Flux Optimization in Passive Membrane Systems with Air Sparging and Relaxation

by

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B.Sc. Civil Engineering, Bangladesh University of Engineering and Technology, 2012

A THESIS SUBMITTED IN PARTIAL FULFILLMENT OF

THE REQUIREMENTS FOR THE DEGREE OF

MASTER OF APPLIED SCIENCE

in

THE FACULTY OF GRADUATE AND POSTDOCTORAL STUDIES

(Civil Engineering)

THE UNIVERSITY OF BRITISH COLUMBIA

(Vancouver)

MARCH 2016

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Abstract

Traditional membrane filtration plants for drinking water require uninterrupted electricity for pumps and fouling control, thus making it unsuitable for small/rural communities and developing countries. Gravity driven passive membrane filtration systems can be a possible solution to this problem. Previous studies demonstrated that frequent air sparging is beneficial to maintaining a high permeate flux in passive membrane systems. Previous studies also reported that forward flushing after relaxation is also beneficial to maintaining a high permeate flux. Air sparging is an alternate solution to forward flushing after relaxation and considered in the present study.

Four different air sparging rates were considered: no air sparging, continuous air sparging, 5 min/day and 5 min/2 days. Periodic air sparging significantly increased the steady-state permeability (0.39±0.003 B/B_i and 0.37±0.007 B/B_i for 5 min/ day and 5 min/2 days respectively) compared to conditions with no air sparging (0.21±0.006 B/B_i). The highest permeability (0.56±0.036 B/B_i) was achieved with continuous air sparging. Three different relaxation periods prior to periodic air sparging (5 min/day) were tested (1 hr, 4 hrs and 8 hrs). Relaxation prior to periodic air sparging increased the steady-state permeability (0.47±0.012 B/B_i and 0.41±0.023 B/B_i for 1 hr and 4 hrs relaxation period respectively) compared to condition without relaxation prior to air sparging (0.39±0.003 B/B_i). However, lower permeability was observed when a longer relaxation period (0.25±0.002 B/B_i for 8 hrs relaxation period) was considered.
Preface

This dissertation is original, unpublished, independent work by the author, Md Nesar Khadem.
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Nomenclature

$\Delta P$ Transmembrane pressure

$\mu$ Dynamic viscosity of water

$\mu_{20}$ Dynamic viscosity of water at 20°C

$\mu_T$ Dynamic viscosity of water at T°C

AOC Assimilable Organic Carbon

ASTM American Society for Testing and Materials

ATP Adenosine Triphosphate

B Membrane permeability

$B_s$ Normalized membrane permeability at steady-state

$B_i$ Initial membrane permeability normalized to 1

$B/B_i$ Normalized membrane permeability

CLSM Confocal laser scanning microscopy

NF Nanofiltration

Da Dalton

DI De-ionized water

Dia. Diameter

DOC Dissolved organic carbon

F Feed

EDTA Ethylenediaminetetraacetic acid

EPS Extracellular polymeric substances

GE General Electric

HCl Hydrochloric acid

HDPE High density polyethylene
HS  Humic substances
HMW  High molecular weight
ID  Inside diameter
J  Flux
J_o  Initial operating flux
LC  Liquid chromatography
L/m²·h  Litres per meter square per hour
LMW  Low molecular weight acid
MBR  Membrane biological reactor
MF  Microfiltration
MW  Molecular weight
NaClO  Sodium Hypochlorite
NOM  Natural organic matter
OCD  Organic carbon detection
OD  Outside diameter
PBBR  Packed Bed Biofilm Reactor
PES  Polyether Sulfone
PI  Propidium Iodide
PT  Pressure transducer
PVDF  Polyvinylidene Fluoride
R  Total resistance
R_{IR}  Irreversible fouling resistance coefficient
R_{M}  Membrane resistance coefficient
RO  Reverse osmosis
R_{R}  Reversible fouling resistance coefficient
R^2  Coefficient of determination
SEM  Scanning Electron Microscope
SMR  Submerged Membrane Reactor
SSC  Saline-Sodium Citrate buffer
SSF  Slow Sand Filter
\( t \)  Time
\( t \text{ATP} \)  total Adenosine Triphosphate
TBE  Tris Borate EDTA
TMP  Transmembrane pressure
Tris  Tris (hydroxymethyl) aminomethane
UF   Ultrafiltration
\( V \)   Volume
ZW-1000 ZeeWeed®-1000 membrane
ZW-500 ZeeWeed®-500 membrane
Acknowledgements

The completion of this thesis would not be possible without the contributions of many people that I truly respect and care about. For that, I would like to take the opportunity to express my sincere gratitude to them. I would like to thank my supervisor, Dr. Pierre Bérubé, for always asking me the question “what do want to say, and does that answer your question”. I am grateful for his passion in teaching and dedication to his students for make topics simplified. His attendance and patience in all stages of my learning curve and research experience is very much appreciated. I would like to thank Paula Parkinson and Tim Ma for their valuable training in the lab. Their tireless assistance and amicable friendships gave incredible values to my laboratory experience. I would also like to thank Bill, Harold and Scott from Rusty Hut for their assistances and suggestions, which enabled this research project successful. I would also like to thank for the sincere friendships and support from all of my cohort in the graduate student office (Room M107).

Special thank you is forwarded to Patricia Oka, who trained me to run instruments, which was used in this research. While working with her as a research assistant, it was she who made me interested in this topic. I am also grateful for the assistance of François St-Pierre, during my research who periodically helped me with water analyses and sampling. And off course, Jörg Winter for helping me collecting water from Jericho Pond in cold or sun. Also talking to him about my research came very handy in every stages of the research. I also like to thank Syed Zaki and Sardar Muntasir who were very kind to contribute their time and expertise.
Dedication

This thesis is dedicated to my parents, for their devotion and effort for me.

This thesis is also dedicated to my friend and wife Tuba, for loving me.

This thesis is also dedicated to all the great people I met during staying in Vancouver.
1 INTRODUCTION

Traditional drinking water treatment methods generally use physical separation processes to remove particles, as well as chemical or biological treatment processes to remove dissolved pollutants and organic matter. Unlike traditional methods, membrane filtration technologies can potentially achieve equivalent or even better contamination removal than traditional methods in a single step. However, the auxiliary systems generally required for sustainable operation of membrane systems are mechanically complex and require uninterrupted electricity. As a result, membrane filtration technology is often too complex and/or expensive to implement in small/rural communities and developing countries.

Recent studies have indicated that under sub-critical flux conditions, membrane systems can operate with minimal or even without any auxiliary fouling control systems (Peter-Varbanets et al. 2010). Long term operation has been attributed to the establishment of microbial community on the membrane surface which causes steady-state flow over the filtration period. However, operation with some auxiliary systems, such as frequent air sparging (Oka 2015), relaxation (Peter-Varbanets et al. 2012), relaxation with forward flushing (Peter-Varbanets et al. 2012), backwashing (Oka 2015), feed water modification (Peter-Varbanets et al. 2011; Chomiak et al. 2014) and pre-treatment of feed water (Derlon et al. 2014), has been reported to improve the operation of passive membrane systems.

1.1 Objectives

The objectives of the present research were to develop a sustainable passive filtration system using hollow fiber UF membranes that can operate with minimum energy and operational complexities. In meeting the objectives, hypotheses were made that: a) fouling layer that formed during passive UF hollow fibre membrane filtration contains a microbial community, b) the biomass concentration in the fouling layer changes under different operating conditions and affects permeate flux, c) periodic air
sparging increases permeate flux, d) relaxation prior to air sparging helps to increase permeate flux and e) increasing hydrostatic pressure increases permeability.

1.2 Thesis Structure

Chapter 2 discusses the basic properties of membrane filtration and related studies by others. Chapter 3 presents the knowledge gaps with current knowledge that provided the objectives of the present study. Chapter 4 describes in detail the materials and methods that were used to answer the knowledge gaps. The data analyses and results obtained are presented and discussed in Chapter 5. Chapter 6 discusses the conclusions and engineering significance of the research.
2 LITERATURE REVIEW

2.1 Membrane Filtration

Membrane filtration involves the use of a selective barrier for the removal of selected contaminants from a solution. Depending on the contaminants they can remove from solution, membranes are classified into two broad groups: 1) High pressure systems (Reverse Osmosis and Nanofiltration) which are used for removal of dissolved contaminants and 2) Low pressure systems (Microfiltration and Ultrafiltration) which are used for removal of suspended and colloidal particles. A summary of the categories of membranes typically used and their basic properties is presented in Table 1 and Figure 1. Ultrafiltration membranes (UF) can generally provide up to 4 log removal of biological contaminants and colloids in drinking water treatment. UF membranes are more popular in filtration industries for drinking water. The present study focuses on UF membranes.

Table 1 Membrane classification

<table>
<thead>
<tr>
<th>Classification</th>
<th>Reverse Osmosis (RO)</th>
<th>Nanofiltration (NF)</th>
<th>Ultrafiltration (UF)</th>
<th>Microfiltration (MF)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pore size (nm)</td>
<td>0.1-1</td>
<td>1-10</td>
<td>10-1</td>
<td>100-1000</td>
</tr>
<tr>
<td>Contaminant Removal Mechanism</td>
<td>Sieving Diffusion</td>
<td>Sieving Diffusion</td>
<td>Sieving</td>
<td>Sieving</td>
</tr>
<tr>
<td>Operating Pressure (Bar)</td>
<td>10-30</td>
<td>4-20</td>
<td>0.1-2</td>
<td>0.1-2</td>
</tr>
<tr>
<td>Permeate Flux (L/m²-h)</td>
<td>4-40</td>
<td>8-80</td>
<td>20-200</td>
<td>50-500</td>
</tr>
</tbody>
</table>

(Metcalf and Eddy 1991)
Membrane filtration can be operated under constant flux or constant pressure conditions (Sagle and Freeman 2004). In constant pressure operation, the pressure increases over time due to the accumulation of foulants on the membrane, resulting in an increase in transmembrane pressure (TMP). For constant pressure operation, permeate flux declines over time with accumulation of foulants. The present study focuses on constant pressure operation.

Based on the flow direction of feed water, membrane systems used for drinking water treatment are defined as either an external system or a submerged system (Figure 2). In external systems the feed to be filtered is pumped from a feed tank through the membrane filter (Positive Pressure). Permeate (filtered water) is collected from the outside (permeate side). The pump creates the TMP differential that allows the feed to pass through the membrane (inside-out flow). Fouling from the feed side is recirculated or purged periodically to maintain a high permeate flux.

In submerged systems, the membrane is submerged inside a feed tank. A suction pump is connected to the permeate side of the membrane. The pump creates a negative pressure differential that
allows the feed to pass through the membrane (outside-in flow). Fouling from the feed side is purged or sparged periodically to maintain a high permeate flux. The present study focuses only on submerged membrane systems.

Figure 2 Schematic of typical membrane operation. (a) external system; (b) submerged system

Four types of membrane modules are commonly used based on the geometry of the membrane: plate and frame, tubular, spiral wound, and hollow fiber (Sagle and Freeman 2004). Hollow fibers are relatively long and narrow, with fiber length in the range of 1 to 2 m and fiber diameter between 0.5 to 2 mm (EPA 2005). A typical hollow fiber module may consist of anywhere from a few hundred to thousands of fibers (EPA 2005). Hollow fibers may be operated in an inside-out or outside-in configuration, as indicated in Figure 3. Inside-out fibers are more disadvantageous compared to outside-in fibers because of the susceptibility to clogging of the lumen during operation. The filtration
area is also lower in inside-out configuration compared to outside-in configuration. Additionally, there is more head loss in the inside-out than the outside-in configuration, since all the feed flows through the lumen in the inside-out configuration, while only the permeate flows through the lumen in the outside-in configuration (Chan 2010). Hollow fibers are generally in an inside-out configuration for external pressure driven operation, and in outside-in configuration for submerged vacuum–driven operation (Chan 2010). The present study focuses on outside-in hollow fiber membranes.

![Diagram of inside-out and outside-in configuration in hollow fiber membrane](image)

Figure 3 Schematic of the inside-out and outside-in configuration in hollow fiber membrane

### 2.2 Membrane Fouling and Fouling Control

Membrane fouling is one of the major limitations of membrane technologies (Le Clech et al. 2003; Marroquin et al. 2014). Membrane fouling can be described as the pressure accumulation of retained materials on or in the membrane leading to pore constriction, pore blockage, and/or cake/gel layer formation (Lee et al. 2005). In water filtration, fouling results from a combination of different mechanisms simultaneously or individually depending on the characteristics of the feed water and operational parameters (Bolton et al. 2006; Tang, et al. 2011). Fouling causes an increase in the TMP (constant flux operation) or reduces the permeate flux (constant pressure operation) by increasing the resistance to permeate flow through the membrane.
Fouling can be reversible or irreversible in nature. Reversible fouling is of two types, hydraulically reversible and chemically reversible. Hydraulically reversible fouling can be minimized by applying back flush, tangential flow or shear flow (air sparging) (Jermann et al. 2007). Chemically reversible fouling can be minimized by chemical cleaning methods. Irreversible fouling cannot be eliminated and causes a permanent decrease in permeability and eventually leads to membrane replacement.

The relationship between permeate flux and resistance is presented in Equation 1.

\[
J = \frac{\Delta P}{\mu (R_M + R_{IR} + R_R)}
\]

where, \(J\) is volumetric permeate water flux through the membrane (L/m²·h), \(\Delta P\) is the transmembrane pressure (TMP) (bar), \(\mu\) is dynamic viscosity of water (kg/m·s), \(R_M\) is membrane resistance coefficient (1/m), \(R_{IR}\) is irreversible fouling resistance coefficient (1/m), \(R_R\) is reversible fouling resistance coefficient (1/m). This equation can be rewritten as

\[
B = \frac{\mu J}{\Delta P}
\]

where, \(B\) is membrane permeability (m).

Fouling can generally be modeled based on one of the four mechanisms: adsorption, pore blocking, cake deposition or bio-film formation (Peter-Varbanets et al. 2010) as illustrated in Figure 4.

Figure 4 Schematic of the membrane fouling mechanisms during filtration (a) adsorption, b) pore blocking, c) cake deposition and d) bio-film
Adsorption occurs due to physical and chemical interactions of foulant present in water with the membrane. Over time, adsorption reduces the size of pore openings in the membrane. During filtration, accumulation of foulant particles which are larger than the pore size of the membrane deposits on the membrane surface forming a cake structure. Pore blocking causes complete or partial plugging of the pores by entrapment of foulant. Biofilm formation occurs due to the retention and growth of microorganisms on the membrane surface.

To prevent excessive accumulation of foulant in membrane filtration systems, various foulant prevention and removal techniques, such as forward flush, backwash, purging (wasting), air sparging (scouring), chemical cleaning, pre-treatment of the feed are used individually or simultaneously. During forward flushing the feed side of a membrane is flushed with feed water more rapidly than during the production phase. This causes turbulence resulting in the release of the particles that are on the membrane surface. Forward flushing requires pumping the feed and a large volume of feed is wasted during this process. Backwash of a membrane is the reversal of permeation flow with clean water (i.e. filtered water). Filtered water is flushed through the feed water side of the system under pressure to remove foulant from the membrane pores. Backwash also requires high energy pumps to create high pressure and requires extra piping. Air sparging or air scouring generates turbulence in the water to remove foulant from the membrane surface. Usually air bubbles are introduced from the bottom of the membrane reactor. These rising air bubbles create turbulent conditions near and at the membrane surface and cause scouring of the foulant. To add air to the system, air blowers are generally used. Chemical cleaning uses different chemical agents (e.g. bleach, caustic soda, citric acid) to remove foulants which are not removable by physical processes. Pre-treating the feed prior to filtration reduces the foulant loading on the membrane and decreases the use of other fouling prevention techniques. Fouling controls increase the operational complexities and cost of operation.
2.3 Gravity Driven Passive Membrane Filtration

Gravity or hydrostatic pressure head can be an alternative to pumps to create the transmembrane pressure. Figure 5 represents a basic process for a gravity driven membrane filtration.

![Figure 5 Schematic of the gravity driven filtration system](image)

Fouling control requires electricity for the pumps and chemicals. Thus gravity filtration alone does not completely eliminate the energy required for membrane filtration. Operating at a low flow mode or passive condition reduces the mass transfer towards the membrane substantially. This reduces the energy and or chemicals used for fouling control. A passive membrane system is a sustainable membrane operation method which operates below critical (sub-critical) condition. Critical flux is defined as the theoretical permeate flux of a membrane system under which no or little fouling is observed (Le Clech et al. 2003; Choi and Dempsey 2005). Critical conditions can be obtained in the low pressure membranes with low permeate flux rate (Bérubé et al. 2008; Le Clech et al. 2003) (Figure 6).
Peter-Varbanets et al. in 2010 investigated the use of gravity driven passive filtration in a flat-sheet membrane at pilot scale. Four different hydrostatic pressures (4 kPa, 15 kPa, 25 kPa and 50 kPa) were used. Permeate flux data collected over volume indicated that the permeate flux comes to a steady-state over time (5 days) and continues without any fouling control measures. An increase in biomass was observed on the membrane surface over time. Systems with no microbial communities did not reach steady-state over the same filtration period. They concluded that the microbial growth results in a porous fouling layer and responsible for the sustainable steady-state permeate flux. The presence of a microbial community was confirmed by using ATP (measuring actively growing microorganisms through detection of adenosine triphosphate) measurements and CLSM imaging (detection of a microbial community by confocal laser scanning microscopy). CLSM imaging was also used to observe the physical structure of the fouling layer. Similar low steady-state permeate flux with a microbial community was observed by Oka (2015); Akhondi et al. (2014) and Derlon et al. (2014). This flux stabilization and resistance occurs when the increase in the resistance due to structural changes in the fouling layer containing a microbial community counteracts the decrease in resistance due to deposition and irremovable fouling (Peter-Varbanets et al. 2011). The presence of a microbial community on the membrane surface also increases the quality of permeate in terms of assimilable organic carbon (AOC).
content (Derlon et al. 2014). The experiment also indicated that flux stabilization was independent of hydrostatic pressure (Peter-Varbanets et al. 2010) while the steady-state permeate flux depends on the feed characteristics (NOM) (matter composed of organic compounds) (Peter-Varbanets et al. 2010; Peter-Varbanets et al. 2011). The system proposed by Peter-Varbanets et al. (2010) was field tested in Annet-sur-Marne, France and Ogunjini, South Africa to provide sufficient drinking water for 100-200 people (Boulestreau 2010). Portable Aqua Unit for Lifesaving (PAUL) introduced in 2010 by University of Kassel (Noubactep 2013) based on this process provides decentralized supply of clean water in emergency and disaster situations.

2.4 Fouling Control in Gravity Driven Passive Membrane Filtration

Deposition of non-dissolved material, structural changes in the fouling layer and the development of irremovable fouling over time are the three major causes of fouling in gravity driven passive membrane filtration (Peter-Varbanets et al. 2011). Operating without any physical or chemical fouling control causes steady-state at low flux over the filtration period for gravity driven passive membrane filtration system. To optimize the system in terms of permeate flux, different fouling control mechanisms have been studied: air sparging (cleaning membranes by rising air bubbles) (Oka 2015), relaxation (Peter-Varbanets et al. 2012), relaxation with forward flushing (Peter-Varbanets et al. 2012), backwashing (Oka 2015), feed water modification (Peter-Varbanets et al. 2011; Chomiak et al. 2014) and pre-treatment of feed water (Derlon et al. 2014).

Oka (2015) investigated five different air sparging approaches with hollow fiber membranes (no air sparging, continuous, 25 mins off/5 mins on, 4 hrs off/5 mins on and 4 hrs off/30 mins on). 3.8 L/min airflow was used for systems with air sparging. A hydrostatic pressure of 40 cm was used and 10% of the operating volume was purged daily for all experiments. Compared to systems with no air sparging, systems with air sparging had higher steady-state flux indicating the sparging was beneficial to improve permeate flux. It was also observed that daily purging (wasting of feed from the reactor) improved permeate quality.
Peter-Varbanets et al. (2012) considered flat sheet membrane filtration systems with five different relaxation periods (0 hrs, 3 hrs, 6 hrs, 12 hrs and 19 hrs relaxation period). Results from the experiment indicated that an increase in periodic operation increased the average permeate flux. This was possibly due to the fact that at relatively low pressures, the biofouling does not lead to clogging, but the structure of the fouling layer remains heterogeneous and porous (Peter-Varbanets et al. 2012). Swelling of the foulant layer was observed during relaxation (Valladares Linares et al. 2015). When relaxation with forward flushing was introduced into the system (12 hrs relaxation with and without forward flushing), higher permeate flux was observed compared to system with no forward flushing. The increase in permeability was caused by removal of the swelled foulant layer with forward flushing.

The effect of backwashing on a passive filtration system was investigated by Oka (2015). In her study, she operated three systems with different initial permeate fluxes (30 L/m²·h, 20 L/m²·h and 10 L/m²·h) and with and without backwash. The systems were operated over a 2 month period with a 4 hrs filtration and 10 mins backwash mode. Results from the experiments indicated that significant improvement in permeate flux before and after backwashing with higher initial permeate flux (30 L/m²·h) and no significant improvement in systems with low initial permeate flux (20 L/m²·h and 10 L/m²·h). This indicates that fouling control by backwashing was not significant when the initial permeate flux was low.

Feed water modification by adding inorganic particles to the feed can change the structure of the foulant layer. Peter-Varbanets et al. (2011) operated two gravity driven flat sheet passive membrane systems using river water spiked with Kaolin (30 mg/L and 300 mg/L). They concluded that the presence of inorganic particles did not affect flux stabilization because NOM causes the significant resistance. Chomiak et al. (2014) operated two gravity driven systems with Kaolin (300 mg/L) and Diatomite (300 mg/L) with flat sheet membranes. From the experiments it was observed that small particles (Kaolin) caused a homogenous fouling layer that reduced the permeate flux while large particles (Diatomite) created a heterogeneous fouling layer that increased the permeate flux. They
concluded that a microbial community was the significant source of resistance while heterogeneity in the fouling layer by inorganic particles improves permeability. To quantify the microbial community they used ATP measurements and for the fouling layer structure, a SEM (scanning electron microscope) technique was used.

Derlon et al. (2014) operated three gravity driven passive membrane filtration systems with three different pretreatments: no pretreatment, a slow sand filter (SSF) and a packed bed biofilm reactor (PBBR). Results from their research indicated that feed water with pretreatments helped to increase the mean permeate flux. Mean permeate fluxes of 7.5±1.3 L/m²·h, 8.4±1.3 L/m²·h and 8.9±1.2 L/m²·h were measured for the system with no pretreatment, system with PBBR and system with SSF respectively. Both PBBR and SSF increased the permeate flux at the beginning but after reaching steady-state permeability the difference was less compared to the system with no pretreatment.
3 KNOWLEDGE GAP AND OBJECTIVES

The objective of the present study was to develop a simplified gravity driven passive membrane filtration system that can be used in remote areas or developing countries with minimum maintenance. Knowledge gaps generated from the literature review are the following.

a. A microbial community is significant for achieving the steady-state permeability (Peter-Varbanets et al. 2010). Then, what are the changes that can occur to microbial community under different hydrodynamic conditions that may alter the steady-state?

b. Previous studies indicated that air sparging is beneficial to fouling control for the passive membrane filtration (25 mins off/5 mins on, 4 hrs off/5 mins on and 4 hrs off/ 30 mins on) (Oka 2015). What happens to the permeability if the frequency of air sparging (periodic air sparging) is much lower?

c. Forward flushing after relaxation results in significant improvement in permeability (Peter-Varbanets et al. 2011). What happens to the permeability if forward flushing is replaced by air sparging?

d. Under the above two conditions (periodic air sparging and relaxation prior to periodic air sparging) what happens to the permeability if the hydrostatic pressure is increased?

In meeting the objective, the present study will attempt to address the following research questions:

1. Are there any changes in biomass concentration under different hydrodynamic conditions and what are the effects on the permeate flux?

a. What are the differences in biomass concentration when no purging and no air sparging, daily purging, continuous air sparging, periodic air sparging and air sparging prior to relaxation conditions are applied?

b. Does permeate flux increase or reduce in absence of microbial community?
2. Does periodic air sparging improve permeate flux compared to no air sparging and continuous air sparging?
   a. Does prolonged periodic air sparging affect permeate flux compared to daily air sparging?
3. How effective is relaxation prior to periodic air sparging to improve permeate flux?
   a. Does relaxation affect permeate flux if used before air sparging?
   b. Does relaxation duration affect permeate flux?
4. Does an increase in hydrostatic pressure affect permeate flux?
4 METHODOLOGY

4.1 Primary Experimental Setup

For answering the questions generated from the knowledge gaps, bench scale submerged membrane reactors were used for the present research. Individual components of the primary submerged membrane reactors are listed below.

- Feed system
- Submerged membrane reactor system (SMR)
- Permeate collection System
- Air sparging system
- Purging system

Figure 7 illustrates the simplified layout of the submerged membrane reactor system (R1-R10) and its components, and Figure 12 illustrates the complete layout of the submerged membrane reactor.

![Figure 7 Schematic of the submerged membrane reactor system](image)

Depending on the questions under consideration, the submerged membrane reactor configuration was modified to meet the experimental requirements (Figure 8). Table 2 summarizes the complete experimental program of the study with primary experimental setup.
Table 2 Summery of the primary experimental program

<table>
<thead>
<tr>
<th>System name</th>
<th>Purging</th>
<th>Air sparging</th>
<th>Relaxation</th>
<th>Operation period</th>
<th>Special Note</th>
</tr>
</thead>
<tbody>
<tr>
<td>R1</td>
<td>none</td>
<td>none</td>
<td>none</td>
<td>2014-09-24 to 2014-11-25</td>
<td></td>
</tr>
<tr>
<td>R2</td>
<td>none</td>
<td>none</td>
<td>none</td>
<td>2014-09-24 to 2014-11-25</td>
<td>1 gm/L Sodium Azide in feed water</td>
</tr>
<tr>
<td>R3</td>
<td>10% v/day</td>
<td>none</td>
<td>none</td>
<td>2014-12-11 to 2015-01-30</td>
<td></td>
</tr>
<tr>
<td>R4</td>
<td>none</td>
<td>11.5 L/min, continuous</td>
<td>none</td>
<td>2014-12-11 to 2015-01-29</td>
<td></td>
</tr>
<tr>
<td>R5</td>
<td>10% v/day</td>
<td>11.5 L/min, 5 min/day</td>
<td>none</td>
<td>2015-03-16 to 2015-04-17</td>
<td></td>
</tr>
<tr>
<td>R6</td>
<td>10% v/day</td>
<td>11.5 L/min, 5 min/2 days</td>
<td>none</td>
<td>2015-03-16 to 2015-04-17</td>
<td></td>
</tr>
<tr>
<td>R7 &amp; R7' (replicates)</td>
<td>10% v/day</td>
<td>11.5 L/min, 5 min/day</td>
<td>1 hr/day</td>
<td>2015-03-16 to 2015-04-17</td>
<td></td>
</tr>
<tr>
<td>R8</td>
<td>10% v/day</td>
<td>11.5 L/min, 5 min/day</td>
<td>4 hr/day</td>
<td>2015-06-04 to 2015-07-02</td>
<td></td>
</tr>
<tr>
<td>R9</td>
<td>10% v/day</td>
<td>11.5 L/min, 5 min/day</td>
<td>8 hr/day</td>
<td>2015-06-04 to 2015-07-02</td>
<td></td>
</tr>
<tr>
<td>R10</td>
<td>10% v/day</td>
<td>11.5 L/min, 5 min/day</td>
<td>1 hr/day</td>
<td>2015-06-04 to 2015-07-02</td>
<td>25 cm hydrostatic pressure</td>
</tr>
</tbody>
</table>

Microbial community under different hydrodynamic condition

Permeability flux

Figure 8 Parameters of interest in primary experimental setup

Sodium Azide is a microbial inhibitor. At 1gm/L concentration of Sodium Azide with raw sewage, no microbial community was observed in the coliform plate count after 1, 3 and 7 days. The raw sewage was collected from the UBC Waste Water Pilot Plant.
4.1.1 Feed System

The feed system supplied synthetic raw water into the submerged membrane reactor. The feed system consisted of the following components: feed water, feed tank, feed pump (MasterFlex Pump series: 7520-35) and float switch.

A synthetic raw water consisting of a mixture of pond and tap water was used for the study. Pond water collected from Jericho Pond was used to prepare the feed water. Jericho Pond is located approximately 1 km south of the Jericho beach (Latitude: 49.271 and longitude: -123.195). The pond water is characterized by surface run-off during the winter season and by no flow during the summer season. The TOC (total organic carbon) and DOC (dissolved organic carbon) value varied between 8.7 - 37 ppm and 7.9 - 29 ppm, respectively, during the study period. High algal bloom was observed in the water during dry summer seasons. This pond also serves as a natural habitat for ducks, birds and small fishes.

Collected pond water was sieved through a 1 mm screen to remove large suspended solids. The sieved water was then transported to the UBC Environmental Engineering Laboratory and stored at 4°C until use. Prior to use, the raw water was warmed to room temperature and filtered through a 10 µm filter (Keystone Filtration Solution, Model CGB10). The final filtered pond water was then diluted with tap water to achieve a DOC concentration of approximately 5 ppm to produce the synthetic raw water.

The feed tank was made of high density polyethylene and had a volume of 125 liters. A magnetic float switch and a peristaltic pump (MasterFlex Pump series: 7520-35) were used to add raw water to the submerged membrane reactor and to maintain a constant water level in the feed tank. The feed tank was manually cleaned every two weeks during the operation to remove any precipitated solids.
4.1.2 Submerged Membrane Reactor System (SMR)

The submerged membrane reactor consisted of a system tank, membrane modules, a module connection manifold and a permeate line. The system tank was made of HDPE (high density polyethylene). Two different sizes of tanks were used within the duration of the experiment. Dimensions of the reactor tanks are listed in Table 3. R10 was operated in system tank type 2 and an operating volume of 92.14 liter was used.

Table 3 System tank dimensions

<table>
<thead>
<tr>
<th></th>
<th>Length</th>
<th>Width</th>
<th>Height</th>
<th>Operating Volume</th>
</tr>
</thead>
<tbody>
<tr>
<td>System Tank, type 1</td>
<td>800</td>
<td>508</td>
<td>654</td>
<td>162.5 liters</td>
</tr>
<tr>
<td>System Tank, type 2</td>
<td>450</td>
<td>450</td>
<td>620</td>
<td>81 liters</td>
</tr>
</tbody>
</table>

The membranes used in the present study were ZW-500 (GE Water and Process Technologies, Canada) outside-in flow UF hollow fiber membranes. The physical properties of the membrane fibers are summarized in Table 4.

Table 4 Membrane properties

<table>
<thead>
<tr>
<th>Membrane properties</th>
<th>Unit</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Material</td>
<td>-</td>
<td>PVDF</td>
</tr>
<tr>
<td>Outside diameter (OD)</td>
<td>mm</td>
<td>1.9</td>
</tr>
<tr>
<td>Inside diameter (ID)</td>
<td>mm</td>
<td>0.8</td>
</tr>
<tr>
<td>Nominal pore size</td>
<td>µm</td>
<td>0.04</td>
</tr>
<tr>
<td>TMP range</td>
<td>kPa</td>
<td>-55 to 55</td>
</tr>
<tr>
<td>Max. operating temperature</td>
<td>ºC</td>
<td>40</td>
</tr>
<tr>
<td>Operating pH range</td>
<td>-</td>
<td>5.0-9.5</td>
</tr>
<tr>
<td>Max. cleaning temperature</td>
<td>ºC</td>
<td>40</td>
</tr>
<tr>
<td>Membrane properties</td>
<td>Unit</td>
<td>Value</td>
</tr>
<tr>
<td>-----------------------------</td>
<td>------</td>
<td>--------</td>
</tr>
<tr>
<td>Cleaning pH range</td>
<td>-</td>
<td>2.0-10.5</td>
</tr>
<tr>
<td>Max. Cl₂ concentration</td>
<td>ppm</td>
<td>1,000</td>
</tr>
</tbody>
</table>

Each membrane module consisted of two 230 mm long membrane fibers that were potted together into bulkheads made of 6.35 mm (¼") OD rigid plastic tubing. One end of the fibers was open to allow permeate flow, whereas the other was sealed with epoxy. Epoxy glue was used for potting and sealing the membranes. Each module had an effective filtration area of 0.002387 m². Figure 9 details the configuration of the membrane module.

Figure 9 Schematic and picture of the membrane module: a) membrane module and b) picture of membrane module

The module connection manifold consisted of a 4x4 array with total of 16 connection points (Figure 10). This setup gave the opportunity to work with 16 membrane modules simultaneously. The bulkheads with the open end of the membrane fibers were linked to the connection points using 6.35 mm (¼") OD push in connections with valves. The valves enabled each membrane module to be isolated from the connection manifold for membrane fiber harvesting. A permeate line was connected
to the module connection manifold to collect the combined permeate from all membrane modules. The top of the membrane modules were secured vertically using nylon string with 95% membrane looseness allowing slight movement of the membrane modules during air sparging.

Figure 10 Schematic and picture of the module connection manifold (plan view): a) module connection manifold and b) module connection manifold with membrane modules

### 4.1.3 Air Sparging and Purging System

Sparged air was added at the base of the submerged membrane reactor system through 9.525 mm (\(\frac{3}{8}\)”) diffusers. The openings of the diffusers were located below the module connection manifold. A timer device (ConTrolX) and solenoid valve were used to control the sparged air on/off conditions. A pressure control valve (NORGREN Co. R06-221-NNKA) with a flowmeter (Key Instruments, model no: GS10810AVB) was used to control and monitor the air flow. Oka (2015) used 3.8 L/min air sparging for the membrane reactors. Based on the tank cross sectional area used in the research, an air sparging flow rate of 11.6 L/min air was used for the present study. This ensured similar volume of air bubbles were used for cleaning the membranes by gas sparging. Purging was also done from the base of the submerged membrane reactor system, controlled by a timer and solenoid valve. During purging, a feed volume equivalent to the 10% of the tank operating volume was wasted from the submerged membrane reactor. Purging and air sparging were done in parallel to reduce the
accumulation of solids in the submerged membrane reactor system. The mixing induced by air sparging ensured that settable solids did not accumulate in the reactor tank over time.

4.1.4 Permeate Collection System

The permeate collection system consisted of two parts: permeate flow control and permeate collection system (Figure 11). A solenoid valve equipped with a timer (ControlX) was attached to the permeate collection tube and used to control the permeate flow from the membrane modules. In open valve position permeation occurred and in closed valve position relaxation or no filtration occurred. To maintain a desired hydrostatic pressure on the membrane, the permeate collection tube was connected to a constant liquid level beaker. The constant liquid level beaker was arranged to serve the purpose of constant hydrostatic pressure on the membrane modules. Water level in the constant liquid level beaker was kept constant by draining excess permeate to a permeate collection beaker. Periodically the weight of water in the collection beaker was measured and the collected permeate was further analyzed for different parameters. To measure the permeate flux, permeated volume was divided by permeation time for effective permeating surface and reported as L/m²·h. The density of water was corrected for temperature using Table A1 in Appendix A.

Figure 11 Schematic of the permeate collection system
Figure 12 Detailed schematic of the primary submerged membrane reactor system (R1-R10)
4.2 Secondary Experimental Setup

The R0, R1' and R2' system (Figure 13) were separately built to assess the effect of the presence of a microbial community on the permeate flux. Individual components of the secondary submerged membrane reactors are listed below.

- Feed system
- Submerged membrane reactor system (SMR)
- Membrane module
- Permeate collection system

Table 5 Summery of the secondary experimental program

<table>
<thead>
<tr>
<th>System name</th>
<th>Purging, sparging and relaxation</th>
<th>Operation period</th>
<th>Special note</th>
<th>Parameters of interest</th>
</tr>
</thead>
<tbody>
<tr>
<td>R0</td>
<td>none</td>
<td>2015-09-17 to 2015-09-19</td>
<td>1 gm/L Sodium Azide in RO water</td>
<td>Contribution of the microbial community to the extent of fouling</td>
</tr>
<tr>
<td>R1'</td>
<td>none</td>
<td>2015-09-17 to 2015-09-19</td>
<td></td>
<td></td>
</tr>
<tr>
<td>R2'</td>
<td></td>
<td>2015-09-17 to 2015-09-19</td>
<td>1 gm/L Sodium Azide in feed water</td>
<td></td>
</tr>
</tbody>
</table>

4.2.1 Feed System

The feed system supplied synthetic raw water into the submerged membrane reactor. The feed system consisted of the following components: feed water, feed tank, feed pump (MasterFlex Pump series: 7520-35). Similar procedures were followed to make the feed water as those described in Section 4.1.1. The feed tank was made of HDPE and had a volume of 5 liters.

4.2.2 Secondary Submerged Membrane Reactor System (SMR)

The secondary submerged membrane reactor consisted of a system tank, membrane modules, and a permeate line. The system tank was made of HDPE. A 10 L cylindrical plastic tank was used for
housing the single membrane module and filled up to 8 L. Single membrane module was used in each of the reactors. Similar procedures were followed to make the membrane module as those described in Section 4.1.2.

4.2.3 Permeate Collection System

The outlet of the submerged membrane module was connected to a peristaltic pump (MasterFlex Pump series: 7520-35) to create a negative pressure of 1.96±0.1 kPa. To measure the in line pressure, a pressure gauge (Ashcroft DG25) was used. All three reactors were operated at a similar flow rate of 10 L/m²-h. A similar flow rate was used for the feed peristaltic pumps to keep the water level in the reactor constant. Periodically the weight of water in the collection beaker was measured (Mettler Toledo PB3002-S) and the collected permeate was further analyzed for different parameters.

![Diagram of the permeate collection system](image)

Figure 13 Detailed schematic of the secondary submerged membrane reactor system (R0, R1' and R2')
4.3 Quality Control

4.3.1 Membrane Cleaning

Before each experiment, the membrane modules were cleaned using the following procedure (Bérubé et al. 2008).

1. New membrane modules were soaked in 750 ppm of Sodium Hypochlorite (NaClO) solution (diluted from 10.3% industrial grade bleach with RO water) for 24 hrs;
2. The 750 ppm solution of NaClO was then filtered through the membrane module at -28 kPa vacuum for a period of 20 minutes;
3. The membrane modules were transferred to 50 ppm NaClO solution which was filtered through the membrane at -28 kPa vacuum for a period of 20 minutes;
4. The clean membrane modules were preserved in 50 ppm NaClO solution.

Prior to use, each module was rinsed by soaking in RO water for 1 hour. For membranes in the secondary submerged reactor, membranes were soaked in RO water and Sodium Azide (1 gm/L) and prior to use were cleaned with RO water.

4.3.2 Membrane Integrity Testing

Before starting each new experiment, the integrity of the cleaned membrane modules as verified using the pressure hold (bubble) test as per the following procedure.

1. Compressed air (41.4 kPa) was applied to the inside of a submerged hollow fiber.
2. The pressurized submerged membrane module was then monitored for 2 minutes using a pressure gauge (Wika DG-10-S).

If any significant (i.e., >10%) pressure drop or bubble formation was detected, the module was considered breached and discarded. The setup is illustrated as in Figure 14.
Before starting each new experiment, the integrity of the module connection manifold as verified using the pressure hold (bubble) test as per the following procedure.

1. Compressed air (41.4 kPa) was applied to the inside of a submerged module connection manifold.
2. The pressurized submerged module connection manifold was then monitored for 10 minutes using a pressure gauge (Wika DG-10-S).

If any bubble formation was detected, the module connection manifold was repaired and tested again. During operation the integrity of the module connection manifold and membrane modules

4.3.4 Clean Water Permeation and Priming

A clean water permeation test using RO water was conducted at the start of each experiment to measure the intrinsic resistance of the membrane modules. Prior to the test, a pump was used to draw RO filtered water through the membranes for a period of 24 hours at 10 L/m²·h. This ensured removal of bubbles from the membrane modules and the module connection manifold. Following the initial priming period, a constant hydrostatic pressure of 1.96 kPa (20 cm hydrostatic pressure) provided the driving force for the permeation through the modules. The permeate flow rate was measured for a period of one hour. The initial membrane resistance was calculated using Equation 3.
Equation 3  \[ R = \frac{\Delta P}{\mu J} \]

where, \( R \) is the membrane resistance (1/m).

All experiments were performed with hydrostatic pressure head providing the 1.96 kPa transmembrane pressure required for permeation. As for the clean water permeate tests, prior to the start of each experiment, the system tank was filled with RO water and a pump (MasterFlex Pump series: 7520-35) was used to provide a transmembrane pressure equivalent to that provided by hydrostatic pressure head (i.e. 1.96 kPa) for a period of 2 hours. This was observed to be required to remove small bubbles from within the module connection manifold. The density and viscosity of water were estimated as required for different temperatures (Appendix A).

4.3.5 Sampling

Samples for analysis were collected from the feed tank, system reactor and permeate collection beaker every 1 to 5 days throughout the duration of each experiment. The feed tank was sampled at mid-depth after fully stirring the contents of the tank. The system tank with no purging and no air sparging was sampled 30 cm below the height of the liquid level after purging. The system tank with air sparging and/or purging was sampled after air sparging and purging. The system tank with relaxation was sampled after relaxation. The permeate samples were taken directly from the permeate collection beaker. All samples were collected in a beaker that had previously been rinsed with the sample.

To harvest a membrane module for biomass analysis and imaging, the corresponding membrane module valve was closed when the module was to be harvested. To harvest, fibers in a module were cut at the lower base of the modules above the bulkhead. Care was taken not to disturb other membrane modules in the system tank when fibers were harvested. Harvesting was done before purging, air sparging and relaxation.
4.4 Analytical Methods

4.4.1 Adenosine Triphosphate (ATP)

Adenosine Triphosphate (ATP) was used as a surrogate measurement approach to quantify the amount of biomass in the microbial community in the foulant layer on the membrane surface. After harvesting of individual membrane modules, a 50.8 mm (2 inch) long membrane coupon was collected from the module for biomass analysis. The membrane coupon was then put into UltraLyse 7 (Extraction) tube and a mechanical vortex (Fisher Scientific, catalogue no: 02215365, speed: 10) was used for 3 minutes to separate the foulant layer (containing the microbial community) from the membrane (Appendix A). Biomass analysis was performed as described by the manufacturer LuminUltra (LuminUltra Technologies Ltd., 520 King Street, Fredericton, New Brunswick, Canada, E3B 6G3) and the test kit was ASTM D4012 compliant. An incubation time of 5 minutes was used for the separated fouling layer inside the Extraction tube. A 1 mL aliquot from the extraction tube was transferred to an UltraLute (dilution) tube. The tube was then capped and inverted three times to mix. Then 100 µL from the dilution tube was transferred to a 12x55 test tube (Assay tube). 100 µL of Luminase (enzyme Luciferase) was added to the assay tube and swirled gently for five times. This was immediately inserted into the luminometer (Lumitester C-110 Luminometer) and tATP (Total Adenosine Triphosphate) was recorded using LumiCalc software. The light detected by the luminometer was converted to an electronic signal (Equation 4).

\[
\text{Equation 4} \quad \text{ATP} + \text{O}_2 + \text{Mg}^{2+} + \text{Luciferase} \rightarrow \text{sAMP} + \text{PPi} + \text{oxyluciferin} + \text{light}
\]

\(t\text{ATP}\) represents the accumulation of sessile biomass in the fouling layer including the intracellular and extracellular (living or dead) ATP of a microbial community. Single point calibration (UltraCheck™ 1) was used per day or for each set of samples analyzed at the same time.
4.4.2 Total and Dissolved Organic Carbon (TOC and DOC)

Quantification of organic carbon (total organic carbon, TOC and dissolved organic carbon, DOC) was conducted by UV persulfate method (Standards Methods: 5310C) with instrument Phoenix 8000 by DOHRMANN and run by software TOCTalk/Phoneix 8000, according to their user manuals. TOC and DOC samples were analyzed in a batch every 2 weeks. For (DOC) quantification, 25 mL sample was filtered with a 0.45 µm nylon filter prior to the analysis. Three drops of 10% hydrochloric acid (HCl) solution was added to convert all bicarbonate and carbonate ions to carbon dioxide (CO₂).

Three blanks and single standard solution of 5 ppm were included in each batch of the DOC/TOC analyses. The standard was made from diluting a 1,000 ppm stock solution of 0.5312 grams of potassium hydrogen phthalate (C₈H₅KO₄) preserved with 250 mL of H₃PO₄ to the concentrations of interest with DI water. During the analysis, a total volume of 10 mL was extracted twice from each sample vial for carbon detection. In system calibration was used to calculate the carbon concentration in the samples (Appendix A).

4.4.3 Confocal Laser Scanning Microscopy (CLSM)

To analyze the microbial community qualitatively CLSM was used. Staining process and stock solution preparation were derived from manufacture company Invitrogen (Life Technologies Inc. 5250 Mainway, Burlington, Ontario, Canada L7L 5Z1). SYBR® Gold Nucleic Acid gel (diluted to 1:10,000) stain was used to detect DNA and RNA. Peak excitation and emission wavelength for this stain is 500 nm and 550 nm respectively. Propidium Iodide (PI) (diluted to 1:3000) marker was used to detect the dead cell populations by counterstaining. Peak excitation and emission wavelengths for the stain are 535 nm and 617 nm respectively. The excitation wavelength for both stains are adjacent, so they were not used together on same sample.
Stock solution preparation for SYBR® Gold Nucleic Acid gel.

- Stock SYBR® Gold stain was diluted to 10,000-fold to make a 1X staining solution.
- Dilution was done by adding it into TE (10 mM Tris-HCl, 1 mM EDTA, pH 7.5–8.0), TBE (89 mM Tris base, 89 mM boric acid, 1 mM EDTA, pH 8.0), or TAE (40 mM Tris-acetate, 1 mM EDTA, pH 7.5–8.0) buffer. The stocked solution was preserved at 4º C in brown color bottle, this solution was stable for at least 4 weeks.
- Staining with SYBR® Gold stain is somewhat pH sensitive. For optimal sensitivity, the pH of the staining solution at the temperature used for staining was controlled between 7.0 and 8.5 by the buffer solution.

Stock solution preparation for Propidium Iodide (PI):

- To make a stock solution from the solid form, PI (MW = 668.4) was dissolved in deionized water (DI) at 1 mg/mL (1.5 mM). The stocked solution was preserved at 4º C in brown color bottle, this solution was stable for at least six months.
- Stock solution for counterstaining was prepared by equilibrating the sample briefly in 2X SSC (0.3 M NaCl, 0.03 M sodium citrate, pH 7.0). A 500 nM solution of PI was prepared by diluting the 1 mg/mL (1.5 mM) stock solution 1:3000 in 2X SSC.

For staining, a 40 mm long fouled membrane was used from a harvested module. The sample membrane was placed into a petri dish and stained three times with the particular stock solution for staining. An incubation time of 10 minutes for SYBR® Gold Nucleic Acid gel or 5 minutes for PI was used between each staining. Excess staining was washed off using DI water for 3 times and kept in the dark before imaging. Care was taken not to disturb the fouling during the staining and washing processes. The sample was then fixed in a petri dish using 1% agar solution (Figure 15). Both staining and imaging were done within 2 hours after harvesting.
CLSM images of the stained membrane samples were acquired using an Olympus FV1000-MPE Multiphoton Microscope. For CLSM, the system was equipped with 5 continuous wave (CW) lasers at: 405 nm, 440 nm, 473 nm, 559 nm, 635 nm. For CLSM applications, the system offers fluorescence spectral analysis via the grating gated detectors. Two objectives were used during scanning. The 10X Air objective (model no: UPLSAPO10X2, numerical aperture: 0.95 and working distance: 3.1 mm) was used to focus the sample area that needed to be scanned. The 40X Water Dipping objective (model no: LUMPLFLN40XW, numerical aperture: 0.8 and working distance: 3.3 mm) was used for CLSM imaging. Prior to imaging with the 40X Water Dipping objective the samples were submerged with DI water.

For optical sectioning using CLSM, a vertical interval of 5 µm was used with 400x400 pixel rate. The aperture time for individual capture was no greater that 10 seconds to avoid image bleaching. The direction of lens was along Z axis and from the bottom to the top to reduce mechanical jittering. No refractive corrections were applied for agar solution. The acquired images were analyzed using Olympus FV version 4.1 software.
To measure the fouling layer thickness, optical sectioning data were used. During imaging, the first and the last image with microbial community were identified. Vertical positions of lens were recorded for these two positions and the distance between these two positions was the fouling layer thickness. From the imaging the microbial community also can be classified based on three basic shapes: Coccus (spherical), Bacillus (rod-shaped) and Spiral (twisted).

### 4.4.4 Data Analysis for Permeability

Performance of individual systems in terms of permeability was presented through plotting the normalized permeability ($B/B_i$) (unitless) over filtered volume or time. $B$ is the membrane permeability (m) at any time or volume and $B_i$ is the initial membrane permeability after cleaning. Equation 2 was used to calculate the membrane permeability for constant hydrostatic pressure and $B_i$ was calculated during filtering RO water.

The steady-state of the normalized permeability over volume filtered was calculated through curve fitting using Microsoft Excel Student Edition. The steady-state of the normalized permeability over time was calculated by analyzing the permeability values after 3 weeks of operation. This steady-state over time was defined by the time period during which the first order derivative of normalized permeability over time was $<$0.01. From the data analysis, all systems with active microbial growth reached steady-state after 3 weeks of filtration. A confidence level of 95% was used to report the steady-state permeate flux (L/m²·h) over volume filtered.

No pairwise comparison was done since different experiments were conducted over different time periods which affected the feed characteristics significantly (Peter-Varbanets et al. 2011), nonetheless overall trends can be assessed.
5 RESULTS, DATA ANALYSIS AND DISCUSSION

The present chapter is divided into five sections and each section discusses: 1) the microbial community observed in a gravity driven passive membrane filtration under different hydrodynamic conditions and 2) the optimization of gravity driven passive membrane filtration in terms of permeability.

5.1 Conditions with No Purging and No Air Sparging

The characteristics of the microbial community during gravity driven passive membrane filtration with no purging and no air sparging were assessed through a comparative study of two systems: one with (R1) and one (R2) without a microbial community (Table 2). Both systems were similarly configured except the feed to R2 contained Sodium Azide (NaN₃) (1 gm/L). Figure 16 presents tATP data which indicate that the foulant layer was highly bio-active during the filtration period for the system without Sodium Azide. A mass balance analysis estimated that the accumulated biomass on the membrane surface was higher than the theoretical biomass that can be retained by the membrane (Appendix C). In contrast to the system without Sodium Azide, minimal/marginal biomass was found on the membrane surface in the presence of Sodium Azide. The system with Sodium Azide could only be operated for 7 days because the permeate flow decreased to essentially zero after this period (Figure 16).

![Figure 16 Biomass concentration on membrane surface with and without Sodium Azide](image-url)
An image of the foulant layer is presented in Figure 17a for the system without Sodium Azide. CLSM images (Figure 17b and c) of the fouled layer indicated presence of both active and inactive microorganisms in the fouling layer during filtration for the system without Sodium Azide. Based on the shape of the microorganisms, SYBR® Gold Nucleic Acid gel stain revealed that (Figure 17c) Cocci and Bacilli bacteria were dominant in the microbial community. No higher order grazers which feed on the microbial communities were observed from the CLSM images for membranes from the system without Sodium Azide. No microbial community was detected in the system with Sodium Azide by CLSM imaging Figure 17d.

Figure 17 Microbial community in fouling layer on membrane surface during gravity driven passive filtration: a) visible brown color foulant layer formed on membrane surface with and without Sodium Azide system, b) PI stain- indicating presence of dead bacterial cells without Sodium Azide, c) SYBR® Gold stain, indicating presence of all bacterial cells without Sodium Azide and d) SYBR® Gold stain, indicating presence of no bacterial cells with Sodium Azide
Optical cross sections indicated a relatively consistent thickness of foulant layer containing microbial community in the system without Sodium Azide and the thickness varied over time (Figure 18a). Variation in this foulant layer thickness was also reported by Peter-Varbanets et al. (2010) for flat sheet UF membranes which were also gravity driven passive membrane filtration systems. In their study, an increase to 27 µm in foulant layer thickness was observed within the first 9 days of operation and the thickness reached a steady-state condition. This trend was not observed in the present study, as the thickness of the fouling layer reached a maximum thickness (210.6 µm) at 36 days of operation and gradually reduced to 170 µm by 50 days of operation. This can be explained possibly due to different feed type, hydrodynamic conditions and swelling of the fouling layer during staining. During harvesting of the membrane modules it was observed that the foulant layer tended to readily slough off from the membrane surface (Figure 18b). This suggests that flow interruption/relaxation, which removes the driving pressure required for filtration, could be an effective means of membrane surface cleaning (see Section 5.4).

![Graph](image)

**Figure 18** Characteristics of the foulant layer containing a microbial community in the systems with and without Sodium Azide; a) thickness of foulant layer over the filtration period and b) fouling layer sloughing off

The contribution of the active microbial community to fouling control during passive filtration was assessed through a comparative study of three systems: one with feed water (R1’), one with feed water containing Sodium Azide (R2’) (1 gm/L) and one with RO water containing Sodium Azide (R0)
While no fouling was observed in Figure 19 comparison test for RO filtration with Sodium Azide, an extensive fouling was observed in the filtration of feed water with Sodium Azide compared to filtration of feed water without Sodium Azide. These results indicate that an active microbial community reduces the extent of fouling. Possibly, the active microbial community (Appendix C) present in the fouling layer created a porous structure which increased the permeability of the fouled membrane as suggested by Peter-Varbanets et al. (2011) and Derlon et al. (2014).

Figure 19 Normalized permeability under different microbial growth conditions over volume filtered

5.2 Conditions with Purging or Continuous Air Sparging

The contribution of air sparging and daily purging on the growth of a microbial community during gravity driven passive filtration was assessed through a comparative study of three systems: one with no purging and no air sparging (R1), one with daily purging (R3) (10% of operating volume) and one with continuous air sparging (R4). Figure 20 presents the biomass concentrations on the membrane surface under the different hydrodynamic conditions considered. The results indicate that a microbial community was present on the foulant layer that was formed on the membrane surface over the filtration period for all systems. Within approximately 3 weeks of operation, the biomass concentration reached
a steady-state for all systems in spite of the changes in the hydrodynamic conditions. Possibly, the time to reach the steady-state for a microbial community was not affected by hydrodynamic conditions but affects the biomass concentration. A higher biomass content was observed for the system with purging (6.02E8±1.11E8 pg/m²) compared to the system with no purging (7.48E7±4.97E7 pg/m²) during first 3 weeks of operation. It is possible that purging helped to bring new substrate to the microbial community in the system with purging, which resulted in a reduction of the stabilization period for the microbial community or simply a higher biomass concentration in the feed water. A similar trend but of different magnitude of active biomass concentration was reported by Peter-Varbanets et al. (2010) for gravity driven flat sheet UF membranes after 3 weeks of operation (approximately 2E8 pg/m²). At steady-state of the biomass, the rate of sloughing of biomass off the membrane was likely equal to the rate of growth of new biomass (3.57E8±8.48E7 pg/m² in system with no purging and no air sparging and 2.72E8±4.14E7 pg/m² in system with purging). In contrast, a steady-state in the system with continuous air sparging was achieved due to air bubbles continuously and effectively cleaning the fouling layer off the membrane surface (Figure 21a). No excessive biomass accumulation was observed during the study. This indicates that the microbial community falling off or released from the membrane surface during the filtration period.

Figure 20 Biomass concentration on membrane surface with different hydrodynamic condition
An inactive microbial community was detected on the membrane surface from CLSM imaging on system with purging (Figure 21b). Imaging for an active microbial community in the system with purging was not possible due to its fragile nature. No strong bio-active signal was observed for membrane from continuous air sparging during CLSM imaging (Figure 17d), possibly as a result of the air bubbles continuously and effectively cleaning the fouling layer of the membrane surface. The optical cross sections indicated that the thickness of the fouling layer on the membrane surface containing a microbial community in the system with purging varied similarly over time as that of the system with no purging and no air sparging (Figure 21c). Both systems reached a maximum thickness value and decreased after that. A greater thickness in the fouling was observed for the system with purging possibly due to differences in feed and or purging that help to bring new substrate to the microbial community.

Figure 21 Fouling layer containing microbial community on membrane surface with different hydrodynamic conditions; a) visible fouling layer on membrane with purging, b) PI stain-indicating presence of dead bacterial cells in CLSM imaging and c) thickness of fouling layer
Systems with no air sparging (both R1 and R3) exhibited a similar and sudden decline in the normalized permeability (~80%) compared to the system with continuous air sparging (R4) (~30%) during the first three weeks of operation (Figure 22). The low drop in permeability in the system with continuous air sparging indicated that the contribution of air sparging in fouling control is substantial. Similar steady-state permeate flux were observed for both conditions with no purging (0.223±0.024 B_s/B_i) and with purging (0.183±0.019 B_s/B_i) systems after 3 weeks of operation (Figure 22) (Table 6). For systems with no purging and with purging, the steady-state permeate flux condition was sustained during the remaining filtration period without any fouling control. Purging was not beneficial to improve permeate fluxes but helped to remove settleable solids and improve reactor conditions.

A similar steady-state permeate flux was also observed in previous studies (Peter-Varbanets et al. 2010; Oka 2015) where steady-state permeability was able to be sustained at approximately 0.2~0.25 of the initial value without air sparging and/or daily purging. The decline in the normalized permeability with respect to the volume filtered could be modeled using the exponential relationship presented in Equation 5 (Oka 2015).

\[
\frac{B}{B_i} = \frac{B_s}{B_i} + \left( \frac{B_s}{B_i} \right) \exp(-KV)
\]

where, \( \frac{B}{B_i} \) was normalized permeability at given volume (unitless), \( \frac{B_s}{B_i} \) was normalized membrane permeability at steady-state (unit less), \( \frac{\Delta B}{B_i} \) was change in normalized permeability prior to reaching steady-state (unit less), K was fouling coefficient (1/L) and V was volume filtered (L).
Figure 22 Normalized permeability under different hydrodynamic conditions over volume filtered

The fitted $B/B_i$ values were equivalent to 2.23 L/m²·h and 1.83 L/m²·h (Table 6) for the system with no purging and no air sparging and the system with only purging respectively. These $B/B_i$ values were similar to those previously reported by Oka (2015) (2.2 L/m²·h), but different than those reported by Peter-Varbanets et al. (2011) (~1.5 L/m²·h). The differences in the $B/B_i$ values are likely due to differences in the organic content of the feed water (Appendix B), the acclimatization period (Oka 2015) and/or differences in the hydrodynamic conditions (Peter-Varbanets et al. 2011) considered in their studies. For the system with continuous air sparging, the steady-state condition was much higher than the systems with no air sparging (0.468±0.019 $B/B_i$), indicating that the contribution of air sparging in membrane cleaning allows the removal of the foulant layer from the membrane surface. During the operation it was also observed that purging helped to remove precipitated solids in the reactor tank.
Table 6 Fitted values of the exponential regression model for purging or continuous air sparging

<table>
<thead>
<tr>
<th>Reactor</th>
<th>Conditions</th>
<th>$\frac{B_s}{B_i}$</th>
<th>Mean</th>
<th>±95%</th>
<th>$\frac{B}{\Delta B}$</th>
<th>Mean</th>
<th>±95%</th>
<th>K</th>
<th>Mean</th>
<th>±95%</th>
</tr>
</thead>
<tbody>
<tr>
<td>R1</td>
<td>no purging and no air sparging</td>
<td>0.223</td>
<td>0.024</td>
<td>0.737</td>
<td>0.220</td>
<td>1.754</td>
<td>0.648</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>R3</td>
<td>purging</td>
<td>0.183</td>
<td>0.019</td>
<td>0.412</td>
<td>0.125</td>
<td>1.407</td>
<td>0.499</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>R4</td>
<td>continuous air sparging</td>
<td>0.468</td>
<td>0.116</td>
<td>0.523</td>
<td>0.111</td>
<td>0.073</td>
<td>0.041</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Overall the results from stage 1 indicate the following.

- Air sparging can effectively remove foulants from the membrane over extended operation, allowing the permeability to remain high,
- Daily purging does not substantially contribute to the removal of foulants from the membrane,
- It was possible to maintain a normalized permeability of 0.2 without any air sparging or purging, and
- The foulant layer easily sloughed off from the membrane when harvested.

5.3 Conditions with Different Air Sparging Frequencies

The contribution of infrequent air sparging to fouling control was assessed through a comparative study of four systems: one with no air sparging (R1), one with continuous air sparging (R4), one with 5 minutes/day air sparging (R5) and one with 5 minutes every 2 days air sparging (R6) in gravity driven passive filtration systems. All systems except the one with continuous air sparging, reached a steady-state permeability (in terms of filtration) condition after 3 weeks of operation (Appendix C).

After 3 weeks of operation, higher permeability was observed for the system with continuous air sparging ($0.468±0.116\ \text{B}_s/\text{B}_i$) than systems with no air sparging ($0.223±0.024\ \text{B}_s/\text{B}_i$) and with periodic air sparging ($0.404±0.020\ \text{B}_s/\text{B}_i$ and $0.377±0.018\ \text{B}_s/\text{B}_i$ for 5 min/ day and 5 min/2 days respectively). After 1 month of operation, the system with continuous air sparging yielded $1.33±0.085$ times higher permeability than systems with periodic air sparging (5 min/ day). Similar permeate flux rate and
fouling coefficient (Figure 7) was observed during the operation period for both 5 min/ day (0.553±0.115 B/Bi) and 5 min/2 days (0.505±0.083 B/Bi) (Equation 5) systems despite different air sparging rates (Figure 23). Both of the systems had B/Bi values higher than those reported by Oka (2015) (25 mins off/5 mins on, 4 hrs off/5 mins on and 4 hrs off/ 30 mins on). This was possibly due to the acclimatization period of the membrane prior to starting the experiment considered by Oka (2015). Without acclimatization, B/Bi values as reported by Oka (2015) (0.50±0.003 B/Bi for 9.8, 0.50±0.006 B/Bi for 10.2 L/m²·h, 0.34±0.002 B/Bi for 10.8 L/m²·h) were similar for those system with periodic air sparging to the B/Bi values observed in the present study. This confirms the benefits of air sparging in passive filtration, where even periodic air sparging can contribute to maintaining a high permeability. An increase in permeability with periodic air sparging can be attributed to a change in the fouling layer structure, possibly a porous fouling layer compared to no air sparging.

Figure 23 Normalized permeability under different air sparging rate over volume filtered
Table 7 Fitted values of the exponential regression model for different purging conditions

<table>
<thead>
<tr>
<th>Reactor</th>
<th>Conditions</th>
<th>$\frac{B_s}{B_i}$</th>
<th>Mean ±95%</th>
<th>$\frac{\Delta B}{B_i}$</th>
<th>Mean ±95%</th>
<th>K</th>
<th>Mean ±95%</th>
</tr>
</thead>
<tbody>
<tr>
<td>R1</td>
<td>no air sparging</td>
<td>0.223</td>
<td>0.024</td>
<td>0.737</td>
<td>0.220</td>
<td>1.754</td>
<td>0.648</td>
</tr>
<tr>
<td>R4</td>
<td>continuous air sparging</td>
<td>0.468</td>
<td>0.116</td>
<td>0.523</td>
<td>0.111</td>
<td>0.073</td>
<td>0.041</td>
</tr>
<tr>
<td>R5</td>
<td>5 min/ day air sparging</td>
<td>0.404</td>
<td>0.020</td>
<td>0.553</td>
<td>0.115</td>
<td>1.382</td>
<td>0.386</td>
</tr>
<tr>
<td>R6</td>
<td>5 min/ 2 days air sparging</td>
<td>0.377</td>
<td>0.018</td>
<td>0.505</td>
<td>0.083</td>
<td>1.088</td>
<td>0.27</td>
</tr>
</tbody>
</table>

A lower amount of biomass was observed on the membrane surface for the system with continuous air sparging (1.72E7±5.84E6 pg/m²) compared to that of the system with periodic air sparging (4.97E8±1.41E8 pg/m² and 5.94±1.65E8 pg/m² for 5 min/ day and 5 min/2 days respectively) during the filtration period (Figure 24). Similar biomass density was observed for systems with no air sparging (3.57E8±8.48E7 pg/m²) or periodic air sparging (3.82E8±9.56E7 pg/m² and 3.90±5.15E7 pg/m² for 5 min/ day and 5 min/2 days respectively) after approximately 3 weeks of operation. This was possibly due to the fact that the shear stress produced by the periodic air sparging was not sufficient to remove certain fouling layer containing microbial community while changing the hydrodynamic conditions of the fouling.

Figure 24 Biomass concentration on membrane surface under different air sparging rates
5.4 Conditions with Different Relaxation Period Prior to Air Sparging

In Section 5.1, it was observed that the biomass can easily slough from the membrane surface during handling. From this it was hypothesized that relaxation could be used to release the accumulated foulant from the membrane surface when combined with air sparging. The contribution of relaxation prior to air sparging was considered for this study during gravity driven passive filtration through a comparative study of four systems: one with no relaxation period (R5), two with daily 1 hr relaxation period (R7 and R7' were duplicates), one with daily 4 hrs relaxation period (R8) and one with daily 8 hrs relaxation period (R9). All daily relaxation periods were followed by a 5 minute air sparging period. Daily purging of 10% volume was also applied in all of the four systems to remove settleable solids in the system tank.

From the experiments it was observed that the 1 hr relaxation (0.541±0.028 B/Bi and 0.502±0.069 B/Bi) prior to purging was more effective to produce higher permeability than 4 hrs (0.444±0.076 B/Bi) and 8 hrs (0.111±0.244 B/Bi) respectively after reaching the steady-state permeability. Both 1 hr and 4 hrs relaxation period yielded higher permeability than no relaxation (0.404±0.020 B/Bi), while 8 hrs relaxation period underperformed. Peter-Varbanets et al. (2012) observed that an increase in relaxation period (permeability from 12 hrs relaxation was higher than permeability from 3 hrs relaxation) enhanced permeability, while flushing improved permeability for the systems with relaxation (permeability from 12 hrs relaxation with flushing was higher than permeability from 12 hrs relaxation). The difference was possibly due to the difference in hydrodynamic conditions (flushing as opposed to air sparging) and feed water characteristics (TOC 5.5-7.6 ppm in contrast to TOC 22.3-35 ppm).
Figure 25 Normalized permeability under different relaxation period over volume filtered

Table 8 Fitted values of the exponential regression model for different relaxation period

<table>
<thead>
<tr>
<th>Reactor</th>
<th>Conditions</th>
<th>$\frac{B_s}{B_i}$ Mean $\pm$95%</th>
<th>$\frac{\Delta B}{B_i}$ Mean $\pm$95%</th>
<th>K Mean $\pm$95%</th>
</tr>
</thead>
<tbody>
<tr>
<td>R5</td>
<td>no relaxation</td>
<td>0.404 $\pm$0.020</td>
<td>0.553 $\pm$0.115</td>
<td>1.382 $\pm$0.386</td>
</tr>
<tr>
<td>R7</td>
<td>1 hr</td>
<td>0.541 $\pm$0.028</td>
<td>0.300 $\pm$0.121</td>
<td>0.882 $\pm$0.560</td>
</tr>
<tr>
<td>R7'</td>
<td>1 hr</td>
<td>0.502 $\pm$0.069</td>
<td>0.481 $\pm$0.249</td>
<td>0.594 $\pm$0.603</td>
</tr>
<tr>
<td>R8</td>
<td>4 hrs</td>
<td>0.444 $\pm$0.076</td>
<td>0.586 $\pm$0.254</td>
<td>0.661 $\pm$0.572</td>
</tr>
<tr>
<td>R9</td>
<td>8 hrs</td>
<td>0.111 $\pm$0.244</td>
<td>0.853 $\pm$0.189</td>
<td>0.290 $\pm$0.360</td>
</tr>
</tbody>
</table>

A higher amount of biomass was observed on the membrane for the systems with one hour relaxation (4.07E8±4.62E7 pg/m² and 3.18E8±9.32E7 pg/m²) compared to other two systems 4 hrs (2.44E7±2.98E6 pg/m²) and 8 hrs (6.95E7±2.97E6 pg/m²). It is possible that extended relaxation reduced the substrate flow towards the microbial community, stressing the organisms in the microbial community, and as a result increasing the fouling. The amount of biomass observed on the systems with 1 hr relaxation period was similar to that with no relaxation (4.97E8±1.34E8 pg/m²).
Figure 26 Biomass concentration on membrane surface under different relaxation period

5.5 Hydrostatic Pressure

The contribution of hydrostatic pressure was considered for this study during gravity driven passive filtration through a comparative study of three systems: two with 19.5 cm hydrostatic pressure and one with 25 cm hydrostatic pressure. R7 and R7’ were operated in duplicate with a hydrostatic pressure of 19.5 cm and R10 was operated with a hydrostatic pressure of 25 cm. Both systems were operated with 1 hr relaxation prior to daily 5 mins air sparging and 10% purging of operating volume.

No significant increase was observed in permeate flux \((0.541\pm0.028 \text{ B}_{r}/\text{B}_{i}, 0.502\pm0.069 \text{ B}_{r}/\text{B}_{i})\) for the system with 19.5 cm hydrostatic pressure and \((0.499\pm0.054 \text{ B}_{r}/\text{B}_{i})\) for the system with 25 cm hydrostatic pressure during the filtration period (Figure 27). Oka (2015) operated three gravity filtration systems \((10 \text{ L/m}^2\cdot\text{h}, 13 \text{ L/m}^2\cdot\text{h} \text{ and } 16 \text{ L/m}^2\cdot\text{h})\) and the results also indicated that the steady-state permeability of 13 L/m²-h and 16 L/m²-h that can be sustained was lower than 10 L/m²-h. An increase in the hydrostatic pressure was not helpful to increase the \(\text{B}_{r}/\text{B}_{i}\) (Table 9).
Table 9 Fitted values of the exponential regression model for different hydrostatic pressure

<table>
<thead>
<tr>
<th>Reactor</th>
<th>Conditions</th>
<th>$\frac{B_s}{B_i}$</th>
<th>Mean ±95%</th>
<th>$\frac{\Delta B}{B_i}$</th>
<th>Mean ±95%</th>
<th>K</th>
<th>Mean ±95%</th>
</tr>
</thead>
<tbody>
<tr>
<td>R7</td>
<td>19.5 cm</td>
<td>0.541</td>
<td>0.028</td>
<td>0.300</td>
<td>0.121</td>
<td>0.882</td>
<td>0.560</td>
</tr>
<tr>
<td>R7'</td>
<td>19.5 cm</td>
<td>0.502</td>
<td>0.069</td>
<td>0.481</td>
<td>0.249</td>
<td>0.594</td>
<td>0.603</td>
</tr>
<tr>
<td>R10</td>
<td>25 cm</td>
<td>0.499</td>
<td>0.054</td>
<td>0.886</td>
<td>0.355</td>
<td>0.872</td>
<td>0.169</td>
</tr>
</tbody>
</table>

Figure 27 Normalized permeability under different hydrostatic pressure over volume filtered

Similarly, biomass concentrations were observed to be insignificantly different for both conditions for the system with 19.5 cm hydrostatic pressure (2.48E8±9.27E7) and for the system with 25 cm hydrostatic pressure (2.86E8±7.20E7).
Overall the results from stage 2 indicate the following.

- No significant change occurred in the biomass concentration after reaching steady-state permeability (3 weeks),
- Daily periodic air sparging increased permeability significantly,
- Relaxation prior to purging helped to remove fouling layer from the membrane surface,
- Extended relaxation was not beneficial, and
- No significant increase in permeate flux was observed with the increase in hydrostatic pressure.

Figure 28 Biomass concentration on membrane surface under different hydrostatic pressure
6 ENGINEERING IMPLICATIONS AND CONCLUSIONS

Analyzing from stage 1 and stage 2 from chapter 5, it can be deduced that continuous air sparging > relaxation prior to periodic sparging > periodic air sparging > no sparging in terms of permeability.

Figure 29 Optimization of permeability

A system with clean water flux of 10 L/m²·h when operated under gravity mode in passive condition with no air sparging and no purging and real feed water will produce 2 L/m²·h water. This system will require no daily maintenance over the filtration period. If daily 5 min air sparging is added, the system will yield 4 L/m²·h filtered water. If a 1 hr relaxation period prior to 5 min air sparging is provided, 5 L/m²·h filtered water will be produced. If continuous air sparging is provided, 6 L/m²·h filtered water will be produced. Increasing the permeability also increases the system complexity (Figure 30) and may increase the manufacturing, operating and maintenance costs.
Conclusions

1. A microbial community played an active role in controlling permeate flux during passive filtration. It helped to minimize the membrane fouling.

2. Different operating conditions affect biomass concentration.

3. Periodic air sparging is beneficial to increase the permeate flux compared to only purging or no air sparging. An increase in permeate flux was observed when 5 mins daily air sparging was added, compared to no air sparging and no purging.

4. Relaxation prior to air sparging increased the permeate flux.

5. Extended relaxation was not beneficial to permeate flux.

6. Steady-state permeate flux was a function of time.
References


Appendix A

Table A1 Temperature correction for density (gm/cm$^3$) of water

<table>
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<th>°C</th>
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Temperature correction for viscosity of water

Equation 6 \[ P_{20°C} = \frac{P_T \mu_{20°C}}{\mu_T} \] (Oka 2015)

Where, \( P_{20°C} \) is the normalized TMP at 20°C, N/m$^2$

\( P_T \) is the obtained TMP at temperature T°C, N/m$^2$

\( \mu_{20°C} \) is the viscosity of water at 20°C, Ns/m$^2$

\( \mu_T \) is the viscosity of water at T°C, Ns/m$^2$

Viscosity of water at T°C was estimated using the three regression models listed below over the range of temperatures observed during the present study (19 to 26°C) (Oka 2015).

Equation 7 \[ \mu_{T_{15-20°C}} = -2.76 \times 10^{-5}T + 1.55 \times 10^{-3} \]

Equation 8 \[ \mu_{T_{20-25°C}} = -2.17 \times 10^{-5}T + 1.44 \times 10^{-3} \]

Equation 9 \[ \mu_{T_{25-30°C}} = -2.00 \times 10^{-5}T + 1.40 \times 10^{-3} \]
Organic Carbon Analysis:

Figure A1: Calibration curve for Phoenix 8000 to measure TOC and DOC

Vortex Mixing:

Optimum mixing time for separating fouling layer containing microbial community from membrane using mechanical vortex mixer.

Figure A2: Vortex mixing time as a function of biomass separation
Flowmeter Calibration:

![Flowmeter Calibration Graph](image)

Figure A3: Calibration curve for air flowmeter (Key Instruments, model no: GS10810AVB)

Pressure Data Logger:

![Pressure Data Logger Graph](image)

Figure A4: Calibration curve for Hobo data logger connected to pressure transducers
Appendix B

Permeate flux (L/m²·h) of individual membrane modules in different membrane reactors and at different location of the membrane connection manifold (A-D, a-d).

**R1**

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**R0, R1' and R2'**

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![Graph A5: Organic content in feed, reactor and permeate for R1, R3 and R4 system](image1)

![Graph A6: Organic content in feed, reactor and permeate for R5, R6 and R7 system](image2)

Figure A5: Organic content in feed, reactor and permeate for R1, R3 and R4 system

Figure A6: Organic content in feed, reactor and permeate for R5, R6 and R7 system
Figure A7: Organic content in feed, reactor and permeate for R7' and R10 system
Appendix C

Mass balance of biomass for R1 system
Assumptions:

- System was completely mixed
- Microbial community was completely retained by the membrane
- Theoretical biomass retained on the membrane surface \( \frac{\sum_{t_i}^{t_f} V_t \times C_t}{A} \)

Where, \( V_t \) was the volume of water filtered (l) between time \( t_i \) and \( t_f \), \( C_t \) was the biomass concentration (pg/L) in the feed water at time \( t \), and \( A \) was the surface area of the membrane (m²).

Figure A8: Comparison of theoretical and actual biomass concentration on membrane surface

Table A2: Presence of microbial community in R1' and R2' system

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<td>Biomass tATP (pg/m²)</td>
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<tr>
<td>25 hrs</td>
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Figure A9: Normalized permeability in system when operated under different air sparging condition over filtration period

Figure A10: Normalized permeability under different air sparging rate over filtration period
Figure A11: Normalized permeability under different relaxation period over filtration period

Figure A12: Normalized permeability under different hydrostatic pressure over filtration period