Optimization of the Woven Fibre-Immersed Membrane Bioreactor (WF-IMBR)

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Master of Engineering in the Department of Chemical Engineering
in the Faculty of Engineering & the Built Environment at
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DECLARATION

I hereby declare that this dissertation is my own work unless stated to the contrary in the text, and that it has not been submitted for a degree to any other University or institution.

[redacted]

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ABSTRACT

In this research, the woven fibre microfiltration (WFMF) fabric which is produced locally in South Africa is used as a membrane material. It is cheaper in price in comparison with the current commercial membrane materials that are in use. The WFMF is also more robust when compared with the commercial membrane materials thus is able to withstand harsh working conditions. From previous studies on the WFMF, it has been shown that it can be used as a membrane material without any compromise to permeate quality. This research seeks to optimize the working conditions of this membrane material (WFMF) with an aim of achieving lower running costs and better anti fouling strategies in comparison to the commercial MBRs.

The objectives and aims of this research was to come up with a MBR system whose running cost is lower than that for the commercial systems, which can be adapted for use in any environment, especially in the hardship regions where its robustness would be an added advantage. The performance of the WFMF submerged MBR was also optimised including antifouling operating regimes.

This study was done in a pilot plant that was set up at Veolia wastewater treatment plant, Durban Metro Southern Works. The feed water for the pilot plant was pumped from the return activated sludge mixing chamber by means of a submersible pump. The MLSS concentration of the feed water was about 12 g/l. The various investigations that were conducted in the course of this research included the effect of spacing between membrane modules, relaxation steps and frequencies, evaluation of aeration rates and evaluation of coarse vs. fine bubbles which were all aimed at optimizing the performance of the immersed WFMF MBR. The permeate was checked for turbidity and COD levels to ensure that they were within the accepted water standards.

From the experiments it is shown that the critical flux increased with an increase in aeration rate which is in concurrence with the literature and a starting flux of 30 LMH

was chosen for the running of the pilot plant for the various experimental runs to be carried out. For the pipe diffuser height effect experimental run, the best results were achieved at a height of 5 cm below the membrane modules and the use of a pipe diffuser gave better results than the use of a disc diffuser. For the membrane module spacing effect the best results were obtained at the smallest possible width i.e. 3.5 mm. The best relaxation step sequence was found to be 9 mins on and 1 min off. COD, turbidity and DO was continuously determined during the course of the experimentation.

Further studies should be done on use of the disc diffuser with increased surface area of aeration holes and also hole sizes of smaller diameters to check on its effectiveness as a means of reducing fouling on the membrane surface.

PREFACE

This project was carried out at the Durban University of Technology (DUT), Department of Chemical Engineering. The pilot scale experiments were conducted at Durban Metro Southern Works, Veolia Waste Water Treatment Plant. This project was supervised by Dr. S. Rathilal and Professor V. L Pillay.

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DEDICATIONS

I would like to dedicate this thesis to my mum who passed on before she saw my full work, my dad for his support despite all the tribulations he faced, my siblings Mercy, Mike, and Rose for your encouragement, my friends, especially those who pushed me daily until I finished this work, my special friend Sylvia for encouraging and supporting me in the course of this work.

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NOMENCLATURE

A	membrane area (m^2)
C	total biomass concentration expressed as total suspended solids
J	permeate flux (LMH)
J_C	critical flux (LMH)
K	cake formation correction factor for cross flow micro-filtration
R	resistance (/m)
R_C	cake resistance (/m)
R_m	membrane resistance (/m)
R_f	fouling resistance (/m)
t	time (hr)
V	permeate volume (L)
V_G	gas superficial velocity (m/s)
V_L	liquid superficial velocity (m/s)
α	specific cake resistance
η	viscosity (kg/m/s)
ηP	dynamic viscosity of permeate (Pa.s)

ABBREVIATIONS

AWT	Advanced Wastewater Treatment
DO	Dissolved Oxygen
DUT	Durban University of Technology
CAGR	Compound Annual Growth Rate
CAS	Conventional Activated Sludge
CIP	Cleaned in Place
CIL	Cleaned in Line
COD	Chemical Oxygen Demand
CSTR	Continuously Stirred Tank Reactor
EPS	Extracellular Polymeric Substances
F/M	Food to Micro-organism ratio
FS	Flat Sheet
HF	Hollow Fibre
HRT	Hydraulic Retention Time
IMBR	immersed membrane bioreactor
MBR	membrane bioreactor
MLSS	Mixed Liquor Suspended Solids
MF	Micro Filtration
MWCO	Molecular Weight Cut Off
NF	Nano Filtration
NTU	Nephelometric turbidity Unit
p.e	persons equivalent

RO	Reverse Osmosis
SMP	Soluble Microbial Products
SMBR	submerged membrane bioreactor
SMU	Submerged Membrane Unit
SND	Simultaneous Nitrification & Denitrification
SRT	Solid Retention Time
TMP	Trans Membrane Pressure
TOC	Total Organic Carbon
TSS	Total Suspended Solids
UF	Ultra Filtration
WFMF	Woven Fibre Micro Filtration
WWTPs	Waste Water Treatment Plants

CHAPTER ONE: INTRODUCTION

1.1 Problem statement

The world is now faced with the problem of change in climate due to the destruction of the natural resources and the quickly growing population. This problem has brought about diminishing water levels which affects availability of drinking water and also water for industrial purposes. Industrial countries have already tackled their hydro climatic vulnerability, and have come far in expanding economic development and improving the quality of life for their populations. Emerging economies, on the other hand, remain hampered by water-related challenges such as flooding, drought, and severe water pollution while the poorest economies in semi-arid climates still remain hostage to water problems and suffer large-scale poverty, disease, and uncurbed population growth (Gleick, 2009).

The World Health Organization (WHO) says that at any time, up to half of humanity has one of the six main diseases -- diarrhoea, schistosomiasis, or trachoma, or infestation with ascaris, guinea worm, or hookworm -- associated with poor drinking water and inadequate sanitation. About 5 million people die each year from poor drinking water, poor sanitation, or a dirty home environment often resulting from water shortage. A WHO and UNICEF report released on 26th August, 2004, reveals that more than 2.6 billion people in the world do not have basic sanitation and more than one billion people still use unsafe drinking water. Of the 1.1 billion people in the world who do not have access to safe drinking water, 300 million are in Africa (Ruphael, 2004). Local water resourcing problems provide sufficient motivation for recycling. Water scarcity is determined by the ratio of total freshwater abstraction to total resources. Water stress occurs when demand for water exceeds the amount available during a certain period or when poor quality restricts the use of available water (Judd, 2011).

Problems arise because the fresh water resources are distributed unevenly over the earth’s surface. There are four major global problems concerning fresh water:

1. shortage of renewable supplies,
2. unequal distribution of supplies,

3. problems of water quality and health, and
4. disastrous effects of unrestrained construction of dams (Gleick, 2004).

The resources in potable water are not extensive. It has become a necessity to treat wastewater in order to reuse it (Gre´lot et al., 2009). Increased public concern for health and the environment, the need to expand existing wastewater treatment plants due to population increases, and increasingly stringent discharge requirements have created a need for innovative technologies that can generate high quality effluent at an affordable cost (Gre´lot et al., 2009, Saikkonen et al., 2004, Radjenovi´c et al., 2008).

Water-management areas in South Africa face a water deficit ecosystem and water resources are already being placed under pressure by various users in the sectors. Available water resources are being affected by decreasing water quality, which, in turn, affects net availability (Naidoo, 2009). A combination of polluted water sources and poor management of dams, sewerage works and treatment plants has led to a situation where South African water supply is under serious threat. The water reserves are limited and the country is among the driest on earth with an average rainfall of about 18 inches a year, which is just above half the world average of 34 inches a year (Brulliard, 2009).

More stringent purification standards, growing public awareness of water issues and shrinking prices have led to a boost in membrane activated sludge technology in recent years (Lahnsteiner et al., 2006, Judd, 2011).

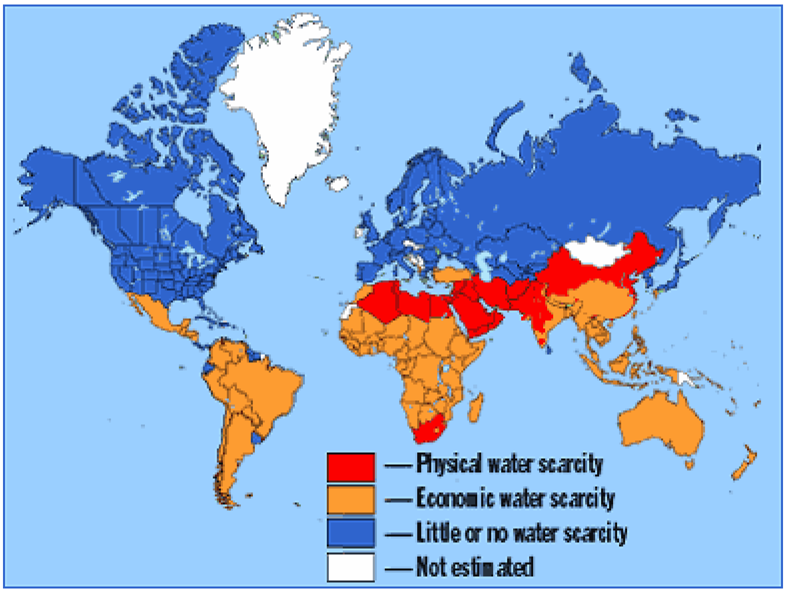


Figure 1.1: Map showing water layout around the world (Judd, 2011)

Wastewater reuse has become increasingly important in water resource management for both economic and environment reasons and has a long history of applications, primarily in agriculture. Additional areas of application include industries, household and urban which are becoming more prevalent. Governments around the world are now laying very stringent standards for treated wastewater effluent in order to enable water recycling. Whilst recycled water is primarily being used for secondary purposes, such as agriculture and industrial use, in some countries like Namibia and Singapore, it is used for human consumption.

One profound reality is that most water treatment technologies are not routinely available in developing countries, especially in rural areas. There is simply not enough money available in rural communities to purchase water treatment technologies and, typically, there is no local resident with the expertise to install and operate such equipment.

There is a need to develop technology that will reclaim wastewater for potable reuse particularly in Southern Africa taking into consideration the water regulating legislatives.

1.2 Background

In the last years a great attention has been addressed towards new technologies to treat wastewater. Legislation and associated regulatory functions exert the greatest influence on the global MBR market, particularly so in the municipal sector. As a consequence of more restrictive legislation, a significant change has been registered in regulation of polluted discharges, specification of both potable and discharge water quality as well as the extent of freshwater resource preservation which influence the choice of water and wastewater treatment technologies (Judd, 2011). Current legislation generally requires an activated sludge process as a major component of a wastewater treatment process. It is widely recognised that a major driver for advancement of municipal water and wastewater treatment technology is legislation, and that two key barriers are cost and perception (Santos et al., 2011, Judd, 2011). Demand for effluents to meet increasingly higher standards is driving optimization of the conventional activated sludge process, which is commonly comprised of a bioreactor and a subsequent clarification process.

An alternative, intensified technology is available with the membrane bioreactor (MBR), which implants the sludge separation process into the bioreactor (Psoch and Schiewer, 2008). In broad terms, a membrane can be defined as an engineered semi-permeable barrier that selectively allows the passage of some materials (e.g. water) while rejecting others (e.g. suspended solids)(Trivedi, 2004). Membrane bioreactors (MBR) are a combination of suspended growth activated sludge process and membrane equipment, where membranes are used for solids/liquid separation which is traditionally accomplished using secondary clarifiers providing an advanced level of organic and suspended solids removal (Fitzgerald, 2008, Trivedi, 2004).

Interest in the MBR technology for wastewater treatment has increased due to increasingly stringent legislation, the opportunity for water reuse/recycling and continuing advancement and decreased costs of membrane technology (Chang and Judd, 2002, Santos et al., 2011, Radjenovi´c et al., 2008, Judd, 2011). MBR technology, which combines a biological-activated sludge process and membrane filtration has become more popular, abundant, and accepted in recent years for the treatment of many types of wastewaters, whereas the conventional activated sludge (CAS) process cannot cope with either composition of wastewater or fluctuations of wastewater flow rate. MBR technology is also used in cases where demand on the quality of effluent exceeds the capability of CAS.

Along with better understanding of emerging contaminants in wastewater, their biodegradability, and with their inclusion in new regulations, MBR may become a necessary upgrade of current existing technology (Activated Sludge Process) in order to fulfil the legal requirements in WWTPs. Although MBR capital and operational costs exceed the costs of conventional process, it seems that the upgrade of conventional process occurs even in cases when conventional treatment works well. This can be related with increase of water price and need for water reuse as well as with more stringent regulations on the effluent quality (Radjenovi´c et al., 2008). The maturing of MBR technology and much wider knowledge of the process, in particular energy optimization and process failure risks, have helped in promoting greater confidence in MBR and also greater willingness to invest in ever larger plants (Judd, 2011). Over the past few decades, the SMBR has been scaled-up from laboratory scale to commercial-scale wastewater treatment technology. Improvements in membrane properties and dramatic reductions in membrane costs make submerged MBRs increasingly competitive with conventional sewage treatment technologies. This technology has proved to have advantages of a small footprint, high removal of COD, effective nitrification/denitrification, less production of excess sludge, and to be a reliable and simple technology to operate (Ndinisa, 2006).

In general, there are two types of membrane equipment: in-pipe cartridge systems that are located external to the bioreactor, and immersed systems that are designed for installation within the bioreactor. Membranes can mainly be configured in four different types of modules i.e. tubular, hollow fibre, flat sheet / plate and frame, and spiral wound. The choice of the configuration is influenced by whether the membrane element is placed within the bioreactor or external to it. Immersed membrane technologies which use hollow fibre or flat sheet membranes are the most popular for MBR applications, since they can more readily accommodate the high concentrations and types of solids found in activated sludge bioreactors (Trivedi, 2004). Membranes can also be classified as either porous or dense. The definition of each type includes a description of the separation mechanism and a cut-off value based on the pore size. Membranes reject material via mechanical separation (sieving) or diffusive mechanisms (osmosis). MF and UF membranes use pressure, also referred to as the TMP, to drive the sieving process whereas NF and RO membranes require pressure to overcome osmosis. MF removes particles down to a size cut-off value of 0.05 microns and at the other end of the spectrum, the RO cut-off is listed in terms of a molecular weight cut-off (MWCO) (Trivedi, 2004). Molecular weight cut-off or MWCO refers to the lowest molecular weight solute (in daltons) in which 90% of the solute is retained by the membrane, or the molecular weight of the molecule (e.g. globular protein) that is 90% retained by the membrane.

1.3 Justification of study

Membrane technology use is growing with new plants being set up mostly due to decreasing capital costs in setting up, decrease in the price of membrane materials and also due to increasingly stringent regulations regarding treatment of effluent of which the MBR is highly suited to deal with. MBR have evolved into a highly efficient wastewater treatment process that is competitive with conventional activated sludge, especially where high effluent quality is required (Hermanowicz, 2011).

The South African MBR market is still in its incipient stages with a few small plants and some that are greater than 1 MLD. Various water authorities in the Western Cape, Eastern Cape and Kwazulu-Natal appear to be embracing MBRs. These are mainly driven by impending water shortage and also due to discharge into environmentally sensitive areas (Judd, 2011). Umgeni Water, in Kwazulu-Natal is currently conducting pilot scale MBR trials using the Toray flat sheet membranes at the Darville Wastewater Treatment Works.

The commercial membranes that are in the market today are still faced with a few problems. In spite of the fact that membrane costs are declining, they still remain relatively high. Another issue is the fact that the membrane materials are very sensitive; cannot be left to dry out thus making storage of the membranes to be a problem, are easily destroyed if/when manhandled by the plant operators which makes them not to be robust. The current commercial MBRs also have high running costs especially in terms of energy due to the levels of aeration that are needed when running. Nevertheless, large progress has been achieved in lowering energy efficiency, primarily through optimization of membrane scouring by aeration in submerged Membrane Bioreactor (sMBR) systems (Hermanowicz, 2011). The MBR also requires skilled personnel to run. Another problem for the MBR is that they are highly susceptible to fouling which affects the performance of the MBR by lowering the flux hence increasing the running costs. Great interest has been shown by researches in trying to minimise the problem of fouling in MBRs.

The next step in development of the MBR would be to develop a membrane bioreactor process that is both robust and efficient for various wastewater applications and with low energy usage in terms of aeration. Research on MBR are on-going in a bid to make it more effective and robust towards the varieties of strength ranges and compounds of wastewater (Mutamim et al., 2013).

IMBRs have major potential applications in South Africa, as well as other developing economies, particularly in terms of point of source treatment of domestic effluents, e.g. small communities, farms, hotels and tourist resorts. The final effluent here could be used directly for agriculture (irrigation), could be used as “grey” water for flushing of toilets, cleaning of grounds, etc. or could be upgraded via reverse osmosis for recycling. If IMBRs are used as point of source treatment for industrial effluents, the product could be used as “grey” water, or could be upgraded for recycling back to the plant (Pillay and Jacobs, 2008).

In this research, the woven fibre microfiltration (WFMF) fabric is used as a membrane material. The WFMF fabric is a material which is manufactured locally in South Africa making it a great membrane material due to its availability and also lower cost in comparison to the commercial membrane materials in use in the market and is also more robust, thus able to withstand harsh working conditions.

Researchers at Durban University of Technology (DUT), Chemical Engineering Department have been using the woven fibre fabric as a membrane material. So far this fabric has shown good promise for use as a membrane material although it has been primarily used for potable water treatment. This research aims to use the WFMF fabric as a membrane material for a submerged flat sheet MBR. The advantages for using WFMF as membrane materials are;

- its relatively lower cost since it is locally produced in South Africa,
- it has a permeate quality which meets WHO drinking water standards and also
- has the ability to withstand harsh treatment i.e. more robust compared to commercial membrane material.

1.4 Research aim

Coming up with a WFMF submerged MBR whose running costs in terms of energy are low, can be easily maintained and easy to run.

1.5 Objectives of research

- Evaluation of the WFMF as a membrane material for commercial use in membrane bioreactors

  Performance comparison with commercial established MBR in terms of

  %COD reduction

  % Solid rejection

- To develop energy reduction strategies in the WF-IMBR

  Effect of module spacing in fouling reduction

  Effect of hydrostatic head between membrane pack and diffuser

  Effect of operating sequence

- Evaluation of periodic filtration (use of relaxation steps) on the performance of the WF-IMBR.

1.6 Approach

This research study was done in a pilot plant that was set up at the Veolia wastewater treatment plant, Durban Metro Southern Works. The feed water for the pilot plant was pumped from the return activated sludge mixing chamber by means of a submersible pump where the MLSS concentration of the feed water was about 12 g/l.

The various aspects investigated in the course of this research included the effect of varying spacing between the membrane modules, effect of relaxation step and frequencies, evaluation of aeration rates and evaluation of coarse vs. fine bubbles which were all aimed at optimizing the performance of the immersed WFMF MBR. The MBR plant was also run at the optimal parameters for an extended period of time. The permeate was checked for turbidity and COD levels to ensure that they were within the accepted water standards.

1.7 Structure of Thesis

Chapter 2 focuses on the literature review into MBR, its history and operations also highlighting some of the commercial membranes in the market today.

Chapter 3 focuses on the research methodology that was used in this research.

Chapter 4 covers results and discussions as per the experimental work done.

Chapter 5 covers conclusions and recommendations based on the results of the experiments.

CHAPTER TWO: LITERATURE REVIEW

Membranes are a key part of chemical technology with MBRs being a key technology. The MBR provides promising technology for wastewater treatment and water reclamation (Yamanoi and Kageyama, 2010b, Baker, 2004, Yamanoi and Kageyama, 2010a)).

A membrane can be defined as:

- An engineered semi-permeable barrier that selectively allows the passage of some materials (e.g. water) while rejecting others (e.g. suspended solids) (Trivedi, 2004)
- A material that forms a thin wall capable of selectively resisting the transfer of different constituents of a fluid and thus effecting a separation of the constituents (Visvanathan and Aim, 2003).
- A material that allows some physical or chemical components to pass more readily through it than others with the degree of selectivity being dependant on the membrane pore sizes (Judd, 2011).

There are several factors that further define or classify a membrane type and they include:

- Pore characteristics such as size, orientation, density and surface charge
- Primary separation mechanism (sieving, diffusion)
- Driving force (pressure, osmosis)
- Material of construction (organic, inorganic)

Of the above factors, pore characteristics have the most significant, or direct impact on the functionality of a membrane (Trivedi, 2004).

Membrane bioreactors combine the activated sludge process with a membrane separation unit for biomass retention (Iversen et al., 2008, Sofia et al., 2004).The separation unit for biomass retention comprises bundles or layers of microfiltration or ultra-filtration membranes (Sofia et al., 2004). Membranes reject material via; mechanical separation (sieving) or diffusive mechanisms (osmosis).

MBRs have been used for some time now as a waste water treatment technology which is a process involving membrane filtration combined with biological treatment (Gupta et al., 2008, Judd, 2007, Ozaki and Yamamoto, 2001). The MBR market has lately strongly risen, either promoting independent process lanes or in parallel with CAS systems (Fenu et al., 2010). In 2005 for instance, about 300 references of industrial applications (>20 m^3/d) and about 100 municipal waste water treatment plants (>500 person equivalent) in Europe were listed (Gre´lot et al., 2009).

Membrane configuration is crucial in determining the overall process performance and is usually in respect to the following (Judd, 2011):

- A high membrane area to module bulk ratio (packing density)
- High degree of turbulence for mass transfer promotion on the feed side
- Low energy expenditure per unit volume of product water
- Low cost per unit membrane area
- Design that facilitates easy cleaning
- Design that permits modularization.

There are two main process configurations of biomass rejection MBRs as shown in Fig 2.1 (Delgado et al., 2011, Judd, 2004);

1. Submerged/immersed (iMBR): is the most widely used in municipal wastewater treatment due to lower associated costs of operation
2. Side stream (sMBR) or externally pressurized cross flow MBR

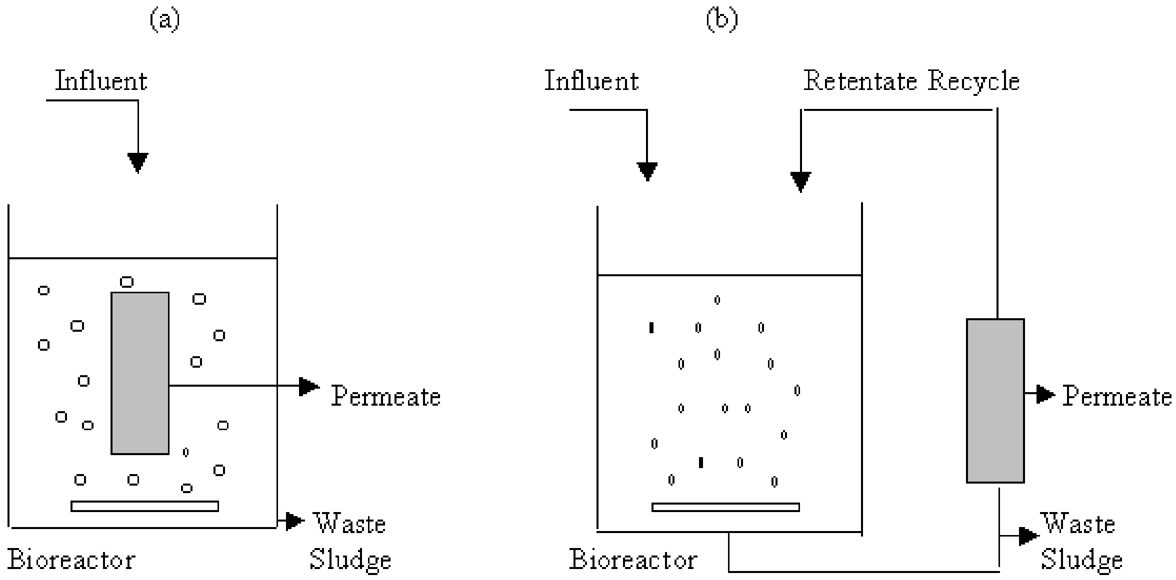


Figure 2.1: (a) Schematic of the immersed membrane and(b) external membrane (Judd, 2004)

In sMBR, membranes are installed outside the bioreactor and mixed liquor from the bioreactor is filtered through them. The mixed liquor is then pumped and pressurized by use of a pump and circulated through the membrane module to the bioreactor. The circulation flow is designed to flow across the membrane surface so as to remove the cake layer. As the mixed liquor circulates, the permeate passes through the membrane and the concentrate is returned to the bioreactor. In the sMBR, the mixed liquor is pumped from the aeration tank to the membrane at flow rates that are 20–30 times the product water flow so as to provide adequate shear for controlling accumulation of solids at the membrane surface (Mohammed et al., 2008). The need for recycling results in high energy consumption making this type of membrane bioreactor less attractive to users and also impractical for full-scale municipal wastewater treatment plants (Mohammed et al., 2008, Sofia et al., 2004).

In the case of a submerged MBR, the membranes are submerged inside the bioreactor and the permeate is suctioned directly by creating a slight vacuum for filtration inside the membrane module which is measured as trans-membrane-pressure (TMP) making this process to be less energy-intensive in comparison to the externally pressurized cross flow MBR (Sofia et al., 2004, Delgado et al., 2011). Upward flow near the membrane surface driven by air bubbles is assigned the role of cross flow to remove the cake layer (Ozaki and Yamamoto, 2001, Mohammed et al., 2008). For the immersed configuration, there are basically two types of commercial membrane modules available i.e. flat sheet (FS), as exemplified by the Kubota technology, and hollow fiber (HF) such as those supplied by GE Zenon or Mitsubishi Rayon (Delgado et al., 2011). Regarding the membrane material used for an iMBR, fluorinated and sulphonated polymers (poly vinylidene difluoride, polyethersulfone) in particular dominate in commercial membrane MBR products (Delgado et al., 2011).

An immersed MBR (iMBR) is superior to an externally pressurized MBR in regard to power consumption and the simplicity of the installation due to the absence of recirculation pumps (Sofia et al., 2004, Radjenovi´c et al., 2008, Ozaki and Yamamoto, 2001). iMBRs are hybrid systems, with the reaction and separation steps located in one unit and the separation steps according to the process intensification principles (Buzatu and Lavric, 2011).

Table 2.1: Advantages and Disadvantages of MBR configurations (Judd, 2004; Radjenov'c et al., 2008)

Submerged membranes systems	External membrane system with high recycling rate and high velocity
• Most recent development (since 1990) • Permeate removed under hydrostatic head; with or without permeate suction, at rate partly determined by aeration. • Aeration cost high (~90%) • Very low pumping costs (higher if suction pump is used (~28%) • Lower flux (larger footprint) • Less frequent cleaning required • Lower operating costs • Higher capital costs	• Longest history (since early 1970) • Pumped system with permeation rate determined by trans membrane pressure and cross flow • Aeration cost low (~20%) • High pumping costs • Higher flux (smaller footprint) • More frequent cleaning required • Higher operating costs • Lower capital costs

2.1 MBR History

The MBR process was introduced in the late 1960s, as soon as commercial scale ultra-filtration (UF) and microfiltration (MF) membranes were available and its driving force is given by a pressure gradient applied between the feed and the permeate sides. Usually pressure values lower than 1 bar are utilised (Guglielmi and Andreottola, 2010).

The original process was introduced by Dorr-Olivier Inc. and combined the use of an activated sludge bioreactor with a cross flow membrane filtration loop. The flat sheet membranes used in this process were polymeric and featured pore size ranging from 0.003 to 0.01µm. Although the idea of replacing the settling tank of the conventional activated sludge process seemed a good one, it was difficult to justify due to the high cost of membranes, low economic value of the product (tertiary effluent) and the potential rapid loss of performance due to fouling. As a result, the focus was on the attainment of high fluxes leading to the need for pumping the MLSS at a high cross flow velocity but at a significant energy penalty (of the order 10 kWh/m^3 product) to reduce fouling. Due to the poor economics of first generation MBRs, they only found applications in niche areas with special needs like isolated trailer parks and ski resorts for example.

Before the 1990s, most installed MBRs were used for industrial water treatment. MBRs did not gain much interest in North America but had considerable success in Japan in the 1970s and 1980s. The MBR breakthrough came when (Yamamoto et al., 1989)proposed a submergence of the membranes into the bioreactor. Until then, MBRs were designed with the separation device located external to the reactor and relied on a high TMP to maintain filtration.

First generation membrane bioreactors were operated with organic or inorganic tubular membranes placed in external recirculation loops. Immersed bioreactors had been developed by Yamamoto et al (1989). in order to simplify their use and to cut down on operating costs (Gre´lot et al., 2009, Radjenovi´c et al., 2008).

The other key steps in the recent MBR development include the acceptance of modest fluxes (25% or less than those of the first generation), and the idea to use two-phase bubbly flow to control fouling. The lower operating cost obtained with the submerged configuration along with the steady decrease in the membrane cost encouraged an exponential increase in MBR plant installations from the mid-1990s.

Since then, further improvements in MBR design and operation have been introduced and incorporated into larger plants.

While early MBRs were operated at a Sludge Retention Time (SRT) as high as 100 days with mixed liquor suspended solids up to 30 g/l, the recent trend is to apply a lower SRT (around 10–20 days), resulting in more manageable mixed liquor suspended solids (MLSS) levels (10–15 g/l). Thanks to these new operating conditions, the fouling propensity in the MBR tends to decrease and overall maintenance has been simplified as less frequent membrane cleaning is necessary.

Currently there is a range of commercially available MBR systems most of which use submerged membranes although some external modules are available; these external systems also use two-phase flow for fouling control. In terms of membrane configurations, mainly hollow fibre and flat sheet membranes are applied for MBR applications (Le-Clech et al., 2006).

There are five principal configurations currently employed in membrane processes all with their inherent practical benefits and limitations. They are based on either a planar or cylindrical geometry and comprise (Stephenson et al., 2000, Radjenovi´c et al., 2008):

- Pleated filter cartridge
- Plate and frame (Flat Sheet-FS)
- Spiral bound
- Tubular
- Hollow fibre (HF)

In the HF module, large amounts of HF membranes make a bundle, and the ends of the fibres are sealed in epoxy block connected with the outside of the housing.

The spiral-bound configuration is mostly used for the NF and RO process. The membranes are wound around a perforated tube through which permeate goes out.

Plate-and-frame membrane modules comprise of FS membranes with separators and/or support membranes. The pieces of these sheets are clamped onto a plate.

The water flows across the membrane and permeate is collected through pipes emerging from the interior of the membrane module in a process that operates under vacuum.

Tubular membranes are encased in pressure vessels and mixed liquor is pumped to them. They are predominantly used for side-stream configurations.

2.2 Membrane separation processes

The membrane filtration is divided into four narrower ranges based on particles size as follows and as shown in Table 2.2 (Judd, 2006, Strathmann, 2001):

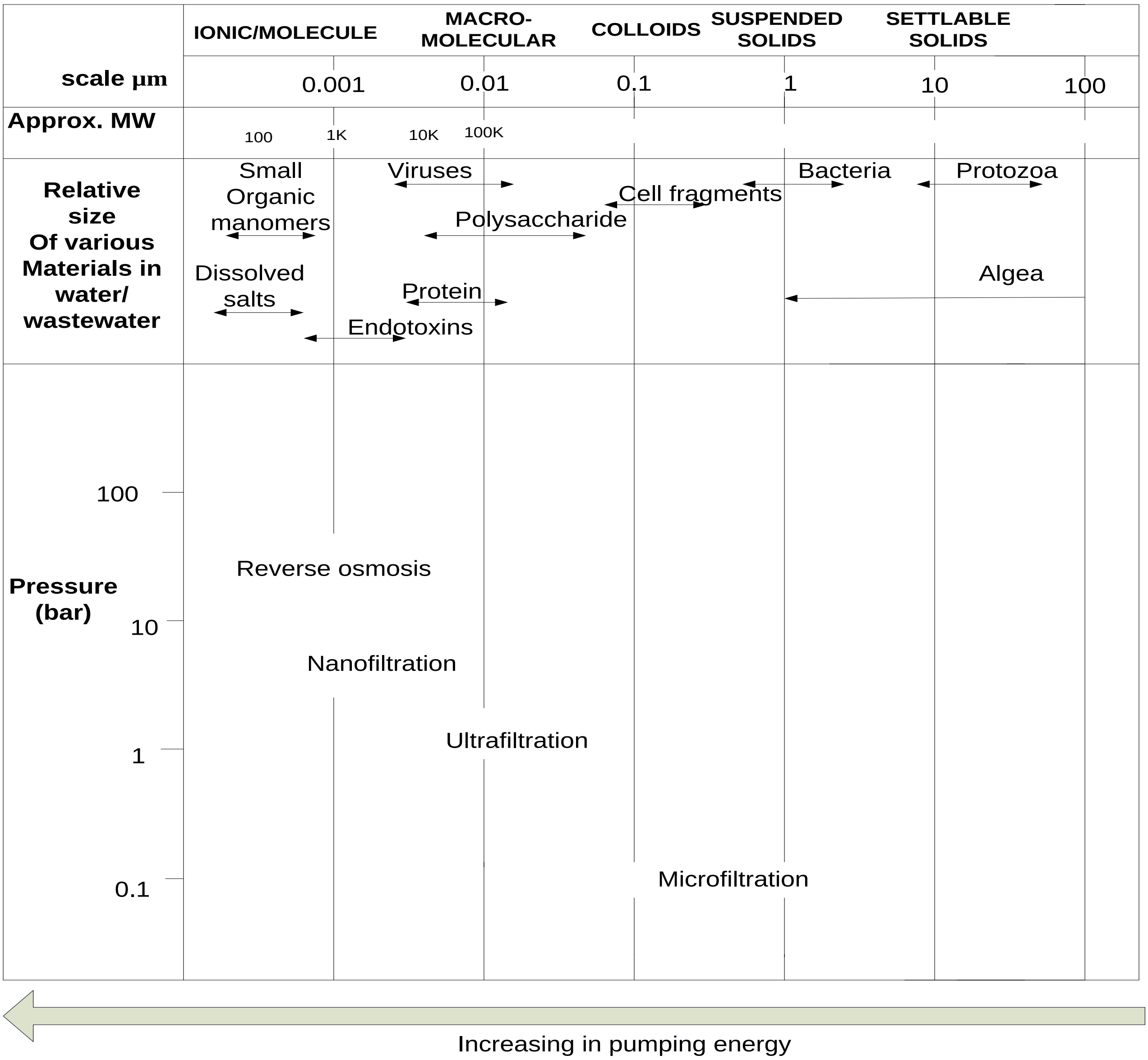
Microfiltration: it is the coarsest size of the membrane filtration classes. Its function is to separate suspended particles from dissolved substances. Microfiltration membranes are classified by pore diameter cut-off (PDCO) which has the diameter of the particles in the range of 0.1 to 10 µm.

Ultra-filtration: is used for separation of large macromolecules such as proteins and starches and all types of microorganisms, such as bacteria and viruses. Ultra-filtration membranes are classified by molecular weight cut-off (MWCO) which is defined as the molecular weight of the smallest molecules. Ultra-filtration covers particles and molecules that range from 1,000 to 500,000 Daltons in molecular weight.

Nano-filtration: membranes retain solute molecules ranging from 100 to 1,000 Daltons in molecular weight. Nano-filtration membranes are classified by molecular weight cut-off like ultra-filtration membranes or by percentage sodium chloride rejection like reverse osmosis membranes. It can also reject contaminants as small 0.001 µ m.

Reverse osmosis: involves the tightest membranes which are capable of separating even the smallest solute molecules or particles with diameter as small as 0.0001 µm. Reverse osmosis membranes are classified by percentage rejection of sodium chloride in an aqueous solution under specified conditions and range from 99-99.5%.

Table 2.2: Classification of membrane and colloidal/micro molecular organic matter in ground and surface water (Judd, 2011)



Microfiltration (MF) and Ultra filtration (UF) membranes use pressure, also referred to as trans membrane pressure (TMP), to drive the sieving process whereas Nano filtration (NF) and Reverse osmosis (RO) membranes require pressure to overcome osmosis. MBRs have proven to be an effective secondary treatment technology with membranes in MF and UF range (Radjenovi´c et al., 2008).

By using micro or ultra-filtration membrane technology (with pore sizes ranging from 0.05 to 0.4 µm), MBR systems allow the complete physical retention of bacterial flocs and virtually all suspended solids within the bioreactor (Yamanoi and Kageyama, 2010b, Le-Clech et al., 2006).

The market value of MBR technology was estimated to be approximately US$217 million in 2005, rising at an average annual growth rate of 10.9%. MBRs are becoming more cost-effective as membrane and membrane process costs continue to decrease and environmental regulations become increasingly more stringent (Judd, 2007). According to (Santos et al., 2011), the market penetration of MBR technology has been reported as growing by an average of 11.6–12.7% per annum since the turn of the millennium , with this rate diminishing marginally towards 2013. As such the MBR market has increased at rates slightly higher than that of the similarly rapidly expanding desalination membrane technologies, for which compound annual growth rates of 9–10% have been observed. In all cases, it must be assumed that the predictions do not account for the economic downturn, which may cause market stagnation over a period of 2–3 years. As with desalination, the market growth in MBR technology is driven by a combination of increasing water scarcity and increasingly stringent legislation. These have tended to promote the more widespread implementation of water reuse technologies (Santos et al., 2011, Zhang et al., 2006).

In Europe, the market of medium and large scale MBR plants (> 5,000 p.e.) for municipal or domestic wastewater treatment is almost entirely shared by two non-European module producers i.e. Zenon Environmental and Kubota Corporation (Lesjean and Luck, 2005). England, Germany, France, Belgium and the Netherlands are the European countries with the highest numbers of full-scale plants for municipal or industrial wastewater treatment (Lesjean and Luck, 2005).

Over the last two decades MBR technology has gained a significant market share of wastewater treatment and is projected to grow at a Compound Annual Growth Rate (CAGR) of 13.2%, higher than that of other advanced technologies and other membrane processes, increasing its market value from $337 million in 2010 to $627million in 2015 (Delgado et al., 2011). According to (Braak et al., 2011a), MBR technology has been experiencing strong growth recently due to the good product water quality with the global market being expected to grow from US$296 million in 2008 to US$488 million by 2013. According to (Cutler, 2011), the global market for MBR systems grew to $838.2 million in 2011 and it is projected to grow up to $3.44 billion by 2018 representing a CAGR of 22.4%. This can be seen as a reaction to global mega trends which are increasing usage of MBRs in the market.

MBRs have been implemented in more than 200 countries. Of interest is the case of China and some of the European countries which have an implementation rate of over 50% and 20%, respectively (Delgado et al., 2011, Judd, 2011). MBR sales have generally grown faster than the GDPs of the countries in which they are installed (especially so in China) as well as more rapidly than the industries that use them (Judd, 2011).

Figure 2.2 shows comparison between conventional activated sludge process and MBR process with MBRs being widely used for wastewater treatment due to it achieving higher effluent quality in comparison to the conventional activated sludge process. The favourable microbiological quality of the effluent in MBRs is a major factor in their frequent selection for water reuse, even though full disinfection cannot be guaranteed, particularly considering the distribution and storage components of a full-scale system, which can be prone to re-growth of microorganisms and contamination from various sources (Melin et al., 2006, Meng et al., 2006).

Membranes are usually made from different plastic and ceramic materials although metallic membranes also exist. The most widely used materials for membranes are celluloses, polyamides, polysulphones, charged polysulphone and other polymeric materials such as polyacrylonitrile (PAN), polyvinylidenedifluoride (PVDF), polyethylsulphone (PES), polyethylene (PE), and polypropylene (PP). All of these polymeric materials have a desirable chemical and physical resistance. They are hydrophobic membranes, known to be more prone to fouling than hydrophilic ones due to the fact that most interactions between the membrane and the foulants are hydrophobic in nature. It is due to this reason that all commercially available membranes are modified by chemical oxidation, organic chemical reaction, plasma treatment, or by grafting in order to achieve a more hydrophilic surface. The modification process and membrane module fabrication method usually differs from one membrane manufacturer to another (Radjenovi´c et al., 2008).

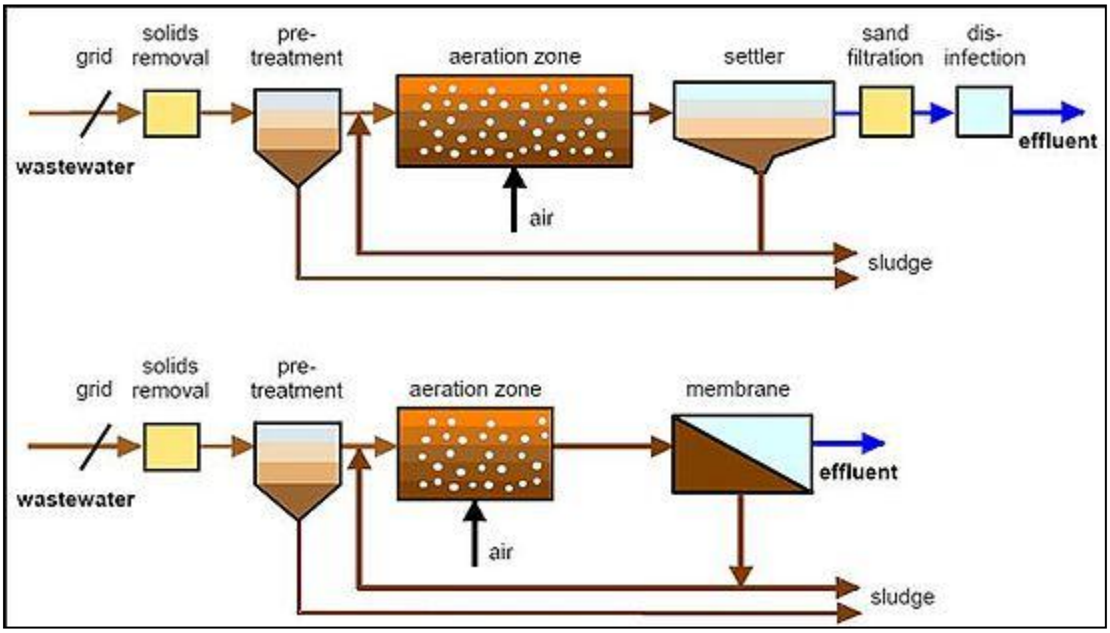


Figure 2.2: Schematic of conventional activated sludge process (top) and membrane bioreactor (bottom) (Atkinson, 2006)

2.3 Commercial Flat Sheet MBR Suppliers/Manufacturers

A review of the share of the municipal market across the MBR membrane product suppliers reveals that it is still dominated by the original three suppliers with Kubota providing around 20-25% of the total number of MBR installations for the top 11 membrane providers and GE Zenon more than 40% of the total global installed capacity for MBR treatment. Mitsubishi Rayon Engineering have an estimated similar number of installations as Kubota though their activities are largely limited to the Far East (Judd, 2011).

2.3.1 Kubota Process

The Kubota Submerged Membrane Unit (SMU) uses flat plate thin-film composite membranes manufactured exclusively for use in wastewater treatment applications. Each sheet of membrane material is made by dipping a non-woven mat of polyethylene terapthalate (PET) into a solution of chlorinated polyethylene and subsequently allowing the wetted mat to dry. During the drying process a thin membrane skin forms over the mat with a nominal pore size of 0.4 microns. Finished sheets of membrane material are cut to size and ultrasonically welded to acrylonitrile butadiene styrene (ABS) panels for mechanical support. This patented ultrasonic welding process is developed by Kubota specifically to provide a fail-safe triple bond between the ABS panel and the membrane itself. Inserted between the membrane and the panel is a polyester spacer material that serves as a plenum to evenly distribute the permeate flow to channels cut into the panel. Each of the channels, or grooves, terminates at a nozzle on top of the panel. The finished product is referred to as a membrane cartridge (as shown in Figure 2.3).

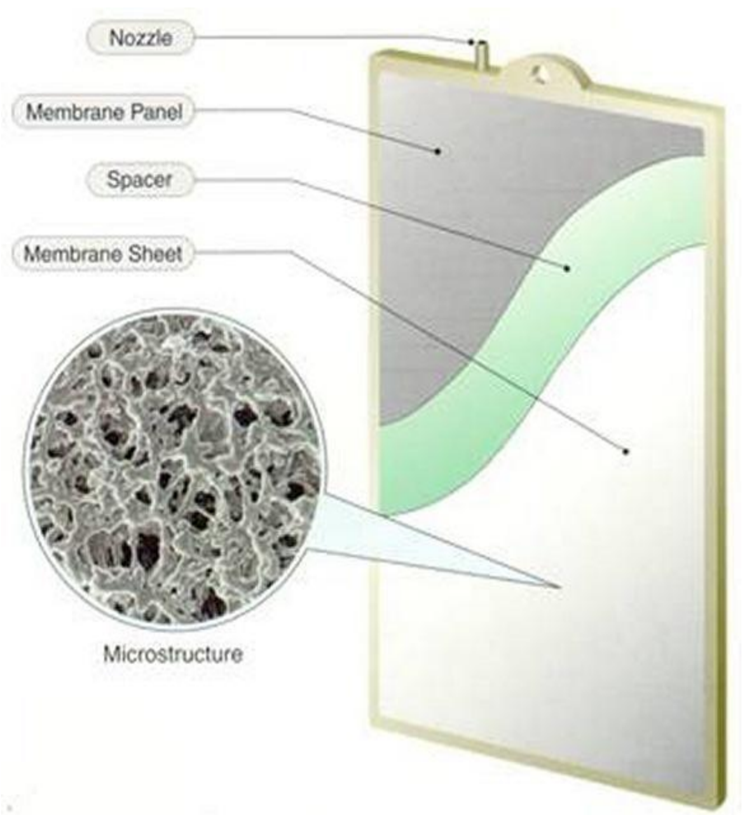


Figure 2.3: A Kubota Membrane Cartridge.

The membrane panels are slotted into the top section of the unit allowing a 7 mm gap between panels to minimise clogging. Kubota successfully pioneered the flat sheet concept in which air is injected from a distributor below a stack of panels. The air then rises through the rectangular channels formed between adjacent pairs of panels. In use, the aerated sludge rises up between the panels and causes recirculation of sludge within the tank volume.

Each membrane panel is connected to a permeate collection manifold, and the treated effluent is removed by low pressure suction. The low pressure suction is produced by available hydraulic gravity head or by a pump. The membrane panels consist of an injection moulded plastic flat plate with the membrane welded ultrasonically to both outer surfaces. Each panel has approximately 0.8 m^2 of membrane surface area. Under magnification the membrane itself can be seen to be composed of a non-woven fibrous support material coated with a porous selective layer of pore size 0.2-0.45 µm (Churchouse, 1997).

Typical membrane cartridge dimensions are 19” x 39” x 0.25”. The total filtration area per cartridge, including membranes on the both side of the plate, is approximately 8.6 ft^2. Multiple cartridges (25-200) are inserted into cassettes and the cassettes in turn stacked on top of integral air diffuser cases. A clear plastic tube connects the nozzle of each cartridge to permeate (filtered effluent) manifold that is connected to the permeate header located outside the MBR. The entire assembly, consisting of one or two membrane cassettes and an air diffuser case, is referred to as a submerged membrane unit or SMU. Kubota membranes are designed to handle a wide range of MLSS concentration (e.g., 8,000 – 35,000 mg/L). This feature provides a unique advantage for Enviroquip’s MBR technology in BNR applications over its competition. The engineered configuration of a Kubota SMU, and specifically the fixed vertical orientation of the cartridges, allow the diffused air to uniformly scour each membrane sheet. In normal operation, equilibrium is established within seconds between the material being brought to the membrane surface through filtration and the material that is scoured away by the cross flow of mixed liquor. The result of this equilibrium is commonly referred to as a bio-film.

On average, it is necessary to chemically clean Kubota membranes every three to six months. The SMUs are cleaned in place quickly and efficiently by simply injecting, or pouring, a dilute solution of bleach or oxalic acid (0.5% typical concentration) into an accessible tee on the permeates suction line. The cleaning solution remains inside the cartridges and soaks the membranes for about an hour, and then normal operation is resumed.

2.3.2 Brightwater Engineering

Established in 1990, the company designs, supplies and commissions plants for the treatment of sewage, industrial wastes, water and sludge.

The Brightwater MEMBRIGHT® system is a flat sheet immersed MBR with 150 kDa polyethylsulphone (PES) membranes mounted on a rigid polypropylene (PP) support. The module is 1120 mm in length, 1215 mm in width for a 50 panel unit (715 mm for 25 panel unit) and 1450 mm high. The respective membrane area provided by the two sizes is 46 and 92 m^2, with each square panel, 950 mm in length, providing an area of 1.84 m^2. The membrane spacing is ~9 mm, the panel support spacing being 10 mm and the panels are clamped in place within a stainless steel frame to form the module. The module is fitted with an integral aerator which ensures even distribution of air across the module (Judd, 2006).

2.3.3 Colloide Engineering Systems (CES)

It is a small-to-medium sized enterprise (SME) which provides various treatment technologies for water and wastewater in both the industrial and municipal sectors, concentrating mainly on small-to-medium scale plants.

The CES Sub Snake system is unusual in that the membrane modules are bespoke and fabricated from a continuous 0.04 µm PES membrane sheet which is cut to size and then glued at the edges to form a FS module. The membrane is then wound ‘snake like’ around a purpose built steel or plastic frame comprising a number of rigid vertical poles at each end to make a multiple FS module with a membrane sheet separation of 10 mm. a single tube is inserted into the permeate channel of each FS element for permeate extraction under suction into a common manifold. The maximum depth of the module is dictated by the width of the sheeting and the total membrane area of the module by its overall length (Judd, 2006).

2.3.4 Huber Technology

The Huber VRM® (Vacuum Rotation Membrane) is a moving membrane module which rotates at a frequency of 1-2 rpm. The small shear created by this, combined with scouring of the membrane surface by air from the central coarse bubble aeration, apparently obliviates all cleaning.

The membrane elements comprise a four plate segment of a hexagon or octagon. The membrane material itself is based on 0.038 µm pore size PES material of around 300 µm in thickness. The individual elements are thus relatively small (0.75 m^2 for four parallel plates, hence 0.19 m^2/element) and since each are fitted with a permeate extraction tube, the permeate flow path is relatively short. The plates are 6 mm thick and separated by a 6 mm channel (Judd, 2006).

2.3.5 Toray Industries

The company launched its flat sheet MBR MF membrane product in 2004. The membrane material used is 0.08 µm rated polyvinylidenediflouride (PVDF) with a standard deviation of 0.03 µm. It is reinforced with a polyethylene terephthalate (PET) non-woven fibre and mounted on an ABS support, into which a number of 1-2 mm permeate channels are cut. The element has dimensions of 515 mm by 1608 mm, providing a membrane area of 1.4 m^2, and is 13.5 mm thick including membrane separation of 6-7 mm.

Elements are mounted in a stainless steel frame to form modules ranging from 70 m^2 total membrane area (50 elements) or 140 m^2 (100 elements). A design flux of 33 litres per m^2 is assumed along with maximum TMP of 0.2 bars. The recommended aeration rate is 0.15-0.25 m^3/h for a single deck unit and half these values for the double decked unit (Judd, 2006).

Some examples of operating conditions for MBRs operated with air sparging are given in Tables 2.3 and 2.4 (INSA, 2006).

Table 2.3: Flat sheet modules performance comparison (INSA, 2006)

Membranes	System Capacity (m^3/day)	Flux ($L.m^{-2}.h^{-1}$)	Aeration Conditions ($m^3.m^{-2}.h^{-1}$)
Kubota	1.9 - 13	20 - 33	0.56 – 1.06
Brightwater	1.2	27	1.28
Toray	0.53 – 1.1	21.6 – 25	0.4 - 0.54
Huber	0.11	24	0.35
Colloide	0.29	25	0.5
Submerged plate 0.2-0.45μm (240m^2)	100 (sewage)		0.92

Table 2.4: Hollow Fibre modules performance comparison (INSA, 2006)

Membranes	System Capacity (m^3/day)	Flux ($L.m^{-2}.h^{-1}$)	Aeration Conditions ($m^3.m^{-2}.h^{-1}$)
Zenon	48 - 50	18 – 25	0.29 – 0.4
M. Rayon	0.38	10	0.65
USF Memcor	0.61	16	0.18
Asahi-kasei	0.9	16	0.24
KMS Puron	0.63	25	0.25
Submerged HF (0.2 μm)	24 (product)	13.3	0.94

2.4 MBR advantages & disadvantages

The immersed MBR has developed from a lab scale set-up to a commercially viable wastewater treatment technology with a scale up to greater than 10,000 m^3/day. MBRs have several advantages over the conventional activated sludge systems as shown below (Gre´lot et al., 2009, Fitzgerald, 2008, Ferraris et al., 2009, Yamanoi and Kageyama, 2010a, Mohammed et al., 2008, Radjenovi´c et al., 2008, Cui et al., 2003, Buzatu and Lavric, 2011, Mutamim et al., 2013, Junjun et al., 2010):

Advantages

- Can be operated with a high concentration of MLSS, 10-15 g/l
- Hydraulic retention time (HRT) of 4-8 hours vs. 16-24 hours,

- Reduced process time
- High effluent quality, capable of meeting stringent discharge requirements, opening the door to direct water reuse
- Small footprint, about 25% of conventional plant,
- The possibility of retaining all bacteria and viruses results in a sterile effluent, eliminating extensive disinfection and the corresponding hazards related to disinfection by products
- Less susceptible to upsets due to flow variations
- As the suspended solids are not lost in the clarification step, total separation and control of SRT and HRT is possible.
- Optimum control of the microbial population and flexibility in operation is possible.
- SRT of 15-365 days, can vary based on flow without negative process impact
- Modular expandability
- Capable of meeting advanced wastewater treatment (AWT) standards for nutrient removal
- Less odour
- Ease of operation

Disadvantages

- The main drawback of the MBR technology still remains the capital and operation costs due to use of the membrane filtration aggregates (first sets and replacements), and the high energy requirement resulting from module aeration.
- Quick membrane fouling and subsequent inefficient membrane cleaning significantly impact operation and membrane replacement costs through reduced membrane modules lifespan.
- Loss of permeate during filtration breaks and back flushing.
- Limitations imposed by pressure, temperature, and pH requirements to meet membrane tolerances. Membranes may be sensitive to some chemicals.
- Less efficient oxygen transfer caused by high MLSS concentrations.

- It is also a “high-tech system” requiring qualified and committed staff, clear operational guidelines, and quick reaction in case of any process or system disturbance (Melin et al., 2006, Mohammed et al., 2008).

2.5 MBR market drivers

Capital and production costs of the membrane modules have reduced significantly. Hence, capital costs of newly built MBR plants are comparable to those of a conventional activated sludge (CAS) plant with the additional costs of the filtration system being compensated for by the halved footprint requirement. Continuous endeavour of suppliers to reduce the operation costs have also led to minimised energy, labour and chemical requirements (Lesjean and Luck, 2005, Braak et al., 2011b).

The key drivers for MBR increased use are summarised as below (Judd, 2011, Cutler, 2011);

- Legislation; Clean Water Protection Act (2009) in US, EU Environmental Policy, Comprehensive Working Program on Energy saving and the Emission Elimination in China (2007) etc.
- Local water scarcity
- Return on investment
- Environmental impact
- Public and political acceptance.
- Decrease in price and energy consumption levels in comparison to the earlier models

In the past MBRs may have been disregarded in favour of conventional treatment plants although now where the footprint is limited and a high product water quality is demanded especially for reuse, the MBR is the technology of choice as is shown in the Table 2.5 showing some of the largest MBR plants (Judd, 2011, Yamanoi and Kageyama, 2010b, Yamato et al., 2006, Judd, 2014).

Table 2.5: Sample of largest MBR plants worldwide (Judd, 2014)

Installation	*Location*	*Technology provider*	*Commissioned date*	*PDF (MLD)*	*ADF (MLD)*
Seine Aval	Acheres, France	GEWPT	2016	357	224
Canton WWTP	Ohio, USA	Ovivo USA	2015-2017	333	159
Macau	China	GEWPT	2014	189	137
Brightwater	Washington, USA	GEWPT	2014	186	124
Qinghe	China	OW/MRC	2011	150	150
Busan City	Korea	GEWPT	2012	102	102
Al Ansab	Muscat, Oman	Kubota	2010	77	55
Cleveland Bay	Australia	GEWPT	2007	77	29
Auapolo	Sao Polo, Brazil	Koch Membrane Systems	June 2013	56	56
Sabadell	Spain	Kubota	2009	55	
Peoria	Arizona, USA	GEWPT	2008	58	38
Broad Run WRF	Virginia USA	GEWPT	2008	73	38
Yellow River	Georgia, USA	GEWPT	2011	114	71

Key: PDF: Peak daily flow

ADF: Average daily flow, Mega-litres per day

GEWPT: GE Water and Process Technologies

OW: (Beijing) Origin Water

MRC: Mitsubishi Rayon Corporation

2.6 Nitrification and Denitrification

The adverse environmental impacts associated with ammonia and nitrogen include promotion of eutrophication, toxicity to aquatic organisms and depletion of dissolved oxygen in receiving water bodies due to bacterial oxidation of ammonia to nitrate. Thus, the removal of nitrogen compounds from wastewater is of increasing importance (He et al., 2009).

Compared to conventional sludge systems, MBRs are known to operate at longer sludge residence times (SRT), low food to microorganism ratios (F/M) and reduced hydraulic residence times (HRT). Longer SRTs and higher biomass concentrations tend to result in larger sludge flocs in aeration tanks. Oxygen diffusion limitations are thus hypothesized to form anoxic zones in the core of flocs larger than 40 μm (Acharya et al., 2006). Simultaneous nitrification and denitrification has also been widely explored in the activated sludge processes and MBRs (Acharya et al., 2006).

There are four types of nitrogen common in wastewater (He et al., 2009, Judd, 2011);

- Organic nitrogen
- Ammonia nitrogen
- Nitrite nitrogen
- Nitrate nitrogen

The above constitute total nitrogen content in wastewater although wastewater predominantly contains organic nitrogen and ammonia nitrogen.

Organic nitrogen is converted to ammonia in the first step of the nitrogen cycle. To remove nitrogen from wastewater, ammonia must be oxidised to nitrate (NO_3). This is called nitrification.

The complete formula for the nitrification process (Judd, 2011):

$2NH_4^+ + 3O_2 \rightarrow 2NO_2^- + 4H^+ + 2H_2O$ (nitroso-bacteria)

$2NO_2^- + O_2 \rightarrow 2NO_3^-$ (nitro-bacteria species)

————————————

$NH_4^+ + 2O_2 \rightarrow NO_3^- + 2H^+ + H_2O$

Nitrification is pH-sensitive and declines significantly at pH values below 6.8. At pH 5.8-6 the nitrification rate may be as low as 10-20% of the rate at pH 7.

After nitrification, nitrogen can be removed from the wastewater by reducing the nitrate to nitrogen gas (N_2) which is then released to the atmosphere in a process called denitrification which requires anoxic conditions as well as an organic carbon source to proceed.

Nitrogen removal in most MBRs is generally not as satisfactory in comparison to COD removal. In conventional activated sludge processes, the loss of nitrifier is always unavoidable due to periodical sludge discharge under shorter sludge retention time (SRT). In comparison, the membrane in the MBR systems not only prevents the washout of nitrifiers, but it also results in the accumulation of recalcitrant compounds and microbial metabolic products. This may eventually lead to the differences in nitrogen removal characteristics between MBR and CAS processes (Li et al., 2008, Schmidt et al., 2003).

$$3NO^{-}_{3} + 3H^{+} + 15[H] \longrightarrow 1.5N_2 + 9\ H_2O$$

Currently, total nitrogen (T-N) removal in MBRs is mostly achieved either in a two-stage continuous flow anoxic/oxic system or in an alternating anoxic/oxic system produced by intermittent aeration (Li et al., 2008). This is based on the principle that nitrification requires an aerobic condition for oxidation of ammonia, whereas denitrification occurs under anoxic conditions in the presence of electron donors.

From studies it has been shown that nitrification and denitrification can indeed occur concurrently in the same reactor through simultaneous nitrification and denitrification (SND) (Li et al., 2008). SND is achieved by allowing nitrifiers and denitrifiers (two species of bacteria with different growth environments) to coexist in a single bioreactor.

Compared to nitrogen removal through conventional nitrification and denitrification, SND offers several advantages:

- It eliminates the need for either two separate tanks operated in series or intermittent aeration operation in a single tank, thus continuous effluent output can be achieved with a smaller footprint.
- It utilizes 22–40% less carbon source and reduces sludge yield by 30%.
- Neutral pH level and less demand for alkalinity can be accomplished in the reactor because alkalinity is consumed during nitrification but produced during denitrification.
- Reduced energy requirement due to the reduction in aeration requirement.

2.7 Factors affecting membrane processes

For pressure driven processes, the driving force for the filtration is TMP, a basilar factor indicating the difference between feed and permeate side pressures providing the actual driving force of the membrane process (Franco, 2009).

$$TMP = P_{feed} - P_{permeate}$$

For submerged modules;

$$P_{feed} = \frac{\rho g h_1 + \rho g h_2}{2}$$

Where ρ= density of feed (kg/m^3)

g= acceleration due to gravity (m/s^2)

h_1&h_2= top and bottom module depths respectively

As the trans-membrane pressure increases during the filtration process, the membrane applies a physical resistance where the total resistance is the sum of the resistance of the fouling layer and the resistance offered by the membrane when it is clean (Franco, 2009)

$$R_t = R_m + R_f$$

Where R_t = Total filtration resistance

R_m = Clean membrane resistance

R_f = Fouling resistance

Flux, *J*, is the quantity of material passing through the membrane surface per time per area. It can also be called permeate or filtration velocity and it can be calculated by Darcy´s law (Franco, 2009). Permeate flux *J* ($M^3/M^2.s$) is the ratio between permeate flow (Q, m^3/s) and actual filtering surface area ($A_{membrane,}$ m^2)

$$J = \frac{Q}{A_{membrane}} = \frac{\Delta P}{\eta R_t}$$

Where ΔP= Trans Membrane Pressure (Pa)

η= Viscosity (kg/m.s^2)

R_t= Total filtration resistance [m^{-1}]

Permeability is obtained by dividing flux by the applied TMP. It is expressed as L/m^2.h.Pa or L.m^2.h.bar.

$$\text{Permeability} = \frac{J}{TMP}$$

Both flux and permeability are referred to a standard temperature in order to consider influence of temperature on permeate dynamic viscosity. Assuming permeate as pure water, (Reid et al., 1997)suggests the following expression to describe the temperature effect on the dynamic viscosity.

$$\eta_T = e^{\left(-24.71+\frac{4.209}{T}+4.527* 10^{-2}T+3.4* 10^{-5}T^2\right)}$$

Whereη_T: dynamic viscosity of permeate, (Pas)

T: temperature (K)

Based on a reference temperature of 20°C, the permeate flux can be standardized with

$$J_{20} = \frac{\eta_T J_T}{\eta_{20}}$$

iMBR is strongly capable of resisting shock loadings and variations in inflow turbidity and organic matter content have no effect on their removal efficiencies. The removal of organic pollutants in terms of COD and SS has also been proven to be very high and a good-quality effluent can be achieved during long-term operation. However, how to operate MBR systems efficiently remains a topic of argument because there is a lack of information on the development of their microbial community structure during nitrification (Radjenovi´c et al., 2008).

2.8 Membrane fouling

Membrane fouling continues to be a major hurdle for MBR designers and operators hence limiting their widespread application. Even though MBRs are an already established commercial application (for more than a decade), fouling remains the most crucial problem limiting wider application of membrane filtration (Psoch and Schiewer, 2005, Mutamim et al., 2013, Yamanoi and Kageyama, 2010a, Gre´lot et al., 2009). Membrane fouling in MBRs may be physical, inorganic, organic or biological. Physical fouling refers to the plugging of membrane pores by colloidal species, such that a certain proportion of the membrane surface is effectively occluded. Inorganic and organic fouling usually refer to scalants and macromolecular species respectively (Judd, 2004). Filtration performances of MBRs can be limited by membrane fouling and the aim of most studies about MBR process is to prevent or to limit fouling in order to enhance system performances (INSA, 2006).

The two most significant components of MBR operational costs are membrane replacement and energy demand, both of which are exacerbated by membrane surface fouling and concentration polarization (Verrecht et al., 2008). Fouling is shown by a decrease in permeate flux or an increase in TMP during a membrane process, a consequence of interactions between the membrane and the mixed liquor (components of the activated sludge i.e. feed components, cells, microbial metabolites) and is one of the principal limitations of the MBR process (Radjenovi´c et al., 2008, INSA, 2006).

Membrane fouling is the undesirable deposition and accumulation of microorganisms, colloids, solutes, and cell debris within/on membranes (Menga et al., 2009, INSA, 2006).

Concentration Polarization is the term used to describe solute build up on the membrane surface causing a gradual concentration boundary layer formation near the membrane (polarization layer) where the solute concentration is higher than in the bulk (Guglielmi and Andreottola, 2010, Jie et al., 2012, Radjenovi´c et al., 2008). It is the reversible build-up of dissolved or suspended solute near the surface due to a balance between the convective drag towards and through the membrane, reducing permeation flux. Since liquid velocity within this layer is close to zero, the only mode of mass transport is diffusion, which is significantly slower than convective transport in the bulk solution causing resistance to filtration. The thickness of the boundary layer is dependent on the system's hydrodynamics and can be decreased by promoting the turbulence of liquid flow (Radjenovi´c et al., 2008). At the same time, the hydrodynamic conditions enhance Brownian transport mechanisms which promote a back diffusive flow which physically leads to an additional resistance to the filtration (Guglielmi and Andreottola, 2010).

Membrane fouling occurring in MBRs is attributable to:

- Sludge particle deposition onto the membrane surface
- Adhesion of macromolecules to the membrane surface,
- Pore clogging by small molecules.
- Adsorption of solutes or colloids within/on membranes
- Formation of a cake layer on the membrane surface
- Detachment of foulants attributed mainly to shear forces
- Ageing of the membrane

Sludge particle deposition has been commonly recognized as greatly contributing to membrane fouling (Radjenovi´c et al., 2008, Menga et al., 2009, Gui et al., 2002).
As a measure of fouling, resistance (R), which is inversely related to K (membrane permeability), is often used.
Where R is given by:

$$R = \frac{\Delta P}{\eta J}$$

η = permeate viscosity in (kg.m^{-1} s^{-2}).

Total filtration resistance can be categorised into two (Radjenovi´c et al., 2008):

i. Fouling layer resistance, associated with the filtration mechanism which is dependent on the characteristics of the membrane and filtered solids.
ii. Membrane solution interfacial region resistance, associated with concentration polarization.

Total filtration resistance (R_t) has been divided into three different components i.e.

$$R_t = R_m + R_c + R_f$$

Where:

R_m is the membrane resistance

R_c is the cake resistance

R_f is the fouling resistance.

It is assumed that fouling consists of two separate processes i.e.

i. one being cake fouling caused by suspended particles that form a cake layer on the membrane surface, and
ii. the other type is associated with adsorption of smaller colloid and soluble matter on the membrane surface and in the membrane pores (Radjenovi´c et al., 2008).

According to (Radjenovi´c et al., 2008, Bella et al., 2006), fouling on the membrane occurs in the following manner as illustrated in Figure 2.4:

1. Complete blocking caused by occlusion of pores by the particles with no particle superimposition
2. Intermediate blocking caused by occlusion of pores by particles with particle superimposition
3. Standard blocking where particles smaller than the membrane pore size deposit onto the pore walls thus reducing the pore size
4. Cake filtration where particles larger than the membrane pore size deposit onto the membrane surface

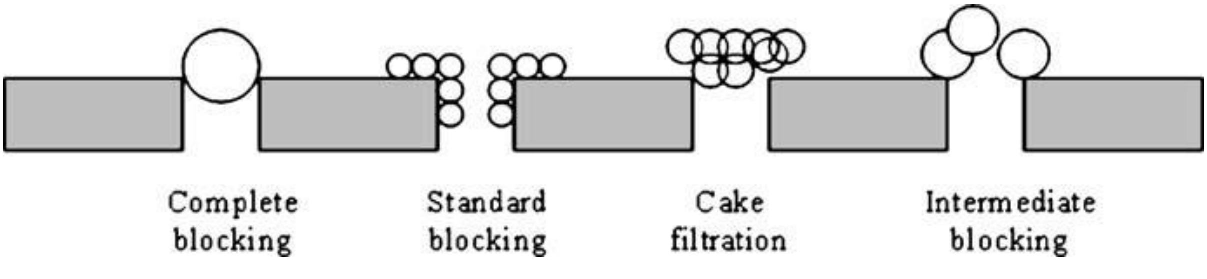


Figure 2.4: Figure showing different types of fouling (Radjenovi´c et al., 2008)

Membrane fouling control remains a major challenge to membrane filtration systems increasing the urgency to minimise their fouling potential and/or develop simple wastewater fouling potential measurement and prediction methods (Johir et al., 2011). This places MBR systems at an economic disadvantage to the CAS process due to high running costs since at least 70% of the MBR total energy demand is used for air scouring as a means of fouling mitigation (Böhm et al., 2012, Prieske et al., 2010).

Majority of commercial submerged MBRs use air scouring to combat membrane fouling. This air is also used by microorganisms in the aerobic biodegradation process. The air bubbles scour the membranes surface generating liquid cross-flow velocity without the need of a recirculation pump. For submerged MBRs energy consumption rates have been reported as ≥1 $kWh.m^{-3}$ with more than 50% of this energy being used for air scouring (Ndinisa et al., 2006a).

Membrane fouling has limited the development and widespread application of MBRs as it leads to filtration flux decline thus increasing the running costs because of the need to clean or replace clogged membranes. Permeate flux decline is influenced by a number of factors i.e. (Chang and Judd, 2002, Bouhabila et al., 2001, Chen et al., 2012, Melin et al., 2006, Yamanoi and Kageyama, 2010b, Ndinisa et al., 2006a, Ndinisa et al., 2006b, Chae et al., 2006, Böhm et al., 2012)

i. the feed wastewater composition (that causes pore blocking, cake formation and development of concentration polarization),
ii. the membrane (element geometry / configuration, area and material composition),
iii. Reactor operation (biological condition and hydrodynamics).

Membrane fouling is significantly influenced by particulate deposition, colloidal or soluble material inside the pores or on the membrane surface, the hydrodynamic conditions, by membrane type and module configuration and also by the presence of higher molecular weight compounds (Melin et al., 2006, Böhm et al., 2012, Chae et al., 2006).

Pollutant particle sizes in wastewater may strongly affect fouling mechanisms in a membrane filtration system. If foulants are of the same size, or smaller than the membrane pores, adsorption and pore blocking may occur. However, if foulants are much larger than the membrane pores, they tend to form a cake layer on the membrane surface.

For the MF membrane process used for wastewater treatment, bio fouling is a major problem because most foulants (microbial flocs) in MBRs are much larger than the membrane pore size. Bio fouling starts with the deposition of individual bacteria on the membrane surface, after which the cells multiply and form a cake layer (Chae et al., 2006).

Fouling can subsequently be characterized according to the nature of the constituent, the mechanism by which it operates, or by the strategy adopted to control it. The fouling constituents could be due to the following combination of chemical and physical interactions (Franco, 2009):

- Particulates (inorganic or organic) can proceed as foulants according to their ability to blind or block the surface
- Organic dissolved components and colloids which can fix on the membrane surface by adsorption.
- Inorganic dissolved components and coagulant residuals which tend to precipitate on the membrane surface.
- Micro-biological organisms, which category covers vegetative matter such as algae and organisms like bacteria which can form colonies, causing bio-fouling

Membrane fouling can be either reversible or irreversible and can be divided into three scales (Johir et al., 2011):

1. Sludge accumulation on the membrane surface, which can be avoided by inducing sufficient local shear stresses (aeration).
2. Irreversible fouling due to physical-chemical interactions (adsorption) of soluble compounds onto the membrane surface which can be cleaned only by chemicals
3. The development of a bio-film due to an accumulation of cells, extracellular polymeric substances (EPS) and soluble microbial product (SMP) on the membrane surface.

In MBRs, the feed to the membrane is activated sludge and membrane fouling is a result of the interaction between the membrane and this complex feed. The characteristics of the mixed liquor are controlled by food to microorganism ratio (F/M) and wastewater composition and this may affect membrane fouling. Activated sludge has many different components i.e. extracellular polymeric substances (EPS), soluble microbial products (SMP) and colloids, which interact with the membrane in different ways (Zhang et al., 2006). Membrane filtration of bacterial suspensions provides a feed of bacterial floc, colloidal species and dissolved macro solutes (EPS), all of which could lead to fouling. EPS is believed to be an important foulant, but the flocculated bacteria, which are usually the dominant species volumetrically, are also considered to be significant (Cho and Fane, 2002). It is widely understood that the EPS generated by micro-organisms are largely responsible for organic fouling of membranes. EPS largely comprise of soluble and colloidal macromolecular species, which can then foul the membrane both at its surface and internally (Judd, 2004).

Fouling is the result of complex phenomena but is essentially caused by the exopolymers produced during lysis of bacteria (Bouhabila et al., 2001). These EPS compounds are mainly proteins, polysaccharides and nucleic acids, lipids and other polymeric compounds. EPS represent the total fraction of biopolymers (the soluble ones (SMP) and those that are bound to flocs) and are normally produced by microorganisms as a construction material necessary for the development of microbial aggregates, such as bio films or flocs, or used as a protective barrier around the bacteria. These EPS and SMP are considered to aid in floc formation and enhance microbial attachment to membrane surfaces. They prevent detachment from the membrane surface by mechanically cross-linking and stabilizing the bio-film. Proteins show high affinities with hydrophobic surfaces whereas carbohydrates are relatively much more hydrophilic than proteins. The EPS and SMP are produced by microorganisms that are released in the liquid phase of the activated sludge which can be quantified by COD and TOC parameters. Biomass growth linked to the substrate consumption leads to the synthesis of EPS (INSA, 2006, n-Vivas et al., 2012, Johir et al., 2011, Chae et al., 2006, Radjenovi´c et al., 2008, Franco, 2009).

Nevertheless, little is known about the circumstances that influence EPS production and their possible release to the water phase. Many operating parameters including substrate composition and organic loading rate appear to affect EPS, with SRT probably being the most significant factor (Radjenovi´c et al., 2008).

EPS and SMP are considered to be the major constituent of irreversible membrane foulants. This requires chemical cleaning of the membrane which ultimately decreases the membrane life time leading to increase in the cost of maintenance (Johir et al., 2011). An important parameter in MBR feed water characteristics is their part of slowly and easily biodegradable organic fractions.

Cake layer formation is another important factor responsible for membrane fouling. It has been reported that sludge cake formation, other than membrane pore clogging, is the predominant cause of fouling in submerged MBRs. The composition of a cake layer in the MBR system usually is very complex including microorganisms, organic and inorganic adsorbed particles, EPS and organic fibres (Chen et al., 2012).

MBR filtration performance decreases with filtration time due to deposition of soluble and particulate materials onto and into the membrane which is attributed to the interactions between activated sludge components and the membrane. Filtration fluxes usually experience regular decline and is dependent on the hydrodynamic conditions, cleaning frequency and methods (Le-Clech et al., 2006, Jie et al., 2012). Membrane filterability and the rise in TMP is associated with the changes in the bio-film structure (Chen et al., 2012).

Comprehension and control of membrane fouling mechanisms has been one of the main preoccupations of the constructors and researchers (Lesjean and Luck, 2005).

Filtration flux has often been reported as the main parameter which controls membrane performance. Fouling mechanisms are usually described in three stages for constant flux operation mode (Braak et al., 2011b, Radjenovi´c et al., 2008):

- A fast but short rise in TMP- conditioning fouling. Strong interactions, among which adsorption, between the membrane surface and colloids, including

EPS, cause initial fouling and pore blockage. This fouling is usually rapid (measured in hours), irreversible by nature, and occurs even for zero flux operation.

- A long period during which TMP increases slightly- slow, steady fouling. The particles settle on the membrane surface and are gradually covered by biopolymers such as EPS, which changes the properties of the membrane surface and makes attachment of the microbial flocs to the membrane surface easier forming a cake layer. The duration of the second step, or sustainability time, depends on the permeate flux
- A very strong rise of TMP- TMP jump. During the previous step permeability is not affected much but fouling is not uniform. Some areas suffer stronger fouling because of flux heterogeneities along the membranes. This is a self-accelerating phenomenon, which leads to exponential fouling. TMP jump could also be induced by a sudden change in the bio-film developed in the membrane.

Depending on the filtration flux, two phenomena can be observed. Beyond the "critical flux", related to system geometry, membrane type, aeration mode and sludge characteristics, a quick accumulation of particles occurs onto the membrane (Lesjean and Luck, 2005).

Fouling increases with increasing flux and, in selecting an appropriate flux for operation, a delicate balance exists between minimising the required membrane area, and so the capital cost, and minimising downtime for backwashing and cleaning, which directly impacts on operating cost (Judd, 2004).

Membrane fouling results in a reduction of permeate flux or an increase in TMP depending on the operation mode which is a result of cake formation on the membrane surface (Menga et al., 2009, Ozaki and Yamamoto, 2001).The cake layer formation on the membrane surface can be removed by hydrodynamic forces. These forces, however, not only affects the deposition layer but might also be responsible for negative effects like SMP release and decrease in floc sizes (Böhm et al., 2012).

To control the fouling that inevitably occurs in MBR operation, several key operational parameters can be modified. The most important strategies are concentration polarization suppression, optimization of physical and chemical cleaning protocols, pre-treatment of feed wastewater, and mixed-liquor modification. Fouling related to concentration polarization can be reduced either by promoting turbulence or by reducing flux (Radjenovi´c et al., 2008).

Most MBR plants operate at relatively modest constant flux as a strategy to slow down membrane fouling rate and hence reduce the frequency of membrane chemical cleaning. MBRs are usually operated at constant permeate flow and successful operation requires identification of a sustainable flux level. To be sustainable, the TMP must not rise rapidly, as this would necessitate frequent cleaning.

Various methods have been adopted to control fouling during the operational cycle of the MBR process, most of which in some way increase the shear rate near the membrane solution interface enhancing mass-transfer (Chang and Judd, 2002). Strategies for sustainable operation include hydrodynamic control by cross-flow (membranes in external loop) or bubbling (membranes submerged in the bioreactor) and ‘sub-critical flux’ operation (Cho and Fane, 2002). The fouling potential of MBR feeds is high. In practice, MBR fluxes are selected with the intention of avoiding rapid fouling and TMP rise (Cho and Fane, 2002). As a consequence, careful fouling management is required for consistent operations of membrane plants (Psoch and Schiewer, 2005).

However for wider application of MBRs the problems that are associated with operational and maintenance costs need to be addressed (Buzatu and Lavric, 2011). Filtration performances can be influenced by strong interactions between the membrane system (membrane structure and shape, module design), the biological fluid and its composition, (which are themselves influenced by the bioprocess operation), and the filtration operating parameters (permeate flux, aeration for fouling control or removal, relaxation, backwashes) (INSA, 2006).

Fouling phenomena on the membrane surface and also within the pores of the membranes reduces the long term flux stability necessitating membrane cleaning which add to the overall cost, as does membrane replacement in cases where cleaning fails to produce adequate flux recovery (Schoeberl et al., 2005).

Factors that contribute to membrane fouling in MBR have been classified into four distinct groups (Delgado et al., 2011, Li et al., 2013)

1. nature of the sludge,
2. operating parameters,
3. membrane/module characteristics and
4. feed wastewater composition (Delgado et al., 2011).

The degree of fouling in an MBR will be determined by three basic fouling factors:

(i) the nature of the feed (to the membrane) (Guglielmi and Andreottola, 2010, Zhang et al., 2006)

Due to their typical charge, proteins play a basilar role in membrane fouling. Such effect is further emphasised by their attitude to interact with other chemical species in the feed (Guglielmi and Andreottola, 2010).

(ii) the membrane properties (Guglielmi and Andreottola, 2010, Zhang et al., 2006)

Roughness of membrane surface is a major factor. Uneven membranes are more susceptible to fouling than homogeneous ones.

Hydrophilicity expresses a membrane's propensity to attract water. A material is defined as hydrophilic if it interacts with water and hydrophobic when it is incompatible with water. When aqueous streams are being filtered, the membrane should be hydrophilic. Hydrophobic material absorbs hydrophobic compounds in the feed solution promoting fouling. However hydrophobic membranes are covered with a thin layer of hydrophilic material in order to make optimum use of both the hydrophobic sturdiness and the hydrophilic low fouling propensity (Guglielmi and Andreottola, 2010).

Membrane pore size and more specifically its relation with particle size in the feed also influences fouling (Guglielmi and Andreottola, 2010).

(iii) the hydrodynamic environment experienced by the membrane (Guglielmi and Andreottola, 2010, Zhang et al., 2006).

Temperature effect on membrane behaviour is not yet completely clear. Under a certain TMP, increases in temperature gives a higher permeate flux as a consequence of lower permeate dynamic viscosity. However if there are salts present (the solubility of which decreases when temperature rises) then higher temperature values can cause a decrease in permeability.

High shear stress induced by feed recirculation or feed sided air-pulsing removes deposited material resulting in reduction of the hydraulic resistance associated with the fouling layer (Guglielmi and Andreottola, 2010).

For iMBR, the operating strategy to control membrane fouling (impacting directly or indirectly on CAPEX and OPEX) includes the following (Delgado et al., 2011):

i. selecting an appropriate permeate flux,
ii. scouring of membrane surface by aeration,
iii. applying physical cleaning techniques, like back flushing (when permeate is used to flush the membrane backwards) and relaxation (when no filtration takes place), and
iv. Applying chemical cleaning protocols, with different frequency and intensity (maintenance cleaning and recovery cleaning).

Previous studies have focused on various factors that affect membrane fouling in MBRs, including membrane permeate flux, aeration intensity, mixed liquor suspended solid (MLSS) concentration, solid retention time and food to microorganism (FM) ratio. In addition to these operational factors, characteristics of mixed liquor are also thought to influence membrane fouling in MBRs (Cui et al., 2003, Yamato et al., 2006).

It is recognized that the hydraulic performance of MBRs may be improved by optimization of the operating conditions. Traditional strategies for fouling prevention mostly try to remedy the effects of fouling by optimization of hydrodynamics and air scouring operation parameters (Iversen et al., 2008, Wu et al., 2008a). Technical solutions are usually implemented such as cyclic or continuous aeration, regular backwash (co-current injection of permeate) or regular relaxation (Lesjean and Luck, 2005).

Researchers have recognized that air sparging offers an opportunity to enhance membrane flux for in-out and out-in filtration (Psoch and Schiewer, 2008, Ndinisa et al., 2006b) as it increases shear stress and reduces fouling compared to non-air sparged operation under similar conditions. Air scouring is used to induce favorable hydrodynamic flow fields in the vicinity of a membrane surface to promote fouling control. Air sparging consists of generating an unsteady flow by injecting air into the concentrate stream. A gas-liquid two-phase flow is then obtained in the feed compartment. The relative motion of the liquid and gaseous phases is thought to create hydrodynamic unsteadiness in the feed channel thus limiting formation of the concentration boundary layer and as a consequence minimizes membrane fouling (Böhm et al., 2012, Ndinisa et al., 2006b).

During the filtration operation a material deposit onto the membrane takes place and causes a fouling which increases with an increase in the permeate flux. Thus, the higher the permeate flux, the higher the material deposit leading to higher membrane fouling with the other operating parameters remaining constant. Moreover, according to Darcy's law, increasing the permeate flux leads to an increase in TMP and due to the cake compressibility, to reduce the material deposit porosity and permeability which leads to a reduction in the filtration performance (INSA, 2006).
From the practical point of view, fouling can be divided into the following:

- Reversible fouling: can be removed from the membrane by physical cleaning
- Irreversible fouling: removed by chemical cleaning
- Irrecoverable fouling: cannot be removed by any cleaning

2.9 Membrane cleaning

Economical operation of microfiltration / ultra-filtration (MF/UF) systems relies on optimization of operating conditions which is normally through fouling minimization

(Zsirai et al., 2012). Various techniques are used to reduce fouling such as reduction of flux, promotion of turbulence to limit the thickness of the boundary layer and/or periodical application of cleaning measures to remove the cake layer and foulants. In case of aerobic MBRs, the air-liquid flow used against fouling is different from the one used for mixed liquor aeration and oxygen mass transfer (INSA, 2006). Physical cleaning in MBRs is normally achieved either by back-flushing or by relaxation (stopping the permeate flow and continuing to scour the membrane with air bubbles). Physical cleaning is a simple and short method (usually lasting less than 2 min) of fouling suppression which does not require the use of chemicals and is less likely to affect the membrane material (Radjenovi´c et al., 2008). Sustainable operation relies on physical cleaning through relaxation or back-flushing, or a combination of both, supplemented with periodic chemical cleaning in place (CIP) (Zsirai et al., 2012, Metzger et al., 2007).

Use of gas-liquid two-phase flow can be very effective in combating fouling as has been revealed in most of the recent studies although it could be energy intensive leading to higher running costs if not operated at an optimum point (Ndinisa et al., 2006a). To maintain an efficient process, frequent cleaning of the membranes is required. Physical or chemical means have been incorporated in most MBR designs as standard operating strategies to limit fouling and also to minimize energy consumption. Relaxation and/or backwashing have been incorporated in most MBR designs as standard operating strategies to limit fouling. Intermittent filtration allows relaxation and reduces compression of the cake layer, thus resistance is reduced and better permeability is maintained. Relaxation, which is the intermittent cessation of permeation permits some flux recovery if the membrane is immersed and scoured with air, such as for iMBRs. Back-flushing is the reversing of permeate back through the membrane, although it is routinely applied to hollow fire (HF) iMBRs (Zsirai et al., 2012, Ndinisa et al., 2006a, Bouhabila et al., 2001, Schoeberl et al., 2005, Wu et al., 2008a).

Cake materials removed by physical cleaning comprise “reversible fouling”. So-called “irreversible fouling” which is thought to relate to membrane pore blocking in the early period of the filtration cycle prior to cake formation is removed only by chemical cleaning. Reversible fouling can become irreversible at overly challenging operating fluxes and hydrodynamic conditions (Zsirai et al., 2012).

For iMBRs, the CIP normally employs a combination of hypochlorite (primarily for removing organic polymers through oxidation), and mineral and/or organic acids (for dislodging scales and metal dioxides through solubilisation). It is thus necessary for MBR operation to rely upon appropriate selection of physical and chemical cleaning protocols. The precise physical and chemical cleaning protocol demanded for sustaining operation is dependent to a large extent on the membrane flux and on the air scouring rates that are selected, and also on the fouling propensity of the mixed liquor (Zsirai et al., 2012). The physical cleaning techniques have been incorporated in most MBR designs as standard operating strategies as ways of limiting fouling (Wu et al., 2008a)

Even if most of MBRs in operation at industrial scale are using air inputs for fouling control, optimisation of air sparging is not a solved problem due to the lack of understanding on mechanisms that are induced by bubbles. Moreover, air sparging or aeration is a key problem for the process as it represents one of the main operating costs. The major reason for this misunderstanding is that each module configuration and membrane systems (which are in constant evolution) present specific flow pattern and thus different transfer phenomena.

Reducing the permeate flux leads to a reduction in fouling, but at the same time this strategy demands installation of more membranes, which then contributes to the increase in capital cost of MBR installation. Flux can be maintained below the critical value to ensure stable operation with little or negligible increase in TMP, thus decreasing cleaning frequency and consumption of chemicals. Alternatively, total installed membrane area can be reduced on the behalf of frequent cleaning. The latter strategy is called intermittent operation. In practice, most submerged MBRs treating municipal wastewater operate at net fluxes of 20–30Lm^{-2} h^{-1} with a relaxation period every 10 min and periodical maintenance chemical cleaning every few months (Radjenovi´c et al., 2008).

2.9.1. Back-Flushing

In most cases, turbulent aeration and backwashing techniques are employed to minimize membrane fouling where the turbulent aeration conditions within an MBR promote scouring of the membrane surface to lower fouling layer formation (Schoeberl et al., 2005). This method disrupts the concentration polarization layer by improving the cross-flow hydrodynamics near the membrane surface (Psoch and Schiewer, 2005). Gas sparging (i.e. injecting gas into the feed of a tubular membrane module to generate a gas liquid two phase cross-flow operation) helps maintain a stable permeate flux over longer time periods. It can even, to a minor degree, reduce internal fouling on the membrane surface due to generation of suction pressure (Psoch and Schiewer, 2005).

The term “back-pulsing” is often employed. It refers to a cyclic process of forward filtration followed by backward filtration which consists in an extremely rapid pulse (pulse duration is generally less than 1 second) and it occurs every few seconds throughout the process whereas backwash interval periods last few dozens of minutes and backwash duration lasts few minutes.

It is a common practice in many iMBRs especially so for hollow fiber membrane systems to additionally backwash the membrane on a periodical basis. By using a reverse TMP for certain periods of time, permeate is forced through the membrane in the reverse direction, which helps to dislodge the cake layer and enhance back diffusion of solute deposits (Schoeberl et al., 2005).

Back-flushing pushes back clear water (in most cases permeate) into the feed stream. It is applied to minimize pore blockage i.e. internal fouling in the deeper layers of the membrane and channel clogging near the membrane surface. During this process, the loosely attached cake detaches from the membrane together with some colloidal and soluble material, such as bound EPS that are entrapped in it. Its influence decreases with growing cake layer thickness, as a result of pressure drop and velocity loss. One major disadvantage of this process is product loss, which can severely decrease the recovery rate at higher flow volumes in reverse direction. Backwashes are mainly efficient for the removal of the accumulated deposit on the membrane surface which mostly constitutes the reversible fouling, whereas pore blocking resistance is not completely eliminated and particularly in the case of high forward filtration fluxes. Following a backwashing, the membrane regains a part of its permeability, and permeation flux or TMP are partially restored. Nevertheless, an irreversible loss of productivity might be observed with filtration time (Psoch and Schiewer, 2006, Zsirai et al., 2012, INSA, 2006).

Optimal backwash duration is one which terminates after the entire removal of the reversible layer. Too short a backwash duration will result in the failure of the complete removal of reversible components. Moreover, the other problem of such a short duration is that any foulants removed from the membrane surface are not sufficiently propelled from the proximity of the fiber bundle so that at the beginning of the following filtration cycle these foulants are very close to the membrane surface and the probability they immediately deposit onto the membrane surface is high which results in a continued TMP increase. A too long backwash duration on the other hand removes effectively the entire reversible layer but in terms of permeate production, an unnecessary additional quantity of permeate is used leading to reduced productivity of the system which also leads to higher energy consumption (INSA, 2006).

The productivity of the system is represented by the following equation:

$$J_{net} = \frac{J_P T_P - J_B T_B}{(T_P + T_B)A}$$

Where J_{net} (L/h.m²) is the permeate productivity of the system, J_p is the permeate flow rate, J_B is the backwash flow rate, T_p is the duration of the permeate production cycle, T_B is the duration of the backwash sequence and A is the surface area of the membrane.

The problem in optimising the backwash interval is similar to that of the backwash duration. Too long a backwash interval involves an inefficient backwashing with regards to the removal of the reversible layer and too short a backwash interval results in a productivity loss in term of permeate production. A long filtration period induces the development of a thick cake layer on the membrane surface leading to the compression of this external layer which becomes increasingly an irreversible foulant layer (INSA, 2006). Most backwashes are operated at higher pressure or flux than forward filtration.

Backwash sequences generally improve the filtration behaviour in an MBR system. For long term filtration they induce a better flux or TMP conservation. Key parameters for a successful backwash operation are backwash duration, backwash interval and backwash flux. The foulant propensity of the biological solution, the permeate flux and the shear stress along the membrane are the factors that determine the operational parameters of the backwash operation.

Several studies show that a combined use of air induced cross flow and backwash is particularly favourable towards fouling suppression. However, proposed optimum conditions vary in a wide range. On the other hand investigations towards simultaneous effect of those operational parameters on fouling have been mainly focused on short term experiments and the unsatisfactory prediction of long term fouling behaviour is still a major obstacle towards development of design concepts for submerged MBR systems (Schoeberl et al., 2005).

For long term filtration, both backwashing and back-pulsing have a positive effect in terms of TMP increasing or permeation flux conservation. It has been reported that backwashing during membrane filtration removes most of the reversible foulants leading to reduced TMP increases and permeate flux decline. So that, backwashing appears to be a key for longer filtration time before intensive physical and/or chemical cleaning. In a general manner, the backwashing frequency, duration and backward flux are important parameters for successful long term operation of filtration. Nevertheless, these parameters are impossible exactly to pre-determine in MBR processes. They depend upon many parameters such as the permeate flux, membrane properties and foulant concentration.

Back-washing sequences are mostly used for HF iMBRs as they are able to withstand the high pressures that are generated by permeate being backwashed during the process. Permeate is usually pumped back at a speed of up to four times the speed of the forward motion.

2.9.2 Relaxation

Application of relaxation sequences during the filtration time (intermittent filtration) in an MBR is another physical method which mitigates membrane fouling, is one of the operating strategies most widely used in a sub-merged membrane for fouling control. In this method, a period of filtration is followed by a period of non-filtration during which the suction pump is switched off and the TMP drops to zero ensuring that the back-transport of poorly attached foulants is enhanced by diffusion from the membrane surface. Intermittent filtration leads to a reduction in the compression of the cake layer, thus resistance is reduced and better permeability is maintained. When the filtration is stopped, the process of gas bubbling which causes shearing on the membrane surface is allowed to continue. This combination makes it easier for the deposited particles to be removed from the membrane surface. When the filtration cycle is resumed, the membrane is relatively clean compared to what it was when the filtration cycle was stopped. Nevertheless, after each relaxation sequence, the membranes permeability is only partially recovered which indicates that irreversible fouling is occurring or that longer relaxation times would be necessary. Thus, pressure relaxation is only able to remove the reversible foulant layer, reversibility being considered at an acceptable time scale. The recovery of permeate flux can be increased if this strategy is combined with air scouring during relaxation periods. However, long and frequent relaxation could cause fouling because of the high instantaneous fluxes needed to mainatain water production. (Ndinisa et al., 2006a, Radjenovi´c et al., 2008, n-Vivas et al., 2012, Braak et al., 2011a).

Deposition on the membrane surface depends on the balance between the velocity toward the membrane surface due to membrane flux and the back transport velocity induced by the shear force. This phenomenon is explained in the Figure 2.5 below. The backward transport of feed particles is mainly related to the cross flow which is strongly dependent on the aeration rate near the membrane and the particle concentration. From conventional brief, long relaxation duration and short relaxation interval are beneficial for fouling control. However, they are not beneficial for retarding TMP increase during permeation. This implies that relaxation duration and intervals should be carefully adjusted to meet the requirements for the two effects (reducing fouling resistance during relaxation and retarding TMP increase during filtration) during both the relaxation and permeation processes. They should be moderate rather than too long or too short. Long and frequent relaxation could cause fouling because of the high instantaneous fluxes needed to maintain water production (Wu et al., 2008a, Braak et al., 2011b).

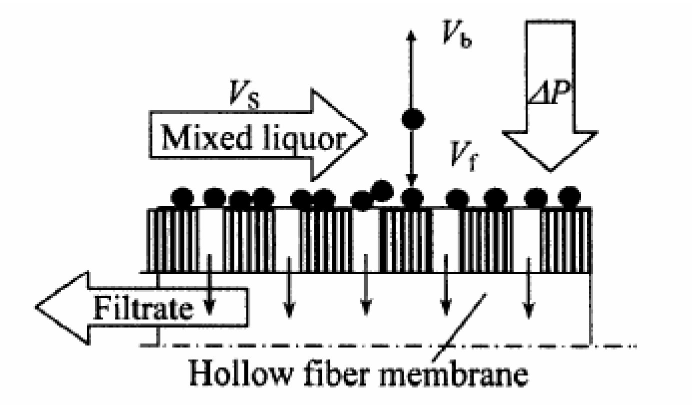


Fig. 2.5. Membrane filtration process, adapted from (INSA, 2006)

Where V_b is the backward transport velocity, V_f is the forward transport velocity and V_s is the cross flow velocity over the membrane surface.

During a filtration operation and when permeation flux is above the critical flux, V_f is higher than V_b and suspended solids accumulate onto the membrane surface, thus constructing the cake layer. Consequently, during relaxation time, V_b is the main phenomenon which makes the particles move and hence back-transport removes reversible foulants from the membrane to the bulk solution. The backward transport is due to concentration gradient in suspended solids between the cake layer and mixed liquor: particles not irreversibly attached to the membrane surface diffused away from the membrane surface thanks to air scouring. It has been reported that in order to maintain certain steady membrane permeability during the filtration operation of a MBR, the cake layer needs to be removed by a shear. In most of MBR systems, it is implemented by an uplifting flow of bubbling air which is supplied by air diffusers located at the bottom of the membrane. This method is called “air scouring”. In this manner, the back transport of reversible foulant during relaxation sequences is strongly enhanced by the air scouring the concentration gradient (INSA, 2006).

(Gui et al., 2002), showed that the TMP increase rate decreases with an increase of non-suction time. For relaxation sequence in the same way as for backwash sequence, there exists an optimal duration after which pressure relaxation has no effect and constitutes a permeate productivity loss. This optimal time occurs after the complete removal of the reversible foulant layer.

On the other hand, the longer the filtration time, the more difficult it is to remove the cake layer. Long filtration time periods are likely to compress the cake layer and hence form an irreversible deposit which accelerates fouling of the MBR. This phenomenon is the same as that which occurs during backwash intervals. From the work of (Hong et al., 2002), in the case of an MBR that is fed with synthetic wastewater, a 15 min non suction time had a limiting suction time of 145 min and when the suction time was prolonged beyond this value the permeate flux was not recovered (INSA, 2006).

From conventional brief, long relaxation duration and short relaxation interval are beneficial for fouling control. However, they are not beneficial for retarding TMP increase during permeation. This implies that relaxation duration and interval should be carefully adjusted to meet the requirements for the two effects (reducing fouling resistance during relaxation and retarding TMP increase during filtration) during both the relaxation and permeation processes. They should be moderate rather than too long or too short (Wu et al., 2008a).

Compared to backwash sequences, relaxation sequences do not involve permeate consumption and hence its use reduces the productivity loss with regard to permeate production. Moreover, relaxation sequences do not require much energy consumption and hence is a more efficient process than backwashing (INSA, 2006, Braak et al., 2011b).

2.10 Membrane flux

Membrane flux, a key parameter in MBR design and application, is closely related to membrane filtration characteristics and it has been reported that it has significant effects on membrane fouling and MBR operation (Wu et al., 2008b). A majority of studies have focused on the concept of critical flux, including the effect of cross-flow velocity (CFV) (which is related to aeration intensity) and sludge concentration on critical flux, and also the fouling characteristics under sub-critical flux operation (Wu et al., 2008b). For most iMBR systems in which the permeate flux is typically in the order of 20 L/(m^2h)–50 L/(m^2 h) for wastewater and drinking water applications respectively, and the bulk liquid velocity at the membrane surface is typically greater than 0.1 m/s (i.e. resulting in a suction rate, defined as the ratio of uniform suction velocity to bulk inlet velocity, of less than 0.01%), the effect of permeation on the hydrodynamic conditions at the membrane surface is expected to be minimal (Böhm et al., 2012). MBRs are usually operated under a constant flux. As the fouling rate increases roughly and exponentially with the flux, MBR plants usually operate at modest fluxes and preferably below the so-called critical flux (Radjenovi´c et al., 2008).

Recovery is normally close to 100% for dead-end filtration while it varies significantly for cross-flow filtration depending on the nature and design of the membrane process. Permeate flux (usually denoted as *J*) is the volume of water passed through a unit area of membrane per unit of time and is often normalized to a standard temperature. The common unit for *J* is usually Lm^{-2} day^{-1}. Most of the available data for MBR is usually given in that manner rather than in SI units. MBRs in most cases operate at fluxes between 10 and 100 Lm^{-2} $h^{-1}$and it is related to its driving force which is TMP or ΔP while the membrane performance can be estimated from the membrane permeability (K), which is calculated as permeate flux per unit of TMP and is usually given as Lm^{-2} h^{-1} bar^{-1} (Radjenovi´c et al., 2008).

Flux selection provides the most significant factor that affects fouling rate. Convective transport of solutes, colloids and suspended matter at higher rates result in a rapid formation of a cake layer as well as internal and/or irreversible fouling phenomena. It is in respect to this that the critical flux concept was introduced. The hypothesis of this concept is that a flux exists, below which a flux decline does not occur (critical flux). Operation below the critical flux is expected to lead to little or even no fouling. At these conditions, convection of material towards the membrane surface is suggested to be outbalanced by back-diffusion or tangential removal of rejected matter. Critical flux is a function of hydraulic conditions, tending to increase with increase in cross-flow. For iMBRs, air induced cross-flow proved to increase the critical flux (Schoeberl et al., 2005, Radjenovi´c et al., 2008).

From the works of (Judd, 2004), it has been implicated that critical flux applies only to suspended solids and has no bearing upon the sustained operation in the presence of dissolved and colloidal macromolecular species, and EPS specifically.

Most MBR plants operate at relatively modest constant flux as a strategy to slow down the membrane-fouling rate reducing the frequency of membrane chemical cleaning. However, due to the complexity of the mixed liquor, some irreversible fouling constantly occurs making it impossible to achieve the sub-critical conditions as for the strong form of the critical flux (Radjenovi´c et al., 2008).

Critical flux (*J*) was originally defined for micro filtration as a flux below which a decline of permeability with time does not occur and above which fouling is observed. However, sub-critical flux fouling in membrane filtration has recently been reported and MBRs operated at sub-critical fluxes appear to be affected by a similar phenomenon over the long term (Pollice et al., 2005, Li et al., 2013). The authors (Pollice et al., 2005, Li et al., 2013) have suggested that critical flux actually represents the boundary between fouling by the dissolved/colloidal components and suspended matter of the biomass. In other words, the critical flux in an MBR can be thought of as marking the transition between constant and non-constant permeability, and hence the terms reversible and irreversible fouling. From the work of (Wang et al., 2008), it has been indicated that a number of researchers even at very low fluxes have experienced an increase of TMP with time which has led to the introduction of a so-called sustainable flux. This represents the flux value at which the fouling rate is operationally and economically acceptable for MBR operation. Flux sustainability is typically assessed by long-term trials in which a flux lower than the short-term critical one, also known as sub-critical flux, is employed and TMP is continuously monitored.

Fouling under sub-critical flux operation is normally attributed to the accumulation of organic macromolecules in the pores and/or on the membrane surface leading to the progressive increase of resistance to filtration. Adsorption is considered to be responsible for the initial rapid and mostly irreversible decline of membrane permeability, which is independent of the hydrodynamic conditions and establishes a gap between clean water filtration and real operating conditions (Pollice et al., 2005).

It is widely accepted that sub-critical fouling in MBRs is mainly caused by organic macromolecules such as soluble microbial products (SMP), extracellular polymeric substances (EPS) and possibly other substances resulting from cell lysis or lost during cell synthesis. Although the precise definition of these groups of compounds is still open to debate, SMP may be identified as cellular components excreted through the cellular membrane due to metabolism under normal or stressed conditions. EPS have been defined as a complex mixture of polysaccharides, proteins, lipids and nucleic acids which form highly hydrated gel or fibrillar matrices in the range of a few nanometres, providing the dominant bridging mechanism between cells in activated sludge (Li et al., 2013, Pollice et al., 2005, Wang et al., 2008).

To suppress fouling of the flat sheet membrane, Ozaki and Yamamoto (2011) investigated the dependency of the sludge accumulation on the variation of hydraulic conditions resulting from the change of the cross flow velocity and clearance between flat membrane sheets. They showed that the scouring effect by bubbles depends on aeration intensity which is defined as air flow rate per cross-sectional area between flat membrane sheets, and can be explained by estimated shear stress (Yamanoi and Kageyama, 2010b). To be sustainable the TMP must not rise rapidly, which would necessitate frequent cleaning. Various methods have been adapted to control fouling during the operational cycle of the MBR process, most of which in some way increase the shear rate near the membrane–solution interface enhancing mass-transfer. Filtration performances can be influenced by strong interactions between the membrane system (membrane structure and shape, module design), the biological fluid and its composition, (which are themselves influenced by the bioprocess operation) and the filtration operating parameters (permeate flux, aeration for fouling control or removal, relaxation, backwashes). Operational and maintenance costs of MBRs are high, mainly due to membrane fouling. Indeed, filtration performances can be limited by membrane fouling and the aim of most studies about MBR process is to prevent or to limit fouling in order to enhance system performances. Membrane fouling deteriorates the permeability of the membrane and consequently increases energy consumption in an MBR. Fouling of membranes in MBR systems can be minimized by:

i. Reduction of flux,
ii. Promotion of turbulence to limit the thickness of the boundary layer
iii. Periodical application of cleaning measures to remove the cake layer and foulants.

2.11 Critical flux evaluation

The combination of membrane filtration with a suspended growth bioreactor is now widely used for municipal and industrial waste treatment. Key to the identification of appropriate operating conditions is the so-called “critical flux” (J_c), a concept originally presented by Field et al (1995) who stated that: “The critical flux hypothesis for microfiltration is that on start-up there exists a flux below which a decline of flux with time does not occur; above it, fouling is observed” (Clech et al., 2003, Wang et al., 2008, Wu et al., 2008b). Since then, the critical flux has been extensively applied to membrane processes from microfiltration (MF) to reverse osmosis (RO).

Membrane flux, a key parameter in MBR design and application, is closely related to membrane filtration characteristics and membrane fouling. It has been reported that membrane flux has significant effects on membrane fouling and MBR operation (Wu et al., 2008b).

Problems related to operational high costs, mainly due to membrane fouling, still need to be addressed. The deposition of biomass and suspended solid on the membrane surface and within the membrane pores leads to an increase of the hydraulic resistance and a permeate flux decline (Bottino et al., 2009). Fouling can be limited, and, consequently, cleaning cycles can be considerably reduced by using air bubble sparging or, as more commonly done, by maintaining the permeate flux below the so-called “critical flux”, J_c, (Bottino et al., 2009, Marel et al., 2009).

The critical flux is a quantitative parameter for the filterability of different membranes and/or different activated sludge mixtures and is generally regarded as the flux above which cake or gel formation by particles or colloids occurs, i.e. convection of these materials towards the membrane by the permeate drag flow exceeds the back transport velocity of material from the membrane induced by shear and Brownian diffusion (Marel et al., 2009).

The sub-critical flux is the flux rapidly established and maintained during the start-up of the filtration, but does not necessarily equate to the clean water flux. Stable filtration operation, i.e. constant permeability (K) for an extended period of time, has been defined as sub-critical operation even when preceded by an initial decline in flux due to solute adsorption (Clech et al., 2003). Given the limitations of applying particle hydrodynamics to the identification of J_c in real systems where both solute adsorption and transient cake behaviour can be important, recourse generally has to be made to experimental determination. By plotting flux against TMP it is possible to observe the transition between constant and non-constant permeability at the onset of fouling. The flux at this transition has been termed "secondary critical flux", but is also defined as the weak form of J_c. Recent studies suggest that this transition marks the boundary between fouling by EPS and sludge solids. Whilst potentially useful in providing a guide value for the appropriate operating flux, the absolute value of J_c obtained by this method is likely to be dependent on the exact method employed and specifically the rate at which the flux is varied with time (Clech et al., 2003).

MBRs are normally operated at a constant flux, preferably below the so called critical flux, as the fouling rate increases roughly and exponentially with the flux, In MBR operations, critical flux is normally defined as the highest flux under which a prolonged filtration with constant permeability is possible (Radjenovi´c et al., 2008).

The critical flux of the WFMF is to be determined by use of the flux step method (Field et al., 1995). The flux is increased in chosen increments and the TMP is observed. For each flux step, the two TMP values are considered. The initial TMP corresponding to the initial sudden increase of filtration resistance and the final TMP, for example the TMP at end of the step. From these two values, the average TMP is evaluated and the flux at which the TMP starts to increase. The critical flux is assumed to be the flux at which sudden increase of TMP is observed and above that point more fouling occurs (Bottino et al., 2009). In the constant flux process, critical flux is often defined as the flux above which TMP starts to increase rapidly with time.

Knowledge of critical flux is important for commercial plants. Operating below critical flux implies that the membrane will be fouled less significantly and this will allow operation for a longer period of time before cleaning may be required. Operating below critical flux may mean that the production rate is low, but on the other hand prolonging the period between cleanings may mean reduced expenditure on cleaning and prolonging the membrane life.

Two distinct forms of the critical flux concept have been defined. In the strong form, the flux obtained during sub-critical flux is equated to the clean water flux obtained under the same conditions. However, clean water fluxes are rarely attained for most real feed waters due to irreversible adsorption of components not removed other than by chemical cleaning. In the alternative weak form, the sub-critical flux is the flux rapidly established and maintained during the start-up of the filtration, but does not necessarily equate to the clean water flux. Alternatively, stable filtration operation, i.e. constant permeability (K) for an extended period of time, has been defined as sub-critical operation even when preceded by an initial decline in flux due to solute adsorption (Clech et al., 2003, Mutamim et al., 2013).

The critical flux usually is determined by flux- or pressure-step methods. Flux-stepping is preferred as the control of the flux is easier. Moreover, at fixed flux a constant flow of foulants towards the membrane is established (Marel et al., 2009). The flux is incrementally increased in a number of steps with fixed duration, and the increase in TMP is recorded. It is then possible to observe the apparent flux where fouling occurs, observed as a significant TMP increase or deviation in linearity of K occurs.This method is preferred over the TMP-step method since it provides better control of the flow of material deposition on the membrane surface, as the convective flow of solute towards the membrane is constant during the run (Clech et al., 2003, Radjenovi´c et al., 2008).

Fouling could happen when the flux is above critical flux. From critical flux, the suitable flux for the operation can be defined based on the TMP sustainability. There is no standard method to find the critical flux due to the difficulties during reporting data. One practical method that can be used is the flux-step method as shown in the

Figure 2.6. This method is relevant for short term critical flux operation and not for long term operation.

There are two concepts of flux, strong and weak. In the strong concept the flux obtained during sub-critical is equal to clean water flux but this concept is not relevant with MBR due to high sludge found in the reactor. In the weak concept, the flux is obtained during operation start-up and is maintained for period of time but is not necessarily equal to clean water flux. The highest flux can be determined when the flux is increased and there is no TMP increment or less permeates. It is shown when fouling is about to happen. Equations 2.1–2.4 are used to define the fouling performance for each flux step (Mutamim et al., 2013).

Initial TMP increase: $\Delta P_0 = TMP_t^n - TMP_f^{n-1}$ 2.1

Rate of TMP increase: $\frac{dP}{dt} = \frac{TMP_f^n - TMP_t^n}{t_f^n - t_t^n}$ 2.2

Average TMP: $\frac{TMP_f^n - TMP_t^n}{2}$ 2.3

Permeability of the system: $K = \frac{1}{P_{avg}}$ 2.4

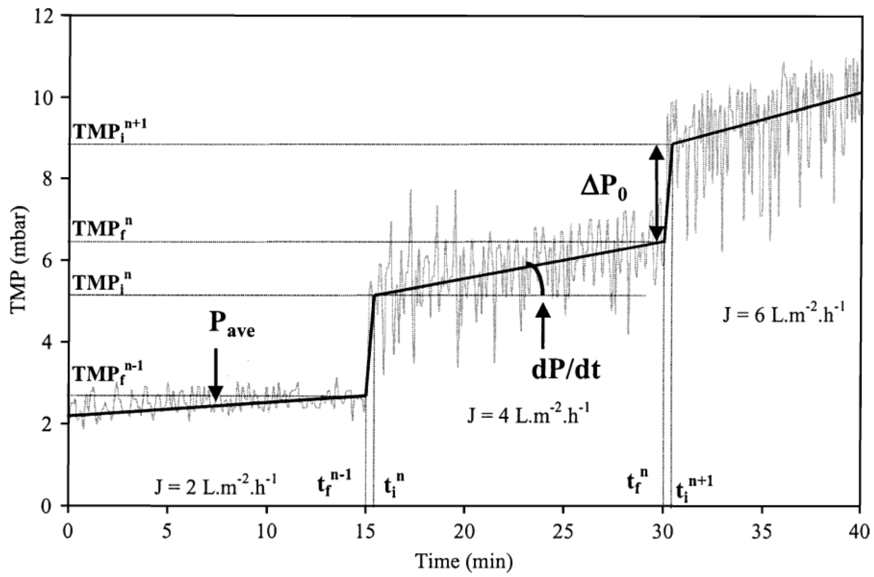


Figure 2.6: Schematic representation of critical flux determination by flux step method (Clech et al., 2003)

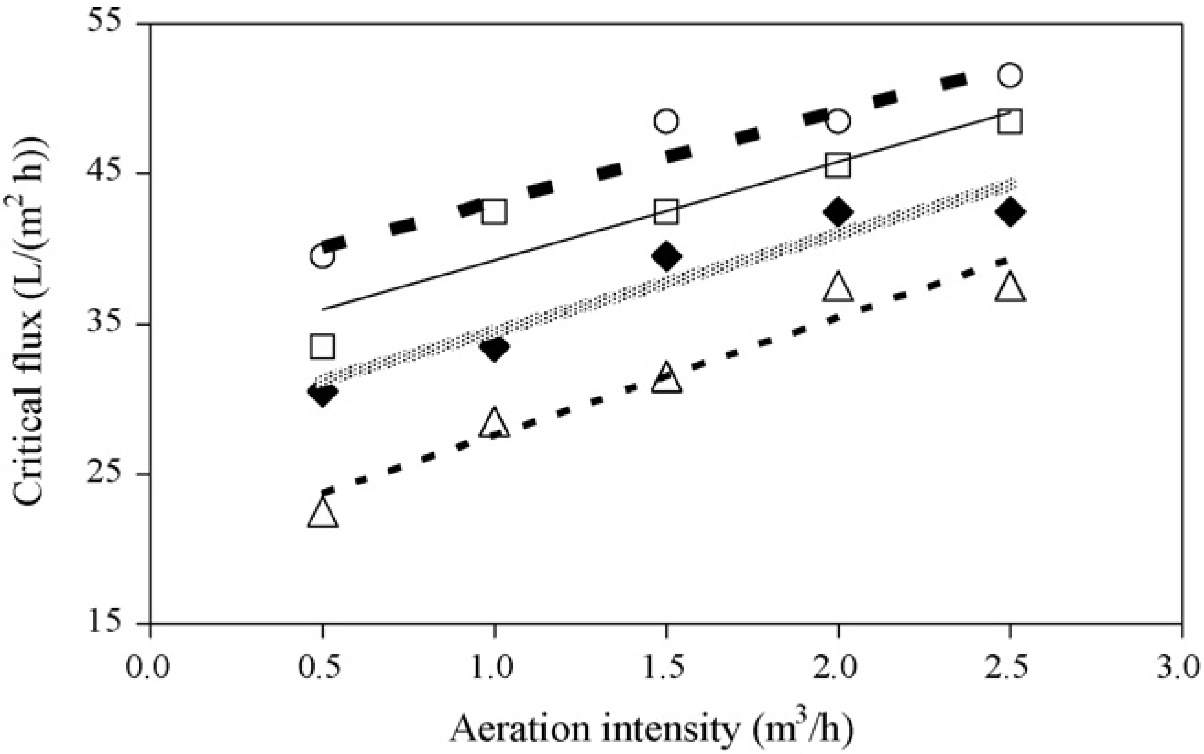


Figure 2.7: Relationship between aeration intensity and critical flux under various sludge concentration (Wu et al., 2008b).

The contribution of aeration intensity to the enhancement of critical flux is becoming greater with the sludge concentration increasing higher as shown in Figure 2.7. In practical application, the enhancement of critical flux by improving aeration intensity can help reduce the number of membrane modules installed in the reactor and in turn decrease the investment of the process; however, the increase of aeration intensity will increase the energy cost of the system. Thus, according to the analysis mentioned above, there should be an optimized sludge concentration for the MBR process and optimal aeration intensity as well. It is also reported that an increase in aeration rate and thus CFV suppressed fouling and increased critical flux (Wu et al., 2008b).

2.12 Chemical cleaning

Despite all known counter-measures, fouling cannot be entirely avoided and sooner or later the membranes have to be cleaned. It is not possible to remove all the material deposited on the membrane by physical cleaning only. Chemical cleaning which is able to remove more strongly the adsorbed deposits can thus be used (Radjenovi´c et al., 2008).

Depending on the technology, the constructor would recommend to perform maintenance cleaning (regular cleaning with low chemical concentrations, or “chemical backwash”, once a week to once a month) which lasts 30–60 min (DeCarolis and Adham, 2007, Radjenovi´c et al., 2008) or curative cleanings (infrequent cleaning with strong chemicals, once a quarter to once a year). Regular maintenance cleaning is also thought to prevent bio-fouling or bio-film formation on the membrane. The chemicals that are usually applied are basic or oxidising when cleaning fouling of biological origin, and acidic when removing salt precipitation. Chemical cleaning is carried out mostly with sodium hypochlorite and sodium hydroxide for organic deposits removal, or with acidic solutions for removal of lime or other inorganic deposits (Radjenovi´c et al., 2008). The most widely used chemical for MBR cleaning is chlorine (sodium hypochlorite), however its use in wastewater treatment is not well tolerated in countries like Germany, and cost-effective alternative chemicals are sought (Lesjean and Luck, 2005). A deposit that cannot be removed by available methods of cleaning is called “irrecoverable fouling”. This form of fouling usually builds up over the years of operation and eventually determines the membrane life-time (Radjenovi´c et al., 2008).

Recovery cleanings are performed on MBR systems when a significant loss of membrane permeability has occurred. Though the cleaning frequency can vary based on operation and water quality, for municipal wastewater treatment applications, it is typically recommended to perform this level of cleaning once or twice per year. The cleaning protocols from MBR suppliers specify the use of similar chemicals, including chlorine (2 to 3 g/L), followed by citric or oxalic acid (2%). In general, MBR systems are either CIP (Zenon and US Filter) or CIL (Kubota, Mitsubishi). During a CIP, the membranes are isolated from the MLSS, and chemicals are continuously circulated through the membranes before soaking. During CIL, the membranes are not isolated from the MLSS, and a chemical is allowed to slowly flow by gravity from the inside to the outside of the membranes. This procedure introduces chemicals to come in direct contact with activated sludge. Both of these cleaning methods are effective at restoring specific flux and TMP to values near those established at the onset of testing of new membranes (DeCarolis and Adham, 2007).

2.13 Operating costs

There are three main elements of a membrane bioreactor (MBR) contributing to operating costs (ignoring membrane costs) (Judd, 2011);

- liquid pumping
- membrane maintenance
- aeration

2.13.1 Liquid pumping

Key hydraulic operating parameters for MBR operation are flux and TMP (ΔP_m) from which permeability K ($J/\Delta P_m$) is obtained (Judd, 2011).

Recycle ratio = $\frac{Q_{dn}}{Q_p}$ where Q_{dn} = recycle flow for denitrification (2.5)

Q_p = feed flow

Pumping power $W_p = \frac{\rho g H}{1000\varepsilon}$ where W_p = power input to pump (kW) (2.6)

H = pump head

ρ = density of pumped fluid

ϵ = pump efficiency

The above equation 2.6 yields the theoretical power requirement for all liquid pumping operations.

2.13.2 Membrane maintenance

This relates to physical and chemical membrane cleaning which incurs process downtime, loss of permeates product and membrane replacement.

Physical and chemical backwashing requirements are dependent primarily on the membrane and process configuration and also feed water quality.

Key design parameters relating to membrane cleaning are (Judd, 2006);

- Period between physical cleans (t_p), either backwashing or relaxation.
- Duration of physical cleaning (T_p)

- Period between chemical cleans (t_c)
- Duration of chemical cleans (T_c)
- Back flushing flux(J_b)
- Cleaning reagent concentration (C_c) and volume (V_c) normalised to membrane area

If it is assumed that a complete chemical cleaning cycle (which will contain a number of physical cleaning cycles) restores membrane permeability to a sustainable level then the net flux is given by equation 2.7 below (Judd, 2011);

$J_{net}=\frac{n(Jt_p - J_bT_p)}{t_c+T_c}$ where n = no of physical cleaning cycles per chemical cleaning (2.7)

$n = \frac{t_c}{t_p+T_p}$

t_c and t_p is determined by threshold parameter values i.e. maximum operating pressure or minimum operating membrane permeability. Net flux is most appropriate to use for energy demand calculations and the pumping energy can be modified as shown in equation 2.8 to (Judd, 2011)

$W_{h,net} = W_h\frac{t_c}{t_c+\tau_c}$ (2.8)

For a periodic chemical clean in place, either maintenance or recovery, the specific mass per unit permeate product as shown in equation 2.9 is (Judd, 2011)

$M_c= \frac{C_c\vartheta_c}{J_{net}A_{m(t_c+\tau_c)}}$ (2.9) where A_m = membrane area (m^2)

v_c = cleaning reagent volume (m^3)

C_c = cleaning reagent concentration

If the cleaning reagent is flushed through the membrane in situ then the volume of cleaning reagent used is as shown in equation 2.10

$V_c = J_cA_mT_c$ (2.10)

where J_c = cleaning flux (LMH)

Thus $M_c = C_c \frac{J_c}{J'_{net}} \frac{\tau_c}{(t_c+\tau_c)}$

Applicable to chemically cleaning enhanced backwashing (CEB) and chemical cleaning in place (CIP).

Frequency of chemical cleaning can be determined from the rate of decline of permeability, where permeability decline rate at constant flux is determined from the pressure, *P* and time, *t* at two points within the cleaning cycle as shown in equation 2.11 below (Judd, 2011).

$$\frac{\Delta K}{\Delta t} = \frac{J_{net}}{t_2 - t_1}\left[\frac{P_2 - P_1}{P_2 P_1}\right] \qquad (2.11)$$

Where P_1 = recovered permeate after cleaning

P_2 = minimum acceptable permeability

2.13.3 Aeration

The main power requirement of an MBR comes from aeration, which is used for supply of DO for metabolism and also to maintain solids in suspension. The biological aeration requirements in MBR processes are higher in comparison to the conventional activated sludge process due to the higher oxygen demand initiated by highly concentrated biomass (Germain and Stephenson, 2005). Aeration is also used for membrane cleaning purposes in submerged MBRs. Membrane cleaning is provided by air scouring in flat sheet membranes and by fibre agitation in the case of hollow fibre membranes.

As biomass and aeration characteristics have an effect on each other, two things should be considered when studying aeration operations in MBRs:

- The effects of biomass components on aeration efficiency, represented by the oxygen transfer parameters,
- The effects of aeration (intensity and type of diffusers) on biomass characteristics.

As in all aerobic biological systems, biomass contained in the MBR requires oxygen to perform diverse chemical reactions. The right amount of oxygen needs to be provided to the micro-organisms and wastewater, in response to their three specific demands:

- Carbonaceous biochemical oxygen demand (BOD): conversion of the carbonaceous organic matter in wastewater to cell tissue and various gaseous end products,
- Nitrogenous BOD: ammoniacal nitrogen is oxidised to the intermediate product nitrite, which is then converted to nitrate (this process is nitrification),
- Inorganic chemical oxygen demand (COD): oxidation of reduced inorganic compounds within the wastewater.

The amount of oxygen diffusing in the mixed liquor is characterised by the oxygen mass transfer coefficient. This mass transfer coefficient is one of the general parameters used to describe the diffusion of particles from regions of high concentration into regions of lower concentration. This approach assumes that the diffusion occurs across an interface. The basic model for mass transfer is:

(Rate of mass transfer) = κ (interfacial area) (concentration difference)

Where κ is the mass transfer coefficient.

The main parameter used to characterise the oxygen transfer in aeration processes is the overall mass transfer coefficient, k_La; where k_L represents the mass transfer coefficient based on the liquid film resistance and a, the interfacial area. Another parameter commonly used to describe the oxygen transfer in biological aerated systems is the α-factor. This correction factor is defined as the ratio between k_La in the process solution and k_La in clean water. It accounts for the effect of process water characteristics on the oxygen transfer coefficient (Germain and Stephenson, 2005, Rodríguez et al., 2012).

In MBRs, like in all aerobic wastewater processes, both the biomass characteristics and the design of the aeration system affect the oxygen transfer. Biomass is a heterogeneous mixture of particles, micro-organisms, colloids, organic polymers and cations, all which have different shapes, sizes and densities. All of these parameters have an impact on oxygen transfer. Mass transfer is also linked with contact area between gas and liquid phases i.e. bubble shape and solids concentration. Bubble characteristics will differ depending on the kind of aerator used and the bubble coalescence effect created by the biomass characteristics. The aeration in MBRs is usually provided by fine bubble aerators and is used to keep the content of the aerobic tank well mixed and also to provide oxygen to the biomass. In addition, in submerged MBRs, coarse bubble aerators situated under the membrane modules are used to scour and/or gently agitate the membranes in order to control membrane fouling.

On the other hand, MBR properties are affected by aeration. Changes in airflow rate affect the biological and physical characteristics of the mixed liquor. Species diversities differ depending on the amount of oxygen available in the solution. Mixing intensity usually as a result of aeration, affects the shape and size of particles by breaking-up sludge flocs.

Aeration plays an important role in MBR operations and represents its major power input (Germain and Stephenson, 2005, Rodríguez et al., 2012) and is used to supply dissolved oxygen to the biomass, to maintain solids in suspension and to improve membrane cleaning in submerged MBRs (Rodríguez et al., 2012). To allow MBRs to be competitive with regards to conventional wastewater treatment plants, these additional costs have to be reduced. Particle concentration, particle size and viscosity are the main parameters characterising the biomass and are known to have an effect on the oxygen transfer. The aeration intensity affects the particle size and the viscosity, while solids concentration modifies the viscosity. Their individual effects on oxygen transfer can be modified by the added effect of another parameter, especially for particle size and concentration. Particle concentration, particle size and viscosity are the main parameters characterizing the biomass and they are known to have an effect on oxygen transfer. These three biomass parameters and aeration are interrelated. The aeration intensity affects particle size and viscosity, while the concentration of solids modifies the viscosity. Also mass transfer is influenced by the area of contact between the gas and liquid phases, so particle concentration and particle size affect oxygen transfer (Rodríguez et al., 2012).

In addition to aeration for the biomass, air can be injected at the membrane bottom in order to limit fouling of the membrane surface or clogging of the bundle. Air is introduced below the membrane assembly and supposedly is ideally distributed to optimise the air scouring action across the membrane surface. The ideal air flow and flow pattern for this purpose are still not clearly known today (INSA, 2006).

The position of the aerators used for scouring also affects the bulk liquid velocity in a submerged FS system, e.g. locating the aerators at the bottom of the tank and not at the entrance to the draft tube where they block the available cross-section and slow down the flow increases the internal circulation (Böhm et al., 2012). Considering that a significant portion of operation and maintenance costs in wastewater treatment plants is originated from aeration costs, excess and extensive aeration to control membrane fouling should be modified and an alternative strategy for membrane fouling control is needed. Nevertheless, it seems that influences of membrane surfaces properties and thus of membrane material are important mainly in the case of short filtration duration. (Ma et al., 2000) reported that after 1 hour of filtration, membranes are sufficiently fouled so that deposit controls the filtration behaviour and in this way, membrane properties have less influence over longer durations (INSA, 2006).

The beneficial effects of air scouring to control fouling at the membrane surface of flat sheet membranes are well documented. Typically, it is assumed that there is a linear relationship between membrane flux (L_{water}/m^2h) and the air scouring rate (Nm^3/hm^2), within some limits above which this positive influence of higher air volume per unit membrane area is no longer observed. Aeration as a means for controlling fouling is related to the cross-flow velocity of air bubbles (Sofia et al., 2004, Böhm et al., 2012).

Besides aeration rate, diffuser port size and the correlating bubble size, module and tank geometry (membrane spacing, fibre slackness, liquid level, cross-sectional areas of riser and down-comer, etc.) have decisive effects on the achieved cross-flow velocity, shear stress and bubble-membrane-contact (Böhm et al., 2012). Air is introduced below the membrane assembly and is supposed to be ideally distributed to optimise the air scouring action across the membrane surface. The ideal air flow and flow pattern for this purpose are not clearly known today (INSA, 2006). Uplift arising from the air bubble rising along the membrane surface helps to induce shear stress which generates the back transport of deposited flocs from the membrane surface. The air diffuser's role is, therefore, important as it governs the condition of the cross-flow velocity stream and thus affecting the rate of fouling and it is a function of membrane configuration and aeration intensity (Sofia et al., 2004).

For membrane processes, air bubbling provides substantial flux enhancement as demonstrated by a considerable amount of data which show the beneficial effects in a number of model systems and on a commercial scale for sewage treatment. The most prevalent application of this technique is for waste water treatment using the MBR. The channeled bubbles accomplish three important objectives (Trivedi, 2004):

- provide adequate oxygen to maintain cell respiration at design MLSS,
- scour the membranes to prevent fouling, and
- create a pressure gradient between the top and bottom of the membrane unit

The pressure gradient created by the rising bubbles induces upward cross-flow of mixed liquor over the membranes. The liquor is filtered as it flows across the membrane, due to the TMP gradient created by the hydrostatic head of the water above the membrane cassettes. The flux, or filter flow rate per area, is directly proportional to the TMP gradient induced by the head of the water over the membranes (i.e. by the water level in the tank) (Trivedi, 2004).

Aeration intensity is defined as airflow rate per unit floor area of a membrane unit. Cross-flow velocities, however, do not necessarily increase proportionately with increasing aeration intensity. There would be a critical airflow rate beyond which further increase in aeration intensity would virtually show no additional shearing effect. In order to achieve a good scouring effect, membrane manufacturers have recommended the use of a coarse bubble diffuser (Sofia et al., 2004).

Aeration for fouling prevention can be operated using specific air injectors, independently to the air injection for biomass. Operation with sequential air injection is also being used on an industrial scale, for example in the ZENON systems where the objective of this sequencing is to decrease operating costs due to the energy consumption induced by air sparging (INSA, 2006). Recent systems use two different aeration systems: one which injects fine dispersed bubbles that are necessary for biomass growth and one which generates coarse bubbles to control fouling on the membrane surface. The liquid flow close to the membrane is only due to permeate flow through the membrane wall and to a liquid motion induced by the bubbles. This process is commercialised by industrial firms for instance with flat sheet membranes (Kubota) or tubular membranes (Milleniumpore) (INSA, 2006).

Bubbling is an obvious strategy to induce flow and produce shear at the membrane surface in submerged systems which is particularly attractive in MBRs. In MBR process, the use of bubbling to control fouling is of paramount importance. Two complementary strategies may be involved: air sparging to prevent membrane fouling during filtration runs and/or air sparging to remove fouling (sequential filtration/gas sparging steps). In case of aerobic MBRs, the air-liquid flow used against fouling is different from the one used for mixed liquor aeration and oxygen mass transfer (INSA, 2006).

Even if most MBRs in operation at industrial scale are using air inputs for fouling control, optimization of air sparging is not a solved problem due to the lack of understanding of mechanisms induced by bubbles. Moreover, air sparging or aeration is a key problem for the process as it represents one of the main operating costs. The major reason for this misunderstanding is that each module configuration and membrane system (which are in constant evolution) presents specific flow patterns and thus transfer phenomena (INSA, 2006). Membrane surface fouling, and the less well investigated phenomenon of membrane channel clogging, are both ameliorated ostensibly through the use of coarse bubble aeration, applied beneath the MBR membrane module. For an immersed MBR approximately 30–40% of the energy demand arises from aeration of the membrane with a further 10–50% depending on feed water strength. It is the membrane aeration which is primarily responsible for promoting permeate flux and/or maintaining membrane permeability (Verrecht et al., 2008).

The effect of aeration on filtration improvement is higher for a higher MLSS concentration in raw water. In the same way, a higher effect of aeration at higher

TMP was observed. These two results have been explained by the reduction of the polarization layer induced by air and liquid slug: the impact of aeration is all the more evident when membrane fouling is high (INSA, 2006).

Bubble size is also an important factor for MBRs. When gas is injected into a stationary liquid, such as the situation in immersed membrane systems, bubbles are formed and move upward driven by buoyancy. The bubble motion also generates a secondary flow behind the bubble, i.e. the wake region. The size of the formed bubbles depends on the way the gas is introduced, the sparger type and gas flow rate. Depending on the size of the bubble, it tends to take different shapes which determine the strength and extent of the wake region (Cui et al., 2003).

Fine bubbles have large oxygen transfer efficiency due to their large surface area. Coarse bubbles are recognized empirically for their ability to scour the membrane effectively. For a fixed flat sheet membrane, the bubble size, bubble frequency, and bubble distribution will influence the hydrodynamics and hence the mass transfer. (Zhang et al., 2009) showed that the shear stress on the flat membrane with a fixed clearance of 20mm increased with bubble size up to a value of 60 mL but was insensitive to size beyond that (Yamanoi and Kageyama, 2010b). A potential advantage of the flat sheet arrangement is that the membrane surfaces are precisely located, unlike hollow fibre bundles, and more accessible to well directed bubbles (Cui et al., 2003).

For the membranes of UF and MF, the thickness and compactness of the external fouling layer are greatly affected by the fluid dynamics of feed flow. The use of gas bubbling to create better fluid dynamics, so as to control external fouling, was proposed by (Tajima and Yamamoto, 1988). In 1990s, a number of studies demonstrated that creating gas–liquid two-phase flow at the feed side was an effective method to limit membrane fouling, with the trans-membrane flux being increased by more than 110%. It was shown that the bubbling induced secondary flow improved greatly the shear stress near the membrane surface, thus the foulant deposition was limited. Several studies found that the efficiency of fouling limitation is greatly affected by the size of induced bubbles, which mainly depends on the size of employed nozzles. Uniformly distributed fine bubbles provided better fouling control than coarse bubbles hence prolonging the membrane operation for hollow fibre submerged membranes (Ding et al., 2011, Lu et al., 2008). About optimal aeration characteristics, Li et al. (1997) studied the influence of air bubble frequency and size (bubbles volume from 2.2 to 8.3 mL and frequency from 0.05 to 1 s^{-1}) in 12.7 mm tubes for HAS et β IgG proteins solutions. An increase of the bubble volume induced an improvement of permeate flux but there was a limiting bubble volume size above which no more increase of permeation flux could be obtained. Same kind of results have also been obtained about the influence of the bubble frequency (INSA, 2006). Possible explanations are based on different phenomena:

- an increase of shear stress on the membrane surface,
- intermittence of air and liquid slugs which inverts the direction of the stress
- pressure variations in the slipstream of slugs,

With a flat sheet module immersed directly in the bioreactor, the dependency of sludge accumulation is on aeration intensity for different flow channel widths. They concluded that sludge accumulation and filtration resistance are dependent on aeration intensity, and are less dependent on flow channel width and MLSS concentration (INSA, 2006).

Besides aeration rate and bubble size (or diffuser ports), module and tank geometry (membrane spacing, liquid level, cross-sectional areas of riser and down-comer, etc.) have decisive effects on the achieved cross flow velocity, shear stress and bubble-membrane contact (Prieske et al., 2010).

Fouling in MBRs is a complex problem caused by interacting biological, chemical and physical phenomena. Recent research has led to the emergence of engineering design tools to optimize module and tank geometry as well as operating parameters based on a more fundamental understanding of the effect of the hydrodynamic conditions in MBRs on fouling control. These have contributed to the development of new sparging strategies that have resulted in up to 70% reduction in power costs for fouling control. However no valid model yet exists that can comprehensively describe the relationship between fouling rate and the hydrodynamic conditions (Böhm et al., 2012).

2.14 Membrane module spacing

A key advantage of the MBR is that it can attain complete solid–liquid separation independent of the quality of the mixed liquor. However, the running cost and energy consumption are still higher in comparison with conventional systems. It is important to develop a cost-effective design to make the technology more applicable to full-scale wastewater treatment plants. It is also necessary to develop a more compact module design that can attain higher flux combined with lower energy consumption.

The main drawback of MBR systems in comparison to the CAS processes remains the high operating cost of MBRs. This is mainly due to the fact that up to 70% of the total energy demand for MBR systems is for fouling mitigation by air scouring. Operating data from full-scale systems suggest that the energy used for fouling mitigation is only optimally used 10% of the time, and therefore, there are significant opportunities to reduce total energy demand for MBR systems.

The major challenge in the membrane filtration systems is the control of membrane fouling and its minimization during operation. There is a pressing need to minimise the fouling potential and/or develop a simple method to measure and predict the fouling potential of wastewater.

Before the membrane bioreactor is fully ready for field application, however, some of its limitations must be addressed. First of all, there is often a rapid decline in flux due to membrane fouling as a result of the high biomass concentration in the reactor. Membrane fouling is characterized as a reduction of permeate flux through the membrane resulting from increased membrane resistance due to pore blocking, concentration polarization, and cake formation. Permeate flux decline is influenced by a number of factors relating to the feed wastewater (composition), the membrane (element geometry/configuration, area and material composition), and reactor operation (biological condition and hydrodynamics). As characteristics of the membrane are determined by membrane manufacturers, researchers have focused on various strategies to reduce membrane fouling.

In general, the membrane fouling occurring in membrane bioreactors is attributable to three processes: sludge particle deposition, adhesion of macromolecules to the membrane surface, and pore clogging by small molecules

Fouling leads to a decline in permeate flux, requiring more frequent membrane cleaning leading to more frequent replacement, which then increases operating costs. Nowadays, academic and commercial research tends to focus on the ways to improve the operation of membrane equipment and to save energy, in particular, by decreasing the aeration demand.

Since the fouling rate increases roughly and exponentially with the flux, MBR plants operate at modest fluxes and preferably below the so-called critical flux. In MBR operation, however, due to the complexity of the mixed liquor, some irreversible fouling constantly occurs, which makes it impossible to achieve the sub-critical conditions as for the strong form of the critical flux (Radjenovi´c et al., 2008).

Two of the most significant components of MBR opex are membrane replacement and energy demand, both of which are made more onerous by membrane surface fouling. Membrane surface fouling, and the less well investigated phenomenon of membrane channel clogging, are both ameliorated ostensibly through the use of coarse bubble aeration, applied beneath the MBR membrane module. For an immersed MBR approximately 30–40% of the energy demand arises from aeration of the membrane with a further 10–50% – depending on feed water strength – demanded for bio-treatment it is the membrane aeration which is primarily responsible for promoting permeate flux and/or maintaining membrane permeability. Reducing the energy demand in a conventional MBR thus relies on an understanding of the total aeration requirements of the process (Verrecht et al., 2008).

Fouling is caused by the deposition of particulate, colloidal or soluble material inside the pores or on the membrane surface. The layers which deposit on the surface can be removed by hydrodynamic forces. These forces, however, do not only affect the deposition layer but might also be responsible for negative effects like SMP release and decrease in floc sizes (Böhm et al., 2012).

One important factor that reduces the filtration performance is flux decline due to cake formation on the membrane surface. It is important to consider how cross flow and/or aeration removes cake layer accumulated on the membrane surface.

It has generally been observed that the sustainable flux increases (roughly linearly) with aeration rate up to some threshold value, beyond which little or no further improvement in permeability is observed. The increased flux has generally been attributed to the associated increase in cross flow velocity of the air-lifted liquid. However, it is also known that the local flow pattern around an air bubble rising through a channel is very complex, exerting significant transient shear at the membrane surface and increasing the flux attained over that from liquid flow alone (Verrecht et al., 2008).

It is commonly accepted that air bubbling close to the membrane is one of the most efficient means of minimizing reversible fouling and ensuring sustainable operation. Bubbling induces local shear stress, which controls fouling and creates a favourable hydraulic distribution. Membrane aeration forms an important part of the operating cost of the MBR and thus is important to optimise the membrane aeration (Johir et al., 2011).

Membrane fouling can be prevented by hydraulic effects. In external-module MBRs, whose membranes are located external to the bioreactors, a cross flow fluid provided by a recirculation pump reduces the membrane fouling. More recently, submerged MBRs have been developed. Their membranes are submerged in aeration tanks, and aeration bubbles not only supply oxygen to the activated sludge, but also scour the membrane surfaces. These bubble flows can suppress the membrane fouling effectively.

For membrane processes, air bubbling provides substantial flux enhancement as demonstrated by a considerable amount of data showing beneficial effects in a number of model systems and on a commercial scale for sewage treatment. The most prevalent application of this technique is for waste water treatment in MBRs. There are currently many suppliers of MBR systems including Kubota who successfully pioneered the flat sheet concept in which air is injected from a distributor below a stack of panels. The air then rises through the rectangular channels formed between adjacent pairs of panels. For a fixed flat sheet membrane, the bubble size, bubble frequency, and bubble distribution will influence the hydrodynamics and hence the mass transfer (Zhang et al., 2009).

For most submerged membrane systems, for which the permeate flux is typically in the order of 20–50 L/(m2 h) for wastewater and drinking water applications, respectively, and the bulk liquid velocity at the membrane surface is typically greater than 0.1 m/s (i.e. resulting in a suction rate, defined as the ratio of uniform suction velocity to bulk inlet velocity, of less than 0.01%), the effect of permeation on the hydrodynamic conditions at the membrane surface is expected to be minimal (Böhm et al., 2012).

Air scouring is used to induce favourable hydrodynamic flow fields in the vicinity of a membrane surface to promote fouling control. These flow fields are induced by the complex interactions between orthogonal as well as cross flows and turbulent eddies created by the rising bubbles. The sparging intensity (i.e. gas flow rate) can significantly affect fouling control in submerged membrane systems. Increasing the sparging intensity results in better fouling control, however, a plateau is typically reported above which further increasing the sparging intensity does not further improve the extent of fouling control. Besides aeration rate, diffuser port size and the correlating bubble size, module and tank geometry (membrane spacing, fibre slackness, liquid level, cross-sectional areas of riser and down comer, etc.) have decisive effects on the achieved cross flow velocity, shear stress and bubble-membrane-contact (Böhm et al., 2012, Prieske et al., 2010).

For flat sheet systems, many geometrical and operational design parameters still need to be optimized (e.g. bubble size and membrane spacing). The presence of the walls can drastically change the bubble shape. When the bubble diameter equals the membrane spacing a further increased bubble size leads to flat cap bubbles. Although rise velocity is independent of channel width, the plate distance influences the maximum possible stable bubble size. In comparison to bubbles rising in unconfined geometries, in narrow channels smaller bubbles break due to the apparent higher shear. An increase in the size of the bubble above a critical diameter does not yield higher shear stress. This is in agreement with (Ndinisa et al., 2006a) who observed that as bubble size increased, so did the cleaning effect. However, when bubbles became larger than the channel width, a further increase in size only had a minor effect on fouling control.

The module geometry as well as the sparging approach can significantly affect the magnitude and distribution of shear stress in FS membrane systems. A number of different channel spacing and module heights are used in commercial flat sheet membrane systems indicating that optimization in this area still needs to be done (Böhm et al., 2012).

Since many fundamentals of multiphase flow in MBRs are still unknown and difficult to access experimentally, there is no common way to construct and operate flat sheet modules as yet which leads to a wide range of specific aeration demand (SADm) values and waste of energy.

Almost all commercial submerged MBRs use the technique of air scouring in order to combat membrane fouling. The air is also used by microorganisms in the aerobic biodegradation process. The air bubbles scour the surface of the membrane and generate the liquid cross flow velocity without the need for a recirculation pump. For submerged MBRs energy consumption rates have been reported as ≥1 kWh .m-3. More than 50% of this energy was used for air scouring. There is a need to reduce this fraction of air used for air scouring in order to make the MBR technology an even more commercially viable alternative to conventional wastewater treatment processes.

Most commercial MBRs operate below critical fluxes. However, even at such conditions, slow fouling still occurs due to the interaction of the membrane with certain constituents of the MLSS such extracellular polymer substances.

For submerged flat sheet membranes, the gap between adjacent membranes can have a significant effect on hydrodynamic conditions.

2.15 Woven Fibre Micro Filtration Membrane

Woven fibre microfiltration (WFMF) technology has undergone significant development at the Pollution Research Group of the University of KwaZulu-Natal (UKZN) during the early 1980’s. It is currently being further developed by the Durban University of Technology (DUT). Due to the spaces between the fibres being in the range of micrometres, this material could be considered as a microfiltration membrane.

At the beginning, woven fibre microfiltration (WFMF) contained two woven polymer layers glued together to form the rows of parallel filter tubes that was termed curtain. Currently, support can be inserted between flat sheets. DUT in collaboration with Water Research Commission of South Africa and Stellenbosch University is developing inexpensive flat sheet modules for ideal conditions of developing economies. The woven fibre microfiltration (WFMF) fabric is developed by a South African company, Gelvenor and offers various potential advantages over the current commercial inversion-cast flat-sheet membranes (Pillay and Jacobs, 2008).

The woven fibre micro filtration module consists of three elements: a PVC frame that incorporates a permeate outlet; two sheets of fabric glued to either side of the frame; and a spacer between the sheets of fabric to facilitate fluid flow to permeate outlet as shown in Figure 2.9. Figure 2.8 shows magnification of Woven Microfiltration sheet (The total path length of the membrane was 7.57 m. yielding a filtration area of 0.6 m^2)

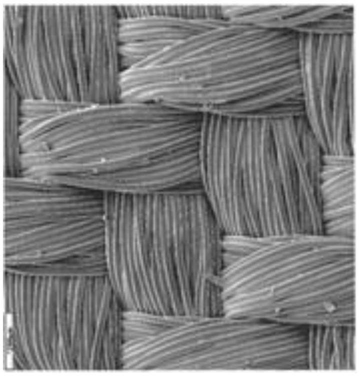


Figure 2.8: woven fabric sheet

Figure 2.9: single WF membrane module

WF technology has been applied intensively on potable water for the last decade and flat sheet modules have been the most tested modules. Most of the investigated systems were designed ideally for rural areas where there is no supply of electricity and lack of skills (Buckley and Naidoo, 2001, Pikwa et al., 2009). In Asia this technology has been tested in water treatment for emergency situation. This work was conducted at the Asian Institute of Technology (AIT). Very little has been reported in wastewater application. In Asia this technology has been tested in decentralised Wastewater treatment systems. However, this project looked at the application of WF membrane module in membrane bioreactors for wastewater treatment for reuse.

2.15.1 Woven Fibre Microfiltration performance

Woven Fibre microfiltration flat sheet membranes have proved to have a good potential in potable water treatment with Table 2.6 below showing some of the performance indicators for the WFMF. Small scale Remote Rural Treatment Systems (RRTS) achieved consistently a turbidity less than 1NTU from a feed ranging from 20 to 300 NTU.

The WFMF membranes modules remove about 95 % of the bacteria. Even for raw waters with very high levels of contamination; the RRTS produces a final product that is completely safe for human consumption.

Table 2.6: Performance of WFMF (adapted from Pikwa et al., 2009)

Water Source	E.Coli in raw water (counts/100 ml)	E.Coli in permeate from RRWTS (before exposure to disinfectant) (counts/100 ml)	E.Coli in product container after disinfection (counts/100 ml)
River 1	4838	980	0
River 2	8160	185	0
River 3	11191	23	0

Advantages of WFMF membrane technology over current commercial modules include:

- Ease of membrane cleaning
- Membrane does not get damaged if it dries
- It is robust
- It can be easily repaired if it is accidently damaged
- It achieves the permeate quality of less than 1NTU turbidity (Pillay and Jacobs 2005)
- It is inexpensive

A significant portion of operation and maintenance costs in wastewater treatment plants originate from aeration costs. Excess and extensive aeration to control membrane fouling should be modified and an alternative strategy for membrane fouling control is needed (Kim et al., 2008).

There are several factors that can be investigated when trying to optimise the geometric setting of the membrane pack. These include

- Height of water above membrane pack
- Diffuser height from membrane modules
- Membrane module spacing
- Type of diffuser used

CHAPTER THREE: METHODOLOGY

3.1 Equipment set up

The pilot plant was set up at the Veolia Waste Water treatment plant located at the Durban South Treatment works. A P&I diagram of the experimental rig used for the experiments is depicted in the Figure 10 below. Figure 11 – 15 below show the equipment set up and the membrane modules that were used in the running of the experiments.

A pilot scale WF-IMBR set up with 18 flat sheet Woven fiber micro filtration (WFMF) membrane modules of area 6.426 m^2 was constructed based on the Wiese pilot scale version (as shown in Figure 3.1). Figure 3.2 shows the equipment set up having the feed tank, the filtration tank which will have the submerged WFMF MBR as shown in Figures 3.3, 3.4 and 3.5. The total area of a single membrane module was designed to be 0.357 m^2 with 0.1785 m^2 on each side of the membrane module. The pore size is estimated to be 0.4 µm. Figure 3.6 shows the fine bubble disc diffuser that will be used to aerate the feed water in the feed tank so as to maintain the DO levels at above 2 mg/l. The membrane module dimensions, membrane pack handle dimensions and membrane pack casing dimensions that were used in the construction of the equipment are also shown below.

The system was fed with real activated sludge abstracted from the return line of activated sludge of the wastewater treatment plant. This was done by means of a submersible pump which was located in the return activated sludge mixing chamber. The MBR permeate flow rate was controlled by using a manipulated peristaltic pump. Permeate was compensated by the raw feed of the oxidation zones. Raw feed was abstracted prior to the anaerobic mixing chambers. The pressure reading (TMP) was recorded by use of a pressure traducer. The SCL-KO4 blower of capacity of 100 m^3/hr per meter head was used for aeration.

The process feed tank, filtration tank and permeate tank were open to the atmosphere. The hydraulic retention time of the pilot plant was 24 hours.

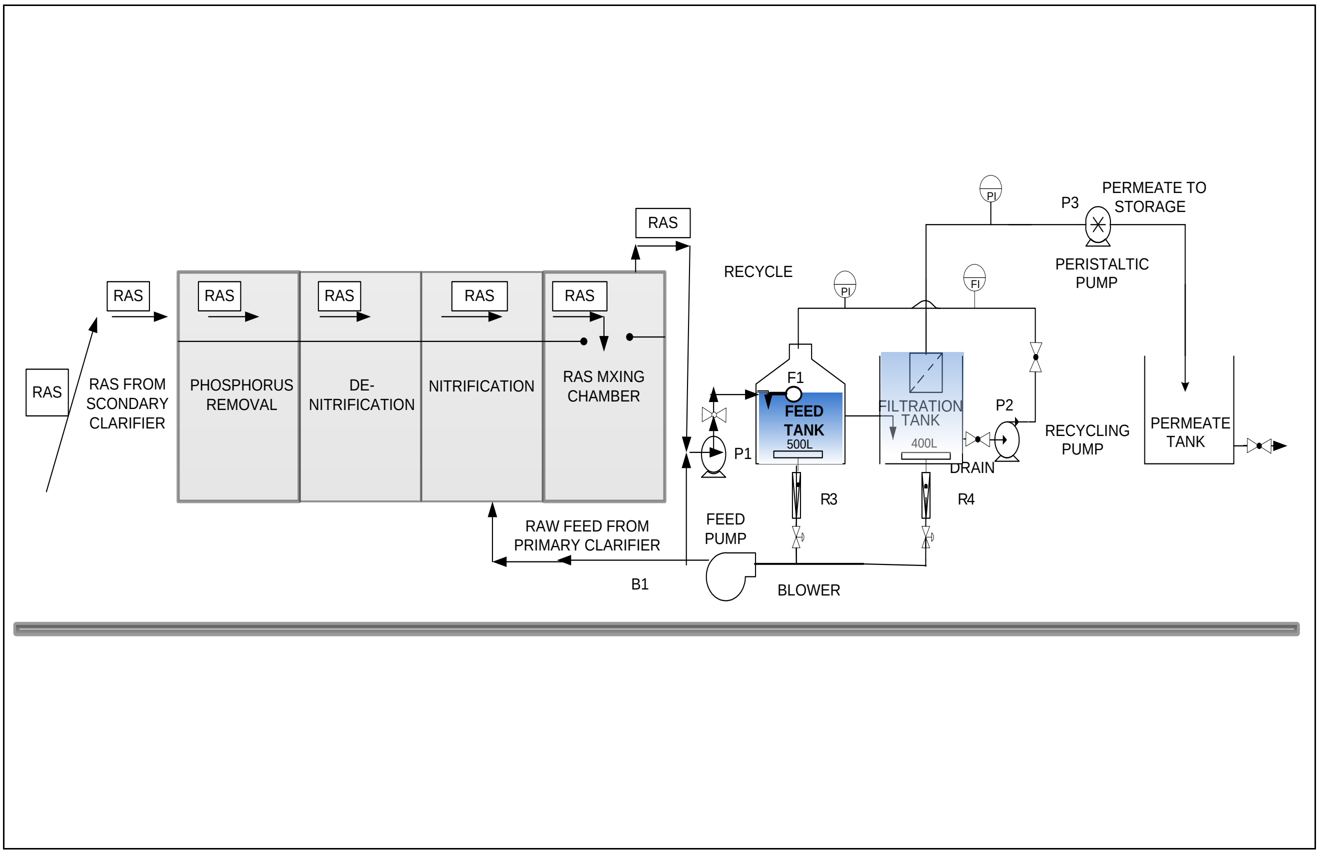


Figure 3.1: P&I diagram of immersed woven fibre microfiltration membrane bioreactors

Figure 3.2: Equipment set-up photo

Figure 3.3: Membrane modules set up

Figure 3.4: Top view of membrane pack

Figure 3.5: Membrane pack with the modules inserted

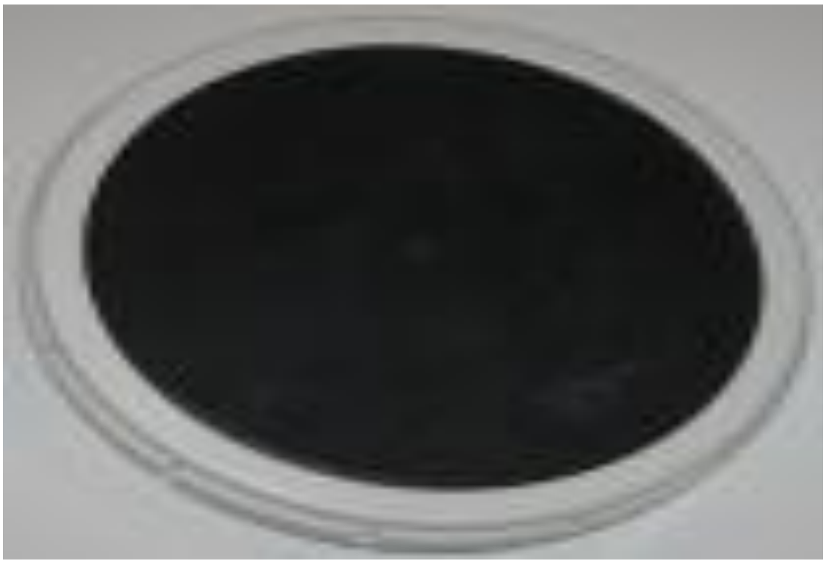


Figure 3.6: **Photo of fine bubble disc diffuser**

Figures 3.7, 3.8, 3.9 and 3.10 show the membrane modules and membrane pack dimensions that were used in the construction of the experimental equipment used in the pilot plant set up at Veolia Wastewater Treatment plant.

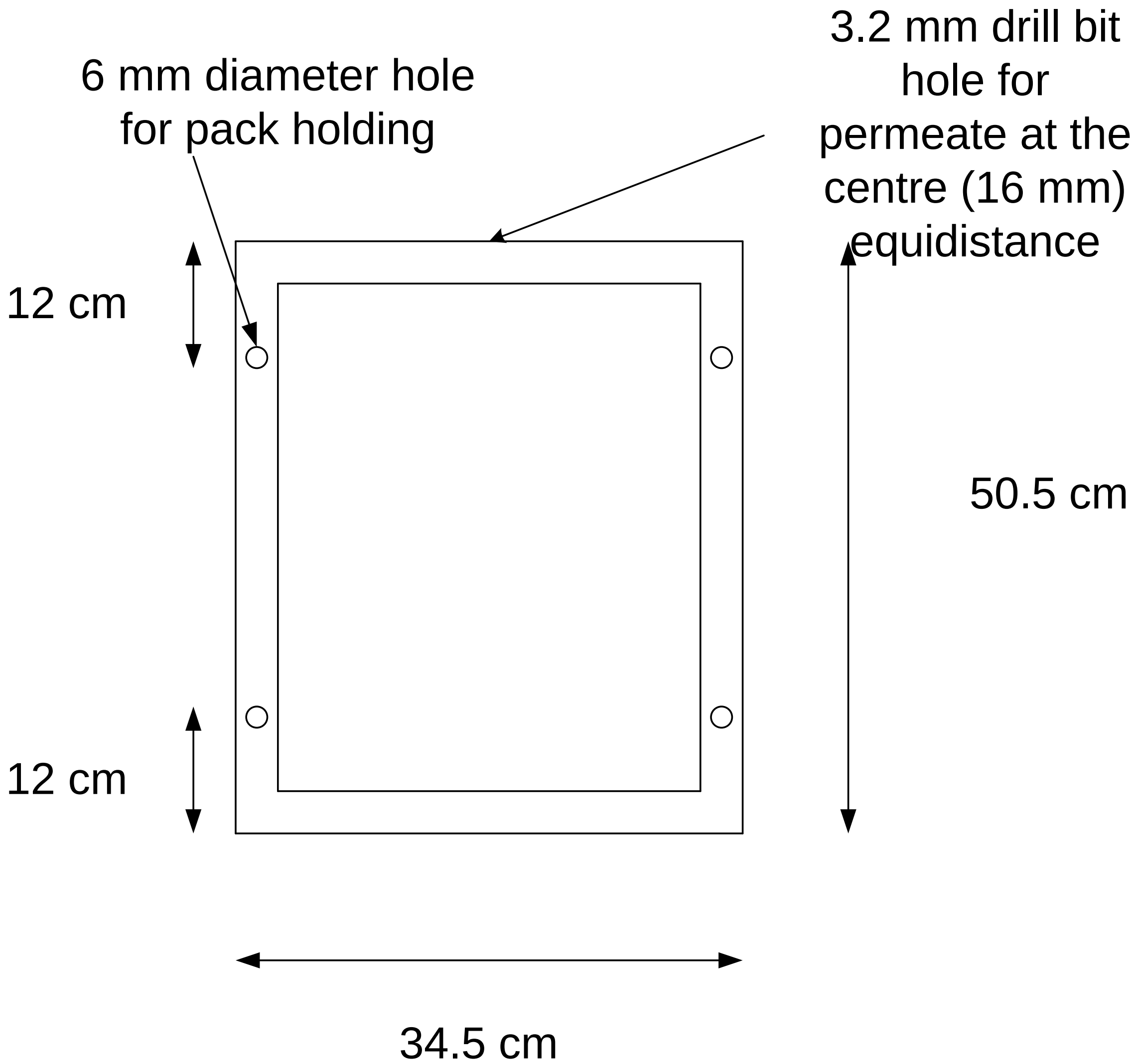


Figure 3.7: Membrane module dimensions

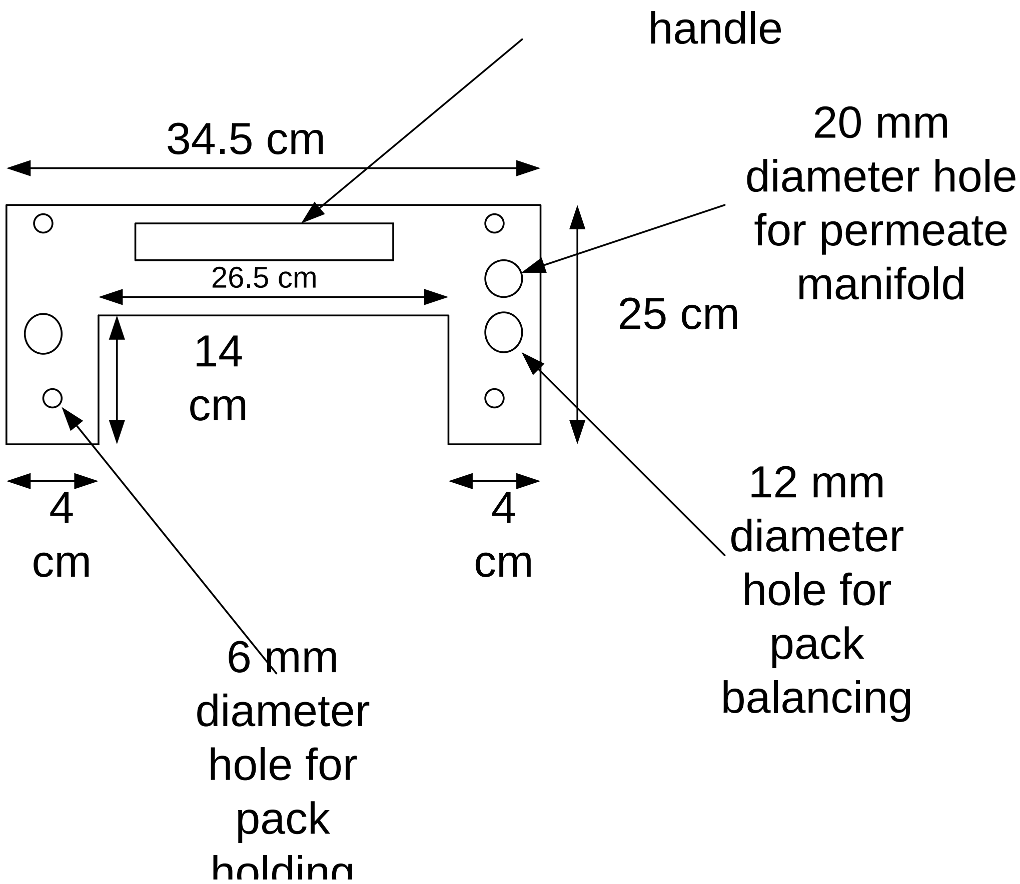


Figure 3.8: Membrane pack handle dimensions

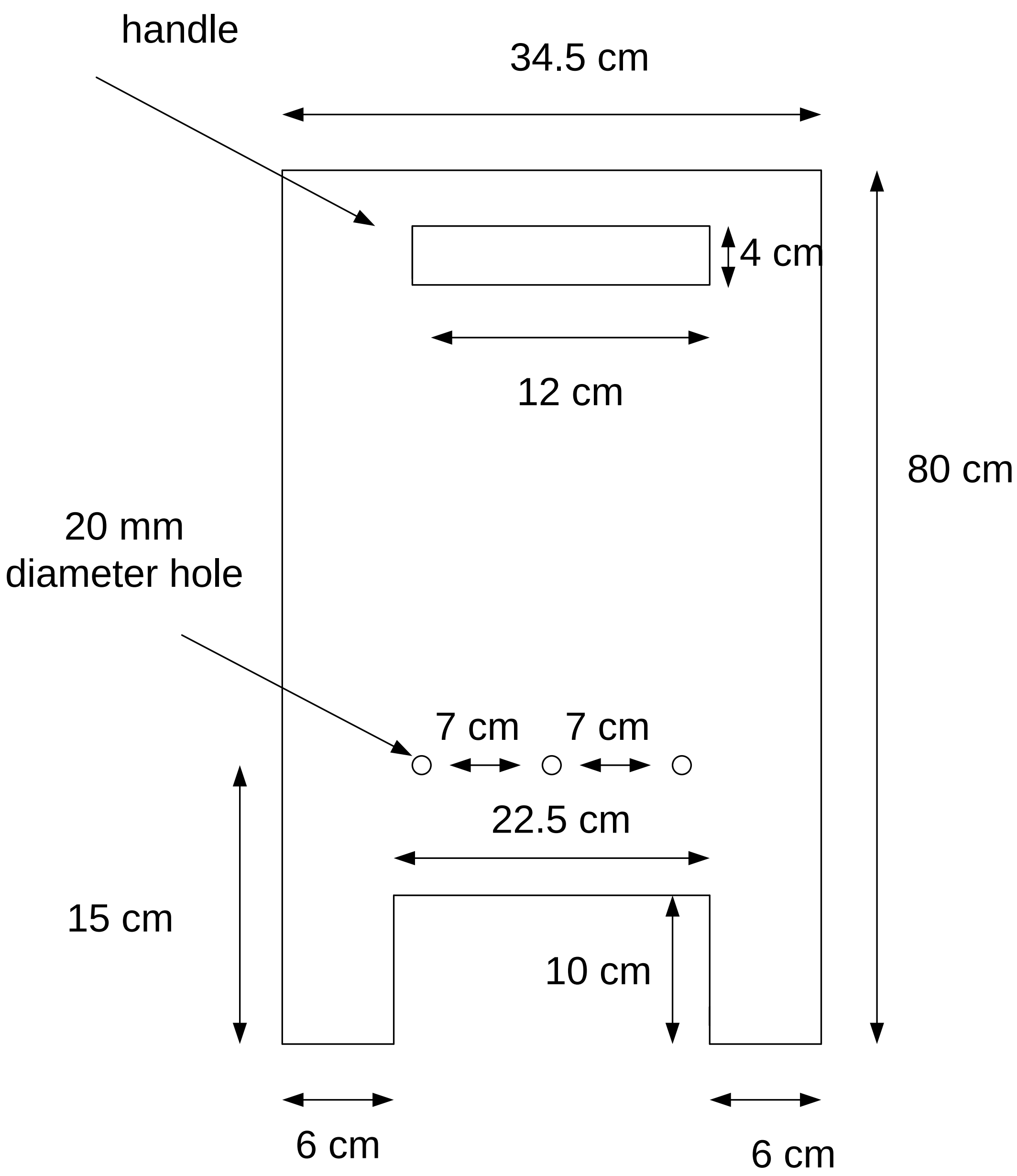


Figure 3.9: Membrane pack casing dimensions

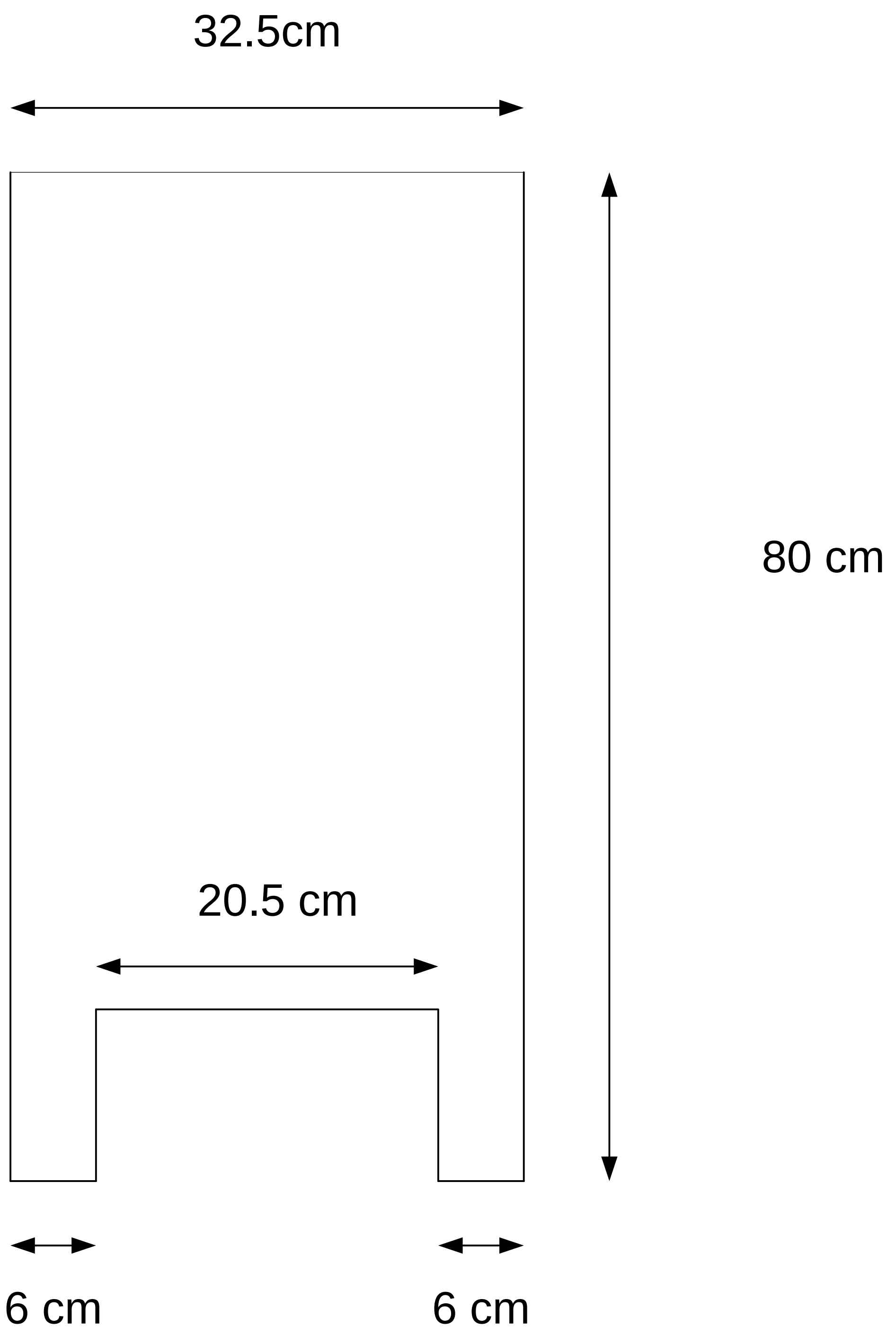


Figure 3.10: Membrane pack casing dimensions

Objectives to be met

There are several parameters that can be investigated when trying to optimise the geometric settings so as to reduce on energy utilisation in terms of aeration. The parameters to be investigated in this research are;

i. Pipe diffuser height effect
ii. Comparison of pipe and disc diffuser
iii. Membrane module spacing

3.2 Critical flux experimental procedure

In this experiment, the critical flux of the wastewater at MLSS concentration of 4 g/l, 8 g/l and 12 g/l are to be evaluated. As indicated earlier, the feed water for the pilot plant was extracted from the return activated sludge mixing chamber which was of about MLSS concentration of 12 g/l.

Critical flux was obtained by use of flux step method as this gives a more accurate data as compared to the pressure step method as has been indicated in various literatures.

As the pump speed is indicated in terms of percentages, a curve of pump speed vs. Flux was initially generated prior to commencement of the experiments. Based on these reading, the flux steps used were based on increasing the permeate pump speed at 10% interval.

Aeration was done by means of the SCL-KO4 blower. The aeration rates are regulated by means of rota-meters with a maximum capacity of 200 l/min. The aeration rates used during this experiment were 20 l/min, 30 l/min and 40 l/min, 60 l/min, 90 l/min, 108 l/min, 126 l/min, 144 l/min also represented as 1.11 l/min per module, 1.67 l/min per module and 2.22 l/min per module, 3 l/min per module, 5 l/min per module, 6 l/min per module, 7 l/min per module and 8 l/min per module respectively.

All experiments were repeated so as to ensure repeatability of results. The dissolved oxygen of the wastewater and turbidity of permeate were also monitored continuously at the different concentrations.

3.3 Pipe diffuser height effect experimental procedure

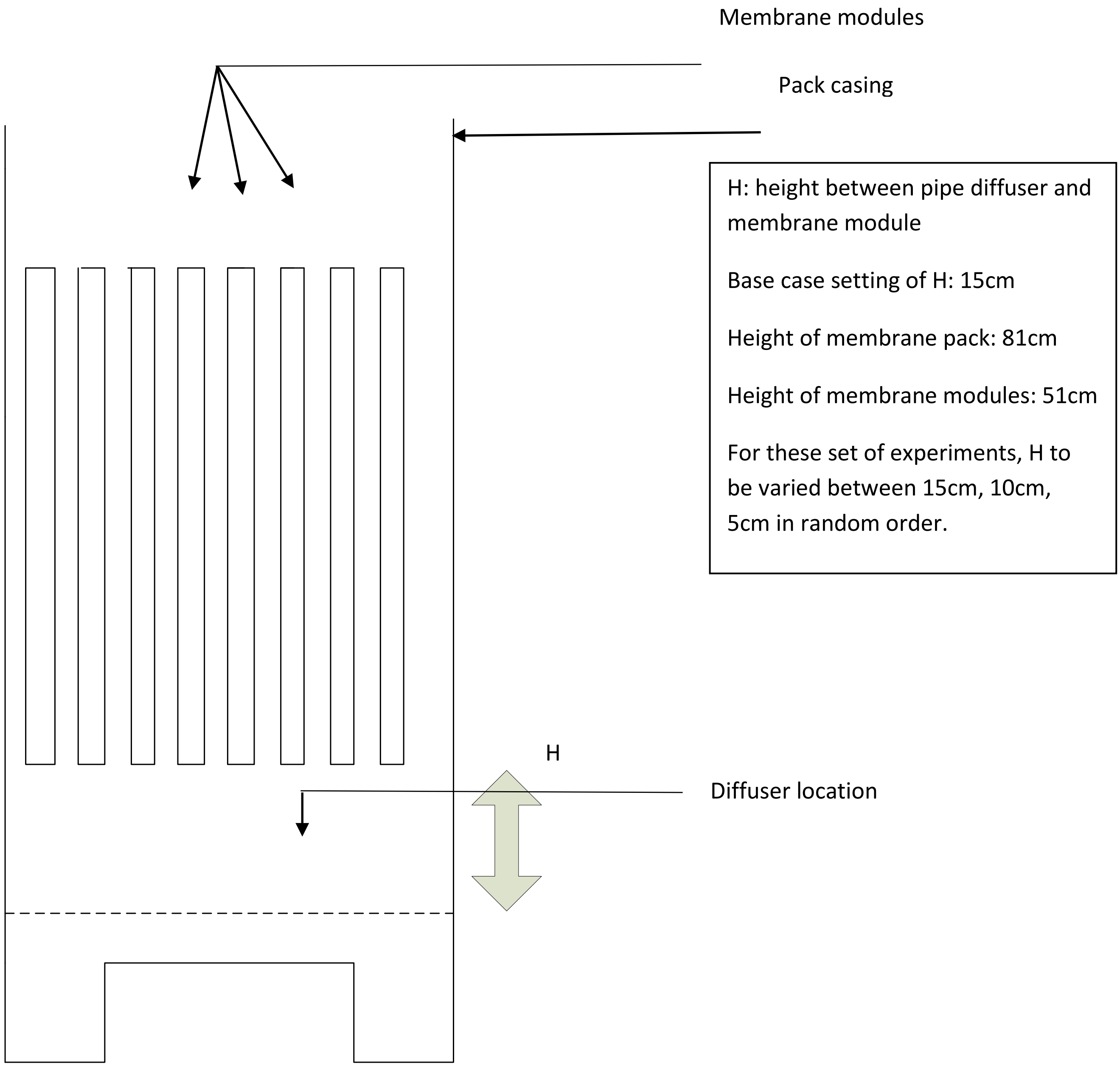


Figure 3.11: membrane pack dimensions for pipe diffuser height effect experiment

The membrane pack as illustrated in Figure 3.11 above was immersed in a bioreactor tank with feed flowing from the Veolia wastewater treatment plant by means of a submersible pump. The membrane pack had a height of 81 cm and the membrane module height is 51 cm. The experiments conducted for this part of the research, the diffusers were located at a height of 5 cm, 10 cm and 15 cm below the membrane modules with the heights being varied in random order in the course of the experimentation period. Holes through which the pipe diffusers were fitted were drilled on the membrane pack at the different heights of 5 cm, 10 cm and 15 cm below the membrane modules. The pipe diffusers had aeration hole sizes of 2 mm. At every height on the membrane pack three pipe diffusers were installed so as to ensure equal distribution of air bubbles across all the membrane modules. Aeration was done by means of a blower and the aeration rate regulated by means of rota meters.

The MLSS of feed water was 12 g/l. The water level was kept constant through continuous pumping of feed from the wastewater treatment plant. Three different aeration rates were used during this set of experiments i.e. 6 L/min per module, 7 L/min per module and 9 L/min per module. The aeration rates were chosen in a manner as to indicate the performance of the MBR at aeration rate for the critical flux chosen and also at aeration rates lower and higher than the aeration for the flux chosen. For the experiments conducted, a starting flux of 30 LMH was chosen as this is the approximate flux that most commercial MBR plants run at. The aeration rate that was used to determine this critical flux of 30LMH was 7 l/min per module.

The aeration rates used were held constant at the three different diffuser heights. The experiments done at the different diffuser heights of 5 cm, 10 cm and 15 cm were done twice so as to ensure repeatability of results. To be monitored is the TMP increase with time and also the flux drop with time.

The experiments were conducted over a four-hour period with the TMP and flux readings being taken every 20 minutes. The membrane modules were removed and cleaned by means of spraying water and scrubbing of the membrane surface at the end of every experimental run and assembled back. This was done so as to ensure integrity of results by beginning all experiments with a membrane surface that is not fouled. Permeate samples were also collected every 20 minutes to be checked for turbidity. Time was taken by means of a stop watch. The dissolved oxygen of the wastewater and turbidity of permeate were also being monitored continuously.

3.4 Comparison of disc and pipe diffuser experimental procedure

A disc diffuser was fabricated by use of PVC sheet and PVC pipes being used as the stands. Two PVC sheets were cut and glued together so as to ensure that air only comes out from the top part. A small circular hole was cut at the bottom through which air from the blower was passed. Fabrication of the disc diffuser was necessitated by the fact that the commercial disc aerators available had a larger diameter than could fit inside the membrane pack being used in the running of experiments. The fabricated disc diffuser had a diameter of 290 mm. Hole sizes for aeration purposes of 1.5 mm were drilled on the top part of the fabricated disc diffuser (as shown in the figures 3.12 and 3.13 below). Holes of an approximate area equal to that provided for by the pipe diffuser were drilled on the disc diffuser

This experiment was for the purpose of knowing if there was a difference when using a disc diffuser with fine bubbles as opposed to using a pipe diffuser and its effect on minimising fouling which is indicated by the rise in TMP as the experimental runs were being conducted.

The membrane pack was immersed in a bioreactor tank with feed flowing from the Veolia wastewater treatment plant by means of a submersible pump. The membrane pack has a height of 81 cm and the membrane module height is 51 cm. For the experiments that were conducted, the diffuser was located at a height of 10 cm below the membrane modules.

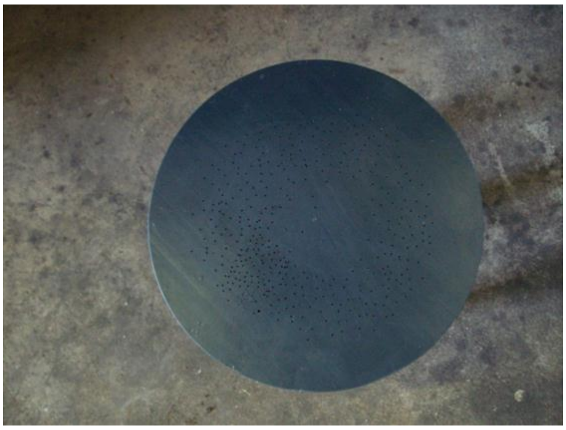


Figure 3.12: **Disc diffuser top view**

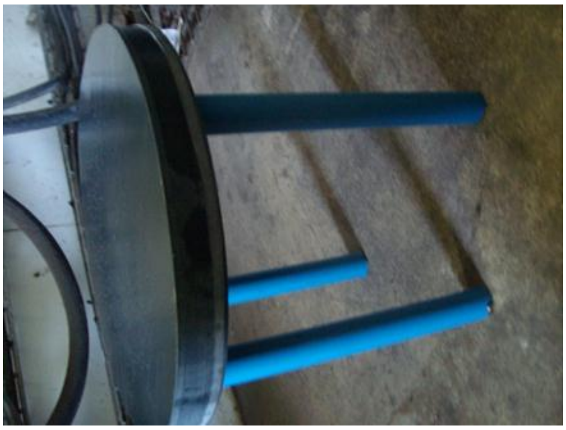


Figure 3.13: **Disc diffuser side view**

The experiments were conducted over a four-hour period with TMP and flux readings being recorded every 20 minutes. The membrane modules were removed and cleaned by means of spraying water and scrubbing of the membrane surface at the end of every experimental run and assembled back. This was done so as to ensure integrity of results by beginning all experiments with a membrane surface that is not fouled. Permeate samples were also collected every 20 minutes to be checked for turbidity. The experiments were done twice so as to ensure repeatability of results.

3.5 Membrane module spacing effect experimental procedure

The membrane module pack was assembled with the gap between the membrane modules being measured by use of a vernier calliper. The gap sizes that were investigated were 3.5 mm, 5 mm, and 6.5 mm. The base case membrane module spacing was 5 mm. This spacing of 5 mm was used due to the fact that most flat sheet commercial MBRs use this membrane module spacing. The aim of this experiment was to find out if this membrane module spacing leads to the most optimal running conditions thus varying of the module spacing by ±1.5 mm. At the end of every set of experiment at a specific gap size, the pack was disassembled and assembled again to ensure appropriate gap size between the modules. Pipe diffusers were used for aeration purposes and were located 15 cm below the membrane module which is the base case height of the diffusers. Aeration was done by means of a blower and the aeration rate regulated by means of rota meters. MLSS of feed water was 12 g/l. The water level was kept at a constant level through continuous pumping of feed from the wastewater treatment plant by use of a submersible pump.

Three different aeration rates were used during the experiments i.e. 6 L/min/module, 7 L/min/module and 9 L/min/module. Most commercial MBRs are run at aeration rates of 10 L/min/module and above. For these experiments it was decided to run at a rate lower than 10 L/min/module as the main aim of this work is to reduce the operational costs of running MBRs. The membranes were run at a starting flux of 30 LMH and the aeration rates used were held constant throughout the period of experimentation to monitor where the TMP increases with time and flux drops with time. The experiments were conducted over a four-hour period with TMP and flux readings being taken every 20 minutes. Time was measured by means of a stop watch.

The membrane modules were removed and cleaned by means of spraying water and scrubbing of the membrane surface at the end of every experimental run. This was done so as to ensure integrity of results by beginning all experiments with a membrane surface that was not fouled. Permeate samples were also collected every 20 minutes to be checked for turbidity. The dissolved oxygen of the wastewater and turbidity of permeate were also being monitored continuously. Experiments were done twice so as to ensure repeatability of results.

3.6 Operating Sequences Experimental Procedure

The membrane pack with dimensions as shown in Figure 3.14 below was used for the experimentation. The dimensions used were those found to have given the best results during the previous experiments done as explained in sections 3.2, 3.3, 3.4 and 3.5 whose main aim was to find the settings that would lead to a lower rate of fouling.

The pack was immersed in a bioreactor and fed with feed water from the Veolia wastewater treatment plant return activated sludge mixing tank by means of a submersible pump. The water level in the feed tank was kept constant by continuous pumping.

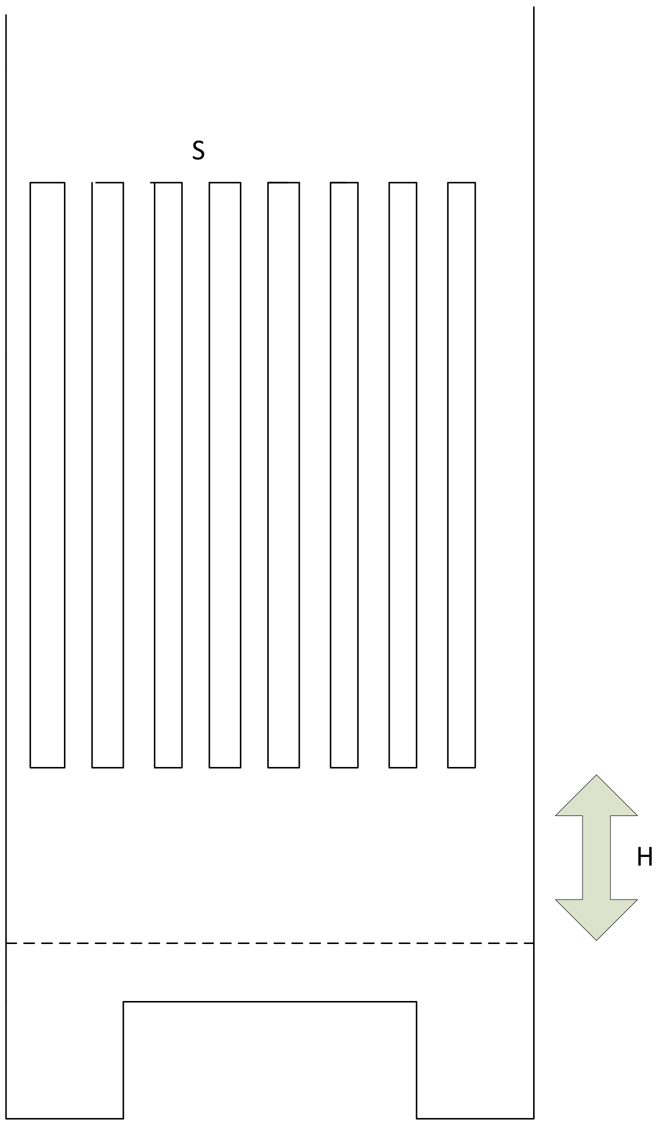
The pilot plant was run at a starting flux of 30 LMH and aeration of 7 L/Min/Module which was held constant throughout the duration of the experiments by means of a rota meter. Aeration was done by means of a blower.

The height of water above the membrane modules was held constant by continuous feeding from the wastewater treatment plant by means of a submersible pump.

The pilot plant was first run continuously without any stoppage for two weeks. After this, the effect of relaxation step and the frequency of the relaxation step were investigated. Three relaxation step frequencies were chosen i.e. 9 minutes on 1 min off, 15 minutes on 1 min off and 30 minutes on 2 minutes off. The relaxation step frequencies were chosen with regards to the commercial plants operation settings and from literature. The TMP profile with time was then observed. Timing was done by means of a stopwatch. The COD and turbidity of permeate was also analysed frequently. COD and dissolved oxygen of the feed were also analysed.

The relaxation step and frequency experiments were repeated so as to ensure repeatability of results. The data were collected over a period of four hours.

Membrane pack representation



S: membrane module gap

H: height between membrane modules and diffuser

No of Modules: 18

S: 5mm

H: 5cm

Diffuser Type: Pipe

Figure 3.14: membrane pack representation showing dimensions used for operating sequence experimental procedure

CHAPTER FOUR: RESULTS AND DISCUSSIONS

4.1 Critical Flux

A series of curves were generated for critical flux at different aeration rates and at different MLSS concentrations.

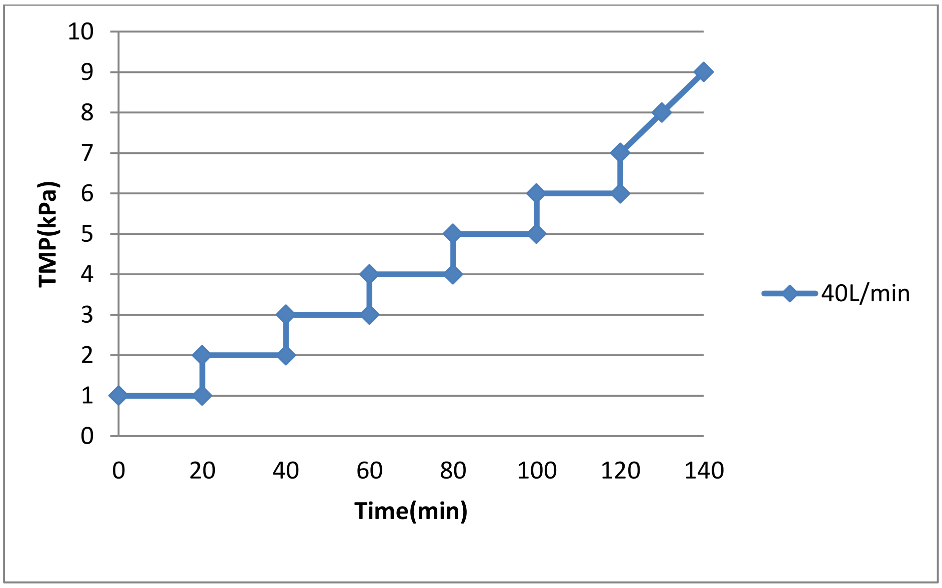


Figure 4.1: Critical flux at MLSS of 4 g/l, aeration of 40 L/min (2.22 l/min per module) determined to be 21 LMH.

A total of 20 curves such as those shown in Figure 4.1 above were generated for the critical flux at the different concentrations and aeration rates. The experiments were also repeated so as to ensure repeatability of results.

A curve of average critical flux vs. aeration rates was also generated for the different MLSS concentrations as shown in Figure 4.2.

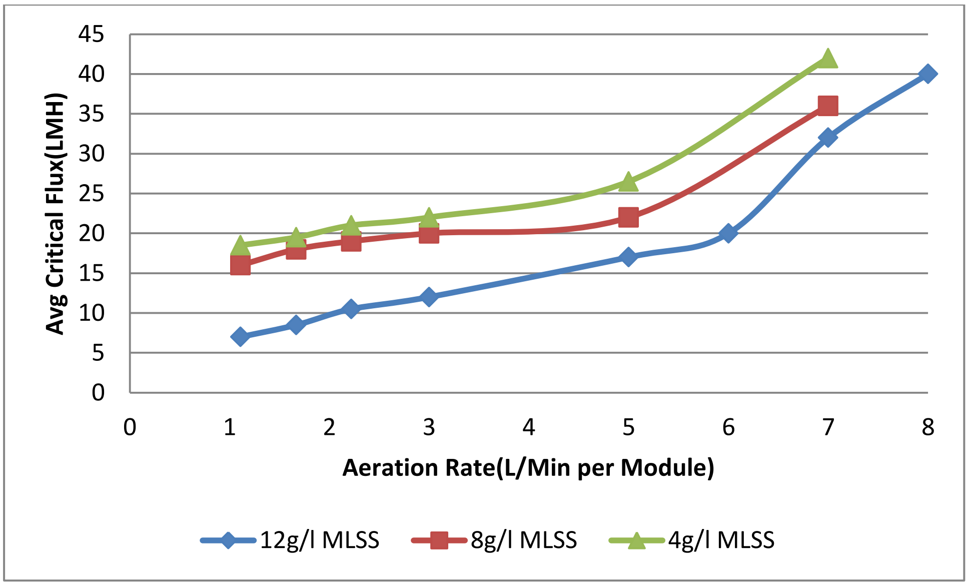


Figure 4.2: Graph of average critical flux vs. aeration rate

Discussion

The membrane modules used in the experimental runs were 18 in number. From the curve of average critical flux vs. aeration rates as shown in Figure 4.2, it can be seen that the critical flux increases with an increase in the aeration rates. It can also be seen that the critical flux increased with a decrease in MLSS.

This is explained by the fact that at higher aeration rates minimal fouling occurs as the air scours away the foulants from the membrane surface at a faster rate in comparison to low aeration rates. This is consistent with literature that says critical flux increases with an increase in aeration rate (Judd, 2011, Pollice et al., 2005, Cele and Pillay).

It is further expected that at some point even with an increase in aeration rate, the critical flux will not increase as the rate of deposition of foulants on the membrane material will be in equilibrium with the rate of scouring of the foulants by the air bubbles. However, in this experiment this was not explored due to limitation of the blower and the rota meters that were used.

4.2 Pipe diffuser height effect

A series of TMP vs. time and flux vs. time curves as shown in Figures 4.3, 4.4, 4.5, 4.6, 4.7 and 4.8 below were generated for the different aeration rates at the different diffuser heights that were investigated.

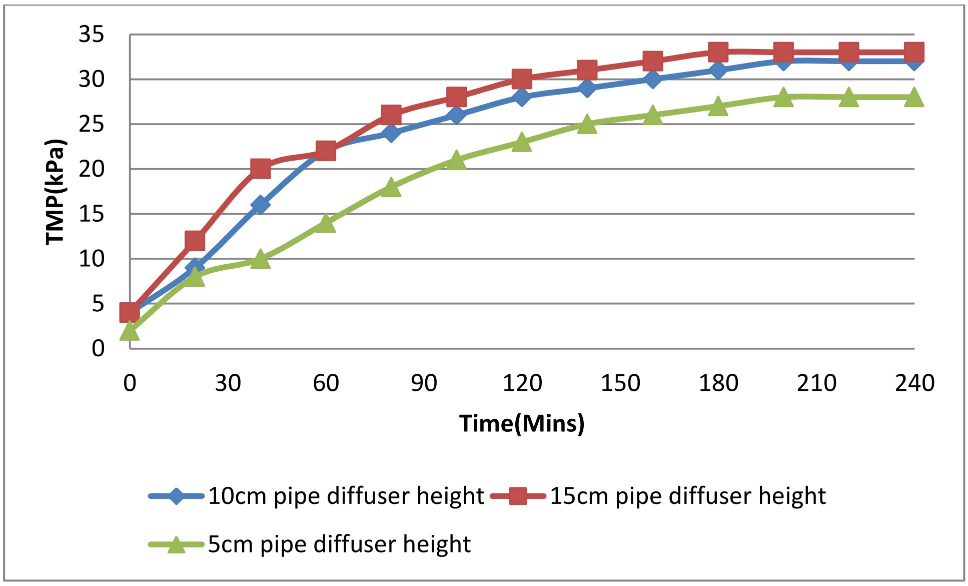


Figure 4.3: TMP vs. time at 7 L/Min per Module aeration

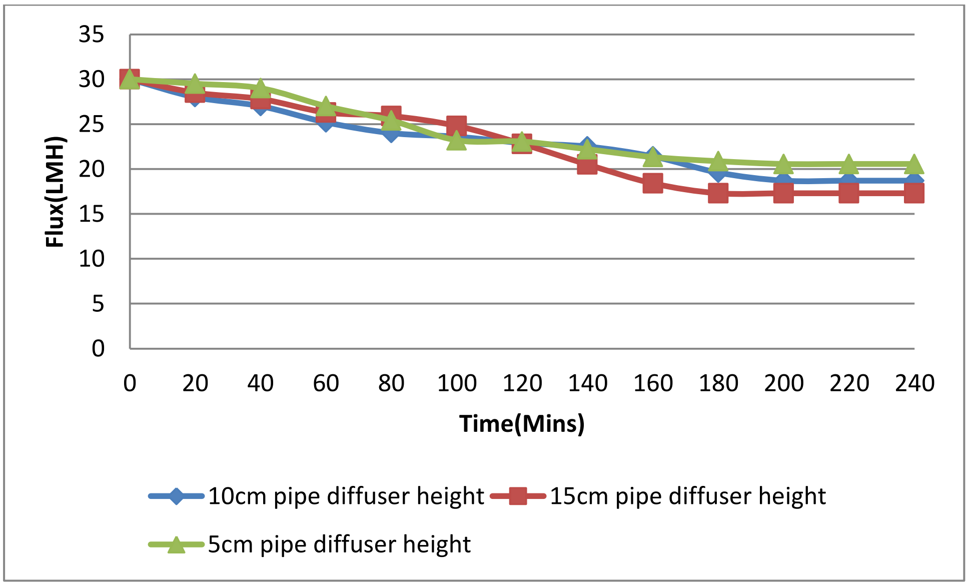


Figure 4.4: Flux vs. Time at 7 L/Min per Module aeration

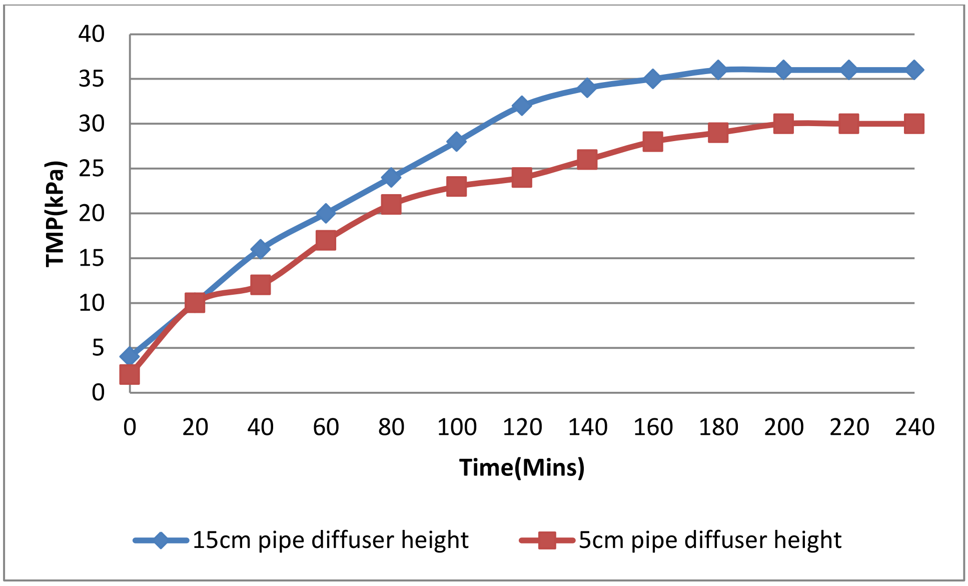


Figure 4.5: TMP vs. Time at 6 L/Min per Module aeration

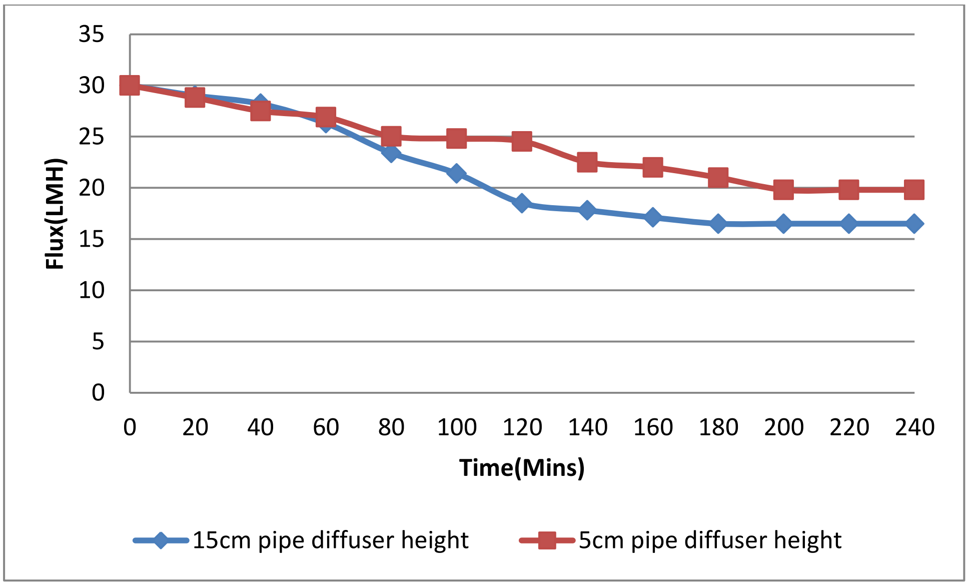


Figure 4.6: Flux vs. Time at 6 L/Min per Module aeration

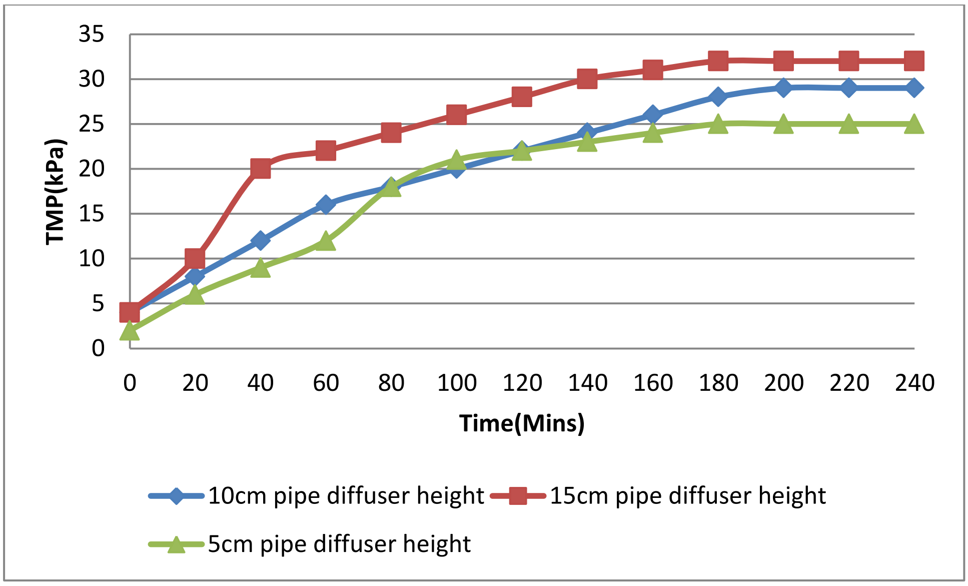


Figure 4.7: TMP vs. Time at 9 L/Min per Module aeration

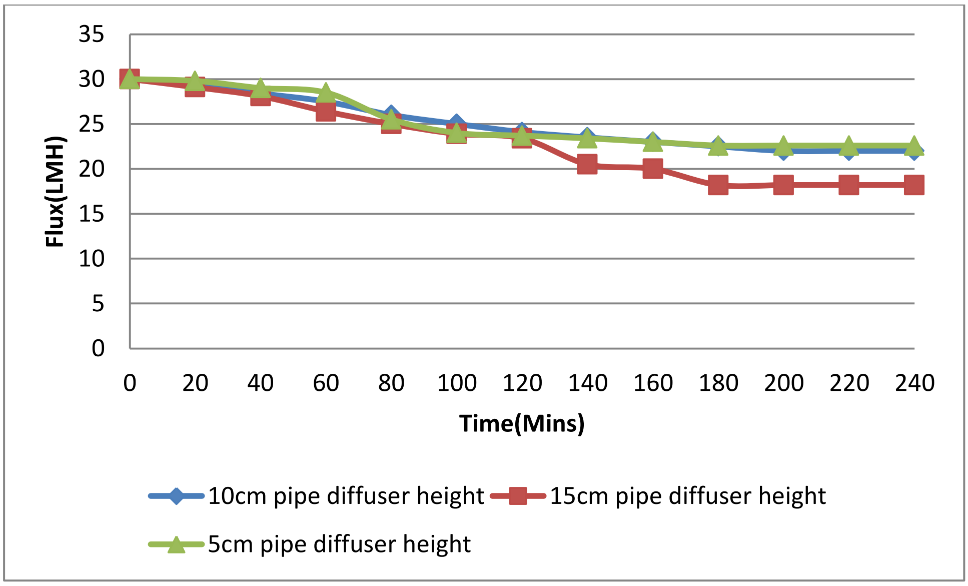


Figure 4.8: Flux vs. Time at 9 L/Min per Module aeration

Discussion

From Figures 4.3, 4.5, 4.7 shown above, it can be seen that the TMP rises until it reaches a point at which it stabilizes while Figures 4.4, 4.6, 4.8 show reduction in flux up to a point at which it stabilizes for the period of the experimental runs carried out. This can be adduced to the initial accumulation of fouling layer on start-up which can occur even under sub critical flux operation as has been highlighted in the literature review section of this work.

Increase in the aeration rates results in a lower point at which the TMP stabilizes at and also a higher point at which the flux stabilizes at for the different pipe diffuser heights that were being experimented on. This shows the increase in effectiveness of the air bubbles in scouring away the foulants from the membrane surface.

Decrease in height between the membrane modules and the pipe diffusers shows lower TMP at which the suction pressure stabilizes and also higher point at which the flux stabilizes. This can be explained by the increasing effectiveness of the air bubbles at scouring the membrane surface with the increase in aeration rates and also decrease in height between the membrane modules and the pipe diffusers.

The stable TMP can be deduced as the point at which the rate of scouring away the foulants is equal to the rate at which the foulants are deposited to the membrane surface. The higher the aeration rate and the lower the height between the membrane modules and the pipe diffusers the lower the point at which the TMP stabilizes showing that the membrane is less fouled.

The turbidity of permeate sampled were all less than 1NTU by the end of each experimental run which is considered satisfactory for drinking water.

In the conclusion of Clech et al. (2003), it has been shown that even under sub critical conditions there is still some fouling that occurs. This explains why at the beginning of every run there is an increase in TMP and a subsequent drop in flux before they stabilise.

Conclusion

From the data curves shown above, it can be seen that the best results are obtained at a pipe diffuser height of 5cm below the membrane module. This is the height that will be used when running long term experimentation.

4.3 Comparison of disc and pipe diffuser

A series of TMP vs. Time and Flux vs. Time curves were generated for the different aeration rates as shown in Figures 4.9, 4.10, 4.11, 4.12, 4.13, 4.14 and 4.15 below.

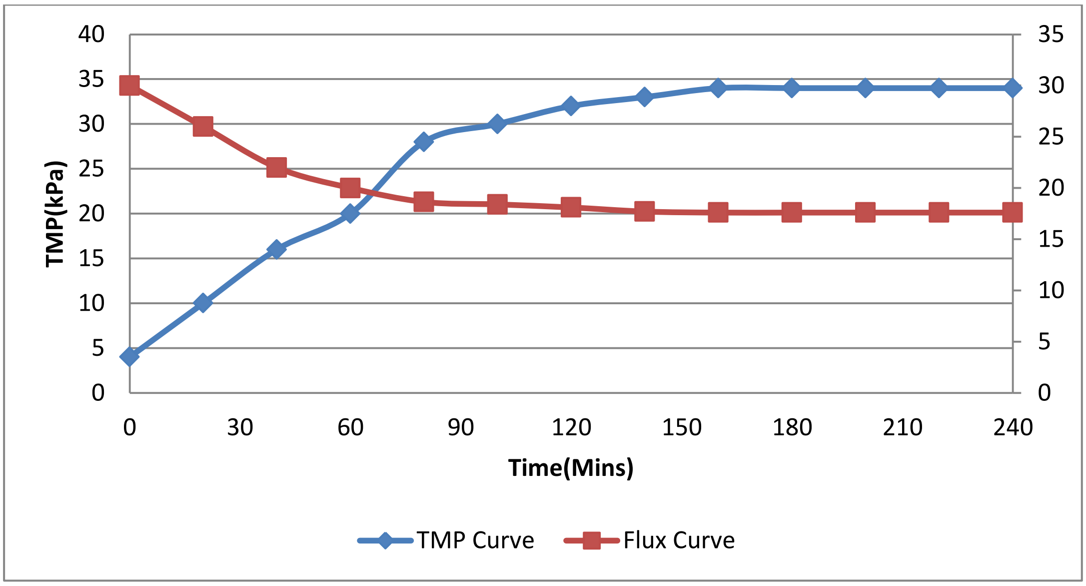


Figure 4.9: TMP & Flux vs. Time Chart 9L/Min per Mod aeration, 10cm disc diffuser height

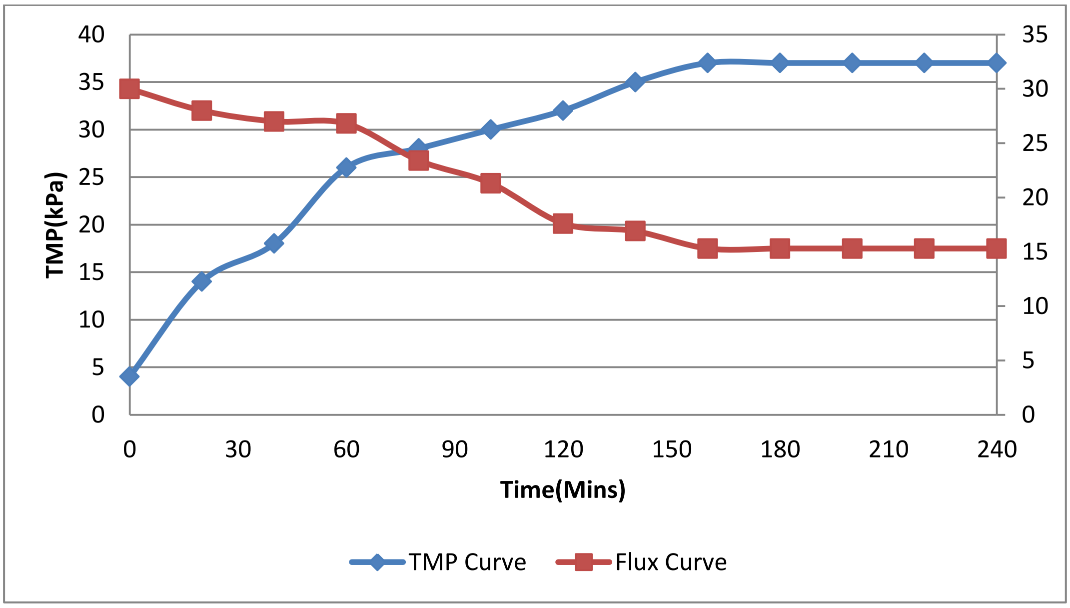


Figure 4.10: TMP & Flux vs. Time Chart 7L/Min per Mod aeration, 10cm disc diffuser height

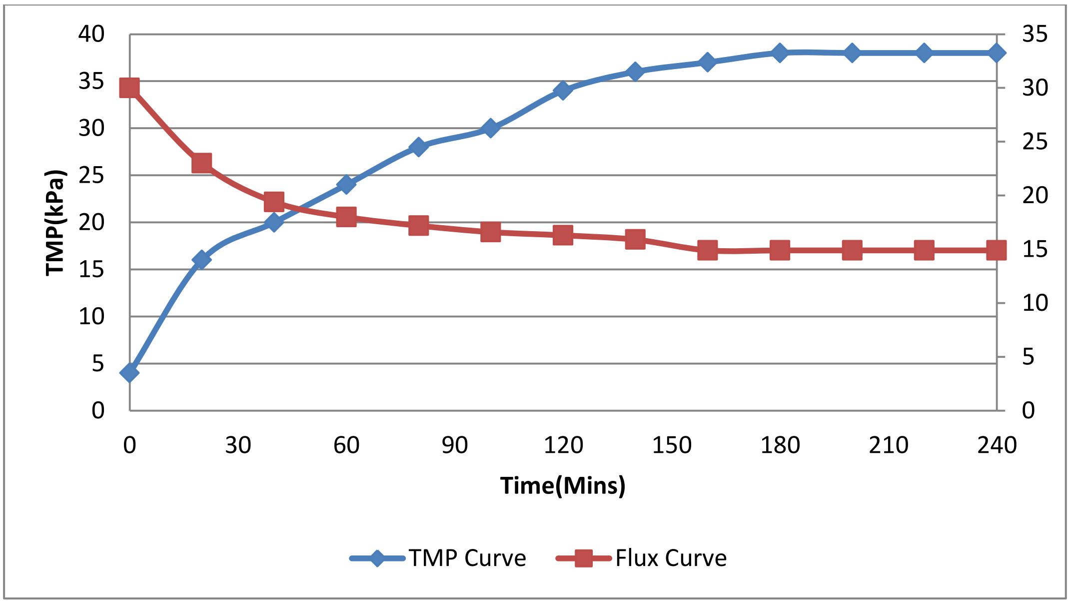


Figure 4.11: TMP & Flux vs. Time Chart 6L/Min per Mod aeration, 10cm disc diffuser height

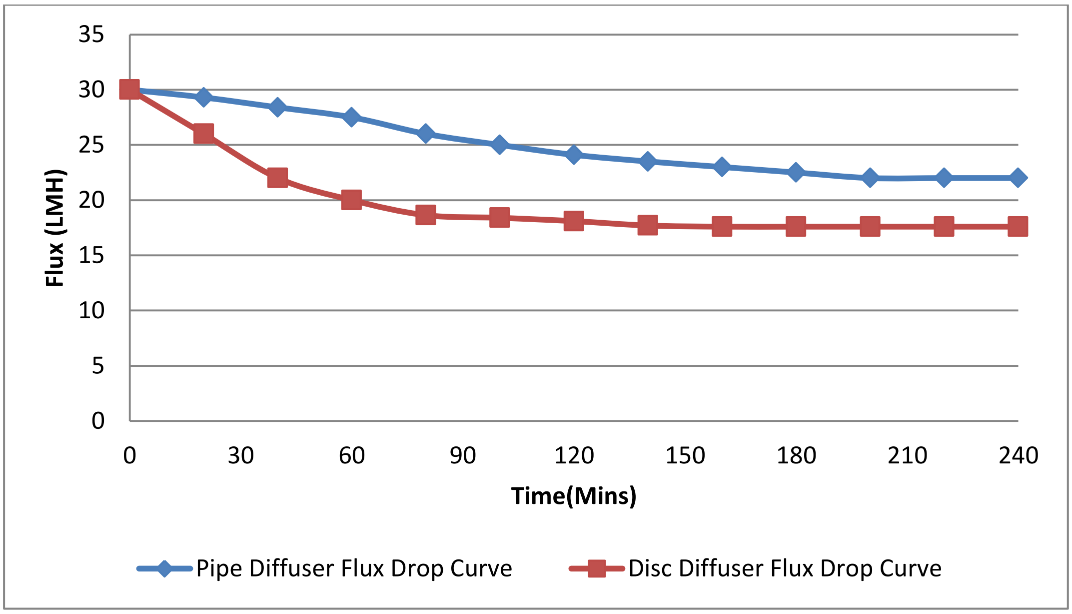


Figure 4.12: Flux vs. Time Comparison Chart 9L/Min per Mod Pipe & Disc Diffuser

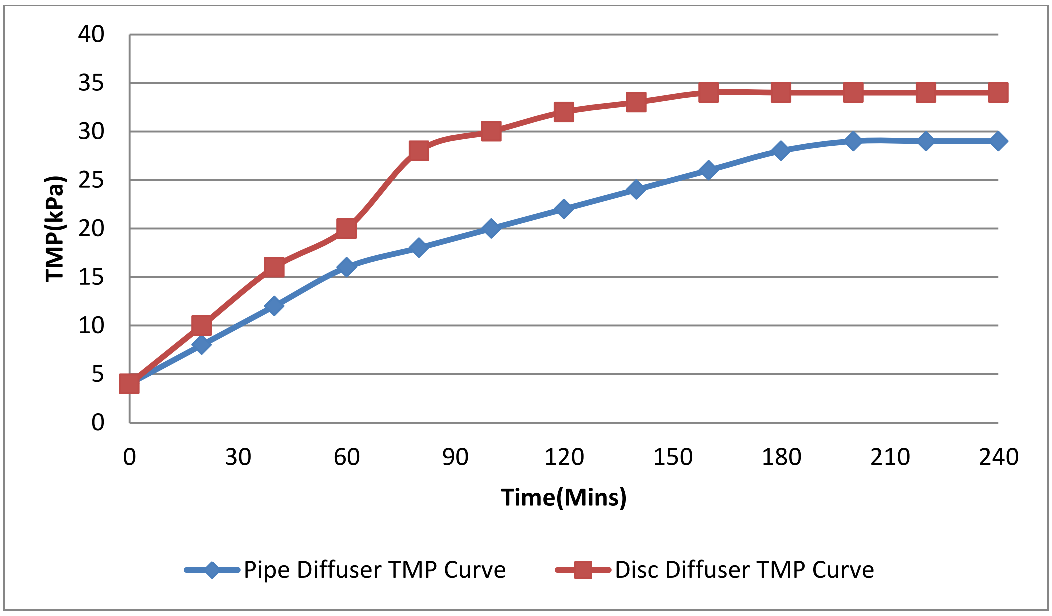


Figure 4.13: TMP vs. Time Comparison Chart 9L/Min per Mod Pipe & Disc Diffuser

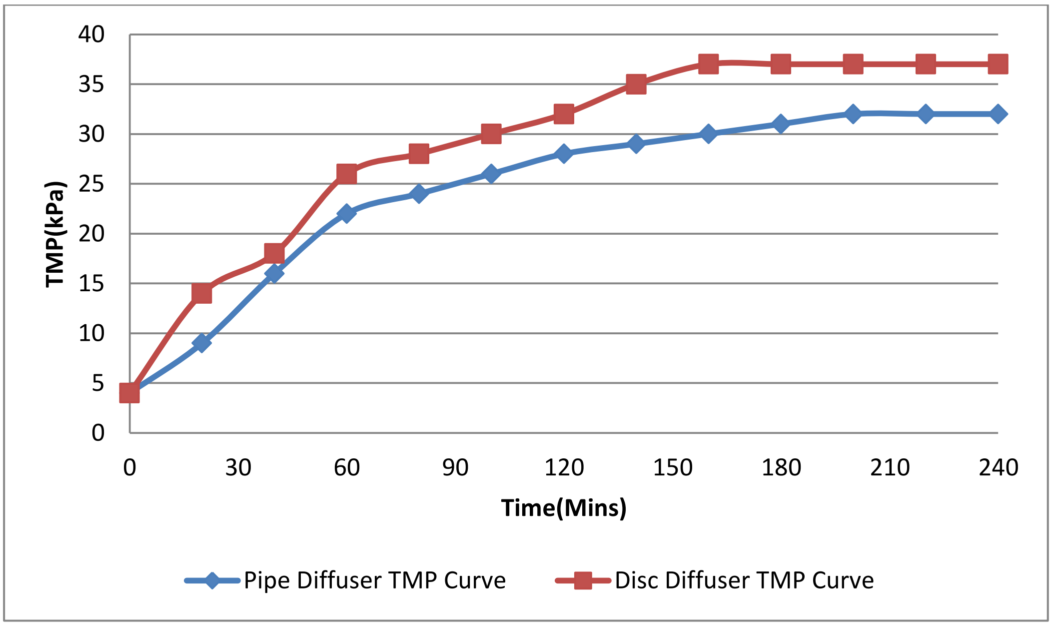


Figure 4.14: TMP vs. Time Comparison Chart 7L/Min per Mod Pipe & Disc Diffuser

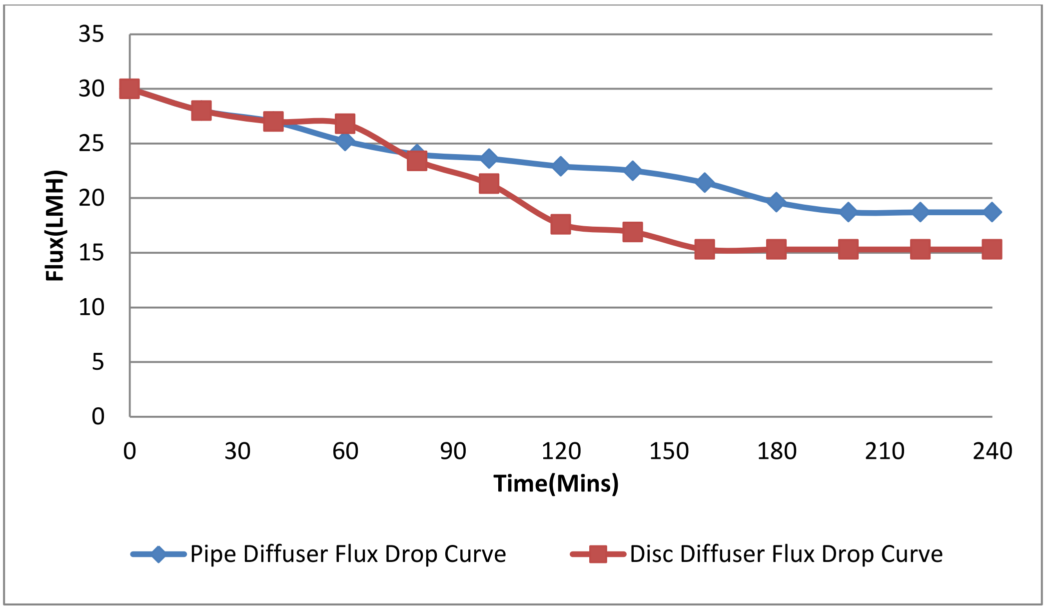


Figure 4.15: Flux vs. Time Comparison Chart 7L/Min per Mod Pipe & Disc Diffuser

Discussion

From the Figures 4.9, 4.10 and 4.11 above showing TMP & Flux vs. Time, it is seen that the TMP rises then it reaches a point at which it stabilizes and the flux drops with time before it stabilises. This can be adduced to the initial accumulation of fouling layer on start-up which can occur even under sub critical flux operation as has been highlighted on in the literature review section of this work.

Increase in the aeration rates leads to a lower point at which the TMP raises and stabilizes at. This can be explained by the increasing effectiveness of the air bubbles at scouring the membrane surface with the increase in aeration rates. The stable TMP can be deduced as the point at which the rate of scouring away the foulants is equal to the rate at which the foulants are deposited to the membrane surface.

Figures 4.12, 4.13, 4.14 and 4.15 show TMP vs. Time and Flux vs. Time comparison charts. In these Figures, it is shown that the performance of the pipe diffuser is better in comparison to that of the disc diffuser. This could be due to the fact that the fine bubbles generated by the disc diffuser are of insufficient surface area to effectively scour the membrane surface of foulants.

It can be further investigated to see if increasing the number of aeration holes on the disc diffuser leads to better reading in comparison to the pipe diffuser.

Conclusion

From the curves above it can be concluded that the performance of the pipe diffuser is better than for the disc diffuser that was fabricated hence it is recommended that pipe diffusers be used for experimentation.

4.4 Membrane module spacing effect

A series of TMP and Flux vs. Time curves were generated as shown in Figures 4.16, 4.17, 4.18, 4.19, 4.20 and 4.21.

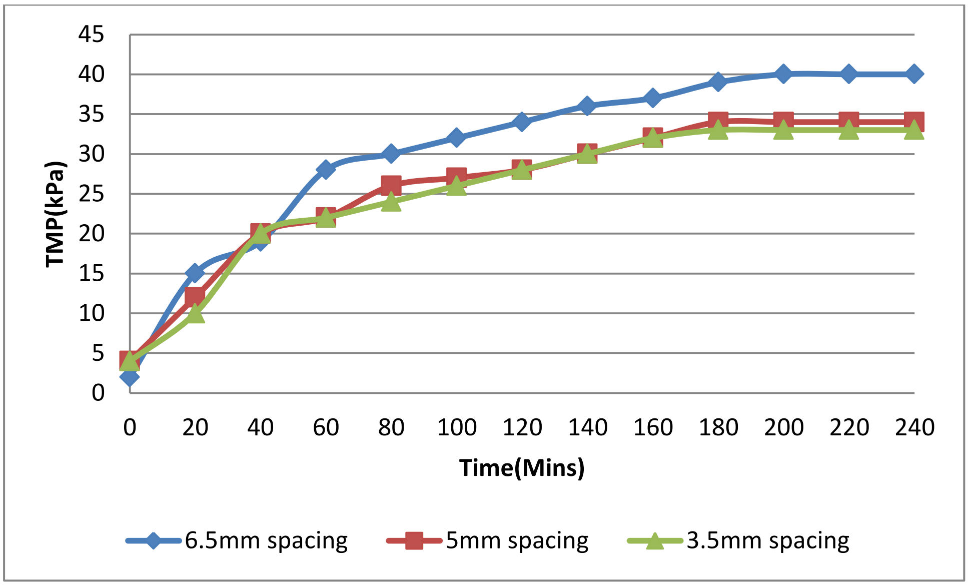


Figure 4.16: TMP vs. Time at 7L/Min per Module aeration

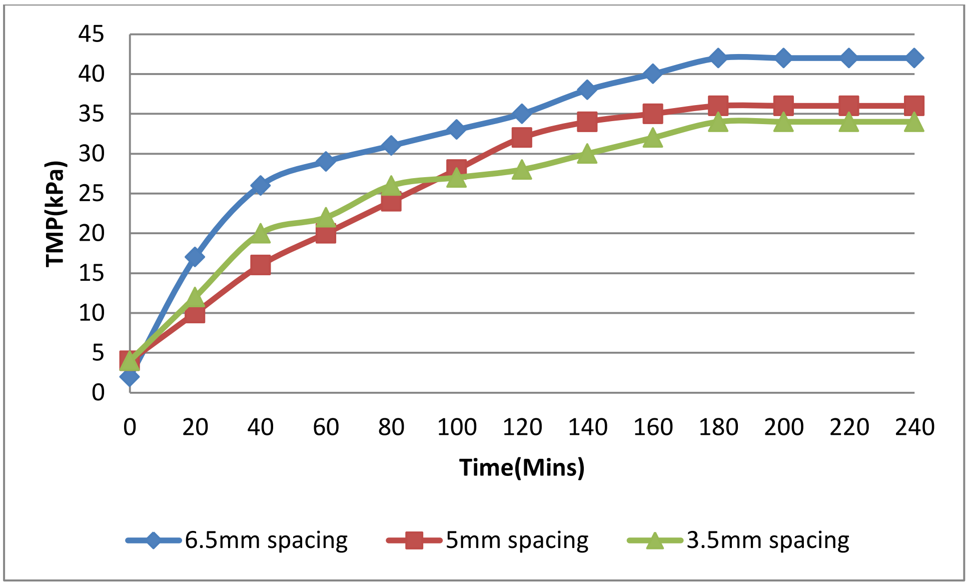


Figure 4.17: TMP vs. Time at 6L/Min per Module aeration

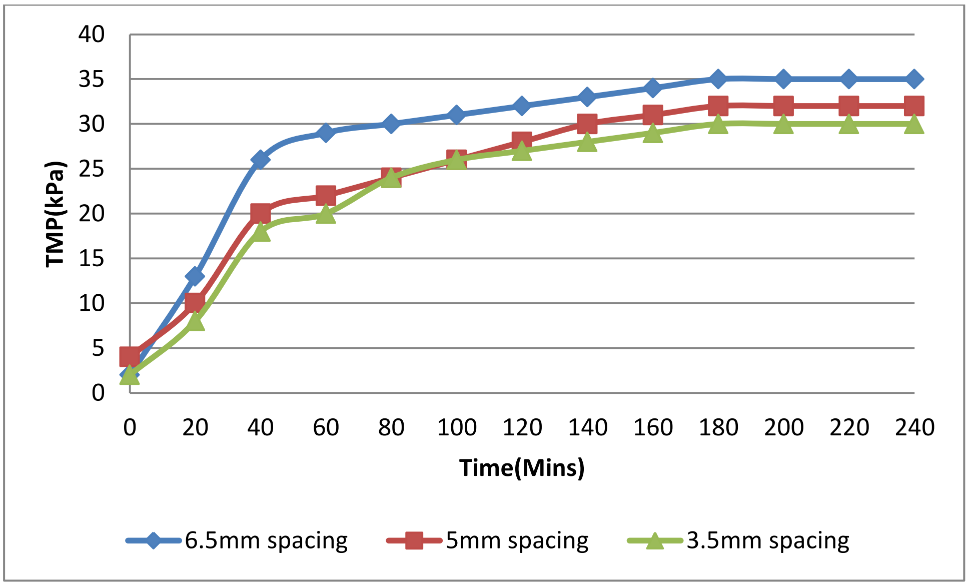


Figure 4.18: TMP vs. Time at 9L/Min per Module aeration

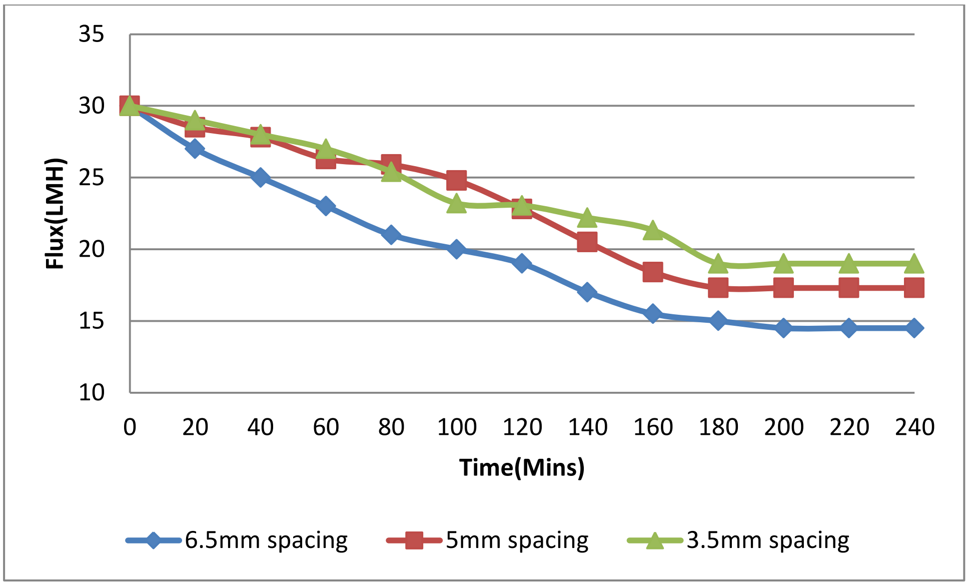


Figure 4.19: Flux vs. Time at 7L/Min per Module aeration

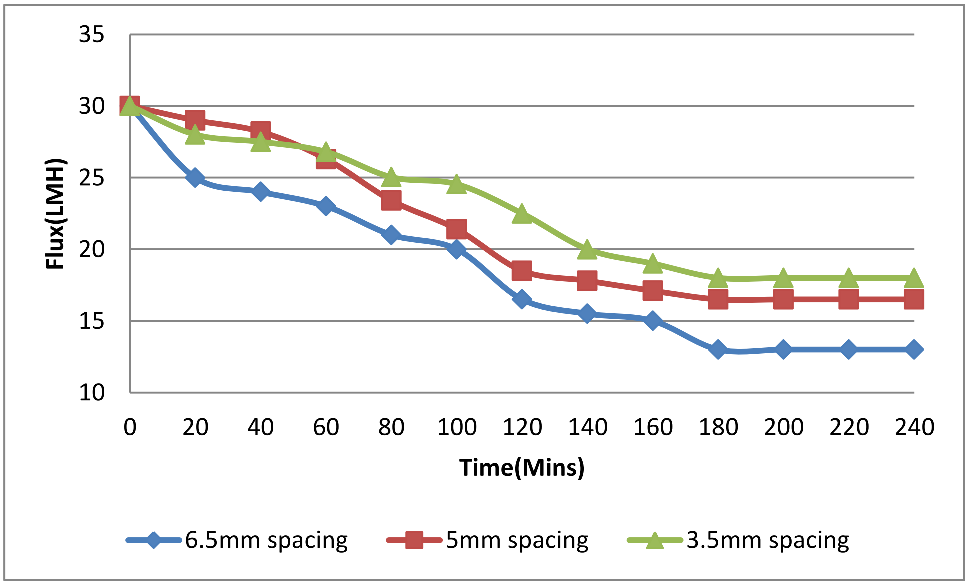


Figure 4.20: Flux vs. Time at 6L/Min per Module aeration

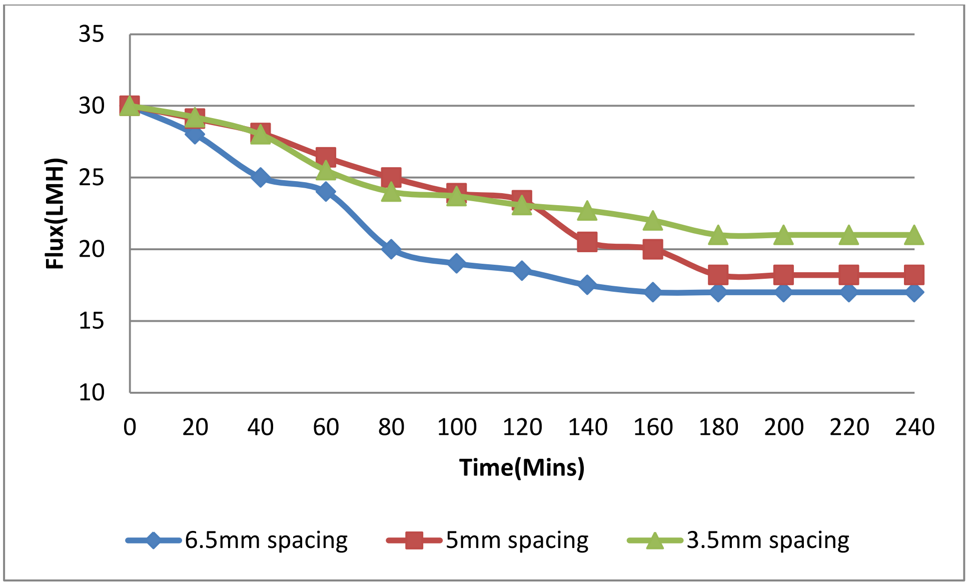


Figure 4.21: Flux vs. Time at 9L/Min per Module aeration

Discussion

From Figures 4.16, 4.17 and 4.18 above, the TMP is seen to rise before eventually stabilizing while from figures 4.19, 4.20 and 4.21, the flux is seen to decrease with time before eventually stabilizing. From the trends shown in the curves, it can be deduced that at this point of stabilisation, the rate of deposition of foulants onto the membrane surface is equal to the rate at which the air bubbles scour away the foulants.

It can also be seen from the above charts that with an increase in aeration rate, there is a decrease at the point at which the TMP stabilises and an increase at the point at which flux stabilises. This shows the increasing effectiveness of the aeration with an increase in the aeration rate.

When increasing the membrane module spacing to 6.5mm, the TMP readings are seen to increase and the opposite when the gap is reduced to 3.5mm. This is because when increasing the gap size, the air bubbles are less effective in scouring the membrane surface due to the reduced contact area between the air bubbles and the membrane surface. When the gap size is reduced the bubbles are more effective due to the increased contact area that they have with the membrane surface hence increased effectiveness in scouring away the foulants from the membrane surface.

The turbidity of permeate was less than 1NTU for the different experimental runs undertaken.

Conclusion

Best results in regard to TMP and Flux are obtained at membrane module spacing of 3.5mm. However, at this spacing, the modules are not held in place hence can be easily moved about. For this reason, membrane module spacing of 5mm is to be used as the membrane panels are fixed in place, is easier to operate at and will thus give more accurate readings when being used in long term operations.

4.5 Operating sequences results

Figure 4.22 below shows data chart for continuous running of pilot plant.

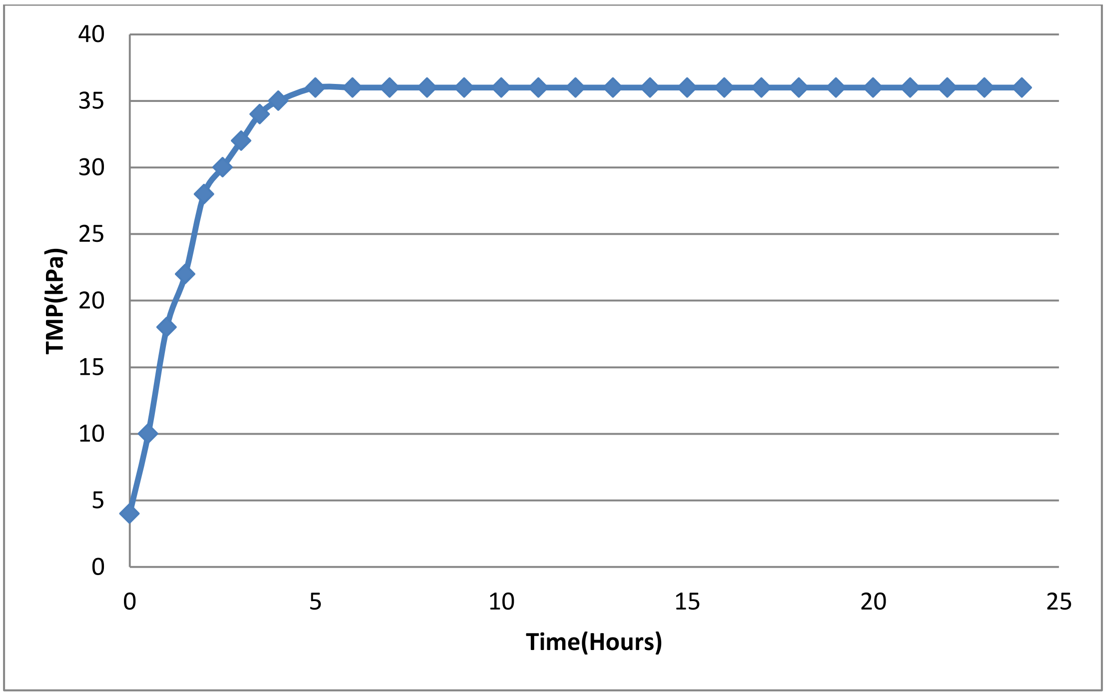


Figure 4.22: TMP vs. Time Chart during continuous operation

The Figures 4.23, 4.24 and 4.25 below show data charts for various relaxation step frequencies.

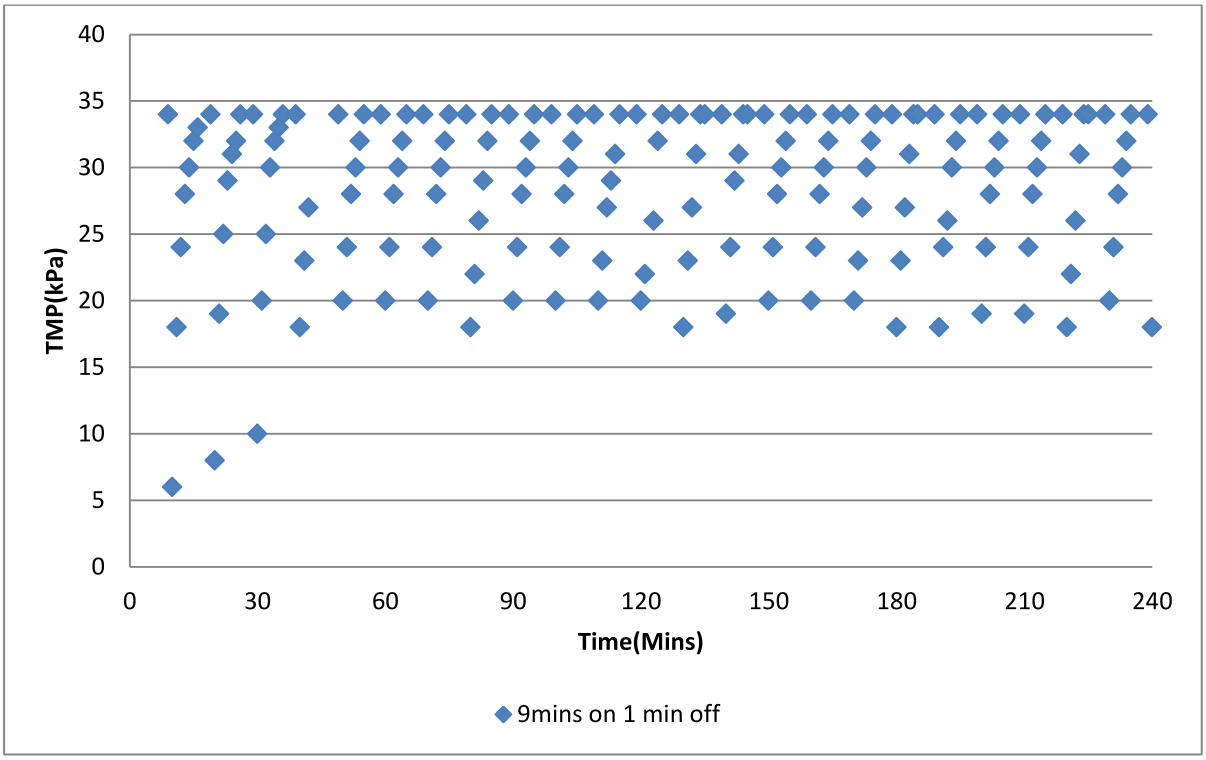


Figure 4.23: 9minutes on 1 min off relaxation step TMP vs. Time Chart

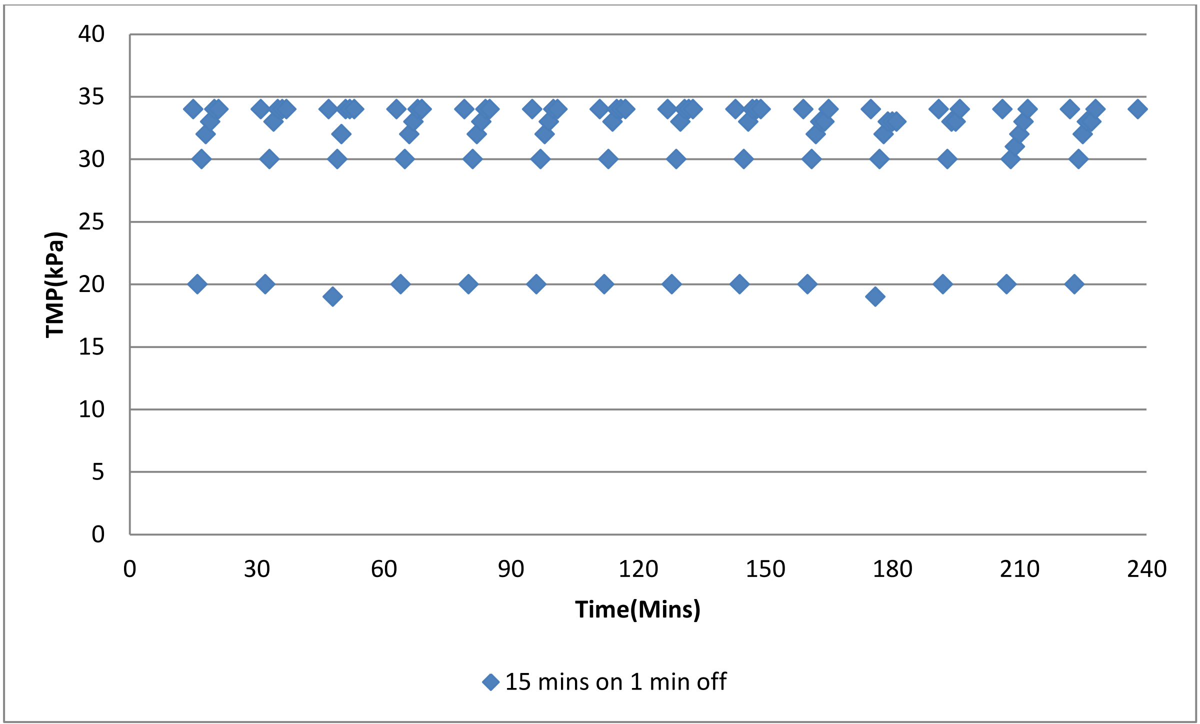


Figure 4.24: 15 minutes on 1 min off relaxation step TMP vs. Time Chart

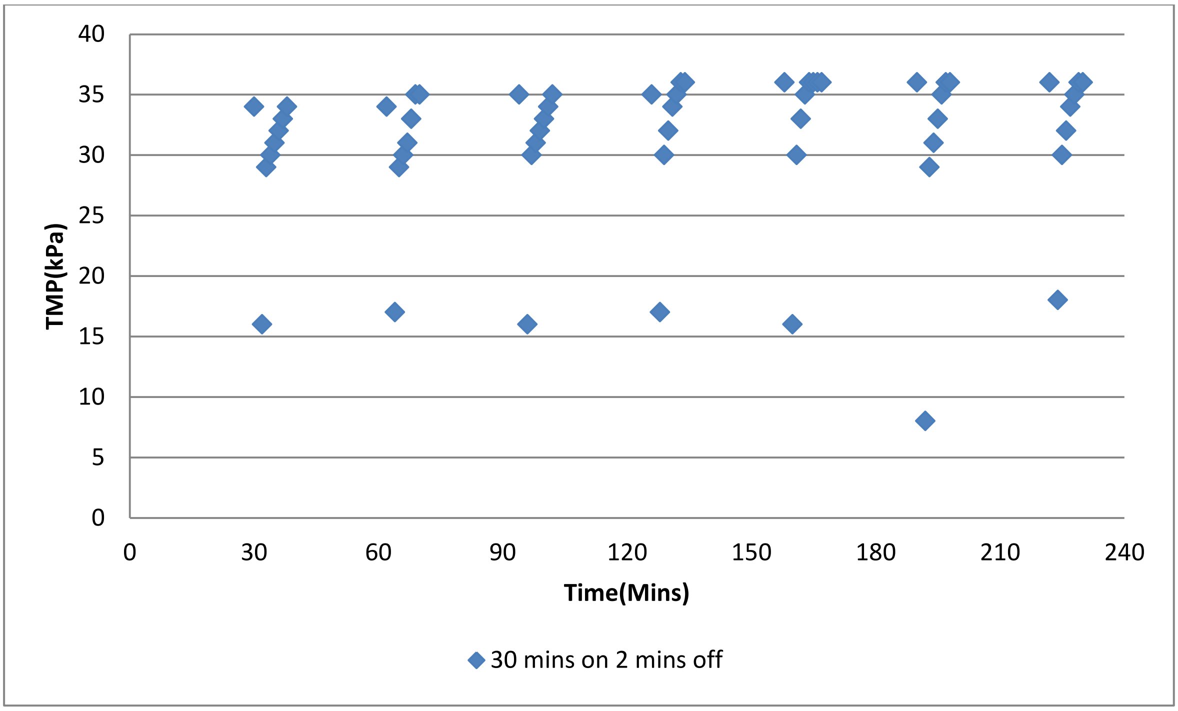


Figure 4.25: 30 minutes on 2 minutes off relaxation step TMP vs. Time Chart

Figure 4.26 below shows the permeate turbidity for the samples collected during the running of the pilot plant.

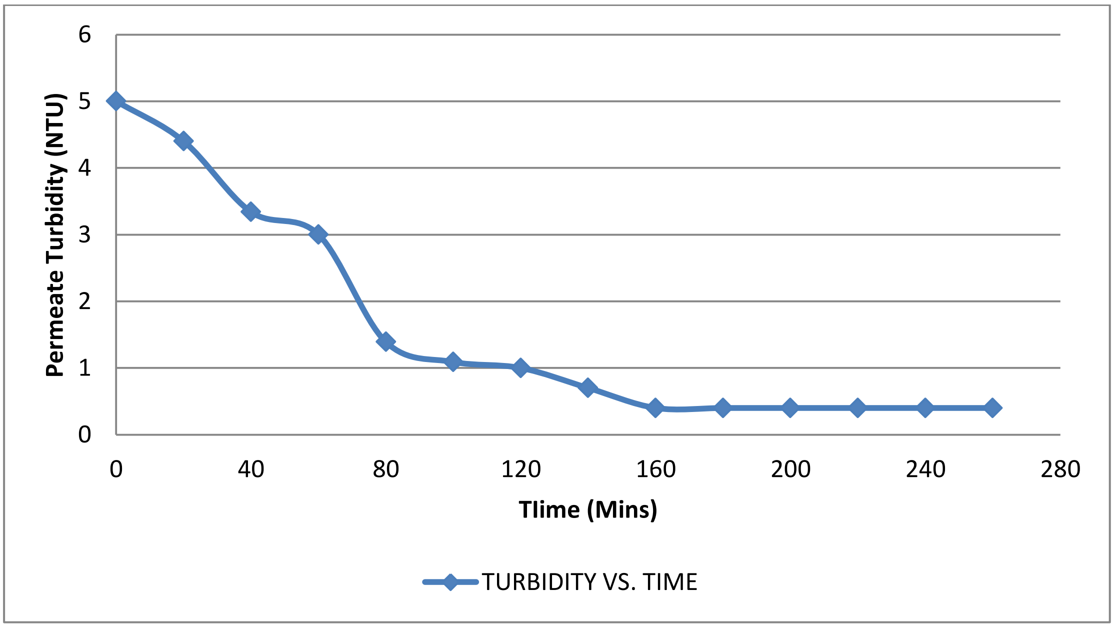


Figure 4.26: Turbidity vs. Time chart for long term experiments

Discussion

During the continuous-running experimental phase, it was noted that TMP stabilized at 36 kPa for the duration of the experiment as is indicated in Figure 4.22, which is a sample reading showing the trend of the TMP profile.

It was also noted during this stage of experimentation that the TMP for the continuous run is higher in comparison to when running with the relaxation steps as has been attested to by Wu et al. (2008a) and Braak et al. (2011a) and as shown in Figures 4.23 and 4.24. However in Figure 4.25 with the relaxation step of 30 minutes on 2 minutes off, it was noted that the final TMP was higher that when conducting the continuous running which can be adduced as the relaxation step of 30 minutes on and 2 minutes off as being ineffective in reducing fouling resulting in higher TMPs. This is in line with what has been covered in the literature review showing that too long a period between relaxation times is ineffective in reducing fouling.

During the experimental runs, permeate samples were collected for testing of COD removal rate and the turbidity. The COD removal rate for the woven fibre microfiltration MBR was found to be greater than 95% which meets the South African standards for wastewater treatment. The turbidity of the permeate was found to be less than 1 NTU for all samples collected which adheres to the South African and WHO standards.

Conclusion

Long periods between relaxation steps is not an effective way of reducing fouling as shown when using relaxation sequence of 30 mins on, 2 mins off.

CHAPTER FIVE: CONCLUSIONS AND RECOMMENDATIONS

Conclusions

- Critical flux values increase with an increase in aeration rate which is consistent with literature. A starting flux of 30LMH was decided upon for running of the pilot plant on subsequent investigations.
- For the pipe diffuser investigations, the best results are obtained at a pipe diffuser height of 5cm below the membrane module due to the lower TMP achieved which was used when running long term experimentation.
- When comparing the performance of the fabricated disc diffuser to the pipe diffuser, the performance of the pipe diffuser is better than of the disc diffuser. The pipe diffusers were used for experimentation.
- For the membrane module spacing effect, the best results in regard to TMP and Flux are obtained at membrane module spacing of 3.5mm. However, at this spacing, the modules are not held in place hence can easily move about especially when running for long periods of time. Hence membrane module spacing of 5mm were when running long term as the membrane modules are fixed in place, is easier to operate at and the difference in performance in comparison with the 3.5mm spacing is not that great.
- For the period of time when the plant was running continuously the TMP held steady. Use of Relaxation step as a means of minimising fouling was investigated. One of the results of this investigation found that long periods between relaxation steps are not an effective way of reducing fouling on the membrane surface as is shown when using the relaxation sequence of 30 mins on, 2 mins off. Most effective relaxation sequence in terms of the time taken for the TMP to climb to the stabilization point is the relaxation sequence of 9 mins on, 1 min off.
- COD reduction was found to be ≥95% for all samples taken. The turbidity of permeate was less than 1NTU for all samples taken. DO was greater than 2mg/l.

Recommendations

Further studies should be done on use of the disc diffuser with increased surface area of aeration holes and also hole sizes of smaller diameters to check on its effectiveness as a means of reducing fouling on the membrane surface.

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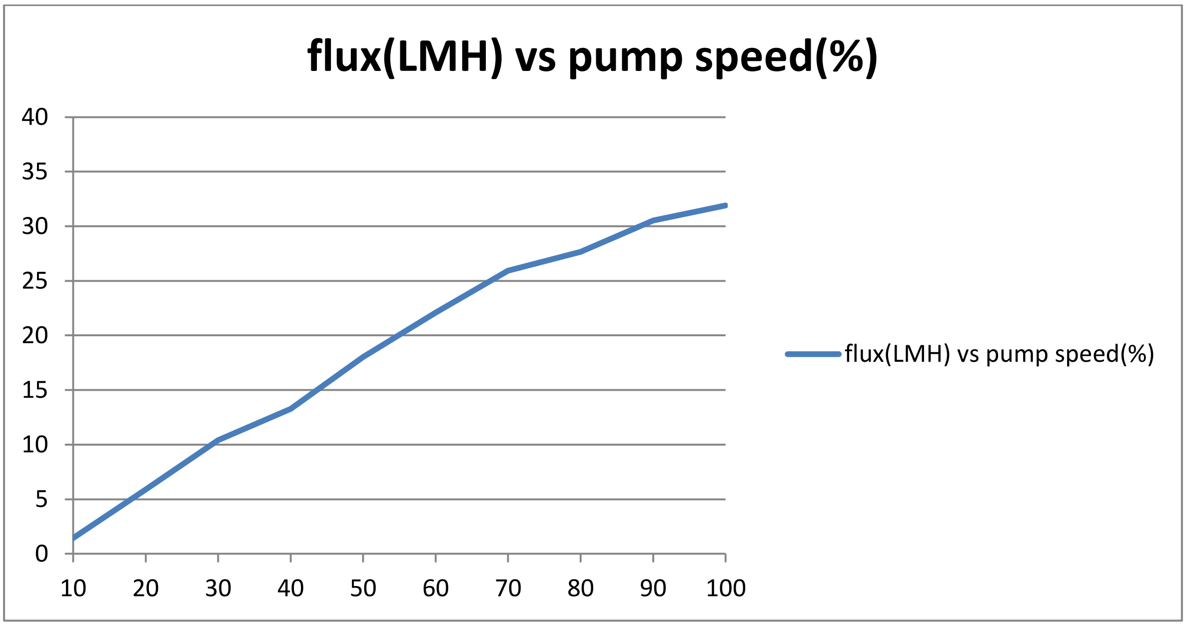
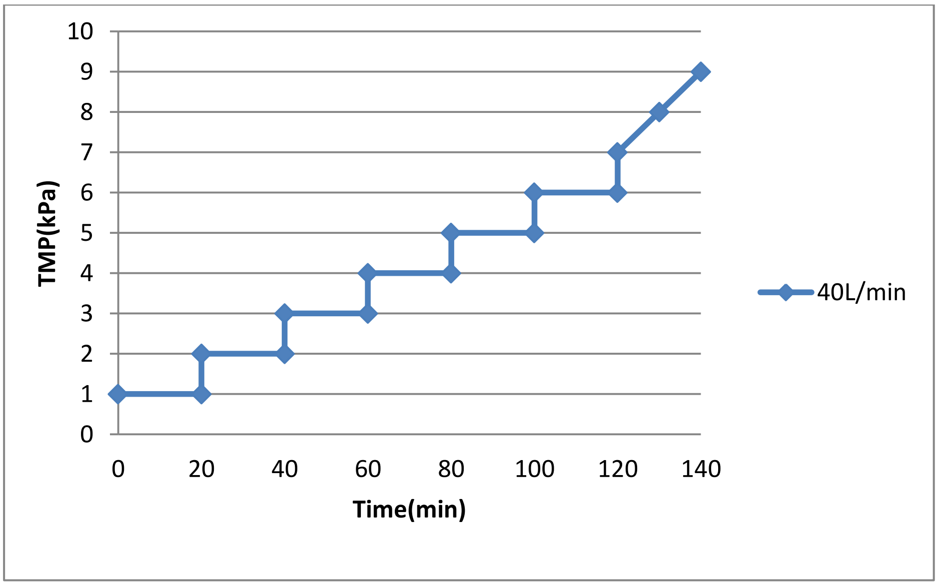
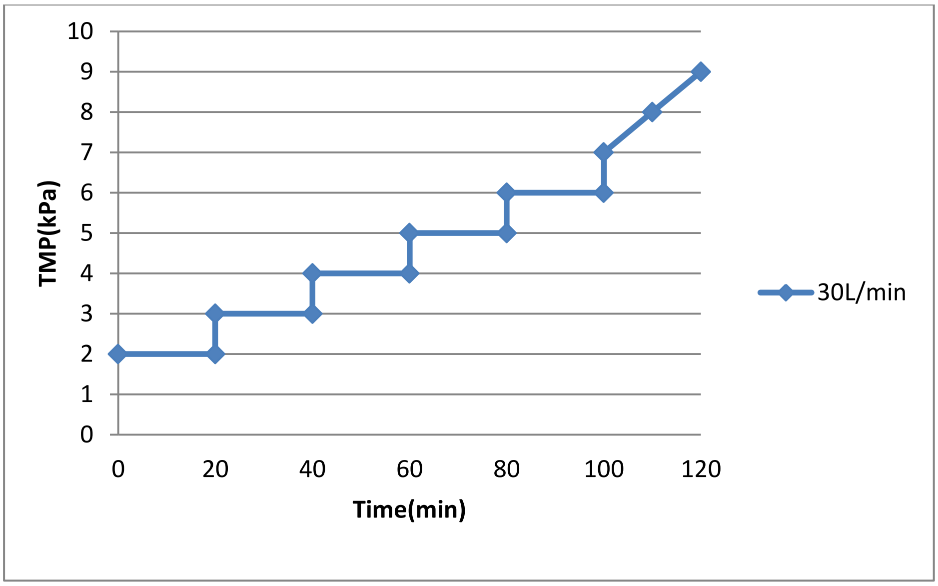
APPENDIX

Calibration chart for pump speed vs flux

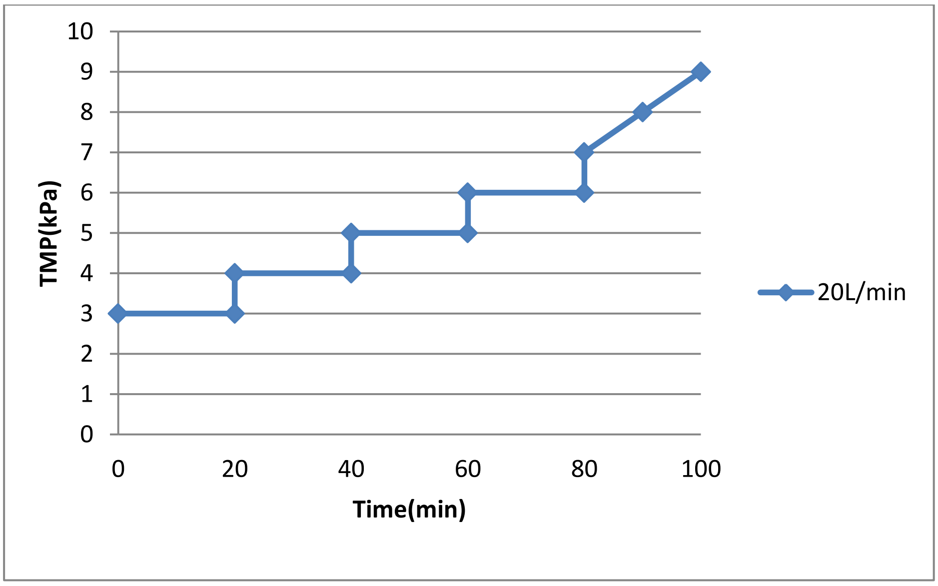




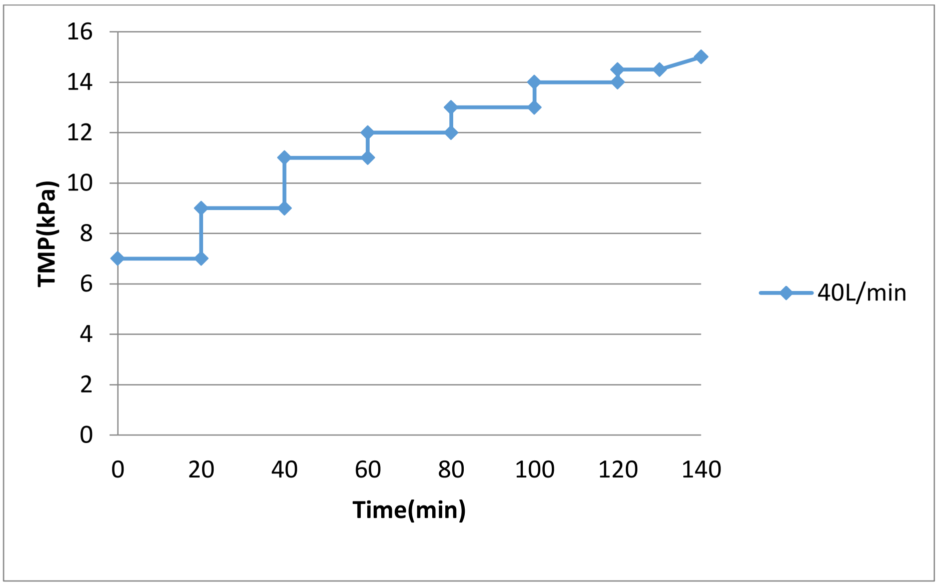
Critical flux at MLSS of 4g/l, aeration of 40L/min (2.22l/min/module) determined to be 21LMH.



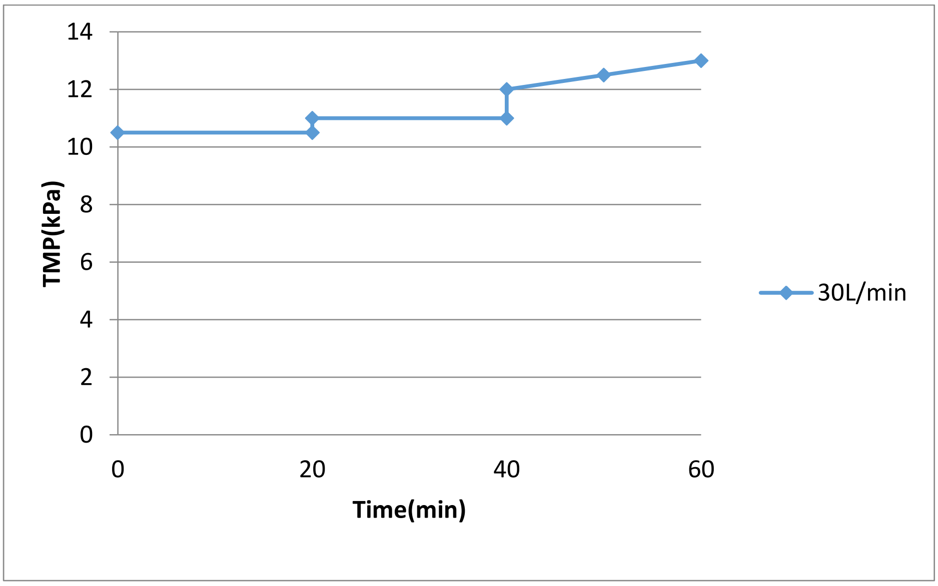
Critical flux at MLSS of 4g/l, aeration of 30L/min (1.67 L/min per module) determined to be 19.5 LMH



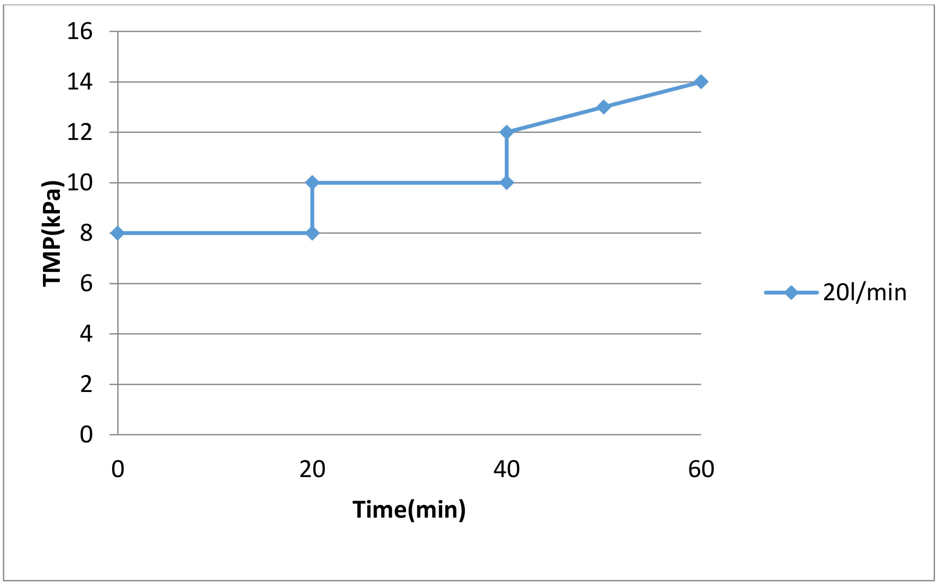
Critical flux at MLSSS of 4 g/l, aeration of 20l/min (1.11 l/min per module) determined to be 18.5 LMH



Critical flux at MLSS of 12g/l, aeration rate of 40L/min (2.22l/min/module) determined to be 10.5 LMH



Critical flux at MLSS of 12 g/l, aeration rate of 30l/min (1.67 l/min per module) determined to be 8.5 LMH



Critical flux at MLSS of 12 g/l, aeration rate of 20 l/min (1.11 l/min per module) determined to be 7 LMH