Membrane Bioreactors for Wastewater Treatment

Course Number: CH-02-212

PDH: 4

Approved for: AK, AL, AR, GA, IA, IL, IN, KS, KY, LA, MD, ME, MI, MN, MO, MS, MT, NC, ND, NE, NH, NJ, NM, NV, OH, OK, OR, PA, SC, SD, TN, TX, UT, VA, VT, WI, WV, and WY

New Jersey Professional Competency Approval #24GP00025600
North Carolina Approved Sponsor #S-0695
Maryland Approved Provider of Continuing Professional Competency
Indiana Continuing Education Provider #CE21800088

This document is the course text. You may review this material at your leisure before or after you purchase the course. In order to obtain credit for this course, complete the following steps:

1) Log in to My Account and purchase the course. If you don't have an account, go to New User to create an account.

2) After the course has been purchased, review the technical material and then complete the quiz at your convenience.

3) A Certificate of Completion is available once you pass the exam (70% or greater). If a passing grade is not obtained, you may take the quiz as many times as necessary until a passing grade is obtained (up to one year from the purchase date).

If you have any questions or technical difficulties, please call (508) 298-4787 or email us at admin@PDH-Pro.com.

RECLAMATION

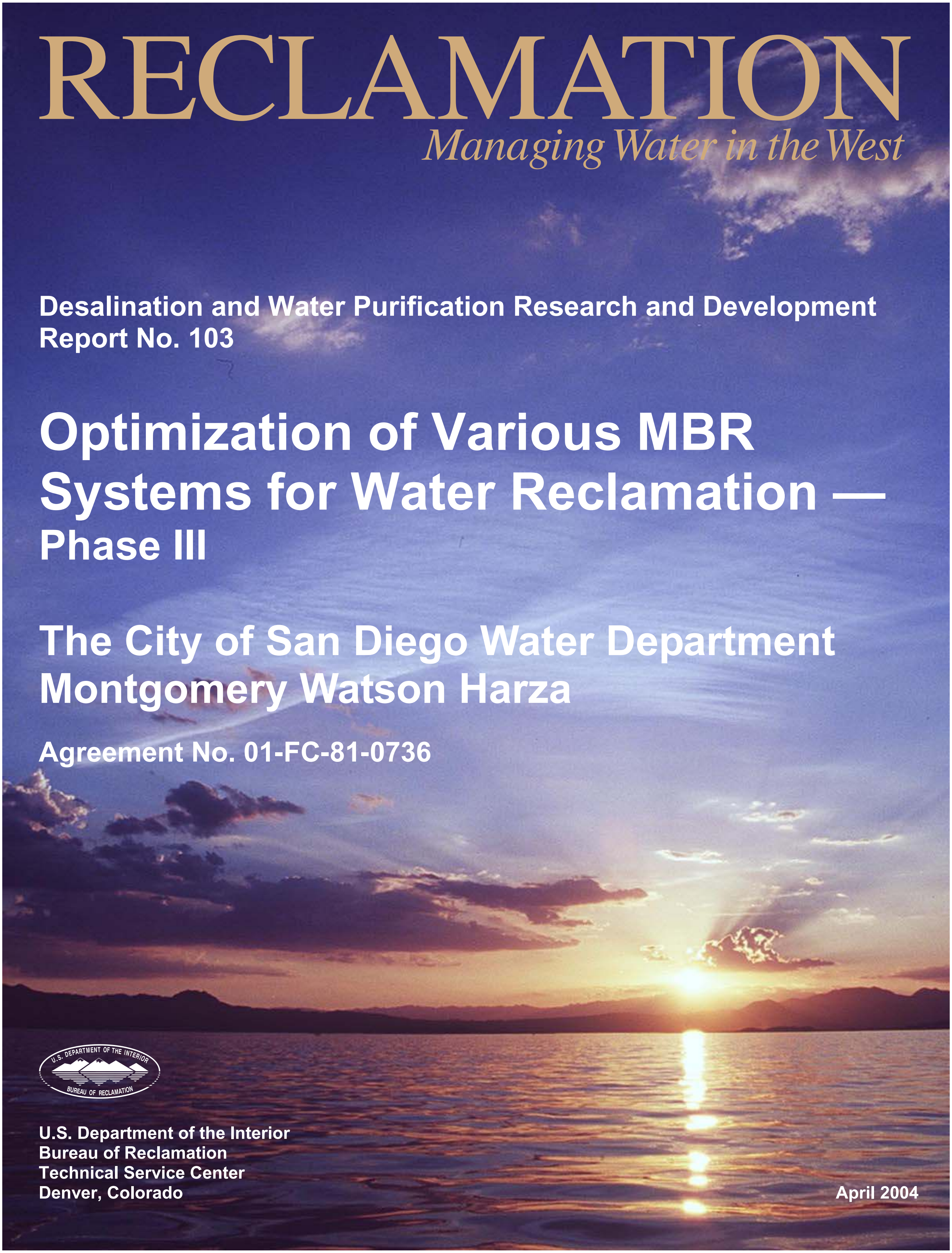
Managing Water in the West

Desalination and Water Purification Research and Development Report No. 103

Optimization of Various MBR Systems for Water Reclamation — Phase III

The City of San Diego Water Department
Montgomery Watson Harza

Agreement No. 01-FC-81-0736



U.S. Department of the Interior
Bureau of Reclamation
Technical Service Center
Denver, Colorado

April 2004

REPORT DOCUMENTATION PAGE

Form Approved
OMB No. 0704-0188

1. AGENCY USE ONLY *(Leave Blank)*

2. REPORT DATE
May 2004

3. REPORT TYPE AND DATES COVERED
Final—April 2002 to April 2004

4. TITLE AND SUBTITLE
OPTIMIZATION OF VARIOUS MBR SYSTEMS FOR WATER RECLAMATION – PHASE III
FINAL TECHNICAL REPORT

5. FUNDING NUMBERS
Agreement No.
01-FC-81-0736

6. AUTHOR(S)
Samer Adham, Ph.D.
James F. DeCarolis

7. PERFORMING ORGANIZATION NAME(S) AND ADDRESS(ES)
The City of San Diego Water Department
600 B Street Suite 700 MS 907
San Diego, CA 92101-4587

8. PERFORMING ORGANIZATION REPORT NUMBER

9. SPONSORING/MONITORING AGENCY NAME(S) AND ADDRESS(ES)
Bureau of Reclamation
Denver Federal Center
P.O. Box 25007
Denver, CO 80225-0007

10. SPONSORING/MONITORING AGENCY REPORT NUMBER
DWPR No. 103

11. SUPPLEMENTARY NOTES
Desalination and Water Purification Research and Development (DWPR) Program

12a. DISTRIBUTION/AVAILABILITY STATEMENT
Available from the National Technical Information Service, Operations Division,
5285 Port Royal Road, Springfield, Virginia 22161

12b. DISTRIBUTION CODE

13. ABSTRACT (Maximum 200 words)
This project evaluated the performance of MBR systems from four leading suppliers including US Filter Corporation/Jet Tech Products Group, Zenon Environmental Inc., Ionics/Mitsubishi Rayon Corporation, and Enviroquip Inc./Kubota Corporation during reclamation of municipal wastewater. During pilot testing, the affect of various operating conditions including feed wastewater source (e.g. raw or advanced primary effluent) flux, HRT, and SRT on MBR performance was evaluated. In addition, latest generation RO membranes from two leading suppliers (Saehan Industries Inc. and Hydranautics) were tested during operation on MBR effluent as a final polishing step for salt removal. Lastly, a cost analysis was performed to determine capital and O&M costs associated with MBR water reclamation facilities ranging in capacity from 0.2-10 MGD.

Results from this study showed that each MBR system operated successful on advanced primary effluent wastewater containing polymer and coagulant. This finding is significant as it increases the number of suppliers and feed water sources municipalities can choose to meet reclamation needs using the MBR process. In addition, each MBR achieved excellent removal of particulate (effluent turbidity <0.1 NTU), organic (effluent BOD5<2 mg/L) and microbial contaminates (total and fecal coliform LRV 4-6). In addition, valuable cost information was generated showing a significant cost savings for MBR systems designed to operate on advanced primary effluent as opposed to raw sewage.

.

14. SUBJECT TERMS—Membrane Bioreactor

15. NUMBER OF PAGES
58 in Main Report
119 in Appendices

16. PRICE CODE

17. SECURITY CLASSIFICATION OF REPORT	18. SECURITY CLASSIFICATION OF THIS PAGE	19. SECURITY CLASSIFICATION OF ABSTRACT	20. LIMITATION OF ABSTRACT
UL	UL	UL	UL

Optimization of Various MBR Systems for Water Reclamation — Phase III

Agreement No. 01-FC-81-0736

Desalination and Water Purification Research and Development Program Final Report No. 103

The City of San Diego Water Department
600 B Street, Suite 700
San Diego, CA 92101

MWH
301 N. Lake Ave, Suite 600
Pasadena, CA 91101

Samer Adham, Ph.D. (MWH)
James F. DeCarolis (MWH)
William Pearce (City of San Diego)

U.S. Department of the Interior
Bureau of Reclamation
Technical Service Center
Environmental Resources Team
Water Treatment Engineering and Research Group
Denver, Colorado

April 2004

Table of Contents

Acronyms and Abbreviations

BOD	biochemical oxygen demand
BOD_5	five-day biochemical oxygen demand
BP	back pulse
cm	centimeter
CAS	conventional activated sludge
CDHS	California Department of Health Services
CEL	Calscience Environmental Laboratory
CIL	cleaned in line
CIP	cleaned in place
COD	chemical oxygen demand
CSTR	continuously stirred tank reactor
DO	dissolved oxygen
DD	double deck
EBPR	enhanced biological phosphorus removal
EDC	endocrine disrupting compounds
ENRCCI	Engineering News-Record Construction Cost Index
F/M ratio	food to microorganism ratio
ft^2	square foot
gfd	gallons per square foot per day
g/L	grams per liter
gpd	gallons per day
gpm	gallons per minute
hr	hour
in	inch
kg	kilograms
L	liter
L/hr-m2	liters per hour per square meter
L/min	liters per minute
L/s	liters per second
mJ/cm^2	millijules per square centimeter
m^2	square meter
m^3	cubic meter
m^3/min	cubic meter per minute
m^3/day	cubic meter per day
MBR	membrane bioreactor
MF	microfiltration
mg	milligram
MGD	million gallons per day
mg/L	milligrams per liter
mg/L-N	milligrams per liter as nitrogen
min	minute

mL	milliliter
MLSS	mixed liquor suspended solids
MLVSS	mixed liquor volatile suspended solids
mm	millimeter
MPN	Most Probable Number
MWH	Montgomery Watson Harza
NaOCl	sodium hypochlorite
NCWRP	North City Water Reclamation Plant
ND	non-detect
NH_3-N	ammonia as nitrogen
NH_4Cl	ammonium chloride
NO_2-N	nitrite as nitrogen
NO_3-N	nitrate as nitrogen
NTU	Nephelometric Turbidity Units
NWRI	National Water Research Institute
O&M	operations and maintenance
OOS	out of service
PFU	plaque forming units
PLC	Programmable Logic Controller
PLWTP	Point Loma Wastewater Treatment Plant
PL Lab	Point Loma Laboratories
PO_4-P	Ortho-phosphate as phosphorus
PO_4	ortho phosphate
ppm	parts per million
psi	pounds per square inch
QA/QC	quality assurance/quality control
RAS	return activated sludge
Reclamation	Bureau of Reclamation
RO	reverse osmosis
RR	recycle ratio
s	seconds
scfm	standard cubic feet per minute
SBWRP	South Bay Water Reclamation Plant
SDI	silt density index
TDS	total dissolved solids
TFC	thin film composite
TKN	total Kjeldahl nitrogen
TOC	total organic carbon
TP	total phosphorus
TSS	total suspended solids
UF	ultrafiltration
UV_{254}	ultraviolet absorbance at 254 nanometer
UVT	ultraviolet transmittance
VFD	variable frequency drive
VSS	volatile suspended solids
WAS	waste-activated sludge

Symbols

$\Delta\pi$	Net Osmotic Pressure of the Feed and Permeate
ºC	Degrees Celsius
$K	Thousands of Dollars
µg	Microgram
µmhos	Micromhos
µm	Micron
πf	Osmotic pressure of the feedstream (psi)

Calculated Parameters

HRT	Hydraulic Retention Time (hours)
IAF	Integrated Averaging Factor
J	Membrane Flux (gfd)
J@20ºC	Temperature Corrected Membrane Flux (gfd)
J_{SP}	Specific Flux (gfd/psi)
P_{NET}	Net Operating Pressure (psi)
Q_{NET}	Net Permeate Rate (gpm)
R	Salt Rejection (%)
RR	Recycle Ratio
SRT	Sludge Retention Time (days)
$SRT_{7\text{-}day}$	Average Sludge Retention Time over 7 days
TMP	Transmembrane Pressure (psi)

1. Executive Summary

Wastewater reclamation is gaining popularity worldwide as a means of conserving natural resources used for drinking water supply. The use of membrane bioreactor (MBR) technology, which combines conventional activated sludge treatment with low pressure membrane filtration, has been proven to be a feasible and efficient method of producing reclaimed water. The membrane component of the MBR process eliminates the need for a clarifier and is performed using low-pressure membranes such as microfiltration (MF) or ultrafiltration (UF). MBR technology offers several advantages to conventional wastewater treatment including reduced footprint, consistent and superior effluent water quality and ease of operation. For many areas, it is necessary to further treat reclaimed wastewater to reduce its inherent salinity making it useable for irrigation and industrial use. The superior effluent water quality of the MBR process makes it suitable for further treatment by reverse osmosis (RO) membranes as a final polishing step in reducing the salinity of reclaimed water.

The City of San Diego and its research consultant, Montgomery Watson Harza, MWH, have been evaluating the MBR process through various research projects since 1997 (Adham et al., 1998, 2000, 2001). Previous research has primarily focused on the feasibility of using MBR technology to produce reclaimed water. In 2001, the City of San Diego was awarded a cooperative agreement by the Bureau of Reclamation to further evaluate the MBR technology for its potential application to water reclamation. The main purpose of the study was to evaluate several leading manufacturers in an effort to encourage competition within the MBR industry. In addition, the study focused on optimizing MBR operation for water reclamation. Accordingly, the project team performed a parallel comparison of four leading MBR suppliers including US Filter Corporation/Jet Tech Products Group, Zenon Environmental, Inc., Ionics/Mitsubishi Rayon Corporation, and Enviroquip Inc./Kubota Corporation.

The four MBR systems were evaluated at the pilot-scale level while operating on wastewater from the Point Loma Wastewater Treatment Plant (PLWTP) located in San Diego, CA.
Phase I testing consisted of the operation of the Kubota and US Filter MBR systems on raw wastewater for over 3,500 hours (146 days) and operation on advanced primary effluent for over 1,200 hours (50 days). During Phase II testing, the Zenon and Mitsubishi MBR systems were operated on advanced primary effluent for over 4,000 hours (187 days). As part of Phase I testing, effluent from the Kubota MBR system was further treated using reverse osmosis (RO) membranes provided by two leading RO manufacturers. The RO membranes were operated for over 1,700 hours (70 days) and 780 hours (32 days) with Kubota MBR effluent produced from raw wastewater and advanced primary effluent, respectively. The RO pilot unit consisted of two single pass trains, which were configured to allow operation at 50 percent recovery. Based on results of this testing, the project team is confident that RO membranes operating on MBR effluent could be successfully operated with a recovery between 75 percent to 90 percent which is the typical operating range for brackish groundwater.

The MBR systems tested were evaluated for their ability to produce high quality effluent and to operate with minimum fouling for a reasonable time between chemical cleanings. Furthermore, operation was optimized by evaluating performance on various types of wastewater (raw and advanced primary) and at different Hydraulic Retention Time (HRT) and flux rates. Overall the four MBR systems were capable of operating on advanced primary effluent, containing coagulant and polymer residual, with little fouling. In addition, each system successfully removed organic biochemical oxygen demand (BOD) consistently below 2 milligrams per liter [mg/L]), particulate (turbidity < 0.1 Nephelometric Turbidity Units [NTU]) and microbial contaminants (up to 6 log removal of total and fecal coliform). Also each system consistently achieved nitrification throughout the testing period with influent wastewater ammonia as nitrogen (NH_3-N) averaging 30 mg/L and MBR effluent ammonia <1 mg/L. Though it was not a goal of this study, the Kubota MBR achieved denitrification by the inclusion of an anoxic zone, which is a required portion of their system. It was also determined that it is feasible to operate the MBR processes at flux rates exceeding 20 gallons per square foot per day (gfd) and HRT as low as 2 hours. Lastly, effluent from the Kubota MBR, the only MBR tested upstream of RO, was shown to be suitable feed water for different types of RO membranes tested.

Cost estimates were developed for full-scale MBR reclamation systems ranging from 0.2-10 million gallons per day (MGD). These estimates included both capital and operational costs related to the MBR process and subsequent disinfection. The costs associated with the membrane portion of the MBR systems were obtained from the four participating MBR suppliers. All other costs including headworks, biological process and disinfection costs were estimated from preliminary design calculations performed by MWH. Results of the cost analysis ($/1000 gal) revealed that 1-MGD MBR water reclamation systems, designed to operate on raw wastewater, ranged from $1.81-$2.23. Cost estimates ($/1000 gal) for 1-million gallon (MGD) MBR water reclamation systems designed to operate on advanced primary effluent ranged from $1.57-$2.00.

2. Introduction

2.1 Background

Due to diminishing water supplies and increasing population, wastewater reclamation is becoming necessary throughout the world to conserve natural water resources used for drinking water supply. The membrane bioreactor (MBR) is a leading edge technology currently being used in countries around the world for water reclamation. Due to advances in the technology and declining costs, the application of MBR technology for water reclamation has increased sharply over the past several years. With the rapid growth of MBR technology, it is important to address the following issues: qualification of new suppliers, evaluation of operating parameters, and refinement of full-scale MBR cost estimates.

Worldwide market research has identified the following suppliers to be established manufacturers of MBR systems for the treatment of municipal and industrial wastewater: Zenon, Mitsubishi, Suez-Lyonniase-des-Eaux, Kubota and X-flow (Adham et al., 1998; van der Roest et al., 2002; WERF 2003). Of these suppliers, only two (Zenon and Mitsubishi) were established in the US market at the onset of this study. However, several new MBR manufacturers have recently entered the US market, which offer systems for the treatment of municipal wastewater. The two most prominent new comers include US Filter/Jet Tech Products Group and the Kubota Corporation. Both manufacturers provide systems, which have unique design innovations that are different from the MBR systems currently established in the US market. For example, the Kubota MBR uses flat sheet membranes rather than hollow fiber membranes for solid-liquid separation while the US Filter MBR system uses a jet aeration process to mitigate membrane fouling. Increasing the selection pool of MBR manufacturers in the US market is important because choosing the best supplier often comes down to specific site requirements and/or limitations (Wallis-Lage et al., 2003).

Some of the key operating parameters, which effect the footprint and cost of full-scale MBR systems, include SRT, HRT and membrane flux. A survey of full-scale MBR plants (Adham et al., 1998) revealed typical values for each of these parameters: SRT >30 days; flux (continuous) =15 gfd; and HRT = 20 hr. However, recent pilot studies have demonstrated MBRs could operate with limited success under more optimal conditions including: HRT 2-4 hours (Adham et al., 2000); SRT 8 days (McInnis, 2003); flux 35 gfd (van der Roest et al., 2002). Further research is necessary to determine the limitations of the design parameters to provide guidelines to the wastewater industry.

To date the application of MBR systems for municipal wastewater reclamation has focused on the treatment of two sources of wastewater: raw sewage and primary effluent. Another potential source of reclaimed water is advanced primary effluent. Advanced primary treatment differs from primary wastewater treatment in that it typically includes the addition of coagulants and/or polymers for solid and organic removal. Some of the potential benefits of operating MBR systems on advanced primary effluent, as opposed to raw sewage, includes: reduction of process air requirements due to reduced solid/organic loading, lower pre-screen maintenance requirements, and reduced foot-print. A major drawback to operation on advanced primary effluent is the potential for chemical addition (particularly organic polymers) to negatively impact the performance of the membranes. The impacts of these chemicals on MBR performance have not been previously studied.

In many water reclamation facilities across the world, it is necessary to reduce (total dissolved solids) TDS to make the water usable for irrigation and industrial applications. A recent survey sponsored by WERF (Foussereau, et al., 2002) evaluated over 100 full-scale water reclamation facilities worldwide, which use membrane technology as tertiary treatment. Results from the study revealed that over 97 percent of the surveyed plants, which required TDS and organic removal, used RO membrane technology. Furthermore, greater than 40 percent of the plants identified in the study used MBR technology for tertiary treatment with a limited number of these plants being used as pretreatment to RO.

A recent literature review identified MBR technology as the newest method of pretreatment for secondary effluent prior to RO treatment (Paranjape et al., 2003). The authors reported only two full-scale facilities in North America currently implement MBR as pretreatment to RO: City of Colony Key (0.4 MGD) and City of Laguna, Santa Maria, California (0.5 MGD). However, the ability of MBRs to produce effluent suitable as RO feedwater has been demonstrated at the pilot scale level. For instance, (Lozier et al., 1999), had moderate success operating RO on effluent produced by a Zenon MBR for indirect potable reuse. Adham et al., 2000 also demonstrated successful operation of RO on MBR effluent; however, this study was limited to one type of RO membrane (Dow/Film Tec LF/LE thin film composite [TFC]) under conservative operating conditions. Since that time, new generation thin film composite membranes have been developed and the number of suppliers has dramatically increased. As a result, further testing is necessary to increase the pool of membranes and membrane suppliers that can provide RO membranes capable of operating on MBR effluent. Due to advances in technology, it may also be feasible to operate these new generation RO membranes with lower pressure and higher production and recovery rates resulting in cost reduction of full-scale facilities.

An important responsibility of the wastewater treatment industry is to develop guidelines for estimating capital and operations and maintenance (O&M) costs associated with MBR technology. Currently a limited amount information exists in literature regarding the cost of full-scale MBR systems. For example, Adham et al., 2002 reported MBR costs were comparable to oxidation ditch and conventional activated sludge processes for 1-MGD capacities. Furthermore, when compared to various conventional treatment processes MBR was the least costly method of producing RO feed water. Churchouse and Wildgoose 1999, evaluated trends in the cost of the membrane component of MBR systems. The authors reported that the membrane replacement cost for Kubota membranes dropped by 75 percent between 1992 and 2000; decreasing the membrane portion of the overall cost from 54 percent to 9 percent. Davies et al., 1998 evaluated the economy of scale of MBR systems. The authors compared the capital costs of MBR to conventional activated sludge for installations of 2,350 and 37,5000 population equivalents. Results showed the MBR costs to be approximately 60 percent lower than conventional treatment for the smaller capacity but 46 percent higher for the larger capacity. Due to the dynamic nature of the MBR industry (including membrane development, increasing number of suppliers, increasing capacity and changing design criteria), it is imperative to periodically update MBR cost estimates.

2.2 Study Objectives

The City of San Diego was awarded a cooperative agreement by the Bureau of Reclamation to evaluate the application of MBR technology for water reclamation. The main objectives of the project were to evaluate manufacturers new to the US MBR market and optimize the application of the MBR processes for water reclamation. Accordingly, four MBR systems provided by US Filter Corporation/Jet Tech Products Group, Zenon Environmental, Inc., Ionics/Mitsubishi Rayon Corporation, and Enviroquip Inc./Kubota Corporation were evaluated at the pilot scale level over a 16-month period. During this time, performance of the MBR process was evaluated under a variety of operating conditions including feed wastewater (raw municipal wastewater and advanced primary effluent), permeate flux and HRT. In addition, the feasibility of using MBR effluent as feed water to newly developed RO membranes supplied by several leading manufacturers, including Saehan and Hydranautics, was evaluated. Lastly, cost estimates were conducted for full-scale MBR water reclamation systems ranging from 0.2-10 MGD (800-40,000 cubic meter per day [m^3/day]). The specific study objectives were to:

- Evaluate the feasibility of new MBR systems for water reclamation
- Assess the impact of coagulant and polymer addition to the MBR feed water
- Optimize the MBR process operation (pre-treatment/post-treatment)
- Evaluate the suitability of various newly developed RO membranes on MBR effluent
- Develop and refine cost estimates for full-scale MBR systems used to produce reclaimed water

3. Conclusions and Recommendations

3.1 Operational Performance

3.1.1 MBR Systems

3.1.1.1 US Filter

- The US Filter MBR system operated on both raw wastewater and advanced primary effluent with minimal increase in Transmembrane Pressure (TMP) with a flux of 15 gfd and HRT of 6 hours.

- The rate of fouling observed on the US Filter membrane, as measured by the rise in TMP, increased with flux.

- On numerous occasions during operation on raw wastewater, several components of the US Filter MBR pilot clogged with debris and hair resulting in temporary shut down.

3.1.1.2 Kubota

- The Kubota MBR system operated on both raw wastewater and advanced primary effluent with minimal increase in TMP with a flux of 15 gfd and HRT of 5 hours.

- Post cleaning, moderate foaming occurred in the aerobic tank of the system. However, normal process operation, which included the transfer of mixed liquor suspended solids (MLSS) from the aerobic zone to the anoxic zone, mitigated the foaming completely with in 1 or 2 days after cleaning.

- On two occasions during testing, the camlock fitting on the discharge side of the submersible transfer of the Kubota MBR pilot became detached due to oxidation resulting in temporary shutdown.

3.1.1.3 Zenon

- The Zenon MBR system operated with minimal increase in TMP under extreme operating conditions including flux of 22gfd and HRT of 2 hours; during this time maintenance cleans were employed three times per week to mitigate membrane fouling.

- The vacuum pressure of the Zenon membrane increased sharply when a partial loss of nitrification occurred.

- During the initial testing period, the variable frequency drive (VFD) controlling feed water flow rate to the Zenon MBR pilot failed, which resulted in unstable operation.

3.1.1.4 Mitsubishi

- The Mitsubishi MBR system operated with moderate increase in TMP with flux of 15 gfd and HRT of 2.8 hours.

- It was necessary to modify the blower system of the Mitsubishi MBR pilot system in order to maintain sufficient dissolved oxygen (DO) during operation at 15 gfd.

- Post cleaning, the biological portion of the Mitsubishi MBR was unstable, which caused a significant amount of foaming to occur in the aerobic tank.

3.1.2 RO Membranes

- The Saehan 4040 BL and Hydranautics LFC3 RO membranes operated with minimal fouling on effluent from the Kubota MBR during operation on raw wastewater and advanced primary effluent.

- The average net operating pressure of the Saehan 4040 BL (low-pressure) RO membranes measured during testing was 45 pounds per square inch (psi).

- The average net operating pressure of the Hydranautics LFC3 (fouling resistant) RO membranes measured during testing was 120 psi.

- A 1-2 mg/L dose of chloramine in the RO feed was effective at mitigating biofouling.

- The RO membranes tested achieved excellent salt rejection ranging from 96 percent-98 percent.

3.1.3 Screening

- Operational issues were experienced with the wedge-wire prescreen equipped on the US Filter MBR Pilot during operation on raw wastewater.

- Minimal maintenance was required on the rotary brush prescreen equipped on the Kubota pilot during operation on raw wastewater.

- The Roto-Sieve (RS) Model 6013-11 drum screen operated with minimal maintenance for more than 4,000 hours of operation on primary effluent.

3.2 Water Quality

3.2.1 Particulate Removal

- All MBR systems tested achieved excellent turbidity removal. Feed turbidity ranged from 36-210 NTU; average effluent turbidity of all systems was < 0.1 NTU.

3.2.2 Organic Removal

- All MBR systems tested achieved excellent organic removal with average effluent BOD, total organic carbon (TOC) and chemical oxygen demand (COD) concentrations ≤2 mg/L, ≤9 mg/L and ≤31 mg/L, respectively.

3.2.3 Biological Nutrient Removal

- All MBR systems tested achieved complete nitrification (NH_3<1 milligrams per liter as nitrogen [mg/L-N]) throughout testing.

- The Kubota MBR system achieved complete denitrification during Part 1 testing and partial denitrification during Part 2 testing.

- Partial enhanced biological phosphorus removal (EBPR) occurred in the Kubota MBR system during Part 1 testing.

3.2.4 Microbial Removal

- Total and fecal coliforms measured in the US Filter permeate ranged from 230 to 3,000 most probable number (MPN)/100 milliliter (mL) and 22 to 230 MPN/100 mL, respectively.

- Repetitive sampling of the US Filter permeate indicated that the observed coliform breakthrough may have resulted from the pore size distribution of the membranes and contamination during backwashing.

- Total and fecal coliform measured in the Kubota permeate were consistently ≤2.2 MPN/100 mL.

- Total and fecal coliforms measured in the Zenon permeate ranged from 14 to 5,000 MPN/100 mL and <2.2 MPN/100 mL, respectively; however, after disinfecting the permeate side of the membranes all total and fecal coliform measurements in the Zenon permeate were ≤2.2 MPN/100 mL.

- Total and fecal coliform measured in the Mitsubishi permeate were consistently ≤2.2 MPN/100 mL.

3.3 CDHS Approval

- As part of the California Department Health Services (CDHS) requirements, the project team conducted virus challenge experiments on the Kubota MBR system towards the end of Phase I pilot testing. These results were presented in a report to CDHS (Adham and DeCarolis, 2003).

- During this study, the Kubota and US Filter MBR systems were approved to meet the requirements of the CDHS for Title 22 Water Recycling Criteria.

- The Zenon and Mitsubishi were approved to meet the CDHS Title 22 Water Recycling Criteria based on previous testing conducted by the project team (Adham et al., 2001a & 2001b).

3.4 Costing Analysis

- Cost estimates ($/1000 gal) for 1-MGD MBR water reclamation systems designed to operate on raw wastewater ranged from $1.81-$2.24.

- Cost estimates ($/1000 gal) for 1-MGD MBR water reclamation systems designed to operate on advanced primary effluent ranged from $1.48-$1.91.

3.5 Other Conclusions

- All MBR systems tested operated successfully on advanced primary effluent containing polymer and coagulant residual.

- The MBR systems tested operated with reasonable cleaning intervals.

- The Kubota and US Filter MBR systems operated successfully on raw wastewater and were capable of producing effluent water quality suitable for RO.

- Cleaning with 2 percent citric acid was found to be the most effective method for membrane cleaning due to the presence of ferric chloride in the MBR feed water.

- O&M associated with pre-screening was significantly reduced during operation on advanced primary effluent as opposed to raw wastewater.

- MBR operational characteristics and performance varied among the four MBR suppliers.

3.6 Recommended Future Work

This project has built on previous knowledge gained by the project team with regards to the application of the MBR process for wastewater reclamation. Through pilot-scale testing, it has been demonstrated that MBR systems from 4 major suppliers can successfully operate on advanced primary effluent containing polymer and coagulant residual. This finding is significant as it increases the number of suppliers and feed water sources municipalities can choose to meet reclamation needs using the MBR process. In addition, valuable cost information was generated showing a significant cost savings for MBR systems designed to operate on advanced primary effluent as opposed to raw sewage.

Due to the increasing application of MBR technology for wastewater reuse in the United States, the project team has identified the following future research needs that can be evaluated at the pilot-scale level:

- Optimization of MBR systems to achieve high-level phosphorus removal (e.g., effluent Total Phosphorus ≤0.1 mg-P/L).

- Evaluation of anaerobic MBR systems to demonstrate process advantages including reduced energy and biomass production.

- Testing of future MBR suppliers to the US municipal wastewater treatment market including Pall Corporation, Norit, Hydranautics, Dynatec, and Huber to meet Title 22 reclaimed water standards.

- Testing of RO membranes on MBR effluent under more aggressive operating conditions such as higher flux (12-14 gfd) and recovery rates (75-90 percent).

- Evaluation and optimization of MBR systems to remove endocrine disrupting compounds (EDCs) and pharmaceuticals present in municipal wastewater.

In addition, the project team strongly recommends that the City of San Diego and Reclamation apply the knowledge gained from this and previous research studies to build and implement a 1-5 MGD MBR water reclamation demonstration facility. This demo-scale facility would provide valuable information regarding potential scale-up and reliability issues of MBR systems as applied to wastewater reclamation. In addition, the facility will provide valuable information to the City of San Diego and other municipalities across the US, regarding the operation and maintenance of full-scale MBR systems for water reclamation.

4. Materials and Methods

4.1 Testing Site

The pilot site used for this study was the Point Loma Wastewater Treatment Plant (PLWTP) located in San Diego, California. Treatment at PLWTP consists of advanced primary treatment, which includes influent screening and grit removal followed by chemical coagulation, flocculation, sedimentation and effluent screening. PLWTP currently uses between 0.5-1.0 MGD of potable water for on-site irrigation and industrial use. The City of San Diego is considering building a MBR system to reclaim wastewater onsite to meet these needs.

Pilot testing was conducted on a concrete slab located at PLWTP. The site had access to sufficient wastewater supply, electrical power, and discharge channels. Proper drainage lines were provided by the City to meet the needs of all pilot equipment. A schematic of PLWTP, showing the location of the feed water supply used for pilot testing, is provided in Figure 4-1 (See appendix A for figures)..

4.2 Feed Water Quality Characteristics

A primary objective of this study was to determine the impacts of various treatment and chemical addition processes on MBR performance. Thus, the MBR pilot systems were operated using two distinct wastewater sources: municipal raw sewage and advanced primary treated effluent. Advanced primary treatment typically includes the addition of coagulants and/or polymers for solids and nutrients removal. The impacts of these chemicals on MBR performance have not been previously studied.

Municipal raw sewage used for the study was passed through an influent screening and grit removal process; a portion of this screened sewage was diverted to the MBR systems during the early portion of the pilot study. Sewage treatment at PLWTP includes chemical coagulation/flocculation/sedimentation. Chemical addition includes ferric chloride (27 mg/L, average dose) and a long chain, high molecular weight anionic polymer1 (0.15 mg/L, average dose). A portion of the advanced primary effluent from PLWTP was diverted to the MBR pilot units during the later portion of the pilot study. Both source waters were further screened prior to MBR treatment.

4.3 Experimental Set-Up

Figures 4-2, and 4-3 are schematic diagrams of the pilot treatment trains for Phase I (Part 1 & 2) and Phase II, respectively. During Phase I testing, Kubota and US Filter were operated on raw wastewater (Part 1) and advanced primary effluent (Part 2). Phase II testing included operation of Zenon and Mitsubishi on advanced primary effluent.

4.3.1 Kubota MBR

A general process flow schematic of the Kubota MBR pilot system is provided in Figure 4-4. As shown, the operating volume of each zone within the process tank is as follows:

- Denitrification/anoxic zone = 1,695-gal (6.42 cubic meters [m^3])
- Pre-nitrification zone = 638-gal (2.41 m^3)
- Nitrification zone = 2,664-gal (10.09 m^3)

Feed water passed through a 3.2 millimeter (mm) traveling band screen before entering the feed holding tank. Next, the feed water was pumped from the feed holding tank to the denitrification zone using a submerged pump with a programmable logic controller (PLC). Water was then pumped to the pre-nitrification zone where it was aerated with fine bubble air. Mixed liquor then flowed by gravity to the nitrification zone where filtration occurred. Constant coarse bubble aeration was provided in the nitrification zone to minimize fouling. This aeration generated an upward sludge crossflow over the membrane surface of approximately 0.5 m/s. Mixed liquor overflowed back to the denitrification zone at a rate which is approximately 4 times the permeate flow rate. Lastly, sludge was wasted daily from the pre-nitrification zone to maintain a constant sludge age.

As shown at the top of Figure 4-4, the nitrification zone contained an upper and lower membrane cassette. This double deck (DD) configuration offers several benefits (van der Roest et al, 2002) including reduction of the membrane foot print, reduction of the biological volume consumed by the membrane system and reduced air consumption used for membrane cleaning. The DD also yields a more controllable biological process and reduces the possibility of short circuiting

Each membrane cassette contained 100 individual Type 510 flat membrane sheets to provide a total membrane area of 1,721 square feet (ft^2) (160 square meter [m^2]). The use of flat sheet membranes to separate activated sludge into solid and liquid is a unique feature of the Kubota MBR system. Specifications of the Kubota Type 510 flat sheet membrane are provided in Table 4-1. Photos of the Kubota pilot unit and the Type 510 flat sheet membrane are provided in Appendix D.

During Phase I (Part 1 and Part 2) pilot testing, the Kubota membrane was operated at a flux equal to 15 gfd (25 liters per hour per square meter [$L/hr\text{-}m^2$]) and a constant coarse bubble airflow of 55 standard cubic feet minutes (scfm) (1.6 cubic meter per minute [m^3/min]). During Part 1, fine air bubble airflow of 10 scfm (0.3 m^3/min) was applied, as necessary, to maintain DO in the nitrification tank at a concentration of 2.0 mg/L. The membrane was operated using a filtration cycle of 9 minutes followed by a 1 minute relaxation period. During relaxation filtration stopped; coarse bubble aeration continued. Nitrified mixed liquor was circulated at approximately 80 gpm (303 liters per minute [L/min]).

4.3.2 US Filter MBR

A general process flow schematic of the US Filter MBR system is provided in Figure 4-5. The US Filter/Jet Tec MBR pilot was equipped with a 1,000-gallon (3.79 m^3) anoxic tank (not shown), 1,500-gallon (5.7 m^3) aerobic tank, 90-gallon (0.34 m^3) membrane tank and a 163 gallon (0.6 m^3) filtrate tank. Throughout testing, the system was operated with nitrification only.

Accordingly, before start up, the system was modified to allow feed water to bypass the anoxic tank. During pilot operation, feed water was center fed through a 1.0 mm wedge wire slotted rotary screen. Screened wastewater then flowed by gravity to a feed equalization tank, which controlled flow to the aerobic tank. The equalization tank contained a submersible pump, which transferred wastewater to the aerobic tank. Level control float switches, placed in the aerobic tank, were used to turn the pump on/off. These level switches were adjusted shortly after start up to lower the average operating level of the tank in order to reduce the HRT. Because of the reduced depth in the aerobic tank, it was also necessary to modify the blower supplied on the pilot to maintain adequate DO. The mixed liquor (MLSS) in the aerobic tank was aerated using fine bubble diffusers located at the bottom of the tank. Next, MLSS was transferred to the membrane tank using a self-priming pump. A portion of the mixed liquor from the membrane tank was filtered by the membranes under a light suction while the remaining portion was overflowed/recycled back to the aerobic tank. The filtered water was then stored in a holding tank, which overflowed to waste. MLSS were wasted daily from the aerobic tank to maintain a target sludge age.

Four US Filter MemJet B10 R membranes were submerged in the membrane tank for a total membrane area of 99 ft^2 (37 m^2). Each membrane module was made of hollow fibers with a nominal pore size of 0.2 micron. During operation, air and mixed liquor were continuously injected near the bottom of the membrane tank to scrub and shake the membrane fibers. Such operation allowed for a crossflow velocity to be established on the membrane surface and minimized membrane fouling. Specifications of the US Filter MemJet B10 R membrane are provided in Table 4-1. Photos of the US Filter MBR pilot unit and the MemJet B10 R membranes are provided in Appendix D.

During the initial period of Phase I (Part 1) pilot testing, the US Filter membranes were operated at a flux of 11.5 gfd (19.2 L/hr-m2) and a constant fine bubble airflow to the aerobic tank of 25 scfm (0.7 m^3/min). For the remainder of Part 1 testing, the flux was increased to 14.5 gfd (24.2 L/hr-m2) and the fine bubble airflow to the aerobic tank was increased to 45 scfm (1.3 m^3/min).

In Part 2 testing, the US Filter membranes were operated at flux rates between 14.5–24 gfd (24.2-40 L/hr-m2) to assess the affect of increased flux on membrane performance. During both Part 1 and Part 2 testing, the coarse bubble airflow to the membrane tank was 8.5 scfm (0.24 m^3/min). Throughout the pilot testing a minimum DO level of 1.0 mg/L was targeted in the aerobic. Lastly, the MLSS overflow rate was between 14-22 gpm (0.88-1.4 liters per second [L/s]) throughout the pilot testing.

During Phase I testing, the US Filter membrane was operated using a filtration cycle of 12 minutes followed by a 1 minute backwash period. During backwashing, the membranes were allowed to relax for 45 seconds. Next, filtrate water was pumped from the inside to the outside of the membrane fibers for a 15 second period. Although forward filtration of wastewater was stopped during backwashing, coarse bubble airflow to the membrane fibers was continued. Also during backwashing, a portion of the air going to aerobic tank was diverted to the bottom (below the jets) of the membrane tank to prevent solids from accumulating.

4.3.3 Zenon MBR

A schematic of the Zenon MBR pilot unit is shown in Figure 4-6. The Zenon MBR pilot unit had a capacity of 10,000 gallons per day (gpd) (38 m^3/day). The pilot unit came equipped with a 1,300-gallon (4.92 m^3) aerobic tank and a 185-gallon (0.7 m^3) ZenoGem membrane unit. A submersible pump, placed in the primary effluent break tank, was controlled by a PLC. This pump fed the MBR. As previously mentioned, the primary effluent passed through a 0.75-mm perforated rotary drum screen before entering the aerobic zone. Activated sludge from the aerobic tank was continuously recirculated to the ZenoGem unit using a submersible pump. The overflow from the ZenoGem unit flowed back to the aerobic tank by gravity. Batch wasting was performed from the aerobic tank to maintain a constant sludge age. Fine bubble diffusers were installed in the aerobic tank to supply adequate DO to the bioreactor to maintain a minimum level of
1.0 mg/L. A photo of the diffuser grid taken before the system was seeded is provided in Appendix D.

One ZW 500d membrane cassette, containing 3 membrane elements for a total area of 720 ft^2 (69 m^2), was submerged in the ZenoGem unit. During operation, coarse air was used to scour the membranes and was cycled on/off at 10 s intervals. The 500d membrane is a reinforced hollow fiber membrane with nominal pore size of 0.04 micron. Membrane specification for the Zenon 500d membrane were obtained from the manufacturer and presented in Table 4-1. A photograph of ZW 500d is provided in Appendix D.

During Phase II pilot testing, the Zenon MBR was tested with a target flux of 22 gfd (37.3 L/hr-m2). The operation cycle was set for 10 minutes production and 30 seconds relaxation for the entire testing period. Coarse air bubble flow rate to the ZenoGem tank was set at 21 scfm (0.6 m^3/min) and the fine air bubble flow rate to the aerobic tank was set at 56 scfm (1.6 m^3/min) for the entire testing period.

As part of the optimization process, the Zenon MBR system was operated under aggressive operating conditions including high permeate flux rate (>20 gfd) and low HRT (2.0 hours). During such operation, the manufacturer recommended that maintenance cleans be performed three times per week to help mitigate membrane fouling. In accordance, maintenance cleans were performed by backpulsing chlorine (250 parts per million [ppm]) or citric acid (2 percent) to the inside of the membrane fibers to remove any build up on the membrane surface. A single maintenance clean consisted of four such back pulses with a 30 s soak time between cycles.

4.3.4 Mitsubishi MBR

A schematic of the Mitsubishi MBR system used in this study is given in Figure 4-7. The pilot unit was equipped with a 1,600-gallon (6.06 m^3) aerobic tank and 250-gallon (0.95) filtrate tank. A submersible pump, placed in the primary effluent break tank, was controlled by the programmable logic controller (PLC) and fed the MBR. As previously mentioned, the primary effluent passed through a 0.75-mm perforated screen before entering the aerobic zone. In the event of tank overflow, an overflow line from the aerobic tank was drained to waste. Sludge was batch wasted daily from the aerobic tank.

Two membrane banks were submerged in the aerobic tank, where coarse air diffusers continuously agitated the membranes and aerated the biomass. To account for the increasing oxygen demands during more aggressive conditions (i.e. low HRT), the tank was retrofitted with fine bubble diffusers. Each membrane bank consisted of 50, 1 m^2 (10.76 ft^2) Mitsubishi Sterapore HF microfiltration membranes, for a total membrane area of 100 m^2 (1,076 ft^2). The hollow fibers were arranged horizontally and attached at both ends to permeate lines. A complete list of membrane specifications is given in Table 4-1. (See Appendix A for tables.) A photograph of the membrane cassette taken before installation is provided in Appendix D.

During Phase II Testing, the Mitsubishi MBR system was operated at target flux between 11.8-14.8 gfd (20 -25 L/hr-m2). The operating cycle was set at 12 minutes production and 2 minutesrelaxation for the entire testing period. Initially, the coarse bubble air flow rate was 26 scfm (0.76 m^3/min) but was later increased to 41 scfm (1.2 m^3/min) by modifying the blower. Fine air flow of 10 scfm (0.29 m^3/min) was used during operation at flux of 14.8 gfd (25 L/hr-m2) to ensure adequate DO (e.g. $\geq$1.0 mg/L) for biological oxidation.

4.3.5 Screening Equipment

4.3.5.1 US Filter

The US Filter MBR pilot was equipped with a Contra-Shear Mini-milli Model 450M screen. This screen is a center feed rotating drum unit with 1.0 mm wedge wire slots. During operation, feed wastewater entered the system through an infeed tank assembly, which directs the liquid tangentially to the rotating drum. The screen was cleaned by external/internal spray nozzles and was enclosed by splash guards to ensure filtrate discharges below the drum. Solids removed during the screening process fall by gravity into a collection bin. The Contra-Shear was operated during Phase I of the US Filter MBR testing at a flow rate of 10 gallons/min. During Part 1 testing on raw wastewater the screening experienced several operational problems.

4.3.5.2 Kubota

The Kubota MBR pilot was equipped with an OR-TEC rotary brush screen type C. Main components of the screening system included: brush assembly, scraper assembly and perforated screen. The screen was made with #304 stainless steel and has 1/8” perforations.

4.3.5.3 Roto-Sieve

A Roto-Sieve (RS) Model 6013-11 drum screen was tested during this study. The RS 11 screen is a rotating drum screen with 0.8 mm perforation. During operation, the wastewater is fed into the drum through an inlet pipe, which distributes the water over the surface of the screen. The wastewater is then filtered through the screen and discharged at the opposite end of the feed inlet. Solids too large to pass through the screen are moved to the inlet side of the screen by the rotating motion of the drum. These particles then exit the system via a discharge collection hopper located below the feed inlet. The screen was also equipped with a counter rotating roller brush which serves to continually clean the screen to prevent clogging of the perforated slots. The brush is fixed against the outside of the drum and rotates by friction between the drum and the brush. The system was also equipped with a sprayer head and spray nozzle to provide cleaning of the sieve drum.

The Roto-Sieve screen was operated for more than 4,000 hours (167 days) during Phase II testing of the Zenon and Mitsubishi MBR systems. The flow rate to the screen was between 20-30 gpm (1.3–1.9 L/s) throughout the testing period. A photograph of the RS 11 screen is provided in Appendix D.

4.3.6 RO System

The RO pilot used in this study compared the performance of two single pass membrane trains, one provided by Saehan, the second Hydranautics. Both trains consisted of two pressure vessels configured in series. Each vessel contained 3 spiral wound 4 by 40 inch thin film composite (TFC) RO elements with a membrane surface area of 85 ft^2 (7.9 m^2) per element. In the first train, six Saehan Model RE 4040 BL membranes were tested. The RE4040 BL represents Saehan's newest generation of RO membranes designed to treat low salinity waters using low pressure. The second train consisted of six Hydranautics LFC3 RO membranes. LFC3 represents the newest generation RO membranes ideal for treatment of municipal wastewater. LFC3 is characterized as a low fouling membrane capable of achieving efficient flow and salt rejection. Specifications for the RE4040 BL and LFC3 RO membranes are provided in Table 4-2. A photograph of the RO pilot unit is provided in Appendix D.

Both RO membrane trains were operated simultaneously on identical source waters at a constant flux of 10 gfd (17 L/hr-m2) and feed water recovery of 50 percent. The source water for Part 1 and Part 2 testing was raw wastewater and advanced primary effluent treated by MBR, respectively. Per the manufacturer's recommendation, the RO influent was dosed with 2 mg/L antiscalant to slow the precipitation of sparingly soluble salts. The antiscalant used was Pre-treat Plus (King Lee Technologies, San Diego, CA). Next, feedwater passed through a 5-µm cartridge filter before being pressurized and introduced into the RO membranes. Two different methods of pre-treatment were tested during the study to mitigate biofouling of the RO membranes. During the first part of testing, RO feed water was dosed with low pressure UV. However, for the remainder of testing, chloramine was dosed to the RO feed water to maintain a 1-2 mg/L total chlorine residual concentration; no free chlorine was allowed onto the RO units.
See Section 4.3.7 for specific details of the UV pilot system.

4.3.7 UV Pilot

The Professional Line UV-system, provided by Aquionics (Erlanger, KY) was tested during this study as a pretreatment to RO. The UV pilot consisted of a disinfection chamber, single low pressure UV lamp, power supply and control panel. A photograph of the UV system is provided in Appendix D. Specifications and operating conditions for the UV system are provided in Table 4-3. During operation, the UV-Output (%) was monitored from the display panel of the system. In addition, ultraviolet transmittance (UVT) of the feed water was measured weekly and averaged 70 percent. Flow to the UV system was 14.4 gpm (0.9 L/s) throughout testing. Based on the feed water transmittance and feed flow the manufacturer estimates the effective dose to be approximately 40 millijules per square centimeter (mJ/cm^2).

4.3.8 Determination of Calculated Parameters

4.3.8.1 Pressure Calculations

The net operating pressure (P_{net}) for the RO systems was calculated according to the following equation:

$$P_{net} = \frac{(P_i - P_o)}{2} - P_p - \Delta\pi \qquad (1)$$

Where,

P_{net} = net operating pressure (psi)
P_i = pressure at the inlet of the pressure vessel (psi)
P_o = pressure at the outlet of the pressure vessel (psi)
P_p = permeate pressure
$\Delta\pi$ = net osmotic pressure of the feed and permeate (psi)

The integrated averaging factor (IAF) assuming 100 percent salt rejection can be used to estimate the osmotic pressure as follows:

$$\Delta\pi = IAF \times \pi_f$$

Where,

π_f = osmotic pressure of the feed stream (psi)
IAF = 1.386 (for 50 percent recovery)

For the RO membranes, the following approximate rule of thumb can be used:

- 1,000 mg/L NaCl solution ≈ 11.5 psi of osmotic pressure, π
- A correlation between NaCl and conductivity can be assumed (1µmho of conductivity = 1 mg/L NaCl).

The transmembrane pressure (TMP) for the Kubota MBR provided in this report is the average driving pressure required to filter water through the upper and lower membrane banks at the given flow rate plus piping resistance. The TMP of each membrane bank was calculated by subtracting the dynamic pressure measured during filtration from the static head in the membrane tank.

$$TMP = (P_{d\text{-upper}} + P_{d\text{-lower}})/2 - P_s \qquad (2)$$

Where:
$P_{d\text{-upper}}$ = Dynamic Pressure measured in the upper membrane bank at given flow rate (psi)
$P_{d\text{-lower}}$ = Dynamic Pressure measured in the lower membrane bank at given flow rate (psi)
P_s = Static Pressure Measured during Relaxation (psi)

TMP for US Filter MBR System was based on:

$$TMP = (P_{suction} - P_{permeate}) \qquad (3)$$

Where:
$P_{suction}$ = Pressure measured at point X in membrane tank on the suction side of the membrane (psi)
$P_{permeate}$ = Pressure measured at point X in the membrane tank on the permeate side of the membrane (psi)

4.3.8.2 Flow Calculations

The net permeate rate for the Mitsubishi, Zenon and Kubota MBR can be calculated using the equation:

$$Q_{Net} = \left(\frac{t_{ON} - t_{OFF}}{t_{ON}} \right) \times Q_P \qquad (4)$$

Where,
Q_{NET} = net permeate rate (gpm)
t_{ON} = the time the MBR membrane is in production (min)
t_{OFF} = the time the MBR membrane is in relaxation (min)
Q_p = Permeate flow rate (gpm)

Please note: this calculation assumes the loss of flow during cleaning in place (CIP) and intermittent maintenance cleans is negligible.

The US Filter/Jet Tech MBR employed backpulsing to minimize fouling. The net permeate rate for this system was calculated with the equation:

$$Q_{NET} = \frac{Q_P t_{ON} - V_{BP}}{t_{ON} + t_{BP}} \qquad (5)$$

Where,
V_{BP} = volume of water backpulsed (gallons)
t_{BP} = time of backpulse (min)

4.3.8.3 Flux Calculation

The flux of the RO membranes and the MBR membranes can be calculated as follows:

$$J = \frac{Q_p \times 1440}{A} \qquad (6)$$

Where,
J = Membrane flux (gfd)
A = Total membrane surface area (ft^2)

4.3.8.4 Temperature Correction

Low-pressure membrane fluxes are normally temperature corrected to 20ºC, and RO membranes are corrected to 25ºC. The membrane fluxes for the MBR membranes can be temperature corrected with the following formula:

$$J @ 20°C = J \times e^{-0.0239(T-20)} \qquad (7)$$

Where,
T = feed water temperature (ºC)

The RO membranes were temperature corrected according to manufacturer's correction factors.

4.3.8.5 Specific Flux

The specific flux is the relationship between flux and the net operating pressure. The relationship is defined by the formula:

$$J_{SP} = \frac{J}{P_{Net}} \qquad (8)$$

Where,

J_{SP} = specific flux (gfd/psi)

Likewise, the temperature-corrected specific flux can be calculated using the temperature corrected flux.

4.3.8.6 Salt Rejection

The salt rejection for the RO membranes was calculated using the following equation:

$$R = 100\left(1 - \frac{c_P}{c_f}\right) \qquad (9)$$

Where,

R = rejection (%)

c_p = permeate conductivity (μmhos)

c_f = feed conductivity (μmhos)

4.3.8.7 Hydraulic Retention Time

The hydraulic retention time (HRT) for the MBR pilot units was calculated using the formula:

$$HRT = \frac{V}{Q_{NET} \times 60} \qquad (10)$$

Where,

HRT = Hydraulic retention time (hours)

V = MBR volume (gallons)

4.3.8.8 Sludge Retention Time

The sludge retention time (SRT) is defined as the total mass of activated sludge in the MBR divided by the mass flow rate of activated sludge being removed. In order to calculate the SRT of the MBRs, the reactors are treated as an ideal continuously stirred tank reactor (CSTR). Under this assumption, concentration of activated sludge in the MBR will be the same as the concentration in the waste stream and the equation will simplify as follows:

$$SRT = \frac{VX_R}{Q_W X_W} = \frac{V}{Q_W} \qquad (11)$$

Assuming that XR is equal to X_W.

Where,

SRT	=	sludge retention time (days)
X_R	=	volatile suspended solids in the reactor (mg/L)
X_W	=	volatile suspended solids in the waste stream (mg/L)
Q_W	=	waste stream flow rate (gpd)

The seven-day SRT ($SRT_{7\text{-day}}$) is calculated by averaging the SRT over 7 previous days as follows:

$$SRT_{7-day} = \frac{SRT_{n=1} + SRT_{n=2} + \ldots + SRT_{n=7}}{7} \qquad (12)$$

Where,

$SRT_{7\text{-day}}$	=	the 7 day average SRT
N	=	day

4.3.8.9 Recycle Ratio

The recycle ratio (RR) for MBR systems operating with anoxic and aerobic tanks is defined as the ratio of the flow of MLSS from the aerobic tank to the anoxic tank, divided by the net permeate rate. The Kubota MBR was the only MBR system operated with an anoxic and aerobic tank. During operation of the Kubota MBR, MLSS was pumped from the anoxic tank to the aerobic tank and returned to the anoxic tank by gravity. Accordingly, only the flow rate from the anoxic to aerobic tank was recorded. As a result, the RR for Kubota MBR was calculated as follows:

$$RR = \frac{Q_R - Q_{NET}}{Q_{NET}} = \frac{Q_R}{Q_{NET}} - 1 \qquad (13)$$

Where,
RR = Recycle Ratio
Q_R = Flow Rate from the anoxic tank (gpm)

Because the US Filter and Zenon MBR systems were equipped with an aerobic tank and separate membrane tank, the RR was determined as the ratio of the flow rate of MLSS from the membrane tank to the aerobic tank divided by the net permeate rate. The RR for these two MBR systems were calculated as follows:

$$RR = \frac{Q_{R-membrane} - Q_{NET}}{Q_{NET}} = \frac{Q_R}{Q_{NET}} - 1 \qquad (14)$$

Where,
RR = Recycle Ratio
$Q_{R\text{-membrane}}$ = Flow Rate from the membrane tank to the aerobic tank (gpm)

4.3.9 Chemical Additions

4.3.9.1 Antiscalant Addition for RO Membranes

In order to control inorganic scaling on the RO membranes an antiscalant product was used[1]. The antiscalant was added in-line; upstream of the RO membranes at the manufacturer's recommended dosage of 2.0 ppm using a chemical-metering pump[2].

4.3.9.2 Chloramine Addition for RO Membranes

In order to control biological fouling on the RO membranes, a 1.0 mg/L chloramine residual was maintained in the MBR effluent during portions of the study. Chloramines were formed in-situ by dosing free chlorine, followed by ammonia (3.9/1 Cl_2/NH_4 ratio). The chemicals were added using chemical metering pumps[3].

4.3.10 Chemical Cleaning of Membranes

All chemical cleanings were performed in accordance to the manufacturers recommended protocol. These protocols are provided in Appendix B.

Mitsubishi and Kubota MBR systems were cleaned in-line (CIL) the presence of MLSS by introducing chemicals to the inside of the membranes through the permeate lines. Chemicals passed from the inside to the outside of the membranes by gravity.

Zenon and US Filter membranes were cleaned in place (CIP) by first transferring MLSS present in the membrane tank to the aerobic tank. This allowed for the membranes to be soaked in the direct presence of chemicals. Maintenance cleans were performed on the Zenon membranes twice per week using a 250 ppm NaOCl and once per week using 2 percent citric acid solution.

The RO membranes were cleaned using 0.1 percent sodium hydroxide. The chemical solution was mixed using RO permeate in an external cleaning skid which consisted of a 100 gallon chemical tank, a heating element and a centrifugal pump. The solution was recycled through the RO concentrate line back to the membrane cleaning tank at a rate of 4-6 gpm for 1 hour. Next, the membranes were allowed to soak for 1 hour. Finally, the cleaning solution was completely drained from the membranes and the system was brought back on-line.

---

[1] King Lee Technologies, Pretreatment Plus 0100, San Diego, CA
[2] LMI Milton Roy, Model P121, Acton, MA
[3] LMI Milton Roy, Model P121, Acton, MA

4.4 Water Quality

4.4.1 On-site water quality analyses

4.4.1.1 Temperature

The temperatures of the aerobic tank of the MBR systems were monitored using in-line temperature gages and a DO probe[4], these values were periodically field verified using an alcohol thermometer. The temperature of the RO influent was determined using an in-line temperature gauge[5].

4.4.1.2 pH

A desktop pH meter[6], was used throughout the study to determine pH of the raw wastewater, primary effluent, MBR effluent and MLSS. The meter was calibrated daily using a 3 point calibration with buffers 4, 7, and 10. The calibration was confirmed daily using a laboratory check standard.

4.4.1.3 Turbidity

The turbidity of the MBR effluents was determined using an on-line turbidimeter[7]. On-line measurements were periodically verified using a bench top turbidimeter[8].

4.4.1.4 Silt Density Index (SDI)

Silt density index (SDI) analyses were performed on the MBR effluents using a SDI machine[9]. The SDI machine filtered water through a disposable 0.45-µm filter. The SDI value was determined by periodic monitoring of the flow rate through the filter, at a constant pressure, over a 15-minute period.

---

[4] YSI Model 55, Yellow Springs, OH
[5] ReoTemp, San Diego, CA
[6] Fisher Scientific International Inc. Accumet Research AR15, Hampton NH
[7] Hach Co., Model 1720D, Loveland, CO
[8] Hach Co, Model 2100N, Loveland, CO
[9] Chemetek, FPA-2000, Portland, OR

4.4.1.5 UV-254 Absorbency

Samples collected for TOC analysis were also analyzed for UV-254 absorbency using a spectrophotometer[10].

4.4.1.6 Conductivity

On-line conductivity of the RO influent and effluent was also monitored using on-line conductivity meters[11]. Measured values were compared with daily conductivity results from the laboratory to ensure continued accuracy.

4.4.1.7 Free and Total Chlorine Residual

The total chlorine residual of RO influent was monitored using grab samples and a colorimetric test kit[12] .

4.4.2 Laboratory Water Quality Analyses

All laboratory water quality analysis were performed at one of the following locations: Point Loma Laboratory (PL Lab), the City of San Diego Water Quality Laboratory at Alvarado, Calscience Environmental Laboratories (CEL Lab) or the City of San Diego Marine Micro Lab. Table 4-4 summarizes the detection limits and methods used for all of the laboratory analyses that were performed.

4.4.3 Sampling Protocol/Frequency

All water quality samples were collected as grab samples using sample containers provided from the corresponding laboratory. All samples were transported to the lab in a cooler and were processed within the allowable holding period. During sampling, sample ports were allowed to flush before samples were collected. All microbial samples were collected using aseptic techniques. The sample ports were flamed and flushed before a sample was collected.

4.4.4 Quality Assurance/Quality Control

Appropriate measures were taken at the pilot site in order to attain the highest amount of quality control and quality assurance. Appendix C contains a technical memorandum documenting the quality assurance/quality control (QA/QC) that was performed throughout the study.

---

[10] Hach Co., DR/4000U spectrophotometer, Loveland, CO
[11] Myron L Company, Series 750
[12] Hach Co., Test Kit Model CN-80, Loveland, CO

5. Results and Discussion Phase-I: Operation of New MBR Systems

5.1 MBR Operating Conditions Phase I (Part 1)

During Phase I (Part 1) pilot testing, the US Filter and Kubota MBR systems were operated on raw wastewater from the PLWTP. The US Filter system was initially operated with an aerobic tank and membrane tank having a combined HRT of 7.6 hours at a flux of 11.5 gfd (19.8 L/hr-m2). Later, the operating level of the aerobic tank was lowered and the flux was increased to 14.5 (24.9 L/hr-m2). These changes reduced the combined HRT to 6.0 hours. Throughout Part 1 investigations, US Filter MBR was operated with an average internal recycle ratio (RR) of 6.

A mixed liquor wasting routine was implemented to allow an $SRT_{7\text{-day}}$ of 9 days and MLSS concentrations of 9–12 grams per liter (g/L). The HRT and $SRT_{7\text{-day}}$ data are presented in Figure 5-1; mixed liquor suspended solids (MLSS) and mixed liquor volatile suspended solids (MLVSS) concentrations are presented in Figure 5-2. As shown, on several occasions during Part 1 testing the MLSS measured in aerobic tank dropped below 4,000 mg/L. This was due to a glitch in the pilot system, which allowed feed water to fill the aerobic tank above the high level set point. After each occurrence, it was necessary to drain the aerobic tank back to the normal operating level before bringing the system back on line, effectively wasting the accumulated solids.

The DO measured in the aerobic tank and system air flow rates are presented in Figure 5-3. The upper graph shows the DO was consistently between 2–4 mg/L during the first 1,349 hours (56 days) of operation. Following this period, the membrane flux was increased. This resulted in a steady decrease in the DO to values < 0.5 mg/L. To avoid anoxic conditions, the flux was reduced back to 11.5 gfd (19.8 L/hr-m2). Accordingly, the DO in the aerobic tank resumed to values between 3-5 mg/L. After 2,620 hours (109 days) of operation, the blower on the system was modified to increase the fine air flow rate to the aerobic tank from 25 to 45 scfm (0.7 to 1.3 m^3/min). Following this modification, the flux was increased back to 14.5 (24.9 L/hr-m2) and the DO was maintained between 2-4 mg/L. The increase of fine air flow to the membrane tank is illustrated in the lower graph of Figure 5-3. Also shown, the coarse air flow to the membrane tank was steady at 9 scfm for the entire testing period. The DO was reduced again after 4,302 hours
(179 days) of operation when the flux was increased to values ranging from 19-24 gfd (32.2-40.7 L/hr-m2).

The Kubota MBR was operated with aerobic and anoxic tanks using Type 510 flat sheet membranes under following conditions: flux= 14.5 gfd (24.9 L/hr-m2); HRT=5.1 hours; RR= 4. The mixed liquor wasting rate was set to achieve an SRT of 11days and a MLSS concentration between 12-14 g/L. The HRT and $SRT_{7\text{-d}}$ values are presented in Figure 5-4. The DO concentrations measured in the aerobic tank are presented in Figure 5-5. As shown, DO in the aerobic tank was consistently between 1.2 and 2.4 mg/L. The MLSS and MLVSS concentrations measured in the aerobic tank and the MLVSS wasting rate are presented in Figure 5-6. The upper graph shows that after 1,670 hours (70 days) of operation, the system was drained resulting in a significant decrease in the MLSS concentration. As a result, the DO measured in the aerobic tank increased to 5.4 mg/L. However, as shown in Figure 5-5, the DO gradually decreased to target of 2 mg/L due to growth of MLSS. After this occurrence, the system was restarted and the MLSS were allowed to increase to 17.8 g/L. At that time, the daily wasting schedule was resumed to meet the target MLSS concentration of 12-14 g/L. As shown in the lower graph, the normal wasting rate required to meet the solids target was between 16-20 kilograms (kg) VSS / day.

5.2 MBR Operating Conditions Phase I (Part 2)

At the end of Part 1 testing, the feed piping to the US Filter and Kubota MBR systems were modified to supply advanced primary effluent. Next the membranes from each MBR system were cleaned in accordance to the manufacturer's protocol. The RO membranes on the Kubota MBR RO skid were also cleaned prior to beginning Part 2 testing. Due to the lower organic content of the advanced primary effluent it was necessary to establish new sludge wasting rates for each MBR system to maintain the target MLSS during Part 2 testing.

The US Filter MBR system was operated under similar operating conditions as Part 1 testing, including combined HRT = 6.0 hours; Flux = 14.5 gfd, RR = 6; fine air flow rate = 45 scfm; coarse air flow rate = 9 scfm and MLSS= 9-12 g/L. However, the mixed liquor wasting rate required to maintain the target MLSS gave an $SRT_{7\text{-day}}$ between 30-40 days. The MLSS and normal sludge wasting rate for Part 1 and Part 2 testing are presented in Figure 5-2. As shown in the lower graph, the normal sludge wasting necessary to maintain the MLSS between 9-12 g/L was much less during Part 2 (0.7-2.0 kg VSS/day) than Part 1 (4.0-6.0 kg VSS/day). Also, the DO during Part 2 was consistently measured to be between 5-7 mg/L. The decreased wasting rate necessary to maintain target MLSS and the increase in DO are both associated with the lower organic content of advanced primary effluent as compared to raw wastewater.

The Kubota MBR system was also operated under similar operating conditions as Part 1, including combined HRT = 5.1 hours; Flux = 14.5 gfd, and RR = 4. However, the target MLSS was reduced to 9-12 g/L and the mixed liquor wasting rate required to meet this goal resulted in a SRT of 18 days (9 days Part 1). As shown in Figure 5-5, the DO in the aerobic tank during Part 2 was much higher than Part 1 and ranged from 3.5 –5.5 mg/L.

5.3 Membrane Performance

5.3.1 MBR Pilot Plants

The membrane performance of the US Filter MBR during Phase I testing is presented in Figure 5-7. As indicated, after 1,182 hours (49 days) of operation, several chlorinated backwashes were employed to disinfect the permeate piping. This reduced the TMP, measured at 11.5 gfd (19.8 L/hr-m2), from 1.3 to 0.93 psi (0.09 to 0.06 bar). Over the next 1,438 hours (60 days) of operation, the TMP increased to 1.62 psi (0.11 bar). As indicated, after 2,620 hours (109 days) of operation, the permeate flux was increased from 11.5 to 14.5 gfd (19.8 to 24.9 L/hr-m2). This caused the TMP to increase from 1.62 to 2.14 psi (0.11 to 0.15 bar). After 2,954 hours (123 days) of operation the system was cleaned using chlorine, which reduced the TMP, measured at 14.5 gfd, from 3.17 to 2.04 psi (0.22 to 0.14 bar). Post cleaning, the system was operated for approximately 11 days during which time no fouling was observed. At this time, the system was cleaned again using both acid and chlorine. This cleaning reduced the TMP from 2.12 to 1.34 psi. Such results indicate that acid was more effective than chlorine in cleaning the membranes. This is expected due to the presence of ferric chloride in the raw wastewater. In the presence of alkalinity, ferric chloride undergoes a hydrolysis reaction, which forms ferric hydroxide causing a red precipitate. When discharging the spent acid solution, it was observed to have a reddish color indicating ferric chloride. Following the cleaning, the MBR was operated at 14.5 gfd (24.9 L/hr-m2) for nearly 1,000 hours (42 days), during which time the TMP increased from 1.34 to 2.7 psi. The system was then cleaned again using acid and chlorine which reduced the TMP to 0.9 psi. After the cleaning, the system was brought back on line and the flux was increased to 19-24 gfd. During operation at high flux rates the TMP increased dramatically. Such results indicate the rate of fouling observed on the US Filter membranes, as measured by rate of TMP increase, increased with increased flux. This data also suggests the critical flux of the membrane is ± 15 gfd. Lastly, during Part 2 testing on advanced primary, the US Filter MBR system was operated for 1,000 hours (42 days) at 14.5 gfd during which time minimal fouling was observed. Such results indicate the US Filter system can operate successfully on advanced primary effluent containing polymer and coagulant residual.

Membrane performance data of the Kubota MBR system measured during Phase I testing is presented in Figure 5-8. As shown in the upper graph, a sharp increase in TMP was observed during the initial 788 hour (33 days) of operation following the start up period. During this time the TMP increased from 1.38 psi (.095 bar) to 5.76 psi (0.4 bar). The manufacturer was notified and recommended the bottom membrane bank be immediately taken offline to avoid damaging the membranes. As indicated, this reduced the TMP to 2.52 psi (0.17 bar). Shortly thereafter, the manufacturer sent field technicians to the pilot site to assess the cause of the fouling. Accordingly, both membrane banks were removed from the system for observation. Visual inspection revealed that the flat sheet membranes were covered with reddish-orange precipitate, indicating the presence of ferric hydroxide. A photograph taken during the inspection is provided in Appendix D. In addition, the ¾ inch permeate line originally used on the pilot unit was replaced with 2 inch line which is used in standard design of full scale Kubota MBR systems. It is believed the ¾ inch piping may have resulted in flow restriction which increased the pressure loss on the permeate side of the membranes. The membrane cassettes were replaced and the membranes were cleaned using chlorine and acid before bringing the system back in service. As shown, following the cleaning, the Kubota MBR operated for over 2,000 hours (83 days) at a flux of 15 gfd (25.41 L/hr-m2) with a TMP between 1-3 psi (.07-0.21 bar) with little or no membrane fouling. During Part 2 testing on advanced primary, the Kubota system was operated for 1,800 hours (75 days) at 15 gfd (25.41 L/hr-m2) during which time TMP was between 1-2 psi (.07-0.14 bar) with no fouling observed. Such results indicate the Kubota MBR system can operate successfully on advanced primary effluent containing polymer and coagulant residual.

5.3.2 RO Pilot Unit

The performance of the Saehan RE 4040 BL RO membranes operating at 10 gfd (16.7 L/hr-m2) and 50 percent recovery on Kubota MBR effluent is shown in Figure 5-9. As shown, during the first 252 hours (10.5 days) the system was operated at 12.4 gfd (21 L/hr-m2). However, in order to simultaneously operate two membrane trains, it was necessary to reduce the flux to 10 gfd (16.7 L/hr-m2) due to limitations on the quantity of available feed water. During the next 500 hours of operation the net operating pressure increased from 44.6 to 57.5 psi (3.1 to 4.0 bar) indicating the membranes had fouled. At that time, the membranes were cleaned according to the manufacturer's recommendation using 0.1 percent sodium hydroxide (pH 13). The cleaning was very effective; reducing the net operating pressure to 34.8 psi at 10 gfd. A similar fouling trend was observed over the next 300 hours (12.5 days) as the net operating pressure increased to 49.0 psi (3.4 bar). The membranes were cleaned again which reduced the net operating pressure to 37.7 psi. Prior to this cleaning, the pre-filters on the RO skid were removed from the system for inspection. It was observed that the pre-filters had undergone a severe discoloration due to an excessive amount of algae growth, which occurred in the Kubota MBR permeate. A photo showing the used pre-filters and a new pre-filter is provided in Appendix D. As a result, two steps were taken to prevent the algae growth in the RO feed water: First, the clear storage tank and permeate piping of the Kubota MBR system were replaced with opaque material to block sunlight. Secondly, a dosing pump was installed to allow for the addition of 1-2 mg/L chloramine to the feed water prior to reaching the RO membranes; prior to this a low pressure UV system was used as pretreatment. After the changes, the system was cleaned and put in service at run hour 1,150. As shown, the system operated for over 818 hours (34 days) during which time the net operating pressure increased from 37.7 to 50 psi indicating chloramine addition was successful in mitigating RO membrane fouling. The membranes were then cleaned one last time prior to Part 2 testing. The net operating pressure increased from 37.2 to 46.7 psi over 700 hours (29.1 days) of operation on Kubota MBR effluent produced from advanced primary effluent.

The performance data of the Hydranautics LFC3 RO membrane operating on Kubota MBR permeate at 50 percent recovery during Phase I testing is shown in Figure 5-10. As shown during Part 1 testing, the flux was reduced to 10 gfd after 24 hours (1 day) of operation, which lowered the net operating pressure to 112.5 psi (7.8 bar). The net operating pressure remained constant for the next 539 hours (22.5 days) of operation. However, over the next 217 hours (9 days) of operation the pressure increased sharply resulting in a final net operating pressure of 188 psi (13 bar). Following the changes described above to reduce algae growth, the LFC3 operated for over 800 hours (33 days) during which time the net operating pressure only increased slightly (113 psi to 131 psi). Lastly, during Part 2 testing the LFC3 operated for over 700 hours with minimal fouling.

5.4 Water Quality

5.4.1 Raw Wastewater

The results of raw wastewater grab sample analyses conducted by the Point Loma Satellite and Alvarado Water Treatment Facility Laboratories are presented in Table 5-1. The values shown are typical of municipal wastewater.

5.4.2 Advanced Primary Effluent

The results of the advanced primary effluent wastewater grab sample analyses conducted by the Point Loma Satellite and Alvarado Water Treatment Facility Laboratories are presented in Table 5-2.

5.4.3 MBR Pilot Systems

5.4.3.1 Turbidity and Silt Density Index (SDI)

The US Filter MBR effluent on-line turbidity data is provided in Figure 5-11. During Part 1 the raw wastewater turbidity was between 58-210 NTU. The MBR effluent ranged from 0.01 to 0.12 NTU with average value of 0.03 NTU. During Part 2, the advanced primary effluent turbidity ranged from 36-130 NTU. MBR effluent ranged from 0.02 to 0.06 NTU with average value of 0.05 NTU.

The Kubota MBR effluent on-line turbidity data is provided in Figure 5-12. During Part 1, the raw wastewater turbidity was between 58-210 NTU. The Kubota MBR effluent ranged from 0.05 to 0.13 NTU with average value of 0.08 NTU. During Part 2, the advanced primary effluent turbidity raged from 36-130 NTU. MBR effluent ranged from 0.06 to 0.13 NTU with average value of 0.08 NTU. Kubota MBR SDI values measured during Phase I ranged from 0.9-1.1.

5.4.3.2 BOD_5, COD and TOC

The five-day biochemical oxygen demand (BOD_5), COD and TOC values for raw wastewater, advanced primary effluent and the US Filter MBR effluent are shown in Figure 5-13. The median value of BOD_5, COD and TOC measured in the raw wastewater was 213 mg/L, 463 mg/L and 40 mg/L, respectively. The organic content of the advanced primary effluent was significantly lower with median values of BOD_5 and COD measuring 97 mg/L and 216 mg/L, respectively. The BOD_5 of the US filter effluent was < 2 mg/L for all samples; except at 768 hours of operation when BOD_5 was measured to be 6.7 mg/L. All US Filter MBR effluent TOC samples were < 10 mg/L and the majority of COD samples measured by the Point Loma Satellite Lab were < 50 mg/L. Previous studies indicate MBR effluent COD < 20 mg/L. As a result, on several occasions COD samples were sent to a commercial lab for analysis. The results showed the average COD in US Filter MBR effluent was 21 mg/L. The discrepancy in COD results maybe due to the presence of chloride which can elevate results.

The BOD_5, COD and TOC values for raw wastewater, advanced primary effluent and the Kubota MBR permeate are shown in Figure 5-14. The BOD_5 of the Kubota MBR effluent was ≤ 2 mg/L for all samples. All Kubota MBR effluent TOC samples were < 10 mg/L and the majority of COD samples measured by the Point Loma Satellite Lab were < 55 mg/L. The average value of COD measured by Calscience Laboratories was 15 mg/L.

5.4.3.3 Biological Nutrient Removal

The inorganic nitrogen results including ammonia, nitrate/nitrite and nitrite from the raw wastewater, advanced primary effluent and US Filter MBR effluent are shown in Figure 5-15. As shown, the NH_3-N content of the raw wastewater and advanced primary effluent were essentially the same with an average value of 27 mg/L. All of the US Filter MBR effluent samples measured for NH_3-N during the study were < 2 mg/L with many values below the detection limit of 0.2 mg/L. Also, the (NO_3/NO_2)-N of MBR effluent was consistently above 20 mg/L. Such results indicate the system was completely nitrifying throughout the testing. Figure 5-17 shows Ortho-phosphate as phosphorus (PO_4-P) results for analyses conducted on the raw wastewater, advanced primary effluent and US Filter MBR effluent. As shown, the Ortho-phosphate (PO_4) content of the raw wastewater and advanced primary effluent was very low with values measuring between 0.054- 2.24 mg/L. The US Filter permeate PO_4 ranged from 0.12-0.65 mg/L. Because the US Filter system was only operating with an aerobic zone it was not possible for BPR (biological phosphorus removal) to occur.

The inorganic nitrogen results for the Kubota MBR system are shown in Figure 5-16. As shown, during Part 1, the Kubota MBR successfully removed ammonia, nitrate and nitrite to values < 1 mg/L – N. Such results indicate the system was fully nitrifying and denitrifying during this time period. However, during Part 2 testing the amount of NO_3/NO_2 in the Kubota effluent increased. For example, during Part 1 all values were < 1 mg/L but during Part 2 values ranged from 3.4 – 6.8 mg/L. Such results indicate that denitrification was decreased during operation on advanced primary effluent. This observation is believed to have resulted from excess DO in the MBR system due to the lower organic content of the advanced primary effluent. During Part 2 testing, the minimum air required for membrane scouring resulted in DO measured in the aerobic to be between 3-5 mg/L. Introduction of DO into the anoxic zone would slow down the denitrification process. Figure 5-18 shows PO-P results for analyses conducted on the Kubota MBR system. During Part 1, the majority of the feed wastewater samples ranged from 0.2 – 1.5 mg/L, while the Kubota effluent was consistently below 0.1 mg/L. These results indicate BPR was occurring in the anoxic zone of the Kubota MBR system. However, during Part 2 the PO_4 in the Kubota effluent increased to values ranging from 0.2-0.4 indicating a decline in BPR. The decrease in BPR is directly related to the partial loss of denitrification also observed during Part 2 testing. The presence of NO_3 in the anoxic tank created an anoxic environment that was not conducive to BPR.

5.4.3.4 Total Coliform, Fecal Coliform, Total Coliphage

The results of total coliform, fecal coliform and total coliphage analyses conducted on the feed wastewater and US Filter MBR effluent are presented in Figure 5-19. Initial results from Part 1 testing showed total and fecal coliform rejections (3-5 log) were obtained with total coliform permeate levels (MPN/100 ml) ranging from 230 to 3,000 and fecal coliform permeate levels (MPN/100 ml) ranging from 22 to 230. However, after 1700 hours of operation, higher total and fecal coliform rejections (4-7 log) were achieved, with total coliform permeate levels ranging from 2 to 240 MPN/100 ml and fecal coliform permeate levels below 10 MPN/100 ml. The enhanced removal may be due to pore plugging of a portion of the larger pores within the membrane pore size distribution. Lastly, the US Filter MBR obtained 3-4 log rejection of natural coliphage throughout the testing period.

Several measures were taken during the study to determine the cause of high total and fecal coliform counts measured in the US filter MBR effluent. These included: disinfecting the permeate side of the membrane, replacing the permeate sample location; taking samples at different times in the filtration cycle and taking samples just after cleaning the membranes. Overall, results showed that total and fecal counts were higher in samples taken just after a backwash and just following a membrane cleaning. A possible explanation of the results follows. First, the permeate piping became contaminated during backwashing due to the presence of algae and bacterial growth which occurred in the permeate storage tank. Second, cleaning the membranes removed the dynamic layer formed on the membrane surface, reducing the sieving ability of the membranes.

The results of total coliform, fecal coliform and total coliphage analyses conducted on the Kubota MBR system are presented in Figure 5-20. As indicated, samples were analyzed from both the upper and lower membrane cassettes. Total and fecal coliform rejections (5-7 log) were obtained with most permeate levels at or below the detection limit (2.2 MPN/100 ml). In addition, significant rejections (3-5) of total coliphage virus were also obtained by the Kubota MBR system.

5.4.3.5 Other Water Quality Parameters

The results of the US Filter MBR system analyzed by the Point Loma Satellite and Alvarado Water Treatment Facility Laboratories during Part 1 and Part 2 testing are presented in Tables 5-3 and 5-4, respectively. Laboratory results for the Kubota MBR system are presented in Tables 5-5 and 5-6, respectively.

5.4.4 RO Pilot Unit

5.4.4.1 Inorganic Nitrogen and Ortho-Phosphate Removal

The Saehan 4040 BL RO feed and permeate inorganic nitrogen species are shown in Figure 5-21. The RO permeate NH_3-N values were all below 0.3 mg/L with many values below detection; the NO_3-N values were between 0.1 and 1.9 mg/L; and the NO_2-N values ranged from 0.006 to 0.021 mg/L with many values below detection. Ortho-phosphate measured in the Saehan RO feed and permeate is shown in Figure 5-22. All PO_4 measurements in the RO permeate were below 0.03 mg-P/L with majority below the detection limit of 0.02 mg-P/L.

The Hydranautics LFC3 RO feed and permeate inorganic nitrogen species are shown in Figure 5-23. The RO permeate NH_3-N values were all below ≤ 0.3 mg/L with many values below detection; the NO_3-N values were between 0.1 and 0.8 mg/L; and the NO_2-N values ranged from 0.005-0.019 mg/L with many values below detection. Ortho-phosphate measurements in the LFC3 RO permeate are shown in Figure 5-24. All PO_4 measurements in the RO permeate were below 0.04 mg-P/L with majority below the detection limit of 0.02 mg-P/L.

5.4.4.2 TOC Removal

All TOC measurements in the effluent of the Saehan 4040 BL and Hydranautics LFC3 RO membranes were below detection of 0.5 mg/L.

5.4.4.3 Salt Rejection

The conductivity measured in the feed and Saehan RO permeate is provided in Figure 5-25. The Saehan RO membranes achieved greater than 96 percent reduction in conductivity throughout the testing.

The conductivity measured in the feed and Hydranautics LFC-3 RO permeate is provided in Figure 5-26. The LFC-3 RO membranes achieved greater than 98 percent reduction in conductivity throughout the testing.

5.4.4.4 Other Water Quality Parameters

The results of the Kubota Saehan RO samples analyzed by the Point Loma Satellite and Alvarado Water Treatment Facility Laboratories are presented in Table 5-7. Laboratory results for the Kubota Hydranautics RO pilot unit can be found in Table 5-8.

6. Results and Discussion- Phase II: Optimization of MBR Systems

6.1 MBR Operating Conditions

Upon completion of Phase I testing, the Kubota and US Filter MBR systems were decommissioned and removed from the pilot site. Next, the site was completely cleared and prepared to accommodate the Zenon and Mitsubishi MBR pilot systems. Representatives from each manufacturer came to the site to commission their MBR systems and assist the project team in preparing hydraulic and electrical connections. Past research by the project team demonstrated that the Zenon and Mitsubishi MBR systems could operate successfully on raw municipal wastewater (Adham et al., 2000). Therefore, during Phase II testing, both systems were connected to receive advanced primary effluent to assess the affect of polymer and coagulant addition on their performance. Furthermore, the project team worked closely with Zenon, the current market leader in MBR technology, to test their system under extreme operating conditions. Both systems were seeded using activated sludge from the nearby South Bay Water Reclamation Plant (SBWRP). During start up, the systems were operated without wasting to allow the MLSS to increase to target values of 10-12 g/L, the operation goal. At this time, a daily wasting routine was implemented to maintain the MLSS concentration in the aeration tanks.

Zenon. The Zenon MBR system was operated with an aerobic tank and ZenoGem tank having a combined HRT of 2 hours at a flux of 22 gfd (37.3 L/hr-m2). The ZW 500 d membrane was operated with a 10 minutesfiltration cycle, followed by a 30 s relaxation period. Maintenance cleans were performed three times per week (2/week, 250 mg/L-NaOCl, 1/week, citric acid-2 percent). A mixed liquor wasting routine was implemented to give an SRT 7-d of 18-21 days and MLSS concentration of 10–12 g/L.

HRT and SRT 7-d data are presented in Figure 6-1. The Zenon MBR concentrations of MLSS and MLVSS are presented in Figure 6-2. As indicated, after 2,300 hours of operation the MLSS concentration decreased from 10.7 to 3.3 g/L. This occurred due to failure of the wasting valve on the pilot system. The valve was repaired and the system was re-seeded with a fresh batch of MLSS of approximately 5.0 g/L from the SBWRP. As shown, the MLSS reached the target 10-12 g/L shortly thereafter. As presented in Figure 6-2, the normal sludge-wasting rate required to maintain the target solids concentration was 2-3 kg VSS/day. The course bubble aeration in the ZenoGem tank was operated at 21 scfm (0.6 m3/min) intermittently (10 s on, 10 s off). The aerobic tank air flow rate was constant at 56 scfm (1.6 m3/min). The DO measured in the aerobic tank is presented in Figure 6-3. As shown during stable operation the DO was between 0.5 to 1.5 mg/L.

Mitsubishi. The Mitsubishi MBR system was initially operated with an aerobic tank having a HRT of 3.5 hours at a flux of 11.8 gfd (20 L/hr-m2). Later, the flux was increased to 14.7 gfd (24.9), which reduced the HRT to 2.8 h. The Mitsubishi Sterapore HF membrane was operated with a 12 minutes filtration cycle, followed by a 2-minute relaxation period; air stayed on during relax. Mixed liquor was wasted daily to give an SRT 7-d of 25-37 days and MLSS concentration of 11–15 g/L. HRT and SRT 7-d data are presented in Figure 6-4. The Mitsubishi MBR concentrations of MLSS and MLVSS are presented in Figure 6-5.

As shown, the MLSS was allowed to increase from 3.6 to 12.9 g/L during start up. After 2,352 hours of operation, the MLSS were severely diluted resulting in a decrease in concentration from 10.3 to 2.2 g/L. The MLSS dilution occurred during an attempt to control foam using a spray nozzle attached near the overflow of the aeration tank. Lastly at 2,568 hours of operation the blower on the system failed and the system was shut off. During the down time, the solids in the aeration tank received no air and therefore went anoxic. A new blower was installed and the system was re-seeded and brought back online at 3,000 hours of operation.

The total and fine air flow rate provided to the aerobic tank is provided in Figure 6-6. Initially, the system was only operated with coarse air bubble aeration at a rate of 45 scfm (1.3 m^3/min). However, as the flux was increased, the DO in the aerobic tank dropped below values conducive for nitrification. Several modifications were made to the aeration system to combat the low DO levels. First, at 489 hours of operation an in-line check valve was removed from the blower which increased the air flow to 52 scfm (1.5 m^3/min). Second, after 650 hours of operation, the blower was modified which further increased the air flow to 67 scfm (1.9 m^3/min). Lastly, after 1,447 hours of operation fine bubble diffusers were added to the aeration tank. The fine bubble air flow rate was set between 10-16 scfm (0.28-0.45 m^3/min) for the remainder of the testing. The DO measured in the aerobic tank is provided in Figure 6-7. During steady periods of operation the DO was maintained around 0.5 mg/L (with coarse air only) and 1-2 mg/L (with coarse and fine air).

6.2 Membrane Performance

6.2.1 MBR Pilot Plants

Zenon. The membrane performance data from the Zenon MBR system is presented in Figure 6-8. As indicated at 674 hours of operation, the variable frequency drive (VFD) controlling the influent feed pump failed. Over the next 150 hours of operation the overall vacuum pressure increased from 1.0 (0.069 bar) to 3.3 psi (0.23 bar). This sharp increase in vacuum pressure resulted from the system being operated with no input of feed water. As filtration continued, the operating HRT decreased with the level in the aerobic tank. Eventually, the DO in the aerobic tank became insufficient for biological oxidation of organic material to occur which caused membrane fouling. At 800 hours (33 days) of operation, the membranes were cleaned using acid and chlorine. The cleaning reduced the overall vacuum pressure from 3.3 to 1.9 psi
(0.23 to 0.13 bar). During the next operational period the VFD continued to fail; resulting in unstable operation and continued membrane fouling. It should be noted that during this time period the on-site engineer worked with a representative from Zenon to change the settings on the VFD in hopes of correcting the problem. It was finally decided to remove the VFD from the

system and operate using a constant feed pump, controlled by level sensors placed in the aerobic tank. At this time, a representative from Zenon came to the site to make appropriate changes to the system to allow such operation.

A membrane cleaning was performed again at 1,350 hours (56 days) of operation which reduced the overall vacuum pressure from 7.5 to 2.3 psi (0.52 to 0.16 bar). The system was then placed in operation at a flux of 22 gfd (37.3 L/hr-m2) and HRT of 2 hours. As shown, the overall vacuum pressure remained near 2.5 psi (0.17 bar) for nearly 550 hours (23 days) of operation. However, at 1,850 hours of operation nitrification was partially lost; which caused the overall vacuum pressure to quickly increase from 2.5 to 9.5 psi (0.17 to 0.66 bar). Based on discussions with Zenon, it was determined that the decrease in nitrification and subsequent membrane fouling may have resulted from a low, operating food to microorganism ratio (F/M ratio). The manufacturer recommended increasing the MLSS in the aeration to approximately 11 g/L to maintain F/M ratio < 0.4 day -1. The system was cleaned again at time of operation 1,950 hours and placed back into operation. As previously explained the mixed liquor was diluted at 2,350 hours of operation. The system was cleaned again after 2,450 hours of operation which reduced the overall vacuum pressure to 1.8 psi (0.12 bar). Post cleaning, the system was brought back on line at flux of 17 gfd
(28.8 L/hr-m2). At 2,639 hours of operation the flux was increased to 22 gfd (37.3 L/hr-m2) and HRT of 2 h. The 500 d membrane was operated for over 1,800 hours (75 days) at these conditions during which time the overall vacuum pressure only increased from 1.3 (0.09 bar) to 3.0 psi (0.21 bar).

Mitsubishi. The membrane performance of the Mitsubishi MBR system is shown in Figure 6-9. As indicated the flux was increased to 11.8 gfd (20 L/hr-m2) after 128 hours (5.3 days) of operation. The system was operated for over 1,500 hours (62 days) during which time the overall vacuum pressure increased from 0.71 (0.05 bar) psi to 2.0 psi (0.14 bar). As part of optimizing the Mitsubishi MBR system, the flux was increased to 14.8 gfd (25.1 L/hr-m2) after 1,700 hours (70.8 days) of operation. Over the next 150 hours (6.3 days) of operation the overall vacuum pressure was stable at 2.5 psi and the DO in the system remained above 0.5 mg/L. As indicated, the membranes were cleaned after 1,850 hours (77 days) of operation. The cleaning reduced the vacuum pressure from 2.6 (0.18 bar) to 1.3 psi (0.09 bar). Post cleaning, the system was brought back on-line at the target flux of 14.8 gfd (25.1 L/hr-m2). However, as shown in the upper graph, after 1,990 hours of operation it was necessary to decrease the operating flux because of excessive foaming in the aerobic tank. The foaming is believed to have resulted from the membrane cleaning because chlorine was brought into direct contact with the MLSS. As shown during the next 724 hours (30 days) of operation foaming continued and resulted in unstable operation of the MBR system. During this time period, several tactics were employed to mitigate foaming and resume stable operation. For example, a sprayer system was constructed and installed on the pilot system. The sprayer system consisted of ½ inch tubing which surrounds the perimeter and passes across the center of the aeration tank. Spray nozzles which produce a fine mist were placed every 6 inches along the tubing. The sprayer system was operated using a timer which can be set to turn on/off up to 6 times per day.

The Mitsubishi MBR system was shut down after 2,568 hours (107 days) operation when the blower failed. A new blower was immediately ordered and installed within 2 weeks of the

occurrence. The aerobic tank was drained, flushed with potable water and reseeded prior to start up because the MLSS was without aeration for several weeks. The system was brought back on line after 3,000 hours of operation at a flux of 11.8 gfd (20 L/hr-m2) and the MLSS was allowed to reach
14 g/L. At this time the flux was increased to 14.8 gfd (25.1 L/hr-m2) and the Mitsubishi Sterapore HF membrane was operated for over 800 hours (30 days) during which time vacuum pressure increased from 2.13 to 3.27 psi.

6.3 Water Quality

6.3.1 Advanced Primary Effluent

The results of advanced primary effluent wastewater grab sample analyses conducted by the Point Loma Satellite and Alvarado Water Treatment Facility Laboratories are presented in Table 6-1.

6.3.2 MBR Pilot Systems

6.3.2.1 Turbidity

The Zenon MBR effluent on-line turbidity data is provided in Figure 6-10. As shown, the advanced primary effluent turbidity measured during Phase II testing ranged from 23-63 NTU. During the entire testing period, the Zenon MBR effluent turbidity ranged from 0.03 to 0.1 NTU with average value of 0.06 NTU. As shown, at 2700 hours of operation the MBR effluent turbidity decreased from 0.06 to 0.04 NTU after the turbidimeter was cleaned.

The Mitsubishi MBR effluent on-line turbidity data is provided in Figure 6-11. During the entire testing period, the Mitsubishi MBR effluent turbidity ranged from 0.04 to 0.10 NTU with average value of 0.07 NTU. As shown, the turbidimeter cleaning performed at 3,772 hours of operation reduced the turbidity from 0.08 NTU to 0.05 NTU.

6.3.2.2 BOD_5, COD and TOC

The BOD_5, COD and TOC values for advanced primary effluent and the Zenon MBR effluent are shown in Figure 6-12. The median value of BOD_5, COD and TOC measured in the advanced primary effluent during Phase II testing was 112 mg/L, 237 mg/L and 44 mg/L, respectively. The BOD_5 in the Zenon MBR effluent was below the detection limit of 2 mg/L in all samples. Zenon MBR effluent TOC samples were all < 9 mg/L and all COD samples < 28 mg/L.

The BOD_5, COD and TOC values for the advanced primary effluent and the Mitsubishi MBR effluent are shown in Figure 6-13. The BOD_5 in the Mitsubishi MBR effluent samples were all < 2 mg/L. All Mitsubishi MBR effluent TOC samples were < 10 mg/L and the COD ranged from 18-31 mg/L with median value of 21 mg/L.

6.3.2.3 Biological Nutrient Removal

The inorganic nitrogen results including ammonia, nitrate/nitrite and nitrite from the Zenon MBR effluent are shown in Figure 6-14. All of the Zenon MBR samples measured for NH_3-N were < 2 mg/L except the sample taken at 1,872 hours of operation which measured 5.0 mg/L. At this time, the MBR system was being operated with F/M > 0.4 day-1 which is above the manufacturer's recommendation. Following this event, the F/M was decreased by increasing the MLSS concentration and nitrification was resumed. As shown in the middle graph, the (NO_3/NO_2)-N of MBR effluent was consistently above 18 mg/L indicating complete nitrification. Figure 6-15 shows PO_4-P results for analyses conducted on the advanced primary effluent and Zenon MBR effluent. As shown, the PO_4 content of the advanced primary effluent was between 0.035- 1.23 mg/L and Zenon MBR effluent ranged from 0.3 to 1.24 mg/L. BPR will not occur in MBR systems operating with only an aerobic tank.

The inorganic nitrogen results for the Mitsubishi MBR system are shown in Figure 6-16. The high levels of ammonia (>5 mg/L) present in the Mitsubishi permeate during the initial 600 hours of operation indicate the system was not achieving complete nitrification. Such data is expected as the seed sludge was growing during this time period and the nitrifying bacteria are relatively slow growers. However, after 850 hours of operation all ammonia samples from the Mitsubishi permeate were below 1.0 mg/L as N. The achievement of nitrification can also be seen in the plot of nitrite/nitrate which shows a trend of increasing NO_2/NO_3 concentration in the Mitsubishi permeate with an increase in time of operation. Figure 6-17 shows PO_4-P results for analyses conducted on the advanced primary effluent and Mitsubishi effluent.

6.3.2.4 Total Coliform, Fecal Coliform, Total Coliphage

The results of total coliform, fecal coliform and total coliphage analyses conducted on the feed wastewater and Zenon MBR effluent are presented in Figure 6-18. As shown, the Zenon system achieved total and fecal coliform rejections ranging from 3-7 log. The total coliform measured in the Zenon permeate were quite high ranging from 14 to 5,000 MPN/100 mL while the fecal coliform were consistently below the detection limit of 2.2 MPN/100 mL. Also shown, the Zenon MBR system achieved total coliphage rejections (4.0-5.5 log) with all values in the permeate at or below the detection limit of 1.0 plaque forming units (PFU)/100 mL.

The fact that the Zenon permeate showed high total coliform counts, despite low counts of fecal coliform and total coliphage suggested the presence of contamination on the permeate side of the membrane. Accordingly, the entire permeate piping system was disinfected after 2,900 hours of operation. As shown, all post disinfection total and fecal coliform measurements in the Zenon MBR permeate were ≤2.2 MPN/100 ml.

Figure 6-19 presents the total and fecal coliform and total coliphage in the influent and effluent of the Mitsubishi measured during Phase II testing. As shown, the Mitsubishi system achieved excellent rejections (5.5-7.0 log) of total and fecal coliform with permeate levels consistently below the detection limit of 2.2 MPN/100 mL. Such data indicates the Mitsubishi MBR is an excellent barrier to bacteria present in the feed wastewater.

Also shown, the Mitsubishi system achieved between 3-5 log rejection of total coliphage with many measurements in the permeate below the detection level of 1.0 PFU/100 mL. The data collected during the first 1,100 hours of operation, clearly shows a trend of decreased permeate coliphage with increasing time of operation. This would be expected for two reasons. First, the amount of total coliphage absorbed to the MLSS increases with increased solids concentration. Secondly, as the membranes become clogged the pore size is decreased which results in removal of virus and other particles which could normally pass through the membrane. It should be noted new membranes were installed on the pilot system before beginning the study.

6.3.2.5 Other Water Quality Parameters

The results of water quality analysis conducted on Zenon MBR effluent by the Point Loma Satellite and Alvarado Water Treatment Facility Laboratories is presented in Tables 6-2. Likewise, laboratory results for the Mitsubishi MBR system from Phase II testing are presented in Table 6-3.

7. Title 22 Approval of MBR Systems

7.1 Zenon and Mitsubishi

In March 2000, the project team met with the CDHS to develop a specific testing protocol for the approving MBR systems as an acceptable filtration technology for compliance with the State of California's Water Recycling Criteria (Title 22). It was decided approval would be based on the systems ability to meet the following criteria:

- Turbidity performance (not to exceed 0.2 NTU more than 5 percent of the time within 24-hour period; and 0.5 NTU anytime)
- Long Term Operational Data (approval to be based on flux and vacuum pressure range)
- Approval to be membrane specific
- Demonstrate ability of the system to achieve 1-log virus reduction at the 50th percentile

Shortly after these criteria were established, the project team performed long term testing on the Zenon and Mitsubishi MBR systems under grant funding from the Reclamation (Adham et al., 2000). Following this testing, the project team conducted virus challenge studies on these systems through funding provided by the National Water Research Institute. Based on the results from these two research projects, both the Zenon and Mitsubishi MBR systems received Title 22 approval in April 2001 (Adham et al., 2001 a and b).

7.2 Kubota and US Filter

At the end of Phase I pilot testing of the current study, representatives from Enviroquip Inc./Kubota Corporation expressed interest in obtaining regulatory approval for the use of the Kubota MBR to meet California's Title 22 Water Recycling Criteria. Accordingly, the project team conducted additional testing on the Kubota MBR system at PLWTP to meet the requirements established by the CDHS. The project team prepared a report summarizing the results of the virus challenge experiments and operational performance data collected from the evaluation of the Kubota MBR system pilot system at PLWTP. This report was submitted to the CDHS in February 2003 (Adham and DeCarolis, 2003). In March 2003, the CDHS sent an approval letter to Kubota stating their acceptance of the Kubota Type 510 flat sheet membrane to meet Title 22 water recycling criteria. A copy of the approval letter is presented in Appendix E.

Also during the current study, representatives from the US Filter Corporation/Jet Tech Products Group contacted the project team and the CDHS regarding Title 22 approval requirements for MBR systems. After reviewing MBR operational data collected from Point Loma, the CDHS accepted the MemJet B10 R membrane to meet the Title 22 water recycling criteria. The virus rejection data of the MemJet B10 R membrane was collected during CDHS approval testing of the membrane for drinking water applications conducted at the Aqua 2000 research center. (Adham and Gramith, 2001).

8. MBR Performance Comparison

8.1 MBR Operating Experience

The four MBR pilot systems were all completely automated, however; a varying degree of operator attention was required for each pilot system. The following summarizes operational experiences with each system.

8.1.1 US Filter MBR System

On numerous occasions during operation on raw wastewater several components of the US filter MBR pilot were clogged with debris and hair which ultimately caused the system to enter into "alarm mode" and shut down. In particular, clogging occurred in the following areas: pre-screen, piping from the aerobic tank to the membrane tank and the rotameter located before the membrane tank. Another operational problem experienced with the US Filter MBR system was the level control system equipped in the aerobic tank. On several occasions during testing the tank overflowed causing the MLSS to be severely diluted. This made it difficult to maintain a steady SRT. It was later discovered that the wiring for the high/low level switched was reversed. As a result, when the aerobic tank level reached a high level the feed pump would continue as if the level was low. These reoccurring incidences throughout the testing made it necessary to provide a significant amount of operator attention to keep the US Filter MBR operating at steady state.

8.1.2 Kubota MBR System

On two occasions during testing, the stainless steal camlock fitting on the discharge side of the submersible transfer pump located in anoxic zone deteriorated and became detached. The transfer pump is used to transfer wastewater from the anoxic tank to the aerobic tank where membrane filtration occurs. Once the camlock became detached, the membrane tank received no further input of feed wastewater and therefore the level dropped as filtration continued. Because the transfer pump was submerged, it was necessary to use a fork lift to remove the pump from the system in order to replace the camlock fitting. Also, as mentioned in Section 5.3, the pilot system was originally equipped with ¾" permeate piping lines. This appears to have caused an increase in pressure loss associated with piping loss. As a result, the system was taken offline. During this time it was necessary to remove the membrane cassettes using a crane and replace perm piping with 2" line. Once these modifications were made, the Kubota MBR pilot operated smoothly with little operator attention.

8.1.3 Zenon MBR System

During the initial period of Phase II testing, the VFD controlling the feed water flow rate to the aerobic tank failed. The VFD was removed from the system and replaced with level control switches. From this point forward the Zenon MBR required minimal operator attention.

8.1.4 Mitsubishi MBR System

Upon increasing the flux rate it was necessary to modify the blower system equipped on the Mitsubishi pilot to provide adequate DO for biological oxidation. Also, after the first membrane cleaning event the biological system was unstable causing a significant amount of foam to form in the aerobic tank. During periods of foaming the Mitsubishi MBR system required a lot of operator attention to prevent MLSS from spilling over the top of the aeration tank. This included building and installing a sprayer system to help control foaming.

8.2 Operating Conditions

8.2.1 Flux, HRT, SRT and MLSS

The US Filter and Kubota MBR systems were operated with flux and HRT values typical of full scale MBR processes. These include flux of 15 gfd and HRT ranging from 4-8 h. However, the Zenon and Mitsubishi systems were operated under more extreme operating conditions in effort to optimize the MBR process for water reclamation. For example, the flux of the Zenon and Mitsubishi MBRs systems was increased and sustained at 22 gfd and 15 gfd, respectively. Such flux values exceed the manufacturers recommended membrane flux. The HRT of the Zenon and Mitsubishi systems were ultimately reduced to 2 hours and 2.8 h, respectfully. All four systems were operated with typical SRT (11-20 days) and MLSS concentrations (9-14 g/L) used in full scale MBR processes.

8.2.2 Frequent Relaxation/Backpulsing

The Kubota, Mitsubishi and Zenon MBR systems relaxed during operation to prevent membrane fouling while the US Filter system used backwashing. The frequency and duration of relaxations ranged from 9-12 minutes and 0.5– 2 minutes, respectively. The backwash frequency and duration of the US Filter MBR was 12 minutes and 1 minute, respectively. The use of relaxation, as opposed to backpulsing, eliminates the need for additional permeate storage tanks and/or valves and piping. In addition, total coliform results collected from the US filter MBR system also suggest that backpulsing can introduce contamination into the permeate piping due to algae growth in the permeate tank. Lastly, as reported by Adham et al., 1998, membrane integrity is another factor to consider in systems that use backpulsing. The authors explained that during filtration the applied vacuum pressure typically causes the solids to clog broken fibers. However, on systems that backpulse these seals can become broken over time which makes it necessary to replace the compromised fiber(s).

8.2.3 Air Usage (Membrane Scour and Biological Requirements)

Each MBR system used coarse bubble aeration to reduce membrane fouling. Membrane airflow rates per membrane area (scfm/ft^2) for the US Filter, Kubota, Zenon and Mitsubishi MBR systems were 0.023, 0.033, 0.030, and 0.028, respectively. The Zenon MBR system was the only system operated with intermittent coarse air, which reduced the total air usage by 50 percent. Each system also used fine bubble aeration to provide sufficient DO to the activated sludge. The fine air to the Kubota MBR was applied intermittently as necessary to maintain 2 mg/L DO in the aerobic tank. All other MBR systems tested were operated with constant fine bubble aeration.

8.2.4 Membrane Cleaning

The cleaning procedures for all four MBR systems used chlorine (2-3 g/L) followed by citric or oxalic acid (2 percent). However, the Zenon and US Filter membranes were cleaned in place (CIP) while the Mitsubishi and Kubota membranes were cleaned in-line (CIL). During a CIP, the membranes were isolated from the MLSS and chemicals were recirculated through the membranes prior to soaking. During a CIL, the membranes were not isolated from the MLSS and chemical was allowed to slowly flow by gravity from the inside to the outside of the membranes. This procedure introduces chemicals into direct contact with activated sludge. Such contact caused a significant amount of foaming to occur in the Kubota and Mitsubishi MBR systems during post cleaning operation. For the Kubota MBR system, which transferred MLSS from anoxic to aerobic zone foaming was mitigated within 1 or 2 days following a cleaning event with no added foam control. However in the case of Mitsubishi, which was only operated with an aerobic tank, foaming persisted for several weeks after cleaning. As a result, it was necessary to install a sprayer system for foam control. Other than foaming issues, both methods of cleaning were effective at reducing vacuum pressure. Furthermore, acid was the most effective cleaning chemical for all four systems.

8.3 Membrane Performance

All four MBR systems demonstrated good membrane performance throughout the testing. The Kubota and US filter membranes required minimal cleaning during operation on both raw wastewater (Part 1) and advanced primary effluent (Part 2). In Part 1 testing, the Kubota MBR was only cleaned once after 788 hours (33 days). Post cleaning the membrane was operated for over 2,000 hours (83 days) at 15 gfd without cleaning. During Part 2, no cleanings were performed and the membrane operated for over 1,800 hours (75 days) at 15 gfd on advanced primary effluent with no fouling. During Part 1 testing, the US Filter was cleaned with chlorine after
2,954 hours (123 days) at 11.5 gfd. Shortly after, the system was cleaned using acid which further reduced the TMP. Following this cleaning the membrane operated for nearly 1,000 hours (42 days) at 14.5 gfd during which time little fouling occurred. The system was cleaned again after 4,201 hours of operation to begin testing at increased flux rates ranging from 19-24 gfd. In Part 2 testing no cleanings were performed and the US Filter membrane was operated for 1,000 hours (42 days) at 14.5 gfd during which time little membrane fouling was observed.

The Zenon membrane was cleaned after 800 hours (33 days) and 1,350 (56 days) of operation due to problems associated with the feed pump which caused membrane fouling. After this problem was fixed the membrane was operated for 600 hours (25 days) at 22 gfd before another cleaning was performed. A final cleaning was necessary at time of 2,500 hours (104 days) due to dilution of the solids. The system was then operated for 1,800 hours (75 days) at 22 gfd without cleaning and with minimal fouling. Throughout testing, maintenance cleans were performed three times per week on the Zenon membrane to mitigate fouling. The Mitsubishi MBR system was cleaned after 1,850 hours (77 days) while operating at 11.8 gfd. Following this cleaning the system was operated at low flux due to excessive problems with foaming. The system was brought back online at 3,000 hours of operation and operated for 800 hours (30 days) at a flux of 15 gfd without cleaning. During this time minimal membrane fouling occurred.

Each MBR system demonstrated the ability to operate for a run time of 1800 hours (75 days) between membrane cleanings with minimal to moderate increases in vacuum pressure. The increased amount of cleaning necessary on the Zenon system was largely due to the reduced HRT and increased flux under which the system was operated.

8.4 MBR Effluent Water Quality (Phase I, Part 1: Kubota and US Filter)

8.4.1 Particulate Removal

The US Filter and Kubota MF membranes produced turbidity values ≤ 0.06 NTU and ≤ 0.10 NTU, respectively, in 90 percent of the samples as shown in Figure 8-1. The slightly lower turbidity values measured in the US Filter MBR effluent as compared to Kubota MBR effluent may be due to differences in the on-line turbidity instrumentation equipped on each MBR system. As presented in Appendix C, the US Filter was equipped with a GLI Accu4 turbidimeter while the Kubota MBR system was equipped with Hach 1720D turbidimeter. This was confirmed by analyzing a series of grab samples from each membrane using a desktop turbidimeter. Results showed the average turbidity of the US filter and Kubota membranes to be 0.06 NTU and 0.07 NTU, respectively.

8.4.2 Organics Removal

Both MBR systems produced excellent removal of organic constituents while operating on raw wastewater. For instance, as shown in Figure 8-2, the BOD_5 measured in the MBR effluent was below the detection limit of 2 mg/L in 92 percent of the US filter samples and 100 percent of the Kubota samples. Figure 8-3 shows a probability plot of TOC measured in the raw wastewater and effluent from both MBR systems. As shown, both the US filter and Kubota MBR systems produced TOC ≤ 9 mg/L in 90 percent of all sample measurements.

8.4.3 Biological Nutrient Removal

During Part 1 testing the US Filter and Kubota MBR systems produced effluent with $NH_3 < 2$ mg-N/L in 90 percent of all samples measured as shown in Figure 8-4. Such results indicate that the both MBR systems successfully achieved nitrification during operation on raw wastewater. Also, shown the total inorganic nitrogen in the US filter system was < 27 mg-N/L in 80 percent of samples while the Kubota MBR effluent was <2 mg/L in 100 percent of the samples. Such results indicate that Kubota system successfully achieved complete denitrification throughout Part 1 testing. As expected denitrification was not observed in he US filter MBR because the anoxic tank was bypassed and the system was only operated with an aerobic zone. Both MBR systems showed removal of Ortho-phosphate during Part 1 testing as shown in Figure 8-5. The MBR PO_4 measured < 0.5 mg-P/L and < 0.1 mg/L in 90 percent samples of the US filter and Kubota MBR systems respectively. The higher removal of Ortho-phosphate removal by the Kubota system was attributed to the presence of the anoxic zone which provides a conducive environment for BPR occur once.

8.4.4 Total Coliform, Fecal Coliform, Total Coliphage Removal

Both MBR systems removed total and fecal coliform throughout Part 1 Testing as shown in Figures 8-6 and 8-7, respectively. Total coliform analysis showed the US Filter MBR effluent contained ≤1,000 MPN/100 mL in 90 percent of all samples and Kubota MBR effluent was ≤2 MPN/100 mL in 100 percent of the samples measured.

Fecal coliforms in the US filter MBR effluent were ≤100 MPN/100 mL in 90 percent of samples and Kubota MBR effluent ≤2 MPN/100 mL in 100 percent of the samples measured.

Figure 8-8 shows the US Filter MBR total coliphage was ≤ 30 PFU/100 mL in 80 percent of samples and Kubota MBR total coliphage effluent was <10 PFU/100 mL in 80 percent of the samples.

8.5 MBR Effluent Water Quality (Phase II: Zenon and Mitsubishi)

8.5.1 Particulate Removal

The Zenon UF and Mitsubishi MF membranes produced turbidity values < 0.10 NTU, respectively, in 90 percent of the samples as shown in Figure 8-9.

8.5.2 Organic Removal

Both MBR systems produced excellent removal of organic constituents during Phase II testing. As shown in Figure 8-10, the BOD_5 measured in the MBR effluent was below the detection limit of 2 mg/L in 100 percent of the samples measured in the Zenon and Mitsubishi MBR effluent samples. Figure 8-11 shows a probability plot of TOC measured in the primary effluent and effluent from both MBR systems. As shown both the Zenon and Mitsubishi MBR systems produced TOC ≤9 mg/L in 90 percent of all sample measurements.

8.5.3 Biological Nutrient Removal

During Phase II testing the Zenon and Mitsubishi MBR systems produced effluent with NH_3 < 1 mg-N/L in 90 percent and 70 percent, respectively, as shown in Figure 8-12. The samples of Mitsubishi effluent that were >1 mg/L-N were measured during the start up period, prior to the establishment of the nitrifying bacteria in the MLSS. Such results indicate that both MBR systems successfully achieved nitrification during operation on primary effluent. As shown in Figure 8-13 PO_4 was measured to be ≤0.65 mg-P/L and ≤0.75 mg/L in 70 percent of the samples from Zenon and Mitsubishi MBR systems, respectively. Accordingly, the primary effluent contained ≤ 0.9 mg-P/L in 70 percent of the samples. Such data indicates that Ortho-phosphate removal was not significant during Phase II testing.

8.5.4 Total Coliform, Fecal Coliform, Total Coliphage Removal

Both MBR systems removed total and fecal coliforms throughout Phase II testing as shown in Figures 8-14 and 8-15 respectively. The Zenon MBR produced ≤1,100 MPN/100 mL in 80 percent of all effluent samples and Mitsubishi MBR effluent was ≤10 MPN/100 mL in 80 percent of the samples measured. Higher counts of total coliform measured in the Zenon permeate was shown to be a result of contamination on the permeate side of the membrane.

Fecal coliforms in the Zenon MBR effluent were ≤2 MPN/100 mL in 80 percent of samples and ≤2 MPN/100 mL in 100 percent of the samples measured from the Mitsubishi MBR.

Figure 8-16 shows the Zenon UF membrane total coliphage were ≤ 1 PFU/100 mL in 100 percent of samples and Mitsubishi MF membrane produced total coliphage effluent ≤20 PFU/100 mL in 80 percent of the samples.

9. Cost Analysis

9.1 Costing Approach

A cost analysis was performed to determine capital and operational costs of full-scale MBR water reclamation systems for treatment capacities ranging from 0.2-10 MGD (800-40,000 m^3/day). The specific approach used to perform the costs analysis is outlined in Figure 9-1. As shown, the project team began by first organizing a workshop with all participating MBR manufacturers including: US Filter Corporation/Jet Tech Products Group, Zenon Environmental, Inc., Ionics/ Mitsubishi Rayon Corporation, Enviroquip Inc./Kubota Corporation. During this workshop members of the project team met with representatives from each manufacturer to discuss the major factors affecting the cost and operation of full-scale MBR systems. Based on information gathered during this workshop, the project team developed a specific list of operational and design criteria to be used in preparing the cost estimates. Key parameters included flux, HRT, SRT, and MLSS. In addition, items such as cleaning interval, membrane replacement period and warranty information was established based on discussions with the manufacturers.

Following the workshop, the project team developed and sent a memo to each manufacturer requesting budgetary cost estimates of capital and O&M costs for the membrane portion of MBR systems for the capacities being considered. A modified version of this memo which contains information specific to the current study is provided in Appendix F. As described, the manufacturers were requested to provide membrane costs based on specific operating parameters such as flux, TMP, loss of active membrane area and redundancy. At the same time, the MWH design team completed cost estimates for complete MBR water reclamation systems (excluding membrane costs) including headworks, process basins, mechanical equipment, blower and pump building, chlorination system and effluent storage. Initial cost estimates were based on the operation of raw wastewater. These costs were further refined for 1 and 5 MGD installations to determine the cost savings associated with operation on advanced primary effluent.

9.2 Operation on Raw Wastewater

9.2.1 Design Criteria

Cost analyses were performed for 0.2, 0.5, 1.0, 5 and 10 MGD (800, 2,000, 4,000, 20,000, and 40,000 m^3/day) installations. All systems were assumed to be sewer mining or scalping facilities built on a clean plot of land and designed to operate on raw municipal wastewater. Such facilities differ from "end-of-pipe" systems as raw wastewater is acquired directly from a sewer pipe and all residuals (screenings, grit and waste-activated sludge [WAS]) are returned to the same pipe which eliminates the need for sludge handling and disposal. The following wastewater characteristics, typical of raw municipal wastewater, were used to model the MBR systems:

BOD_5	290 mg/L
COD	700 mg/L
TSS (total suspended solids)	320 mg/L
VSS	260 mg/L
NH_3-N	30 mg/L
TKN (total Kjeldahl nitrogen)	60 mg/L
TP (total phosphorus)	2 mg/L
TDS	1,200 mg/L
Alkalinity	245 mg/L
Temperature	20 ºC

The MBR systems were designed using the following criteria:

Flux	15 gfd @ 15 ºC
MLSS	8,000 mg/L
F/M	0.13 day-1
HRT	6 h
SRT	10 days

Furthermore, all installations were designed to meet the following effluent water conditions:

- Complete nitrification (i.e. NH_4+-N<1.0 mg/L)
- Denitrification (i.e. NO_3-N<10 mg/L)
- Biochemical Oxygen Demand (BOD_5) < 2.0 mg/L
- Biological Phosphorus Removal (i.e. Total Phosphorus-P <0.2 mg/L)

A schematic of the MBR reclaimed water system is provided in Figure 9-2. As shown, the system included a biological reactor with three distinct zones (anoxic, anaerobic and oxic), membrane bays and a chlorine contact chamber. The system was designed to allow screened and degritted wastewater to enter the anoxic zone. Next, the wastewater would pass through the anaerobic and oxic zone before entering the membrane bays. As shown, solids would then be re-circulated from the membrane bay to the oxic zone. This would provide a crossflow velocity on the membrane surface, which would help mitigate fouling and allow excess DO to be consumed. As shown, MLSS was also re-circulated from the oxic zone to the anoxic zone. This allows nitrates produced from the nitrification process to be brought into the anoxic environment, which is conducive for denitrification. Lastly, re-circulation from the anoxic zone to the aerobic zone promotes enhanced biological phosphorous removal (EBPR).

9.2.2 Capital Costs

Table 9-1 provides the capital costs for each capacity designed to operate on raw wastewater. The table includes total capital costs ($K) and amortized capital costs ($K/year) assuming a 5 percent interest rate over a 30 year period. As shown, the total capital cost estimate for the 1.0-MGD installation ranged from $7,710– $9,280, while the amortized cost ($/yr) ranged from $502-$604. The range in capital costs directly reflects the range of membrane costs acquired from the four participating MBR manufacturers.

The headworks for all installations consisted of bar screening (6 mm), vortex grit removal, lift pumps and odor control. All capital costs associated with headworks were taken from standard budgetary costs used by the MWH.

Basin costs include concrete and ancillary costs associated with the aerobic/membrane, anoxic and anaerobic components of the MBR system. In addition, the costs include basin excavation, structural fill, back fill and waste dirt to haul off site. Lastly, for the 1, 5 and 10 MGD installations the basin costs includes a 5-ton bridge crane; for smaller capacities it was assumed the bridge crane would be rented “as needed” and therefore was included in the O&M costs (See Section 9.2.3).

Mechanical costs shown include fine screening, mixers, aeration equipment, and recirculation pumps and piping. Fine screening costs were provided by Waste Tech Inc (Libertyville, IL). The costs were based on Roto-Sieve (RS) perforated drum screens and includes costs of both duty and stand screens as recommended by the manufacturer. A factor of 25 percent was included in the mechanical cost to account for equipment installation.

Membrane system costs including membranes, pumps, blowers and miscellaneous equipment were developed from budgetary cost proposals provided by the participating manufactures. Each manufacturer was requested to provide membrane costs to include a 5-year non-prorated warranty.

Blower and pump building costs shown are based on two-story building and include all capital costs associated with process blowers, blower piping and valving and blower instrumentation. A factor of 25 percent was added the cost to account for equipment installation.

9.2.3 Operation and Maintenance Costs

Table 9-2 provides the annual O&M costs and the total estimated O&M costs (5 percent interest rate over a 30-year period) for all MBR installations considered. Membrane replacement costs were provided by the participating manufacturers and are based on an 8-yr membrane life. The MWH design team provided all other annual costs. Unit cost assumptions for these annual costs are provided in Appendix F. The table shows the annually O&M cost ($K/yr) for the 1-MGD installation ranges from $158-$212.

9.2.4 Total Costs

Table 9-3 provides a summary of the capital and O&M costs for all capacities operating on raw wastewater. The total capital costs and estimated O&M costs were summed to provide present worth values of each installation. The present worth values shown are based on a 5 percent interest rate over a 30-year period. As shown, the present worth ($K) for the 1-MGD was estimated between $10,139-$12,539. Table 9-4 provides total costs ($/1000 gallon) for each capacity. These costs were derived from the amortized capital cost and the annual O&M cost associated with each capacity. The table shows the total cost ($/1000 gallon) for the 1-MGD capacity ranged from $1.81-$2.24.

Figure 9-3 illustrates the range of total costs ($/1000 gallon) based on the various membrane suppliers for 0.2-10 MGD installations operating on raw wastewater. The shaded area on the graph shows the difference between the high and low end of the range. As shown, the range is greatest for 0.2-MGD facilities and decreases with capacity.

9.3 Consideration of Advanced Primary Treatment

The above cost estimates were tailored for municipalities which, like the City of San Diego, are considering using the MBR process to reclaim wastewater at an existing advanced primary wastewater treatment facility. The major factors considered when performing the cost estimates for such facilities were:

- Access to advanced primary effluent
- Ability to use the existing headworks

Reclaimed water generated at an existing facility, such as PLWTP, can be used to meet industrial and irrigation demands on-site reducing the use of imported potable water.

9.3.1 Design Criteria

To accommodate for on-site demand and the potential for increased demand from adjacent areas, costing was performed for 1 and 5-MGD (4,000 and 20,000 m^3/day) MBR facilities. All facilities were assumed to be scalping facilities built on a clean plot of land and designed to operate on advanced primary effluent. The following wastewater characteristics, typical of advanced primary treated municipal wastewater, were used to perform the process design of the MBR systems:

BOD_5	130 mg/L
COD	280 mg/L
TSS	65 mg/L
VSS	50 mg/L
NH_3-N	30 mg/L
TKN	40 mg/L

TP	2 mg/L
TDS	1,200 mg/L
Alkalinity	230 mg/L
Temperature	20 ºC

Comparing the above water quality to the raw wastewater quality (Section 9.2.1) it is evident there is a significant reduction of organic (BOD/COD) and particulate (TSS/VSS) contaminates by the advanced primary treatment process.

All installations on advanced primary effluent were based on following criteria:

Flux	15 gfd @ 15 ºC
MLSS	8,000 mg/L
F/M	0.13 day-1
HRT	3 h
SRT	10 days

It should be noted the HRT was reduced by 50 percent (i.e. 6 hours to 3 h) as compared to MBR installations designed to operate on raw wastewater. This reduction is attributed to the lower organic and solid loading rate to the MBR system during operation on advanced primary effluent.

9.3.2 Capital Costs

Table 9-5 provides capital costs for MBR installations designed to operate on advanced primary effluent. The table shows the total capital cost ($K) and the amortized cost ($K/yr) assuming 5 percent interest rate over a 30-yr period for each capacity. As shown, the total capital cost for the 1.0-MGD installations range from $6,150– $7,730, while the amortized costs ($K/yr) range from $400-$503. A comparison of these costs to the capital cost estimates for the 1.0 MGD installations operating on raw wastewater, indicates a savings between 17 percent-20 percent is realized by designing MBRs to operate on advanced primary effluent when no headworks costs are considered.

The specific capital cost items reduced for MBR plants operating on advanced primary effluent include basin costs, mechanical equipment costs and blower and pump building costs. All of the costs listed above are related to the solid and organic loading rate to the MBR system, which is reduced during operation on advanced primary effluent. In addition, the capital cost is reduced due to the exclusion of a headworks system, which would be necessary if the facility was not being built at an existing plant.

9.3.3 Operation and Maintenance Costs

Table 9-6 provides the annual O&M costs for MBR facilities designed to operate on advanced primary effluent. As shown, the O&M costs are provided for both the first year and total estimated costs, assuming a 5 percent interest rate over 30-yr period. Membrane replacement costs were provided by the participating manufacturers and are based on 8-yr membrane life. All other annual costs were provided by MWH. Specific unit costs and assumptions regarding annual costs are provided in Appendix F. As shown in Table 9-6, the annual O&M costs ($K/yr) for the 1-MGD installations range from $139-$194. A comparison of these costs to the annual O&M cost estimates for the 1.0 MGD installations operating on raw wastewater, indicates savings of 8 percent-12 percent is realized by designing MBRs to operate on advanced primary effluent when no headworks costs are included.

O&M cost reduction for systems operating on primary effluent is attributed to lower electrical requirements for the biological process, reduced equipment repair and reduced diffuser replacement.

9.3.4 Total Costs

Table 9-7 provides a summary of the capital and O&M costs for 1.0 and 5.0 MGD capacities designed to operating on primary effluent. The total capital costs and estimated O&M costs were summed to provide present worth values of each installation. The present worth values shown are based on a 5 percent interest rate over a 30-year period. As shown, the present worth ($K) for the 1-MGD was estimated between $8,287-$10,712. Table 9-8 provides total costs ($/1000 gallon) for each capacity. These costs were derived from the amortized capital cost and the annual O&M cost associated with each capacity. The table shows the total cost ($/1000 gallon) for the 1-MGD capacity ranged from $1.48-$1.91.

Figure 9-4 shows the total costs ($/1000 gallon) for 1.0 and 5.0 MGD installations operating on raw wastewater and advanced primary effluent. The costs in this graph are based on median total cost values presented in Section 9.2.4 and 9.3.4. As shown, the total cost is reduced for both capacities based on design using advanced primary effluent (excluding headwork costs) and the reduction increases with capacity.

9.4 Economy of Scale Analysis

Figure 9-5 presents total costs ($K/MGD) for 1.) complete MBR systems and 2.) membranes only. Costs shown are for systems designed to operate on raw wastewater and are based on median values determined in Sections 9.2.4. The plot shows both costs decrease with increasing capacity indicating an economy of scale; however, the scale is more profound for the complete MBR system costs.