, ammonia-nitrogen ( $NH_3$ -N), nitrite-nitrogen ( $NO_2^{-}$ -N), nitrate-nitrogen ( $NO_3^{-}$ -N), and biomass (measured as volatile suspended solids, VSS). Chemical parameters were measured in the feed stream, within the bioreactor during reaction cycles, and in the final effluent stream using accepted analytical methods.

Microbial populations and ecology (see below) were monitored at reasonable long-term intervals (i.e., 15-30 days). To be effective at achieving complete treatment, cycle duration and switching between aerobic and anoxic conditions were optimized. While these processes were relatively easy to optimize individually, it was technically challenging to engineer these processes such that they occurred in a single system and consumed a minimum of resources (specifically, oxygen and exogenous resources, such as energy for mixing, internal recirculation, and membrane cleaning).

Shortcut denitrification is a process where the nitrogen in ammonia is only oxidized to nitrite  $(NO_2)$  which is subsequently reduced to nitrogen gas during denitrification. This process relies on less oxygen (and consequently less energy) than traditional nitrification-denitrification sequences. As such, a major thrust was to perform fundamental research on a single-sludge process and study the microbiological community shifts that occurred throughout the system. The central hypothesis that was addressed during this research phase was: a balanced nitrificationdenitrification treatment process can be achieved with carefully tailored oxic and anoxic reaction times based on chemical signatures in the bioreactor. Thus, a major component of this task was to identify proper chemical cues (including water quality and biochemical indicators) that were required to switch between operational modes. Specifically, oxidation-reduction potential (ORP), DO, and pH were continuously analyzed along with water quality parameters (i.e., ammonia, nitrite, nitrate, total organic carbon (TOC), and chemical oxygen demand (COD)) to determine appropriate switching times between aerobic and anaerobic reaction sequences.

Simulated FOB wastewater was developed to challenge the microbial reactor system and enable researchers to rigorously identify the switching requirements for achieving maximum nutrient removal. The key FOB wastewater constituents that were mimicked in the laboratory study were organic carbon load (measured as total organic carbon and COD), nitrogen speciation and respective constituent concentrations (urea and ammonia), typical nutrients and micronutrients (phosphorus, iron, sulfur), and suspended solid concentrations (measured as total suspended solids (TSS) and VSS).

# Development of an Integrated Sensor-Based Control System for the Bioreactor

A control and operating program for the biological process of the OMBR system was developed; specifically, the program allowed biological nutrient removal in a single reactor. The program relied on DO, oxidation-reduction potential, pH, and time to switch between oxic and anoxic phases to achieve nitrification and denitrification, respectively. This subtask was built upon data collected in the previous task, which describe changes in biochemical reaction during wastewater treatment.

A LabVIEW-controlled system was added to the bioreactor system used during the previous task and a computercontrolled system was developed to allow the aeration system to automatically turn on and off as a function of bioreactor probe outputs. Briefly, experimental data provide a suite of ORP, DO, and pH performance curves that were correlated with biochemical environment (specifically dissolved nitrogen speciation and organic carbon concentrations). Based on these data, real time switching criteria were incorporated into the LabVIEW-controlled system to control and maintain proper air delivery to the bioreactor.

#### FO Membrane Fouling, Process Modeling, and Control

FO uses osmotic pressure differential across a semipermeable membrane, rather than hydraulic pressure differential (as in RO), as the driving force for water transport through the membrane. The FO process results in concentration of a feed stream and dilution of an osmotic agent (i.e., a draw solution).

There is evidence that in some cases, fouling occurs to a lesser extent in FO than in RO [18]. In other cases, it has been found that fouling occurs at similar rates in FO and RO; however, the fouling is more reversible in FO – possibly due to the absence of hydraulic pressure in the FO feed stream [27]. Although there is indication in the literature that the presence of salt at the membrane surface may play a role in membrane cleaning [29], the basis for the more reversible fouling has not been proven. This was explored to provide critical insight into FO transport phenomena and fouling reduction. FO membranes were used to conduct bench-scale fouling and cleaning experiments. A range of foulants (organic, inorganic, colloidal, and biological) was used to assess the ability to remove these foulants using rinsing and osmotic backwashing. During osmotic backwashing, water passed from the support side of the membrane to the active side, thereby reversing the direction of flow through the FO membrane and likely removing foulants. Foulant removal was determined by hydraulic testing and scanning electron microscopy (SEM) imaging.

#### MD Membrane Hydrophilization, Process Modeling, and Control

Maintaining membrane hydrophobicity is absolutely essential for assuring that the pores of MD membranes do not become wet. If hydrophobicity is lost, the pores can wet with water and salts and other contaminants can diffuse into the distillate [30]. Pore wetting is considered irreversible, in other words, after the pressure that caused pore wetting is removed, the pores remain wet. In order to restore permeate quality, removal of fluids and drying is required. Pore wetting results in a decline in distillate quality; downtime and prevention of this phenomenon is absolutely critical for successful MD scale-up.

The interrelationship between hydrophobicity and pore size distribution is described in terms of the LEP. LEP is given by the difference between liquid pressure and vapor pressure:

$$P_{l} - P_{V} \le -\frac{2\gamma_{L}\cos\theta}{r} = LEP$$
 (Equation 6)

where  $P_1$  is the pressure of the liquid in the channels of the MD module,  $P_y$  is the pressure of the vapor inside the membrane pores,  $\gamma$  is liquid surface tension,  $\theta$  is contact angle, and r is pore radius (for non-cylindrical pores, a geometry correction factor needs to be introduced). Theoretically, the membrane pores remain dry as long as the LEP is not exceeded.

#### **Energy Considerations for MD**

In this task, the MD system was analyzed using commercial computational fluid dynamics (CFD) tools to determine energy and flow requirements and to optimize MD system design. The thermal and hydraulic characteristics of the MD system and its heat exchanger system were numerically analyzed using ANSYS Fluent, a commercial CFD package. The operational conditions and dimensions based on the small flat-sheet MD system under testing were considered for the CFD analysis to calculate energy and flow requirements of the MD system.

The heat available in the engine cooling system, q, is given by Equation 7:

 $q = \left(\frac{\eta_{EG}}{\eta_{P}}\right)P$ (Equation 7)

The energy balance in the MD system is given by Equation 8:

 $q = q_e + q_c$ (Equation 8)

where, q is the heat carried by the feed water stream for the phase change and conduction heat transfer,  $q_e$  is the heat transfer required for the phase changes (evaporation and condensation) in the membrane,  $q_c$  is the conduction heat transfer through the membrane, Equations 9-13 give  $q_e$ ,  $q_c$ , and  $q_s$ , respectively.

 $q_e = \dot{m}_m \Delta h_{fg} \tag{Equation 9}$ 

$$q_c = k_{eff} A_m \frac{\Delta T}{\delta_m}$$
 (Equation 10)

$$q = \dot{m}_{F} c_{p,F} T_{F,i} - (\dot{m}_{F} - \dot{m}_{m}) c_{p,F} T_{F,o}$$
(Equation 11)

where  $m_m$  is defined with Equation 12 as:

 $H = \frac{N}{2} \left( a + 2\delta_p \right)$ 

$$\dot{m}_m = JA_m \rho_F \tag{Equation 12}$$

where J is the permeate flux across the membrane. J was assumed to be equal to 15 L/s-m<sup>2</sup>.  $A_m$  is the area of the membrane,  $\rho_f$  is the density of the feed water,  $h_{fg}$  is the latent heat of the feed water,  $k_{eff}$  is the effective thermal conductivity of the membrane,  $A_m$  is the thickness of the membrane,  $m_f$  is the flowrate of the feed water,  $c_{p,F}$  is the specific heat capacity of the feed water, and  $T_{F,i}$  and  $T_{F,o}$  are the inlet and outlet temperatures of the feed water stream, respectively. The effective thermal conductivity of the membrane is given by Equation 13:

$$k_{eff} = (1 - \varepsilon)k_m \qquad (Equation 13)$$

where  $\varepsilon$  is the porosity of the membrane and was assumed to be 0.7 and  $k_m$  is the thermal conductivity of the membrane and was assumed to be equal to thermal conductivity of Teflon.

The Reynolds and Prandtl numbers for the feed water (cold fluid),  $Re_c$ ,  $Pr_{ic}$  and engine coolant (hot fluid),  $Re_h$ ,  $Pr_{ih}$  are given by Equations 14 and 15:

$$\operatorname{Re}_{i} = \frac{\left(\frac{2\dot{m}_{i}}{N}\right)D_{h}}{\mu_{i}A_{c}}$$
(Equation 14)  
$$\operatorname{Pr}_{i} = \frac{c_{p,i}\mu_{i}}{k_{i}}$$
(Equation 15)

where i can be either c, the cold stream, or h, the hot stream,  $\mu$  is the dynamic viscosity [70]. A<sub>c</sub> is the cross-sectional area of the channel, D<sub>h</sub> is the hydraulic diameter, c<sub>p</sub> is the specific heat capacity and k is the thermal conductivity of the fluid. The cross-sectional area of each channel and height of the heat exchanger are given by Equations 16 and 17:

 $A_c = aW \tag{Equation 16}$ 

(Equation 17)

The Nusselt number (Nu) and friction factor (f) of the flat plate heat exchanger is given by Equations 18 and 19 [71]:

Nu = 
$$\begin{cases} 1.67 \,\text{Re}^{0.44} \,\text{Pr}^{0.5} & 45 < \text{Re} < 300 \\ 0.405 \,\text{Re}^{0.7} \,\text{Pr}^{0.5} & 300 < \text{Re} < 2000 \\ 0.84 \,\text{Re}^{0.6} \,\text{Pr}^{0.5} & 2000 < \text{Re} < 20000 \end{cases}$$
(Equation 18)

$$f = \begin{cases} 0.3025 + 91.75 / \text{Re} & 150 < \text{Re} < 1800 \\ 1.46 \,\text{Re}^{-0.177} & 1800 < \text{Re} < 30000 \end{cases}$$
(Equation 19)

The effectiveness,  $\varepsilon_{HX}$ , of a heat exchanger is the ratio of the actual heat transfer rate to the maximum possible heat transfer rate of the heat exchanger. As the effectiveness reaches 1, its performance becomes closer to the theoretical maximum. The effectiveness of the heat exchanger is given by Equation 20:

$$\varepsilon_{HX} = \frac{C_h(T_{h,i} - T_{h,o})}{C_{\min}(T_{h,i} - T_{c,i})}$$
(Equation 20)

Where  $C_h$  is the heat capacity rate ( $C_h = m_h \times c_{p,h}$ ) of the hot fluid and  $C_{min}$  is the minimum heat capacity rate with respect to the hot and cold fluid heat capacity ( $C_c = m_c \times c_{p,c}$ ) rates (minimum of  $C_h$  and  $C_c$ ) [71].

The DCMD process is relatively energy-intensive because the process is based on evaporation and condensation [73-75]. Hot feed water is brought into contact with one side of the membrane, while cold distilled water is in contact with the other side. Evaporation occurs at the feed side of the membrane, while condensation occurs at the distillate side of the membrane. The hydrophobic nature of the membrane prevents the water from penetrating the pores of the membrane. The vapor flow in the membrane pores is driven by the pressure difference between the evaporation and condensation.

#### Flow Regimes and Governing Equations

The mass transfer in MD systems includes three steps. The water vaporizes at the menisci at the hot side (feed stream) of the membrane. It then passes through the nano/micro porous membrane and condenses at the menisci at the cold side (distill stream). Researchers have developed several correlations to model the mass transfer through the nano/micro porous membrane using the vapor pressure difference across the membrane [76]:

$$\dot{m}^{\prime\prime} = C[P_{sat}(T_f) - P_{sat}(T_p)]$$
(Equation 21)

where C is the membrane distillation coefficient.

In nano/micro-scale systems, the characteristic dimension of the flow channel is comparable to the mean free path of the gas. The mean free path is given by

$$\lambda = \frac{k_B T}{\sqrt{2\pi} d^2 P} \tag{Equation 22}$$

where  $k_B$  is the Boltzmann constant, d is the hard shell diameter of the molecule, T is the temperature, and P is the pressure. Under such conditions the continuum assumption breaks down. Deviation from the continuum hypothesis is defined with Knudsen number, *Kn*, which is the ratio of mean free path and the length scale:

$$Kn = \frac{\lambda}{D}$$
 (Equation 23)

The flow regimes can be classified into various regimes depending on the Knudsen number. These regimes are: continuum, slip, transition, and free-molecular flow. The discrete particle or molecular-based model utilizes Boltzmann's equation, which is an integral-differential equation and the solution of this equation is limited to a few cases. The continuum based models are governed by the Navier-Stokes equations. Euler's equation corresponds to an inviscid continuum limit that shows a singular limit since the fluid is assumed to be inviscid and non-conducting. Euler flow corresponds to Kn=0.0. The Navier-Stokes equations can be derived from the Boltzmann equation using the Chapman-Enskog expansion.

When the Knudsen number is larger than 0.1, the Navier-Stokes equations break-down and a higher level of approximation is obtained by carrying second order terms (in Kn) in the Chapman-Enskog expansion. A special form of such an equation is called the Burnett equation, for which the solution requires accurate second-order slip boundary conditions in the Kn. The Burnett equation and consistent second-order slip boundary conditions are subject to some controversy and a better way of solving high Knudsen number flow is through molecular-based direct simulation techniques such as the Direct Simulation Monte Carlo method (DSMC).

The process of mass transfer through membrane can be modeled as gas transport through nano/micro-scale porous media, which can be divided into the following three mechanisms depending on the flow regimes.

- 1. Knudsen-diffusion the gas density is significantly low and the pore diameter is very small, therefore the collision between molecules is ignored and only the collisions of the gas molecules with the walls are considered.
- 2. Molecular-diffusion the collisions occurring between molecules dominate over the collisions occurring between the molecules and walls.
- 3. Poiseuille-flow in this mode the gas behaves as a continuous fluid that is driven by a pressure gradient. The collisions between molecules dominate over the molecule-wall collisions.

Schofield presented the membrane distillation coefficient based on these three modes of gas transport in porous media [77]: for Knudsen-diffusion,

$$C_1 = 1.064 \frac{R\varepsilon}{\tau\delta} \left(\frac{M}{R_u T_m}\right)^{0.5}$$
 (Equation 24)

for molecular diffusion.

$$C_2 = \frac{1}{y_{ln}} \frac{D\varepsilon}{\tau \delta} \left( \frac{M}{R_u T_m} \right)$$
(Equation 25)

and for Poiseuille-flow,

$$C_3 = 0.125 \frac{R^2 \varepsilon}{\tau \delta} \left(\frac{M P_m}{\mu R_u T_m}\right)^{0.5}$$
(Equation 26)

where r is the pore radius,  $\varepsilon$  is porosity,  $\tau$  is the tortuosity,  $\delta$  is the thickness of the membrane, M is the molecular weight of the vapor, R is the universal gas constant,  $T_m$  is the mean temperature across the thickness of the membrane,  $y_{in}$  is the mole fraction of air in the membrane pore, D is the pore diameter,  $P_m$  mean pressure across the thickness of the membrane, and  $\mu$  is the viscosity of the vapor. It is clear that the mass transfer in membrane is affected by the characteristics of the membrane and also the membrane temperature.

#### Analytical Solution

Here, the porous structure of the membrane is assumed to consist of microchannels with circular cross sections. It is assumed that a rarefied gas flows in the microchannel. The flow can be modeled using Navier-Stokes equations with slip-boundary conditions because the flows in microchannels are typically in slip and early transitional flow regimes.

 $\mu \frac{1}{r} \frac{\partial}{\partial r} \left( r \frac{\partial u}{\partial r} \right) = \frac{dP}{dx}$ (Equation 27)

where r is the variable in radial coordinate and x is the variable in axial coordinate. The slip boundary condition is given by

$$u_{g,w} - u_w = \frac{2 - \sigma_v}{\sigma_v} \lambda \left(\frac{\partial u}{\partial r}\right)_w$$
(Equation 28)

where  $u_{g,w}$  is the velocity of fluid near the wall,  $u_w$  is the wall velocity, and  $\sigma_v$  is the momentum accommodation coefficient, which is assumed to be unity.

The momentum equation given in Eq. (27) is solved using the velocity slip boundary condition, Eq. (28). The velocity profile is given by

$$u_g(r) = -\frac{1}{\mu} \frac{dP}{dx} R^2 \left[ \frac{1}{4} \left( 1 - \left( \frac{r^2}{R^2} \right)^2 \right) + \frac{2 - \sigma_v}{\sigma_v} \operatorname{Kn} \right]$$
(Equation 29)

The mass flux can be derived by integrating the velocity over the cross-sectional area. The mass flux is given by

$$\dot{m}^{\prime\prime} = \frac{1}{\pi R^2} \int \rho \, u_g(r) 2\pi r dr \qquad (Equation 30)$$

#### CFD Modeling

The flow of the rarefied gas inside a circular microchannel was modeled using ANSYS Fluent 12.1 to find the flow characteristics considering a pressure-driven flow with velocity slip boundary conditions. Knowing that the shear stress at the wall is

$$\tau_w = \mu \left(\frac{\partial u}{\partial r}\right)_w \tag{Equation 31}$$

Substituting the boundary condition, Eq. (28), yields the shear stress for the model being considered.

$$\tau_{w} = \mu \frac{u_{c}}{\delta_{n} + \frac{2 - \sigma_{y}}{\sigma_{y}} D \mathrm{Kn}}$$
(Equation 32)

where the variables used in the above correlations are defined in Figure 6.



Figure 6. Definition of slip velocities for the cell adjacent to the wall

The properties of the membrane used for the analytical and CFD analyses are presented in Table 3.

Table 3. Physical properties of the membrane considered for this analysis

Membrane	Pore diameter $[\mu m]$	Thickness [µm]	Porosity[%]	Tortuosity
QM022	0.44	40	62	3.07

It is assumed that the distillate stream temperature is fixed at 20°C. This value was deemed reasonable based on two available experimental results for the feed stream temperature. The numerical results will be compared with the results of the three different models of mass transfer for the MD membrane and also with the experimental data. The overall mass flux that was provided from experimental data was modified by the porosity of the membrane to get the mass flux per channel,

$$\dot{m}_{ch}^{\prime\prime} = \frac{\dot{m}_{ex}^{\prime\prime}}{\varepsilon}$$
(Equation 33)

# Heat Exchanger Design/Sizing

The continuous fin heat exchanger was sized based on the effectiveness – number of transfer units ( $\varepsilon$ -NTU) method. A simplified model of the heat exchanger is shown in Figure 7 (a) along with the characteristic dimensions used in this analysis. NTU is a dimensionless parameter that is widely used in heat exchanger analysis. The number of transfer units is also a function of the heat exchanger's effectiveness,  $\varepsilon_{HX2}$ . For a single pass, cross-flow heat exchanger the NTU can be expressed as the following:

$$NTU = -\ln\left[1 + (1/C_r)\ln(1 - \varepsilon C_r)\right], \text{ for } C_{\max} \text{ (mixed), } C_{\min} \text{ (unmixed)} \qquad (Equation 34)$$
$$NTU = -(1/C_r)\ln\left[C_r\ln(1 - \varepsilon) + \right], \text{ for } C_{\min} \text{ (mixed), } C_{\max} \text{ (unmixed)} \qquad (Equation 35)$$

where the feed water is the unmixed fluid and the exhaust gas is the mixed fluid.  $C_{min}$  is equal the hot or cold capacity rate,  $C_c$  or  $C_h$ , whichever is smaller,  $C_{max}$  is equal the hot or cold capacity rate,  $C_c$  or  $C_h$ , whichever is larger, and ,  $C_r$  is the heat capacity ratio  $(C_{min}/C_{max})$ .

The effectiveness of a heat exchanger is the ratio of heat transfer to the theoretical maximum heat transfer and is given by

$$\varepsilon_{HX2} = \frac{q_2}{C_{\min} \left( T_{h2,i} - T_{F2,i} \right)}$$
(Equation 36)

The feed water convection coefficient was determined using the appropriate correlation for laminar or turbulent flow inside a circular tube. The convection coefficient in a circular tube with constant wall temperature and laminar or turbulent flow is given by

Nu<sub>D</sub> = 
$$hD/k = 3.66$$
, for constant wall temperature and Re<sub>D</sub>  $\leq 2300$  (Equation 37)

$$\operatorname{Nu}_{D} = hD/k = \frac{(f_{t}/8)(\operatorname{Re}_{D} - 1000)\operatorname{Pr}}{1 + 12.7(f_{t}/8)^{1/2}(\operatorname{Pr}^{2/3} - 1)}, \ 0.5 \le \operatorname{Pr} \le 2000 \ \& \ 3000 \le \operatorname{Re}_{D} \le 5 \times 10^{6} \quad (Equation \ 38)$$

where  $f_t$  is the fanning friction factor of the tubes, Pr is the Prandtl number of the feed water, and Re<sub>D</sub> is the Reynolds number of the feed water.



**Figure 7.** (a) Circular tube–continuous fin heat exchanger, (b) Friction factor and Colburn j factor for a circular tube-continuous fin heat exchanger.

The convection coefficient for the exhaust gas was determined using empirical correlations provided by Kays and London. The correlation is shown in Figure 7 (b) and is given in terms of the Colburn j-factor. The Colburn-j factor is given by

$$j = \frac{h_{h2} (\Pr_{h2})^{2/3}}{Gc_{p,h2}}$$
(Equation 39)

where  $h_{h2}$  is the convection coefficient of the exhaust gas,  $Pr_{h2}$  is the Prandtl, and  $c_{p,h2}$  is the specific heat capacity. G is the maximum mass velocity and is given by

$$G = \dot{m}_{h2} / \left( \sigma A_{fr} \right)$$
 (Equation 40)

where  $\sigma$  is defined as the ratio of the minimum free flow area of the finned passages,  $A_{ff}$ , to the frontal area,  $A_{fr} = L_2 \times L_3$  [5].

The overall fin efficiency is given as  

$$\eta_o = 1 - (A_f / A)(1 - \eta_f)$$
 (Equation 41)

where  $A_f$  is the fin surface area, A is the total heat transfer area, and  $\eta_f$  is the efficiency of a single fin.

#### **Pressure Drop**

The pressure drop occurring across HX1 was derived using the geometry discussed earlier. The pressure drops occurring across HX2 were determined for specific geometry shown in Figure 8.

The pressure drop for the feed water and exhaust stream in the continuous fin heat exchanger are given by Eq. 42 and Eq. 43, respectively.

$$\Delta P_{HX2,c} = 2f_f \left(\frac{L_1}{D_i}\right) \frac{\left(\dot{m}_{F,i}/N_t\right)^2}{2\rho_F A_{ct}}$$
(Equation 42)

$$\Delta P_{HX2,h} = \frac{G^2 v_i}{2} \left[ \left( 1 + \sigma^2 \left( \frac{v_o}{v_i} - 1 \right) + f_f \frac{A}{A_{ff}} \frac{v_m}{v_i} \right]$$
(Equation 43)

where  $\Delta P_{HX2,c}$  is the pressure drop in the in the feed stream across HX2,  $\Delta P_{HX2,h}$  is the pressure drop in the exhaust stream across HX2,  $\rho_F$  is the density of the feed water,  $A_{c,t}$  is the inner cross sectional area of the tubes,  $f_f$  is the fin friction factor,  $v_i$  and  $v_o$  are the inlet and outlet specific volumes, and  $v_m$  is the mean specific volume.

The ratio  $A/A_{ff}$  can be calculated by

$$\frac{A}{A_{ff}} = \frac{\alpha V_{HX2}}{\sigma A_{fr}}$$
(Equation 44)

 $\alpha$  is the ratio of A to the total volume of HX2,  $V_{HX,2}$ .

The fin friction factor was derived using the data provided by Kays and London shown in Figure 9. In deriving the friction factor for the tubes, it was assumed that the tube surface was smooth. Depending on the flow regime of the feed water, the tube friction factor is given by Eq. 45 or 46.

$$f_t = 0.316 \text{ Re}_D^{-1/4}$$
, for ReD  $\le 2 \times 10^4$  (Equation 45)

$$f_t = 0.184 \text{ Re}_D^{-1/5}$$
, for ReD > 2×10<sup>4</sup>

**Figure 8.** Circular tube–continuous fin heat exchanger.  $\alpha$  587 m<sup>2</sup>/m

(Equation 46)



Figure 9. Friction factor and Colburn j factor for a circular tube-continuous fin heat exchanger.

#### Testing of the FO + MD Sub-System

Integration of the FO and MD processes centered on selection of an appropriate draw solution. The draw solution for the FO process was the feed solution for the MD process; the ideal draw solution was one that optimized flux in FO and was easily reconcentrated by MD. It was also one that did not reduce MD membrane hydrophobicity after exposure to high temperatures over longer time periods. After screening the draw solutions, bench-scale FO experiments were carried out to test the water flux produced by each draw solution and quantify the ability to reconcentrate the selected draw solutions.

# **Technical Approach**

As stated above, the main objective of this research was to design, fabricate, and test a pilot-scale OMBR-MD system consisting of a BIO, FO, and MD process and driven by on-site waste heat. To carry out the objectives, several tasks were performed, as outlined below.

# Task 1: Bioreactor Design, Instrumentation, and Control

The objective of Task 1 was to test the biological process coupled with the FO membrane. The bioreactor was equipped with electronic sensors connected to a control system for automated operation, process control, and data collection.

Research was focused on developing a process flow diagram for the membrane bioreactor system. This is an important aspect that was useful for optimizing operations and control of the system, particularly once the optimal probes for the MBR system were selected. Simultaneous attempts were made to develop a bench scale oxic-anoxic bioreactor using the aerobic reactor of an existing membrane bioreactor system. Data were collected for the system over the long-term to identify overall system performance and steady-state operation as well as during select oxic-anoxic cycles to identify changes in water quality during the different phases. An ongoing aspect of the research was to identify the optimal oxic and anoxic cycle length to obtain satisfactory nitrification (during oxic operation), denitrification (during anoxic operation), and organic carbon removal (during both oxic and anoxic reaction periods). The system was tested at a 3:1 oxic:anoxic reaction period during startup; however, this ratio was varied during subsequent research phases to obtain optimal contaminant removal.

# Task 2: Bench-Scale FO Membrane Fouling

The objective of Task 2 was to determine the underlying mechanisms of FO membrane fouling and reversibility as they apply to the common range of foulants typically encountered in traditional or FOB wastewater treatment systems. The underlying mechanisms of FO membrane fouling and reversibility were determined as they apply to the common range of foulants encountered in wastewater treatment systems. The information for the membrane cleaning procedures that are necessary for long-term FO membrane operation within the OMBR were provided as a results of this task.

#### Task 3: Bench-Scale MD Process Modeling

The objective of Task 3 was to develop a DCMD process model to mitigate wetting of the MD membrane.

Experiments were initially focused on determining the best membrane to use for studying the DCMD process, the optimal size of spacers to use with the membrane cell, and the optimal operating temperatures. Modeling efforts were focused on the development and validation of a steady-state heat and mass transfer DCMD model, implemented with the FORTRAN 90 programming language. The DCMD model is based on an iterative application of a steady state heat and mass balance over successive differential sections of a specified membrane. Film theory is used in the model to account for temperature and concentration polarization so that the temperatures and salinities present at the membrane surfaces can be determined. The model was first used to determine the membrane vapor permeability coefficient (MVPC) using experimental data from bench-scale experiments.

#### Task 4: Heat Exchanger Design and CFD Modeling for MD

The objective of Task 4 was to design a heat exchanger system to recover the waste heat from the engine coolant and exhaust of the diesel power generator that serves as the energy source for the MD system.

A computational fluid dynamics (CFD) model was employed to analyze the performance of the MD system for various operating conditions and to determine the energy and flow requirements of the MD system. The thermal and hydraulic characteristics of the MD system and its heat exchanger system were numerically analyzed using ANSYS Fluent, commercial CFD package. The operational conditions and dimensions based on the small flat-sheet MD system under testing were considered for the CFD analysis to calculate the energy and flow requirements of the MD system. For the CFD analysis, the continuity, momentum, and energy equations were solved using. A two-dimensional model was

used for the CFD analysis considering the symmetrical condition in the direction perpendicular to the water flow direction. It was assumed that the top and bottom walls of the MD system are insulated.

# Task 5: Lab-Scale BIO+FO Sub-System Testing

The objective of Task 5 was to test the overall performance of the OMBR biological process coupled with the FO membrane at laboratory scale. Performance evaluations focused on system parameters such as bioreactor and product water quality, membrane water production and fouling, and mixing and aeration requirements. The sub-system was equipped with electronic sensors connected to a control system.

# Task 6: Design of Lab-Scale MD + Heat Exchanger Sub-System

The objective of Task 6 was to design and construct the laboratory-scale MD sub-system coupled with heat exchanger (HX). The sub-system was equipped with electronic sensors connected to a control system.

# Task 7: Lab-Scale FO-BIO and MD-Heat Exchanger Sub-System Assembly

The objective of Task 7 was to integrate the FO+BIO and MD+HX sub-systems. This included physical connections and logic/control, as well as safety aspects of each sub-system. A water-resistant case to house the monitoring and control system electronics was installed. Long-term experiments were performed to evaluate the specific operating conditions required for biological nutrient removal and MD reconcentration of the draw solution. A comprehensive development/troubleshoot task was performed for the LabVIEW programs of each loop, sub-system, and the whole OMBR system. Additionally, an automated backwash system was developed for the system.

#### Task 8: Lab-Scale OMBR System Testing

The objective of Task 8 was to scale-up and apply process control from the bench-scale OMBR to the lab-scale OMBR. The automated laboratory-scale operation carried out was capable of optimizing treatment based on real-time reactor conditions. Lessons learned from the bench-scale system were proven valuable at the lab-scale system despite differences in hydrodynamics and FO module design.

#### Task 9: OMBR System Scale-Up Testing

The objective of Task 9 was to test the combined system under various stress or perturbation events, evaluate alternative membranes and membrane module configurations, and incorporate additional technologies and ideas arising from the results.

# Results

# Task 1: Bioreactor Design, Instrumentation, and Control

 $SubTask\ 1.1-Develop\ an\ oxic-anoxic\ bioreactor\ for\ treating\ FOB\ wastewater$ 

Work was performed to develop and construct a single cell, alternating oxic-anoxic reactor from the existing MBR reactor system. This bench-scale single cell reactor was constructed using the aerobic reactor that had a working volume of 6 L. In this system, influent was pumped directly into the reactor using a peristaltic pump at a constant rate of 3.47 mL/min and an effluent line connected to the membrane cassette was pumped at the same flow rate to achieve a hydraulic retention time of 1.2 days. Air was intermittently supplied to create alternating aerobic and anoxic conditions using a solenoid valve. Sludge inside the reactor was recycled using another peristaltic pump at a rate 3 to 4 times greater than that of the influent supply rate. A schematic diagram of the single cell reactor is shown in Figure 10. The influent was made in batches of 15-20 L (sufficient volume to last 3-4 days), stored in a refrigerator at 4°C to avoid biological growth, and pumped directly into reactor. During aerobic operation, air was pumped at a rate of 1 L/min.



Figure 10. Single cell alternating oxic-anoxic reactor

Several physical and water quality parameters were monitored during bioreactor operation, namely: temperature, pH, (DO), oxidation-reduction potential (ORP), ammonia-nitrogen (NH<sub>3</sub>-N), nitrate-nitrogen (NO<sub>3</sub>-N), orthophosphate (PO<sub>4</sub>-P), COD, and mixed liquor suspended solids (MLSS). Data were collected for concentrations of NH<sub>3</sub>-N, NO<sub>3</sub>-N, PO<sub>4</sub>-P, and COD over the period of June 15-June 29, 2013. In general, the reactor system was effective at removing ammonia from the influent feed stream with complete removal on all sampling days except for one. However, the system was not effective at removing the nitrate formed during the oxic period. Based on the data shown, the denitrification performance does appear to improve over time (effluent values decreasing from about 31 mg/L to about 24 mg/L). It was hypothesized that the alternating sequence timing used does not allow for sufficient anoxic periods to allow complete denitrification to occur. This hypothesis was later tested by changing the oxic:anoxic cycle period from 3:1 to 1.5:1. While phosphorus removal is not a primary objective, effluent orthophosphate levels were consistently below influent levels with a range of 15-50% removal. The primary mechanism for phosphorus removal

is believed to be incorporation in biological solids. COD removal was somewhat varied, with effluent levels consistently below 60 mg/L and an average removal of greater than 90%.

Data were also collected during individual oxic-anoxic cycles to identify reactor changes between individual cycles. In general, the concentrations of ammonia, nitrate, and phosphorus were relatively steady between cycles and did not vary more than 5% on average between cycles. DO values did exhibit a marked difference between oxic and anoxic cycles. During aerobic operation, the DO level increased to a maximum value of about 2.8 mg/L while the minimum DO during anoxic operation was about 1.7 mg/L. For optimum denitrification, the DO concentration should be less than 1 mg/L during anoxic periods and this may be a compounding factor in the poor level of denitrification observed. To alleviate this issue, the duration of the anoxic period was increased. However, to obtain complete denitrification there needs to be sufficient organic carbon to drive the nitrate reduction process. Based on the low COD values measured, the system may be organic carbon limited. Experiments were subsequently performed to identify whether the system was limited by low amounts of organic carbon by conducting spike additions of organic load should lead to a rapid decrease in DO during the anoxic period, and (2) the lower DO and high level of organic carbon should enhance nitrate reduction. If the system turns out to be carbon limited, then the cycle length can be further adjusted to enhance organic carbon degradation during anoxic conditions. Another possibility would be to increase the feed flow in the system, thereby increasing the overall loading rate.

# SubTask 1.2 – Develop an integrated sensor-based control system for the bioreactor

A controlled operating program for the biological process of the OMBR system was developed to specifically allow biological nutrient removal in a single reactor. The control and operating program for the biological process of the OMBR system relied on DO, ORP, pH, and cycling time between oxic and anoxic (DO < 1 mg/L) phases to achieve nitrification and denitrification, respectively. DO, ORP and pH probes will give an indication of the control and operation of the biological process of the system. The total nitrogen (TN) sensor will give a real-time indication of the biological performance of the OMBR system and the MLSS probe will measure the sludge concentration of the system, which is critical for the performance of the biologicator.

Two TN sensors were selected for consideration. One of the sensors is from HACH (NO<sub>3</sub>D <sub>SC</sub>) and the other is from WTW (IQ sensor Net, VARiON® *Plus* 700 IQ). The sensor from the HACH company can measure only nitrate while other measures both nitrate and ammonium. For MLSS measurement, three sensors had been selected for considerations. But out of these three, only Zulling Cosmos 25 probe is available in US. Yokogawa MLSS analyzer (SS 400) and DYS MLSS analyzer (DWA 3000B-MLSS) are not available in the US market.

LabView programming for data acquisition and monitoring and a minor reconfiguration of the bioreactor and its accompanying ports and probes was completed. The new configuration of the bioreactor had an 18-L operating volume that was maintained via an elevated feed reservoir comprised of chilled primary effluent. The system operated with an initial solids retention time (SRT) of 15 days. Additionally, the single stage anoxic/aerobic cycling was achieved by alternating a 15-minute aeration period (~1.5 LPM) with a 45-minute anoxic period.

Initial control trials were run to determine the clean water flux using DI water as the feed solution and a 0.6 M NaCl draw solution circulated at 1,000 mL/minute. The average clean water flux was 7.2 L/m<sup>2</sup>-hr. After characterizing the clean water flux of the submerged membrane module, focus shifted towards the inoculation of the bioreactor. The reactor seed was 2 L (~10% of the total reactor operating volume) of return activated sludge (RAS) collected from the South Truckee Meadows Water Reclamation Facility. The remainder of the operating volume, as well as the elevated feed tank, was filled with primary effluent collected from the Truckee Meadows Water Reclamation Facility. As expected, the actual operating flux decreased slightly compared to the clean water flux, and fluctuated between 6 and 7 LMH under the given operating conditions. The resulting water flux over 40 hours of operation is illustrated in Figure 11.



Figure 11. OMBR flux with clean water only (dotted) and with inoculated primary effluent (solid).

In addition to the flux data acquisition in this trial, continuous monitoring of the bioreactor conductivity, pH, and temperature was performed. The pH throughout the 40-hr trial remained between 7-7.5. The initial conductivity of the feed solution was recorded as 126  $\mu$ S and progressively increased to 4 mS as a result of reverse salt flux from the draw solution. A more detailed examination of the progression of the feed conductivity unveiled a resolution error with the conductivity controller and probe. This error resulted in a uniform step increase output in the conductivity reading.

A sampling procedure and a collective analysis plan of the primary effluent feed, the OMBR, as well as the draw solution from the FO operation was implemented. The OMBR performance was monitored in terms of the resulting product water from the FO operations, its performance in treating the primary effluent feed, as well as the influence of the FO operations on the functionality of the microbial community in the reactor.

Subsequently, continuous operation of the bench-scale FO membrane bioreactor and the analysis of the wastewater treatment and FO membrane performances within the reactor were investigated. The reactor configuration used is illustrated in Figure 12. The 18-L operating volume of the reactor was continuously maintained for 65 days following the initial reactor inoculation.



Figure 12. Bench-scale OMBR configuration.

The operational flux of the reactor was maintained between 2-4 L/m<sup>2</sup>-hr during experiments. Throughout the 65 days of operation, two membranes modules failed, causing a dramatic increase in the conductivity of the reactor mixed liquor. After the membranes were replaced, the flux was observed to gradually decrease from 6 L/m<sup>2</sup>-hr to the 2-4 L/m<sup>2</sup>-hr range, with the reactor conductivity stabilizing at a value of ~8 mS with daily wasting. Periodic backwashes were employed to mitigate membrane fouling and maintain a flux of 2-4 L/m<sup>2</sup>-hr.

A solids retention time (SRT) of 15 days was maintained through daily sludge wasting in the OMBR. An initial anoxic/aerobic cycling of 15 min aeration (~1.5 LPM of air) with a 45 min anoxic period was initially selected for the single stage reactor. The FO unit processes were driven by circulating a 0.6 M NaCl draw solution at 1,000 mL/minutes through the submerged membrane module. The DO and nutrient profile of the reactor during a 1-hr cycle with 15 min on/45 min off aeration (corresponding to Day 2 of reactor operation) is given in Figure 13.

Closer examination of the DO profile displays the reactor failing to reach anoxic conditions under the given aeration cycle (15 min on/45 min off aeration at 1.5 LPM airflow). This has perhaps caused the denitrification processes to be hindered as no noticeable decline in the  $NO_x$  species was observed during the anoxic periods. This prompted the aeration cycle to be adjusted to 10 min on/50 min off at 1.5 LPM. In addition to adjusting the aeration cycle, the reactor was dosed with an inoculant comprised of biomass samples collected from the nitrification- and denitrification-specific unit processes at the Truckee Meadows Water Reclamation Facility (TMWRF). This adjustment was made on Day 20 of the continuous operation which can be viewed in the compiled nutrient data (Figure 14). A single aeration cycle using the newly adjusted 10 min on/50 min off DO scheme (Figure 15) and the accompanying nutrient data correspond to Day 26 of reactor operation.



**Figure 13.** OMBR DO and nitrogen profile for 1-hr aeration cycle (15 min aerobic/45 min anoxic). (Average NH<sub>3</sub>-N Influent = 42.7 mg/L, Average NO<sub>3</sub><sup>-</sup>-N Influent = 1.22 mg/L, Influent DO = 0.26 mg/L)



Figure 1.3.

**Figure 14.** OMBR compiled NH<sub>3</sub>-N and NO<sub>x</sub>-N data for 65 days of operation. (the system was re-inoculated at day 20).



**Figure 15.** OMBR DO and nitrogen profile for 1-hr aeration cycle (10 min aerobic/50 min anoxic). (Average NH<sub>3</sub>-N Influent = 42.4 mg/L, Average NO<sub>3</sub><sup>-</sup>-N Influent = 0.07 mg/L, Influent DO = 0.26 mg/L)

The 65 days of operation following the initial inoculation had promising signs of nitrification as the influent ammonia concentration was reduced substantially in the reactor to nitrate/nitrite ( $NO_x$ ) species (Figure 14). As previously mentioned, the denitrification processes did not appear to be operating effectively, and in order to address this issue several parameters were examined. Initial investigation began with examining whether the absence of true anoxic conditions was hindering denitrification. DO suppression trials were performed to drive down the DO via N<sub>2</sub> gas sparging to see if any reduction in NO<sub>x</sub> concentrations occurred. These trials yielded no significant reduction in NO<sub>x</sub> concentrations during anoxic periods, therefore the focus was shifted to considering whether there was an adequate carbon food source present in the reactor feed for the denitrifying microbial community to utilize.

Another issue arising with the 65 days of reactor operation was the low and declining MLSS concentration observed since the initial inoculation. Figure 16 illustrates the low and declining MLSS concentration with time.

Maintaining the high MLSS concentration that is normally present in a membrane bioreactor was a challenge in the operation of the bench-scale OMBR. The hydraulic residence time (HRT) of the reactor can fluctuate depending on the flux through the FO membrane. As OMBR operation progresses, flux decline attributed to membrane fouling occurs, resulting in an increased HRT in the reactor. The desired SRT:HRT ratio (12:1), which strongly influences biomass concentration, can become difficult to achieve under these conditions. Increasing the SRT also runs the risk of increased salt accumulation in the reactor caused by reverse salt flux from the draw solution. The increased salt accumulation may then adversely affect the microbial activity as well as decrease the osmotic driving force between the reactor and the draw solution. Approaching this issue from an FO operational standpoint may also be limited, as any increase in the draw solution salinity would also lead to an increase in reverse salt flux through the membrane. Not having a full characterization of the substrate and microbial food source in the influent wastewater also made it difficult to determine which parameters were limiting biomass growth. Many of the potential parameters that can be adjusted from one unit process are pinned by the constraints of the other unit process.



Figure 16. OMBR MLSS concentrations over 65 days of operation.

In an effort to increase the MLSS concentration in the reactor, as well as improve reactor performance, several of the remaining operational parameters that would not be detrimental to either unit process were targeted. To help decrease the HRT, the volume of the reactor was decreased from 18 L to 11.2 L, and the surface area of the membrane was expanded from 0.028 m<sup>2</sup> to 0.043 m<sup>2</sup> by re-configuring the FO membrane cassette. Special care was taken to remove an edge of the cassette suspected to be the cause of the broken membranes during the previous operation. Additionally, the SRT was extended from 15 days to 30 days to improve the SRT:HRT ratio. The reactor was reinoculated and operation commenced again under these new operating conditions. The 10 min on/50 min off aeration cycle at 1.5 LPM airflow was maintained in the new reactor configuration. Nitrogen species during the initial 12 days of the newly configured reactor are provided in Figure 17. Initial results were promising as the ammonia and total organic carbon concentrations were observed to decrease with some conversion of the ammonia to nitrate.

The MLSS concentrations of the newly inoculated reactor are illustrated in Figure 18, where the biomass concentration was observed to stabilize between 620-630 mg/L. The initial operational flux and the bioreactor conductivity following the inoculation of the newly configured reactor are given in Figure 19.



Figure 17. OMBR compiled NH<sub>3</sub>-N and NO<sub>x</sub>-N data for new reactor configuration.



Figure 18. Initial 15 days of OMBR MLSS concentrations.



Figure 19. OMBR flux (blue) and reactor conductivity (red).

Subsequently, strategies to best optimize the biomass production and performance in the OMBR were examined. Also approaches to better characterize the fate of nitrogen species in the reactor and draw solution through a mass balance were attempted.

Maintaining adequate biomass concentrations and subsequently achieving observable denitrification has been one of the major challenges facing the coupled physicochemical-biological system. Attempts to address this issue have been focused around three main approaches: (1) the presence of appropriate microorganism species (2) reconfiguration of reactor to improve the SRT:HRT ratio, and (3) the addition of a supplemental carbon source for microorganism growth.

Following the first inoculation of the bench-scale OMBR, the MLSS concentrations declined rapidly (Figure 20, black) and little or no denitrification was observed (Figure 21, black). This initial performance led to questions of whether the appropriate microorganisms were present in the reactor to achieve both nitrification and denitrification, so the reactor was spiked with nitrifying and denitrifying specific biomass obtained from the local WWTP facility. Again, a rapid decline in MLSS (Figure 20, red) and minimal denitrification (Figure 21, red) was observed. It was hypothesized that the operating conditions of the OMBR needed adjustment, specifically the SRT:HRT ratio (optimal being 12:1), so the reactor was reconfigured to promote biomass growth (i.e., the volume reduced, membrane area expanded, and SRT increased). Once again, the MLSS levels fell quickly (Figure 21, green) and the denitrification performance was poor (Figure 21, green). At this point, the issue of whether an adequate carbon source was available for the microorganisms to utilize for the denitrification process was evaluated. To do this, the primary effluent feed into the reactor was spiked with 300 mg/L glucose. Following the addition of the glucose to the reactor feed solution, a noticeable increase in MLSS (Figure 20, blue) and decrease in the reactor NO<sub>x</sub> species was observed (Figure 21, blue). This was attributed to increased denitrification rates amongst the microorganisms as well as a lowered HRT as a result of a newly installed membrane. The gradual accumulation of the NOx species following the initial period of denitrification is thought to be the result of an increasing HRT due to biofouling lowering the operational flux. Because the bioreactor influent is controlled via a float valve, the decrease in membrane flux results in a decreased influent feed rate and thus, an increase in the HRT. The MLSS concentrations were able to be maintained following the glucose addition, although they were still well below the targeted concentrations for an OMBR system (~8,000 mg/L).



Figure 20. OMBR MLSS response to system modifications.



Figure 21. OMBR nutrient profiles. Triangles - NH<sub>3</sub>-N; Circles - NO<sub>x</sub>-N.

The passage of nitrogen species through the membrane was also examined. Successfully tracking and characterizing the behavior of both  $NO_x$  and  $NH_3$  species in the reactor is vital to closing the mass balance for nitrogen and for reporting the efficacy of the system's treatment performance.

Figure 22 illustrates an FO rejection trial conducted using an abiotic SEPA cell system. The purpose of this experiment was to identify how different nitrogen species transport through the FO membrane. Since the experiment was performed without any microbial inocula, data could be collected on the rejection of nitrogenous species by the FO membrane irrespective of biological activity. The feed solution was prepared using ammonium nitrate to render a 35 mg/L NH<sub>3</sub>-N concentration (pH = 7.3) to mimic NH<sub>3</sub> concentrations typical of the primary effluent used as the OMBR

feed. A 35 g/L NaCl draw solution was used having an initial conductivity of 43.5 mS. The draw solution was gradually diluted throughout the course of the 4-hour trial by the flux of water from the feed solution. The feed and draw solutions were circulated counter-current at 1,000 mL/minute via two gear-drive pumps. An HTI OsMem<sup>TM</sup>Cellulose Triacetate (CTA)-ES FO membrane having a membrane area of 139 cm<sup>2</sup> was utilized with an active side-feed side (AS-FS) orientation; this is the same membrane and orientation utilized in both the bench- and lab-scale OMBR systems. The feed tank was placed on a scale to periodically record the mass of the tank in order to calculate the flux through the membrane. The system was operated for a total of 4 hours, with 6 mL of sample being collected every 20 minutes. Following the 4-hour trial, the samples collected were analyzed with a Lachat Flow Injection Analysis System to quantify the NH<sub>3</sub>-N and NO<sub>x</sub>-N concentrations in the feed solution. A mass loss in both NH<sub>3</sub> and NO<sub>x</sub> species were normalized to the original mass of each in solution. A mass loss in both NH<sub>3</sub> and NO<sub>x</sub> species was observed (Figure 22), indicating some degree of nitrogen species passage through the FO membrane. Based on the theoretical NH<sub>3</sub> and NO<sub>x</sub> concentrations calculated assuming full membrane rejection, the percent recoveries for NH<sub>3</sub> and NO<sub>x</sub> were 91.0% and 76.3%, respectively.



Figure 22. Normalized NH<sub>3</sub>/NO<sub>x</sub> mass as a function of FO operation time.

As revealed in Figures 23 and 24, the addition of glucose to the feed resulted in stabilization of MLSS and rapid utilization of nitrate-nitrogen. To better understand the processes occurring during this experimental phase, a separate investigation of carbon source utilization in the OMBR was performed. In particular, the bioavailability of organic carbon in the reactor and the associated influence on the efficacy of denitrification and nitrification processes were investigated. Strategies to optimize the biomass production and performance in the OMBR continued, with a shifted emphasis on higher strength wastewater from synthetic and alternative sources to compliment the currently employed primary effluent feed.

As previously discussed, the addition of glucose to the primary effluent feed solution resulted in higher biomass levels in the system as well as enhanced denitrification in the OMBR. The extent of this enhancement was further studied in a series of subset batch experiments. For these experiments, wasted OMBR mixed liquor was used to inoculate four sets of batch reactors, one spiked with both 60 mg/L glucose and 30 mg/L nitrate and the other with 30 mg/L nitrate alone. This experimental design was developed to identify the role of readily biodegradable organic substrate (glucose) on denitrification compared to organic carbon in the OMBR mixed liquor. One set of the batch reactors was maintained under aerobic conditions by continuous-mixing on a shake table in open 125-mL Erlenmeyer flasks. The second set of batch reactors was held under anoxic conditions in sealed 70-mL vials that were initially bubbled with nitrogen gas to de-oxygenate the samples. Since the vials were sealed, in-situ ORP or DO data were not able to be measured in the anoxic batch reactors; however, since the microbial inoculum had already been exposed to denitrifying conditions under aerobic-anoxic cycles in the OMBR, it was assumed that providing oxygen-free conditions would support denitrification. Each set also included both an OMBR mixed liquor control sample (OMBR mixed liquor with no glucose or NO<sub>3</sub><sup>-</sup> spike) as well as a blank sample (DDW spiked with 30 mg/L NO<sub>3</sub><sup>-</sup> or both 60 mg/L glucose and 30 mg/L NO<sub>3</sub><sup>-</sup>). Six samples were collected over the course of 3 days, and samples were analyzed for nitrate, total nitrogen, and TOC. Conclusions were based on the experimental conditions (e.g., oxic or anoxic) and specific reactor preparation (e.g., OMBR mixed liquor only, addition of glucose, abiotic controls, or DDW with nitrate and/or glucose). Figure 23 (a) and (b) illustrates the rates of NO<sub>3</sub><sup>-</sup> removal from the system under aerobic conditions, while Figure 24 (a) and (b) displays NO<sub>3</sub><sup>-</sup> removal under anoxic conditions. As expected, denitrification was readily observed under anoxic conditions. Looking closer at Figure 24 (a and b), the rates of denitrification are quantified via linear regression applied to the data points demonstrating a decreasing trend before leveling off. Here the contrasted rates of NO<sub>3</sub><sup>-</sup> removal between reactor configurations with OMBR mixed liquor, OMBR mixed liquor spiked with 30 mg/L NO<sub>3</sub><sup>-</sup>, and 60 mg/L glucose validated the carbon limited status, as the denitrification rate increased from 0.102 to 0.214 to 0.522 mg-N/L-hr, respectively.



Figure 23. (a) Aerobic batch reactors (b) Aerobic wastewater control and spike blanks



Figure 24. (a) Anoxic batch reactors (b) Anoxic wastewater control and spike blanks

Direct quantification of carbon source exhaustion was hindered by TOC instrumental malfunctions. Despite the lack of TOC data, the fivefold increase in denitrification observed in the anoxic reactor spiked with both  $NO_3^-$  and glucose further confirms the influence additional carbon and nitrogen substrates have on increased reactor performance.

Incorporating the concept of increasing substrate availability to the microorganisms in the reactor, focus has shifted to utilizing a stronger wastewater source to increase both carbon and nitrogen loading in the reactor. A higher strength wastewater modeled after a previously used synthetic wastewater feed was explored, where primary effluent continued to serve as the base for the feed solution. This higher strength wastewater was appropriately adjusted to: 600 mg/L BOD, 1,100 mg/L TSS, 150 mg/L NH<sub>4</sub>-N, and 1,600 mg/L COD, which is representative of the wastewater characteristics expected at FOBs (per Dr. Judith Barry, Noblis). Table 4 summarizes the various components comprising the newly developed feed and their observed contribution towards the target loading parameters.

Constituent	mg/L	COD	TOC	TN	NH <sub>4</sub> -N	PO4 <sup>3-</sup>	SO4 <sup>2-</sup>	BOD*
Peptone	370.0	510.6	148	48.3	1.7	0.1		204
Glucose	450.0	569.3	165					228
Yeast	370.0	416.3	106	25.2	2.0			167
Ammonium Sulfate	208.0			39.6	45.0		151.5	
Ammonium Bicarbonate	368.0			67.8	75.0			
Potassium Phosphate	66.0					11.8		
Primary Effluent		120	45.2	38.4	36.5	13.3	50.0	48
Target TOTAL		1,600			150	25.0	202	600
Projected Total		1,616	464	219	160	25.2	201.5	646.4

Table 4. Synthetic and Primary Effluent Feed Characterization.

\*BOD: based on the 4th Edition Metcalf & Eddy Wastewater Treatment and Reuse ratios (0.4 mg BOD/mg COD)

Sodium bicarbonate was also initially added to provide buffering capacity for the solution; however, its addition was eliminated as a notable increase in the feed conductivity was observed. Regardless of the sodium bicarbonate omission, the pH of the feed solution was maintained at 7.4 with a conductivity of 1.8 mS. The newly combined synthetic and primary effluent feed incorporates all parameters of the target loading concentrations with the exception of the target TSS concentration. This will require additional consideration in terms of selecting an appropriate surrogate TSS source that will not be detrimental to membrane and reactor performance. The reactor response and performance following the new synthetic and primary effluent feed were monitored. Additionally, a simultaneous spike of fresh return activated sludge collected from the local wastewater treatment plant (South Truckee Meadows Water Reclamation Facility) was added to the reactor to help boost the tailing MLSS as the new synthetic and primary effluent feed was phased in. The OMBR response to a higher strength wastewater that increased both carbon and nitrogen loading to the reactor were then studied.

#### Implementation of Synthetic Feed and Biomass Maintenance

The stronger wastewater was introduced to the system on Day 191 of operation, indicated by the green dashed-line in Figure 25. To increase MLSS levels in the OMBR, a re-inoculation of RAS/WAS from a local wastewater treatment plant (South Truckee Meadows Water Reclamation Facility) was also performed on Day 191. Following this spike, MLSS values in the reactor increased to ~1,900 mg/L before equilibrating to near 750 mg/L (Figure 26). After increasing the feed strength, ammonia levels increased to ~130 mg/L NH<sub>3</sub>-N before dropping back to 15 mg/L NH<sub>3</sub>-N. During the same time, nitrate/nitrite levels did not increase or accumulate, suggesting that the system was effectively denitrifying.

Because of the high ammonia levels and low nitrate/nitrite levels observed around Day 200, it was decided to increase the amount of nitrifying bacteria in the OMBR. A mesh filter was set up at an effluent outfall point of the nitrification trickling filters at TMWRF for two weeks to establish an attached growth culture of nitrifying bacteria that could be transferred to the OMBR system. The biomass collected on the mesh was re-suspended in primary effluent, settled

and decanted, and then introduced into the OMBR (250 mL of settled biomass). This nitrifying bacteria spike occurred on Day 245 of operation, and is indicated by the purple dashed line in Figure 25. Following the spike, near complete removal of ammonia was observed in the OMBR for an extended period of time. Although a slight increase in ammonia concentration was observed after Day 255, removal was still reasonable as the incoming concentrations (~160 mg/L NH<sub>3</sub>-N) were significantly increased with the stronger wastewater. The level of ammonia removal coupled with the consistently low nitrate/nitrite concentrations in the OMBR is indicative of both nitrification and denitrification occurring.

The MLSS concentrations following the initial re-inoculation (Figure 26, green dashed line) leveled off near 750 mg/L, and remained fairly consistent after the nitrifying bacteria spike (Figure 26, purple dashed line). Subsequent MLSS concentrations tailed slightly to levels around 560 mg/L. Despite these low levels of suspended solids, the membrane unit had exhibited attached growth, so the actual TSS levels in the reactor were not well known. The implications of the increased membrane fouling due to higher MLSS and resulting operational flux decrease are discussed in the following section.

# Membrane Condition and Operation Flux

After the implementation of the higher strength wastewater, membrane fouling increased dramatically (Figure 27). The increase in fouling led to the operational flux decreasing from  $\sim 2.2$  to  $\sim 0.4$  L/m<sup>2</sup>-hr, corresponding to an increase in the system HRT from  $\sim 4$  to  $\sim 14$  days. The SRT of the system is maintained at 30 days with solids wasting, which means the SRT:HRT ratio dropped to  $\sim 2:1$ . This is undesirable in terms of maintaining sufficient biomass concentrations, as observed in the decreasing MLSS levels (Figure 26). A more aggressive membrane cleaning and maintenance regime was evaluated to return the operational flux, improve the SRT:HRT ratio, and ultimately promote higher biomass concentrations.

# Nitrifying Bacteria Subcultures

In addition to the nitrifying biomass used to spike the OMBR, a portion of the biomass collected with the mesh catchment system was grown in subset batch cultures. These cultures were concentrated and then stored in the freezer to provide readily available nitrifying bacteria culture stocks for further OMBR spikes.



**Figure 25.** a) OMBR NOx-N and NH<sub>3</sub>-N Profile. b) TOC and TN Profile. (Red - glucose + primary effluent, Green - synthetic feed + primary effluent, Purple - nitrifying bacteria spike)



**Figure 26.** OMBR MLSS Profile. (Red - glucose + primary effluent, Green - synthetic feed + primary effluent, Purple - nitrifying bacteria spike)



Figure 27. a) Clean FO membrane before use. b) Observed FO membrane fouling intensified by higher strength wastewater.

# Task 2: Bench-Scale FO Membrane Fouling

Research focused on continued experimental design and preliminary fouling/cleaning with the bench-scale FO system. Early fouling tests of the bench-scale FO system focused on abiotic fouling of the FO membrane. In lieu of model foulants discussed earlier, the synthetic nutrient broth used in the bench-scale OMBR of Task 1 was used initially. No fouling was observed using the synthetic broth, as determined by a constant flux, even with extended run times of 20 hr (Figure 28). Based on these observations it was hypothesized that the biologically-mediated fouling would be the most problematic.

To evaluate the biologic fouling and the efficacy of osmotic backwashing, 3 L of culture were removed from the aerobic reactor of the OMBR from Task 1 and mixed with 2.2 L of fresh synthetic nutrient broth. The culture was transferred to the feed tank of the bench-scale FO system and gently bubbled with air throughout the fouling/cleaning experiment to mimic the aerobic chamber of the OMBR. The culture was circulated at 1.5 L/min through the FO membrane cell with a draw solution (35 g/L NaCl, 20°C, 1.5 L/min) circulated counter-current to the feed until ~50% flux decline was reached (~20 hr). After fouling, osmotic backwashing was performed by replacing the draw solution with DI water and circulating for 30 minutes. After backwashing, the draw solution was returned and the system operated again to determine the level of flux recovery achieved from the cleaning. The osmotic backwashing restored flux to 80% of the original (Figure 29), confirming that the biologically-mediated fouling can be at least partially reversed with osmotic backwashing.



Figure 28. Water flux in the bench-scale FO system to evaluate abiotic membrane fouling. Synthetic nutrient broth used for the bench-scale OMBR system was used as the feed solution.



**Figure 29.** Water flux in the bench-scale FO system to evaluate biologically-mediated fouling and the efficay of osmotic backwashing. Culture from the aerobic chamber of the OMBR mixed with synthetic nutrient broth as the feed solution.

Based on the positive osmotic backwashing results with culture from the bench-scale OMBR, it was decided that actual domestic sewage would be tested. A sewage sample was obtained from the Truckee Meadows Water Reclamation Facility (TMWRD); the sample was taken from the inlet to the primary settling tank as the head works station was under repair at the time of sampling. The sewage was mixed with synthetic nutrient broth and aerated in the feed tank of the bench-scale FO system and the system was operated as described above. However, after ~44 hr of operation, minimal flux decline was observed, indicating negligible fouling (data not shown). An additional sample was obtained from a second sampling date, but again no fouling was observed after >36 hr of operation, so it was decided that an alternative sewage source should be used. Testing is underway using diluted waste activated sludge from the aeration ditch of the South Truckee Meadows Water Reclamation Facility (STMWRF), which was the inoculum source used to seed the bench-scale OMBR. Modifications to the bench-scale FO system have been required, including the isolation of the float valve, because of the elevated foaming with the concentrated waste source.

In addition to osmotic backwashing, the effects of air scouring as a cleaning mechanism for the FO membrane will be evaluated. To test scouring effects with the bench-scale FO system, a new feed-side half-cell was designed that allows for aeration of the membrane surface (Figure 30). Air is introduced into the feed-side channel through a separate manifold at the bottom of the cell. Both air and feed solution travel vertically up through the cell, with the mesh spacer material preventing the air bubbles from coalescing into a single, large air slug. Preliminary testing with DI water as a feed confirmed that the aeration of the feed side only slightly decreases membrane flux due to air displacing water at the membrane surface (Figure 31). It is also noted that the additional bubblng in the feed tank from the membrane aeration does result in a noisier flux, however the encased float valve installed in the system for the new sewage source should minimize or alleviate that noise.



Figure 30. Modified feed-side half cell for the bench-scale FO system to allow for air scrubbing of the membrane surface.



Figure 31. Water flux in the bench-scale FO system with (blue) and without (red) aeration of the membrane surface on the feed-side. Distilled water was used as a feed source.

Sodium alginate was later used to simulate organic macromolecules that are generally associated with fouling. The membrane was stabilized and a clean water flux was determined during a 1-hr period where only DDW was used on the feed side and circulated counter currently to the draw solution, which was ~2.5 M NaCl. Both feed and draw were at 20°C and circulated at a flow rate of 0.45 L/min. The exact strength of the draw solution was adjusted for each run (to account for slight variation in membrane samples) to achieve a flux of ~11 L/m<sup>2</sup>-hr. After the stabilization period, the feed solution, 200 mg/L sodium alginate, 50 mM NaCl, and 0.5 mM CaCl<sub>2</sub> in DDW, was circulated countercurrent (0.45 L/min, 8.5 cm/s cross flow velocity) to the draw solution. The system was run continuously until the flux was stable for 3 hours, at which point osmotic backwashing was induced by replacing the draw solution with DDW for 30 min. Following the backwashing period, the draw solution was returned and the system was operated again at the

starting conditions to evaluate flux recovery and the time it took for the membrane to return to fouled state. The procedure continued through three osmotic backwashes per experiment. In one instance, a 30-minute experimental air scouring trial (with air flowrate of 3.5 L/min) was implemented to gain insight into air scouring experiments. This air scouring experiment can be seen in relation to the osmotic backwashes in Figure 32.



Figure 32. Progression of water flux in the bench-scale FO system as membrane fouling with sodium alginate occurs, as well as the flux recovery achieved via three osmotic backwashes and one air scour trial.

Based on the results from the osmotic backwashing, fouling reversal and flux recovery is achievable. The flux recovery for each osmotic backwash is promising, as the flux returns to 6-7 after each osmotic backwashing event. The effectiveness of the washing for an extended period of time may be an issue, as the time required to reach fouled state after each osmotic backwash decreased over the experiment. The conditions of the osmotic backwash (e.g., flowrate and intensity) may be adjusted to see how the effectiveness of the osmotic backwash may be improved.

In addition to the bench-scale studies, a preliminary Sensor Control Plan has been developed for the FO subsystem and the entire OMBR system (Appendix A) in collaboration with Dr. Sachi Dash and the rest of the project team.

Later, the implementation of air scouring as an approach to fouling mitigation and reversal was investigated. A modified acrylic SEPA cell was used for these experimental procedures. Custom made to include air ports in the bottom feed channel, this cell makes the incorporation of an air scour into the module convenient and ensures uniform distribution of the air through the module. Several air scour experimental procedures were investigated to determine the most effective method to maintain operational flux and to reverse/prevent fouling. The eventual fate of this research will be to incorporate the FO membrane processes in a submerged osmotic bioreactor, where the active layer of the membrane will be in constant contact with wastewater. Concern for the constituents of the wastewater fouling the model foulant to mimic the predicted fouling behavior of wastewater. Keeping the ultimate application of this research in mind, the air scouring procedures were designed to incorporate the operational conditions of the bioreactor as well as other parameters that could be transposed to the next stages of the project with ease. These experimental air scouring trials examined two primary methodologies for membrane cleaning, fouling prevention and fouling reversal, to evaluate whether a continuous scouring cycle aiming to prevent fouling (fouling prevention) or an air scour following 20 hours of operation (fouling reversal) is more effective for maintaining an adequate operational flux.

#### Air Scour Cycling

In efforts to evaluate fouling prevention through continuous air scouring, an experimental procedure was formulated to include a 15-minute air scour every hour. An electronic timer switch was used to trigger an air vacuum pump to turn on every hour for 15 minutes (1.5 LPM) to provide the air supply for the scour. The FO system was continuously

operated with the feed (sodium alginate solution) and draw (NaCl) solutions circulating counter current at 450 CCM, and no operating conditions were altered during the air scouring period. Figure 33 illustrates the flux in the air scour cycling trial over the 50-hour operational time period with a 10-hour DDW control that was operated with the same air scour cycling conditions.



Figure 33. Sodium alginate air scour cycling trial (15 min on at 1.5 LPM, 45 min off).

Figure 34 provides a 10-hour section of the sodium alginate air scour cycling trial. This depicts the flux depression that occurs during the 15-minute air scouring period. This observed decrease can be attributed to the decrease in water-to-membrane contact due to the air bubbles from the scour displacing a volume of the water being circulated in the module. This results in a decrease in the volume of water in contact with the membrane, subsequently decreasing the flux.



Figure 34. 10-hour dection of the sodium alginate air scour cycling trial.

From the resulting data, the implementation of this air scour cycling was observed to effectively maintain the flux above 8.5  $L/m^2$ -hr over the 50-hour fouling period. The trends observed potentially implicate that the aeration conditions and cycling in a bioreactor may adequately act as a membrane cleaning mechanism while the system is continuously operated, reducing and possibly eliminating the need to take the unit offline for frequent cleaning.

#### 20-hour Fouling Period & Air Scour

In addition to evaluating air scour cycling as a method for fouling prevention, experimental procedures similar to those applied with osmotic backwashing were also evaluated to analyze the effectiveness of air scouring as a fouling reversal mechanism. In these trials the feed (sodium alginate solution) and draw (NaCl) solutions were circulated (450 CCM) countercurrent for a 20-hour fouling period. Keeping the system under normal operating conditions, a 15-minute air scour (1.5 LPM) was introduced into the module following the 20-hour fouling period, after which the air supply was

stopped and the system returned to normal operating conditions. The system was continuously operated for 80 hours, and a total of three 15-minute air scours were performed. Figure 35 demonstrates the fouling behavior during the 20-hour fouling period, as well as the flux response and recovery following the 15-minute air scour.



Figure 35. 15-min air scour (1.5 LPM) following 20 hours of operation.

Applying the 15-minute air scour every 20 hours for fouling reversal was observed to maintain an operational flux around 6  $L/m^2$ -hr in the system. The overall performance and flux maintenance of this cleaning procedure is out performed by the air scour cycling trials that were observed to maintain a flux of 8.5  $L/m^2$ -hr. Hydraulic conditions and module constraints are both considerable factors separating this experimental procedure from the conditions of an osmotic membrane bioreactor; however, these trials yield promising results in terms of utilizing aeration already present in the bioreactor to mitigate and reverse membrane fouling.

#### Task 3: Bench-Scale MD Process Modeling

Experiments were initially focused on determining the best membrane to use for studying the DCMD process. Three membranes were tested for performance with and without the use of spacers. The three membranes tested were the PTFE45, QM022, and QL211100 membranes. The PTFE45 membrane is a composite membrane with a Polytetrafluoroethylene (PTFE) active layer and polypropylene (PP) support layer. The QM022 membrane is a single layer membrane with a PTFE active layer and no support layer. The QL211100 membranes allowed reverse salt flux to occur in the experiments without the spacers, while the PTFE45 membrane did not allow reverse salt flux. The PTFE45 membrane was subsequently used in all experiments until the supply of that membrane ran out.

Experiments were performed using the PTFE45 membrane to determine the optimal size of spacers to use with the membrane cell. The membrane cell is made from two acrylic plates, fastened together over the membrane by ten stainless steel bolts. Holes have been drilled into the sides of the acrylic plates so that the feed and distillate solutions can be pumped into each of the cell's two channels on either side of the membrane. A water tight seal is obtained within the cell by the use of rubber O-rings, which also prevent the membrane sheet from moving inside the cell. The size of each channel in the membrane cell was measured, and they were both determined to be 0.79 mm. The active membrane area within the cell is  $0.0130 \text{ m}^2$ . Six different sized spacers were tested, four of which were larger than the membrane channels and two of which were smaller than the membrane channels. Additional rubber gaskets were placed in the membrane cell for the experiments using spacers that were bigger than the channels. Experiments were performed using each set of spacers with a 35 g/L feed solution, distilled water as the distillate solution, and

temperatures of 20 and 40°C for the distillate and feed streams, respectively. The results of the experiments indicated that the membrane performed best with the 0.68 mm spacers.

The performance of the DCMD system was also evaluated for different operating temperatures. The operating temperatures were determined by choosing three sets of temperatures that correspond to the same vapor pressure difference across the membrane. The three temperature sets were chosen to be 20.0 and 40.0°C, 46.9 and 58.0°C, and 65.6 and 70°C for the distillate and feed streams, respectively. Each of the temperature pairs corresponds to a vapor pressure difference across the membrane of 4.9 kPa, when using a feed solution with a salinity of 35 g/L and distilled water for the distillate stream. Experiments were performed with the three temperature pairs at flow rates of 0.5, 1.0, and 1.5 LPM for both the feed and distillate streams (Figure 36). The results showed that water flux increased with the higher temperature pairs at 0.5 LPM, and that the 46.9 and 58.0°C temperature pair showed the highest flux for the 1.0 and 1.5 LPM flow rates. The results also showed that the flux increased with flow rate for the 20.0 and 40.0°C, and 40.0°C, and 46.9 and 58.0°C temperature pairs, but decreased with flow rate for the 65.6 and 70.0 temperature pair. The results indicate diminishing increases in water flux for the two higher temperature pairs with increasing flow rate and consistent increases in water flux with flow rate for the 20.0 and 40.0°C temperature pairs.



Figure 36. Water flux at of 0.5, 1.0, and 1.5 LPM with a vapor pressure difference across the membrane maintained at 4.9 kPa using three different operating temperature pairs.

Methods to reduce the pressure within the system have been investigated to ensure membrane hydrophobicity. High pressures are a concern because if they exceed the membrane's LEP then membrane wetting can occur, causing reverse salt flux: pressures as high as 30 psi have been observed with the QM022 membrane. While the 30 psi pressure is not close to the QM022 membrane's LEP of 56 psi, system pressures should be minimized to maintain a sufficient margin of safety. Tests were performed on the system to determine the cause of the high pressure. Removing a piece of Teflon tape that had partially covered one of the inlets to the membrane cell reduced the pressure to 20 psi on the distillate side, indicating that there was still a problem. After testing the system with one component removed at a time, the source of the problem was determined to be within the membrane cell. A PTFE45 membrane was used in place of the QM022 membrane and the pressure was reduced to an acceptable 5 psi. The results indicate that very high pressures are unavoidable when using the QM022 membrane.

The DCMD process model has been modified to be more versatile and applicable to a larger range of operating conditions. The previous version of the model was implemented in Microsoft Excel but is currently being written with the Fortran programming language to improve its versatility. Writing the program in Fortran allows for much more control over the questions that the model can be used to answer and makes it much easier to update or change the model equations. Additional equations are being included in the new model that allow it to be applicable to a larger range of operating conditions. The Excel model was limited to only turbulent flow conditions, parallel flow mode, and one-dimensional temperature distribution. The new model is based on two-dimensional temperature distribution, which allows for more accurate model predictions. The new model will be able to evaluate the DCMD process in laminar and turbulent flow conditions, and parallel and counter-flow modes. Programming the model in
Fortran allows for implementation of a dynamic model that changes depending on the operating conditions. The use of Fortran also allows the model to have access to a large range of numerical method subroutines that may be required to solve the model under certain operating conditions. Step-by-step modeling and validation procedures are available in the paper by Achilli and co-workers [4].

Once a new supply of membranes was obtained, tests were performed to determine the membrane distillation coefficient (MDC) for the DCMD process model. The MDC represents the portion of the vapor pressure difference driving force that results in water vapor flux across the membrane for a given membrane at some operating temperature. The MDC can be determined experimentally by calculating the flux and the vapor pressure difference from experimental results, and then solving for the MDC. The MDC value will be determined experimentally because the theoretical equations currently used to determine the MDC value do not provide accurate results. Tests have been performed with distilled water being used for the feed and distillate solutions with the 20 and 40°C temperature couple so that the MDC can be determined for the new membrane. The same data were collected for additional temperature pairs so to test a range of MDC values.

Subsequently, efforts were focused on the development and validation of a steady state heat and mass transfer DCMD model, implemented with the FORTRAN 90 programming language. The DCMD model is based on an iterative application of a steady state heat and mass balance over successive differential sections of a specified membrane. Film theory is used in the model to account for temperature and CP so that the temperatures and salinities present at the membrane surfaces can be determined. The temperatures and salinities at the membrane surfaces are different than those of the bulk fluids and must be known to determine the driving force that is available to support the DCMD process. The model can evaluate heat and mass transfer in co-current or counter-current flow modes.

First, the model was used to determine the membrane vapor permeability coefficient (MVPC). The MVPC dictates the fraction of the vapor pressure driving force that translates into water flux (Eq. 47). Many researchers have modeled the MVPC as a complex combination of diffusion models. However, the diffusion models require extensive data on membrane properties for each membrane that is to be modeled. Determining all of the membrane properties requires many time consuming experiments to be performed, and the process of determining the relative contribution of each type of diffusion is very time consuming. For the proposed model, a simplified process for determining the MVPC has been developed that employs only a small number of water flux experiments for each membrane to be modeled.

$$J_w = MVPC(P_{v,f,m} - P_{v,d,m})$$

# (Equation 47)

where  $J_w$  is the transmembrane water flux, *MVPC* is the Membrane Vapor Permeability Coefficient,  $P_{v,f,m}$  is the vapor pressure at the feed side membrane surface, and  $P_{v,d,m}$  is the vapor pressure at the distillate side membrane surface.

The MVPC was determined by using experimental data from bench-scale experiments. Data from experiments using distilled water as the feed solution were used to remove the effect of CP on the relationship between water flux and transmembrane vapor pressure difference. The effect of temperature polarization on the relationship between water flux and transmembrane vapor pressure difference was removed by using the model to find the temperatures at the membrane surface. The average temperature between the inlet and outlet was used in calculations to determine the MVPC because the temperature distribution across the membrane was unknown. Using the average temperature provides a good approximation of the MVPC because the bench scale experiments use a small membrane area. With polarization effects removed, a constant and more accurate relationship between water flux and transmembrane vapor pressure difference is obtained. For example, for membrane QM022, an average MVPC value was obtained from multiple experiments, resulting in an average MVPC value of  $7.37 \times 10^{-7}$  s/m, with a standard deviation of  $2.7 \times 10^{-8}$  s/m.

The model was tested using experimental data operated in multiple flow modes, with different temperature couples, and feed solution salinities. All data was taken using a single-layer polytetrafluoroethylene (PTFE) membrane labeled QM022 in a membrane cell with an effective membrane area of 138.7 cm<sup>2</sup>. Feed outlet temperature and water flux predictions for counter-current experiments are given in Figures 37 and 38, respectively. Water flux predictions for co-current experiments with high salinity feed solutions are given in Figure 39. All low-salinity co-current experimental results were used in obtaining the MVPC so low-salinity co-current data was not used to validate the model.



Figure 37. Experimental and predicted DCMD feed outlet temperature for counter-current experiments.



Figure 38. Experimental and predicted DCMD water flux for counter-current experiments.

Errors are due to the distillate flow rate estimation technique, the presence of high pressures in the membrane cell, the range of validity of water property regressions, and because heat losses to the surroundings were not considered. Because the first step of calculations in counter-current flow mode begins at the outlet of the distillate stream, the flux across all other steps must be approximated to account for their contribution to the flow at the outlet. The approximation technique underestimates the flow rate and introduces error into the prediction. The error in the counter-current experiments can also be attributed to the high pressures in the membrane cell, between 10 and 15 psi above atmospheric pressure. All water properties are calculated as functions of temperature and salinity at atmospheric pressure, so the higher pressures introduce error into the predictions. The high salinity co-current experiments show increasing error for increasing salinity because the range of validity for most of the water properties is between 150 and 180 ppt. The results of the lowest of the high salinity experiments show the best agreement with experimental data is reduced at the higher salinities because they are outside or nearing the upper boundary of validity of some of the water properties.



Figure 39. Experimental and predicted DCMD water flux for co-current experiments with high salinity.

Once the DCMD model was completed and tested with experimental data, it was expanded to include an analysis of heat and mass transfer in the bioreactor, FO draw loop, and DCMD feed loop. The models of the bioreactor and FO and DCMD loops are based on iterative application of a steady state energy balance. The FO module was modeled as a counter-current heat exchanger in addition to a pre-heating heat exchanger (HX1), located just before the waste heat heat exchangers. The bioreactor, FO draw loop, DCMD feed loop, and HX1 sections of the model are all iterated until steady temperatures are obtained.

The results of the model indicate that temperature control is needed on the DCMD distillate loop. All model results indicated that the bioreactor temperature would become almost equal to the DCMD feed outlet temperature and that the DCMD distillate outlet temperature would become almost equal to the DCMD feed inlet temperature. The model was also tested by moving HX1 to the wastewater inlet pipe and increasing the membrane length, but the results were the same. The DCMD distillate temperature must be controlled to maintain desirable system performance.

Later, the DCMD model was used to determine the optimal operating conditions and configuration for the DCMD process by evaluating the system's energy consumption for those conditions. The goal of the analysis was to design an MD system that produces 6 L/hr of product water while minimizing energy consumption. The 6 L/hr production was selected to match the production of the FO system in the OMBR. The temperatures were selected to range between 30 and 70°C, based on the limitations of the equipment and expected incoming wastewater temperature. Specific energy consumption was calculated as the sum of the pump energy and the energy required to change the temperature of each stream from the outlet temperature to the inlet temperature set-point. The pump (electric) energy contributes minimally to the energy consumption in comparison to the energy (heat) required to change the temperature of each stream. All model calculations were done using the QM022 membrane characteristics with feed and distillate stream salinities of 35 and 0 ppt, respectively. In addition to system modeling, the bench scale DCMD heating system has been improved and more data is being gathered to further validate the model.

In-parallel configurations were analyzed first, using a fixed membrane size of 12 in wide by 12 in long and a channel velocity of 0.25 m/s. A three-inch margin was removed from all membrane sizes considered to compute the effective membrane area actually available within the module. The 0.25 m/s velocity was selected as a starting point because it is a commonly used testing condition for DCMD bench scale experiments. The data collected was the distillate inlet temperature required to provide 6 L/hr production when using a 70°C feed inlet temperature, the recovery, specific energy consumption, and power consumption (Table 5). The maximum feed inlet temperature of 70°C was selected because the driving force is greater for the same temperature difference at higher temperatures. The results showed low recoveries (between 0.39 and 0.78%) and high specific energy consumption (average of 1875 kWh/m<sup>3</sup>). The results indicated that to increase the recovery, the number of modules in-parallel must be reduced while still meeting the desired flux. Additionally, when observing the very small change in temperature from inlet to outlet over the short 12 in length, it suggested that the input energy was not being used to its full potential, resulting in low performance. These results indicated that a longer membrane should be used to fully utilize the input energy in each of the streams.

Modules In-Parallel	Feed Inlet Temperature (°C)	Distillate Inlet Temperature (°C)	Feed Outlet Temperature (°C)	Distillate Outlet Temperature (°C)	Water Production (L/hr)	Recovery (%)	Specific Energy (kWh/m³)	Power (kW)
5	70.0	36.4	63.7	42.5	6.0	0.78	1,920	11.6
6	70.0	44.4	65.0	49.3	6.0	0.65	1,873	11.3
7	70.0	49.1	65.8	53.2	6.0	0.56	1,860	11.2
8	70.0	52.2	66.3	55.8	6.0	0.49	1,860	11.2
9	70.0	54.5	66.8	57.6	6.0	0.43	1,868	11.3
10	70.0	56.2	67.1	59.0	6.0	0.39	1,885	11.4

**Table 5.** Performance with modules in-parallel using a 12 in by 12 in membrane with a 0.25 m/s channel velocityand a feed salinity of 35 ppt.

The performance of single membrane module configurations was evaluated by recording performance at various lengths for different channel velocities. Evaluations began by using the maximum feed inlet temperature of 70°C and the minimum distillate inlet temperature of 30°C, providing the maximum allowable temperature difference and therefore the maximum possible production from a given membrane area. System performance was evaluated for channel velocities between 0.1 and 0.5 m/s at lengths between 1 and 12 ft. Using a channel velocity below 0.27 m/s, the desired water production cannot be achieved in a single module because sufficient temperature differences between the feed and distillate cannot be maintained. The performance at 0.5 m/s (Figure 40) represents the conditions where the desired water production is achieved in one module at the point of minimum energy consumption. However, because the channel velocity is high, the recovery is somewhat low at 1.9% with a power consumption of 11.6 kW. The same production can be achieved with a higher recovery of 3.4% and an only slightly higher power consumption of 12.1 kW when using a 12 ft long membrane and a velocity of 0.27 m/s (Figure 41).



**Figure 40.** Performance using a channel velocity of 0.50 m/s, a feed inlet temperature of 70°C, and a distillate inlet temperature of 30°C.



**Figure 41.** Performance using a channel velocity of 0.27 m/s, a feed inlet temperature of 70 °C, and a distillate inlet temperature of 30°C.

Another important feature of the performance with a 0.27 m/s velocity is that the distillate outlet temperature becomes larger than the feed outlet temperature for a shorter membrane length than with the 0.5 m/s velocity. The temperature difference can be used to recover heat from the distillate stream to pre-heat the feed stream using a heat exchanger, reducing the power consumption. The use of a heat exchanger with a 0.27 m/s velocity and a 12 ft long membrane would result in an estimated power reduction of 3.8 kW, bringing the total power consumption to 8.2 kW or a specific energy of 1,366 kWh/m<sup>3</sup>. The power reduction is currently only an estimate but the heat exchanger will be added to the model to verify the estimate.

A concern when using long membranes is the large amount of head loss that occurs, creating large pressure differences across the membrane in excess of the membrane's liquid entry pressure (LEP) could compromise the hydrophobicity of the membrane, causing salt flux into the distillate. However, utilizing the QM022 membrane, the LEP is 55 psi and the maximum transmembrane pressure difference with a 12 ft long membrane and a velocity of 0.27 m/s is 10 psi. Even with a 12 ft long membrane and a 0.50 m/s velocity the maximum transmembrane pressure difference is 20 psi. Pressure losses developed across a long membrane module may be a concern for the membrane module materials or the piping being used. A 12 ft membrane may seem to be prohibitively large but the membrane module could be partitioned into multiple modules in series that are stacked together to conserve space. Based on the current findings, the optimal operating conditions are to use a feed inlet temperature of 70°C, a distillate inlet temperature of 30°C, a velocity of 0.27 m/s, and a single module with a 12 ft long membrane (Table 6).

Table 6.	Characteristics	of the currentl	y optimal	operating	conditions and	configuration.
						0

Channel		Feed Inlet	Distillate Inlet	Feed Outlet	Distillate Outlet	Water		Specific		Maximum
Velocity	Membrane	Temperature	Temperature	Temperature	Temperature	Production	Recovery	Energy	Power	Pressure
(m/s)	Length (ft)	(°C)	(°C)	(°C)	(°C)	(L/hr)	(%)	(kWh/m³)	(kW)	(psi)
0.27	12	70.0	30.0	39.5	58.7	6.0	3.44	1,366	8.2	10.1

DCMD model simulations were performed based on the bench scale data to determine the optimal operating conditions for the pilot scale DCMD system. A design for the membrane modules in-series configuration has been developed that minimizes materials use and space consumption by stacking the membrane modules on top of each other as in Figure 42.

A stacked membrane module system with six 2-foot by 1-foot membranes is currently being designed to simulate the twelve square feet of membrane needed. We selected this orientation as it allows for a compact system footprint, reduces materials use, and eases membrane installation compared with a single long membrane cell. The DCMD cells were being constructed from impact resistant polyethylene, which offers high machinability while maintaining low

cost and weight. The stacked cells will be pressed together using two sheets of the same material, one on each side of the stack. They will be pressed together using 3/8 inch stainless steel bolts, which will be fixed along the width and removable along the length of the DCMD cells. By allowing for the removability of the bolts, cells can be removed and maintained from the apparatus. Each inner cell will be machined to house two membranes, one on each side. The top and bottom cells within the press will have one membrane channel. The plumbing configuration will allow for operation in concurrent and countercurrent mode. As the pilot scale DCMD system is being fabricated, we are also designing a bench-top which will house the system. The bench-top will be designed to make the system mobile and easily operated. Additionally, the configuration of the system will be done so with considerations of coupling with other components of the water treatment system (FO).



Figure 42. Cross-section of the proposed in-series membrane module layout based on a stacked module design.

To demonstrate the effectiveness of the in-series heat recovery exchanger design, a comparison has been made between in-parallel and in-series configurations for the 6 L/h system. All model investigations use modules that are 1 ft by 2 ft, with a margin of 1.5 in removed from the perimeter of the membranes in each module to account for the securing o-rings used in the flat-sheet module design. The least energy intensive operating conditions for the 6 L/h system for in-series and in-parallel configurations when not using a heat recovery exchanger are given in Table 7. The results show that the in-parallel configuration is more energy efficient than the in-series configuration when not using a heat recovery exchanger. However, Table 8 shows the most energy efficient operating conditions when using a heat recovery exchanger and how the use of a heat recovery exchanger results in the lowest power consumption for heating (2.5 kW), highest GOR (1.53), and highest recovery (3.3%).

Table 7. Performance of in-series and in-parallel configurations without a heat recovery	exchanger.

	<b>Q</b> <sub>main</sub>	<b>Q</b> <sub>in,f,d</sub>	S <sub>f,in</sub>	S <sub>d, in</sub>	T <sub>f,in</sub>	T <sub>f,out</sub>	T <sub>d,in</sub>	T <sub>d,out</sub>	Jw		Power			A <sub>m</sub>	A <sub>m,eff</sub>
Mode	(LPM)	(LPM)	(ppt)	(ppt)	(°C)	(°C)	(°C)	(°C)	(L/h)	R	(kW)	GOR	Ν	(ft²)	(ft²)
COU	9.0	2.3	35.0	0.0	70.0	61.4	47.4	55.7	6.0	1.1	5.7	0.68	4	8.0	5.2
COU	3.0	3.0	35.0	0.0	70.0	40.5	30.0	57.9	6.0	3.3	6.1	0.64	6	12.0	7.9
(	Mode COU COU	QmainMode(LPM)COU9.0COU3.0	Qmain         Qin,f,d           Mode         (LPM)         (LPM)           COU         9.0         2.3           COU         3.0         3.0	Qmain         Qin,f,d         Sf,in           Mode         (LPM)         (LPM)         (ppt)           COU         9.0         2.3         35.0           COU         3.0         35.0         35.0	Qmain         Qin,f,d         Sf,in         Sd,in           Mode         (LPM)         (LPM)         (ppt)         (ppt)           COU         9.0         2.3         35.0         0.0           COU         3.0         3.0         35.0         0.0	Q <sub>main</sub> Q <sub>in,f,d</sub> S <sub>f,in</sub> S <sub>d,in</sub> T <sub>f,in</sub> Mode         (LPM)         (LPM)         (ppt)         (o°C)           COU         9.0         2.3         35.0         0.0         70.0           COU         3.0         35.0         0.0         70.0	Q <sub>main</sub> Q <sub>in,f,d</sub> S <sub>f,in</sub> S <sub>d,in</sub> T <sub>f,in</sub> T <sub>f,out</sub> Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (°C)           COU         9.0         2.3         35.0         0.0         70.0         61.4           COU         3.0         3.0         35.0         0.0         70.0         40.5	Qmain         Qin,f,d         Sf,in         Sd,in         Tf,in         Tf,out         Td,in           Mode         (LPM)         (LPM)         (ppt)         (°C)         (°C)         (°C)           COU         9.0         2.3         35.0         0.0         70.0         61.4         47.4           COU         3.0         35.0         0.0         70.0         40.5         30.0	Q <sub>main</sub> Q <sub>in,f,d</sub> S <sub>f,in</sub> S <sub>d,in</sub> T <sub>f,in</sub> T <sub>f,out</sub> T <sub>d,out</sub> Mode         (LPM)         (LPM)         S <sub>f,in</sub> S <sub>d,in</sub> T <sub>f,in</sub> T <sub>f,out</sub> T <sub>d,out</sub> T <sub>d,out</sub> COU         9.0         2.3         35.0         0.0         70.0         61.4         47.4         55.7           COU         3.0         35.0         0.0         70.0         40.5         30.0         57.9	Q <sub>main</sub> Q <sub>in,f,d</sub> S <sub>f,in</sub> S <sub>d,in</sub> T <sub>f,in</sub> T <sub>f,out</sub> T <sub>d,in</sub> T <sub>d,out</sub> J <sub>w</sub> Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (°C)         (°C)         (°C)         (°C)         (°C)         (C)         (C) <td< td=""><td>Qmain         Qin,f,d         Sf,in         Sd,in         Tf,in         Tf,out         Td,in         Td,out         Jw           Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (LL/h)         R           COU         9.0         2.3         35.0         0.0         70.0         61.4         47.4         55.7         6.0         1.1           COU         3.0         35.0         0.0         70.0         40.5         30.0         57.9         6.0         3.3</td><td>Qmain         Qin,f,d         Sf,in         Sd,in         Tf,in         Tf,out         Td,in         Td,out         Jw         Power           Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (°C)</td><td>Qmain         Qin,f,d         Sf,in         Sd,in         Tf,in         Tf,out         Td,in         Td,out         Jw         Power         Power           Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (°C)         (°C)         (°C)         (L/h)         R         Power         GOR           COU         9.0         2.3         35.0         0.0         70.0         61.4         47.4         55.7         6.0         1.1         5.7         0.68           COU         3.0         35.0         0.0         70.0         40.5         30.0         57.9         6.0         3.3         6.1         0.64</td><td>Qmain         Qin,f,d         Sf,in         Sd,in         Tf,in         Tf,out         Td,in         Td,out         Jw         Power         Power         N           Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (°C)</td><td>Qmain         Qin,f,d         Sf,in         Sf,in         Sd,in         Tf,in         Tf,out         Td,in         Td,out         Jw         Power         Mode         Am           Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (°C)         (°C)         (°C)         (L/h)         R         Power         GOR         N         (ft²)           COU         9.0         2.3         35.0         0.0         70.0         61.4         47.4         55.7         6.0         1.1         5.7         0.68         4         8.0           COU         3.0         35.0         0.0         70.0         40.5         30.0         57.9         6.0         3.3         6.1         0.64         6         12.0</td></td<>	Qmain         Qin,f,d         Sf,in         Sd,in         Tf,in         Tf,out         Td,in         Td,out         Jw           Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (LL/h)         R           COU         9.0         2.3         35.0         0.0         70.0         61.4         47.4         55.7         6.0         1.1           COU         3.0         35.0         0.0         70.0         40.5         30.0         57.9         6.0         3.3	Qmain         Qin,f,d         Sf,in         Sd,in         Tf,in         Tf,out         Td,in         Td,out         Jw         Power           Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (°C)	Qmain         Qin,f,d         Sf,in         Sd,in         Tf,in         Tf,out         Td,in         Td,out         Jw         Power         Power           Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (°C)         (°C)         (°C)         (L/h)         R         Power         GOR           COU         9.0         2.3         35.0         0.0         70.0         61.4         47.4         55.7         6.0         1.1         5.7         0.68           COU         3.0         35.0         0.0         70.0         40.5         30.0         57.9         6.0         3.3         6.1         0.64	Qmain         Qin,f,d         Sf,in         Sd,in         Tf,in         Tf,out         Td,in         Td,out         Jw         Power         Power         N           Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (°C)	Qmain         Qin,f,d         Sf,in         Sf,in         Sd,in         Tf,in         Tf,out         Td,in         Td,out         Jw         Power         Mode         Am           Mode         (LPM)         (LPM)         (ppt)         (ppt)         (°C)         (°C)         (°C)         (°C)         (L/h)         R         Power         GOR         N         (ft²)           COU         9.0         2.3         35.0         0.0         70.0         61.4         47.4         55.7         6.0         1.1         5.7         0.68         4         8.0           COU         3.0         35.0         0.0         70.0         40.5         30.0         57.9         6.0         3.3         6.1         0.64         6         12.0

\*\*\* COU = counter-current, N = number of modules

Table 8. System performance when using a heat recovery exchanger.

		V <sub>in</sub>	<b>Q</b> <sub>in, f, d</sub>	T <sub>f,in</sub>	T <sub>f,out</sub>	T <sub>d,in</sub>	T <sub>d,out</sub>	T <sub>f,HX,out</sub>	T <sub>d,HX,out</sub>	٦w		Power			A <sub>m</sub>	A <sub>m,eff</sub>
Configuration	Mode	(m/s)	(LPM)	(°C)	(°C)	(°C)	(°C)	(°C)	(°C)	(L/h)	R	(kW)	GOR	N	(ft <sup>2</sup> )	(ft <sup>2</sup> )
in-series	COU	0.279	3.0	70.0	40.5	30.0	57.9	57.5	40.7	6.0	3.3	2.5	1.53	6	12.0	7.9

\*\*\* COU = counter-current, N = number of modules

Bench-scale experiment results were expanded for use with the model. The bench-scale DCMD system has been fully integrated with LabVIEW for automated data acquisition and experiments have been performed using the new setup. The main focus of experiments has been to expand the existing dataset used to calculate the membrane vapor permeability coefficient (MVPC). The model's predictions for DCMD system performance are highly dependent on the MVPC, so it is important to assure that an accurate value is obtained. A new MVPC value of  $7.95 \times 10^{-7} \text{ kg/(m}^2 \text{ s})$  for membrane QM022 has been determined based on the combination of the new data and the data from the pre-LabVIEW system (Table 9). The assumption of a constant MVPC value is verified by observing the strong linear relationship between the vapor pressure difference and transmembrane water flux in Figure 43, with an R<sup>2</sup> value of 0.9989. The new MVPC value differs slightly from the MVPC value of  $8.13 \times 10^{-7} \text{ kg/(m}^2 \text{ s})$  determined based on pre-LabVIEW data. To assure that the new MVPC value is valid, additional experimental data using the LabVIEW-based system is being collected. A small number of experiments using the LabVIEW-based setup have already been performed but more data is needed to provide a strong basis for the model's accuracy.

 Table 9. MVPC values obtained based on data from the pre-LabVIEW system and the current LabVIEW-based system.

Dataset	MVPC (kg/(m <sup>2</sup> s))
Pre-LabVIEW	7.75E-07
Pre-LabVIEW	8.28E-07
Pre-LabVIEW	8.48E-07
Pre-LabVIEW	8.52E-07
Pre-LabVIEW	7.82E-07
Pre-LabVIEW	7.92E-07
Post-LabVIEW	7.60E-07
Post-LabVIEW	7.54E-07
Post-LabVIEW	7.63E-07
Average	7.95E-07



Figure 43. The relationship between transmembrane water flux and vapor pressure difference.

Data from the lab-scale system will also be used to verify the ability of the model to make accurate predictions for larger-scale DCMD systems based on bench-scale data

In addition to the pilot-scale DCMD fabrication, progress has also been made on the instrumentation and control for the MD subsystem and is focusing on the following tasks:

- 3.1 Instrumentation of the lab scale system for collecting data for model validation
- 3.2 Control system validation from operational dynamic data
- 3.3 Design of experiments to validate steady state model
- 3.4 Scale up of the lab scale model to pilot plant level
- 3.5 Extension of the steady state model to a dynamic model

# 3.1 Instrumentation of the lab scale system for collecting data for model validation

# 3.1.1 Current Bench- Scale Model Physical Design

Figure 44 shows the lab-scale setup for the membrane distillation sub-system at Humboldt University. Feed stream solution is pulled through an immersion heater by a peristaltic pump. The heated feed solution is delivered to the membrane module inlet where the temperature and pressure are measured. At the outlets from the membrane module, temperature and pressure are again measured before feed solution flows through a rotameter and is returned to the feed reservoir.

The temperature of the distillate solution is adjusted through use of a manually controlled heat exchanger before the solution is moved, by peristaltic pump, to the inlet of the membrane module. The temperature and pressure of the distillate stream are measured immediately before entrance to the membrane module and immediately after the stream exits the module. Distillate solution then flows through the rotameter to gage the flow rate before being returned to the distillate bath.

Typically, the feed stream solution is fed to the membrane module at higher temperature than the distillate solution. As the temperature of the feed stream solution is at a higher temperature than the distillate solution, pure water vapor passes through the membrane from the feed stream to the distillate solution stream. The inflows and outflows are measured to at a predetermined frequency to compute the flux.



Figure 44. Lab-scale DCMD system at Humboldt State University: red - feed stream; blue - distillate stream.

The instrumentation at Humboldt University is aimed at collecting data for modeling and control. Table 10 lists the typical instrumentation

Parameter	Device	Range
Temperatures	PT-RTD	1-100 °C
Inlet Pressures	Transducers	-14.7-15
Outlet Pressures	Transducers	1-100 psi
	Probe and	
Conductivities	transmitter	0-199.9 μS
Conductivities	Probe and	
	transmitter	0-199.9 mS
	Probe and	
рН	transmitter	0-14
Mass of permeate	Balance	
produced		0-5 kg

Table 10. Current data acquisition instrumentation for the bench scale DCMD system.

The data is collected at predefined sampled frequency by the LabView system. The collected data primarily include feed inlet and outlet flows, feed inlet and outlet temperatures, feed inlet salinity, distillate inlet & outlet flows, distillate inlet and outlet temperatures and distillate inlet and outlet salinity. The flux is calculated from measuring the weight at 10-minute interval. The sampling frequencies are provided in Table 11.

Table 11. Sampling frequency for the data.

	Frequency					
Parameter	Sampling	Recording				
Temperatures	1 Hz	.02 Hz				
Pressures	1 Hz	.02 Hz				
Salinities	1 Hz	.02 Hz				
рН	1 Hz	.02 Hz				
Weight of permeate produced	As available	.02 Hz				

# 3.2 Control system validation from operational dynamic data.

#### 3.2.1 Existing Control

The current lab scale system operates mostly on manual control except one. A LabVIEW resident PID controller is used to control the feed solution inlet temperature by adjusting the power to the immersion heater. Figure 45 shows the variations of the FO tank temperature and the permeate inlet temperature. The redline shows the time series data for the feed inlet temperature from an initial temperature of around 30 °C. The temperature comes to desired set point at 40 °C within 200 seconds.





The flow rate of both feed and distillate streams is controlled manually by adjusting the speed of the peristaltic pumps. The flow rate of each stream is set at the beginning of an experiment and monitored throughout the experiment. Additional flow adjustments can be made as needed but are rarely necessary after the initial stabilization period. Manual control systems performance seems to be satisfactory for our lab scale operational objectives.

# 3.3 Design of experiments to validate steady state model

#### 3.3.1 Steady State Model

At HSU, Professor Andrea Achilli and his students developed a steady state model from mass and energy balance. The approach was to start from a differential volume across the length of the membrane distillation unit and integrate it across the length of the membrane. Through successive iteration a steady state solution was achieved. Figure 46 shows the quality of model prediction against experimental data. Additional experimental and modeling conditions are described in Table 12.



Figure 46. Results for a variety of conditions show good agreement between bench-scale experimental and predicted values.

Table 12.	Variable ranges	used in e	xperiments	for control	l system	validation.
	0				~	

Variable	Range
FO tank to MD inlet flow	1.0-2.0 LPM
FO tank to MD inlet temperature	40-65 °C
FO tank salinity	0-200 g/L
Permeate Tank to MD inlet flow	1.0-2.0 LPM
Permeate Tank to MD inlet temperature	20-45 °C

A summary of the results of the DCMD modeling are:

- The model predicts outlet temperatures and water fluxes with good accuracy in both co-current and counter current operational modes (Figures 47-49).
- The membrane distillation coefficient can be assumed linear throughout the range of temperatures applicable to DCMD.
- The model can account for multiple types of spacers without specific information about their geometries.
- The accuracy of model predictions decreases with increasing feed solution salinity.



**Figure 47.** Experimental results and model predictions for water flux using a variety of temperature couples. All experiments shown were performed in counter-current flow mode with a flow rate of 1.1 LPM and feed solution concentration of 35 g/L NaCl. This pattern and level of accuracy is representative of the findings of the model validation for other experiments with different experimental conditions.



**Figure 48.** Experimental results and model predictions for water flux as a function of flow rate. All experiments shown were performed in co-current flow mode with a flow rate of 1.1 LPM and feed solution concentration of 35 g/L NaCl. The amount of error incurred is consistent across flow rates.



**Figure 49.** Experimental results and model predictions for outlet temperatures of (a) feed and (b) distillate solutions using a variety of temperature couples. All experiments shown were performed in counter-current flow mode with a flow rate of 1.1 LPM and feed solution concentration of 35 g/L NaCl.

From the experimental results it can be concluded that that the water flux in DCMD dependents on:

- the temperature difference between the distillate and feed streams
- the feed operating temperature
- the flow rates of distillate and feed streams

Increasing any of these parameters results in an increase in fresh water flux across the membrane. Increasing the feed and/or distillate flow rates reduces the residence time in the membrane module, increasing the temperature difference by reducing the temperature drops. Also, increased flow rates reduce the transport resistance for water flux across the membrane. The process design and control objectives should focus on increasing the feed inlet temperatures, reducing the distillate inlet temperatures, and increasing the distillate and feed flow rates within the tolerance of the membrane materials and the availability of waste heat from the diesel generators.

#### Scale-Up Modeling

The DCMD model was used to determine distillate water flux, specific energy consumption and outlet temperatures as a function of membrane length for the scaled up system (Table 13). The target distillate production was 6 L/hr. It was found that a heat recovery exchanger was required to reduce the specific energy of the DCMD process and to reheat the feed when longer membranes are used (Figure 50).

Parameter	Value	Unit
Membrane Length	12	ft.
Inlet Temperature Feed	70	°C
Inlet Temperature Distillate	30	°C
Velocity	0.275	m/s
Heat Required	2.5	kW
Cooling required	TBD	kW
Feed Flow	3.0	LPM
Distillate Flow	3.0	LPM
Specific energy	870	kWh/m <sup>3</sup>

**Table 13.** Design parameters for the scaled up system determined from the DCMD model. These values assume a distillate production of 6 L/hr. and the use of a heat recovery exchanger to reduce energy input to the system.



Figure 50. The results of the DCMD modeling show that the lowest specific energy and greatest production can be achieved with 12 ft. of membrane and the use of a heat recovery exchanger.

#### 3.4 Scale up of the lab scale model to pilot plant level

The next step was to build upon the modeling results to come up with a systems level simulator. The simulator was rated to treat 6 L/h. The scope of simulation was limited to the components shown in Figure 51. The simulator helped in deciding on the piping sizes, tank sizes, pump capacities, heat exchanger specifications, power consumption and membrane distillation area requirements. The simulation model was based on simple mass and energy balances.



Figure 51. Scope of simulation (bio-reactor is not included. just shown for the sake of completeness.)

#### 3.5 Extension of steady state model to dynamic model

The final task was to include dynamics in the pilot plant simulation model. The dynamic model provides information needed for the control systems design. This effort gave insight into the control design at the pilot plant level.

# Task 4: Heat Exchanger Design and CFD Modeling for MD

# SubTask 4.1 – Build computational fluid dynamics model

The flat-sheet MD system considered for an initial CFD analysis consists of two flow channels sandwiching a thin permeable membrane as shown in Figure 52(a). The feed and permeate water streams are in a parallel arrangement and are separated by the membrane. The design and dimensions of the flow channels are shown in Figure 52(b). The dimensions and operating conditions of the MD system used for CFD modeling are shown in Table 13.



**Figure 52.** (a) Membrane distillation system used in the initial CFD analysis. (b) Flow channel design of the membrane distillation system.

In MD systems, the water evaporated from the hot feed water side is transported by the saturation pressure difference through the porous membrane to the cold permeate side where the vapor condenses. The water vapor flow in the nano-scale channels of the membrane is considered as a non-continuum flow because the mean free path of the water vapor is larger than the pore diameter.

The CFD analysis performed focused on the pressure drop and heat transfer calculations in the MD system using a simplified model for water permeation through the membrane. It was assumed that water permeation through the porous membrane is governed by the Darcy-Forchheimer law, which is valid for continuum flow. Because the water permeation rate through the membrane is negligible compared to the water flow rates in flow channels, the impact of water permeation on pressure drop is rather small.

CFD results showed that water from the feed stream slowly permeates through the porous membrane to the permeate stream. The pressure is uniform across the channels due to the small flow rate of the water permeation. The temperature distribution in the MD system forms a thermal boundary layer at the permeate side due to the heat transfer and water permeation.

# SubTask 4.2: Analyze the energy and flow requirements for the MD system

# Energy Requirements for the MD system

The energy requirements of the MD system were determined by assuming that the system could process 90% of the 25 gallon/day of waste water produced by each soldier [66]. The heat exchanger system for the MD system considered for this study is shown in Figure 53 (a).

A flat-plate (liquid-to-liquid) heat exchanger that recuperates waste heat from the diesel engine coolant system was chosen due to its compactness and low pressure drop. The flat plate heat exchanger system was analyzed in order to evaluate its performance in heating the feed water stream to the temperatures required for the MD system. An air-to-liquid heat exchanger, shown in Figure 53(a), which works in series with the liquid-to-liquid heat exchanger, is neglected (i.e.,  $T_{c,o} = T_{F,i}$ ).



Figure 53. (a) Process diagram of the heat exchanger systems for the MD system considering the engine coolant as available heat source, (b) Membrane energy and flow diagram

This analysis focuses on finding the maximum number of soldiers for which the distilled water requirements can be met based on power requirements of 1000 soldiers. It was assumed that 1000 soldiers require 3.25 MW [66]. The thermodynamic efficiency and temperature specifications of the diesel-generator and the power requirements, P, of the soldiers determine the heat that is available in the engine cooling system. According to the diesel engine specifications, the temperature of the engine coolant,  $T_{h,i}$ , from the engine cooling jacket to engine radiator is 104°C [2] and the temperature of the engine coolant returning to the engine,  $T_{h,o}$ , is 80°C [67]. The feed water inlet temperature,  $T_{c,i}$ , was assumed to be 20°C. Approximately 36% of the energy in diesel fuel is converted to electricity and approximately 25% is lost to the engine coolant system [2].

# Heat Exchanger Design/Analysis

A simplified model of a flat plate, counter-flow, heat exchanger is shown in Figure 54 along with its corresponding temperature profile, where T is temperature, and the subscripts h, c, i, and o represent hot, cold, in, and out, respectively.



Figure 54. (a) Flat plate heat exchanger in counter-flow arrangement with relevant dimensions. (b) Temperature profile of a counter-flow heat exchanger

The dimensions of the flat plate heat exchanger (FlatPlate, Model #131002833) are listed in Table 14. The cold fluid stream is the feed water and hot fluid stream is engine coolant. Using the engine coolant temperatures of the diesel engine and desirable temperature of the MD system, the heat exchanger can be designed using the log mean temperature difference (LMTD) method [68].

L[m]	W [m]	$\delta_p [\mathrm{mm}]$	<i>a</i> [mm]
0.3	0.12	0.42	1.6

Table 14. Dimensions of flat plate heat exchanger

At this stage, the study does not consider the effects of the fouling factors on the heat exchanger's performance.

#### **MD** Feed Water Requirements

Considering the energy balance of the membrane system, the maximum number of soldiers whose waste water could be treated was determined for one temperature condition. The average temperature of the feed stream across the membrane was fixed at a constant 60°C. The result is shown in Table 15 along with the calculated GOR (gain output ratio,  $m_F/m_m$ ) values. The pumping power for varying heat exchanger volumes is shown in Figure 51.

Table 15. Temperature conditions and results

<i>T<sub>F,i</sub></i> [°C]	$T_{F,o}$ [°C]	Max Number of Soldiers	Effectiveness, $\varepsilon_{HX}$	GOR	Membrane Area [ <i>cm</i> <sup>2</sup> ]	<i>q<sub>c</sub></i> [kW]	q <sub>e</sub> [kW]
90	30	525	0.84	19.41	105.12	1036	912

The limiting factor is the outlet temperature of the feed water,  $T_{F,o}$ . This is due to the fact that  $T_{F,o}$  can never drop below the inlet temperature of the permeate stream. The temperature condition used exhibits the high effectiveness and reasonable GOR.

Pressure drop increases as the height of the heat exchanger decreases. This is due to the fact that increasing the height corresponds to an increase in the number of channels and a decrease in the mass flow rate per channel. This leads to a transition of the flow from a turbulent regime to a laminar regime. As shown in Figure 55, the pressure drop for the plate heat exchanger is minimal.

The conduction heat transfer is larger than the heat transfer due to evaporation. This is important because larger evaporation heat transfer allows for higher wastewater treatment. For the temperature conditions analyzed, more than



50% of the heat provided by the engine coolant is lost through conduction across the membrane. The distillation process would be more efficient if the evaporation heat transfer were increased.

Figure 55. The feed stream pressure drop across the HX for varying heat exchanger heights

# Conclusions

From the analysis, the waste water from 525 soldiers can be treated utilizing waste heat from the engine coolant of diesel power generators for 1,000 soldiers. The pressure drop across the heat exchanger is minimal. By decreasing the conduction loss across the membrane, the wastewater treatment process can be enhanced.

Half of a micro channel was modeled using ANSYS, Fluent 12.1. The saturation pressures based on the water temperature at the feed and distillate sides of the membrane were used as the inlet and outlet boundary conditions of the channel flow. The temperature of the distillate side was fixed at 20°C. The diameter of the channel is the pore diameter of the membrane and the length of the channel is equal to the membrane thickness.

The velocity contour of the water vapor flow in the microchannel of the membrane is presented in Figure 56 (a). It is shown that the velocity is at maximum in the center of the channel due the shear stress at the wall. As shown in this figure, the velocity at the wall has a non-zero value because of the velocity slip boundary condition. The velocity profiles from the analytical and CFD solutions are compared in Figure 56 (b). It is shown that both profiles are in good agreement.



Figure 56. (a) Velocity contours of the vapor flow in microchannel of the membrane, (b) Comparison of the velocity profile of the vapor flow for analytical and CFD solution ,  $T = 40^{\circ}$ C.

The mass flux results of the membrane are compared for various models for different flow regimes in Figure 57. It is shown in this figure that the mass flux increases by increasing the feed stream temperature. The slip regime model predicts the highest mass flux, while the Poiseuille-flow model predicts the lowest mass flux. It can be also shown that the mass flux results from the slip flow regime and molecular diffusion are very similar.

The figure shows that none of the flow regimes can predict the measured mass flux and all the models shows a significant deviation from the experimental data. Since the Kn of the flow in the membrane being considered lies in the transition regimes, the slip flow regime cannot predict the mass flux of the membrane as evidenced by Figure 57. The mass flux predicted by the slip flow regime is 20 times higher than the experimental data. The Knudsen diffusion model predicts 2.2 times higher mass flux than the experimental data, while the Poiseuille flow model predicts about 0.1 times of the mass flux of the experimental data.



Figure 57. Mass flux of the membrane based on different flow regimes

It can be concluded the mass flux predictions based on different models deviates significantly from the experimental data. The experimental data lies between that predicted by Knudsen diffusion and Poiseuille flow, therefore a combination of the membrane distillation coefficient of these two regimes may provide a reasonable prediction of the actual mass flux of the membrane.

The energy requirements of the MD system were determined by assuming that the system was capable of processing 90% of the 25 gallons/day of waste water produced by each soldier. A two-stage compact heat exchanger system that recuperates waste heat from a diesel engine is considered to heat the feed water to optimum temperature for membrane distillation. The first stage utilizes a flat plate heat exchanger, HX1, which recuperates waste heat from the diesel engine's coolant system. The engineering and design of the flat plate heat exchanger was discussed later. The second stage utilizes a circular tube-continuous fin heat exchanger, HX2, to recuperate waste heat from the engine's exhaust system. The energy balance and sizing of the continuous fin heat exchanger were discussed in later. The objective of this report is to quantify the system pressure drops and membrane performance as a function of HX2's volume. A process diagram of the MD system is shown in Figure 58.

#### Energy Considerations

This analysis focuses on finding the maximum number of soldiers whose waste water can be treated based on the power requirements of 1000 soldiers. It was assumed that 1000 soldiers require 3.25 MW of energy, i.e., 3.25kW per solider. The total energy in the engine available for recuperations is determined by the total power output of the diesel generator, *P*, and its thermodynamic efficiencies. According to the specifications of an industrial diesel generator, approximately 30% of the energy available in diesel fuel is lost through the engine's exhaust gas,  $\eta_{EX}$ , 25% is lost to the engine coolant system,  $\eta_{EC}$ , and only about 36% of the energy in the fuel is converted to electricity,  $\eta_P$ . Therefore, the total energy available in the cooling system is  $q_1=(\eta_{EC}/\eta_P)P$ , and in exhaust stream is  $q_{EX, tot}=(\eta_{EX}/\eta_P)P$ .



Figure 58. Process diagram of the membrane distillation system.

The energy balance equation used to find the mass flow rate of the exhaust gas,  $m_{h2}$ , and is given by

$$q_{EX,tot} = \dot{m}_{h2} c_{p,h2} \left( T_{EX,1} - T_{EX,2} \right)$$
(Equation 48)

where  $c_{p,h2}$  is the specific heat capacity of the exhaust gas,  $T_{EX,1}$  is the exhaust gas temperature inside the combustion chamber, and  $T_{EX,2}$  is the temperature of the exhaust gas at the exit of the tailpipe. The temperature inside the combustion chamber ranges from 750~900°C and the exhaust gas temperature at the tailpipe exit was assumed to be equal to ambient, 25°C. The energy balance across the coolant streams in HX1 is given by

$$q_1 = m_{h1}c_{p,h1}(T_{h1,i} - T_{h1,o})$$
(Equation 49)

where  $\dot{m}_{h1}$  is the coolant mass flow rate passing through the heat exchangers,  $c_{p,h1}$  is the specific heat capacity of the coolant and  $T_{h1,i}$  and  $T_{h1,o}$  are the inlet and outlet temperatures of the engine coolant. The energy balance across the exhaust streams in HX2 is given by

$$q_2 = \dot{m}_{h2}c_{p,h2}(T_{h2,i} - T_{h2,o})$$
(Equation 50)

where  $q_2$  is the amount of heat transferred from the exhaust stream to the feed water stream, and  $T_{h2,i}$  and  $T_{h2,o}$  are the inlet and outlet temperatures of the exhaust gas. The energy balance from the inlet HX1to the outlet HX2 is given by Eq. 51:

$$q_1 + q_2 = \dot{m}_{F,i} c_{P,F} \left( T_{F1,i} - T_{F2,o} \right)$$
(Equation 51)

where  $\dot{m}_{F,i}$  is the feed water mass flow rate and  $c_{p,F}$  is the specific heat capacity of the feed water. Additional feed water must be added to the feed stream,  $\dot{m}_{w,i}$ , in order to replace the feed water that is exiting the feed stream and

entering permeate stream,  $\dot{m}_m$ . The additional energy balance that must be included to determine the effects of adding feed water to the feed stream is given by

$$\dot{m}_{F,o}c_{p,F}T_{F3,o} + \dot{m}_{w,i}c_{p,F}T_{w,i} = \dot{m}_{F,i}c_{p,F}T_{F1,i}$$
(Equation 52)

where  $\dot{m}_{F,o}$  is the mass flow rate of the feed water exiting the membrane,  $T_{F3,o}$  is the outlet temperature of the feed water at the membrane, and  $T_{w,i}$  is the temperature of the feed water that is being added to the feed stream. It was assumed that the temperature of  $T_{w,i}$  was equal to 25°C. The energy balance on the feed water stream at the membrane is given by

$$\dot{m}_{F,i}c_{p,F}T_{F3,i} - \dot{m}_{F,o}c_{p,F}T_{F3,o} - q_c - q_e = 0$$
(Equation 53)

where  $q_c$  and  $q_e$  are the heat lost through the membrane due to conduction and evaporation, respectively.

#### Heat Exchanger Design/Sizing

A parametric study was conducted to determine the effects of the membrane area on the system's performance. Membrane area was considered for this study because it is directly related to the quantity of water that can be processed and it is a variable that can be readily changed in the MD system. It is important that the effects and process limitations be understood. The variables that were fixed for this study are outlined in Table 16.

Table 16. Relevant system variables that were fixed when varying A<sub>mem</sub>.

$T_{F3,i}$	$T_{P,i}$	$T_{w,i}$	$T_{h2,i}$	$T_{h2,o}$
70	30	25	600	200

The MD system is arranged in counter flow configuration and in order to function effectively it requires that the feed water inlet temperature at the membrane be higher than the permeate outlet temperature ( $T_{F3,i} > T_{P,o}$ ). It also requires that the feed water outlet temperature at the membrane be higher than the permeate inlet temperature ( $T_{F3,o} > T_{P,i}$ ). Figure 59 (a) suggests that the system can meet the membrane's temperature requirements and treat the waste water of approximately 2,000 soldiers, using the assumptions listed in Table 16. This figured was derived assuming that the permeate flux, *J*, remained constant and equal to 10 L/(s·m<sup>2</sup>).



Figure 59. (a) System temperatures and the number of soldiers whose waste water can be processed vs. membrane area, (b) Heat exchanger and membrane performance vs. membrane area.

The effectiveness of the heat exchangers remained virtually unchanged as shown in Figure 60 (b). However, the performance of the system improves significantly (lower GOR) with increasing membrane area. This is because the mass transfer into the permeate stream is directly proportional to the membrane area as shown in Eq. 51. Furthermore, the energy balance at the membrane, given by Eq. 52, shows that  $m_{F,i}$  must decrease in order to keep the system balanced.

# The effects of HX volume on HX and membrane performance

A parametric study was conducted to determine the effects that the volume of HX2 had on the system's performance. All studies were conducted utilizing the heat exchanger geometry detailed in Figure 11. The performance of HX2 was assessed by quantifying the effects that HX volume had on the HX's performance criterion (pressure drops and effectiveness) as well as the performance criteria of the membrane (GOR). System pressure drops and HX effectiveness as a function of  $V_{HX2}$  is plotted in Figure 60. For this study, the frontal area ( $L_2$  and  $L_3$ ) was fixed and only the depth of the heat exchanger was adjusted. Additional constraints that were imposed are detailed in Table 17.

Table 17. Relevant system variables that were fixed when varying  $V_{HX2}$ .

$T_{F3,i}$	$T_{P,i}$	$T_{w,i}$	$A_{mem}$	$N_s$
70	30	25	450	1268

It was found that decreasing  $V_{HX2}$  significantly affected the pressure drop experienced by the exhaust gas. This is due to the fact that as  $V_{HX2}$  becomes smaller, the exhaust flow passage becomes more constricted and leads to higher exhaust velocities.

Additionally, the effectiveness of the continuous fin HX rises as  $V_{HX2}$  is increased. The effectiveness of the heat exchangers was discussed earlier. The rise in effectiveness is to be expected because increasing  $V_{HX2}$  leads to a much larger heat transfer area and allows for the HX to approach the theoretical maximum heat transfer. It was also found that the effectiveness of the plate HX,  $\varepsilon_{HX1}$ , decreases when  $V_{HX2}$  is increased. This is primarily due to the fact that as  $V_{HX2}$  increases, the feed water mass flow rate in the heat exchangers,  $\dot{m}_{F,i}$ , must also increase in order to keep  $T_{F3,i}$  constant while accommodating the larger influx of heat at HX2,  $q_2$ . This causes the minimum heat capacity of HX1 to rise, thus decreasing the effectiveness of HX1.



Figure 60. Pressure drop and heat exchanger effectiveness vs. volume of the continuous fin heat exchanger.

Increasing  $V_{HX2}$  was found to be detrimental to the performance of the membrane and led to larger GOR numbers (GOR =  $\dot{m}_{F,i}/\dot{m}_m$ ). Figure 61 shows that the feedwater mass flowrate increases with increasing HX volume. Again, this is due to  $\dot{m}_{F,i}$  having to increase in order to maintain a constant inlet temperature at the membrane.



Figure 61. Membrane performance (GOR) vs. the volume of the continuous fin heat exchanger.

# Task 5: Lab-Scale BIO+FO Sub-System Testing

#### Bench-Scale Biofouling Analysis: Sidestream vs. Submerged Configuration

A key operating consideration associated with OMBRs is the salt concentration within the bioreactor because high concentration can inhibit microbial degradation and nutrient removal. Reverse salt flux through FO membranes may be influenced by the membrane configuration, but there is limited knowledge on the advantages of different FO configurations with respect to water flux, reverse salt flux, and fouling propensity. The bench-scale sidestream and submerged FO modules were developed and tested with activated sludge. The results indicated lower steady-state water flux for the submerged configuration; however, fouling tests did not include the effect of aeration to control membrane fouling of the submerged membranes.

The bench-scale submerged module was modified with a shroud that enabled air scour at the surface of the submerged membranes (Figure 62). Fouling tests were carried out with three different FO membrane configurations for OMBRs; submerged without aeration (SUB), submerged with aeration (SUB\*), and sidestream (SS) configurations were compared using two draw solution concentrations (35 g/L and 100 g/L NaCl). The bench-scale systems were tested using activated sludge from Hyperion wastewater treatment plant in Los Angeles. The MLSS was  $7.32 \pm 0.08$  g/L and the MLVSS was  $3.35 \pm 0.04$  g/L. The bioreactor was continuously aerated to maintain a DO concentration over 3.0 mg/L and the pH was  $7.4 \pm 0.2$  over the course of the experiments. These values are characteristic of OMBR systems with high solids retention for treatment of high-strength domestic wastewaters.



Figure 62. Close-up image of a cell from the bench-scale submerged module modified with a shroud to direct aeration along the membrane surface for fouling control.

For the 35 g/L NaCl draw solution, all configurations began with the same initial water flux of approximately 9 L/m2hr (Figure 63.a). As expected, shear from aeration (SUB\*) or crossflow (SS) effectively mitigated membrane fouling compared to the configuration without shear (SUB). Flux decline from fouling was 81% (Figure 63.a, boxed region) for the SUB configuration and 21% for the SUB\* configuration. Over the same period, flux decline was 11% for the SS configuration. Interestingly, the steady-state water flux with fouled membranes was essentially equal for the SUB\* and SS configurations, indicating that shear from aeration (aeration rate =  $1.5 \text{ m}^3 \text{ m}^{-1} \text{ h}^{-1}$ ) was effectively equal to shear from crossflow (20 cm/s). After removing the membranes and rinsing them with DI water, greater than 98% of the initial water flux was recovered for all configurations (Figure 63.a).



**Figure 63.** Experimental water flux for submerged, submerged\*, and sidestream configurations with (a) 35 g/L and (b) 100 g/L NaCl draw solution concentrations. The boxed regions indicate the data point from which the average steady-state water flux was calculated.

As expected, the 100 g/L NaCl draw solution resulted in greater initial water flux (~19 L/m<sup>2</sup>-hr; Figure 63.b) compared to the 35 g/L NaCl draw solution. However, membrane fouling resulted in greater water flux decline. The greater flux

decline is the result of more foulant deposition due to the greater initial water fluxes. Flux declined 91% for the SUB configuration, 51% for the SUB\* configuration, and 51% for the SS configuration. Interestingly, the steady-state water flux for each configuration was practically identical with *both* draw solution concentrations; in other words, a nearly three-fold increase in the osmotic driving force did not result in a greater steady-state water flux. This suggests that OMBRs may have an osmotically controlled "homeostatic" flux that is analogous to the critical flux in MBRs with UF/MF membranes, and it implies that FO membrane fouling plays a more significant role in homeostatic water flux than the osmotic driving force beyond the  $\Delta \pi$  at which the homeostatic flux occurs.

For MBRs, the critical flux is the point where increasing the transmembrane pressure (TMP) driving force beyond a critical TMP does not result in a stable operation; it only results in greater membrane fouling. For OMBRs, the homeostatic water flux can be defined as the point where increasing the osmotic pressure driving force beyond a critical  $\Delta \pi$  does not result in greater water flux; it does however result in thicker foulant cake layers as measured by cross-sections obtained with scanning electron microscopy (Table 18). The critical flux of an MBR has been well defined as a function of the membrane and the membrane configuration (and cleaning mechanism), the feed solution, the operating conditions (SRT, HRT, crossflow), and the driving force. These same parameters would apply to the homeostatic flux in OMBRs, however, one key difference is the presence of dissolved solutes as well as solids in OMBR systems. Exceeding the  $\Delta \pi$  at which the homeostatic is achieved may result in greater RSF without greater water flux.

 Table 18. Foulant cake layer thickness (µm) for cake layers formed with different OMBR membrane configurations and two draw solution concentrations.

Configuration	35 g/L NaCl Draw Solution			100 g/L NaCl Draw Solution		
Submerged	15.6	+1	3.2	13.8	±	2.3
Submerged*	6.3	+1	1.3	18.3	±	3.1
Sidestream	2.7	±	0.7	12.5	±	2.0

The specific reverse salt flux (SRSF), which has the units of grams of draw solute reverse fluxed per liter of FO permeate, is a useful performance parameter. The SRSF values were calculated for new, fouled, and cleaned membranes with both draw solution concentrations (Figure 64). The initial SRSF with NaCl feed water and 35 g/L NaCl draw solution was smallest to largest in the order of SS, SUB, and SUB\* (Figure 64). Lower initial SRSF in the SS configuration may be the result of hydraulic pressure in the feed side channel (3.5 psi) from recirculation pumping, whereas there is no pumping on the feed side in the SUB and SUB\* configurations (hydrostatic pressure = 0.2 psi). Indeed, lower RSF with increasing pressure on the feed side has been reported in pressure assisted osmosis studies.



**Figure 64.** Specific reverse salt flux during each experimental stage for submerged, submerged\*, and sidestream configurations with (a) 35 g/L and (b) 100 g/L NaCl draw solution concentrations.

For submerged membranes, initial SRSF was greater with aeration. Without aeration, draw solutes that reverse diffuse through the membrane will concentrate on the feed side and mix into the bulk feed solution only by diffusion or from incidental bulk feed solution mixing in the reactor. This salinity gradient at the selective layer surface, which is similar to concentrative external concentration polarization, reduces RSF into the bulk solution. Continuous aeration along the membrane surface increases the mixing intensity at the membrane surface and induces a feed solution crossflow velocity through the aeration channel that transports draw solutes into the bulk solution, ultimately leading to greater SRSF. Initial SRSF with NaCl feed solution was similar for all configurations with both draw solution concentrations (Figure 64).

SRSF increased with fouled membranes for all configurations. Previous studies have attributed this phenomenon to a charged-based mechanism wherein a negatively charged foulant layer increases reverse diffusion of positively charged draw solutes (Na<sup>+</sup> in this case). Upon cleaning the membranes, SRSF was similar to the initial stage for all configurations, which indicates that membrane fouling was generally reversible and that decoupling of water flux and RSF with fouled membranes was not due to a chemical changes of the membranes.

A mass balance approach using the experimentally measured forward water flux and reverse salt flux with fouled membranes was used to model the salinity in an OMBR. In OMBRs, the SRT is decoupled from the HRT. For COD removal and nitrification, a 10-day SRT is recommended and for reuse applications where nutrient removal includes a denitrification step, longer SRTs are required. In the current study, we chose the lower steady-state SRSF from the 35 g/L membrane fouling tests to model OMBR salinity using Equation 54. Two SRTs for different reuse scenarios were chosen; 10 days and 40 days. The volume of the model reactor was set to 1,000 L and water production was fixed at 100 L/d to ensure the HRT was less than one day. To achieve this, the membrane area was adjusted based on the steady-state water flux from the experimental configuration testing. A 120-day period was modeled and the influent wastewater TDS was 1,000 mg/L as NaCl.

$$C_{R} = \left[Q_{in}C_{in} + J_{S} - Q_{W}C_{R_{(i-1)}}\right] \frac{t - t_{(i-1)}}{V_{R}} + C_{R_{(i-1)}}$$
(Equation 54)

where  $C_R$  is the NaCl concentration of the bioreactor;  $Q_{in}$  is the FO flux plus the wasting rate (the total wastewater influent flow is equal to the FO flux plus the wastage flow);  $C_{in}$  is the influent concentration;  $Q_W$  is the wasting rate;  $J_S$  is the reverse salt flux; and  $V_R$  is the reactor volume.

The membrane area required to meet the 100 L/d water production target was 2.39, 0.56, and 0.52 m<sup>2</sup> for the SUB, SUB\*, and SS configurations, respectively (Figure 65). Salinity reached a steady-state concentration when the incoming salinity (from RSF plus wastewater) was balanced by the sludge wasting rate (determined by the SRT). For both SRTs, the steady-state salinity increased in the order of SS, SUB\*, and SUB. Elevated bioreactor salinity may be inhibitory to biological nitrogen removal. Inhibition of nitrification has been reported to occur at salinities as low as 2 g/L NaCl and as high as 10 g/L NaCl in a sequencing batch reactor.



Figure 65. Model OMBR salinity as a function of time for submerged, submerged\*, and sidestream configurations using 35 g/L NaCl draw solution with (a) 10-day and (b) 40 day solids retention times.

Conversely, Achilli et al. (2009) reported no inhibition of nitrification with a steady-salinity of ~2.5 g/L NaCl and others have observed an inhibition period followed by a shift in the microbial community to halotolerant microorganisms that recovered nitrification in a bioreactor with a steady-state salinity of ~1 g/L NaCl. Nevertheless, maintaining low salinity is desirable because 1) no halotolerant microbial community acclimation period is required, 2) higher bioreactor salinity may lead to scaling and increased membrane fouling, and 3) lower RSF decreases the requirement for draw solution replenishment. Of the three OMBR configurations tested in this study, the SS configuration offers the advantage of lower bioreactor salinity, and ultimately more efficient carbon and nitrogen removal compared to the SUB or SUB\* configurations.

#### Bench-Scale MD Analysis: Long-Term Performance and Surface Characterization Results

The long-term performance results for 100 days of continuous operation with a feed solution concentration of 35 g/L NaCl and feed and distillate inlet operating temperatures of 64 °C and 38 °C, respectively, are shown in Figure 66. Flux declined fairly linearly with time; overall, flux decreased by 16%. Flux declined in large part due to the gradual accumulation of foulants on the surface as discussed in the previous report. After the experiment was completed, the membrane was removed from the module, desiccated, and characterized using scanning electron microscopy coupled with energy-dispersive x-ray spectroscopy (SEM-EDS), goniometry, and atomic force microscopy (AFM).

Representative SEM images and EDS results for the feed side of the unused and used membrane are shown in Figure 67. The heterogeneous fouling layer on the feed side covered a significant area of the membrane surface, completely hiding the fibrous structure in some areas. The fouling layer was largely organic matter indicated by the carbon and oxygen peaks in the EDS spectra. This likely originated from trace quantities in the DI water used for all feed solutions and/or dust accumulation, as the auxiliary tanks were not fully closed to the environment in order to maintain constant concentration and for mixing. Small quantities of fouling are to be expected for long-term operation even when treating

low-fouling solutions, such as FO draw re-concentration. The foulants were relatively strongly adhered to the membrane and, unlike NaCl deposits, were not removed by rinsing with water.



**Figure 66.** Water flux and distillate conductivity versus time for 100 days of continuous operation with 35 g/L NaCl feed solution, 64 °C inlet feed temperature, and 38 °C inlet distillate temperature.



**Figure 67.** Representative SEM images and EDS spectra of the (a) feed side of an unused membrane and the (b) feed side of a used membrane with 35 g/L NaCl feed solution, 64 °C inlet feed temperature, and 38 °C inlet distillate temperature.

Representative SEM images and EDS results for the distillate side of the unused and used membrane are shown in Figure 68. Surprisingly, the distillate side of the membrane also revealed significant visible changes in membrane

morphology. Fibrils appeared to thicken, and the node area increased relative to fibril area. Characterization results for the distillate side of used MD membranes have, to the best of our knowledge, not been reported in the literature. The EDS spectrum confirmed that these were changes in the membrane itself and were not due to fouling, as no elements other than those found on the unused membrane surface appeared.



**Figure 68.** Representative SEM images and EDS spectra of the (a) distillate side of an unused membrane and the (b) distillate side of a used membrane with 35 g/L NaCl feed solution, 64 °C inlet feed temperature, and 38 °C inlet distillate temperature.

Contact angle measurements for the unused and used membrane samples for the long-term experiment are shown in Table 19. The contact angle on the feed side of the used membrane was reduced by 56%, as expected due to the visible fouling layer on the surface. However, the 26% reduction on the distillate side was surprising given the lack of external deposits. The hydrophilization of the membrane surface can thus be attributed to physical or chemical changes in the membrane itself. Representative AFM images and surface roughness results for the unused and used membrane samples are shown in Figure 69 and Table 20, respectively.

Table 19. Contact angles measured on unused membrane and membrane used in the long-term experiment with 35g/L NaCl feed solution, 64 °C inlet feed temperature, and 38 °C inlet distillate temperature.

Membrane	Side	Water Contact Angle
Unusod	Feed	$140\pm3.96\ensuremath{^\circ}$
Onused	Distillate	$140\pm3.63$ °



Figure 69. Representative 3-D AFM images of the (a) feed side of an unused membrane, the (b) feed side of a used membrane, the (c) distillate side of an unused membrane, and the (d) distillate side of a used membrane.

Table 20. Roughness values for membrane samples used in long-term experiment with 35 g/L NaCl feed solution,64 °C inlet feed temperature, and 38 °C inlet distillate temperature.

Membrane	Side	Root mean square roughness (nm)
Unused	Feed	$31.0 \pm 5.40$
Unused	Distillate	27.1 ± 7.58
Used	Feed	$3.0\pm0.88$
	Distillate	$14.5 \pm 3.11$

Roughness decreased by 92% on the feed side due to the fouling layer. Roughness also decreased by 57% on the distillate side due to differences in the membrane surface itself. As increased roughness correlates with increased hydrophobicity, these results supported the contact angle measurements and qualitatively correspond to the SEM results. The changes observed on the distillate side of the used membrane are changes in the membrane itself and as such are likely to be effectively permanent, reducing membrane lifetime. Also, surface modifications for polymeric membranes (such as antifouling capabilities or increased hydrophobicity via increased roughness) may be less effective over long-term operation due to surface changes caused by use. However, despite the significant changes in membrane surface properties on both sides, distillate conductivity remained very low and high rejection was maintained. This suggests that feed-side surface properties alone are not responsible for long-term salt rejection; furthermore, a substantial hydrophobic layer and/or hydrophobic distillate-side surface may be important to maintain long-term rejection. The role of distillate side surface properties and pore properties in maintaining long-term performance are currently being analyzed.

The design of the scaled up DCMD system was later finalized and included fabrication and assembly of many of the individual components of the system, including the membrane module (Figure 70.a), heater assembly (Figure 70.b), circulation tanks, and heat exchanger assemblies.



Figure 70. Stacked DCMD membrane modules (a) and parallel heating assembly (b).

The stacked membrane module (Figure 70.a) has been designed to operate in series, essentially simulating a single 12 ft<sup>2</sup> membrane. The bolts shown are used to clamp the modules together. The two outer modules do not have a membrane channel; these are used in conjunction with the bolts to compress the remaining interior module O-rings together. The plastic bags shown coming out of the modules are not actual membrane, but rather, they were used in a testing phase to ensure no leaks were present in the membrane module. The final DCMD module stack had a total of eight cells (three shown above). The heating assembly is also shown (Figure 70.b). These units are connected in parallel to provide even heating to the DCMD feed stream. Integration of these system components, as well as others mentioned above, was completed during this task.

Further development of the LabVIEW program that controls the scale up DCMD system was worked on during this phase of research. The final step in writing and adjusting this program occurred in subsequent tasks when the system was fully integrated.

Preliminary efforts were made during this phase of research to identify and plan for emergency system shutoff triggers. Implementing these measures included creating code within the LabVIEW program and mechanical triggers, both of which were further developed as the system was integrated.

Construction of the system has been composed of several subcomponents, including: the water storage and circulation system, the heating system, the electrical system, and the membrane cell system. Construction of the individual water tanks was completed. The water storage and circulation system developed was composed of three tanks; two associated with the feed stream (Figure 71.a) and one for the distillate stream. The design and operation of these tanks was completed such that experiments using the system were able to operate without the addition or subtraction of water, which was accomplished through the use of an actuated control valve (Figure 71.b) that recirculates produced distillate water as clear water to be used on the feed side. The tanks were equipped with instrumentation that measures pressure and salinity, both of which were used for the operation of the system.



Figure 71. (a) Feed clear water and circulation tank and (b) actuated control valve on distillate tank.

The heating system was constructed and attached to the cart. The heating system consisted of three 1.5 kW submersed electrical heaters operated in parallel (Figure 71.a). Parallel arrangement was chosen to avoid the use of a cascading system and to ultimately provide heat in the most stable method possible. The quantity of heat energy provided by the heating system can be changed by unplugging individual heating elements, which allows for experiments to be run at membrane areas smaller than the final membrane size of 12 ft<sup>2</sup>.



Figure 72. (a) Parallel heating system and (b) electronics control box.

The electrical system was constructed and mounted to the cart (Figure 72.b). All electrical components have been placed into a mixed media vented plastic box, which was designed to help protect the electronics from moisture or other hazards present on the cart. The electronics and data analysis system has been designed to allow for 100% automated operation.

The scaled up membrane distillation system is shown in Figure 73. In order to construct a modular and compact system, the membrane module configuration was not designed in a series configuration that was originally modeled. The system developed was a hybrid configuration that had the feed stream configured in parallel and the distillate stream configured in series. The model was upgraded to properly predict fluxes and outlet temperatures for this modified arrangement. Trouble shooting of the physical system was completed and optimization of the control program occurred during this task.



Figure 73. Photograph of scaled-up MD system.

The LabVIEW program that controls the DCMD system was modified to incorporate new hardware included in the scaled-up system. Several safety shutoffs were included in this version of the control program and the program was refined and tested. The new hardware configuration enabled more frequent data sampling, which resulted in less noise through the averaging of data from smaller time steps.

The experiments performed on the scaled-up system allowed the research team to:

- Determine how differences in heat losses in the larger and smaller systems affect model accuracy
- Obtain data to modify the model to address different flow configurations
- Refine the existing model with larger membranes
- Identify potential problems for larger systems

In parallel, a 2-D CFD model using a viscous flow model for the membrane mass transfer was developed to simulate the performance of the scaled-up MD system (Figure 74.a). Figure 74.b shows the 2-D CFD model with the equivalent channel length of the MD system using a counterflow configuration between the feed and distillate flows.



Figure 74. (a) Scale-up MD system and (b) 2-D CFD model with the equivalent channel length of the MD system.

The membrane dimensions used for the CFD simulation are listed in Table 22. The effective dimensions were used to account for the spacers installed in the MD channels. It was assumed that the effect of the spacers on the turbulent characteristics of the channel flow was negligible. Table 22 lists the thermo-physical and geometrical properties which are the calibrated values within the measurement uncertainties of the properties and dimensions of the membrane (QM022) for best agreement with the experimental results.

Table 21.	Dimensions	of the	scale-up	MD	system.
					-

Effective width	9 in	22.86 cm
Effective length	21 in	53.34
Channel height	0.031 in	0.0787 cm
Effective height	0.02532 in	0.0643

 Table 22. Thermo-physical and geometrical properties of the MD system.

ε	τ	δ <sub>m</sub>	r <sub>p</sub>	k
0.64	2.34	<mark>84 [</mark> μm]	0.14 [µm]	0.29 [W/mK]

The CFD simulation results verified that the predicted water flow rate across the membrane was in excellent agreement with the experimental result (within 1.46%). The predicted outlet temperatures ( $T_{\text{F,out}}$ ,  $T_{\text{D,out}}$ ) of the feed and distillate flows were also in a good agreement with the measured temperatures as shown in Table 23.

**Table 23.** Comparison between the experimental and CFD results.

Exp. Mass [kg/s	5] Mass [kg/s	]	error %		
0.000415917	0.000422		-1.46		
Experimental	temperatures		C	FD	
Experimental	temperatures				
T <sub>F,in</sub>	T <sub>D,in</sub>		T <sub>F,in</sub>		T <sub>D, in</sub>
70°C	47.4°C		70°C		47.4°C
T <sub>F,out</sub>	T <sub>D,out</sub>		T <sub>F,out</sub>		T <sub>D, out</sub>

The lab-scale OMBR system was outfitted with a suite of probes capable of continuous data acquisition for a series of water quality parameters. These probes consist of:

- DO probe (IQ SensorNet FDO<sup>®</sup> 701x IQ SW)
- pH and oxidation reduction potential (ORP) probe (IQ Sensor Net SensoLyt<sup>®</sup> 700 IQ SW)
- Ammonium probe (AmmoLyt<sup>®Plus</sup> 700 IQ)
- Nitrite/nitrate and carbon probe (NiCaVis 701 IQ NI)

All of these probes were connected to a controller (IQ Sensor NET MIQ/TC 2020 XT) that was capable of internal logging as well as providing a six-channel output to a data collection interface (i.e., LabVIEW). After installation within the OMBR system, the four probes were placed online and initial checks for all electrodes and sensors were performed. The probes were situated and housed in the lab-scale OMBR as illustrated in Figure 75. Water quality matrix adjustments for applicable probes was conducted following the inoculation of the reactor with return activated sludge, primary effluent, and a synthetic feed solution.



**Figure 75.** Lab-scale OMBR equipped with new probe series. Top from left to right: pH/ORP probe (IQ Sensor Net SensoLyt<sup>®</sup> 700 IQ SW), ammonium probe (AmmoLyt<sup>®Plus</sup> 700 IQ), DO probe (IQ SensorNet FDO<sup>®</sup> 701x IQ SW). Bottom: nitrate/nitrite/carbon probe (NiCaVis 701 IQ NI).

# Task 6: Design of Lab-Scale MD + Heat Exchanger Sub-System

The energy requirements of the lab-scale MD/HX subsystem were determined by assuming that the MD system is required to produce 6~10 L/hr distilled water. In other words, the brine water influx is 6~10 L/hr because the MD system is required to match the same amount of distilled water production to maintain the mass balance of the system. A schematic diagram of the MD/HX subsystem is shown in Figure 76.



Figure 76. Process diagram of the MD and heat exchanger subsystem.

The operating ranges and the baseline conditions considered for the analysis is listed in Table 24. The operating temperature range is determined considering the water boiling temperature and the maximum membrane temperature.

Operating range of the MD system considered for the analysis						
$T_{F,i}[^{\circ}\mathrm{C}]$		$T_{P,i}[^{\circ}\mathrm{C}]$		$T_{w,i}[^{\circ}\mathrm{C}]$		
65~80	65~80 35~50 30					
Baselin	e operating conditions	and MD system dimensi	on considered for the	analysis		
$T_{F,i}[^{\circ}\mathrm{C}]$	$T_{P,i}[^{\circ}\mathrm{C}]$	$T_{w,i}$ [°C]	<i>A<sub>m</sub></i> [76]	$\dot{m}_w$ [L/hr]		
72.5	42.5	30	1.0	6~10		

Table 24. Operating range and baseline operating conditions of the MD and heat exchanger subsystem.

The variation of the required feed inlet flow rates for different feed and permeate inlet temperatures is shown in Figure 77. It can be seen that the lowest feed flow rate is required at the highest feed inlet temperature and lowest permeate inlet temperature. Since the sensible heat of the feed stream is the energy source high flow rates of the feed stream is required to provide the large energy necessary for the latent heat and conduction loss. Consequently, the high flow rate of the feed stream through the very small channels of the membrane system consume large pumping power.


**Figure 77.** Variation of the feed flow rate for different feed and permeate inlet temperatures of the lab-scale MD and heat exchanger subsystem with (a) 10 L/hr and (b) 6 L/hr water.

The variation of the required heat input for different feed and permeate inlet temperatures is presented in Figure 78. It can be seen that the heat input decreases by increasing the feed and permeate inlet temperatures. This result is attributed to the low conduction heat transfer at high feed and permeate inlet temperatures.



**Figure 78.** Variation of the heat input for different feed and permeate inlet temperatures of the lab-scale MD and heat exchanger subsystem with (a) 10 L/hr and (b) 6 L/hr water production.

As shown in Figure 79, the feed flow rate is very high compared to the required distilled water production. The high flow rate creates a high pressure drop in the system causing the excessive pumping power consumption. The results from Figures 77 and 78 suggest that (i) the high feed stream temperature and low permeate stream temperature minimize the feed flow rate and (ii) the high feed and permeate stream temperatures minimize the heat input requirement.

Heat recovery for an MD system is possible if the permeate temperature from the MD system exceeds the feed temperature. Favorable temperature conditions for the heat recovery can happen when the feed and permeate streams are in a counterflow arrangement for the MD system. Figure 79 shows the heat recovery heat exchanger installed between the permeate and feed streams.



Figure 79. Heat recovery heat exchanger for MD system.

To estimate heat transfer rate for the heat recovery HX, the following conditions were assumed for the heat transfer analysis:

- The volumetric flow rates of the feed and permeate streams are equal and range from 1 to 5 LPM.
- Feed inlet temperature  $(T_{f,i})$  to the heat recovery HX is fixed at 40°C which is the average temperature of the feed stream from the MD system.
- Permeate inlet temperature  $(T_{p,i})$  to the heat recovery HX is 40°C+ $\Delta T_i$ . The inlet temperature difference,  $\Delta T_i$  changes from 5 to 30°C, i.e.,  $T_{p,i} = 45 \sim 70$ °C.

The results of the heat transfer rate between the feed and permeate streams are shown in Figure 80. As the flow rate increases, the heat transfer rate increases. The heat transfer rate increases more rapidly with the flow rate, as the inlet temperature ( $\Delta T_i$ ) increases. The outlet temperatures of the feed and permeate streams from the heat exchanger are shown in Figure 81. Note that the inlet temperatures of the feed and permeate streams into the heat exchanger are 40°C and 40°C +  $\Delta T_i$ , respectively. The effectiveness of the heat exchanger as an efficiency of the heat exchanger, is decreased from 0.68 to 0.73 as the flow rate is increased.



Figure 80. Variation of the heat transfer rate of the heat recovery heat exchanger of the MD system for different inlet temperature differences.

The results of the thermal performance of the heat exchanger are based on the favorable temperature conditions ( $\Delta T_i > 0$ ) which is essential to the heat recovery approach. In this case, the MD system can greatly benefit from the heat recovery and therefore it is recommended to purchase and install it in the MD system. If the measured temperatures are less desirable (i.e., small  $\Delta T_i$ ) or unfavorable (i.e.,  $\Delta T_i < 0$ ), the heat recovery heat exchanger should be avoided

because of an additional pressure drop in the heat exchanger which will reduce the flow rate or increase the pumping power and therefore affect negatively the MD performance.



**Figure 81.** Outlet temperatures of the (a) feed and (b) permeate streams from the heat recovery heat exchanger of the MD system for different inlet temperature differences. Note that the inlet temperatures of the feed and permeate streams into the heat exchanger are  $40^{\circ}$ C and  $40^{\circ}$ C +  $\Delta$ T<sub>i</sub>, respectively.

The temperature and velocity distribution in the feed and distillate channels are shown in Figure 82. It is clearly observed that a thin thermal boundary layer is formed near the membrane because of the heat transfer during the phase changes (evaporation on the feed side of the membrane and condensation on the distillate side of the membrane) (Figure 82.a). However, the temperature change across the channel height is very small as compared to the change along the flow directions. It is also observed in Figure 82.b that the velocity distributions in the channels shows a fully-developed (parabolic) velocity profile in each channel and also thin velocity boundary layers near the membrane.



Figure 82. (a) Temperature and (b) velocity distributions in the feed and distillate channels magnified at the center region of the MD system.

The effects of the membrane properties on the membrane mass flow rate are discussed below. It is observed from the simulation results (Figures 83-84) that the viscous flow regime model shows the best agreement with the experimental results, while the Knudsen-diffusion model shows the worst agreement. The molecular-diffusion model and the molecular-/Knudsen-diffusion model shows moderate accuracies.

Figures 83(a) and (b) shows the variations of the membrane mass transfer rate for different porosities and tortuosities, respectively. Larger the porosity and smaller the tortuosity are, larger the mass transport rate is. This is because the larger porosity and short flow passage (tortuosity) decreases the flow resistance of the water vapor flow in the membrane pores.



Figure 83. Variation of the membrane mass transfer rates for different (a) membrane porosities and (b) tortuosities.

Figures 84 (a) and (b) shows the variations of the membrane mass transfer rate for different membrane thicknesses and pore radii, respectively. As can be seen, thinner the membrane is, larger the mass transport rate is. This is because the thinner membrane decreases the flow resistance of the water vapor in the membrane pores.



Figure 84. Variation of the membrane mass transfer rate for different (a) membrane thicknesses and (b) pore radii.

Figures 85 shows the variation of the membrane mass transfer rate for different thermal conductivities of the membrane. Lower the thermal conductivity is, larger the mass transfer rate is. This is because the low thermal conductivity decreases the conduction heat loss across the membrane from the feed flow to distillate flow.



Figure 85. Variation of the membrane mass transfer rate for different membrane thermal conductivities.

#### Lab-Scale Heat Exchanger Design

Diesel power generator used in FOB provides a significant amount of waste heat from engine coolant (20% of fuel energy) and exhaust gas (30% of fuel energy) which can distill wastewater in a MD system to produce distilled water. Heat exchanger systems are used to extract heat from the waste heat sources and supply the extracted heat to the MD system. Due to different duty cycle and thermal operating conditions (temperature, working fluid, energy amount), the integration of the heat exchanger systems with the Diesel power generator would require an temperature control to match the energy supply from the waste sources and demand from the MD system at high thermal efficiencies.

To achieve a reliable and automated operation of the MD system at FOB, first, we propose to use an electrical water heater to mimic the waste heat from Diesel engine coolant. The same conditions (e.g., inlet and exit temperatures and coolant flow rate) of the Diesel engine coolant will be used for the electric water heater. Secondly, we propose to use an electrical air heater or gas-fired furnace to mimic the engine exhaust gas and provide the same conditions (e.g., gas temperature and flow rate). The schematic of the original heat exchanger system using a Diesel generator as waste heat source is shown in Figure 86(a). The heat exchanger system using a mock heat sources is shown in Figure 86(b).

There are two types of heat exchangers used in the heat exchanger system for the Diesel generator: liquid-to-liquid heat exchanger between the engine coolant and MD feed fluid and gas-to-liquid heat exchanger between the engine exhaust gas and MD feed fluid. The typical design of the heat exchangers are shown in Figure 87. The specifications (dimensions, type) of the heat exchanger will be determined based on the energy requirements (3 kW) of the lab-scale MD system. The desirable inlet ad outlet temperatures of the feed water of the MD system are 30°C and 70°C, respectively. Based on the thermal analysis of a 525-solider MD system, it was estimated that the volume of the liquid-to-liquid heat exchanger (design: flat plate heat exchanger) are 0.015 m<sup>3</sup> [=0.4 m (H) × 0.12 m (W) × 0.3 m (L)] and the volume of the gas-to-liquid heat exchanger (design: compact heat exchanger) are 0.324 m<sup>3</sup> [=0.45 m (H) × 1.6 m (W) × 0.45 m (L)]. Note that the engine coolant temperature (< 105°C) and engine exhaust gas temperature (< 600°C) are normal conditions.



**Figure 86.** (a) Original heat exchanger system using Diesel generator. (b) Simplified heat exchanger system using mock heat sources (electrical water heater; electric air heater or gas-fired furnace)



**Figure 87.** (a) Liquid-to-liquid heat exchanger between the engine coolant and MD feed fluid and (b) gas-to-liquid heat exchanger between the engine exhaust gas and MD feed fluid

#### Task 7: Lab-Scale FO-BIO and MD-Heat Exchanger Sub-System Assembly

After the two sub-systems were physically attached, modifications to each sub-system were applied to facilitate the feed transfer between different tanks. Also, a LabVIEW control system was developed for monitoring and controlling the operation of the OMBR system.



Figure 88. Forklift lifting the submerged FO membrane cassettes from the bioreactor.

After restarting the FO-BIO sub-system the team collected several weeks of data to establish baseline flux prior to inoculation with biomass. At that point, a building-level electrical issue damaged several control components. The damage required repurposing MD sub-system components to the FO-BIO sub-system and the implementation of a

secondary method for measuring water flux and reverse salt flux. The team is working with electrical division of the Facilities and Maintenance Department at the university to isolate and solve the problem.

FO bench-scale analyses were conducted to determine and evaluate side-stream and submerged FO configurations for the pilot-scale OMBR, and a salt mass balance was developed to model the steady-state bioreactor salinity. Ongoing MD bench-scale testing was also carried out to test the long-term operation of and performance of MD membranes.

#### FO-BIO Sub-System Modifications

In late July the team scheduled a re-initialization of the FO-BIO sub-system. The reactor was emptied, providing an opportunity to install an automated wasting valve to the bottom of the tank and to replace the FO membranes. The membranes that were removed were visually inspected and fouling was analyzed using confocal microscopy. New membranes were installed in the lab-scale system using silicone adhesive as a sealant. Figure 89 shows a fouled membrane cassette after being removed from the bioreactor (left), installation of a new membrane (center), and confocal imaging of a fouled membrane (right). The fouled membrane was stored in DI water prior to imaging, thus the loosely adhered foulant cake was removed. The pattern of the woven embedded mesh used as support material in the FO membranes and a small degree of biofouling can be seen in the confocal images. The new membranes were wetted during the replacement process by misting for 24 hours to allow the sealant to set up prior to being submerged in the bioreactor. In addition to replacing the membranes, an automated wasting valve was installed and a protective case for the instrumentation and control electronics was mounted to the front of the reactor (Figure 90).



**Figure 89.** A fouled membrane module after removal from the bioreactor (left), membrane replacement using silicone sealant to adhere the membrane to the cassette (center), and an image of the fouled membranes taken using a confocal microscope (right).



Figure 90. A steel cage was mounted to the front of the bioreactor to protect electronics. Acrylic sheeting will be mounted to the frame for protection.

Prior to re-inoculating the bioreactor, clean water flux tests were performed to confirm baseline clean water flux of  $\sim 6 \text{ L/m}^2$ -hr from previously established laboratory tests. These results were verified both manually, and with the LabVIEW program and associated instrumentation and controls.

#### MD System Modifications

Titanium heat exchangers (55,000 BTU; Beta Heat Exchangers model SP-55K) were installed to replace the existing copper braised heat exchangers and prevent corrosion in the MD distillate and feed loops (Figure 91). The distillate side is plumbed to the chiller and the feed side is plumbed to the heaters.



Figure 91. Titanium heat exchangers in the MD feed and distillate loops.

Two manifolds were installed on the inlet and outlet streams of both feed and distillate side of the MD membrane to enable running the MD system with parallel flow pattern in the cassettes (Figure 92). The parallel flow configuration offers higher temperature gradients across all membrane cassettes and a more uniform pressure difference across the membrane within each cassette. The effects of multiple operational parameters on the MD sub-system performance were evaluated, and the framework of these experiments is summarized in Table 25. Comparing the results of experiments 1 and 3, with co-current and counter-current configurations, respectively (Figure 93), shows that the counter-current configuration requires higher energy compared to the co-current scheme. Lower flow rates resulted in lower energy requirements but also resulted in lower distillate fluxes. Overall, different operational conditions seemed to have lower fluxes than the manufacturer's reported flux.



Figure 92. MD module placed in vertical position to improve mixing and minimize channeling.

Exp.	Configuration	Flowrate (L/min)	ΔP (psi)	T <sub>heater</sub> (°C)	T <sub>feed(in)</sub> (°C)	T <sub>feed(out)</sub> (°C)	T <sub>chiller</sub> (°C)	T <sub>distil.(in)</sub> (°C)	T <sub>distil.(in)</sub> (°C)	Permeate flux (L/m <sup>2</sup> -hr)
1	1 <sup>a</sup>	1.5	1	100	60	25	7	13	42	1.6
2	1	0.75	0.5	100	72	22	7	7	42	1
3	2 <sup>b</sup>	1	0.7	80	70	19	2	11	54	1.8
4	2	1	0.7	90	70	35	2	6	27	1.8
5	2	1.5	1	95	70	42	4	10	34	1.8
6	2	1.5	1	93	72	47	3	10	30	1.9
7	2	1.5	1	91	75	55	3	7	23	2.0
8	2	1.75	1.2	100	70	42	7	10	33	1.7
9	2	2	1.6	100	70	44	8	12.5	35	1.6

Table 25. Experimental design for testing MD sub-system (parallel flow configuration)

<sup>a</sup> Configuration 1 in Figure 93

<sup>b</sup> Configuration 2 in Figure 93

<sup>c</sup> Insufficient heat exchange, the temperature was not reached.



Figure 93. MD parallel flow configuration: 1) counter- current flow in the cassettes (A); 2) co-current flow in the cassettes (B)

#### Automating tank transfer control, aeration, wasting, decanting, and backwash routines.

<u>Feed Transfer</u>: The continuous transfer between "FO Draw", "MD Feed", and "Concentrate" tanks was achieved by developing feed transfer logic. The logic comprises three operational modes, "FO to MD transfer", "MD processing", and "MD to Concentrate". The FO draw gets diluted as water from the bioreactor transports through membrane. The dosing from concentrate tank maintains the salt concentration and driving force at a constant value. Once the FO draw tank level reaches the maximum set point, the contents are transferred to the MD feed tank and the MD process then reconcentrate the draw solution, produces distillate water, and recovers the salt. Concentrated MD feed then will be transferred to the concentrate tank. (Figure 94)



<u>Aeration</u>: Aeration logic was modified to control the DO based on the oxygen level in the bioreactor to replace the current On/Off time-based controlling method. With this modification, the DO level is now used to dictate the appropriate time to shut off or turn on the aeration. The logic will be improved to control the DO based on  $NH_4^+$ -N and  $NO_3^-$ -N levels.

<u>*Wasting*</u>: Wasting was also programmed to start concurrent with the backwash routine to control SRT and salinity in the bioreactor. Wasting is performed twice per day, with 4.2 L being wasted each time from the bioreactor. (Figure 94)

<u>Decanting</u>: A conductivity probe (HI3002, Hanna Instruments) was installed in the bioreactor to monitor salinity. Logic was developed to automate the decanting process. A normally closed solenoid valve was installed on the side of the bioreactor above the FO membrane module level. A normally open solenoid valve was also installed on the bioreactor feed tank to stop the feed flow to the bioreactor during decanting. The decanting process is initiated once the salinity reaches a set point (conductivity = 35 mS/cm), at which point the sump pump and aeration will shut off to allow MLSS to settle. The decanting and OMBR feed valves will be switched concurrently to drain the liquid level only to the top of the membrane module to ensure the membranes do not dry out (about 30 gal). The system returns to normal operation immediately after the decanting process is completed by switching the valves back to their original positions and turning on the sump pump (for mixing) and aeration. (Figure 95)

<u>Backwash</u>: Multiple solenoid valves, two (A & B) normally open and four (C, D, E and Distillate to BW) normally closed, were employed in the loop to enable backwash automation via defining 5 different modes for the backwash logic (Figure 95). Valve position at different backwash logic modes are summarized in Table 26.



Figure 95. Backwash, wasting, and decanting loops and valves. (Green and orange valves represent normally open and normally close solenoid valves, respectively).

Mode	Valve A	Valve B	Valve C	Valve D	Valve E	Distillate to BW
Flush	ON	ON	ON	ON	OFF	OFF
Back wash	ON	ON	OFF	ON	ON	OFF
Drain	ON	ON	ON	ON	OFF	OFF
Operation	OFF	OFF	OFF	OFF	OFF	OFF
BW Tank Refill	OFF	OFF	OFF	OFF	OFF	ON

Table 26. Backwash logic modes and the status of the solenoid valves at each mode.

#### Improving maintainability and visualization of the LabVIEW program by making it modular.

Tab controls in the front panel of the main program were introduced to the program. Figure 96 shows the screenshot of the updated front panel. Additionally, new alarms were added and existing alarms were modified for additional safety notification. The main LabVIEW program was redesigned in a modular configuration by introducing new LabVIEW sub-VIs. The main program communicates with the sub VI's using input and output queues in which the main program inserts input data into an input queue and the corresponding control sub-VI then de-queues the input. Similarly, the control sub-VI inserts its output into the output queue, then the main program de-queues the output queue and displays the output. A block diagram of the main program with sub VI's and queues is shown in Figure 97.



Figure 96. Screen shot of main screen front panel



Figure 97. Block diagram of the main OMBR-MD LabVIEW program

#### Task 8: Lab-Scale OMBR System Testing

#### FO-BIO Sub-System Testing and Data Collection

The FO-BIO sub-system has been running for more than one month after modifications. Initially, the reactor was seeded with 100 liters of activated sludge obtained from the South Truckee Meadows Water Reclamation Facility (Reno, NV). Thereafter it has continuously been fed high-strength synthetic wastewater containing ammonium carbonate and sucrose at recommended concentrations. Phosphate dosing to the reactor with monopotassium phosphate ( $KH_2PO_4$ ) was initiated to ensure phosphorus is not a limiting nutrient of biological degradation (Figure 98). To maintain a solids retention time of 30 days, which is sufficient for microbes performing nitrification, the wasting rate was set to 8.4 L/d of reactor mixed liquor.

Daily wasting also helps to regulate the salinity in the bioreactor. Regular backwashing is performed to maintain a consistent water flux and mitigate membrane fouling. Backwashing was initially carried out once a day for one hour, but was augmented to two 30-minute cycles. Since making this change, water flux measurements have been more consistent. Bioreactor conductivity is monitored regularly and used to calculate reverse salt flux (RSF) and specific reverse salt flux (SRSF) in the bioreactor (the SRSF is the RSF normalized to the water flux given in units of grams of reverse salt diffused per liter of permeate.). Water flux, RSF, and SRSF for one-month of continuous OMBR operation is presented in Figure 98.

The LabVIEW system monitors and records the volume, conductivity, and temperature of the FO draw solution and concentrate. The pH and concentrations of ammonia  $(NH_3 - N)$ , nitrate  $(NO_3^- - N)$ , nitrite  $(NO_2^- - N)$ , dissolved COD, and DO in the bioreactor are also recorded. Figure 99 shows the concentrations of chemical species in the OMBR subsystem. The aeration/anoxic cycles are controlled by aerating the bioreactor for 30 minutes followed by 30 minutes of mixing with a submerged pump to maintain biological solids in suspension. It is worth noting that the ammonia level increased when fresh synthetic feed was introduced. The ammonia spikes are likely due to higher feed concentrations, as some ammonia in the feed volatilizes over time.



Figure 98. Water flux, reverse salt flux, and specific reverse salt flux data for OMBR sub-system.



Figure 99. Performance of the OMBR sub-system and concentrations of chemical species over time.

# Task 9: OMBR System Scale-Up Testing

As part of an additional research efforts we have tested the effect of FO configuration on water flux, reverse salt flux, and membrane fouling. As part of this task, the UNR and USC groups have developed new test modules that allow us to evaluate three different membrane coupons with independent draw solutions operating in a single bioreactor environment. By having all three membranes in the same biological environment, they are exposed to the same biofouling potential. With independent draw solution loops, we are able to independently monitor flux through each membrane to evaluate how quickly each membrane is fouling.

### Bench-Scale Biofouling Analysis: Sidestream vs. Submerged Configuration

A key parameter of OMBRs is the salt concentration of the bioreactor because high concentration can inhibit microbial degradation and nutrient removal. The reverse salt diffusion of FO membranes may be influenced by the configuration of the FO membrane, but there is limited knowledge on the advantages of different FO configurations with respect to

water flux, reverse salt flux (RSF), and fouling propensity. Bench-scale experiments were conducted to investigate the optimal configuration for the OMBR.

A bench-scale FO system was used to test performance of sidestream and submerged membrane configurations. During sidestream operation both the feed and draw solutions are pumped across the membrane surfaces. During submerged operation the FO membranes are mounted on a plate and frame cassette and the draw solution is circulated through the interior channel while the module is submerged in the feed solution. The lab-scale FO-BIO subsystem is operated in the submerged configuration.

Bench-scale sidestream and submerged FO modules were tested with activated sludge from the lab-scale FO-BIO subsystem. The results indicated similar steady-state fouled membrane water fluxes for both configurations and lower SRSF for submerged operation. The bench-scale FO system was tested with activated sludge from Hyperion wastewater treatment plant (Los Angeles, CA) in both FO configurations using two draw solution concentrations; 35 and 100 g/L NaCl. Two modifications were applied to the testing method; the fouling layers were analyzed with a scanning electron microscope (SEM) equipped with an elemental X-ray energy dispersive spectrophotometer (EDS) and spacers were not used in the feed side channels during sidestream operation.

Water flux data for the submerged configuration and sidestream configurations with both draw solution concentrations is shown in Figure 100. For the 35 g/L NaCl draw solution (Figure 100 left), both configurations resulted in the same initial flux of ~9 L/m<sup>2</sup>-hr. Water flux dropped significantly, to 1.7 L/m<sup>2</sup>-hr, with the activated sludge feed in the submerged configuration due to membrane fouling. For the sidestream configuration, water flux decreased by only 10% with the activated sludge feed. After removing and cleaning the membranes, both configurations preformed with greater than 96% recovery of initial flux.

The experiment was repeated with 100 g/L NaCl draw solution (Figure 100 right). The initial flux was greater for both configurations compared to the 35 g/L draw solution, and it was higher for the submerged configuration than the sidestream. As before, when the membranes were fouled the water flux declined to  $1.7 \text{ L/m}^2$ -hr for submerged configuration. A considerable flux decline was also observed with the sidestream configuration, although the steady-state water flux was greater. As before, recovery was greater than 96 % for both configurations. However, with the 100 g/L NaCl draw solution, both configurations showed signs of significant membrane fouling.



Figure 100. Water flux data for FO membrane fouling experiments in the submerged and sidestream configurations using 35 g/L (left) and 100 g/L (right) NaCl draw solutions.

The specific reverse salt flux (SRSF), which is the grams of draw solute reverse fluxed per liter of FO permeate, for the submerged and sidestream configurations with both draw solution concentrations is shown in Figure 101. Overall, the SRSF was greater in the submerged configuration, particularly when the membrane was fouled. This high SRSF is the result of low water flux in the submerged configuration, which is partially the result of a thicker foulant cake layer, as determined by SEM.



Figure 101. Specific reverse salt flux for both FO configurations during each experimental testing stage for the 35 g/L (left) and 100 g/L (right) NaCl draw solutions.

The foulant cake layer thickness was investigated using SEM (Figure 102). With the 35 g/L NaCl draw solution, the sidestream and submerged configurations had cake layer thicknesses of  $2.7 \pm 0.7 \mu m$  and  $15.6 \pm 3.2 \mu m$ , respectively. Hydraulic scouring in the sidestream resulted in a thinner fouling layer.

When the draw solution was 100 g/L NaCl, the foulant cake thickness was similar for both configurations. The sidestream and submerged configurations had cake layer thicknesses of  $11.3 \pm 2.0 \,\mu\text{m}$  and  $14.7 \pm 2.3 \,\mu\text{m}$ , respectively. The higher draw solution concentration resulted in a greater initial water flux and brought more foulants to the membrane surface in both configurations, leading to the formation of thicker foulant cake layers.



**Figure 102.** SEM images showing foulant cake layer thickness with 35 g/L NaCl draw solution for (A) sidestream and (B) submerged configurations and using 100 g/L NaCl draw solution for (C) sidestream and (D) submerged configurations.

In the process of reverse salt diffusion, sodium and chloride partition between the foulant cake layer and the bulk feed solution. The elemental cake layer composition was obtained from EDS analysis and the weight percentages of sodium and chloride were used as indicators of reverse salt diffusion into the biofilm. Sodium and chloride were below detection limits for the new membrane, and they were 2.1 and 2.7 % by weight for the submerged and sidestream configurations, respectively, in the cake formed using 35 g/L sodium chloride draw solution (Table 27). The cake layer formed in the sidestream configuration using 100 g/L NaCl draw solution contained 5.4 wt % sodium whereas the submerged cake layer was 1.57 wt % sodium. These data indicate that sodium partitions into the cake layer more in the sidestream configuration than the submerged configuration. The implication here is that less salt is reverse fluxed into the bulk feed solution in the sidestream configuration.

	Submerged		Sidestream		
S Conc. (g/L NaCl)	35	100	35	100	
Sodium (wt %)	2.13	1.57	2.73	5.43	
Chloride (wt %)	2.03	1.85	2.05	6.62	

 Table 27. Sodium and chloride content of biofilms formed with submerged and sidestream FO configurations using 35 g/L and 100 g/L NaCl draw solutions.

Additional SEM images and EDS analysis were carried out on the cake layer formed in the sidestream configuration using 100 g/L NaCl draw solution (Figure 103). Interestingly, mineralization was observed on the surface of the cake layer in the form of sodium chloride crystals. These crystals were absent from the cake layer in the submerged configuration. Crossflow in the sidestream configuration provides an explanation of this phenomenon. Crossflow on the feed side hydraulically scours loosely adhered foulants, leaving behind particles with greater negative charges. These particles arrange themselves into thin fouling layers, as visualized by SEM, which may give rise to a dense, hydrophobic structure that facilitates sodium chloride precipitation. As a consequence, draw solutes are sequestered in the cake layer rather than the bulk feed solution.



Figure 103. SEM image of the cake layer formed in the sidestream configuration using 100 g/L NaCl draw solution (left). The presence of sodium chloride crystals  $\sim 2 \mu m$  in length was detected on the surface (right).

A summary of the performance parameters and biofilm thicknesses for all experiments is provided in Table 28. These results indicate that OMBRs should be operated with low draw solution concentration in the sidestream configuration to minimize reverse salt flux into the bulk solution. However, it is important to consider the total suspended solids concentration of the activated sludge when making a recommendation for FO configuration. The previous study was conducted with an activated sludge TSS of ~3 g/L and found that the submerged configuration had lower SRSF, whereas the current study was extracting water from activated sludge with ~7.3 g/L TSS. This distinction, and the anticipated TSS content at an FOB, will be considered before finalizing a design for scale-up. Another important distinction in the current study is the absence of feed side spacers. In the previous study there was significant fouling on the feed side spacer that may have resulted in decreased water flux and greater SRSF for the sidestream configuration.

		Submerged	S	Sidestream		
DS conc.	35	100	35	100		
Fouled Membrane	1.7	1.7	8.0	8.8		
Reverse Salt Flux	6.8	26.2	4.9	9.1		
Biofilm Thickness (µm)	15.6	14.7	2.7	11.3		
Specific Reverse Salt Flux	3.93	15.7	0.62	1.05		

**Table 28.** Water flux, reverse salt flux, biofilm thickness, and specific reverse salt flux for submerged andsidestream FO configurations with 35 g/L and 100 g/L NaCl draw solutions.

#### Bench-Scale MD Analysis: Long-Term Performance

A long-term bench-scale MD experiment was initiated on August 3, 2016 and was operated for 141 days (Figure 104). The MD feed and distillate temperatures were 65 and 39 °C, respectively. The flow rates were 1.5 L/min and the feed concentration was 35 g/L NaCl. The experiment operated continuously over the 141 days but was stopped due to failure of the feed-side pump. The distillate conductivity was relatively constant throughout the experiment, indicating that long-term MD operation is possible without membrane failure.



Figure 104. Water flux and distillate conductivity for the 141-day experiment with 35 g/L NaCl feed and 1.5 L/min flow rate in the feed and distillate streams.

There was a gradual 30% decrease in water flux over the duration of the experiment. The most likely cause of the decrease in water flux was the formation of a fouling layer on the surface of the membrane (Figure 105). The presence of a thin fouling layer became visually apparent during the initial weeks of the experiment, and this fouling layer became darker and more pronounced over the remainder of the experiment. Analysis is currently under way to determine the content of the fouling layer to determine if it was caused by biofouling or by dust that may have entered the feed. Additional analysis is underway to determine if the reduction in water flux was due to a reduction in membrane hydrophobicity caused by long-term exposure to viscous flow with high salinity and high temperature. Analyses will include AFM, confocal laser scanning microscopy, SEM, and measurement of contact angle. After the feed stream pump is repaired, a new long-term experiment will be performed with the same 65-39 °C temperature couple, the same 1.5 L/min flow rate, and a 200 g/L NaCl feed solution (increased from 35 g/L). Comparison of long-term experiments under different operating conditions will help determine the impacts of long-term exposure of MD membranes to viscous flow with high salinity and temperature on membrane hydrophobicity.



Figure 105. Fouling layer on the MD membrane after the 141-day experiment.

#### Submerged and spiral-wound FO membrane modules with pilot-scale OMBR-MD

Alternative FO and MD membranes and membrane module configurations were implemented and the integrated system was operated for an extended period. A 4040 FO membrane module that contained a spiral-wound CTA membrane element was installed and performance was evaluated. The membrane was evaluated for potential benefits in terms of water flux, reverse salt flux, and contaminant rejection. In effort to improve MD membrane performance, the tricot mesh spacers used previously were replaced with diamond-style spacers, the symmetric polytetrafluoroethylene (PTFE) MD membranes were replaced with asymmetric PTFE membranes with a polypropylene backing for support, and larger heat exchangers were installed to improve heat recovery and energy efficiency. These changes were expected to decrease the pressure drop across the MD membrane module and increase heat transfer into the feed solution; both of which would lead to higher MD water flux.

#### OMBR subsystem

The bioreactor was seeded with activated sludge collected from Cold Springs Water Reclamation Facility (Washoe County, NV). The reactor was fed a high-strength synthetic wastewater solution comprised of 1,350 mg/L COD and 160 mg/L NH<sub>4</sub><sup>+</sup>-N. To achieve nitrification and denitrification in a single reactor, a solids retention time of 30 days was set by wasting 9.2 L/d and the redox environment in the reactor was controlled by alternating aeration and non-aeration cycles. The reactor was allowed to operate under anoxic conditions for 30 minutes and then it was aerated until the DO concentration reached 2 mg/L. The microbial community was acclimated to these conditions for six weeks prior to integration with the FO membrane module.

Prior to this report, the OMBR was operated with a plate-and-frame FO membrane module that was submerged in the bioreactor. For the current reporting period, the OMBR was operated with a spiral-wound FO membrane that was mounted in a side-stream configuration externally to the bioreactor (Figure 106). The spiral-wound configuration required one additional pump to circulate the FO feed solution and a 700 µm mesh pre-filter was used to screen particulate matter from entering the module.



Figure 106. OMBR sub-system with mesh pre-filter and 4040 spiral-wound FO membrane module.

#### Water flux and water production

The FO membrane module was operated continuously for 30 days (Figure 107). For the first nine days of operation, a 20 g/L NaCl draw solution was used and the membrane was osmotically backwashed every 24 hours. Flux declined due to membrane fouling and flux recovery was observed after each backwashing cycle. FO water production decreased from 38.5 to 4.0 L/d over this period (Figure 108). On day 10, the draw solution concentration was increased to 35 g/L NaCl and FO water production increased to between 29.0 and 48.7 L/d over the course of days 10 to 15 (Figure 108). In effort to increase FO water flux and water production, the frequency of backwashing cycles was increased to every 12 hours on day 16. The twice-daily backwash cycles did result in two periods of flux recovery per day; however, net water production for the system did not increase due to flux decline from fouling and also due to the additional downtime required for the second backwash cycle. Bioreactor conductivity increased from 0.4 to 3.9 mS/cm over 30 days of OMBR operation (Figure 107), corresponding to a reverse salt flux less than 0.23 gMH. Reverse salt flux with the spiral-wound membrane was substantially lower than it was with the plate-and-frame module (4 gMH).



Figure 107. FO water flux, MD water flux, and bioreactor conductivity over 30 days of operation.



Figure 108. Daily FO and MD water production.

The MD subsystem was integrated with the OMBR sub-system on day 18. MD was operated with feed inlet and distillate inlet temperatures of 70 and 30 °C, respectively, and feed and distillate solutions were circulated at a flow rate of 2 L/min. Initial MD water flux (day 18) was ~  $3.5 \text{ L/m}^2$ -hr (Figure 110) and flux declined over time to ~ 2 L/m<sup>2</sup>-hr on day 20. Over the same period, feed solution inlet pressure increased (Figure 110), and between days 19 and 21, feed solution inlet pressure increased from 10.4 to 16.8 psi, indicating MD membrane fouling was occurring. On day 21, the MD sub-system was taken offline and a fouling mitigation strategy was developed to mitigate overpressurization in the MD membrane module. Beginning on day 22, a fouling mitigation strategy that consisted of a 30-minute flush with 0.1 M NaOH followed by a 30-minute rinse with tap water was implemented (Figure 109, black arrows). NaOH flushing on day 22 resulted in 85% recovery of initial flux and similar pressure to the initial pressure on day 18. However, after subsequent NaOH flushes on days 23 to 26, the amount of flux recovery decreased and inlet pressures continued to increase. In addition to membrane fouling, decreasing MD water flux over time may also be attributable to pore wetting, and increasing distillate concentration from 0.07 to 0.21 g/L as NaCl over time indicates some pore wetting may have occurred (Figure 109).



Figure 109. MD water flux and distillate conductivity during integrated OMBR-MD operation.



Figure 110. Feed and distillate inlet and outlet pressures during integrated OMBR-MD operation.

#### Water quality

COD removal and COD concentrations in the wastewater, bioreactor, FO draw solution tank, and MD distillate tank are shown in Figure 111. COD concentrations in the bioreactor were lower than those in the FO draw solution. COD accumulation in the FO draw solution is attributable to high rejection by MD membranes. Overall, the integrated system achieved between 98.7 and 100% COD removal. On day 18, when the MD sub-system was brought online, ammonium removal was 96.2%. Due to the closed-loop between the FO draw solution and MD feed solution, ammonium tended to accumulate in the FO draw tank and MD feed tank, leading to a decreasing trend in ammonium removal over time (Figure 112). However, a significant amount of ammonium removal was observed in the bioreactor and the integrated system achieved between 85.8 and 96.2% ammonium removal. Moreover, nitrification in the reactor, and high nitrate rejection by FO and MD membranes, resulted in distillate solution with less than 0.05 mg/L nitrate-nitrogen (Figure 112).



Figure 111. Soluble COD concentrations in the wastewater, bioreactor, FO draw solution tank, and MD distillate tank during integrated operation. Blue circles represent system removal.



Figure 112. Ammonium-nitrogen concentrations in the wastewater, bioreactor, FO draw solution tank, and MD distillate tank during integrated operation. Blue circles represent system removal.



Figure 113. Nitrate-nitrogen concentrations in the bioreactor, FO draw solution tank, and MD distillate tank during integrated operation.

#### MD membrane fouling analysis

The MD membranes were removed on day 26 and samples taken from the inlet, center, and outlet sections of the membrane module were analyzed with scanning electron microscopy (SEM) and elemental dispersive spectroscopy (EDS). Major constituents found on the feed side inlet, which are typically associated with biological fouling, include carbon, oxygen, sodium, magnesium, chloride, and calcium (Figure 114). The presence of fluorine is due to the PTFE membrane material and the presence of aluminum and silicon may be attributable to aluminosilicate, an anti-caking agent in the sucrose used to prepare the synthetic wastewater. Platinum and palladium peaks are due to the membrane coating process required for SEM analysis. Foulant concentrations were highest at the feed side inlet, lower at the middle, and lowest at the outlet, indicating that foulant deposition occurs the in a lengthwise gradient across the membrane. The SEM image and EDS spectra at the outlet are most representative of a virgin PTFE membrane.



Figure 114. SEM images and EDS spectra of used MD membranes taken from the feed side (a,b) inlet, (c,d) middle, and (e,f) outlet sections of the membrane module.

Major constituents indicative of organic membrane fouling on the distillate side of the membrane at the inlet include, carbon, oxygen, sodium, magnesium, and chloride (Fig. 115). Similar to the feed side, SEM images and EDS spectra confirm foulant concentrations were greatest at the distillate side inlet, lower at the middle, and lowest at the outlet. These results suggest that increasing feed and distillate inlet pressure over time was due to foulant accumulation at the inlets.



Figure 115. SEM images and EDS spectra of used MD membranes taken from the distillate side (a,b) inlet, (c,d) middle, and (e,f) outlet sections of the membrane module.

# **Conclusions and Recommendations for Future Research**

An integrated OMBR-MD system was designed, fabricated, and tested at the pilot-scale for wastewater treatment and production of high-quality reuse water. There were very clear synergies in using MD to reconcentrate the draw solution used in the OMBR process, in using low-grade heat/renewable energy to drive the MD process, and in recovering heat from the distillate solution to reduce the overall heat consumption of the system. The abundance of low-grade heat produced by diesel generators at FOBs make the OMBR-MD system an ideal system for such locations.

Coupling the OMBR and MD subsystems was achieved by developing an automated dosing and transfer system to maintain constant FO draw solution concentration and prevent heat from being transferred to the bioreactor, a critical design consideration for maintaining stable biological nitrogen removal. Specifically, the OMBR was coupled to DCMD using a three-tank configuration and automated controls for transferring solutions between the FO and MD processes. This allowed for continuous water production by FO and MD without transferring heat to the bioreactor, which is critical for maintaining stable biological nitrogen removal. Similar productivity for both FO and MD processes was achieved by adjusting the FO membrane area.

During membrane selection and testing, no significant differences in water flux between CTA and TFC membranes were observed when sludge feed was the feed solution. After long-term operation, FO water flux with the CTA membrane was the same for 20 g/L and 35 g/L NaCl draw solution, although bioreactor conductivity increased when 35 g/L was used due to greater RSF. This indicates that utilizing lower draw solution concentration has the benefit of lower bioreactor salinity compared to higher draw solution concentrations.

During long-term OMBR operation, carbon and nitrogen removal was achieved in a single-reactor by alternating between aerobic and anoxic bioreactor conditions. Results from long-term testing using a high-strength wastewater showed 98.4% removal of COD and 90.2% removal of NH4+-N was achieved.

Physical design aspects and process controls for system integration as well as procedures for wastewater treatment with the single-stage OMBR were established. The research results from the SERDP project were published in several peer-reviewed journal articles [refs], with more in preparation.

Several areas for improvement were identified throughout the development and testing phases of this system. Namely, passage of some dissolved, low molecular-weight contaminants into the closed-loop FO-MD solution led to MD membrane fouling and reduced MD water flux, and over time, the ammonium concentration in the distillate solution tended to increase. A suggested improved design will draw upon this and other lessons learned during the current project.

A suggested improved OMBR-MD system is shown in Figure 116. Major components are the bioreactor (TK-001) with submerged FO membrane modules (M-001) and MD membrane modules (M-002). In this system, the influent wastewater enters the bioreactor where it is biologically treated to remove organic, nitrogenous, and phosphorus contaminants. Ammonium passage into the distillate solution is mitigated by improved biological ammonium removal (i.e., nitrification/denitrification) in the OMBR. To accomplish this, it is suggested to configure the OMBR with a dual-chamber tank with separate anoxic and oxic zones. Aeration in the oxic zone would be optimized for nitrification and membrane cleaning; the anoxic zone would be optimized for denitrification. Water from the bioreactor would be drawn through the submerged FO membrane that rejects most dissolved contaminants. Once the water permeates across the FO membrane, it becomes mixed with the draw solution stream used to create the osmotic pressure gradient. This draw solution stream is circulated and, over time, becomes more dilute, triggering the need for a tank switch. Once the tank switch occurs, the diluted draw solution volume will pass to the MD loop and undergo further treatment through a fine screen (F-002), a cartridge filter (F-003), a carbon filter (F-004), and a ultraviolet (UV) reactor (F-005) to reduce the possibility of membrane fouling, membrane damage, or passage through the MD membrane by small organic compounds or disinfection residual. The MD module would operate on the same process stream but at two different temperatures. Before entering the MD module, the draw solution would be cooled by a shell and tube heat exchanger (E-001 provides a cold condensing side of the membrane to collect the distillate that crosses the membrane). Once the cold process stream exits the MD module it would then be heated by another shell-and-tube heat exchanger (E-002) to bring the process stream to a temperature where the stream's vapor pressure is sufficient to produce distillate at the desired rate. The process stream would then be returned to the MD module on the hot side of the

membrane. Inside the MD module, the process stream vaporizes across the MD membrane in a final contaminant rejection step. The resulting MD distillate is collected for potable reuse.



Figure 116. P&ID of suggested revised OMBR-MD treatment system.

In this manner, the revised OBMR-MD system operates is suggested to operate in a semi-batch configuration defined by two separate process loops. One loop operates the OMBR and the other operates multiple MD modules. Each loop operates individually until two triggering conditions are met; the draw solution reaches a target dilution in the OMBR loop and the MD feed solution reaches a target concentration in the MD loop. Once the triggering conditions are met, the solutions in each loop are exchanged; the diluted FO draw solution is then treated by MD and the concentrated MD feed solution is then diluted by FO. This semi-batch process scheme allows for independent and "continuous" operation of the FO and MD loops, reducing the complexity of the necessary control mechanisms to maintain steady-state operation of a dynamic, continuous process and thus implementing a simplified operation that maximizes the performance of each process.

Operation in the described semi-batch mode will result in each loop either diluting or concentrating the process draw stream over time. The draw and feed stream concentrations will dictate when the tanks must switch loops. For example, the OMBR loop draw solution will become more dilute over time, minimizing the driving force for permeate extraction. Once this loop becomes too dilute, the performance will be unsatisfactory, but the draw solution becomes an ideal stream for water recovery and distillation by MD. Simultaneously, a similar change will happen within the MD loop. The salt concentration of the MD loop will increase as distillate is produced and the permeation rate of the MD membrane will decrease due to the increased vapor pressure required for volatilization of distillate, but the MD feed stream will become an ideal concentrated draw stream for the OMBR loop. A dosing system will be paired with one of the tanks to re-introduce salt to the system as some will be lost during the process due to reverse salt flux within the OMBR process loop. By optimizing the dynamics between these two loops, the proposed system will meet the target production rate in the most efficient way possible.

The suggested revised system will be programmed to periodically backwash the FO membranes to extend membrane life and improve performance. The backwash cycle will utilize a portion of the distillate stored in a standalone tank. When FO performance degrades, or after a predetermined time period has elapsed, the system will cease normal operation and change to the necessary valve configuration to route the distillate to the FO module. Backwashing in a

counter-flow regime will reverse the impact of fouling and scaling that occurs over time on the FO membrane. The used backwash solution will be further treated with MD to maximize water recovery.

In the current system, the bioreactor was designed with a submerged plate-and-frame FO membrane module and an aeration baffle was placed below the membrane module to provide air scour for fouling mitigation. In the suggested revised system, a bioreactor with a baffled tank (i.e., with separate anoxic/oxic zones) is suggested for improved biological ammonium removal. Also, in the current project, submerged plate-and-frame and sidestream spiral-wound FO membrane module configurations were tested. In the suggested revised OMBR-MD system, both configurations should be considered. It is suggested to request bids for FO membranes from companies such as Porifera, Inc., Fluid Technology Solutions, Inc. (FTS), and others. Modular systems available from these manufacturers would allow for easy addition of membrane area, which will streamline assembly and maintenance procedures at fixed DoD installations. A key consideration will be to determine how the bioreactor and water recovery performance scales with membrane area.

Regarding the MD subsystem and based on testing subsequent to development of the current OMBR-MD system, it is suggested to considered use of an FO module in the revised system instead of the DCMD module that is used in the current system. In AGMD, heat recovery can be integrated into the module design. It is also suggested to solicit bids for modular MD membranes systems from: Memsys, Aquastill, and Solar Spring, who offer a variety of custom-built hybrid modules tailored to particular applications. A key consideration will be design and construction of the heating and cooling systems based on waste heat availability and desired process configurations determined.

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## Appendices

The "Instrumentation and Controls", which provide detailed information on the pilot-scale OMBR-MD system, were submitted as a standalone document. The operating manual was not included because the system itself was not a DoD deliverable item.