DEVELOPMENT AND EVALUATION OF WOVEN FABRIC IMMERSED MEMBRANE BIOREACTOR FOR TREATMENT OF DOMESTIC WASTE WATER FOR RE-USE

By

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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.

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ABSTRACT

Increased public concern over health and the environment, the need to expand existing wastewater treatment plants due to population increase, and increasingly stringent discharge requirements, have created a need for new innovative technologies that can generate high quality effluent at affordable cost for primary and secondary re-use. The membrane biological reactor (MBR) process is one of the innovative technologies that warrant consideration as a treatment alternative where high quality effluent and/or footprint limitations are a prime consideration.

MBR processes have been applied for the treatment of industrial effluent for over ten years (Harrhoff*,* 1990). In this process, ultrafiltration or microfiltration membranes separate the treated water from the mixed liquor, replacing the secondary settling tanks of the conventional activated sludge process. Historically, energy costs associated with pumping the treated water through the membranes have limited widespread application for the treatment of high volumes of municipal wastewater. However, recent advancements and developments in membrane technology have led to reduced process energy costs and induced wider application for municipal wastewater treatment (Stephenson *et al.,* 2000). This report describes a small and pilot scale demonstration study conducted to test a woven fabric microfiltration immersed membrane bioreactor (WFM-IMBR) process for use in domestic wastewater treatment. The study was conducted at Durban Metro Southern Wastewater Treatment Works, Veolia Plant, South Africa.

The main objective of this project was to develop and evaluate the performance of an aerobic woven fabric microfiltration immersed membrane bioreactor (WFM-IMBR) for small scale domestic wastewater treatment. The experiments were oriented towards three sub objectives: to develop the membrane pack for immersed membrane bioreactor based on WF microfilters; to evaluate the hydrodynamics of WF membrane pack for bioreactor applications; and to evaluate the long-term performance and stability of WFM-IMBR in domestic waste water treatment.

The literature was reviewed on membrane pack design for established commercial IMBR. The data collected from literature was then screened and used to design the WF membrane pack. Critical flux was used as the instrument to measure the WF membrane pack hydrodynamics. Long-term operation of the WFM-IMBR was in two folds: evaluating the performance and long term stability of WFM-IMBR.

The membrane pack of 20 flat sheet rectangular modules (0.56 m by 0.355 m) was developed with the gap of 5 mm between the modules. The effects of parameters such as mixed liquor suspended solids or aeration on critical flux were examined. It was observed that the critical flux decreased with the increase of sludge concentration and it could be enhanced by improving the aeration intensity as expected and in agreement with the literature. Hence the operating point for long term subcritical operation was selected to be at a critical flux of 30 LMH and 7.5 L/min/module of aeration.

Prior to the long term subcritical flux of WFM-IMBR, the operating point was chosen based on the hydrodynamic study of the WF membrane pack. The pilot scale WFM-IMBR demonstrated over a period of 30 days that it can operate for a prolonged period without a need for cleaning. Under subcritical operation, it was observed that there was no rise in TMP over the entire period of experimentation. Theoretically this was expected but it was never investigated before. Good permeate quality was achieved with 95% COD removal and 100% MLSS removal. The permeate turbidity was found to be less than 1 NTU and it decreased with an increase in time and eventually stabilized over a prolonged time.

Woven fibre membranes have demonstrated great potential in wastewater treatment resulting in excellent COD and MLSS removal; low permeate turbidity and long term stability operation. From the literature surveyed, this is the first study which investigated the use of WF membranes in IMBRs. The study found that the small scale WFM-IMBR unit can be employed in fifty equivalence person and generate effluent that is free of suspended solids, having high levels of solid rejection and has acceptable discharge COD for recycle.

Future work should be conducted on energy reduction strategies that can be implemented in WFM-IMBR for wastewater treatment since high energy requirements have been reported by commercial IMBRs.

PREFACE

This project was carried out at the Durban University of Technology (DUT), Department of Chemical Engineering. The pilot scale experiments were conducted at EThekwini Municipality Southern Wastewater Treatment Works (SWWTW), Veolia Plant. This project was supervised by Professor V. L Pillay, and Doctor S Rathilal. The project was completed over a period of 36 months from July 2009 to July 2012.

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DEDICATIONS

I dedicate this thesis to my parents, my girlfriend, my son and friends.

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NOMENCLATURE

Symbol	Definition	Units
J	Flux	LMH
J_C	Flux	LMH
A	membrane area	m^2
Q	Volumetric flow rate	m^3/hr
TMP	Trans membrane pressure	kPa
ΔP	Differential pressure	kPa
µ	Viscosity	Pa.s
R_C	cake resistance	m^{-1}
R_m	membrane resistance	m^{-1}
K	Membrane permeability	$l/m^2/hr/bar$
SRT	Sludge retention time	hr
V	Reactor volume	L
HRT	Hydraulic retention time	hr
VDS	Discharge sludge volume	l/d
k	mass transfer coefficient	
C_b	bulk salt concentration	mg/L
C_m	salt concentration on the membrane	mg/L
C_p	permeate salt concentration	mg/L
U_G	Aeration Velocity	m/s

ABBREVIATIONS

WF	Woven fibre
IMBR	Immersed membrane bioreactor
COD	Chemical Oxygen demand
NTU	Nephelometric Turbidity Unit
MLSS	Mixed Liquor Suspended Solids
DUT	Durban University of Technology
NRF	National Research Fund
UNEP	United Nation environment Program
WWT	Wastewater treatment
MBR	Membrane bioreactor
CASP	Conventional Activated Sludge Process
WFM	Woven fibre microfiltration
WFMF	Woven fibre microfiltration
TMP	transmembrane pressure
BOD	Biochemical oxygen demand
SS	Settleable solids
UV	Ultra violet
F/M	Food to micro-organism ratio
MWCO	Molecular weight cut-off
PDCO	pore size diameter cut off
RO	Reverse osmosis
NF	Nanofiltration
UF	Ultrafiltration
MF	Microfiltration
PVDF	polyvinylidene fluoride
EPS	Extra cellular polymer substances
CFV	Cross-Flow Velocity
HRT	Hydraulic Retention Time
SRT	Sludge retention time

NOM	Natural Organic Matter
SMP	Soluble microbial products
HF	Hollow Fiber
AAGR	Average Annual growth rate
CAS	Conventional activated sludge
DOC	Dissolve organic matter
TSS	Total suspended solids
TKN	Total kjedahl nitrogen
TN	Total Nitrogen
UKZN	University of Kwa Zulu Natal
AIT	Asian Institute of Technology
RRTS	Remote Rural Treatment Systems
RRWTS	Remote Rural Water Treatment Systems
WRC	Water research commission
PVC	Povinnylchloride
DO	Dissolved Oxygen
TSS	Total suspended solids
TOC	Total organic carbon
VSS	Volatile suspended solids
DO	Dissolve oxygen
TSS	Total suspended solids
TOC	Total organic matter
VSS	Volatile suspended solids
SDI	Sludge discharge index
MFI	Modified fouling index
EPDM	Ethylene propylenediene monomer

CHAPTER 1: INTRODUCTION

1.1Background and Problem Statement

Global Environment Outlook of the United Nations Environmental Program (UNEP) wing reports that about one in three of the world's population is currently living in countries suffering from moderate-to-high water stress, where water consumption is more than 10% of renewable freshwater resources. Many countries in Africa and Asia have very low or catastrophically low water availability (UNEP, 2002).

Water scarcity and water pollution are some of the crucial issues that must be addressed within local and global perspectives. One of the ways to reduce the impact of water scarcity as well as water pollution minimization is to expand water and wastewater reuse (UNEP, 2002). However, for wastewater reuse to be successful, it requires the involvement of all stakeholders, enactment of regulations, and availability of appropriate technologies.

Faced with these challenges, there is an urgent need to improve the effectiveness of water usage, and to augment the existing sources of water with more sustainable alternative technologies. Various approaches, modern and traditional, exist throughout the world for efficiency improvement and augmentation. Wastewater reuse is one of these approaches and has become increasingly important in water resource management for both environmental and social economic benefit (Asano, Burton and Tchobanoglous, 2006).

Wastewater reuse conserves freshwater supplies, is environmentally friendly and makes economic sense because the reclaimed water is available near urban areas where water supply reliability is most crucial and water is priced the highest.

Conventional wastewater treatment is currently widely used globally, however it has limitations since its scale is predominantly large and more capital investment is needed. It requires skill to operate which hinders its success in developing countries. Its implementation in remotely developed residential complexes is difficult since more capital investment is required; rural areas also experience a disadvantage from this technology since inhabitants are scattered. These challenges are common in developing countries. Hence, there is a need to develop better waste water treatment (WWT) technologies suitable for rural area's or remote residential complexes in developing countries (Pillay and Jacobs, 2008).

Amongst the technologies that have been developed for wastewater treatment is membrane bioreactors (MBRs). MBRs are increasingly being specified as a viable alternative for the reclamation of wastewater for reuse (Howell, Chua and Arnot, 2004). The MBR refers to a combination of membrane technology and high rate biological process technology for wastewater treatment (Stephenson et al., 2000).

MBRs produce excellent effluent quality using a small footprint. In developed economies, the driving forces behind the use of MBRs are;

1) The strict effluent discharge standards. In many cases, the MBR effluent quality is so good that it can be reused directly in non-potable applications;
2) Their small footprint.
3) The continuous decrease in membrane costs is increasing the competitiveness of MBR compared with conventional activated sludge processes (CASP).

The major limitations in the application of existing MBR systems in developing economies are: lack of robustness of commercial membranes (membranes are destroyed when they dry out) and commercial membranes are generally expensive. Hence there is a need to develop a robust membrane that can overcome the difficult conditions present in developing economies (Pillay and Jacobs, 2008).

Woven fabric microfiltration (WFM) is a new technology developed in South Africa by Durban University of Technology in collaboration with the University of Stellenbosch. The technology is currently applied to drinking water production with success. Hence, WFMF has a good potential to be applied in wastewater.

This project looks at the integrations of WFM with IMBR (WFM-IMBR) for application in wastewater treatment for reuse. Membrane pack development, hydrodynamics and long-term operation stability is the focus area of this study.

1.2 Objectives and Aims

The main aim of this project is to develop and evaluate the performance of woven fabric microfiltration immersed membrane bioreactor (WFM-IMBR) for small scale domestic wastewater treatment for reuse.

The specific objectives of the study are as follows:

1. To develop the membrane pack for immersed membrane bioreactor based on woven fibre (WF) microfilters.
2. To evaluate the hydrodynamics of the WF membrane pack for bioreactor applications.
3. To evaluate the short-term operation of WFM-IMBR in domestic wastewater treatment.
4. To evaluate the long-term performance of WFM-IMBR in domestic wastewater treatment.

1.3 Approach

The literature was reviewed on membrane pack design for commercially established IMBR systems. The data collected from literature was then screened and used to design the WF membrane module and pack. The membrane pack housing was designed based on the membrane pack size and optimum circulation inside the membrane pack housing. The height between the base and the diffuser, the height between the diffuser and membrane pack and the height above the membrane pack was optimised.

Critical flux was used as the instrument to measure the WF membrane pack hydrodynamics. Critical flux was evaluated by the step method. Critical fluxes were evaluated at different sludge concentrations and different aeration rates.

Short term operation of a semi pilot WFM-IMBR was evaluated at subcritical conditions prior to long term operation of pilot WFM-IMBR. This experimental step acted as a feasibility study for long term operation. Short-term stability and performance of WFM-IMBR was monitored. The following parameters viz mixed liquor suspended solids (MLSS), turbidity, chemical oxygen demand (COD) and trans membrane pressure (TMP) were monitored in order to evaluate the stability.

Long-term operation of WFM-IMBR was evaluated at subcritical conditions. The performance of WFM-IMBR was compared to commercial small scale IMBRs. The pilot scale WFM-IMBR design was based on a commercial small scale IMBR version. Long-term stability and performance of WFM-IMBR was evaluated. The same parameters monitored during short term subcritical operation were monitored in this experimental trial namely; MLSS, turbidity, COD and TMP.

1.4 Thesis Outline

Chapter 1 The introduction outlines the purpose of this study. It provides the background information on the membranes, membrane bioreactors and woven fabric microfiltration in membrane bioreactors.

Chapter 2 A literature review which deals with IMBRs and application of IMBRs in wastewater treatment for reuse is reported. Membrane fouling, fouling characteristics, Fouling reduction strategies and the concept of critical flux is presented.

Chapter 3 The membrane module, membrane pack development and the membrane pack housing is reported.

Chapter 4 Presents the hydrodynamics of the developed membrane pack. Information on critical flux evaluation by the step method and the effect of sludge concentration and aeration on critical flux is given.

Chapter 5 Gives general information about Durban Metro Southern Wastewater Treatment Works. It also presents the methodology followed in this study. General measuring protocol is described considering the materials and methods for each laboratory analysis applied. The description of the pilot scale WFM–IMBR is presented. Experiment design and operation conditions are described.

Chapter 6 Presents the summary of findings.

Chapter 7 Presents the conclusions.

CHAPTER 2: LITERATURE REVIEW

2.1 Water scenario in developing economies

Among the natural resources that are indispensable for human welfare and socio-economic development, water is ranked number one. However, scarcity and misuse of this life-supporting resource poses a serious and growing threat to food security, human health and the environment (Ndinisa, 2006). The United Nations International Drinking Water and Sanitation Decade (1981-1990) set its goal to provide safe drinking water and sanitation for the entire world population. However, a decade later more than a billion people in the world still do not have access to clean drinking water and more than 1.7 billion lack access to sanitation (Vigneswaran and Visvanathan, 1995). Problems arise because the fresh water resources are distributed unevenly over the earth's surface (Ndinisa *et al.,* 2005). There are four major global problems concerning fresh water: a) shortage of renewable supplies, b) unequal distribution of supplies, c) problems of water quality and health, and d) disastrous effects of unrestrained construction of dams (Gleick, 2004).

The World Bank stated that the way the human race is dealing with its water supplies needs to change drastically, and that people will have to acknowledge that fresh water is a scarce natural resource that needs to be treated with great care (Brans, 1997). Nowadays sustainable water use has become more important than ever. As the climate changes are accelerating and considering the water shortage in developing countries, in the near future the situation will get worse. Therefore efficient water reuse is undoubtedly one of the answers to this problem.

Consequently, 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 it is used for human consumption for example China (Ndinisa, 2006). South Africa is a water scarce country with an average rainfall of 500 mm per annum, 21% of the country receives less than 20 mm and as such all effluent has to be purified and returned to rivers.

Considering this situation, an efficient and effective technology for wastewater treatment and reuse is of increased interest in this context. One of these technologies is the Membrane Bioreactor (MBR) which allows the separation of treated wastewater from the active biomass, with a solid-liquid separation by membrane filtration.

2.2. Objectives of Wastewater Treatment

The principal objective of wastewater treatment is generally to allow human and industrial effluents to be disposed off without posing danger to human health or causing unacceptable damage to the natural environment.

2.2.1. Constituents of wastewater

The composition of wastewater varies widely. This is a partial list of what it may contain (Asano and Tchobanoglous., 1987):

- Water (> 95%) which is often added during flushing to carry waste down a drain.
- Pathogens such as bacteria, viruses, and parasitic worms.
- Non-pathogenics and bacteria.
- Organic materials such as hair, food, paper fibers, plant material and humus.
- Soluble organic material such as urea, fruit sugars, soluble proteins and drugs.
- Inorganic particles such as sand, grit, metal particles and ceramics.
- Soluble inorganic material such as ammonia, sea-salt and hydrogen sulfide.
- Animals such as protozoa, insects, arthropods and small fish.
- Gases such as hydrogen sulfide, carbon dioxide and methane.
- Emulsions such as paints, adhesives, mayonnaise and emulsified oils.
- Toxins such as pesticides, poisons and herbicides.
- Pharmaceuticals and other hormones.

2.2.2 Constituents to be removed from wastewater by activated sludge process

This following table shows the key wastewater parameters, and its standards, to be treated before it is discharged into receiving environment. It also shows the achievable figures in a secondary treatment before the disinfection stage.

Table 2.1: Waste discharge standards for key sanitary variables of wastewater treatment in South Africa adapted from new South African effluent discharge standards (adapted from Asano and Tchobanoglous, 1987).

Pollutant parameter	Units	Advanced secondary treatment	Proposed waste discharge standard
Chemical oxygen demand	mg/ℓ	50	65
Suspended solids	mg/ℓ	15	20
Ammonia nitrogen	mg N/ℓ	2.0	3.0
Phosphate phosphorus	mg P/ℓ	0.8	1.0

The above table proves that the proposed discharge standards can be achieved at the secondary treatment stage. However the tertiary treatment stage is still important for disinfection.

The following shows the regulated metals in wastewater before discharge into the municipal sewer and standards for long term and short term periods.

Table 2.2: South African effluent discharge standards for metals (adapted from Asano and Tchobanoglous, 1987).

Trace Metal	**Units**	**Technologically Achievable standard**	**Proposed Waste Discharge Standards**	
			Short term	**Long term**
Aluminium	ug/ℓ	45	50	30
Arsenic	ug/ℓ	53	60	30
Cadmium	ug/ℓ	<10	8	1
Chromium III	ug/ℓ	<5	110	110
Chromium VI	ug/ℓ	<5	20	20
Copper	ug/ℓ	<10	6	2
Cyanide	ug/ℓ	N/A	30	6
Iron	ug/ℓ	187	300	300
Lead	ug/ℓ	<50	10	9
Manganese	ug/ℓ	277	400	400
Mercury	ug/ℓ	<10	2	1
Nickel	ug/ℓ	13	N/S	N/S
Selenium	ug/ℓ	<100	50	8
Zinc	ug/ℓ	40	50	50

It is common practice in a developing economy like South Africa for government to grant relaxation of wastewater discharge standards for a short term period while public institution is providing support to achieve the long term discharge standards. This table shows the relaxed short term discharge standards, long term discharge standards and achievable standards.

2.3 Conventional wastewater treatment

Conventional wastewater treatment consists of a combination of physical, chemical, and biological processes and operations to remove solids, organic matter and nutrients from wastewater. The terms used to describe different degrees of treatment in order of increasing treatment level are: preliminary, primary, secondary, and tertiary or advanced wastewater treatment (Asano and Tchobanoglous., 1987). A generalized wastewater treatment block diagram is shown in Figure 2.1.

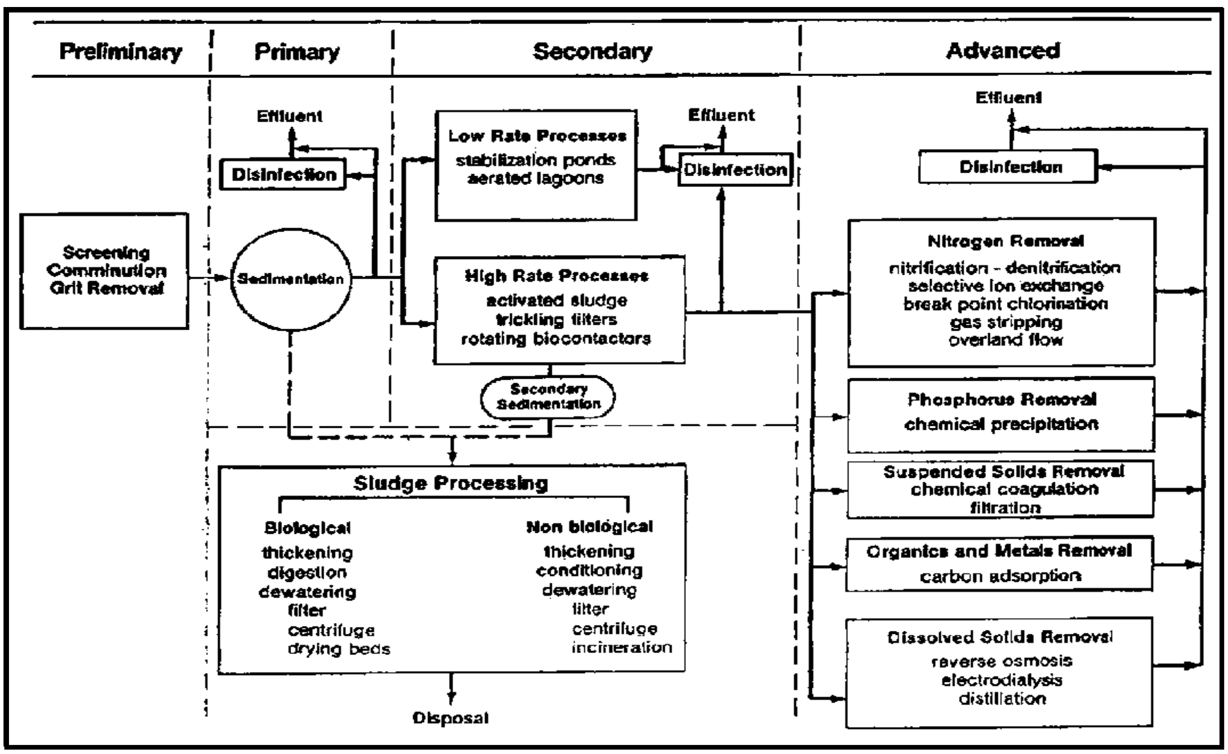


Figure 2.1: Generalized block diagram for municipal wastewater treatment

(Adapted from Asano, Smith and Tchobanoglous, 1985)

2.3.1 Stages of wastewater treatment

Preliminary treatment

The main goal of preliminary treatment stage is the removal of particulate materials and other large materials in raw influent. The removal of these materials improves the operations and maintenance of the treatment units. This stage comprise of screening and grit removal.

In grit chambers, the flow of raw feed is maintained moderately high in order to prevent the settling of most organic solids (Asano and Tchobanoglous, 1987).

Primary treatment

The main goal of primary treatment is the removal of settleable organic and inorganic solids by settling. The removal of floating material in this stage is achieved by skimming. This stage comprises of primary settling tank (PST) and no chemical reaction is taking place on this stage (Asano and Tchobanoglous, 1987).

Typical effluent quality from selected primary treatment plants are tabulated in table E1.1 page 147.

Secondary treatment

This stage is a one big chamber divided into aerobic zones which has the varying aeration intensity depending on the targeted removal compound of interest. Generally this step comprise of five aeration zones. In the first zone, the return sludge from clarifiers is mixed with raw feed from primary treatment stage. This zone maintains the food micro ratio by combining the return sludge which has less organic and high bacteria population with raw effluent from primary treatment stage which is rich in organics. There is no use of air in this zone only mixing of two streams is taking place (Asano and Tchobanoglous, 1987).

In the second zone, the ammonium nitrite is converted into nitrate due to aeration intensity. In the third zone aeration is increased two fold of the intensity of the second zone. Nitrates are converted into ammonia. In the fourth zone the aeration is manipulated and Ammonia is converted into nitrogen gas which escapes into atmosphere (Asano and Tchobanoglous, 1987).

In the fifth zone, the phosphorus removal takes place due the aeration intensity used.

Numerous aerobic biological processes exist for secondary treatment different primarily in the manner in which oxygen intensity to the micro-organism and in the rate at which organism oxidize the organic materials. The common applied high rate processes include activated sludge, trickling filters or biofilters, oxidation ditches and rotating biological contactors (RBC), (Asano and Tchobanoglous, 1987).

Tertiary treatment

Tertiary treatment stage is the further treatment after second treatment stage. It is commonly used when a specific wastewater constituent which cannot be removed by secondary treatment has to be removed in this stage. Individual treatment has the capacity to treat and remove the following: nitrogen, phosphorus, suspended solids, refectory organics, heavy metals and dissolved solids (Asano and Tchobanoglous, 1987).

Disinfection

Disinfection is the polishing stage after tertiary treatment stage. Chlorine gas dosage is commonly used. However, ozone and ultra violet (UV) irradiation can also be used for disinfection but these techniques are very rare in developing economies. The chlorine gas disinfections generally provide contact time of about 30 minutes.

The bactericidal effects of chlorine disinfection and other disinfectants are very dependent upon pH, contact time, organic content, and effluent temperature (Asano and Tchobanoglous, 1987).

2.3.2 Limitations of conventional wastewater treatment

Although the conventional activated sludge process has been successfully employed globally, it has the following limitations especially for application in developing countries;

i. High capital cost investment and high operation cost

This means that the implementation in remote developed complexes and remote rural areas where people are scattered is limited.

ii. Skilled labour is required

Competent labour is required to operate this system. Such skilled labour is not readily available in remote rural areas

iii. Constant supply of electricity is required

Energy is needed for operation of the activated sludge process which limits its implementation in places where there is no electricity especially rural areas of developing countries.

Hence, there is need to develop technologies that can be employed in rural areas and remote developed complexes for wastewater treatment in Southern Africa. Membrane bioreactors (MBR) represent an interesting alternative.

2.4 Historical perspective of IMBRs

The use of biological treatment can be traced back to the late nineteenth century. By the late 1930s, it was a standard method of wastewater treatment (Rittmann *et al*., 1987). During the course of anaerobic or aerobic digestion, soluble organic matter is biodegraded into various end-products such as H_2O, CO_2, CH_4 and biological cells. After removal of the soluble biodegradable matter in the biological process, any biomass formed must be separated from the liquid stream to produce the required effluent quality. A secondary settling tank is traditionally used for the separation and this clarification is often the limiting factor in effluent quality (Benefield and Randal, 1980).

The quality of the final effluent from conventional biological treatment systems is highly dependent on the hydrodynamic conditions in the sedimentation tank and the settling characteristics of the sludge. Consequently, large volume sedimentation tanks offering several hours of residence time are required to obtain adequate solid-liquid separation (Fane, Fell and Nor, 1978). At the same time, a close control of the biological treatment unit is necessary to avoid conditions that lead to poor settling ability or bulking of sludge. A correct food to microorganism ratio (F/M) should be maintained in order to determine proliferation of filamentous bacteria that results in sludge with poor settling properties. Very often, however, economic constraints limit control options such as pH and MLSS. Even with such controls, further treatment, such as filtration or carbon adsorption are needed for most wastewater reuse applications. Therefore, a solid-liquid separation method different from conventional methods is desirable.

Application of membrane separation techniques for bio-solid separation can overcome the disadvantages of the sedimentation tank and biological treatment steps. The membrane offers a complete barrier to suspended solids and yields high quality effluent. Although the concept of activated sludge process coupled with ultrafiltration was commercialised in the late 1960s by Dorr-Oliver (Smith, Gregorio and Talcott,1969), the application has only recently started to attract serious attention and there has been considerable development and applications of membrane processes in combination with biological treatment in the 1990s (Churchouse and Wildgoose, 1999). Full-scale commercial aerobic MBR processes first appeared in North America in the late 1970s and then in Japan in the early 1980. The

MBR technology did not enter Europe on a commercial scale until the mid – 1990s. According to Stephenson *et al.* (2000) there are over 500 commercial units in operation mid 1990s worldwide, 66% of which are in Japan. Around 98% of these MBR combine membranes with an aerobic process rather than anaerobic process. The MBR process in general, offers several advantages over the conventional processes currently available. MBRs produce excellent effluent quality using a small footprint. The MBR effluent quality is so good that it can be reused directly in non-potable applications.

Membrane bioreactor refers to the combination of a membrane process like microfiltration or ultrafiltration with a suspended growth bioreactor, and is now widely used for municipal and industrial wastewater treatment (Judd,2006).

2.4.1 Introduction to membranes

Definition

A membrane can be defined as a thin film that is capable of separating materials as a function of their physical and chemical properties when a driving force is applied across it. The advantages of membrane technology include continuous separation, low energy consumption, easy combination with other existing techniques and easy up-scaling.

Types of membrane filtration

Pressure driven membrane filtration processes are classified as follows based on the particle size of the contaminants to be separated:

Microfiltration is the coarsest size of the pressure driven membrane filtration class. Its application is to separate suspended particles from dissolved substances. A microfiltration membrane is classified by pore diameter cut-off (PDCO) and retains particles with diameters in the range of 0.1 to 10 µm (Cheryan, 1998).

Ultrafiltration is used for separation of large macromolecules such as proteins and starches and all types of micro-organisms, such as bacteria and virus (Aptel and Buckley, 1996). Ultrafiltration membranes are classified by molecular weight cut-off (MWCO) which is defined as the molecular weight of the smallest molecules that may be separated. Ultrafiltration covers particles and molecules that range from 1,000 to 500,000 daltons in molecular weight (Cheryan, 1998).

Nanofiltration membranes retain solute molecules ranging from 100 to 1000 Daltons in molecular weight. Nanofiltration membranes are classified by molecular weight cut-off like ultrafiltration membranes or by percentage sodium chloride rejection like reverse osmosis membranes. It can also reject contaminants as small as 0.001 µm (Taylor and Jacobs, 1996).

Reverse osmosis involves the tightest membranes which are capable of separating even the smallest solute molecules or particles with a diameter as small as 0.0001µm (Taylor and Jacobs, 1996). Reverse osmosis membranes are classified by percentage rejection of sodium chloride in an aqueous solution under specified conditions and range from 99 to 99.5%.

Membranes can be manufactured by a wide variety of materials which include inorganic membrane (sintered metals and ceramics) and organic membranes (polymers). The inorganic membranes have better chemical, mechanical and thermal stabilities, but have disadvantages of being very fragile and more expensive than the organic membranes. The organic membranes are widely used in water and wastewater applications because they are more flexible and can be made from cellulose. All synthetic polymers have relatively good chemical, mechanical and thermal stability tendencies, and also provide the membranes with better antifouling properties through the use of hydrophilic polymers (Cheryan, 1998; Aptel and Buckley, 1996).

2.4.2 Overview of membrane filtration processes

The basic principle of all membrane operations is the separation of a mixture of substances with a selective thin film. The transport of matter through a selective barrier is caused by a chemical potential difference between two phases, i.e., the feed and the permeate (Mulder, 1996). In pressure-driven membrane filtration systems, which have been widely applied in water and wastewater treatment systems, the driving force is a pressure difference across the membrane. Typically, four types of membranes are distinguished according to their separation range (molecular weight cut-off or pore size) and the applied Trans membrane pressure: reverse osmosis (RO), nanofiltration (NF), ultrafiltration (UF), and microfiltration (MF) (Figure 2.2).

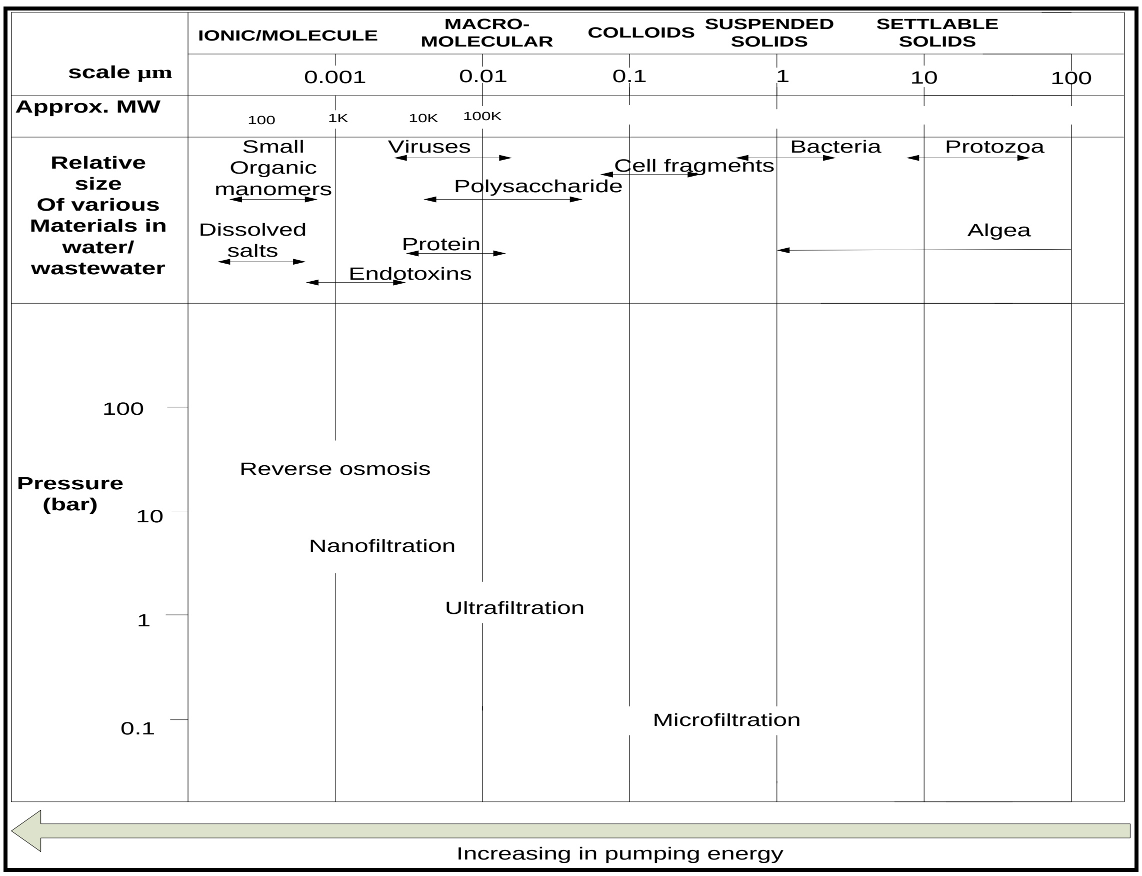


Figure 2.2: Classification of membranes and colloidal/macromolecular organic matter in ground and surface water (Adapted from Mallevialle, Odendaal, and Wiesner, 1996)

In MBRs, a tight microfiltration or a lose ultrafiltration membrane is often applied. The most often used membrane materials in MBRs are organic polymers, e.g., polyethylene, polypropylene and polyvinylidene fluoride (PVDF) membranes (Judd, 2006). Some of them are blended with other materials to change their surface charge or hydrophobicity (Mulder, 1996). Inorganic membranes (e.g., ceramic membranes) are only used in special applications for example where solvent resistance and thermal stability are required (Barker, 2004). The ultra-filtration membrane often has a supporting layer (e.g., a microfiltration membrane), onto which a thin skin layer, i.e., a true ultrafiltration membrane, is attached.

2.4.3 Membrane fouling

Membrane fouling refers to the deposition or adsorption of material on the surface of the membrane or within the pores. It is a common and costly problem in membrane filtration applications. Fouling may cause a decline in permeate flux, increase in TMP, loss of permeate quality and deterioration of the membrane, etc. In conjunction with the forming of a filter cake, a shift in the effective pore size or molecular weight cut-off (MWCO) to a smaller size is common, which can result in a MF process displaying the characteristics of a UF membrane (Lee, Ahn and Lee, 2001).

Although MBRs have numerous advantages, membrane fouling is the major constraint that is encountered (Le-Clech, Jefferson and Judd, 2003a). During filtration operation, performance losses are expected because of the reduction of membrane permeability. This leads to additional operating costs, which contribute to the loss of competitiveness of MBRs. This problem is generally observed by performing a constant transmembrane pressure (TMP) filtration during which a flux decline is expected to occur. As shown in Figure 2.3, the TMP increases when the solution is filtered at a constant flux (Le-Clech, Jefferson and Judd, 2003a).

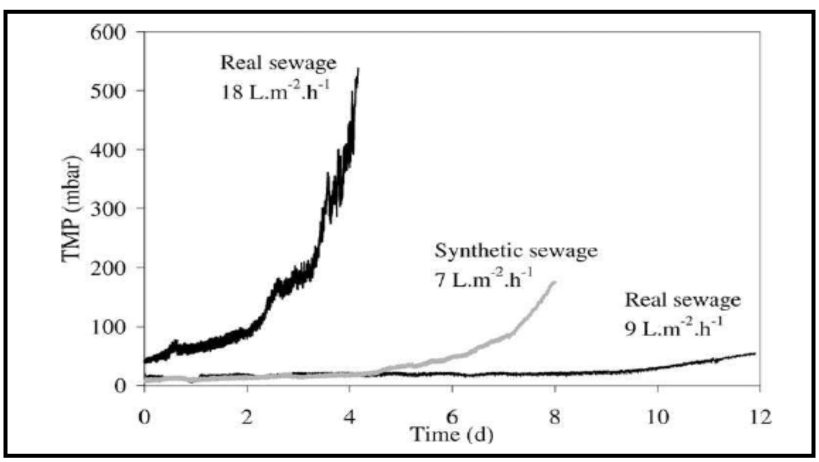


Figure 2.3:Long-term filtration for constant flux operation (Le-Clech, Jefferson and Judd, 2003a)

It is complicated to manage fouling because of various factors as shown in Figure 2.4.

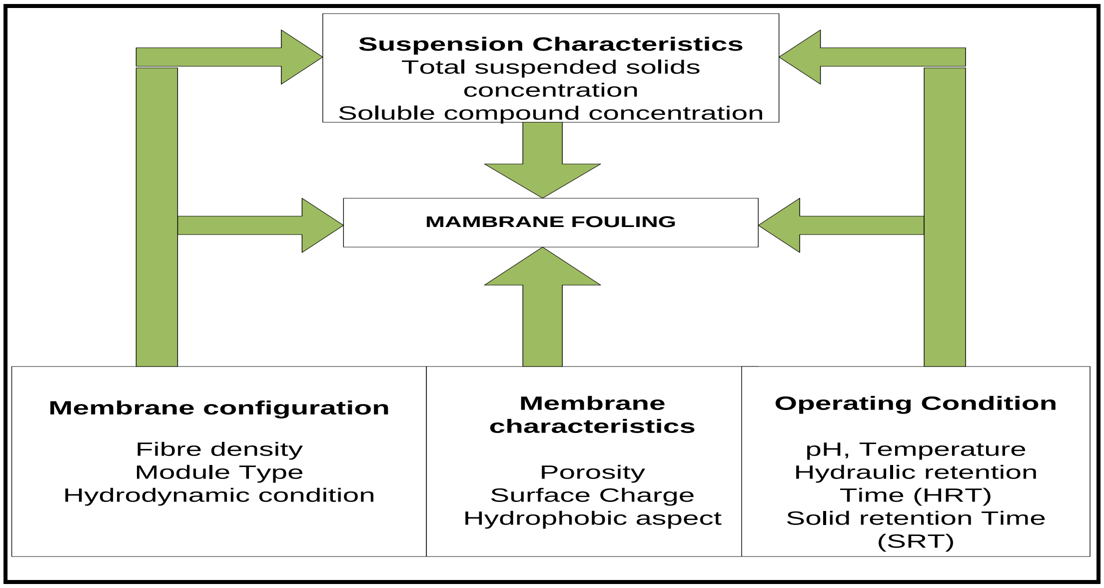


Figure 2.4: Key parameters in membrane fouling (Khongnakorn *et al.,* 2007)

2.4.4 Suspension characteristics

In the bioreactor, the mixed liquor suspension contributes to the membrane permeability reduction because of three types of compounds (particular, colloidal and soluble). Table 2.3 shows the contradiction according to some authors in the relative importance of each of them. However, it is difficult to compare these results because of the operating conditions that are applied (hydrodynamic conditions, solid retention time) are different in each case.

Table 2.3: Relative role of different sludge fractions in membrane fouling (Bouhabila, Ben Aïm and Buisson, 2001; Defrance *et al*., 2000; Wisniewski, 1998).

Fraction	Bouhabila (%)	Defrance (%)	Wisniewski (%)
Solutes	26	5	52
Colloids	50	30	25
Suspended solids	24	65	23

Basically, there exists two types of fouling (Judd, 2006):

- **Reversible fouling,** which can be easily eliminated by several appropriate mechanical operations (relaxation, backwashing).
- **Irreversible fouling,** which is only eliminated by chemical cleaning.

Because this clarification is simplistic, Hermia (1982) proposed to describe the fouling mechanism according to the size of the particles and the membrane pores:

- **A cake layer** is constituted when the particle size is greater than the pore size.
- **Pore narrowing** results from the entrance of the particles into the interior of the membrane when the particle size is slightly smaller than the pore size but does not pass through.
- **Pore blocking** occurs when the particle size is relatively the same as the pore size leading to the diminution of the filtration surface.
- **Adsorption** is irreversible fouling because of the physical-chemical interactions between the flocs and the membrane. This corresponds to the formation of molecular layers at the surface or in the membrane pores, which leads to the reduction of the pore size.

These types are illustrated below in figure 2.5.

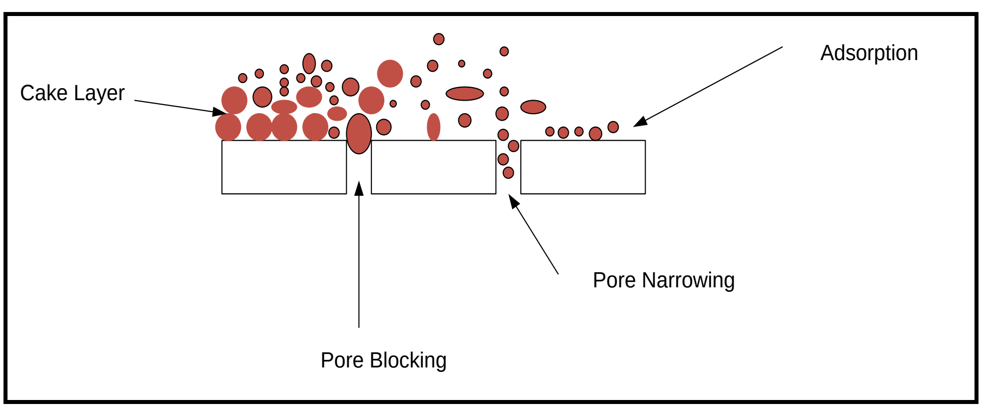


Figure 2.5: The different fouling mechanisms (Bouhabila, Ben Aïm and Buisson, 2001; Defrance *et al*., 2000; Wisniewski, 1998).

2.4.5 Fouling Control

Membrane fouling can be eliminated or limited by two major methods called preventive and curative, by optimizing the operating conditions (Le-Clech *et al.,* 2003b).

The preventive methods

The fouling rate is directly linked to the amount of particles brought to the membrane surface, and therefore to the flux decline. However, a flux which is too low would result in higher operational costs. Thus; an optimal operating flux must be found.

Field *et al.* (1995) introduced the concept of critical flux, which is defined as being the flux below which no membrane fouling is observed.

Figure 2.6 shows the schematic representation of flux, TMP and the critical flux onset point.

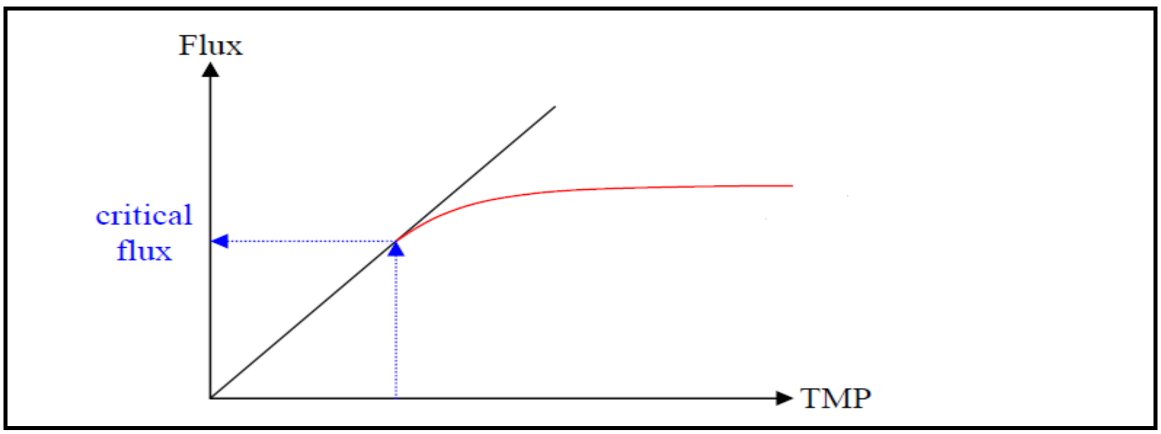


Figure 2.6: Definition of the critical flux from the curve Flux =f(TMP)(Espinosa , 2005)

The determination of the critical flux depends on several factors such as the membrane state, the characteristics of the mixed liquor and the hydrodynamic conditions of the system (Espinosa, 2005).

Le-Clech *et al.* (2003b) even demonstrated that this concept cannot be used for MBR processes and more generally for biological sludge filtration. Whatever the imposed flux, they observed an increase in the transmembrane pressure. This means that membrane fouling occurs even at the lowest flux tested namely 2 $L.h^{-1}.m^{-2}$. However, this augmentation remains low up to 10 $L.h^{-1}.m^{-2}$ in short filtration time conditions (15 min). Thus, Le-Clech *et al., (*2003b) defined these fluxes as being ‘sustainable’ in the studied conditions.

This is in agreement with the findings of (Ognier, Wisniewski and Grasmick, 2004) who tested the concept in much longer filtration time conditions. In spite of an initial filtration flux lower than the critical flux, a weak increase in the TMP is observed at the beginning of the experiment and an exponential increase after a few days of filtration (Figure 2.7). This would lead to a local flux greater than the critical flux. Then, a cake layer is being formed and results in a sudden augmentation of the TMP in the second period.

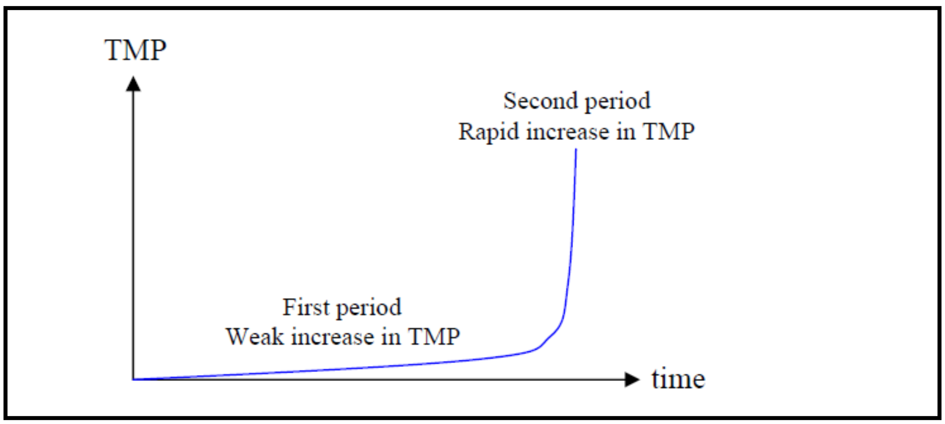


Figure 2.7: Trans membrane pressure change during long-term constant sub-critical flux in the membrane bioreactor (Ognier, Wisniewski and Grasmick, 2004).

Aeration

Good aeration implies turbulent conditions close to the membrane and thus reduces the formation of the cake. Many studies are ongoing to determine the optimum aeration conditions, in terms of air flow and flow patterns. However, Cui, Chang and Fane, (2003) estimated that the aeration could represent 50% of the energy consumption. It is therefore important to define an optimal air flow rate that would limit membrane fouling at a reasonable energy cost.

The curative methods

Backwashing

Backwashing involves pumping of the permeate in the reverse direction through the membrane in order to remove the reversible fouling by unclogging pores and transporting particles back to the reactor.

As backwashing uses treated water, it is important to optimize the parameters of this process such as the frequency, the duration and the ratio of these two parameters. Furthermore, Visvanathan ., (1997) argued that more frequent washing (200s filtration/15s washing) is less efficient than less frequent but longer washing periods (600s filtration/45s washing). However, it would be noted that despite many studies, the design of this process is often empirical (Van Kaam, 2005).

Chemical cleaning

It is generally recognized that relaxation and backwashing efficiency tend to decrease with operation time, as shown in the Figure 2.8 (Ognier, Wisniewski and Grasmick, 2004).

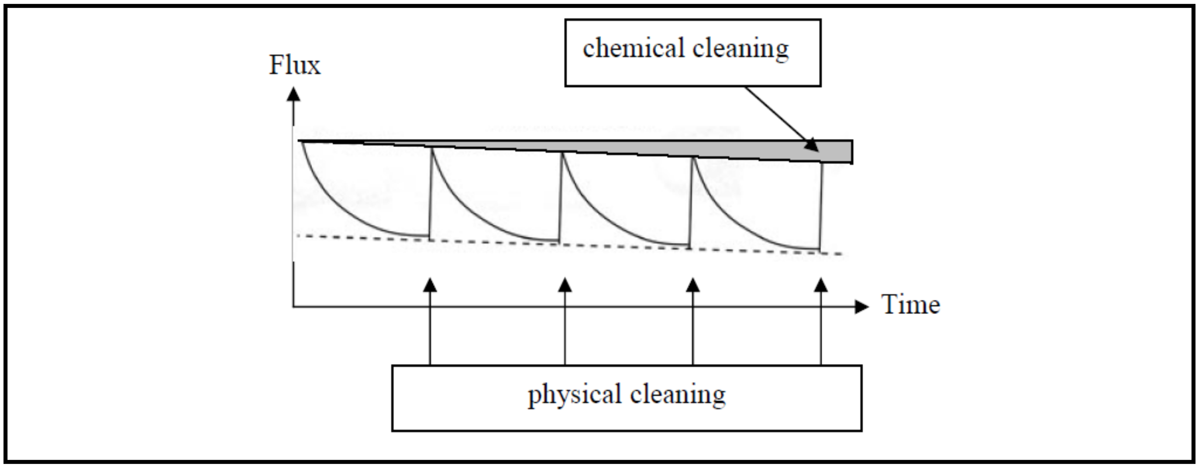


Figure 2.8: Membrane regeneration (Ognier, Wisniewski and Grasmick, 2004)

The chemical cleaning is the only way to eliminate the irreversible fouling. It is done by putting the membranes into a chemical solution that breaks the cohesion strengths between the particles and the membrane. Similar to existing physical cleaning strategies, the frequency of chemical cleaning may vary according to the degree of fouling (Ognier, Wisniewski and Grasmick, 2004).

2.4.6 Membrane fouling and foulant definition

Figure 2.9 summarises the factors affecting membrane fouling in the MBR process. It illustrates that process configuration; membrane features, biomass characteristics and operating conditions are more or less independent although they are interlinked.

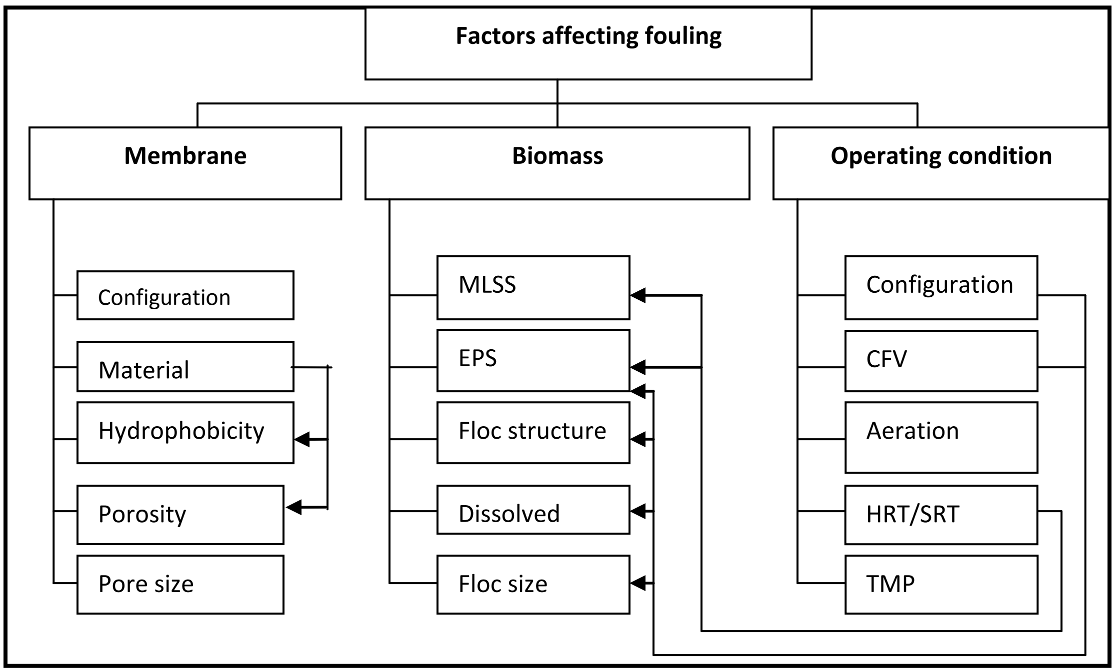


Figure 2. 9: Factors influencing membrane fouling in MBR processes (Adapted from Chang *et al.,* 2002).

2.4.7 Interactions between foulant and membrane

The affinity of foulants to the membrane can significantly influence the membrane fouling and permeate quality. The interaction between the foulant and membrane is more pronounced for the colloidal and macromolecular organic matter rather than the particulates due to the fact that they have smaller sizes. There are many factors which can influence this interaction, e.g., charge, pH, hydrophobicity, multivalent ions (Ca^{2+} and Mg^{2+}) and ionic strength morphology (Lee *et al.,* 2003).

Charge

If the colloids/macro-organics and the membrane surface have the same charge, the colloids/macro-organics will be repelled by the membrane due to electrostatic forces. Consequently, the adsorption of these organics is low (Nyström, Kaipia and Luque, 1995; Hong and Elimelech, 1997; Schafer *et al.,* 2004). Many colloids and macro organics are negatively charged at neutral pH conditions (Lee *et al.,* 2003), therefore, the MF/UF membranes in water and wastewater filtration processes are often manufactured or modified to be negatively charged. However, it should be noted that the charge of the membrane can be modified by adsorption and deposition of colloids/macro organics and eventually, the membrane may have the similar charge as the deposited colloids/macro organics (Hong and Elimelech, 1997).

Hydrophobicity

If the colloids/macro organics and the membrane surface have opposite hydrophobicity, the colloids/macro-organics may be repelled by the membrane (Hong and Elimelech, 1997). Many membranes for drinking water treatment are made hydrophilic (Mulder, 1996), which has the advantages of high membrane permeability and low affinity with the aromatic foulant e.g., many natural organic matters (NOMs). Fang and Shi (2005) conducted the filtration of MBR sludge and reported that the Extra cellular polymer substances (EPS) membrane suffered more pore blocking than the polyvinldene fluoride (PVDF) membranes because the former is more hydrophobic. However, it should be noted that the hydrophobicity of the membrane can be modified by the adsorption and deposition of colloids/macro organics and eventually, the membrane tends to have similar hydrophobicity to the deposited colloids/macro organics (Hong and Elimelech, 1997).

Membrane morphology

Membrane morphology, e.g. pore opening, pore size distribution and surface roughness can affect membrane fouling. Generally, a narrower membrane pore size distribution can reduce the amount of fouling (Mulder, 1996). Fang and Shi, (2005) investigated the filtration of MBR sludge using a few different MF membranes with similar nominal membrane pore size of 0.2 to 0.22 µm. The extra cellular polymer substances (EPS) membrane with large pore openings (18 to 20µm) suffered significantly more pore blocking than other membranes. The latter had a smooth surface and a more uniform pore size distribution, which suffer less pore blocking (Fang and Shi, 2005).

Concentration polarization

Concentration polarisation is the decrease in flux and solution concentration after passing through a selective barrier. Even if the applied force is increased, there is a minimal increase in permeate flux. This flux is known as limiting flux. According to Bowen and Jenner, (1995) this flux tendency can also be explained by gel polarisation phenomenon.

“According to this gel concentration polarisation phenomenon, the concentration at the membrane filter surface increases as the macro-solutes of the solution reaches its solubility limit and precipitates on the membrane surface to form solid gels. While for colloids the gel layer resembles a layer of closely packed spheres” Bowen and Jenner, (1995).

Figure 2.10 shows the schematic representation of concentration polarization and a boundary layer of the flow.

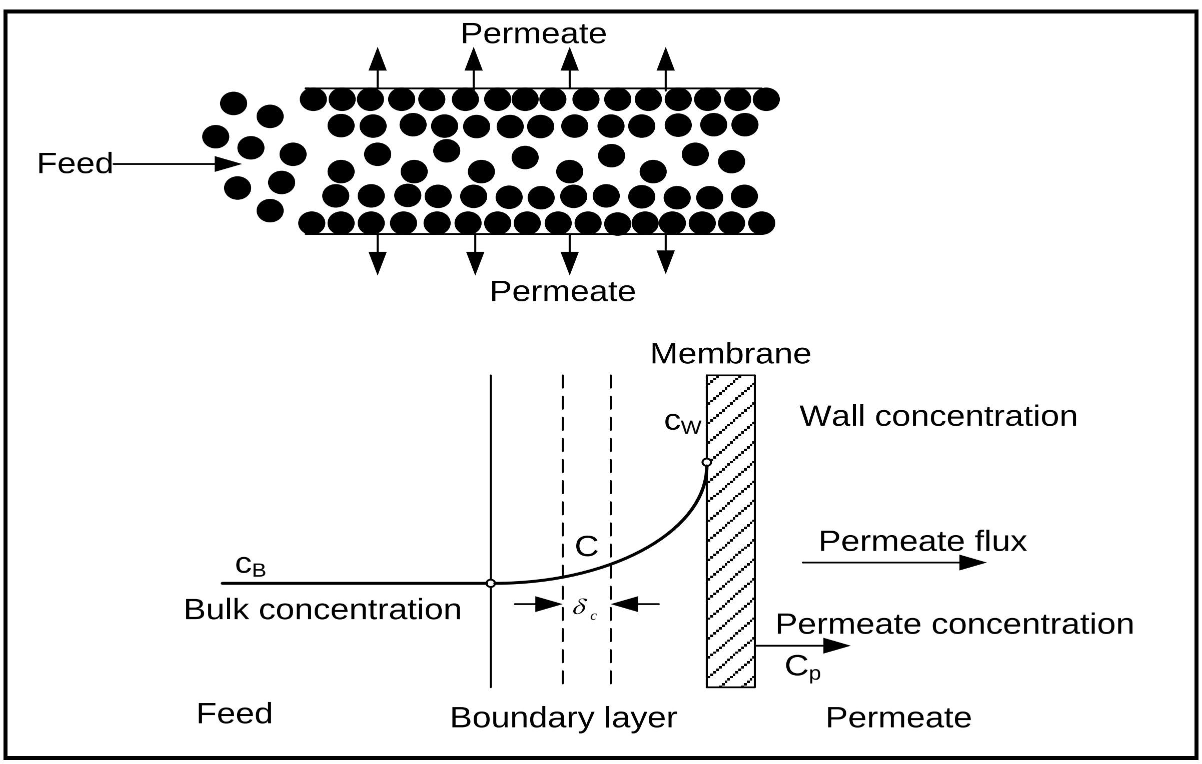


Figure 2.10: The schematic view of concentration polarization profile and the boundary layer (Adapted from Belfort, Davis and Zydney, 1994).

The relationship between flux and concentration is represented below in equation 2.1

$$CP = \frac{Cm - Cp}{Cb - Cp} \equiv \exp\left(\frac{Jv}{k}\right) \qquad \text{(2.1)}$$

Where CP is the concentration polarisation and dimensionless, Jv is the local flux in litres per square meters and hour (LMH), k is the local mass transfer coefficient in $m^3/m^2.kg$, C_m is the salt concentration on the membrane in kg/m^3, C_b is the bulk salt concentration in kg/m^3 and C_p is permeate salt concentration in kg/m^3.

Fouling of pressure driven membrane filtration systems

The common compounds that foul a membrane can be shown in the following four categories: particulate fouling caused by colloids and suspended solids, organic fouling caused by adsorption of organic matter, biofouling caused by deposition or growth of microorganisms, and scaling caused by salt precipitation as shown in table 2.4.

Table 2.4: Characteristics of four types of membrane fouling (Adapted from Belfort, Davis and Zydney, 1994)

	Particulate fouling	**Organic fouling**	**Biofouling**	**Scaling**
Foulants	Colloids Suspended solids	Organic matter	microorganism	Salt Metal cation
Major factors affect fouling	Concentration particle size distribution Compressibility of particles	Concentration Charge Hydrophobicity pH Ionic strength Calcium.	Temperature Nutrients	Temperature Concentration pH
Indicator of fouling prediction	Silt density index (SDI) Modified fouling index (MFI) Specific resistance to fouling(SRF)	DOC UV_{254}	Assimible organic compound (AOC) Biofilm formation rate (BFR)	Solubility
Feed water pre-treatment	Coagulation MF and UF	Adjustment of pH coagulation	Sand filtration Biofilter Coagulation Flocculation UF and MF	Acid Anti-scalent

Particulate fouling

Small particles can accumulate on the membrane surface, thereby forming a filter cake, which is referred to as particulate fouling. The particulates can either be suspended solids, colloids and even microorganisms. Particulate fouling is the dominant type of fouling in most MF and UF systems. However, MBRs using MF and UF membranes suffer more colloidal and organic fouling, which is addressed intensively in this thesis (Zeman and Zydney, 1996).

Organic fouling

Organic fouling refers to the adsorption of dissolved organic substances on the membrane surface or in its pores due to the intermolecular interactions between the membrane and organic matter (Zeman and Zydney, 1996). Natural organic matter (NOM) fouling in drinking water filtration processes is a well-known problem (Combe *et al.*, 1999; Jones *et al.,* 2000; Lee *et al.,* 2004). Humic substances are a major fraction of NOM. However, the filtration of wastewater and activated sludge has been applied more recently and soluble microbial products (SMP) fouling has been the main concern.

Biofouling

Biofouling refers to the adhesion and growth of microorganisms on the membrane surface, i.e., the formation of a biofilm, which results in a loss of membrane performance. Basically a biofilm can occur on all kinds of surfaces, natural and synthetic, due to the fact that bacteria have developed elaborate adhesion mechanisms. RO and NF processes suffer more from biofouling due to their low flux and limited membrane cleaning options (Flemming, Griebe and Schaule, 1996; Flemming, 1997; Baker and Dudley, 1998).

Scaling

The formation of scale on the membrane surface may occur if dissolved salts exceed their solubility product. Typically, over-saturation is of concern in reverse osmosis and nanofiltration operations with regard to $CaCO_3$, $CaSO_4$, $BaSO_4$, $SrSO_4$, $MgCO_3$, and SiO_2 (Barker, 2004). However, RO plants can operate at super-saturation conditions (e.g., $BaSO_4$) without scaling (Bonne *et al.,* 2000). Scaling is not dominant in MBR fouling. However, iron or calcium precipitation may occur in some cases. Acid cleaning should be considered if oxidant cleaning is not sufficient to restore the membrane permeability (tePoele and van der Graaf, 2005).

However, one has to keep in mind that there are overlaps in the above four types of fouling, e.g., organic fouling due to the deposition of suspended solids can be particulate fouling, so is the biofouling due to the seeding and growing of a biofilm. In addition the different types of fouling can occur simultaneously and form hybrid fouling, which can be more difficult to clean (tePoele and van der Graaf, 2005).

The following table shows the major suppliers of membranes around the world.

Table 2.5: Major Membrane Product Suppliers (Modified from Judd, 2006)

Flat Sheet	Hollow Fiber	Multiple Tubular
Kubota	Zenon	Berghof
Huber	Simens Water Technologies-Memco	Norit X-Flow
OrelisPleiade	Kolon	Oreliskerasep
Colloide	Mitsubishi Rayon	Milleniumpore
Brightwater	Motimo	
ITRI non-woven	Polymem	
Microdyn-Nadir	KOCH Puron	
Han-S	Asahi kasei	
	Ultraflo	

2.4.8 Early history of membrane bioreactors (MBRs)

Research on membrane bioreactors started 30 years ago. The technology first entered the Japanese market through a licence agreement between Dorr-Oliver and Sanki Engineering CO. Ltd, and currently they are commonly used in Japan (Stephenson et al., 2000). Most MBR installations are less than 19 years old, therefore the design criteria for this technology is still evolving (Wallis-Lage, 2003). Currently, over 500 membrane bioreactors are being used in processes for treating and reuse of domestic wastewater and industrial wastewater mostly from food and beverage industries (Stephenson *et al*., 2000).

Biomass separation MBRs, which can be both aerobic and anaerobic, are the most common and they have been widely applied in full-scale. Full scale aerobic MBRs were first installed in North America in the late 1970s and early 1980s in Japan. The aerobic MBR did not appear in Europe until the mid of 1990s (Stephenson *et al.,* 2000).

Most of the early membrane bioreactor projects for municipal wastewater were applications for small flows, ski resorts, trailer parks or office complexes, where it was important to handle load variation and where operation was easy. The equipment used for these MBRs were originally external cross-flow ultra-filtration membrane systems, with very long sludge retention time (SRT) of 50 days or more, and high mixed liquid suspended solids (MLSS) in the order of 15,000 mg/L up to 25,000 mg/L (Crawford *et al.,* 2000) compared to a MLSS of 1000-4000 mg/L that are used for activated sludge processes (Metcalf and Eddy, 2003). These applications were biologically robust, due to long SRT and complete Nitrification (Crawford *et al.,* 2000).

Next, MBRs were developed to remove total nitrogen. This was achieved by recirculating the nitrified mixed liquor into an anoxic zone where the nitrate could be reduced to nitrogen gas (Crawford et al., 2000). Phosphorus removal has been achieved by biological means (Le-clech, Chen and Fane, 2006).

2.4.9 Process Configuration in MBRs

Process description

There exist two common types of bioreactors Gander, Jefferson and Judd, (2000): the side stream configuration with an external recirculation loop (Figure 2.11) and the submerged MBR where the membrane is directly immersed into the reactor (Figure 2.13).

Recirculated MBRs

In this configuration, the mixed liquor is pumped from the reactor into the filtration module situated outside the bioreactor. In the module, the suspension is filtered from inside the membrane to outside for obtaining permeate. The retentate is then sent back into the reactor.

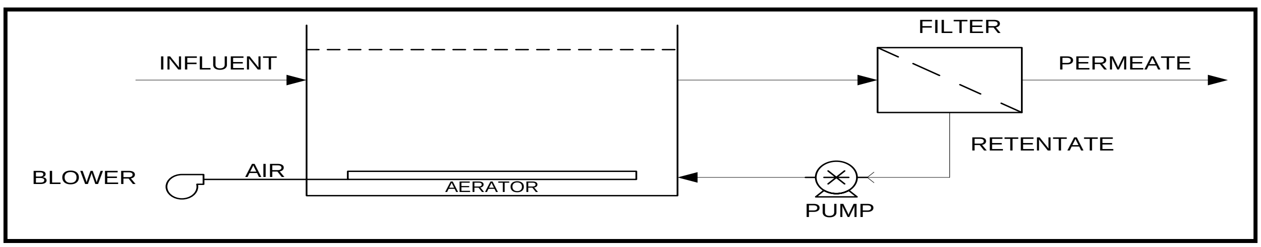


Figure 2.11: MBR with external membranes

The filtration carried out is a cross-flow type (Figure 2.12). This type of filtration generally requires high recirculation velocities into the membrane to make up for the loss of productivity induced by rapid fouling. The high shear-stress allows to reduce particle accumulation and accordingly membrane washing.

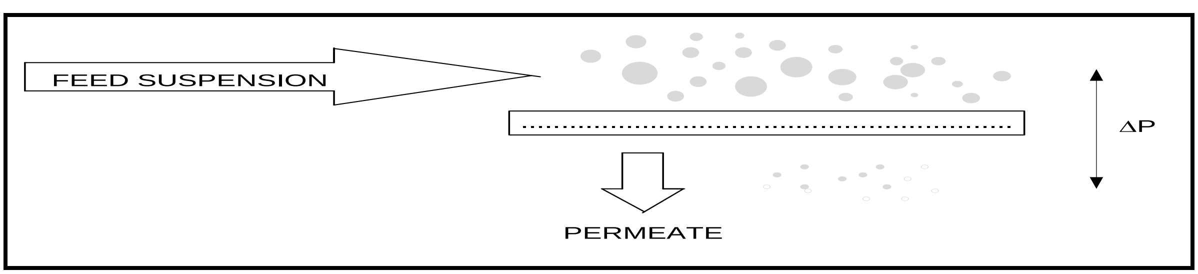


Figure 2. 12: Cross-flow filtration

However, this configuration involves high energy costs because of the need to pump at high velocity. That is why the membrane has directly been immersed into the reactor to save energy costs.

Submerged MBRs

The principle of operation is to immerse the membrane in the biomass and to ensure a filtration from outside the membrane to inside Gander, Jefferson and Judd, (2000).

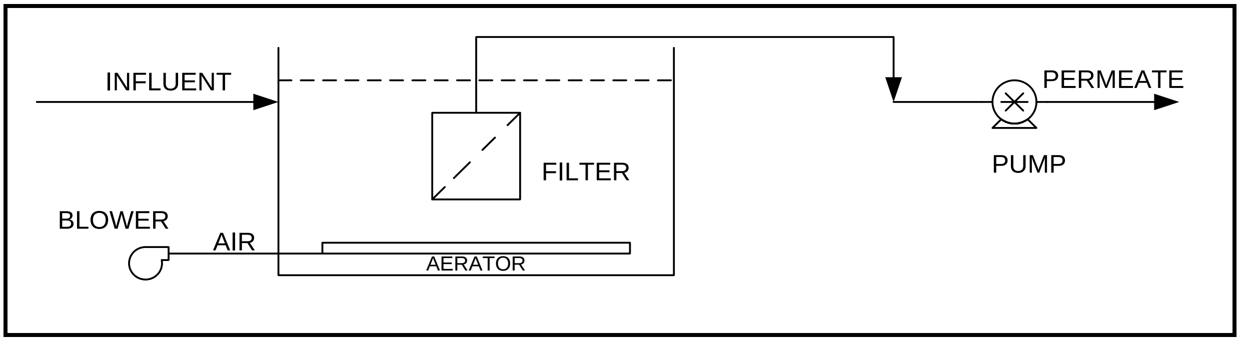


Figure 2.13: MBR with submerged membrane

This time, the filtration is called dead-end filtration because the velocity of the liquid is low at the membrane surface (Figure 2.14).

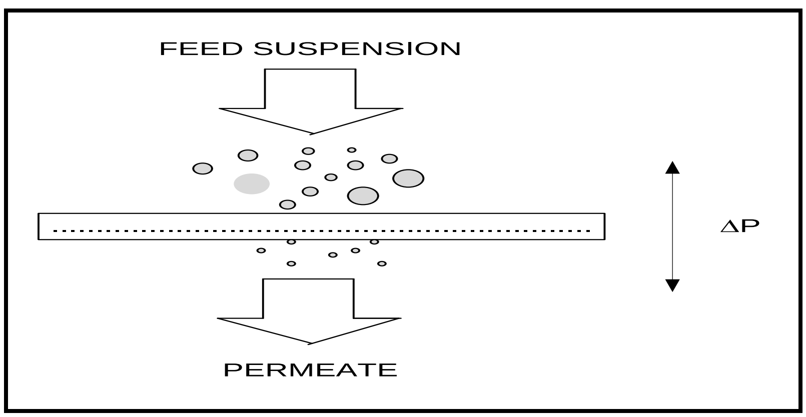


Figure 2.14:Dead-end filtration

Aeration in the case of submerged MBRs is used for biomass growth as well as for controlling membrane fouling. In terms of energy requirements, Gander, Jefferson and Judd, (2000) estimated that the energy needed for the external configuration ranges from 10 to 50 kWh.m unit whereas submerged MBRs need 0.2 to 0.4 kWh.m unit.

Membrane configurations

Three configurations of the membranes dominate in the MBR process: plate and frame/flat-sheet (for example Kota), tubular (used in most side stream configurations), and hollow fibre (e.g. Zenon and Mitsubishi Rayon, 2004).

In particular, hollow fibre (HF) membrane modules provide the highest packing densities. They are very simple with “tows” of fibres potted at one or both ends in a module. The module operates out to in, with the cake layer building up on the outer membrane surface.

The following table shows the specifications taken into consideration for manufacture of Kubota membranes.

Table 2.6: Design Consideration for flat sheet membrane manufacture of Kubota (Modified Wallis-Lage, 2003)

Membrane	
Type	Plate and frame
Configuration	Vertical
Pore size	0.4 µm
Module size	0.8 m^2
Location	Throughout basin
Screening Size	≤3mm
Flux rate	
Average (LMH)	17 to 25
Maintenance	
Clean type	Backpulse
Cleaning of membranes	
Type	Chlorine backwash
Frequency	In situ , ≤6months
Biological Parameters	
SRT, days	15
MLSS, mg/L	≤10 000

Table 2.7: Advantages and disadvantages of the MBR configurations (Modified Stephenson *et al.,* 2000)

Table 2.7 shows the advantages and disadvantages for Submerged and Side stream MBR.

Submerged MBR	**Side-stream MBR**
Advantages • Small footprint • Feed-forward control of O_2 demand • Less frequent cleaning required • Lower operating costs • Low liquid pumping costs (28% of total costs) (Gander *et al.,* 2000) • Low energy consumption (Cote *et al.,* 1998)	*Advantages* • Small footprint • Complete solids removal from effluent • Effluent disinfection • High loading rate capability • Combined COD, solids and nutrient removal in a single unit • Low/zero sludge production • Rapid start up • Sludge bulking not a problem
Disadvantages • Susceptible to membrane fouling • High aeration rates are required	*Disadvantages* • Aeration limitation • Membrane fouling • Membrane costs • High operating costs • High pumping cost (60 to 80) % of total costs) (Gander *et al.,* 2000) • High cleaning requirement • Process complexity

Drivers of MBR to the market

Figure 2.15 shows the factors influencing the MBR market which are generally acknowledged to having the greatest influence:

(a) New, more stringent legislation affecting both sewage treatment and industrial effluent discharge.
(b) Local water scarcity.
(c) The introduction of state incentives to encourage improvements in wastewater technology and particularly recycling.
(d) Decreasing investment cost.
(e) Increasing confidence in and acceptance of MBR technology

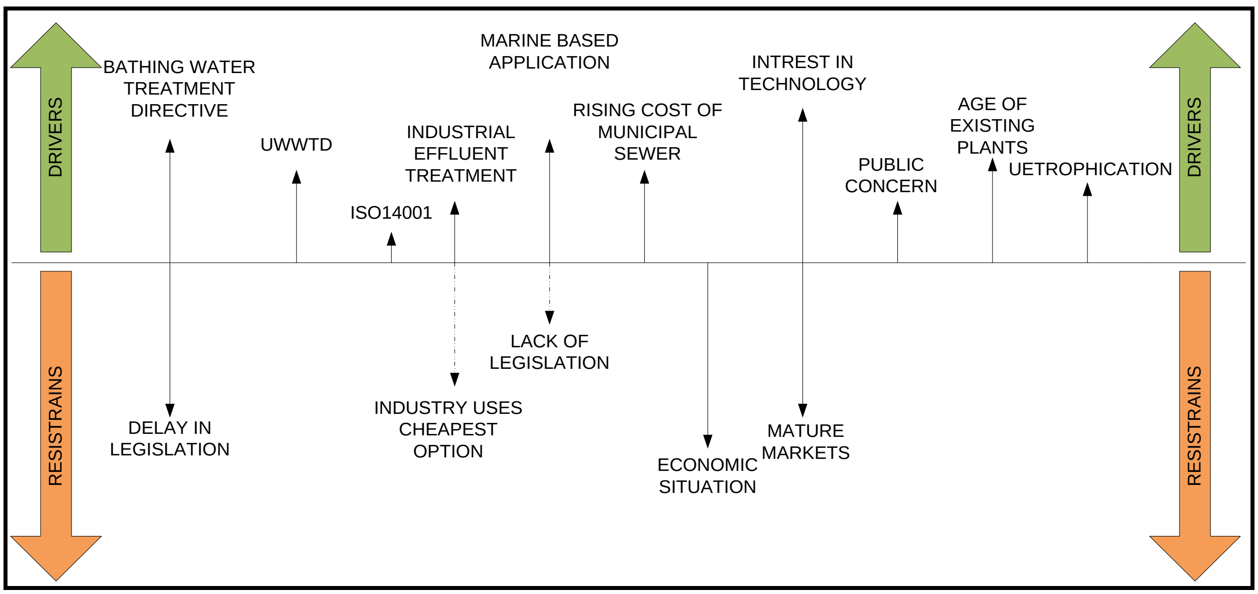


Figure 2.15: Force field analysis, growth drivers and restraints

Factors influencing the market both positively ("drivers") and negatively ("restrains") are shown, the longer arrows indicating the more influential factors. Dotted lines indicate where the influence of a particular factor on the European market is subsiding (Frost and Sullivan, 2003 and Judd, 2006)

Perspective of the future development of MBR

The MBR will not remain just one of many wastewater treatment technologies, but is expected to become a core technology that will be used for various types of wastewater management. Figure 2.16 shows the possible development of MBR.

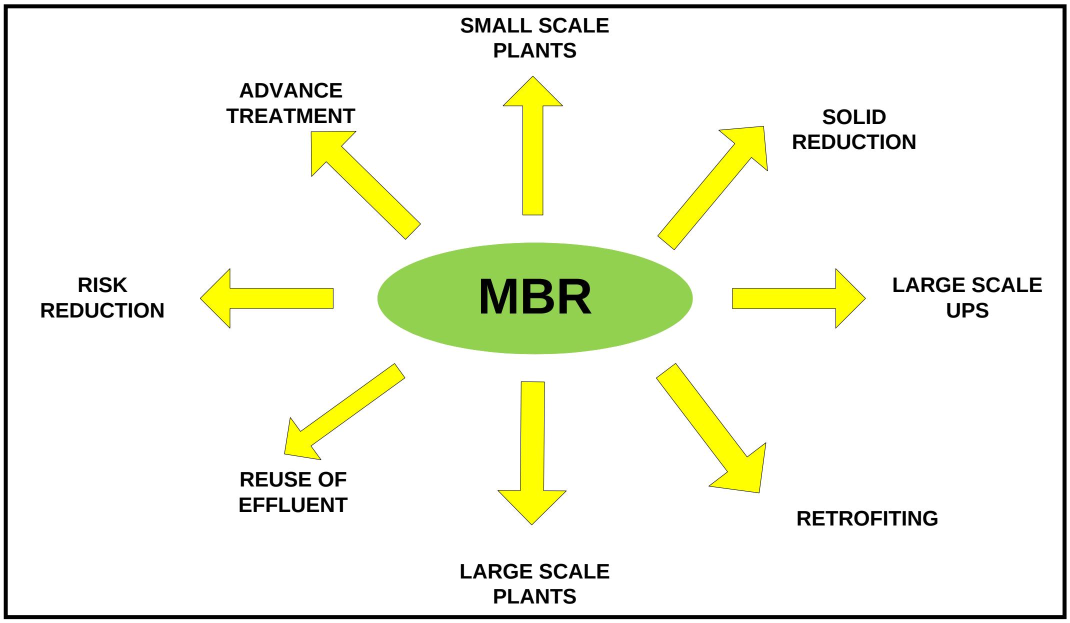


Figure 2.16: The possible deployment of MBR (Adapted from Murakami *et al.,* 2010)

Perspectives of MBR market

According to the technical market research report of a United State-based business Communication CO Inc. (BCC, 2006), the global MBR market was valued at an estimated US$216.6 million at that time, and was rising at an average annual growth rate (AAGR) of 10.9%. It was expected to approach $481.701 million in 2013. This market is growing faster than the larger market for advanced wastewater treatment equipment, at about 5.5% AAGR, and more rapidly than the market for other types of membrane systems, which are increasing at rates of 8 to 10%, depending on technology.

2.4.10 Advantages and disadvantages of MBR

Table 2.8 shows advantages and limitations of IMBRs, the previous table 2.7 showed the advantages and disadvantages of submerged and side stream MBRs.

Table 2.8: Advantages and disadvantages of IMBRs (Stephenson *et al.,* 2000; Metcalf & Eddy, 2003; Crawford *et al.,* 2000).

Advantages	***Disadvantages***
Small footprint	High membrane costs
Complete solids removal	High capital cost
High loading rate capability	Membrane complexity
Combined COD, solids and nutrients	Membrane fouling
Removal in a single unit	Operation and maintenance
Low/zero sludge production	Energy costs
Rapid start up	Aeration limitations
Sludge bulking not a problem	
Effluent disinfection, barrier against pathogens such as the chlorine-resistant organisms, cryptosporidium and Giardia.	

The advantage of low sludge production can be explained by the ability to operate the MBR at high SRT. This creates a condition of substrate limitations and results in a low food to microorganism ratio (F:M) (Stephensen *et al.,* 2000). Because the membrane replaces the clarification process, the cost and the settling problem associated with secondary clarifiers are eliminated. It also allows operation at high SRT without requiring larger aeration volumes, because the MLSS in the system is high. Due to this the MBR creates a small footprint, and requires less space than conventional treatment facilities (McInnes, Alexander and Corey-Schneider, 2001).

MBR produces a high quality effluent; typical suspended solids are less than 1 mg/l. Also nitrification, denitrification and chemical phosphorus removal have been accomplished successfully with MBRs (McInnes, Alexander and Corey-Schneider, 2001). The retention of all suspended matter and the most soluble compounds within the bioreactor leads to excellent effluent quality. The membrane not only retains all biomass, but also prevents the escape of extracellular enzymes and soluble oxidants creating a more active biological mixture capable of degrading a wider range of carbon sources. Since suspended solids are not lost in the clarification step, total separation and control of the SRT and HRT are possible enabling optimum control of the microbial population and flexibility in operation (Cicek, 2003). Because the membranes retain all biomass in the system, long SRT and high MLSS concentration can be achieved. This prevents nitrifying bacteria from being washed out of the bioreactor, improving the nitrifying capability. Complete nitrification has been observed with a HRT as low as 2 hours (Fan *et al.,* 1996).

The MBR is also capable of removing large numbers of bacteria and viruses. Since a thin biofilm is formed on the membrane surface the pore size is decreased, and organisms with greater diameter than the pore size will be removed (Stephenson *et al.,* 2000). The possibility of retaining a high number of bacteria and viruses results in a sterile effluent, eliminating extensive disinfection and corresponding hazards related to disinfection by-products (Cicek, 2003).

The most significant disadvantages with MBRs are the cost, which is caused mainly by the MBR itself and high energy costs due to the need for a pressure gradient and aeration (Cicek, 2003). However, the MBRs also have some operational problems like fouling, which limit the flux and leads to required cleaning which stops the operation. Aeration problems can arise, because of the high biomass concentration (Stephensen *et al.,* 2000). Additionally, when operated at high SRTs inorganic compounds accumulating in the bioreactor can reach concentration levels that can be harmful to the microbial population or membrane structure (Cicek, 2003).

2.4.11 Application of MBR in Municipal wastewater treatment

Municipal wastewater is the major field where MBR technology has been put into practice. Under the anticipation of more stringent wastewater treatment standards, many companies (tabulated in Table 2.5) have been developing and updating a series of commercialized MBRs with different capacities (Judd, 2006). At the same time, a lot of research has focused on the MBR performance test on municipal wastewater treatment. Studies on process innovation and operation optimization have also been dominantly based on municipal wastewater or synthetic domestic wastewater treatment.

2.4.12 Performance of Conventional Activated Sludge and MBRs

Table 2.9 shows the performance of the activated sludge process versus the performance of MBRs.

Table 2.9: Activated sludge and MBR effluent qualities (Adapted from Stephenson[a] *et al.,* 2000 and Cicek[b] *et al.,* 1999)

Parameter	Activated sludge	MBR
Sludge age[b] (d)	20	30
Sludge Production[b](kgVSS/COD.d)	0.22	0.27
Mean Flocsize[b](µm)	20 to 120	Less than 10
COD removal[a], %	94.5	99.0
DOC removal[a], %	92.7	96.9
TSS removal[a], %	60.9	99.9
NH_3-N removal[a], %	98.9	99.2
Total-P removal[a], %	88.5	96.9

MBRs have high performance in terms of effluent quality as compared to the Activated Sludge Process (as shown in table 2.9). MBRs have more than 96% removal in COD, DOC, TSS, NH_3N and total P removal.

Table 2.10 shows the influent and effluent of CASP and MBRs, it also shows comparisons of the two systems.

Table 2.10: Average influent and effluent characteristics and process removal performance under different operating conditions (Soriano *et al.,* 2003).

Process	Parameters		Influent (mg/L)				Effluent (mg/L)				%Performance removal	
	SRT (days)	HRT (days)	NH_4-N	TKN	TN	COD	NH_4-N	TKN	TN	COD	TN	COD
CAS	7.0	11.0	57	66	66	365	9	20	25	54	62	85
	4.2	18.0	53	65	67	397	10	21	27	38	60	91
	2.0	11.0	62	77	77	420	51	57	58	109	25	74
	1.8	10.5	49	64	63	426	29	33	33	51	48	88
MBR	6.5	10.2	74	74	74	404	6	7	17	35	77	91
	3.2	11.5	71	71	71	425	14	15	19	19	73	96
	3.0	5.1	73	73	73	340	30	36	40	28	45	92
	2.2	3.9	74	74	76	410	43	47	49	43	36	90
	2.1	12.7	72	72	72	432	32	34	36	27	50	94
	2.0	8.5	62	62	62	370	24	28	31	30	50	92

The influent to both processes were more or less the same but the effluent for both processes is different, in average MBRs has high performance in terms of quality output as shown in table 2.10. The percentage COD removal of MBR was 93% and 84% for CAS. MBR has the TN removal of 55% while CAS has the TN removal of 48%. Table 2.10 demonstrate that MBRs perform better that CAS in terms of the quality of effluent. It can also be noted that influent was more or less the same quality but the effluent was different. MBR achieved superior quality than CAS.

The following table shows the sludge production for different IMBRs and CAS process.

Table 2.11: Sludge production for various wastewater treatment processes

Treatment Process	**Sludge Production (mgVSS /mgCOD)**	**Reference**
Conventional activated sludge	0.60	Gander, Jefferson and Judd, (2000)
Submerged MBR	0.00 to 0.30	Gander, Jefferson and Judd, (2000)
Submerged MBR	0.16 to 0.31	Bouhabia *et al.,* (2001)
Submerged MBR	0.20 to 0.50	Cicek *et al.,* (2001)
Submerged MBR	0.10 to 0.16	Lee *et al.,* (2003)
Submerged MBR	0.26 to 0.32	Xing *et al*., (2003)

From table 2.11, it can be noted that the submerged MBR produces a lower sludge as compared to the conventional activated sludge process. The comparison of conventional activated sludge and submerged MBR reported by Gender et al., (2000) was using the same feed. From Gender et al., (2000) it can deduced that submerged MBR perform better than conventional activated sludge. The work reported by Bouhabia *et al.,* (2001), Cicek *et al.,* (2001), Lee *et al.,* (2003) and Xing *et al*., (2003) showed the varying sludge production but less than the one produced by conventional activated sludge process. However, different feed was used on these studies as it was reported on different.

Table 2.12 shows different effluent quality parameters from different IMBRs treatment schemes where different influent quality of wastewater were used.

Table 2.12: Summary of operation conditions of aerobic membrane bioreactor processes

Wastewater effluent	**synthetic**	**municipal**	**domestic**	**domestic**	**tannery**
Reactor Volume (L)	7	3900	4.5	66	2.25
Membrane area (m^2)	0.1	13.9	4	0.24	0.27
HRT (d)	7.8	10.4-15.6	5	30	1
SRT(d)	20-60	-	5-40	-	10.-50
MLSS (g/L)	2.4-5.5	18-20	-	-	10-40
COD (mg/L)	280	786	95-400	74-102	1500-2000
COD removal%	>95	90-95	>90	>85	93
Flux (L/m^3.d)	9	432-648	-	-	6.7-3.5
TMP (kPa)	-	18-26	-	-	27
Reference	Lee et al., (2003)	Rosenberger, and Kraume, 2002(a).	Huang, Gui and Qian, (2001)	Jefferson et al., (2001)	Yamamoto and Win, (1991)

The operating parameter of Aerobic Membrane Bioreactor Processes varies from type to type of effluent for example Synthetic, Municipal, Domestic and Tannery. The effluent quality output is more or less the same despite the different influent properties.

Table 2.13 shows quality performance of MBR and Activated Sludge Process.

Table 2. 13: Activated sludge and MBR effluent qualities (Stephenson *et al.,* 2000).

Parameter	**Activated sludge**	**MBR**
COD removal, %	94.5	99.0
DOC removal, %	92.7	96.9
TSS removal, %	60.9	99.9
NH_3-N removal, %	98.9	99.2
Total-P removal, %	88.5	96.9

Table 2.13 demonstrates that MBR produces superior effluent quality compared to Activated Sludge Process. MBR process achieves COD removal of 99% while Activated Sludge Process achieves 94.5%.

It can also be noted that MBR removes Total Suspended Solids by 99.9% while the Activated Sludge Process achieves 60.9% TSS removal.

MBR achieves good removal of Total Phosphorus by 96.9% while Activated Sludge Process achieves 88.5%.

Table 2.14 shows the performance comparison of Activated Sludge Process and MBR.

Table 2. 14: Performance comparison of activated sludge (AS) with MBR (Cicek *et al.,* 1999).

Parameter	AS Process	MBR Process
Sludge age (d)	20	30
COD removal (%)	94.5	99
DOC removal (%)	92.7	96.9
TSS removal (%)	60.9	99.9
Ammonia N removal (%)	98.9	99.2
Total P removal (%)	88.5	96.6
Sludge production (kg VSS/COD. d)	0.22	0.27
Mean floc size(µm)	20 to120	Less than 10

Table 2.14 shows that MBR Process high removal of COD, DOC, TSS, Ammonia-Nitrogen and Total Phosphorus compared to Activated Sludge Process. Table 2.14 also shows that MBR Process is not limited by solids. It achieves sludge age of 30 days and sludge production of 0.27 kg VSS/COD. d.

MBR process handles effluent with mean floc sizes less than 10 µm while Activated Sludge Process can only handle effluent with mean floc sizes ranging from 20to 120 µm.

Table 2.15 shows operating conditions and nutrients removal of CASP and MBRs in municipal wastewater treatment.

Table 2. 15: Nutrient removal and process conditions in MBRs and conventional activated sludge process for municipal wastewater treatment (Kraume *et al.,* 2005).

	Unit	**Conventional ASP**[a,b,c,d]	**MBR**[b]	**MBR**[c]
SRT	d	10.25	>30	30
HRT	h	4-8	>6	8
MLSS	kg/m^3	5	12-16	-
BOD	$kg/ (m^3.d)$	0.25	-	-
Loading rate	**-**	0.32-0.64	-	-
BOD(F/M)	kg/(kg d)	0.05	>0.08	-
BOD removal	**%**	85-95	-	-
Effluent conc.	mg/L	15	-	-
COD removal	**%**	94.5	-	99
Effluent conc.	mg/L	-	>30	-
TSS removal	**%**	-	60.9	-
TSS	mg/L	10-15	-	-
Turbidity	NTU	4	0.7	0.66
N_{total} removal	**%**	-	-	-
Effluent conc.	mg/L	<13	<13	-
NH_4^+ removal	**%**	98.9	-	99.2
P_{total} removal	**%**	88.5	-	96.6
Effluent	mg/L	-	0.8-1	<0.3

[a]Mudrack , 1985; [b]Cui, Chang and Fane, 2003; [c]Cicek *et al.,* 1999; [d]Gander, Jefferson, and Judd, 2000.

MBRs are not limited by MLSS concentrations as compared to the conventional activated sludge process. Conventional activated sludge process can only treat influent with MLSS that range from 4-5 kg/m^3.The MBR proved to have a better quality output as compared to the conventional activated sludge process. MBR and conventional activated sludge process achieve a COD removal of 99% and 94.5% respectively.

2.5 Membrane bioreactor theory

Calculation of flux, J

The key parameter in any membrane process is the flux, J:

$$J = \frac{Q}{A} \tag{2.2}$$

J ($l/m^2/h$ or LMH) is described as the flow rate of permeate Q (m^3/h), per surface area of the membrane A (m^2).

The transmembrane pressure is given by the difference between the static pressure (P_{stat}) and the pressure imposed by the pump through the membrane ($P_{filtration}$):

$$TMP = \Delta P = P_{static} - P_{filtration} \tag{2.3}$$

Calculation of transmembrane pressure

The relationship between TMP and J is provided by:

$$J = \frac{\Delta P}{\mu(Rm + Rc)} \tag{2.4}$$

Where R_m is the resistance of the membrane (m^{-1}), R_c is the resistance of the cake layer (m^{-1}), μ is the fluid viscosity ($Pa.s$).   .

Calculation of permeability

$$K = \frac{J}{\Delta P} \tag{2.5}$$

Where K is the membrane permeability ($l/m^2/h/bar$)

Calculation of sludge retention time (SRT)

$$SRT = V/Q \tag{2.6}$$

SRT is the sludge retention time (day); V is the volume of reactor (l).

Calculation of hydraulic retention time (HRT)

$$HRT=V/VDS \tag{2.7}$$

HRT is the hydraulic retention time (h), and VDS is the discharge sludge volume (l/d).

2.5.1 Critical flux

For microfiltration, critical flux is defined as that flux, at start-up, below which no irreversible fouling takes place (Field *et al.,* 1995). Therefore, when operating below the critical flux the TMP and permeate flux will always be directly proportional in a steady state. The moment the flux is increased above the critical flux, the TMP will continue to increase with constant flux operation or the flux will continue to decrease with constant TMP operation. This has been visually confirmed by Li *et al.,* (1998) when cross-flow of particulate suspensions at low fluxes was observed under a microscope. There were distinct conditions of operation where the particles did not deposit onto the membrane.

Knowledge of critical flux is important for commercial plants. Operating below critical flux implies that the membrane was fouled less significantly and this allowed 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 prolonged membrane life.

The reversibility of the fouling at sub-critical fluxes and the irreversibility at super-critical fluxes was also confirmed by Defrance and Jaffrin (1999). It is believed that the critical flux is a function of the system hydrodynamics and the nature of the membrane, as well as the nature of the material retained at the membrane interface (Stephenson *et al.,* 2000).

From equation 2.3, if there was no fouling, $R_c = 0$ and equation 2.3 will reduce to

$$J = \frac{\Delta P}{\mu(Rm)}$$

There will be no TMP variation at any given flux operation, the TMP and Flux relationship is represented below (see sub-critical zone on the left portion of figure 2.17):

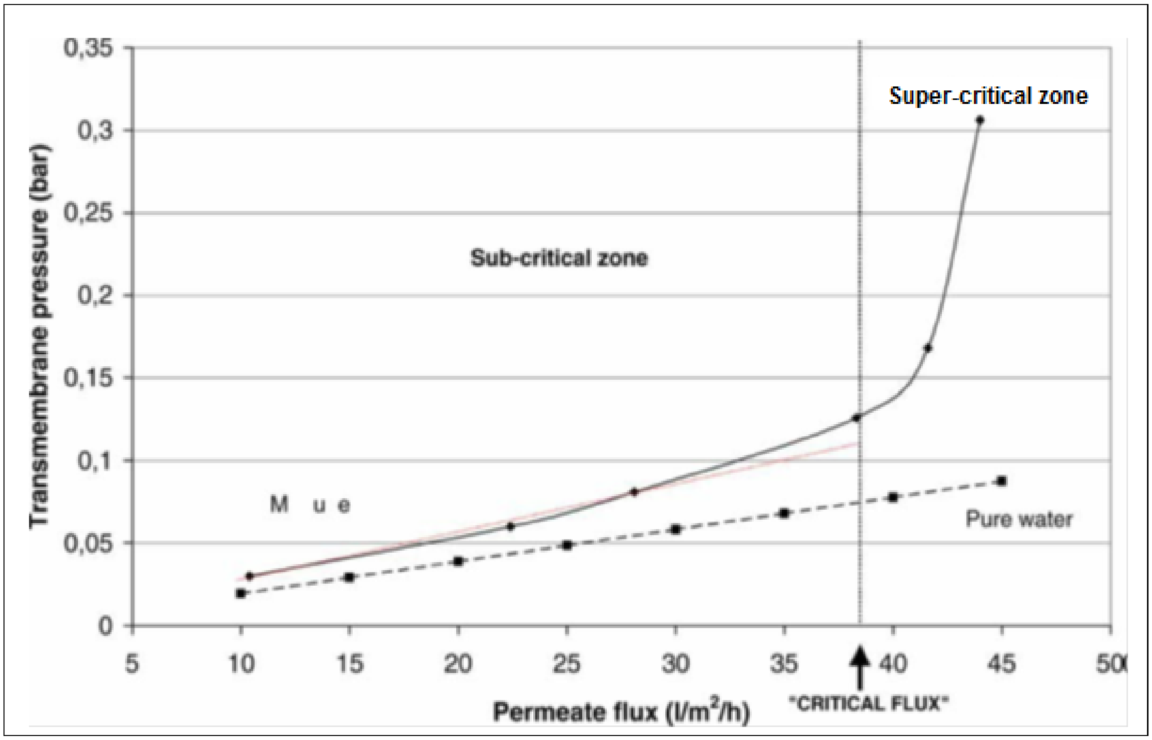


Figure 2.17: Experimental determination of critical flux (Ognier, Wisniewski and Grasmick, 2004)

Figure 2.17 shows the experimental results obtained for different levels of imposed permeate flow for a side stream MBR (Ognier, Wisniewski and Grasmick, 2004). It is found that above a flux value of approximately 40 L/m^{2}h, a clear break occurred in the curve with a significant change in pressure. This shift in TMP trend is characteristic of biological floc deposition on the membrane and thus of super-critical condition exist. Critical flux can be defined as a flux, when exceeded, the TMP increases constantly and below it the Flux and TMP is directly proportional.

When fouling does take place J_c is exceeded and fouling occurs. The cake resistance Rc also increases with increase in time (see super-critical zone on the right portion of figure 2.17):

Therefore

$$J = \frac{\Delta P}{\mu(Rm + Rc)}$$

2.5.2 Relationship between MLSS, aeration and Flux on critical flux

Figure 2.18 shows relationship the between MLSS, flux and aeration rate.

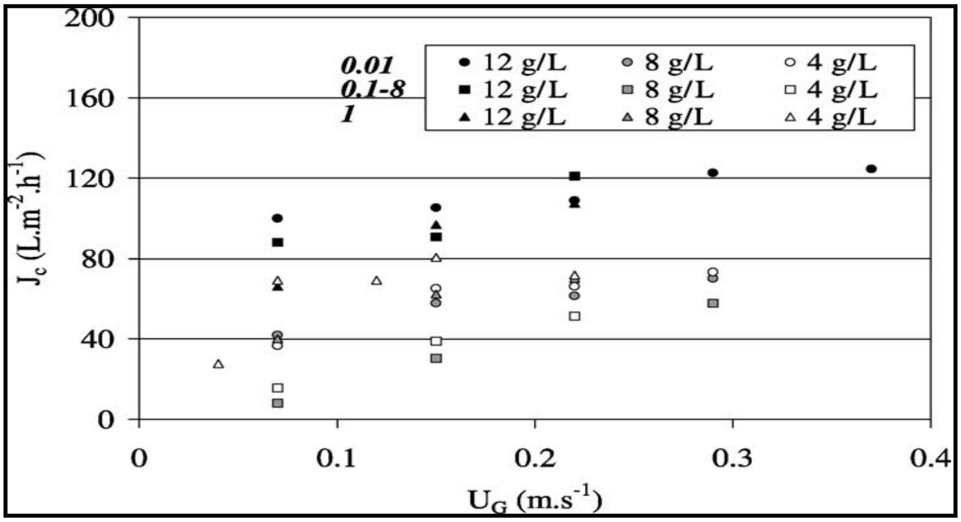


Figure 2.18: Effect of aeration, mixed liquor suspended solids concentration and membrane pore size on the critical flux value. (●) 0.01µm pore membrane, (▲) 0.1–8 µm pore membrane, (■) 1 µm pore membrane. Black: 12 g/L MLSS; grey: 8 g/L MLSS; white: 4 g/LMLSS. (Adapted from Jefferson *et al*., 2001).

It can be observed clearly from Figure 2.18 that aeration, mixed liquor suspended solids concentration and membrane pore size has the effect on critical flux. Although this curve expressed more vividly the effect of membrane pore size however, it can also observed that mixed liquor suspended solids has the significant effect on critical flux. It can also be observed that aeration rate or air velocity has the direct effect on critical flux. The increase of aeration also increases the critical flux (refer to figure 2.18). This is also in agreement with Howell *et al*., (1992), Field et *al.,* (1995) and Le Cletch *et al*., (2005).

2.5.3 Relationship between TMP and Flux at subcritical flux operation

The current trend in MBR is to operate at constant flux and monitor TMP rise. Since fouling rate and cleaning frequency increases with increase in imposed flux, it is favorable to operate the MBR at moderate flux i.e. sub-critical flux. Ognier, Wisniewski and Grasmick, (2004) analyzed long-term variation in membrane permeability under sub-critical flux conditions with no intermediate membrane regeneration in side-stream MBRs. During prolonged runs, two distinct periods were identified as shown in Figure 2.19.

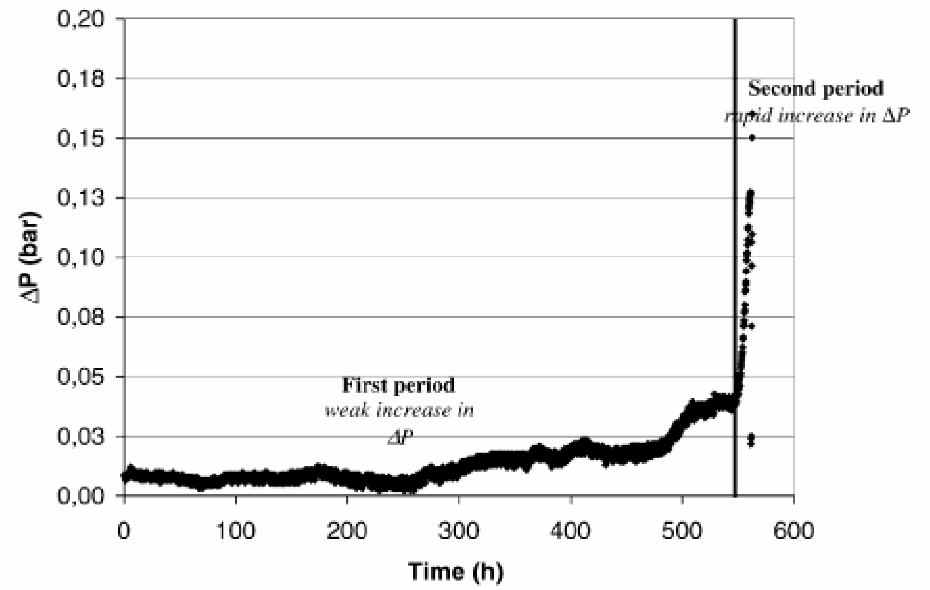


Figure 2.19: TMP change during long term constant sub-critical flux conditions (Ognier, Wisniewski and Grasmick, 2004)

In the first period, it is noticed that the initial choice of sub-critical condition in a long run does not prevent the gradual fouling of the membrane. Moreover, the period during which fouling gradually occurred appeared to be irreversible due to adsorption or colloidal fouling. In the second period, a marked increase in fouling rate was observed which reflected super-critical condition with cake layer formation and hydraulically reversible (Ognier, Wisniewski and Grasmick, 2004).

During the first period, solute-membrane or colloids-membrane interactions provoke a reduction in the number of pores open to the filtrate flow. This reduction of the area open to the flow is expressed as gradual increase in local flux in the pores remaining open. In the absence of regular membrane regeneration, the local flux increase slowly intensifies as the pores close and may lead to the local flux reaching a level equal to critical flux value and leading to a steep rise in the TMP. A deposit then forms on the membrane initiating a very high hydraulic resistance which marks the onset of cake formation. This fouling mechanism known as ‘local filtration flux concept’ is depicted in Figure 2.19.

Fane, (2002) and his research group on membrane fouling has extensively investigated the fouling behavior pattern. In one of their earlier studies, long term experiments in anaerobic side-stream MBRs revealed a two-step pattern (Cho and Fane, 2002) as shown in Figure 2.20.

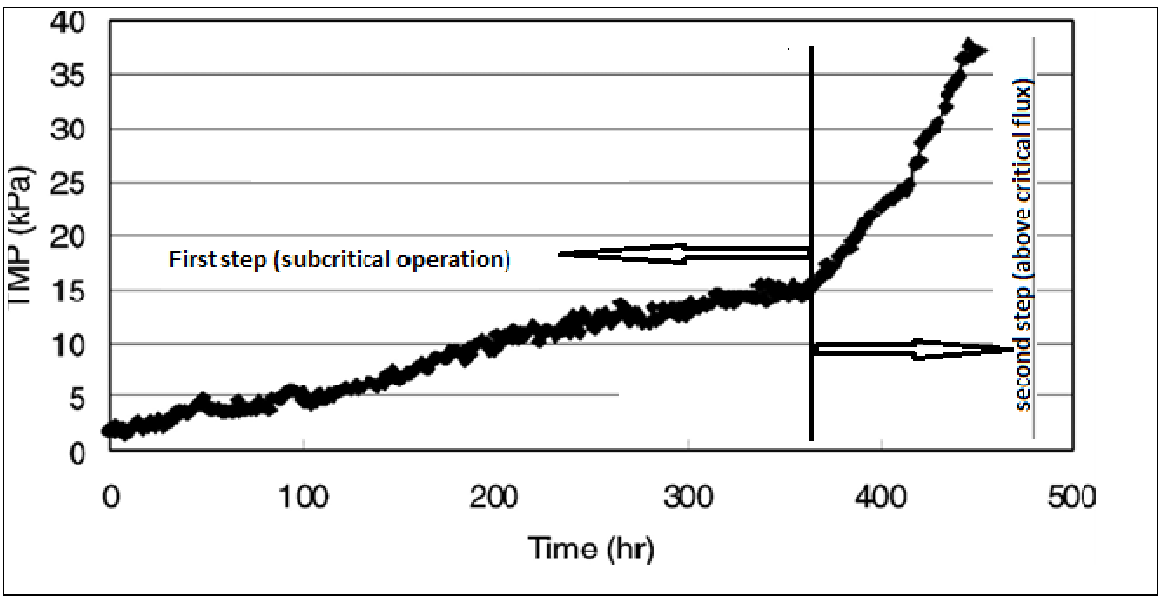


Figure 2.20: Long term TMP profile at imposed flux of 30 L/ (m^2.h) (adapted from Cho and Fane, 2002).

Prior to these two filtration steps, a conditioning step was reported in recent publications referred to as Stage 1 (Zhang *et al.,* 2006). This study by Zhang *et al*. (2006) reported a detailed mechanism involved in the three fouling stages.

The three fouling stages are discussed as follows:

Stage 1-conditioning fouling

"At the beginning stage of MBR operation, there is strong interaction between the new membrane and colloids and solute, mostly SMP present in the mixed liquor resulting in adsorption as well as colloidal pore narrowing or blocking which is mostly irreversible. This initial irreversible resistance is referred to as 'conditioning fouling" (Lapara, Klatt and Chen, 2006). Passive adsorption of colloids and organics has been observed even for zero-flux operation, before any deposition initiates (Zhang *et al*., 2006). The conditioning fouling has been reported to be independent of the shear intensity applied on the membrane surface in the MBR while it is dependent on the membrane pore size diameter distribution and surface chemical structure. However, its contribution to the overall hydraulic resistance at the end of the membrane filtration cycle is insignificant.

Stage 2-Steady fouling

Operating MBRs at sub-critical flux conditions causes the small bioflocs and SMP to steadily deposit on the membrane surface. Moreover, biofilm growth can initiate on the irreversibly attached bioflocs residues on the membrane surface from a first Stage (refer to figure 2.20). The biomass deposition tendency increases leading to gradual jump in trans-membrane pressure. Over a period of time, this phenomenon worsens. This steady fouling is dependent on the shear intensity and their distributions on the membrane surface provoked by the aeration rate in a submerged MBR.

Stage 3-TMP jump

At the end of second stage, with some zones of the membrane more fouled than others, the filtration through these specific zones is predicted to decrease. As a result, permeate productivity redistributes to the less fouled membrane areas or zones, for which the operating flux exceeds the critical flux. At this stage one observes steep TMP rise known as 'TMP jump'. Zhang *et al*. (2006) found that the sudden rise in TMP can also be caused by sudden changes in the biofilm or cake layer structure. It was demonstrated that under low dissolved oxygen concentration and substrate conditions in the biofilm sub-layers, the biomass could release a large amount of sugar monomers and block the membrane area, resulting in a sharp decrease in the membrane filterability.

2.6 Woven fabric microfiltration membrane technology

Woven fibre microfiltration (WFM) technology had undergone significant development at the Pollution Research Group of the University of KwaZulu Natal (UKZN) during the early 1980’s (Visvanathan and Abeynayaka, 2012). It is currently being further developed by 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 (Peter-Verbanets *et al*., 2009).

At the beginning of woven fibre microfiltration, WFM 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.

2.6.1 Woven fibre microfiltration sheet and module

The membrane is a flat-sheet woven fibre micro-filtration fabric produced locally in South Africa by a company called Galvenor (Figure 2.22). The module consists of three elements: a polyvinyl chloride (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. Figure 2.21shows the magnification of the woven microfiltration sheet.

Scanning Electron Micrograph of Woven Fibre Fabric

The following figure 2.21 shows the fifteen times magnification of WF membrane sheet.

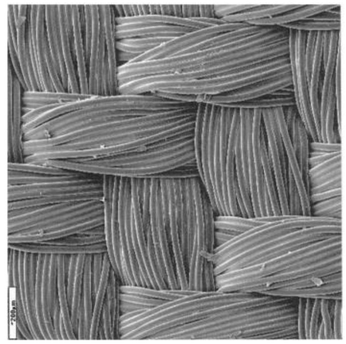


Figure 2.21: Woven fabric sheet

The following figure shows the completed WF membrane module.

Figure 2.22: Single WF membrane module

2.6.2 Historical application of WF membrane technology

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 operational skills, (Pillay, 2011). In Asia this technology has been tested in water treatment for emergency situations (Visvanathan and Abeynayaka, 2010). This work was conducted at Asian Institute of Technology (AIT). Very little has been reported in wastewater application. In South East Asia this technology has been tested in decentralised wastewater treatment systems (Visvanathan and Abeynayaka, 2012), however, this project looked at the application of the WF membrane module in membrane bioreactors for wastewater treatment and reuse.

2.6.3 Woven fibre microfiltration (WFMF) typical performance

Woven Fibre microfiltration flat sheet membranes have proved to have a good potential in potable water treatment. Small scale Remote Rural Treatment Systems (RRTS) achieved turbidity less than 1 NTU consistently from a feed ranging from 20 NTU to 300 NTU (Pillay and Kalu, 2012).

The following figure 2.23 shows the typical performance of WFMF membrane module.

Figure 2.23: Typical feed and permeate samples

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

Table 2.16 shows the performance of WFM membranes in potable water treatment application for remote rural communities in South Africa. It can be noted that WFM membrane removes E. coli by approximately 99.7% prior to disinfection.

Table 2.16: Performance of WFMF (adapted from Pillay and Kalu, 2012)

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 (counts/100 ml)
River 1	4838	980	0
River 2	8160	185	0
River 3	11191	23	0

2.6.4 Advantages of WFMF membrane technology over current commercial membrane modules

The membrane is easy to clean: Cleaning of WFM modules can be achieved by brushing and rinsing with tap water. Most of commercially developed flat sheet modules are made of cellulose material which makes it sophisticated to achieve cleaning of the modules. A good example of these modules is Kubota. WFM membrane can also be cleaned by drying the membrane; this is normally conducted by leaving the membrane in the sunlight and then peeling off the film cake at the surface of the membrane (Pillay and Kalu, 2011).

The membranes do not get damaged if dry out: most commercially developed flat sheet membranes are not user friendly for developing economies' conditions where resources/skills are scarce. A good example of these modules is Kubota once again. Kubota modules if they get scratched they cannot be repaired. WF membrane is a robust membrane that can be easily repaired if it is accidently scratched (Pillay and Kalu, 2011).

It achieves a permeate quality of less than 1 NTU turbidity (Pillay and Jacobs 2005). The WF membrane modules achieve more or less the same performance as commercially developed flat sheet modules in terms of COD removal, MLSS removal and turbidity rejection (Pillay and Dlamini, 2010).

The membranes are inexpensive. WF membranes are produced locally in South Africa. It is cost effective because it can be sourced locally and there will be no waiting for shipping of the material.

CHAPTER 3: DEVELOPMENT OF A FLAT SHEET WOVEN FABRIC MODULE, MEMBRANE PACK AND MEMBRANE PACK HOUSING FOR IMMERSED MEMBRANE BIOREACTORS

3.1 Introduction

WFM-IMBR development and design started in 2008 (Pillay and Jacobs, 2008). This was the collaboration of Durban University of Technology (DUT), University of Stellenbosch and Water Research Commission (WRC) of South Africa.

This collaboration developed the membrane module and membrane module pack. Geometry for membrane pack housing was developed however, there were shortcomings in the spacing between modules (gap) and the membrane pack compactness.

Guidelines for membrane pack housing can still be used today for continuous development of WFM-IMBR (Pillay and Jacobs, 2008). This is because of its outstanding performance in circulation thus keeping the MLSS concentration constant and achieving perfect mixing.

Parameters for membrane pack housing developed in previous studies of WFM-IMBR is being used in this project however, PVC sheet was used as the material of construction because it is inexpensive and user friendly for construction (Pillay and Jacobs, 2008).

A new pack of WFMF flat sheet modules equivalent to the size of a small commercial scale WFM-IMBR was developed, resulting in the membrane pack comprising of twenty modules. Optimum Membrane pack spacing or gap sizing is developed taking into consideration the optimum bubble size from bubble size distribution base curves. The overall objective of this chapter was to develop the WF module, the WF membrane pack and the WF membrane pack housing for the WFM-IMBR applications.

3.2 The flat sheet membrane

The membranes that were used throughout the project were the woven fabric microfiltration (WFM) membranes. The membrane is a flat-sheet woven fabric produced locally in South Africa. It is depicted in Figure 2.23. Figure 2.23 shows magnification of the woven microfiltration sheet. The sheet has an effective pore size of 1 to 2 µm.

3.3 Criteria for the development of WF membrane module, WF membrane pack and WF membrane pack housing

The criteria for the development of flat sheet WF modules:

i. it should be simple and robust, and inexpensive in construction;
ii. it should give a good permeate quality;
iii. it should give a good flux that would be viable for IMBR applications;
iv. it should require easy and inexpensive cleaning strategies;

The criteria for the development of flat sheet membrane pack:

i. the membrane pack should have the capacity to treat domestic wastewater for a target market of hundred person equivalence (at least 20 houses).
ii. the membrane pack should be designed ideally for niche processes, ski resorts, remote residential complexes and hotels.
iii. the membrane pack should be easy to clean using tap water and a brush.
iv. the membrane pack modules should be robust and inexpensive.
v. the membrane pack should have optimum gap spacing that has good scouring output of the membrane pack.
vi. the membrane pack should guarantee good performance in terms of pathogen removal, high solid rejection and high COD removal.
vii. the membrane pack needs to have sufficient free volume around it in order to allow stable circulation of solids.

Criteria for the development of the WF membrane pack housing:

i. the material of construction should be inexpensive
ii. the material of construction should be easy to mold into WFM-IMBR casing

3.4 WF Membrane module

WF membrane modules were made in house at Chemical Engineering Laboratory, Durban University of Technology. The module consisted of three elements: a PVC frame that incorporates a permeate outlet, two sheets of WF membrane glued to either side of the frame, and a spacer between the sheets of fabric to enhance fluid flow internally in the module. The PVC frame has a 2 mm nipple which allows permeate to flow out of the module. The modules dimensions used in this project were 51 cm by 35 cm with a nominal pore size estimated to be 1 to 2 µm. The total surface area of a module was 0.357 m^2 (refer to figure 3.1).

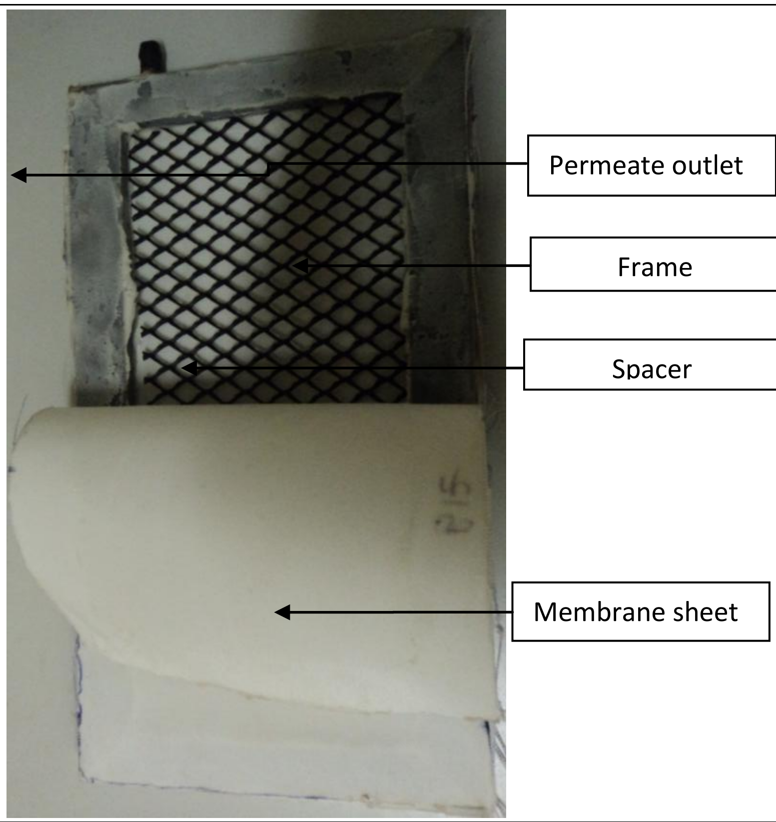


Figure 3.1: Final WF module design.

3.5 WF Membrane pack

Module gap

There is no systematic study of the effect of the gap width that has been reported. The gap between the modules has a major influence on fluid flow patterns, and eventually the effectiveness of the air-scouring. The large-scale IMBRs (Kubota and Toray) use gaps of between 6 mm and 7 mm. Weisse Technology uses gaps of 5.5 mm (small module packs) and 6 mm (large module packs) .

Prior to this development of the WF module and WF membrane pack, two phase flow characterization investigation was conducted for the effect of diffuser pores sizes and aeration flow rate on bubble sizes. The goal of the investigation was to determine the optimum bubble size for membrane scouring. Although there was no repeatability conducted, it was observed that (2 to 5) mm bubbles were dominant at all aeration rates investigated at different diffuser pore sizes used (see appendix L for the procedure used to determine the bubble size diameter). It was also observed that more bubbles were achieved when a 2 mm diffuser pore size was used. Therefore a 5 mm gap was chosen as the optimum gap for WF membrane pack design.

Number of modules

Weise's module packs consisted of 21 modules with a filtration area of 7 m^2. In this project a further constraint was the need to have adequate free volume around the membrane pack for biomass circulation.

Based on the above criteria, it was found that a module pack of 18 to 20 modules was feasible. The pack of 20 modules was chosen, giving a filtration area of 7.14 m^2.

Assembling of WF Membrane pack

The membrane pack was developed by assembling 20 identical WF modules of 0.56 m by 0.355 m into one pack (Figure 3.3). All WF membrane modules were made in house at Chemical Engineering Laboratory. Four holes are drilled into each module. Threaded rods are inserted through the holes, and each module is secured in place by nuts.

Hence, the gap between the modules can be maintained. The individual modules were connected to the permeate manifold by PVC piping (Figure 3.2). Handles were designed to act as shock absorbers in order to protect permeate outlet and manifold during membrane pack cleaning (see Figure 3.2 and 3.3). Handles allow the pack to be easily removed, as well as facilitate the positioning of the product manifold.

The product manifold is simply a pipe with nipples attached to it. Once the pack is assembled, the permeate outlets from each module are connected to the permeate manifold with flexible tubing. The elements and the final membrane pack are shown in Figure 3.3.

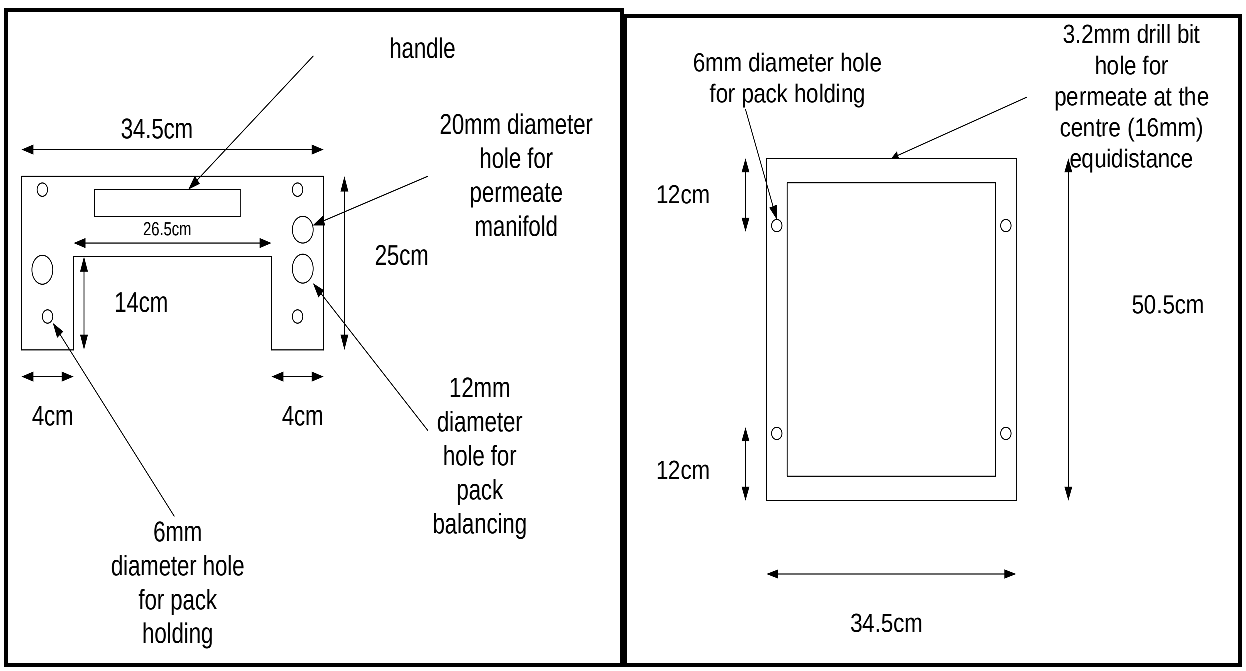


Figure 3.2: Schematic diagram for WF module and handles of membrane pack

Figure 3.3: WF membrane pack showing side view (left) and top view (right)

3.6 WF Membrane pack housing

The membrane housing consists of a frame that holds the membrane pack and the air spargers/diffusers for the air scouring. The design of the membrane housing is critical in determining whether unhindered free circulation of biomass will occur through and around the membrane pack, or whether there will be restrictions leading to poor circulation, frothing etc.

The important geometry aspects of the membrane housing are shown below (refer to Figure 3.4):

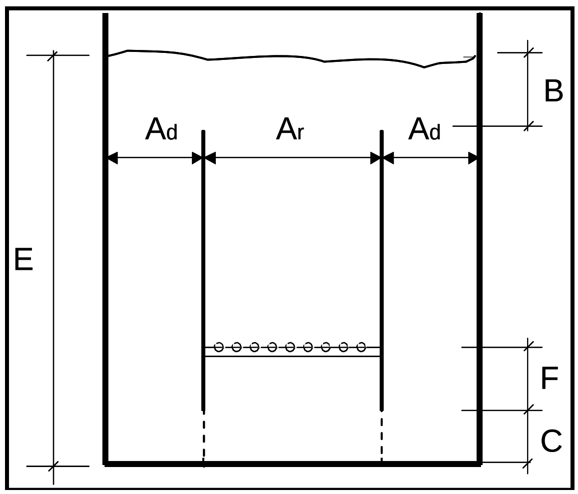


Figure 3.4: Simulated diagram of WF-MP casing

A_d : area of the down-comer section.

A_r : area of the riser section.

B : clearance height between the top of the membrane stack and the liquor level.

C : bottom entry clearance for fluid recirculation.

F : position of the air diffuser relative to the base of the membrane pack casing.

E : total liquor head in reactor vessel.

The effect that each dimension has on stable circulation is presented below in table 3.1 below and has been studied in detail (Jacobs and Pillay, 2009). The works of Jacobs and Pillay also provided guidelines for dimensions and ratios that would ensure stable circulation.

Table 3.1: Important geometry parameters of WFM-IMBR housing

Geometric parameters	**Symbol**	**description**
Area of down comer	A_d	Down-comer section creates stable circulation inside the bioreactor thus maintaining constant concentration and enhancing perfect mixing. It also maintains constant dissolved oxygen concentration.
Area of the riser section	A_r	Riser section allows contact between membrane modules and rising bubbles thus scouring the membranes. The higher the riser sections (height) the higher the degree of scouring.
Clearance heights between the top of the membrane pack and the liquor level	B	Clearance level prevents overflowing of the reactor due to scouring.
Bottom entry clearance for fluid circulation	C	It enhances circulation and makes it easier to withdraw and discharge sludge.
Total liquor head in reactor vessel	E	Total liquor level enhances the formation of bigger bubbles through collisions and pressure difference thus scouring the membrane pack.
Position of the air sparger relative to the base of the membrane pack and air entry clearance	(F)	Air sparger height help in directing aeration to the membrane pack, the higher the height of the sparger, the bigger the bubbles directed to the pack for scouring.

This project followed guidelines that were developed by Pillay and Jacobs in 2008 closely in determining the initial configuration for the membrane housing. The starting design used in this project is summarized below (refer to Figure 3.5).

PVC of 5 mm thickness was selected as a material of construction. The following dimensions were selected based on the works by Pillay and Jacobs, (2008).

- The overall height of the WF membrane pack casing was 0.8 m
- The height between membrane pack and diffuser was 0.25 m
- The height between casing base and diffuser was 0.1 m
- The head above the membrane pack was 0.1 m

Final design output

Schematic representation of final design output of the membrane pack casing

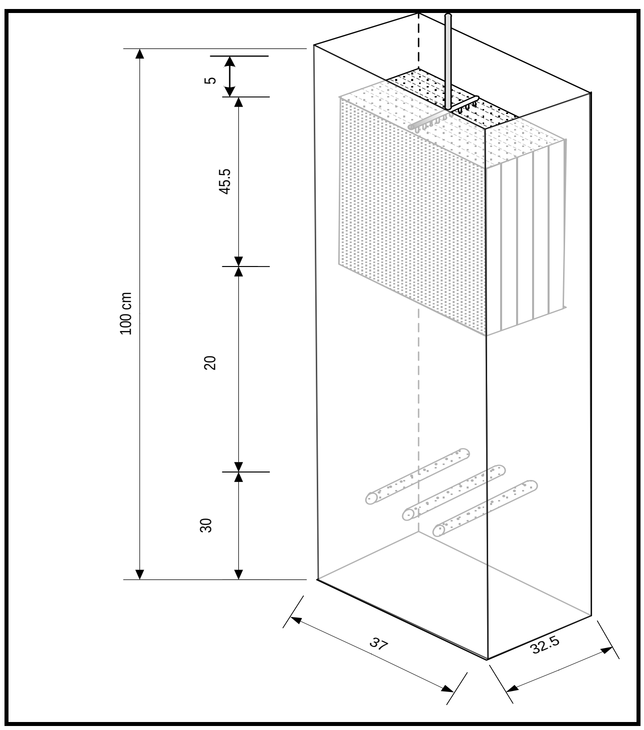


Figure 3.5: Schematic of WF membrane pack casing.

3.7 Diffuser/sparger design

Another important aspect of the membrane housing is the air sparger. The design of the sparger must ensure that:

(i) Bubbles of an appropriate size are produced, to facilitate maximum scouring

(ii) There is an even distribution of scouring bubbles across the membrane pack, to prevent “dead zones”.

Generally, the sparger and the membranes are the most expensive capital elements of an IMBR system. Many IMBRs use exotic imported silicone membrane spargers to ensure that the above criteria are met.

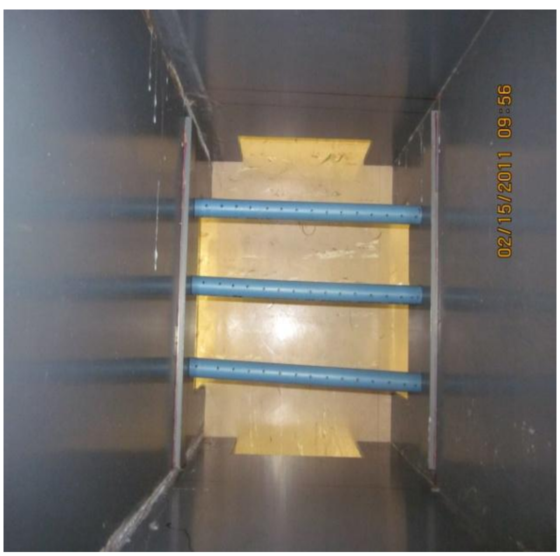
A major aim of this project was to devise an IMBR system from inexpensive and locally available materials. Previous experience indicated that a simple sparger, fabricated from a pipe with holes of the appropriate size drilled into it, worked adequately (refer to section 3.5).

Accordingly, the sparger was fabricated from a 20 mm rod which was drilled with 2 mm diameter holes. Fifteen cylindrical holes were drilled with 15 mm equidistance. The diffuser was inserted 5 cm below the membrane pack, to equally distribute scouring air in all modules. The final membrane housing, with the membrane pack, is shown in Figure 3.6.

The below figure 3.6 shows different views of the final design output of membrane pack casing

Overview of membrane case

View from side, showing air inlet for gas spargers



Top view, showing gas spargers

Permeate off-takes on panels

Figure 3.6: Different views of the WF membrane pack casing

3.8 Summary of the design output

Woven fabric microfiltration (WFM) membranes with pore sizes of 1 to 2 µm were used.

The modules of dimensions of 51 cm by 35 cm were developed in house at the Chemical Engineering laboratory. The total surface area of a module was 0.357 m^2.

The WF membrane pack of twenty modules was developed based on the scale of hundred person equivalence (50 households). A gap of 5 mm between the modules was used to promote optimum scouring of the membrane pack.

The membrane pack housing of 37 cm by 32.5 cm by 80 cm was developed. The PVC sheet of 5 mm thickness was used as a material of construction.

The following dimensions for WF membrane housing were employed:

- The overall height of the WF membrane pack casing was 0.8 m
- The height between membrane pack and diffuser was 0.25 m
- The height between casing base and diffuser was 0.1 m
- The head above the membrane pack was 0.1 m

CHAPTER 4: EVALUATION OF HYDRODYNAMICS OF WFM-MEMBRANE PACK IN AN OPERATED BIOREACTOR

4.1 Introduction

Immersed membrane bioreactor (IMBR) performance can be determined by biological nature of the feed and hydrodynamic operation of the IMBR system. This section concerns the hydrodynamics aspect part of the IMBR system. The instrument used to gauge the hydrodynamics of WFM-IMBR was critical flux and was evaluated by the step method technique. According to Field et al (1995), critical flux can be defined as a flux below which no deposition occurs on the membrane surface. In constant flux process, critical flux is often defined as the flux above which TMP starts to increase rapidly with time. Since WF membrane pack has never been operated in an IMBR there is a need to establish hydrodynamic performance of this membrane pack and to compare to commercially existing small scale IMBR systems.

One of the objectives for this project was to evaluate hydrodynamics and short term stability of the WF membrane pack in a seeded sludge bioreactor. Hence, investigation were carried out: (1) to evaluate the effect of aeration rate and sludge concentration on critical flux; (2) to derive the average critical flux-aeration curve as a function of sludge concentration and (3) to compare critical flux characteristics to other developed small scale IMBR systems.

4.2 Definition and determinations of critical fluxes

4.2.1 Definition of critical flux

Field *et al*. (1995) introduced the concept of critical flux, which is defined as being the flux below which no membrane fouling is observed.

4.2.2 Critical flux evaluation method

The critical flux was determined according to 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. In this technique, if the flux is increased at certain increments and no sign of TMP variation is observed, then the flux is defined as below critical flux at that aeration rate used (refer to Figure 4.1). 5 LMH to 35 LMH were below critical flux. At the last two steps of 40 and 45 LMH the TMP variation was observed. The initial TMP corresponding to the initial sudden increase of filtration resistance and the final TMP at the end of the step were recorded. From these two values, the average TMP can be deemed as critical flux at the given aeration and flux. The critical flux is assumed to be the flux at which there is a sudden increase of TMP is observed and above that point more fouling occurs. 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 membrane life.

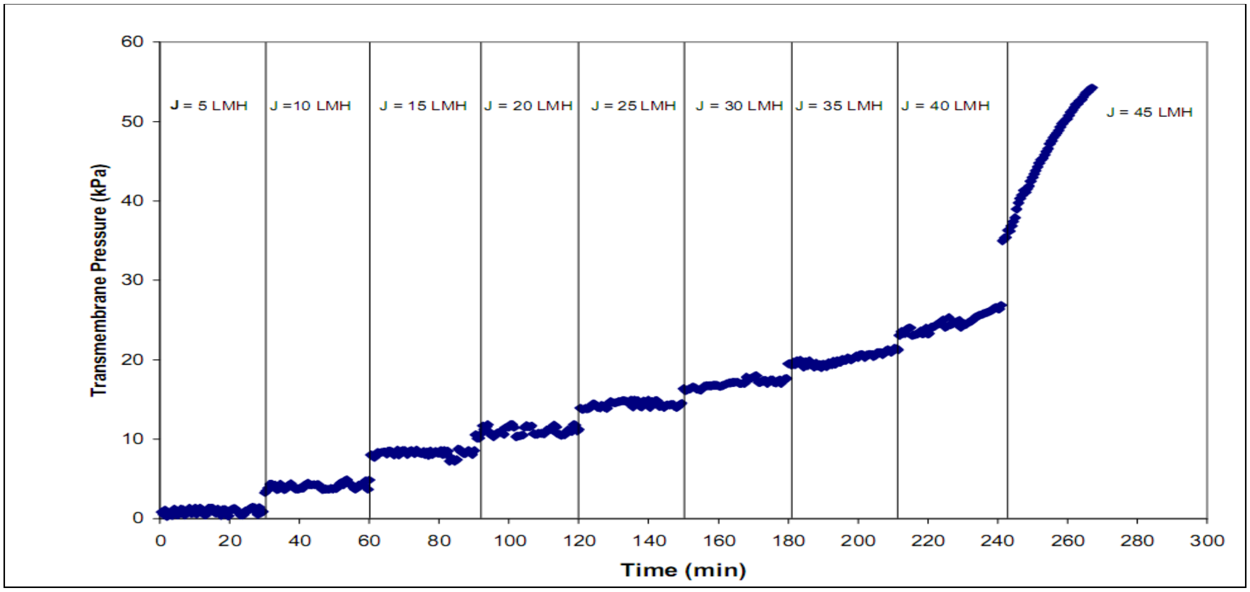


Figure 4.1: Illustration of critical flux determination by flux step method
(Adapted from Ndinisa, 2006)

The relationship between TMP and Flux was discussed in section 2.5.

4.2.3 Equipment set up

A diagram of the experimental rig used for the experiments is depicted in Figure 4.2. A pilot scale WFM-IMBR set up with 20, flat sheet woven fibre micro membrane modules with total area of 7.14 m^2 was constructed. The total area of a single module was designed to be 0.357 m^2 with 0.1785 m^2 on each side.

The system was operated as a constant flux operation and therefore TMP increased with time. The process feed tank, filtration tank and permeate tank were open to the atmosphere. The hydraulic retention time was 24 hours.

The system was fed with real activated sludge extracted from the return line of activated sludge (refer to Figure 4.2). The permeate flow rate was controlled by using a manipulated peristaltic pump. Filtered permeate from the filtration tank was compensated by the raw feed of the WFM-IMBR. Raw feed was harvested prior to the anaerobic mixing chambers. The pressure reading (TMP) was recorded using a pressure gauge meter. The SCL-KO4 blower of capacity of 100 cubic meters per hour was used for aeration. The critical fluxes at four different aeration flow rates were evaluated (2, 4, 6and 8 L/min/module).

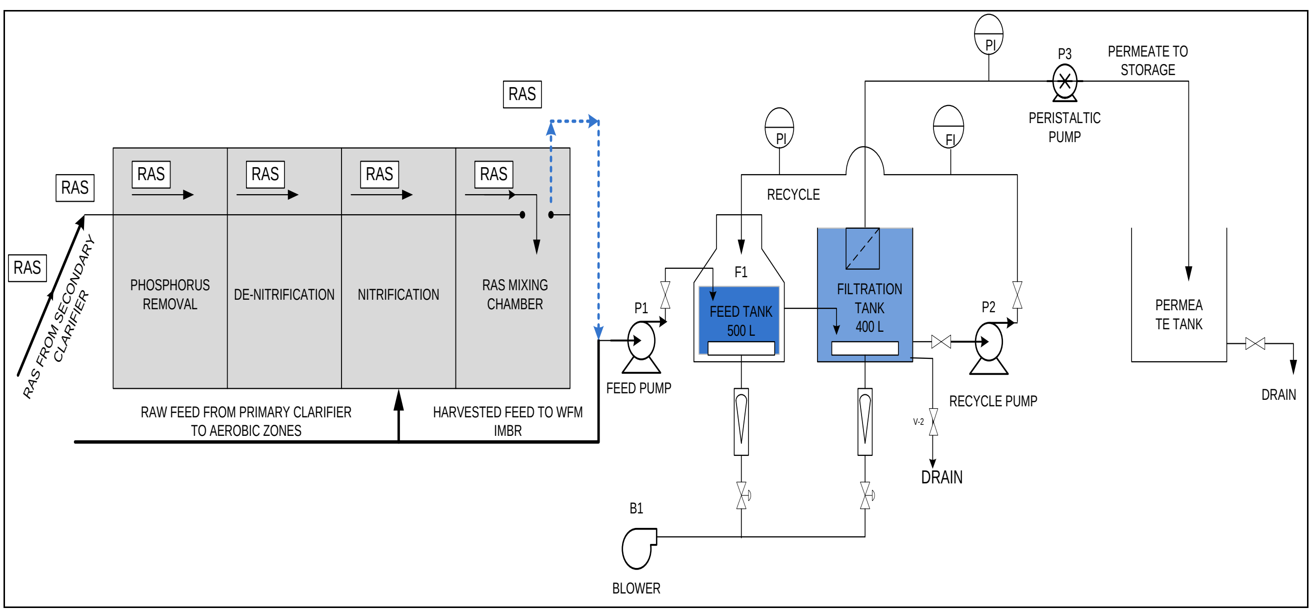


Figure 4.2: PID diagram of immersed woven fibre microfiltration membrane bioreactor.

The Feed

The primary treated raw domestic feed from households was used. It was mixed with activated sludge suspension harvested from the commercial plant. After the desired concentration was achieved, raw feed was supplied continuously to the WFM-IMBR system supplementing the filtrate. The sludge in excess was discharged on a daily basis based on desired concentration for the WFM-IMBR system.

The particle size distribution of activated sludge obtained is shown in Figure 4.3 below. From Figure 4.3, it can be determined that the mean particle diameter of activated sludge is about 100 microns. The nominal membrane pore size is estimated to be 0.1 to 0.2 µm and this implies that the main mechanism of fouling will be due to cake formation.

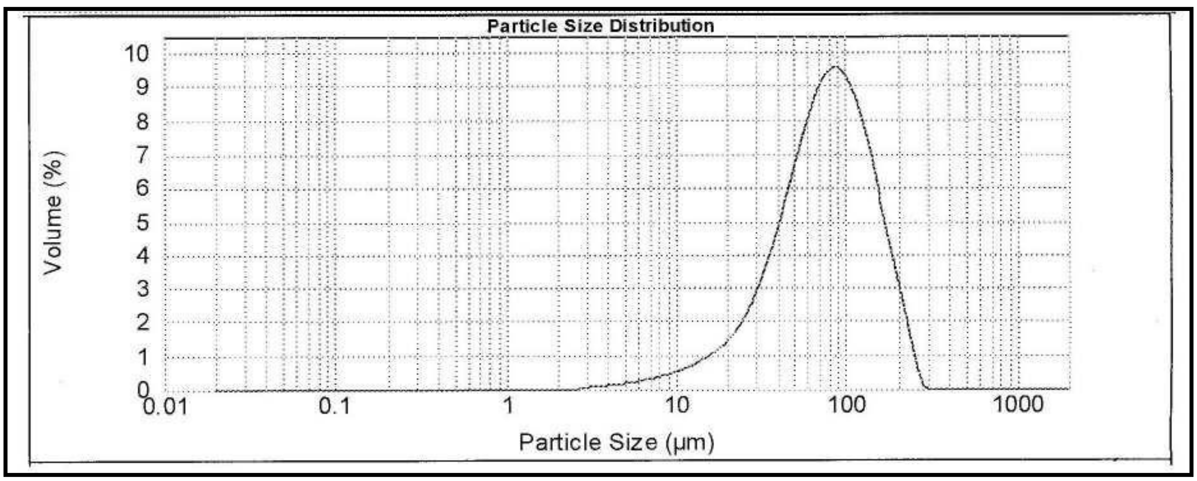


Figure 4.3: Particle size distribution of aerated waste activated sludge

4.2.4 Process description

The MBR system is the combination of a filtration technology (predominantly microfiltration) and operating bioreactor.

The aim of this technology in a conventional activated sludge process is to replace the solid separation process (clarification) by a means of a microfiltration. By replacing the clarifier, the foot print of the process significantly reduce.

This technology is ideally applicable for conventional activated sludge process upgrades where space limitation is a prime consideration.

The MBR process is dived into three tanks. First tank (biological tank) is mainly for aeration of sludge into the desired dissolve oxygen content. No filtration is taking place.

On the second tank, filtration takes place. The membrane pack is connected into the suction duty pump. The suction pump creates a partial vacuum to the immersed membrane pack in order to create pressure drop across the membrane surface.

The filtrate (Permeate) water passes through the membrane pack where the sludge is retained in the filtration tank. Sludge is periodically discharged from the filtration tank in order to maintain the stable quantity of sludge concentration in the reactor.

This technology produces a high quality effluent suitable for recycling for secondary purposes. MBR technology requires a significantly smaller space for installation.

MBR process is not limited by sludge concentration. It can operate with sludge concentration as high as 20 mg/L.

MBR process typical performance achieves acceptable COD suitable for release to the receiving environment and 100% removal of suspended solids.

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4.2.5 Equipment description

The feed, filtration and permeate tank

The feed was contained in one 450 L cylindrical tank, filtration and permeate tank were 400 L in sizes. The tanks were made of PVC material. The tanks were filled with activated sludge from the commercial plant up to 80% of its capacity. The first tank was used as the aerobic zone and the successive tanks were filtration and permeate storage tanks. The filtration tank was coupled to the feed tank and the filtered permeate was stored in the permeate tank (refer to Figure 4.2). Twenty modules were used in this study. The feed tanks were open to atmosphere.

Course bubble and fine bubble diffuser for scouring

Two diffusers were used; the in-house diffuser discussed in 3.5 was a coarse bubble diffuser. A fine diffuser was also used for microbial food supply and to keep dissolved oxygen at the desired concentration.

Figure 4.4 shows the cylindrical in-house made coarse bubble diffuser.

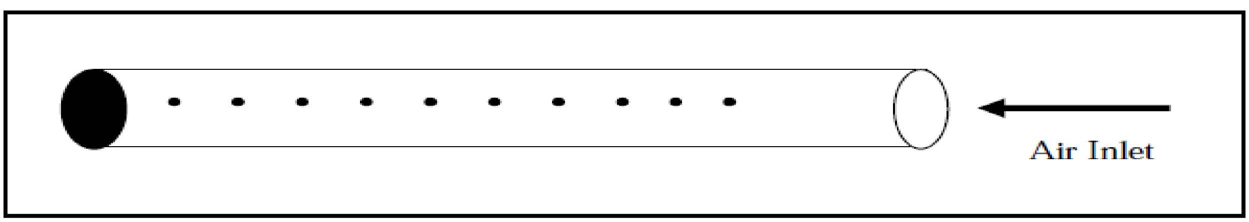

Figure 4.4: Schematic diagram of the in-house made coarse bubble diffuser.

Fine bubble diffuser for biological digestion of sludge

The circular disc for biological aeration was used. The disc was made from ethylene propylenediene monomer (EPDM) and has ability to supply air uniformly which makes it easier for microbes to utilize it as a source of energy. Low membrane plasticizer content reduces shrinkage and hardening, but is enough to avoid creeping from the disc.

Figure 4.5 shows the round disc fine bubble diffuser.

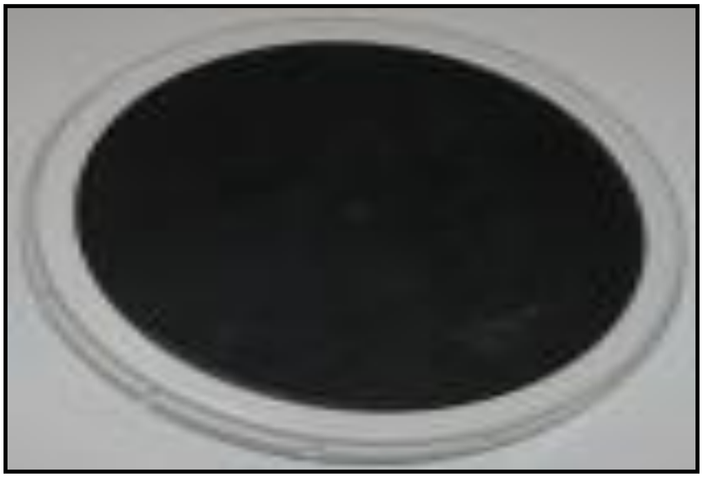

Figure 4.5: The fine bubble diffuser disc

Feed pump

A submersible pump was used for pumping the feed from the return line of activated sludge. The capacity of submersible pump was 14 m^3/hr with and a maximum head of 12 m.

The peristaltic pump

A variable speed peristaltic pump (Watson Marlow 3*S* model) was used as a driving force in sucking of permeates from the membrane pack. Marpen material for pump tubing was used. The pump had a capacity of 0.12 m^3/hr.

Recirculation pump

A single phase MPK 40 centrifugal pump was used as the recirculation pump in order to keep the concentration constant inside the WFM-IMBR system. The capacity of the recirculation pump was 400 l/hr with a maximum head of 9 m.

The air Blower

The single phase air blower SCL-KO4 MS was employed. It was connected to the diffuser passing through a globe valve and rotameter. The valve was used to regulate the air flow rate. The capacity of the blower was 1.1 kW with a maximum flow rate of 155 m^3/hr.

Pressure Transducers

The transmembrane pressure (TMP) was measured via a gauge pressure connected on the line between the membrane and a suction pump. Although the gauge produced negative data, all plot values of pressure were presented as positive values for graphical presentation of the data. The minimum and maximum pressure values were 0 and 100 kPa.

4.2.6 MBR operating parameters

The MBR operating parameters for the performance evaluation are as follows:

- Filtrate (Permeate) Flux
- Suction pressure
- Aeration

These above operating parameters are discussed below.

Flux

Flux is the rate at which permeate is filtered and passes through the membrane per unit surface area. Flux is commonly expressed in liters per square meter per hour (LMH). The operating flux throughout the pilot study was increased in 5 LMH increments to 30 LMH. The startup flux was 10 LMH.

Suction pressure

Suction pressure is used to withdraw permeate through the membrane pack. Suction pressure is expressed in kilopascal (kPa) units. The MBR operates at variable flux.
After a prolonged operation of the MBR, the membranes become fouled. The TMP increases and membrane regeneration is required. Membrane regeneration is achieved by washing of the membrane pack or by backwashing.

Aeration

Aeration is used to scour the membrane as a means of cleaning. Air is injected below the membrane pack by means of a diffuser or sparger. Air bubbles travels towards membrane surface thus combating any cake formation on the membrane surface.
Aeration is used to maintain the required dissolve oxygen content of 2 mg/L on the system. Dissolve oxygen content is important for survival of the bacteria population growth. The aeration was supplied at 80l/min into the biological tank for respiration of the bacteria purposes.

Due to head loss in aeration piping, one blower was found to be insufficient for aeration requirements. One more identical blower was needed for scouring of the membrane pack and to support microbial activities on biological tanks by supplying optimum oxygen.

4.2.7 Experimental procedure

- Sludge was allowed to grow in an operated bioreactor to the desired concentration.
- The critical flux is determined according to the flux step method. This was done by fixing the aeration parameter and varying the flux in 5 LMH increments.
- The TMP was monitored at every given flux, aeration rate and sludge concentration of 4, 8, 12 and 16 g/L.
- The aeration considered in this study was 2, 4, 6, 8 and 10 L/min/Module. Dissolved oxygen (DO) was monitored and was controlled to be above 2 mg/L at all given times.
- All tanks were open to atmosphere. After every critical flux was evaluated, the membrane pack was cleaned by brushing and rinsing using tap water.

Repeatability

Prior to the commencement of experimental investigations, the reproducibility of the results from the experimental rig was tested. Runs were carried out at various conditions, each condition being repeated at random. From these tests the repeatability of the TMP data was found to be within a margin of 10 % or less which was considered to be acceptable. Consequently, all the results presented in the rest of this thesis are only reported for cases where the experimental conditions were within one standard deviation of the mean value.

Leak test on WF membrane modules and pack

The leak test was performed before the startup of any trials. This was done by pumping air inside out of the membrane module within a tank filled with water through bubble point observation technique. The observation of bubbles was the evidence of the membrane module leak.

A result on this leak test was positive with no observation bubbles in water when the membrane module was suppressed up to 1 bar. Pure water flux of 20 LMH was also conducted at standard temperature and pressure to test the permeability of the membrane module.

Permeate data collected showed no breach of membrane integrity, as 100 % of the measurements showed turbidity less than 1 NTU.

4.2.8 Analytical methods

Mixed Liquor Suspended Solids (MLSS)

Mixed liquor Suspended Solids (MLSS) is a mixture of raw or settled wastewater and activated sludge contained in an aeration basin in the activated sludge process. Mixed liquor suspended solids (MLSS) is the concentration of suspended solids in mixed liquor, usually expressed in milligrams per liter (mg/L). It was evaluated by standard method HACH No 8029 (refer to Appendix H). The sample was dried in an oven and the dried solids were measured (mg). It was reported as mg/L.

Dissolved oxygen

Dissolved oxygen (DO) analysis measures the amount of gaseous oxygen (O_2) dissolved in an aqueous solution. Oxygen gets into water by diffusion from the surrounding air, by aeration (rapid movement), and as a waste product of photosynthesis. In this study, DO was measured using a HACH No 6780 standard method (refer to Appendix J). DO concentration was kept above 2 mg/L for the entire experimentation.

4.2.9 Results and discussion

Determination of critical flux by the step method

Figure 4.6 shows how the critical flux was determined by using the step method.

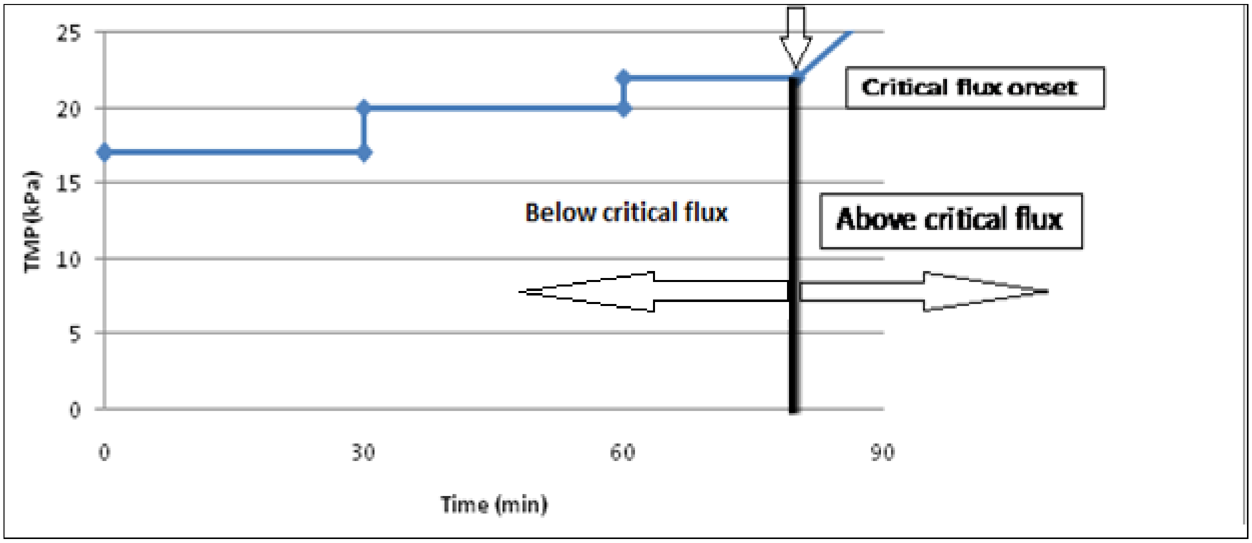

Figure 4.6: Critical flux for membrane with spacer at 10 L/min of aeration (see repeatability's on Appendix 4

Critical flux characteristics at varying aeration and mixed liquor suspended solids

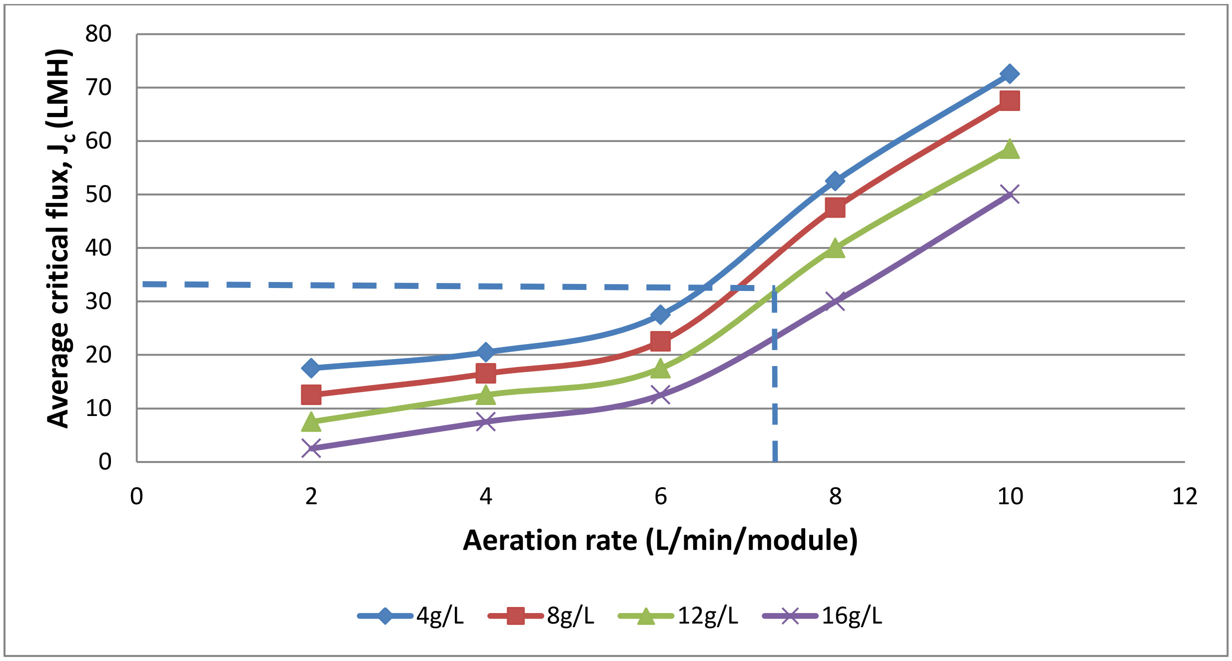

Figure 4.7: The relationship between average critical flux and aeration rates at different concentration of activated sludge.

It was observed that the critical flux decreased with the increase of sludge concentration. It was also observed that critical flux increases with the increase in aeration (refer to Figure 4.7).

These trends were expected and they were consistent with literature and theory (Fane and Ndinisa, 2006). These trends fitted well with the criteria for a selection of subcritical point for short term operation of WFM-IMBR.

The critical flux of 30 LMH at 12 g/L sludge concentration and aeration of 6.8 L/min/module was found. However the aim of this study was to establish the subcritical flux conditions and operate at that point for extended periods. Therefore any value above 6.8 L/min/module was optimum long term subcritical operation of WFM-IMBR.

Thus, the operating point of 30 LMH at 12 g/L sludge concentration and aeration of 10 L/min/module was chosen as the operating point for short term subcritical operation of WFM-IMBR system (refer to figure 4.7) based on established existing IMBR units in developed economies.

This was done for subcritical flux operation purpose. This was lower than commercial small scale systems such as Kubota, Copa IMBR and Huber IMBR which utilizes aeration of 10 L/min/module,17.5 L/min/module and 150 L/min/module respectively.

4.3 Short term sub critical operation of WFM-IMBR on a waste sludge

4.3.1 Equipment set up

The experimental setup used in this section is identical to that used in section 4.2.3 and depicted in Figure 4.2. Section 4.2 evaluated the hydrodynamics of WFM-IMBR. In this section, short term subcritical stability operation and performance of the WFM-IMBR is evaluated.

4.3.2 Experimental procedure

A semi pilot scale WFM-IMBR set up with 20, flat sheet woven fibre micro membrane modules with total area of 7.14 m^2 was constructed based on Kubota small scale recommendations. The total area of a single module was designed to be 0.357 m^2 with 0.1785 m^2 on each side.

The feed (return activated sludge and primary treated raw fresh feed) was prepared to the desired concentration. The activated sludge was aerated at 60 L/min exceeding the required dissolved oxygen concentration of 2 mgO_2 /L in all biological tanks and filtration tanks. The activated sludge concentration of 12 g/L, aeration of 10 L/min/module and flux of 30 LMH was used in this investigation (refer to figure 4.6 for operating point chosen). The experiment was conducted for nine days. The fresh feed was continuously supplied at 0.05 m^3/hr in order to compensate for the filtered permeate. Dissolved oxygen was monitored at 12 hour intervals. The permeate flow rate and trans membrane pressure were continuously recorded. The hydraulic retention time was 24 hours. The MBR was operated under critical flux or subcritical flux of 30 LMH for two weeks. Permeate was withdrawn using a peristaltic pump. The permeate MLSS, COD and turbidity was monitored at 6 hour intervals.

4.3.3 Analytical methods

The quality of feed and permeate is a primary concern in MBRs, the membrane performance in terms of water quality was monitored in this section using three parameters, namely, chemical oxygen demand(COD) removal, suspended solids removal, and turbidity.

Mixed Liquor Suspended Solids (MLSS) and dissolved oxygen

For Mixed Liquor Suspended Solids (MLSS) and dissolved oxygen see section 4.2.8.

Turbidity

Turbidity is a measure of the clarity of water and is commonly expressed in Nephelometric Turbidity Units (NTU). Suspended solids and colloidal matter, such as clay, silt and microscopic organisms cause turbidity. Turbidity was tested using a HACH 2100P meter.

However this turbidity meter had a maximum detection limit of 1200 NTU and it was found that in all the experiments the feed turbidity was always higher than this value and hence could not be measured. Only the permeate turbidity is reported.

Chemical Oxygen Demand

Chemical Oxygen Demand (COD) is defined as the quantity of a specified oxidant that reacts with a sample under controlled conditions. The quantity of oxidant consumed is expressed in terms of its oxygen equivalence. COD is expressed in $mg0_2/L$. COD was measured by the micro-COD method proposed by HACH standard test method no. 5220D (refer to Appendix G) in which COD vials, a COD reactor and a spectrophotometer is used.

4.3.4 Results and discussion

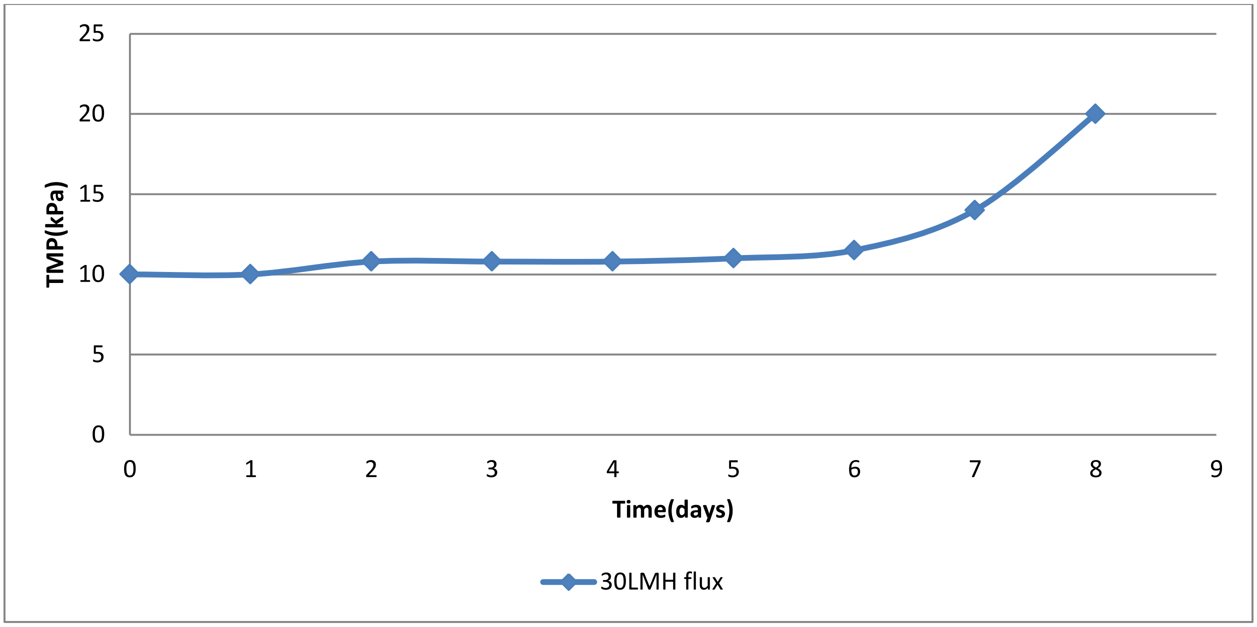

Figure 4.8: The relationship between trans membrane pressure and time at fixed subcritical flux.

The activated sludge concentration was fixed at 12 g/L. For an 8 day periods, the TMP was found to be constant for the first 6 days and there was a sudden increase in TMP for the last two days. The TMP increased exponentially from 10 to 20 kPa, this jump was not expected and it can be associated with internal processes of mixing fresh feed with the return activated sludge or possible contamination of raw feed influent into the WFM-IMBR which in turn resulted in fouling of the membrane. Due to cake formation on the surface of the membrane as a result of fouling the TMP started to vary and exponential increase was observed.

The stable stage for first six days was expected and it was in agreement with literature (refer to figure 2.19). According to Ognier, Wisniewski and Grasmick, (2004) IMBR can be operated for prolonged periods at subcritical conditions and above subcritical conditions fouling is observed. Also from theory (refer to equation 2.4) if there was no fouling, $R_c = 0$,and $J = \dfrac{\Delta P}{\mu(Rm + Rc)}$ will reduce to $J = \dfrac{\Delta P}{\mu(Rm)}$ and there will be no TMP variation at any given flux operation, therefore the TMP will be stable at that given flux. Therefore the stable part of the graph responds to no fouling which is $J = \frac{\Delta P}{\mu(Rm)}$.

However the sudden increase in TMP was not expected and this can be due to the varying quality of sludge where it was harvested or internal conditions of the sludge such as contamination. Sludge with dead microbial organism will cause fouling due to the fact that there is no micro-organism counteracting the formation of cake on the surface of the membrane. This simple means that micro-organisms are important for the success operation of WFM-IMBR. According to equation 2.4, if there is cake formation R_c, the TMP variation can be observed while if $R_c = 0$ there will be no TMP variation. Therefore when there is well alive bacteria, there will be no R_c and TMP will always be stable provided the bioreactor is operated under subcritical conditions.

This trend fitted well with the objective and criteria as it was the testing step to see whether WFM-IMBR can be operated for extended periods. Therefore it can be deduced that WFM-IMBR can be operated for extended periods at subcritical conditions provided it is operated below critical flux although the short term trial was conducted for a limited period however this was sufficient to conclude that WFM-IMBR can be operated for extended periods at subcritical flux conditions.

The COD and MLSS removal in WFM-IMBR

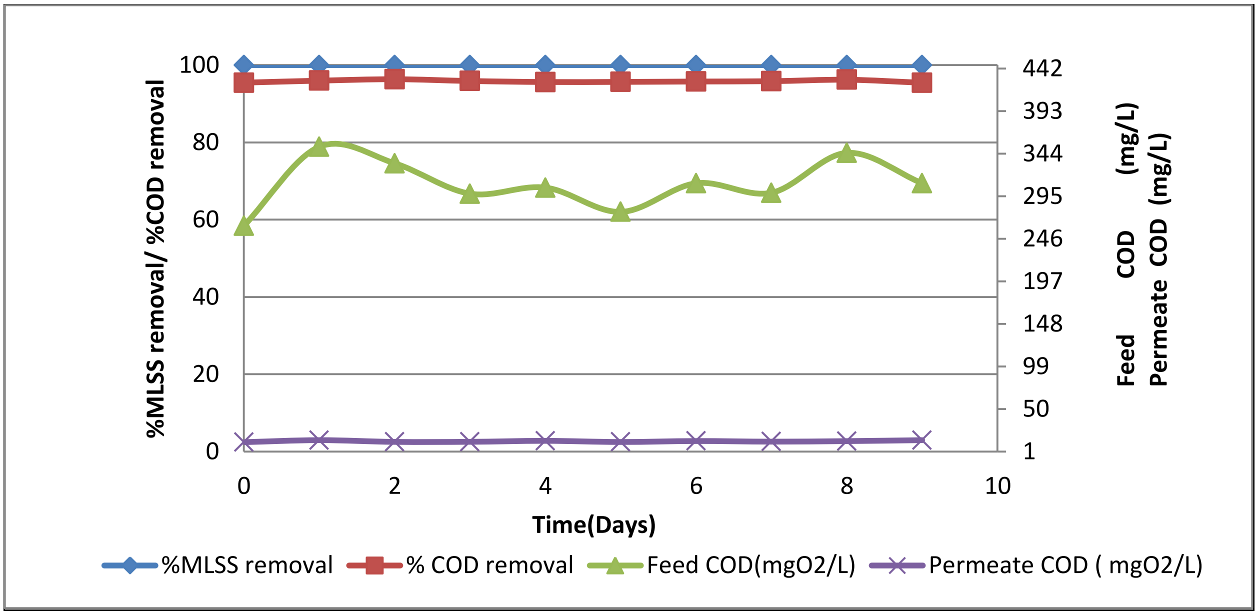

Figure 4.9: The relationship between MLSS and COD removal with time at fixed flux and activated sludge concentration.

MLSS removal was found to be stable for 9 days at 100 % removal of solids. COD removal was varying between 95% and 98% for 9 days.

These trends were expected and meaningful in terms of quality performance of the WFM-IMBR. Refer to tables 2.9 and 2.15 for influent and effluent quality of IMBRs and compare to the trends depicted in figure 4.9. Generally IMBR produces effluent that can be recycled directly for reuse. Effluent quality generated by WFM-IMBR was of the same quality to the one reported by Cicek *et al.* (1999), Stephenson *et al.* (2000), Bouhabila, Ben Aïm and Buisson, (2001), Rosenberger and Kraume, 2002 (b), Soriano *et al.* (2003) and Gander, Jefferson and Judd, (2000).

COD and MLSS removal fitted well with the criteria chosen for WFM-IMBR evaluation since it rejected 100% of solids and had a COD removal of 95%.

According to Mudrack,(1985); Cui, Chang and Fane, (2003); Cicek *et al.* (1999),Gander, Jefferson and Judd, (2000) and Judd , (2006) IMBR generally remove 100% of solids and achieve above 95% of COD removal and therefore data represented in figure 4.9 is in agreement with literature.

Solid rejection performance of WFM-IMBR

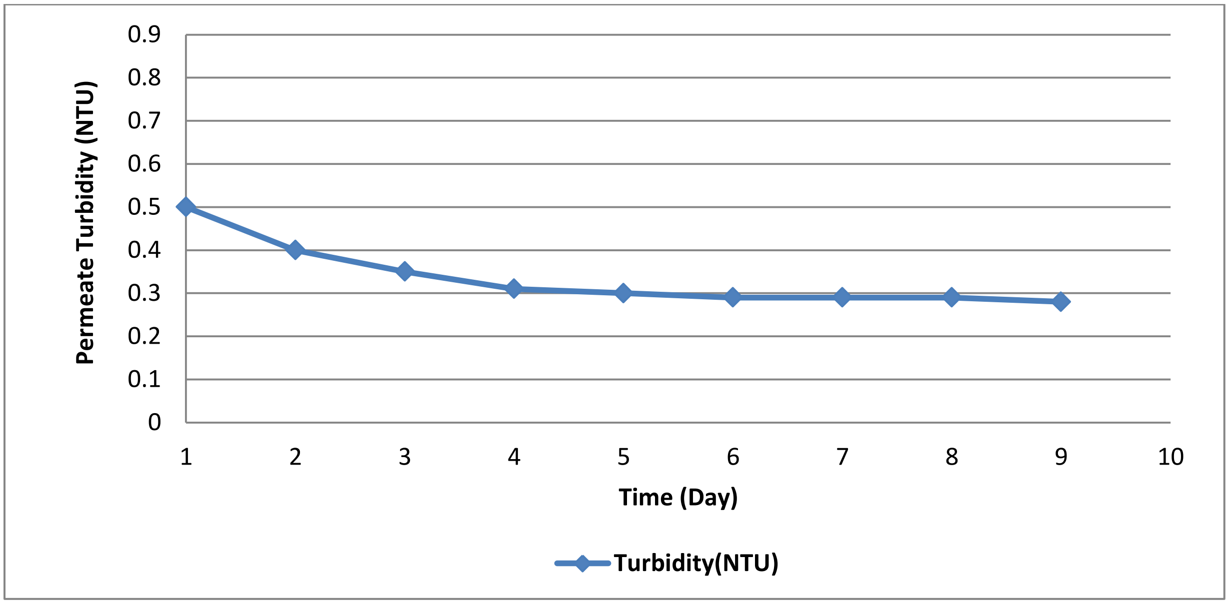

Figure 4.10: The relationship between permeate turbidity and time at fixed activated sludge concentration

The permeate turbidity was found to be less than 1 NTU and it was decreasing with the increase in time and eventually stabilizes from day 4 up to day 9 (refer to figure 4.9).

According to table 2.15 in literature, permeates turbidity was low and was therefore ignored. This is a common trend in IMBR performance (refer to section 2.4). According to Pillay and Kalu (2011), the RRWTS system treated the effluent with turbidities ranging from 20 to 300 NTU to below 1 NTU. Therefore this trend was expected and made sense in terms of quality performance of the WFM-IMBR.

Turbidity rejection fitted well with the criteria chosen for WFM-IMBR evaluation since less than 1 NTU was achieved at all varying points.

This was also in agreement with IMBR literature since generally IMBR in literature achieves less than 1 NTU of turbidity (Pillay and Kalu, 2011; Pillay and Dlamini 2010; Pillay and Jacobs, 2008 and Judd, 2006). Refer to section 2.6.3 for typical performance of WF membranes in a potable water treatment. However, this was a first study to be conducted in wastewater treatment using WF membranes.

4.4 Summary of results for chapter 4

Critical flux was used as a parameter to measure WFM-IMBR hydrodynamics. It was observed that the critical flux decreased with the increase of sludge concentration. It was also observed that critical flux increases with an increase in aeration.

Thus, the operating point of 30 LMH at 12 g/L sludge concentration and an aeration of 10 L/min/module was chosen as the operating point for short term subcritical operation of WFM-IMBR system.

Short term Subcritical operation proved to be stable. This demonstrated that WFM-IMBR can be operated at subcritical conditions for extended periods in an experimentation of 8 days.

The membrane performance in terms of water quality was monitored in this section using three parameters, namely, chemical oxygen demand (COD) removal, suspended solids removal, and turbidity.

MLSS removal was found to be stable for 6 days at 100 % removal of solids. COD removal was varying between 95% and 98% for 9 days. The permeate turbidity was found to be less than 1 NTU and it was decreasing with an increase in time and eventually stabilised in day 4.

The WFM-IMBR data generated in section 4.3.4 was compared to the performance of established small scale IMBRs (refer to the summary of established small scale IMBR in appendix F) and comparison showed that WFM-IMBR has more or less similar performance in terms of effluent quality and hydrodynamics. However there is still a lot to be done to develop WFM-IMBR in terms of energy reduction strategies and understanding of microbial activities in wastewater and its link to fouling.

CHAPTER 5: PILOT SCALE LONG TERM PERFORMANCE EVALUATION OF WFM-IMBR

5.1 Introduction

In chapter 3 good hydrodynamics of WFM-IMBR on waste sludge was demonstrated. The sludge used was seeded to the desired concentration by means of mixing the raw feed with the return activated sludge from the return line. In this chapter, sludge is allowed to grow into the desired concentration. Optimum conditions for sludge growth were provided in order to keep microbial organisms active and alive.

The purpose of this WFM-IMBR pilot plant study was to assess the performance of the process over a long time period to produce a high quality effluent, targeting good turbidity removal, good removal of MLSS and COD in particular. High COD removal, MLSS removal and Turbidity are important to this pilot study because the purpose of the MBR is to reclaim wastewater for reuse by replacing the clarifier with the MBR.

The specific objectives were: (i) To evaluate the performance of WFM-IMBR for domestic wastewater treatment using a well active microbial activated sludge, that was harvested from the aeration zones of the activated sludge process (ii) To determine the long term operation of WFM-IMBR prior to membrane cleaning and (iii) To optimize the WFM-IMBR hydrodynamic parameters of the system.

5.2 Criteria for good IMBR operations

The current commercial flat sheet IMBR achieved turbidity of less than 1 NTU, 99% COD removal, 100% MLSS removal and 4 log bacteria/virus removals. IMBRs are not limited by sludge, however, most IMBRs operate at 12 g/L. Commercial IMBRs operate at 30 MLH and most IMBRs operate at a range of 10 L/min/module to 50 L/min/module aeration rate (Pillay and Jacobs, 2005).

Hence, the criteria adopted for good operations of IMBR in this project were MLSS removal, COD removal, turbidity reduction and long term stability operation at 30 LMH.

5.3 Equipment set up and methodology

The experimental setup used in this chapter is identical to that used in chapter 4 and depicted in Figure 4.2. This section is the extension of the previous chapter.

The equipment set up showing a control panel is presented in Figure 5.1. Four air rotameters, two water flow rotameters, three pressure gauges, two pumps and one blower is presented. Three operating tanks are shown with lines showing the direction of air, raw feed and permeate.

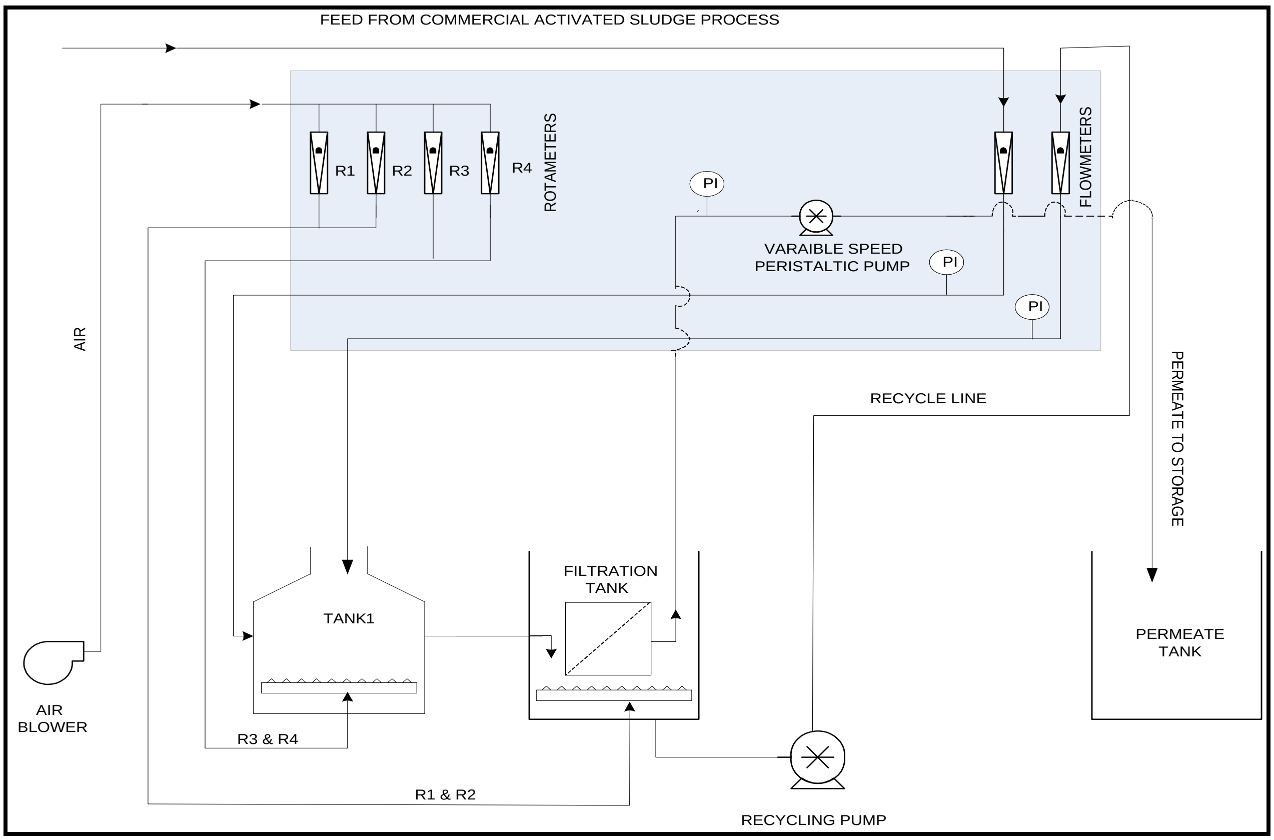

Figure 5.1: Schematic representation of pilot scale WFM-IMBR system equipment set up and control panels

A photograph of the equipment set up is shown below in figure 5.2 with the control panel.

Figure 5.2: Pilot scale WFM-IMBR system equipment set up and its control panels at experimental site

The feed solution

The real raw, pretreated (harvested prior to discharge into aeration zones) domestic wastewater was used. It was mixed with activated sludge suspension harvested from aerobic reactors. After the desired concentration was achieved, primary treated raw feed was supplied continuously to the WFM-IMBR system supplementing the filtrate. The excess sludge was discharged on daily basis based on the desired concentration for the WFM-IMBR system required.

The particle size distribution of activated sludge obtained is shown in Figure 4.3. From Figure 4.3, it can be determined that the mean particle diameter of activated sludge is about 100 microns. Particle size distribution of aerated waste activated sludge was measured with a Malvern 2000 Mastersizer. The nominal membrane pore size is estimated to be 1 to 2 µm.

5.4 Experimental procedure

- The feed (return activated sludge and primary pretreated domestic feed) was prepared to a desired concentration by growing the culture feed (mixture of return activated sludge and primary pretreated domestic feed) in the presence of oxygen.

- The activated sludge was aerated at 60 L/min exceeding the required dissolved oxygen concentration of 2 mg/L in all biological tanks and filtration tanks which is required by microbial organisms for as a source of oxygen.
- No chemical enhancer was used since the treatment of domestic wastewater was achieved by biological means and filtration.
- The system was operated at 10 LMH and it was increased by 5 LMH increments up until 30 LMH was achieved. In so doing the sludge was allowed to grow up to 12 g/L from 3.8 g/L before 30 LMH was achieved.
- After the desired concentration and flux was achieved, the system was continuously operated below critical flux for the purpose of the objectives of this chapter (refer to section 5.1 last paragraph).
- The fresh feed was continuously supplied to compensate for the filtered permeate at 110 L/hr and the recycle was kept at 440 L/hr in order to provide perfect mixing and sludge suspension inside the tanks. This is a common practice for recycle to be four times the feed for good mixing.
- Dissolved oxygen was monitored at 12 hour intervals by 5320AD HACH method to ensure good support for respiratory of microbial organisms.
- The MBR was operated at a subcritical flux of 30 LMH for two months as per the operating point for sub-critical flux chosen in chapter 4 (refer to section 4.2.9).
- The permeate MLSS, COD and turbidity was monitored at 6 hour intervals by the HACH method 4560, 6970 and 6789 respectively in order to measure the quality performance of WFM-IMBR system. (Refer to Appendix G, H and I).

5.5 Three phases of pilot scale operation and MLSS build up

The pilot scale experimentation and monitoring program was developed to achieve the objectives of the long-term stability operation and performance evaluation of the WFM-IMBR. The program consisted of a start-up phase and was planned to have three operational phases.

The main goal of the entire experimentation was to achieve a COD < 60 mg/L and turbidity < 1 mg/L while measuring the amount of MLSS in the treated effluent. Dissolved oxygen was kept above 2 mg/L in the biological tanks.

The phases were as follows:

Phase I, the objective was to harvest sludge from aerobic reactors to pilot scale WFM-IMBR and grow it from 4 g/L to 12 g/L biologically while establishing the wasting technique of sludge after the desired concentration is achieved. For the first month, the pilot was operated in a modified batch mode in order to increase the MLSS concentration in the WFM-IMBR to the target of 12 g/L.

The operation of Phase I can be broken into three periods corresponding to changes in the pilot set system operation regime.

Period 1 was the startup of aeration and monitoring of dissolved oxygen.

Period 2 was the startup of filtration and monitoring of permeate quality.

The feed to the WFM-IMBR was harvested from aeration zones (see figure 5.2) while in figure 4.5 it was harvested from the return line of the activated sludge.

In period 3 the flux was increased from 10 LMH by 5 LMH increments corresponding to an increase in MLSS by 2 g/L increments up until the MLSS was 12 g/L and the flux was 30 LMH.

The Figure 5.3 shows the schematic representation of WFM-IMBR system. It also shows the oxidation zones where feed to the system was harvested. The difference between the previous figure 4.2 and 5.3 is that in this case the sludge was harvested from the aeration zones in aerobic reactor while figure 4.2 the return sludge was used.

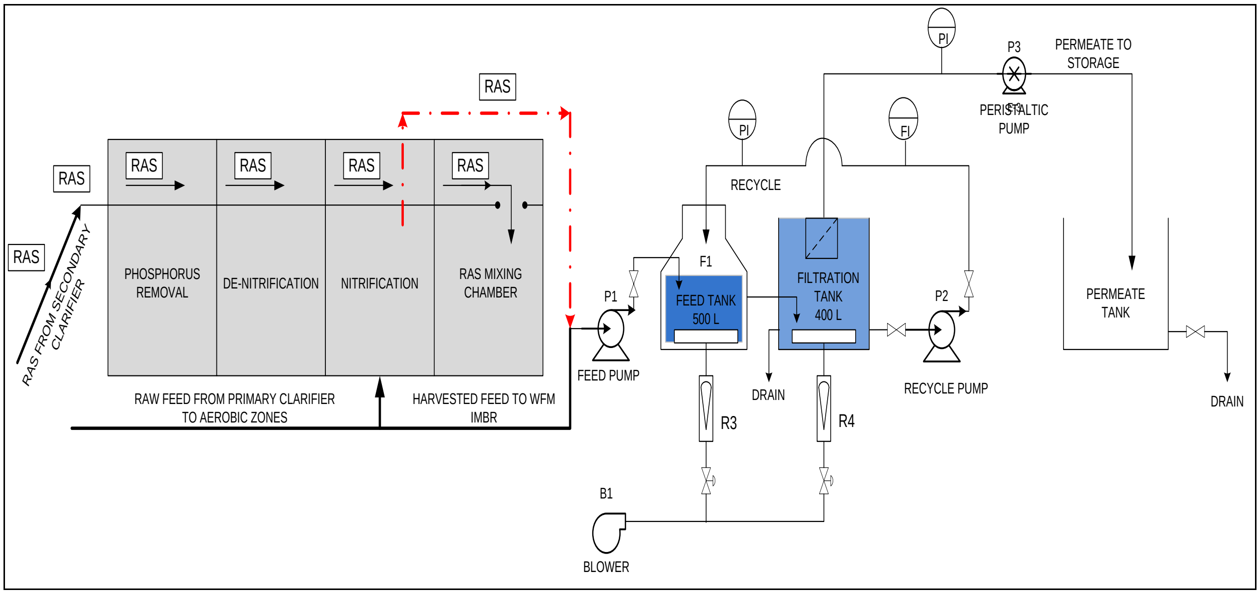

Figure 5.3: Schematic representation for the system employed in phase II pilot scale experimentation.

Phase II: the objective of phase II was to start up the WFM-IMBR including the filters while employing one biological aeration tank (aerobic tank) and filtration tank at subcritical conditions of 10 LMH and 4 g/L MLSS and increasing concentration and flux gradually up until 12 g/L and 30 LMH was achieved, respectively. The circulation was also activated in phase II to keep the sludge concentration uniform in the system.

The operation of Phase II can be broken into 2 periods corresponding to changes in the pilot set system operation regime.

Period 1, was the achievement of 12 g/L MLSS and startup of subcritical operation of WFM-IMBR pilot scale experimentation.
Period 2 was the commencement of super critical operation of WFM-IMBR after 30 days of subcritical operation of WFM-IMBR at 30 LMH and 12 g/L. The flux was increased in 5 LMH increments until there was extensive fouling of the membrane pack.

Phase III: the objective was to increase flux to super critical conditions and monitor the TMP while maintaining the sludge concentration at 12 g/L.

Membrane integrity

The membrane integrity and leak test was repeated. Refer to section 4.2.7 detailed membrane leak testing used in this study.

Membrane cleaning

Membrane modules were cleaned by brushing and rinsing with tap water. Chemicals can also be used especially for biofouling and irreversible fouling cleaning of the membrane pack. Chlorine solution was proposed to wash and soak modules in this study.

5.6 Results and discussion

5.6.1 Determination of the operating point

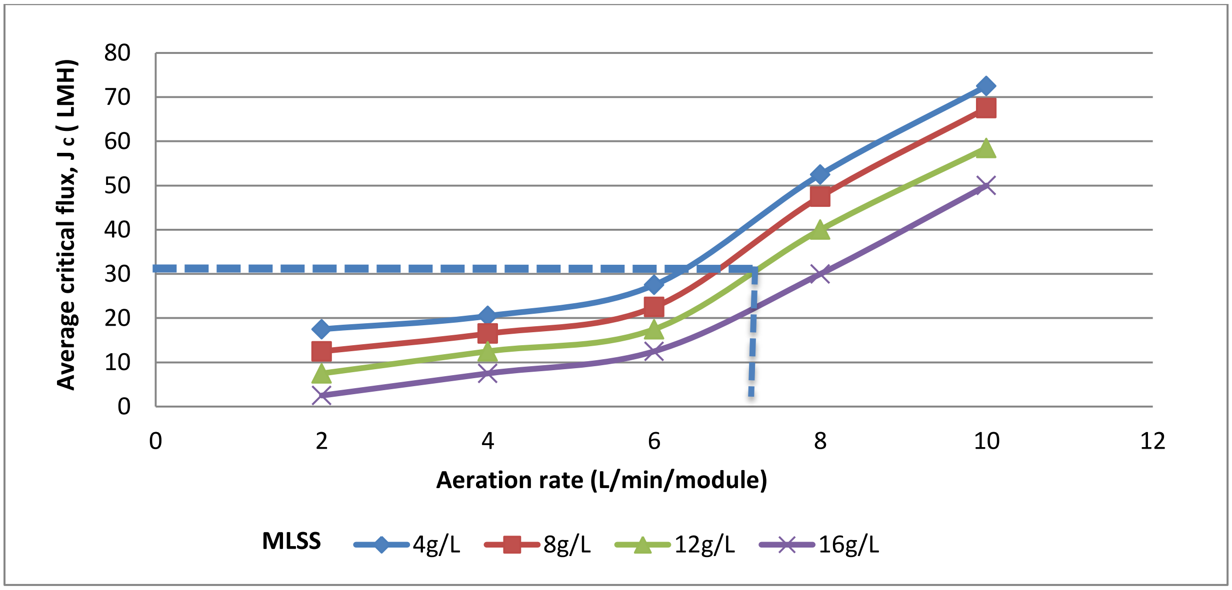

Figure 5.4: Critical flux characteristics at varying aeration and mixed liquor suspended solids

The critical flux of 30 LMH at 12 g/L sludge concentration and aeration of 6.8 L/min/module was found. However the aim of this study was to establish the subcritical flux conditions and operate at that point for extended periods. Therefore any value above 6.8 L/min/module was optimum long term subcritical operation of WFM-IMBR.

Thus, the operating point of 30 LMH at 12 g/L sludge concentration and aeration of 10 L/min/module was chosen as the operating point for short term subcritical operation of WFM-IMBR system (refer to figure 4.7) based on established existing IMBR units in developed economies.

This was done for subcritical flux operation purpose. This was lower than commercial small scale systems such as Kubota, Copa IMBR and Huber IMBR which utilizes aeration of 10 L/min/module,17.5 L/min/module and 150 L/min/module respectively.

The operating point of 30 LMH, 12 g/L MLSS and aeration of 10 L/min/module was chosen as the operating point (refer to figure 5.4).

5.6.2 Results for pilot scale experimentation

The experimentation results of long term operation and stability evaluation of WFM-IMBR system at subcritical conditions.

The TMP, MLSS and FLUX are shown in Figure 5.5.

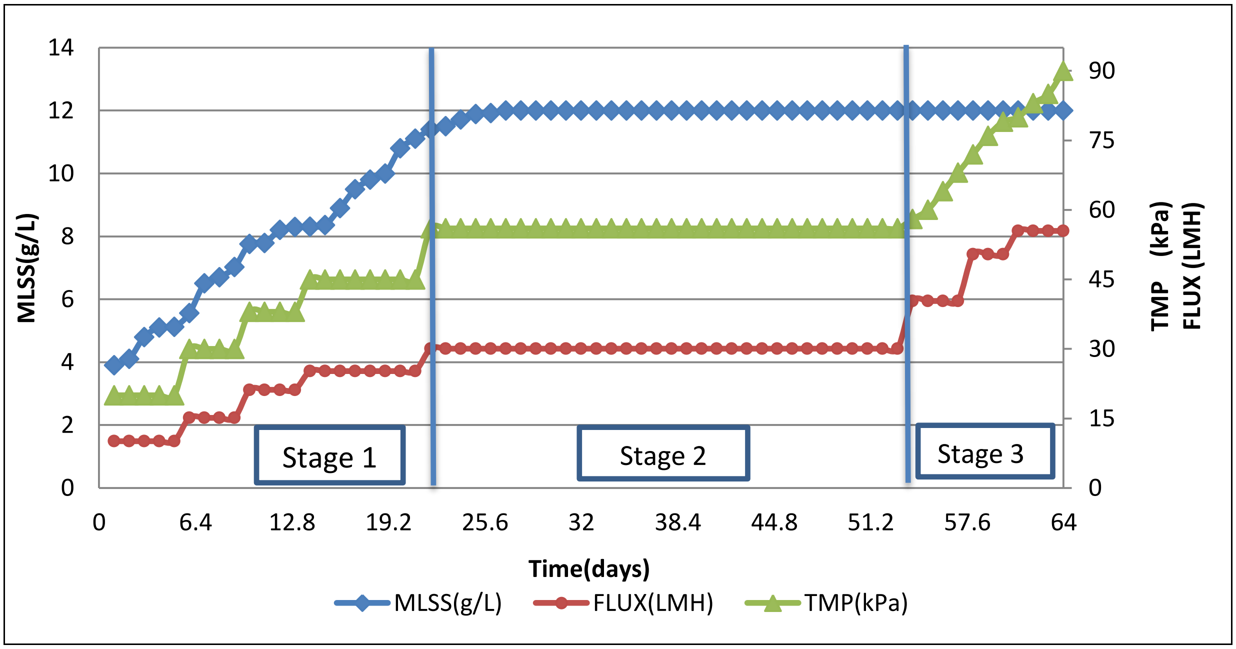

Figure 5.5: The relationship between Transmembrane Pressure and Time at fixed mixed liquor suspended solids of 12 g/L, 30LMH and 10L/min/module.

In stage 1, Flux was increased in 5 LMH increments up until the subcritical operation of 30 LMH at12 g/L and aeration of 10 L/min/module was reached.

In stage 2, the subcritical operation was started after 12 g/L of MLSS was achieved from day 21 to day 53.

The TMP was found to be stable for 30 days (from day 20 to day 50) and there was no variation observed in hydraulic performance (refer to Figure 5.5 stage 2).

In stage 3, super critical operation of WFM-IMBR was started after 32 days of subcritical operation

The flux was increased above the critical flux point and the TMP became unstable and increased exponentially (from day 53 to 63) up until there was a rapid hydraulic loss which was evident of fouling (refer to Figure 5.5 stage 3).

These trends were expected and they were in consistent with theory (refer to equation 2.3).These trends were also in agreement with literature according to Howell, Chua and Arnot, (2004).When operating below the critical flux, the TMP and permeate flux would always be stable and there will be no TMP variation (Howell, Chua and Arnot, 2004).

The moment the flux was increased above the critical flux, the TMP continued to increase with constant flux operation or the flux will continue to decrease with constant TMP operation (Howell, 1995). This has been visually confirmed by Li *et al.* (1998).

The trend responded positive to the criteria selected for this pilot scale trial. The first two stages were expected (refer to section 2.53), according to Ognier, Wisniewski and Grasmick, 2004 (b) the two distinct phases were observed when operating below and above critical flux, however, Ognier, Wisniewski and Grasmick, 2004 (b) did not run at super critical flux. This is the first study to be conducted where critical flux is exceeded to super critical conditions. All stages 1 and 2 were expected and in agreement with theory (refer to figure 2.19, according to Ognier, Wisniewski and Grasmick, 2004 (a) IMBRs can be operated for prolonged periods at subcritical conditions and above subcritical conditions fouling is observed).

If the operating flux is further increased, the super critical flux is achieved. It is safe to conclude that figure 5.8 is in agreement with theory (refer to section 2.5.3).

5.6.3 Quality Performance of WFM-IMBR using real activated sludge

The percent COD and MLSS removal at a varying loading rate is shown in Figure 5.6.

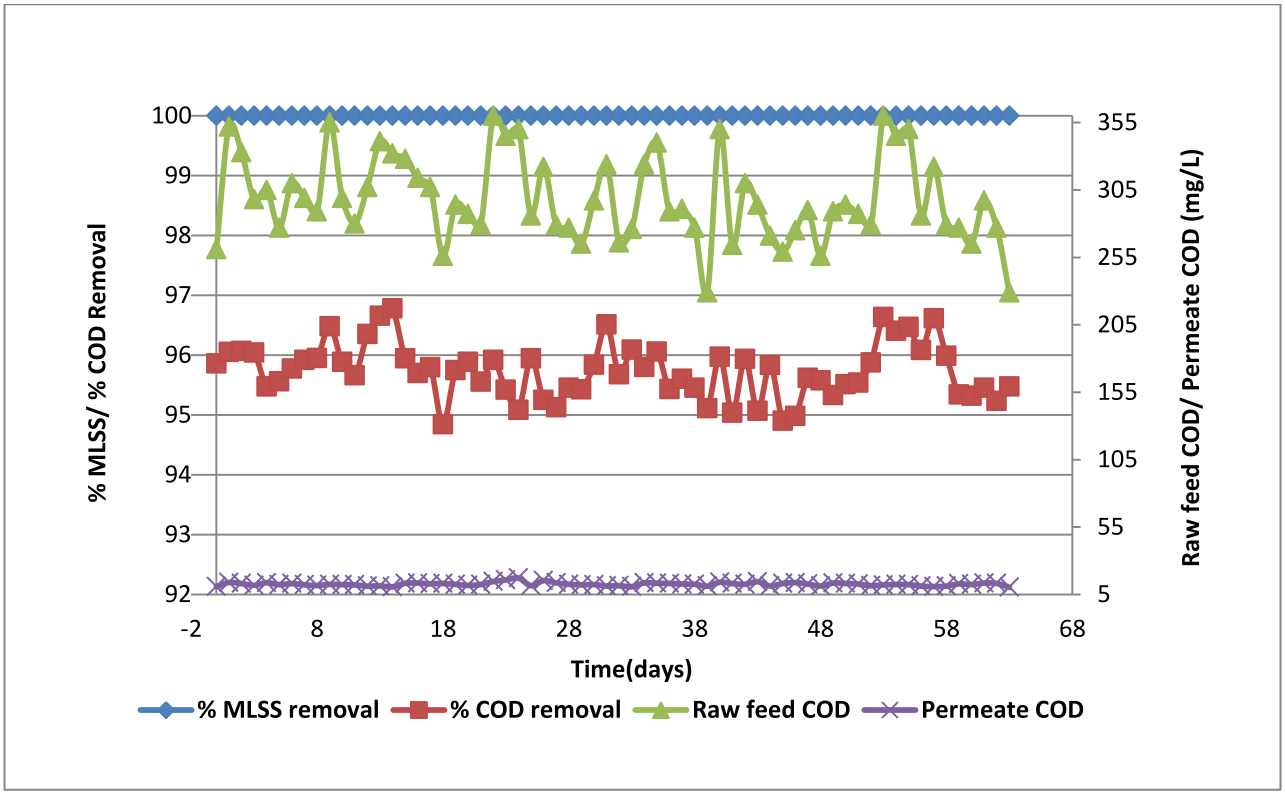

Figure 5.6: The relationship between %COD/MLSS removal and Time at varying sludge concentration.

MLSS removal was found to be stable for 30 days at 100 % removal of solids. COD removal was varying between 95% and 97% for 30 days.

According to Mudrack,(1985); Cui, Chang and Fane, (2003), Cicek *et al.* (1999), Gander, Jefferson and Judd, (2000) and Judd , (2006) IMBRs generally remove 100% of solids and achieve above 95% of COD removal provided there is no interference to the conditions of experimentation and therefore the data represented above in figure 5.6 is in agreement with literature.

The raw feed COD trend was expected since influent quality of domestic wastewater treatment is well documented (refer to table 2.12, according to Huang, Gui and Qian, (2001) the feed influent COD to the wastewater treatment plant ranges from 95 to 400 mg/L).

Permeate COD was found to be 55 mg/L at all given points , refer to table 2.10 for generic performance of IMBR in terms of permeate quality.

COD and MLSS removal fitted well with the criteria chosen for WFM-IMBR evaluation since it was in agreement with literature especially to that reported by the following authors: Cicek *et al.* (1999), Stephenson *et al.* (2000), Bouhabila, Ben Aïm and Buisson, (2001), Rosenberger and Kraume 2002(a), Soriano *et al.* (2003) and Gander, Jefferson and Judd, (2000).

Percentage COD removal and permeate COD quality fitted well with the criteria chosen for WFM-IMBR evaluation since it has the equivalent performance as that of established small scale IMBRs (refer to table F1.2). Therefore WF membranes proved to be competitive in wastewater treatment and the WFM-IMBR demonstrated that it can be operated for extended periods at subcritical conditions.

5.6.4 Solid rejection performance by WFM-IMBR

The permeate turbidity reduction against time for long term pilot scale experimentation is depicted in Figure 5.7.

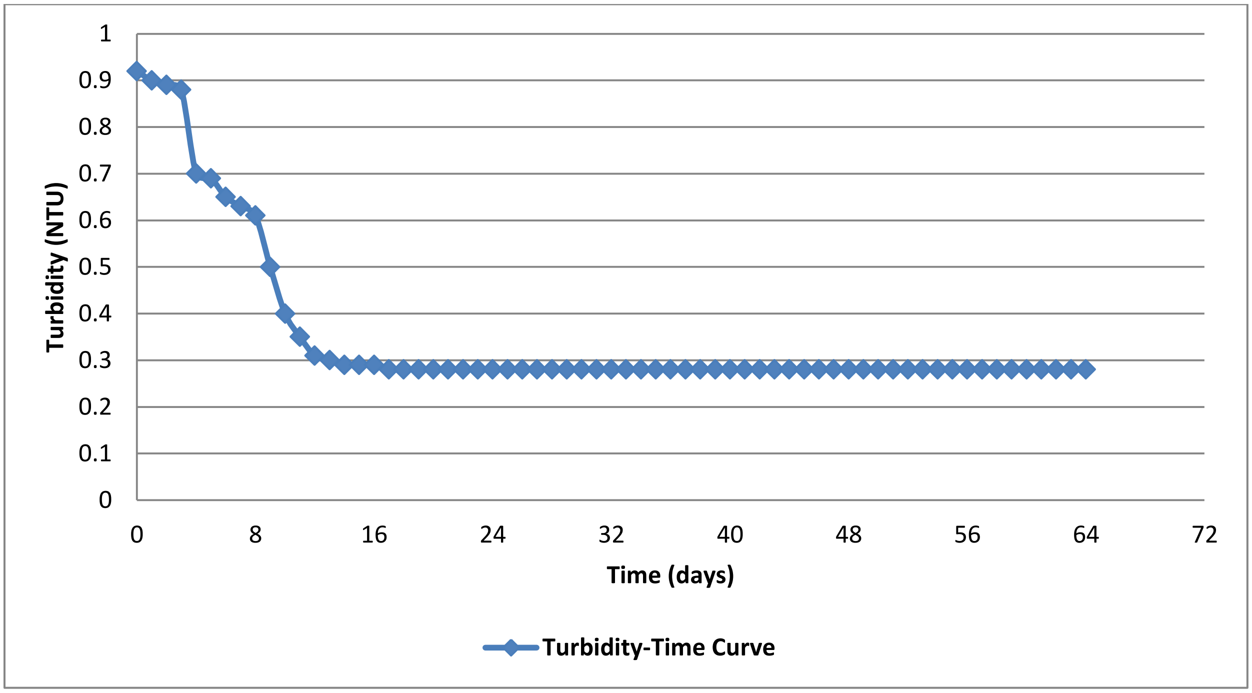

Figure 5.7: The relationship between permeate Turbidity (NTU) and Time (days) for long term pilot scale experimentation at 16 g/L loading.

The permeate turbidity was found to be less than 1 NTU and it was decreasing with an increase in time and eventually stabilizes from day 32 up to day 64 (refer to Figure 5.7).

According to Table 2.15, permeate turbidity in most IMBRs was very low. This is a common trend in IMBR performance (refer to section 2.4 in literature review). According to Pillay and Kalu (2011) the RRWTS used the same WF membranes used in this study for potable water treatment application, the RRWTS system treated effluent with turbidities ranging from 20 to 300 NTU to below 1 NTU. Therefore this trend was expected and made sense in terms of quality performance of the WFM-IMBR.

This was also in agreement with IMBR literature since generally IMBR in literature achieves less than 1 NTU of Turbidity (Pillay and Kalu, 2011; Pillay and Dlamini 2010; Pillay and Jacobs, 2008 and Judd, 2006).

WF membranes have been applied to potable water treatment with success (Refer to section 2.6.3 for the typical performance of WF membranes in a potable water treatment). However this was a first study to be conducted in wastewater treatment using WF membranes.

Turbidity rejection fitted well with the criteria chosen for WFM-IMBR evaluation since it had the same performance as that of established small scale IMBRs (refer to table F1.2). Therefore WF membranes proved to be competitive in wastewater treatment and WFM-IMBR demonstrated that it can be operated for extended periods at subcritical conditions.

5.7 Summary of results

WFM- IMBR Pilot Scale performance

The pilot scale sub-critical flux operation trials demonstrated that the WFM-IMBR can be operated for prolonged time with neither hydraulic loss nor need of membrane cleaning.

The TMP was found to be stable for 30 days (from day 20 to day 50) and there was no variation observed in hydraulic performance (refer to Figure 5.8 stage 2). This showed that the WFM-IMBR can be operated for an extended period at subcritical flux operation.

A pilot scale WFM-IMBR trial also demonstrated that by operating the WFM-IMBR at supercritical conditions, TMP becomes unstable and increases exponentially up until there is a rapid hydraulic loss which is evident of high fouling.

The performance of the WFM-IMBR pilot scale system was found to exceed the design expectation; COD and MLSS removal was found to be 95% and 100%, respectively, and was found to be stable with time. The permeate turbidity was found to be less than 1 NTU and decreasing with an increase in time and eventually stabilised after a prolonged time, from day 32 up to the end of experimentation. This was in line with the objectives and criteria chosen for these investigations.

CHAPTER 6: CONCLUSIONS

The experimental investigations conducted to evaluate the performance and long term stability of WFM-IMBR for treatment of domestic wastewater for reuse provided the following conclusions:

The WF membrane pack proved to have a great potential in membrane technology; it has excellent COD and MLSS removal and acceptable permeate turbidity. The membrane is robust and it is easy to clean.

Since the operational parameters employed in this study were the same as that of small scale commercial established IMBR operation (refer to appendix F for summary of established small scale IMBRs), the WFM-IMBR proved that it has a great potential for application in domestic wastewater treatment for reuse. Good quality performance output and long-term stability operation demonstrated that the WFM-IMBR can be operated for extended periods at subcritical conditions.

Long term subcritical operation proved to be stable for 30 days and there was no variation observed in hydraulic performance.

The operating point of flux of 30 LMH at 12 g/L sludge concentration and aeration of 10 L/min/module chosen as the operating point for long term subcritical operation of the WFM-IMBR system in chapter 4 met the criteria chosen in this study with success.

The TMP was found to be stable for 30 days and there was no variation observed in hydraulic performance. This showed that the IMBR can be operated for extended periods at subcritical flux operation conditions.

The performance of the WFM-IMBR, in terms of quality was found to exceed the expectation with more than 95% COD removal, 100% MLSS removal and permeate turbidity less than 1 NTU at any given time.

CHAPTER 7: RECOMMENDATIONS

The following studies are recommended for further development of WFM-IMBR unit:

Since this was the first study to be conducted by using WF membranes in IMBRs it is recommended to perform energy reduction strategies of the system.

To conduct full scale investigations on another WWT plant in order to verify the pilot-scale results.

Considering its robust material strength, new application fields, such as anaerobic treatment or high strength industrial wastewater treatment should be tested on this WFM-IMBR.

Investigation for the behavior of micro-organisms at varying aeration shear intensity in WFM-IMBRs can be of interest for the understanding of interaction between the bacteria and membrane for proper optimization and control in future.

A further study is recommended to investigate the effect of backwashing of WFM-IMBR for its performance restoration. Backwashing of WFM-IMBR specification like intensity, duration and cycle can be of benefit for achieving the extended period of membrane filtration cycles and maintaining high fluxes of permeate.

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APPENDIX

APPENDIX A: BUBBLE SIZE DETERMINATION AT A VARYING PORE SIZE DIFFUSER

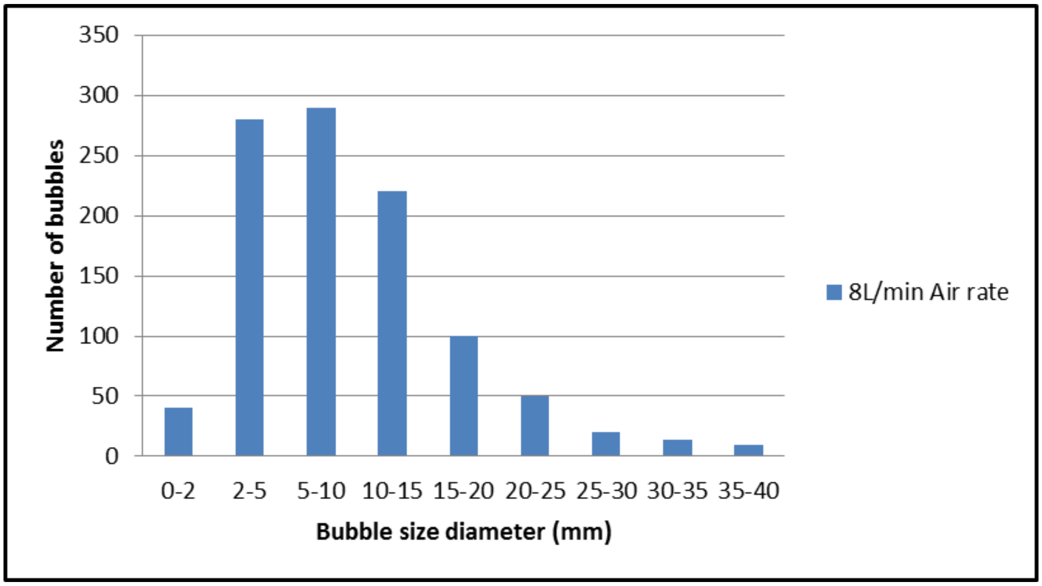

Figure A1.1: The relationship between number of bubbles and bubble size diameter for 1 mm diameter nozzle

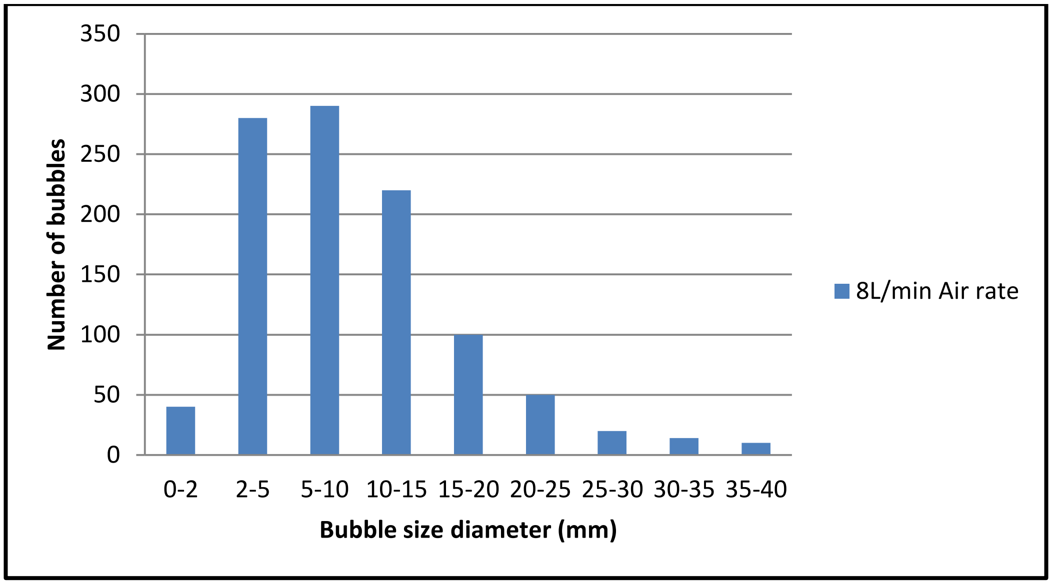

Figure A1.2: The relationship between number of bubbles and bubble size diameter for 1 mm diameter nozzle

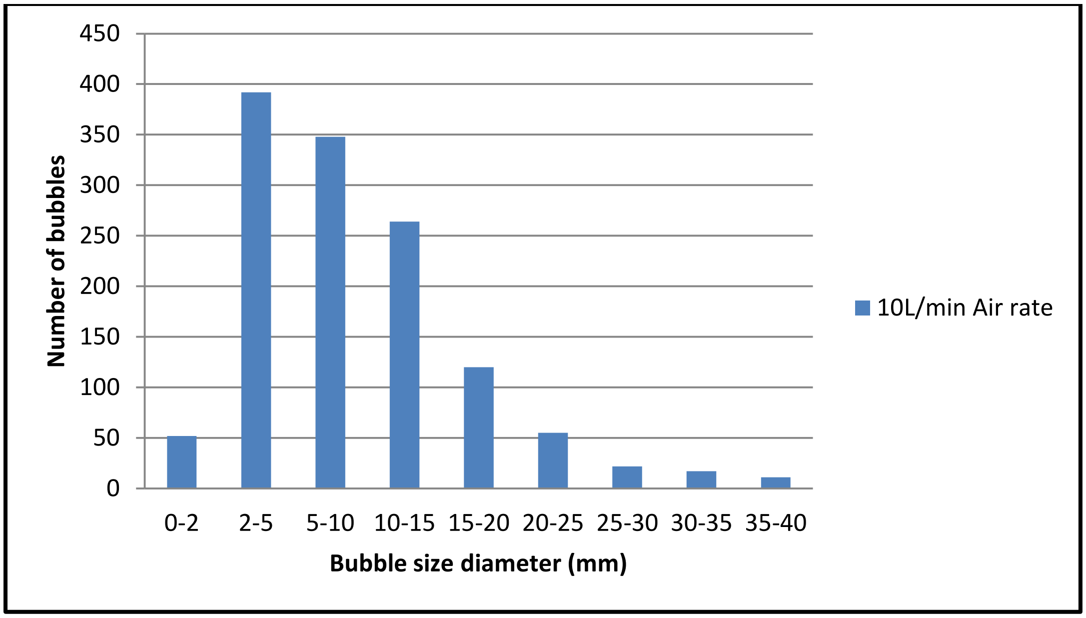

Figure A1.3: The relationship between number of bubbles and bubble size diameter for 1 mm diameter nozzle

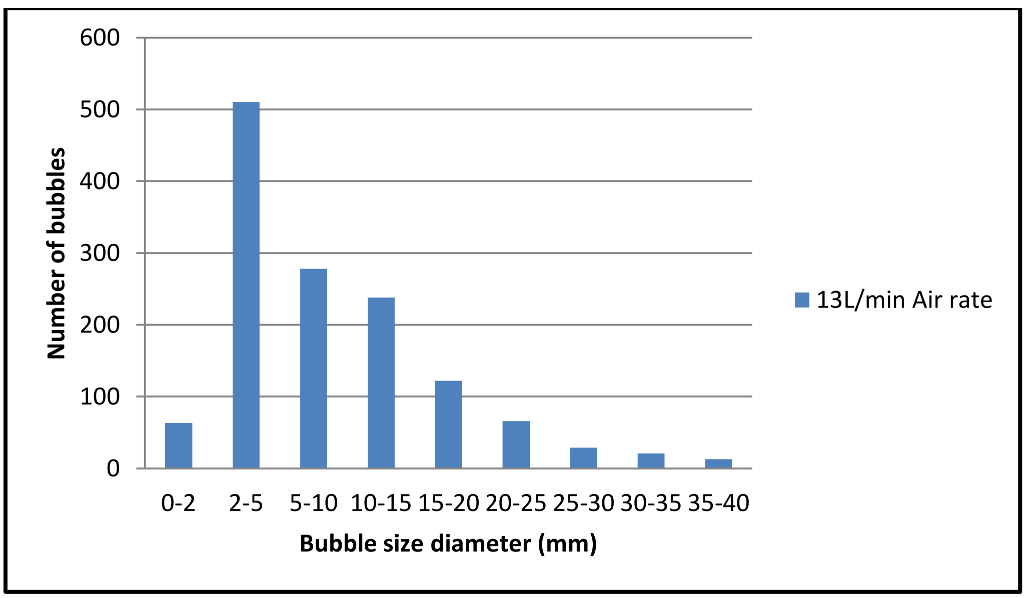

Figure A1.4: The relationship between number of bubbles and bubble size diameter for 1 mm diameter nozzle

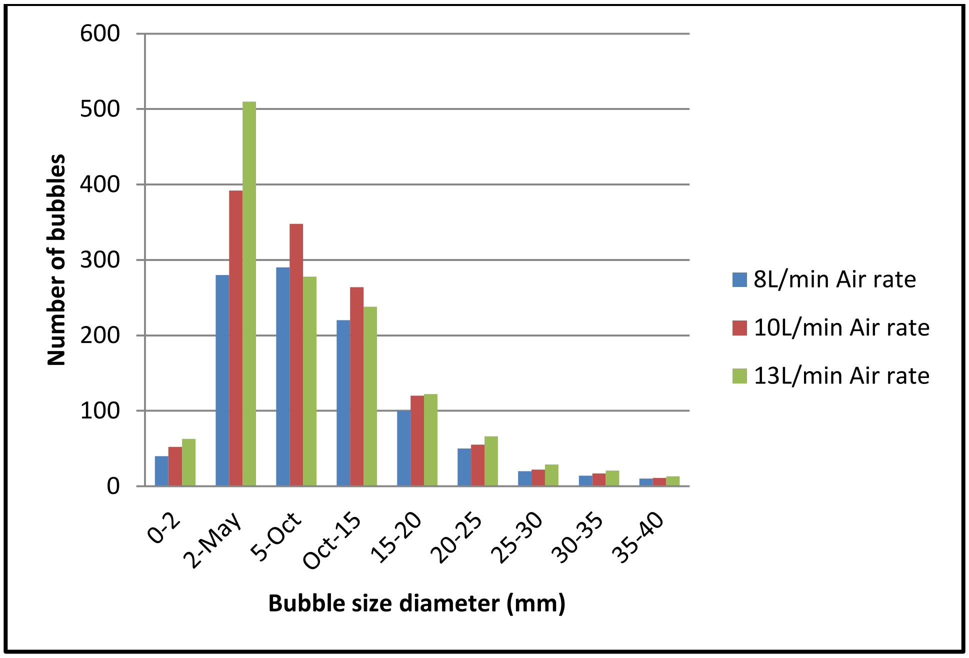

Figure A1.5: The relationship between number of bubbles and bubble size diameter for 1 mm diameter nozzle

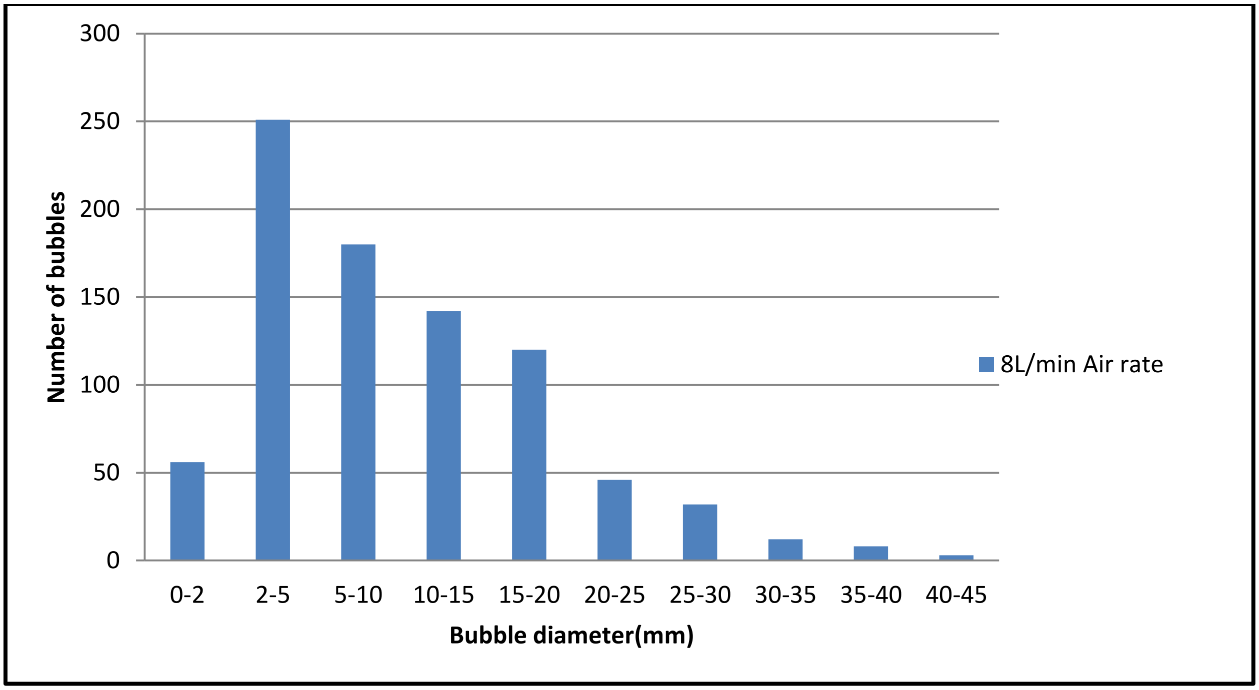

Figure A1.6: The relationship between number of bubbles and bubble size diameter for 1.5 mm diameter nozzle

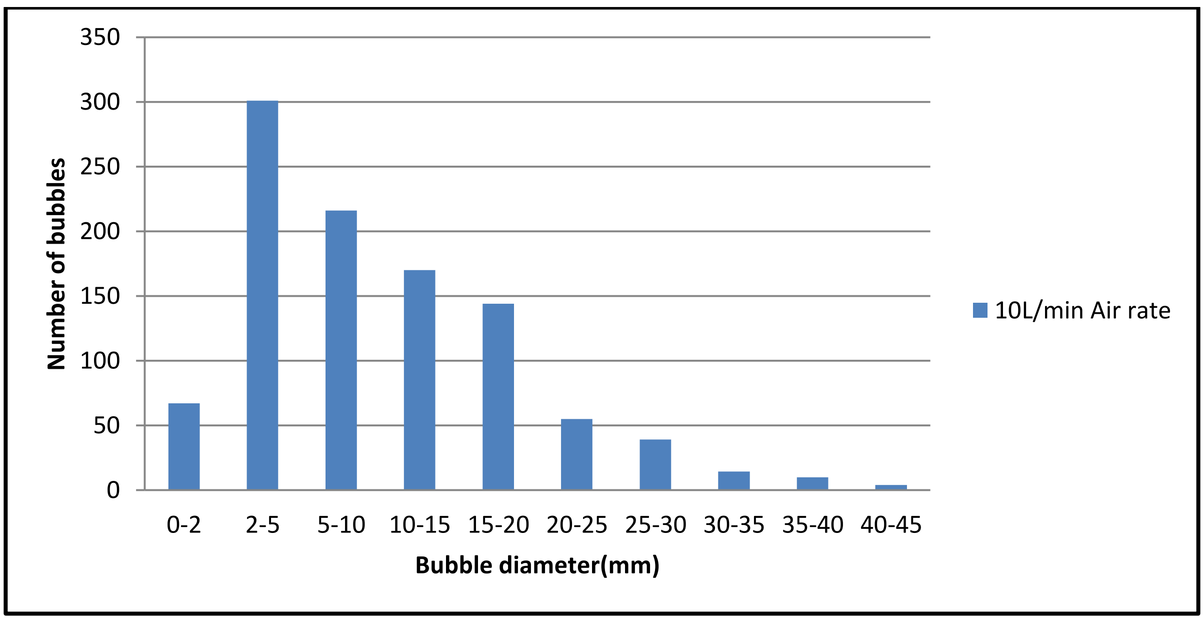

Figure A1.7: The relationship between number of bubbles and bubble size diameter for 1.5 mm diameter nozzle

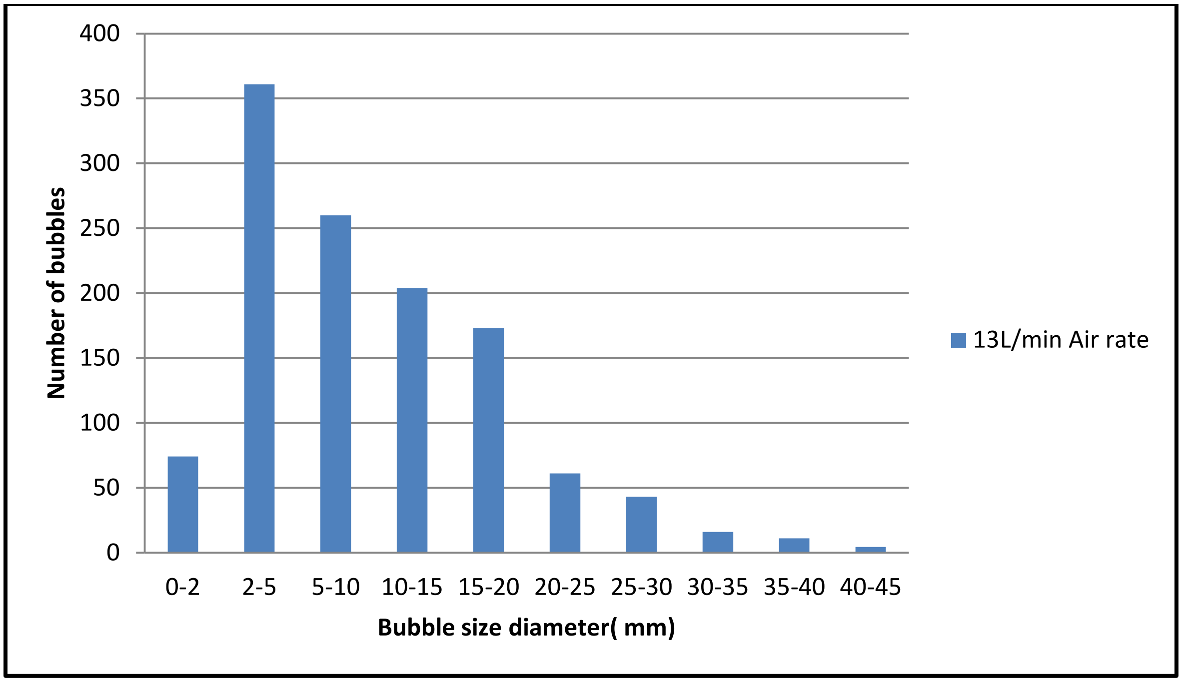

Figure A1.8: The relationship between number of bubbles and bubble size diameter for 1.5 mm diameter nozzle

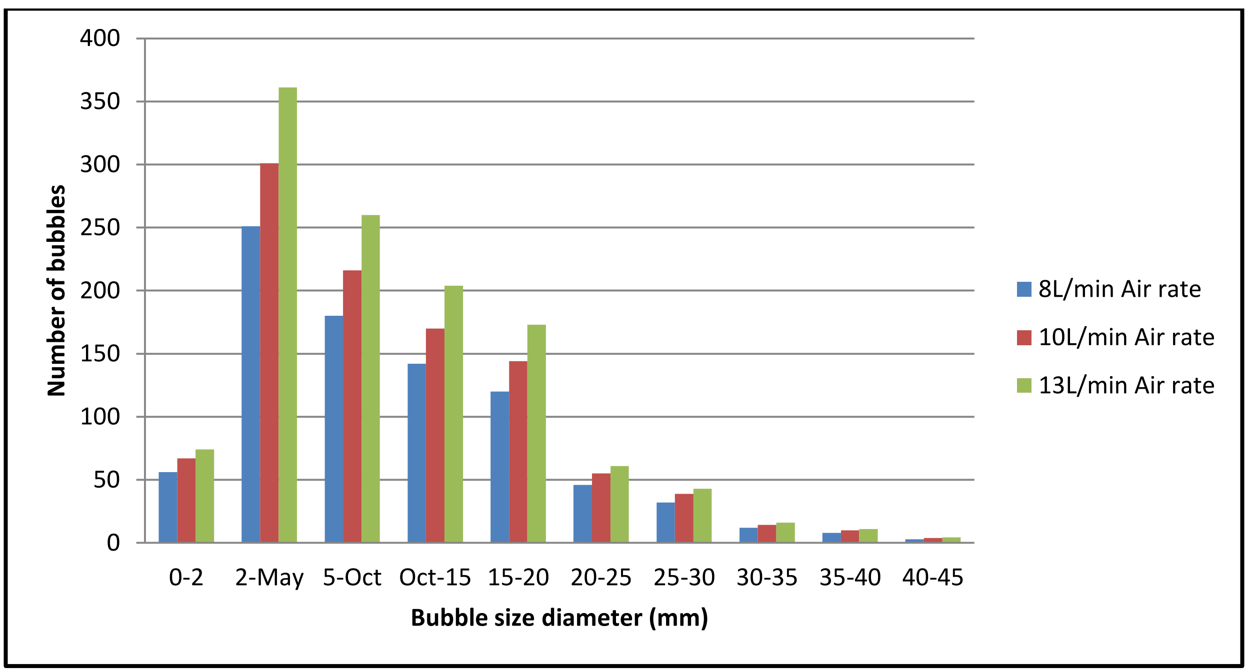

Figure A1. 9: The relationship between number of bubbles and bubble size diameter for 1.5 mm diameter nozzle

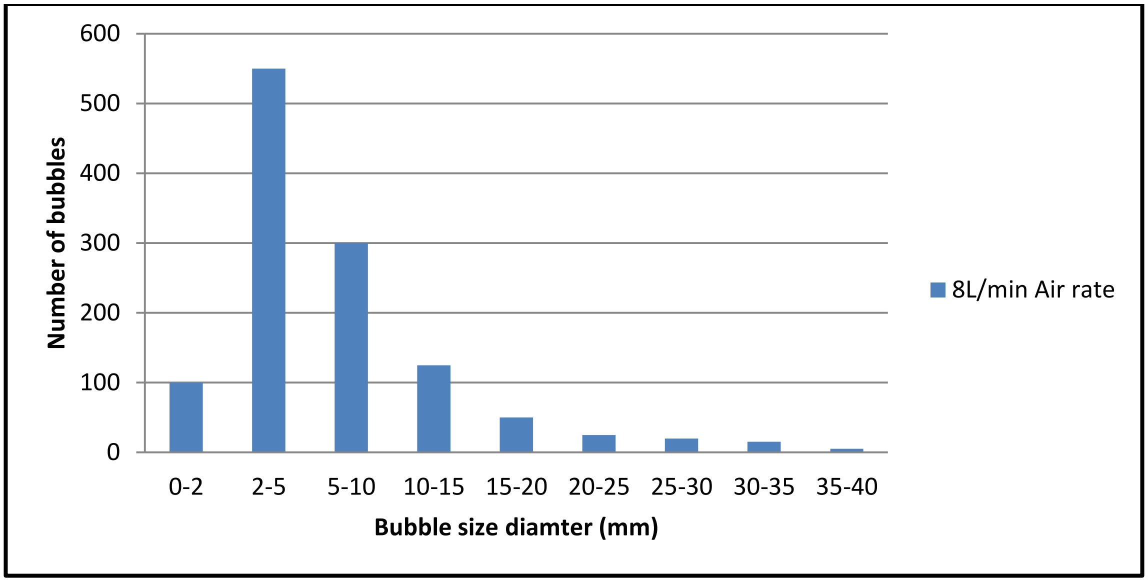

Figure A1.10: The relationship between number of bubbles and bubble size diameter for 2mm diameter nozzle

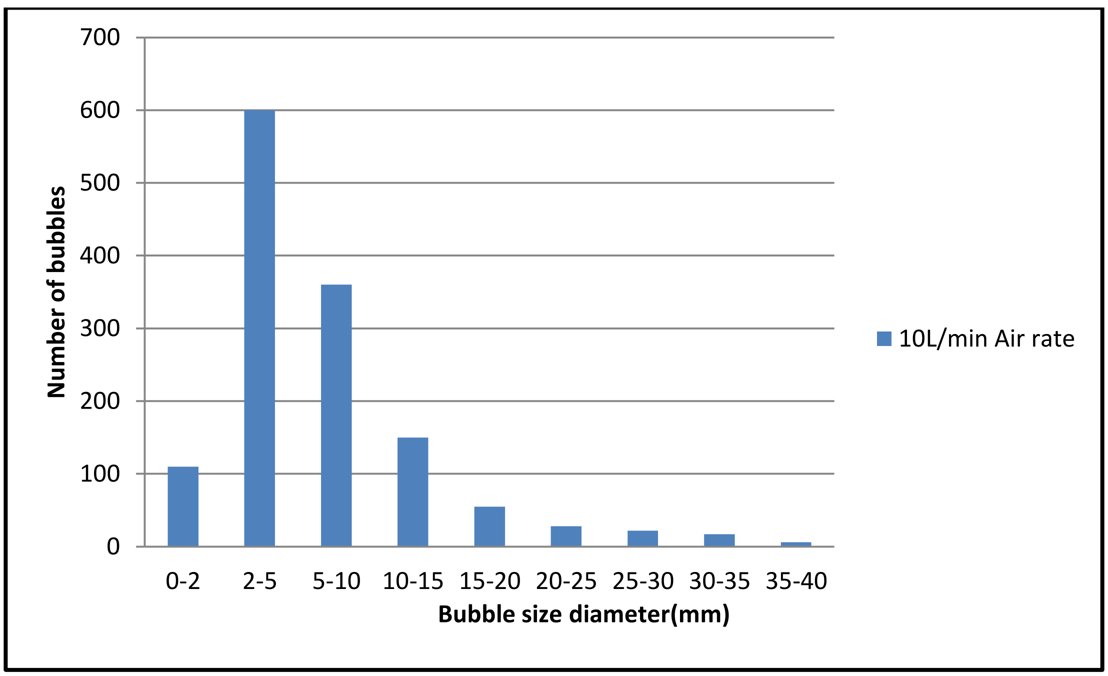

Figure A1.11: The relationship between number of bubbles and bubble size diameter for 2mm diameter nozzle

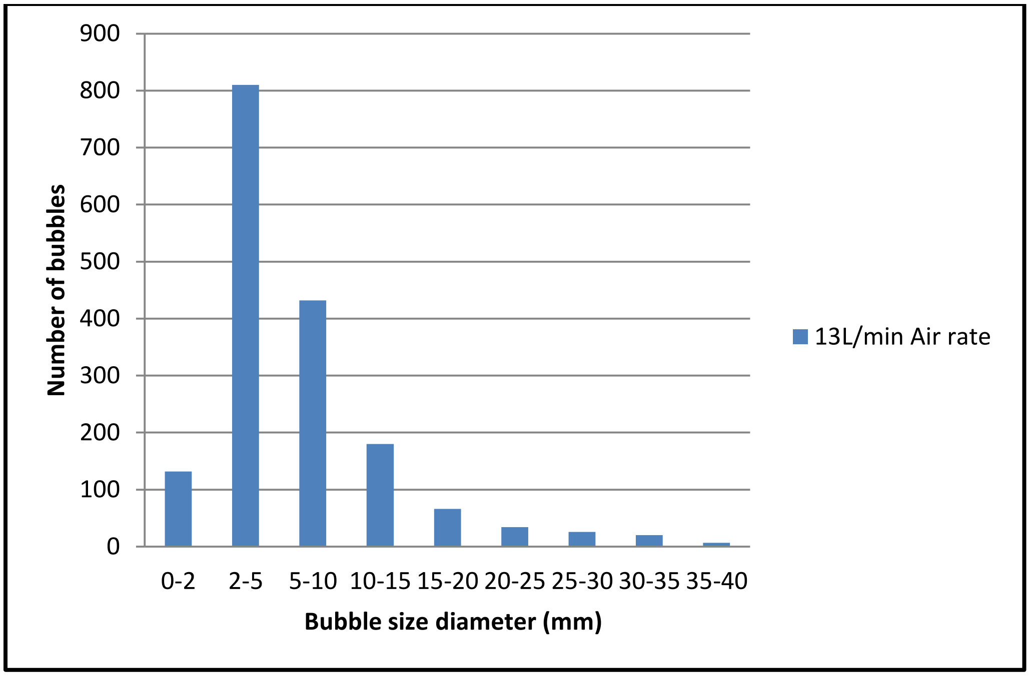

Figure A1.12: The relationship between number of bubbles and bubble size diameter for 2mm diameter nozzle

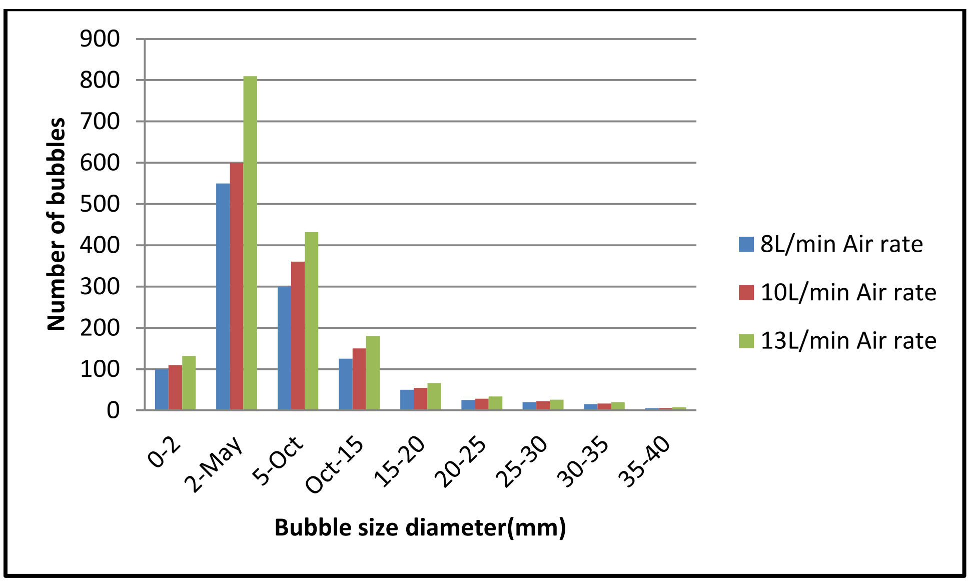

Figure A1.13: The relationship between number of bubbles and bubble size diameter for 2mm diameter nozzle

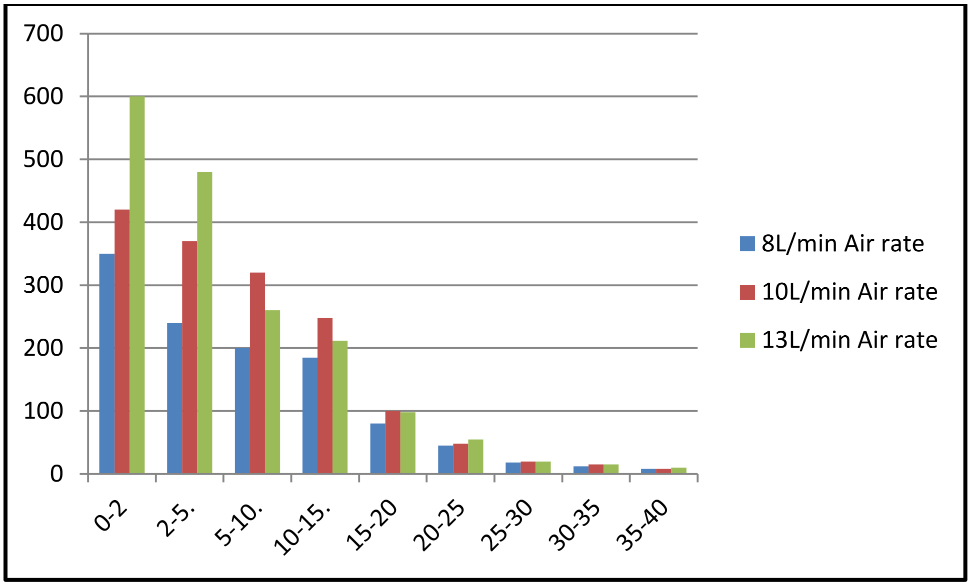

Figure A1.14: Comparisons of round and square nozzles at varying aeration rate

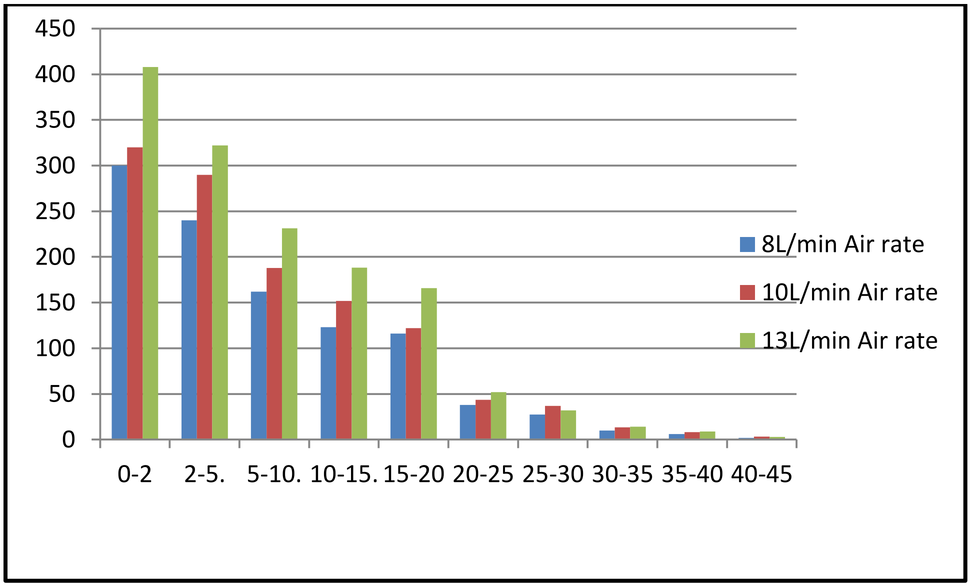

Figure A1.15: Comparisons of round and square nozzles at varying aeration rate

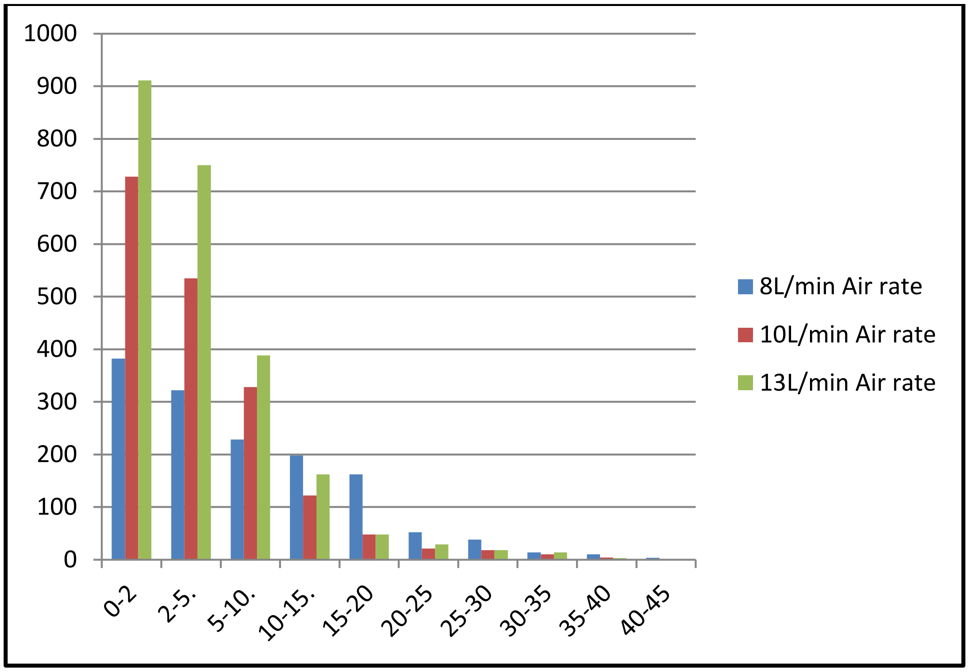

Figure A1.16: Comparisons of round and square nozzles at varying aeration rate

APPENDIX B: CRITICAL FLUX EVALUATION METHOD AND PROCEDURE

WFM-IMBR CRITAL FLUX EVALUATION BY STEP METHOD AND DERIVATION AVERAGE CRITICAL FLUX - AERATION RATE CURVE

DETAILED EXPIRIMINTATION

SET 1:

The main aim is to determine the TMP-Aeration curve, average criticalflux-aeration curve without any flux enhancement strategy

1 **INVESTIGATION II (4DAYS)**

1.1 **Objective**

Determine the critical flux for 2, 4, 6,8,10, L/min/module aeration flow rate

1.2 **Equipment required**

- **i.** 1x basic pilot scale unit of WFM-IMBRsystem
- **ii.** Activated sludge to make up moderate turbidity feed
- **iii.** Measuring cylinder
- **iv.** Stop watch

1.3 **Experimental procedure**

- **i.** Make up high turbidity, high fouling feed
- **ii.** Fill the tank with high fouling feed
- **iii.** Start the aeration in the system, aerate at 2L/min/module
- **iv.** Vary the flux in 5LMH increments and record the TMP at 30 minutes interval up until there is increase in TMP reading
- **v.** Clean the membrane thoroughly using brushing and excess tap water
- **vi.** Repeat the (iii) to (v) to get at least 3 repeatable TMP-TIME curves
- **vii.** Change the aeration at 2L/min/module interval up to 10L/min/module and repeat (iii) to (vi) at each and every aeration used

1.4 Reports

TMP –TIME curves corresponding to air flow rate used (21 curves)

2 INVESTIGATION III (1 DAY)

2.1 Objective

Derivation of the average critical flux –aeration curve

2.2 Method

a) Take the average of lower and upper flux before onset of critical flux

b) Prepare the excel spreadsheet of average critical fluxes and aeration air flow rate

2.3 Reports

Plot average critical flux against corresponding aeration flow rate(x3)

APPENDIX C: CRITICAL FLUX EVALUATION AT VARYING AERATION RATE AND VARYING SLUDGE CONCENTRATION

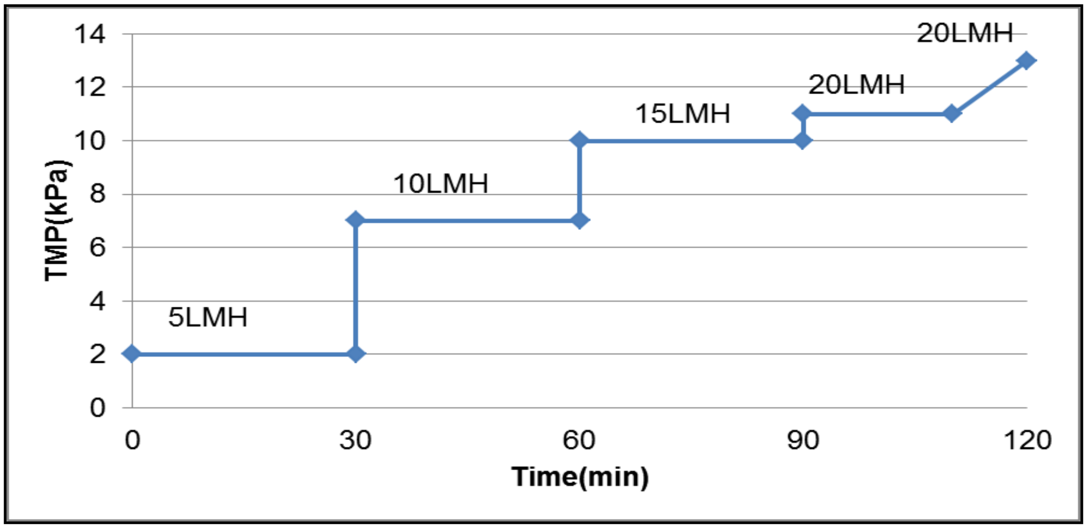

Figure C1.1: Critical flux determination at the aeration of 5L/min/module and 4g/Lsludge concentration

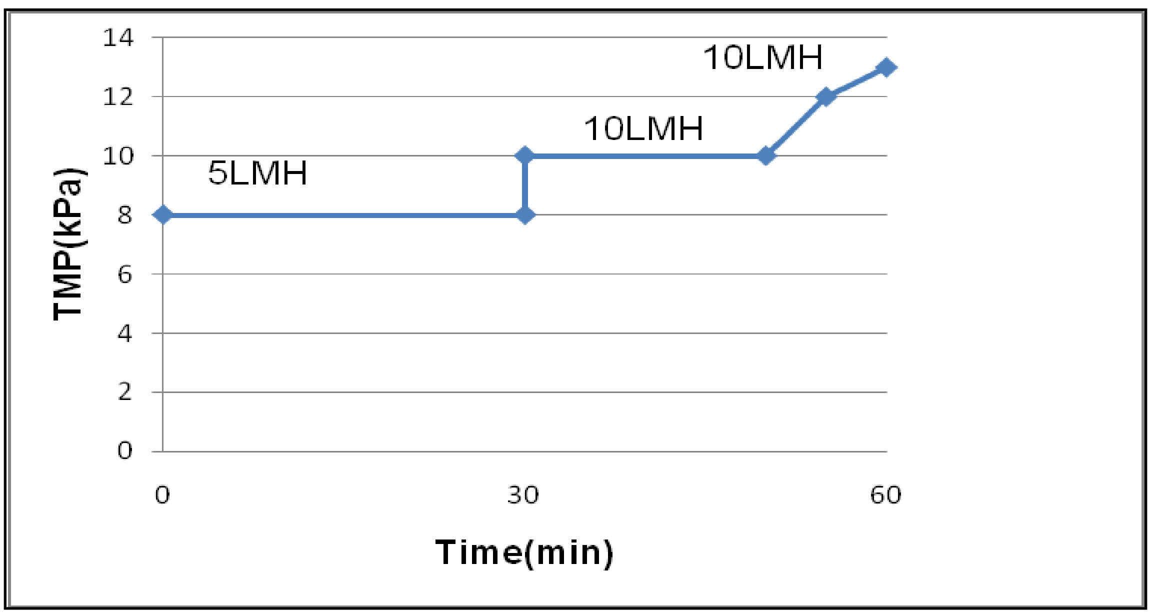

Figure C1.2: Critical flux determination at the aeration of 5L/min/module and 8g/Lsludge concentration

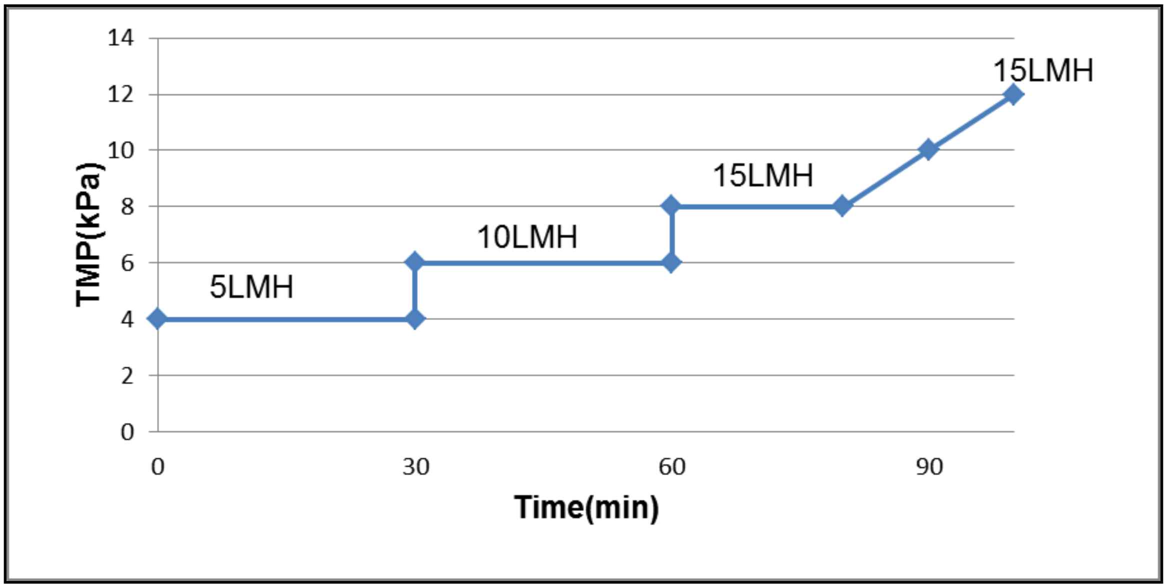

Figure C1.3: Critical flux determination at the aeration of 5L/min/module and 12g/Lsludge concentration

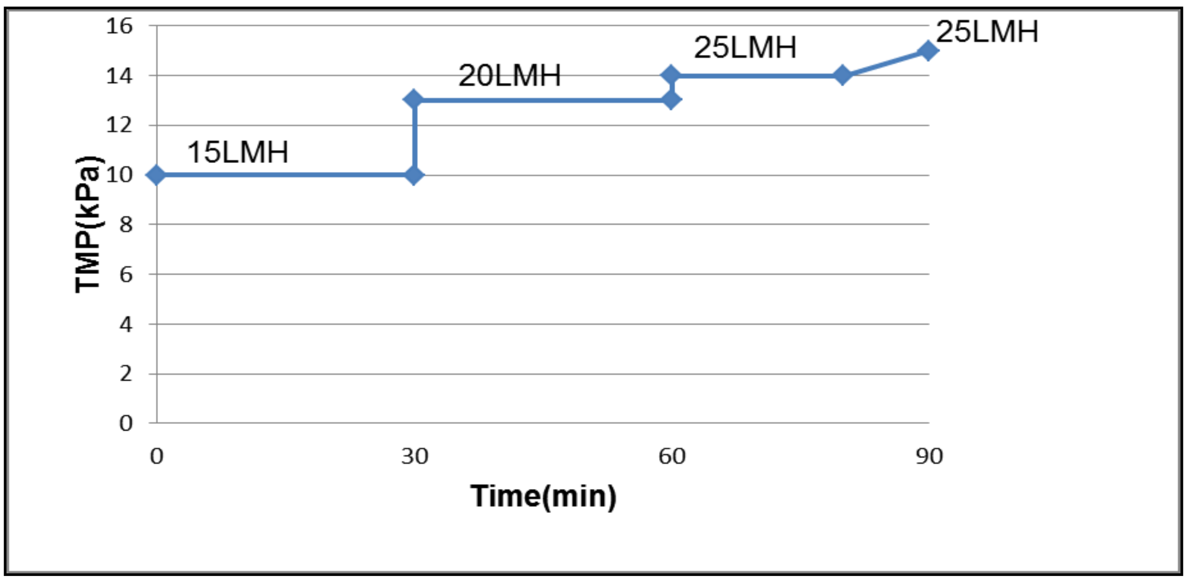

Figure C1.4: Critical flux determination at the aeration of 10L/min/module and 4g/Lsludge concentration

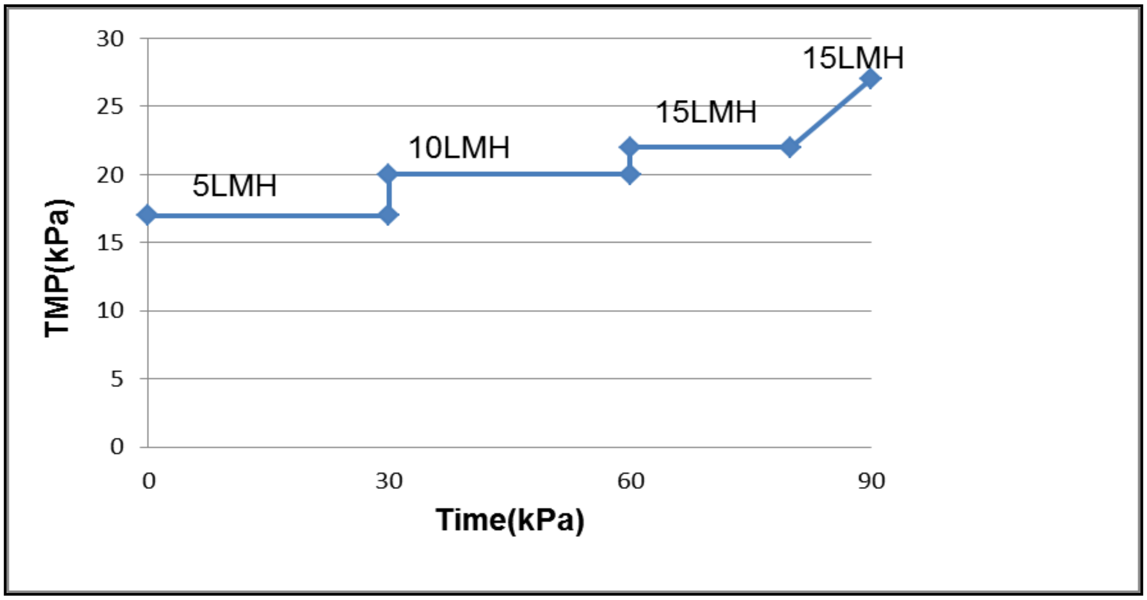

Figure C1.5: Critical flux determination at the aeration of 10L/min/module and 8g/Lsludge concentration

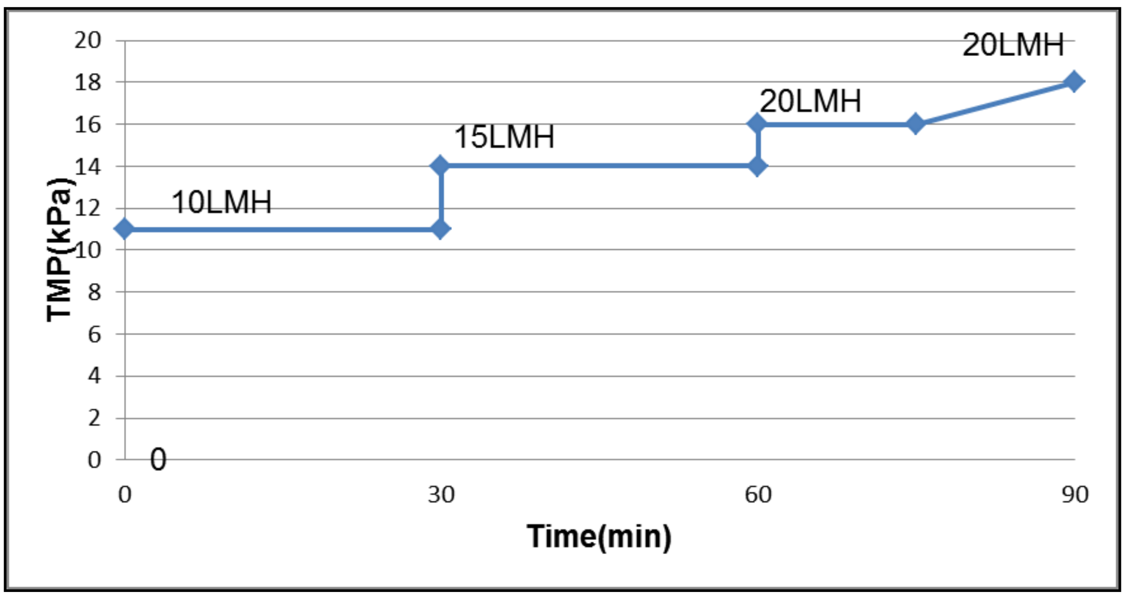

Figure C1.6: Critical flux determination at the aeration of 10L/min/module and 12g/Lsludge concentration

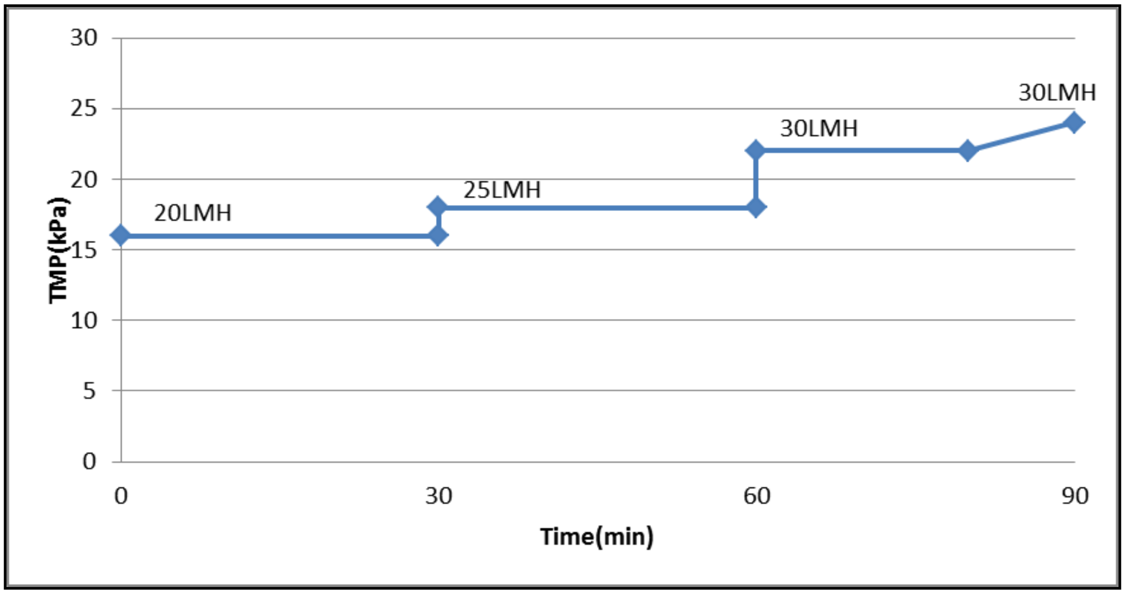

Figure C1.7: Critical flux determination at the aeration of 15L/min/module and 4g/Lsludge concentration

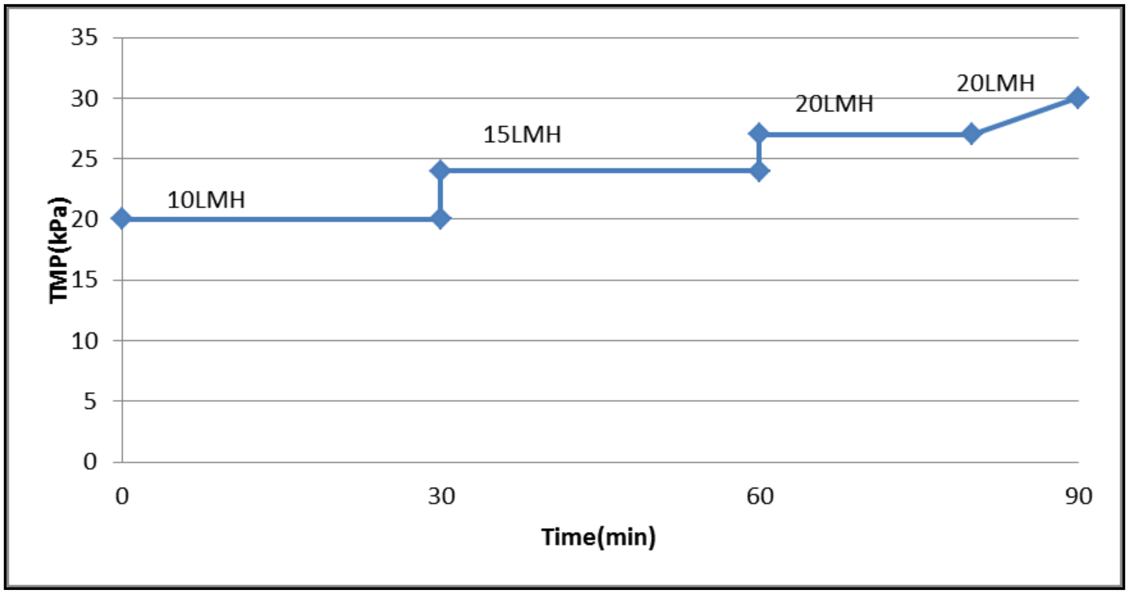

Figure C1.8: Critical flux determination at the aeration of 15L/min/module and 8g/Lsludge concentration

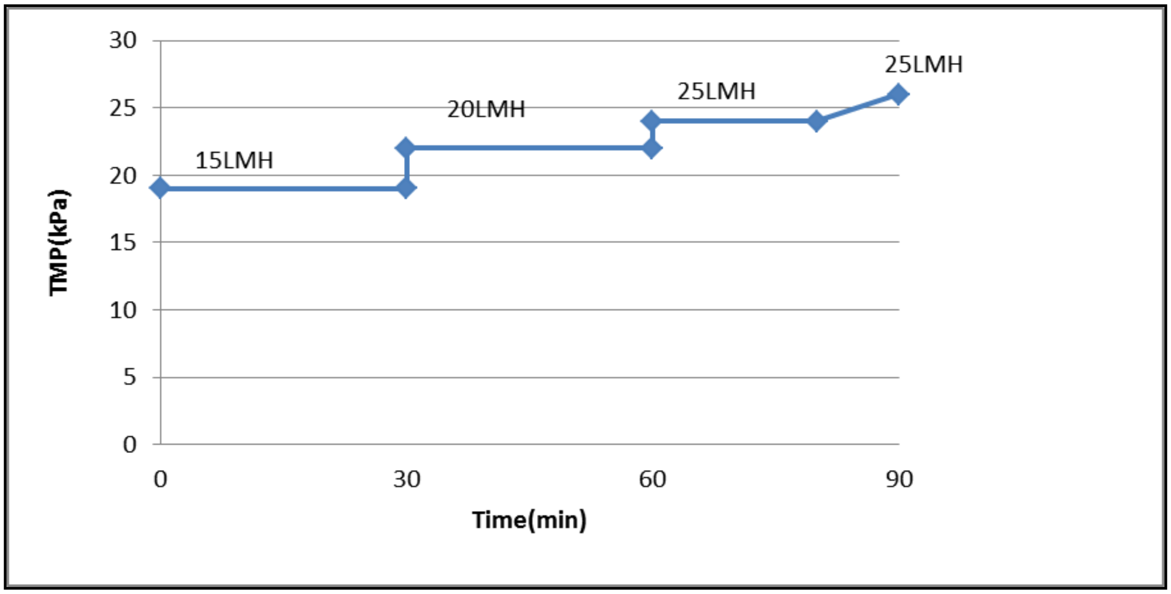

Figure C1.9: Critical flux determination at the aeration of 15L/min/module and 12g/Lsludge concentration

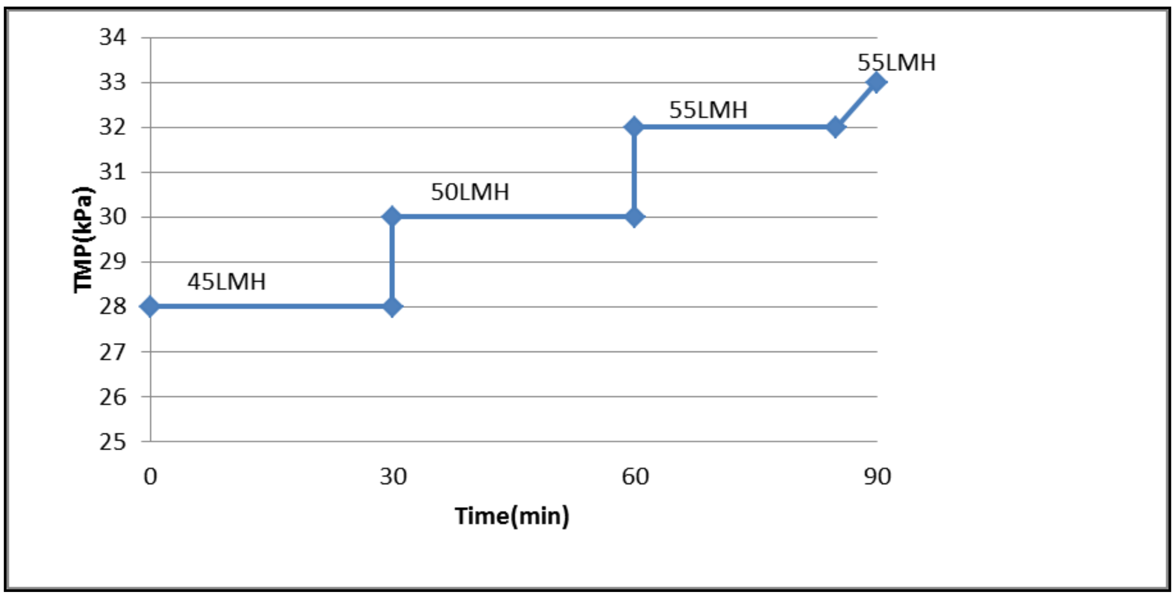

Figure C1.10: Critical flux determination at the aeration of 20L/min/module and 4g/Lsludge concentration

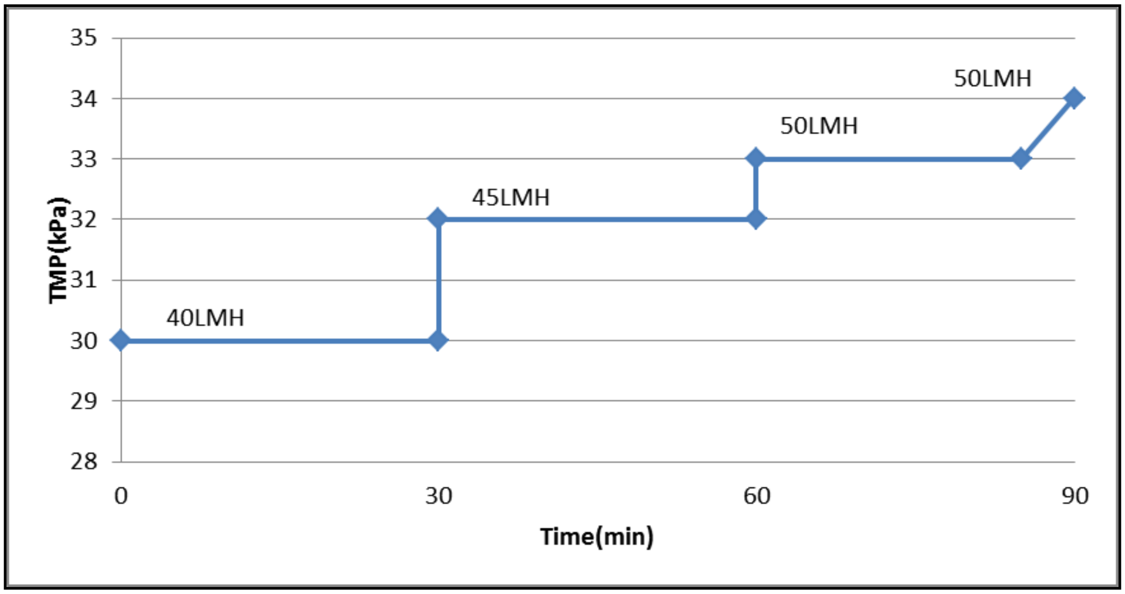

Figure C1.11: Critical flux determination at the aeration of 20L/min/module and 8g/Lsludge concentration

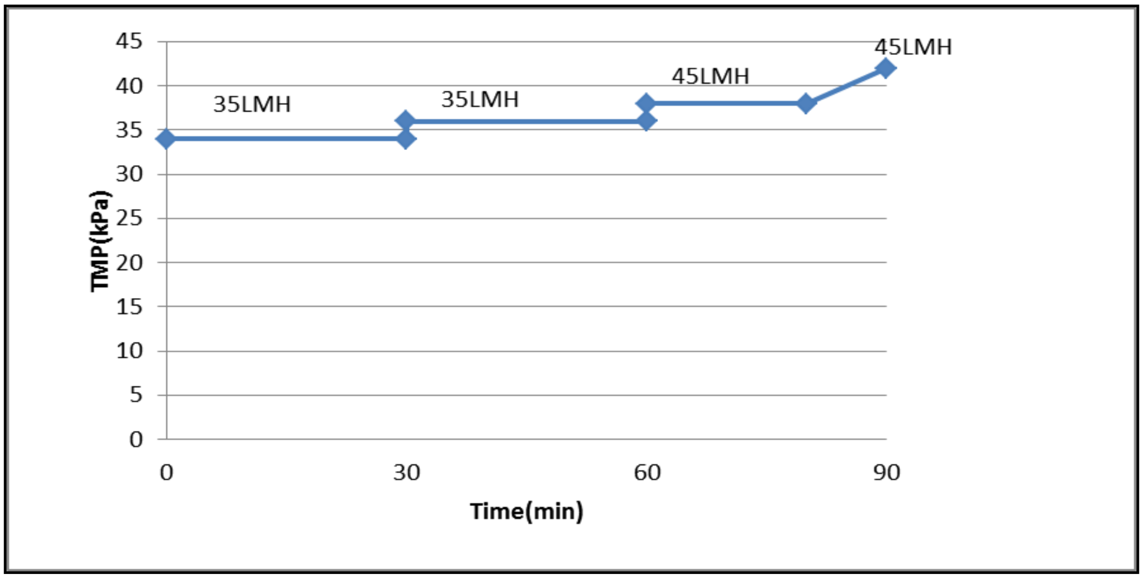

Figure C1.12: Critical flux determination at the aeration of 20L/min/module and 12g/Lsludge concentration

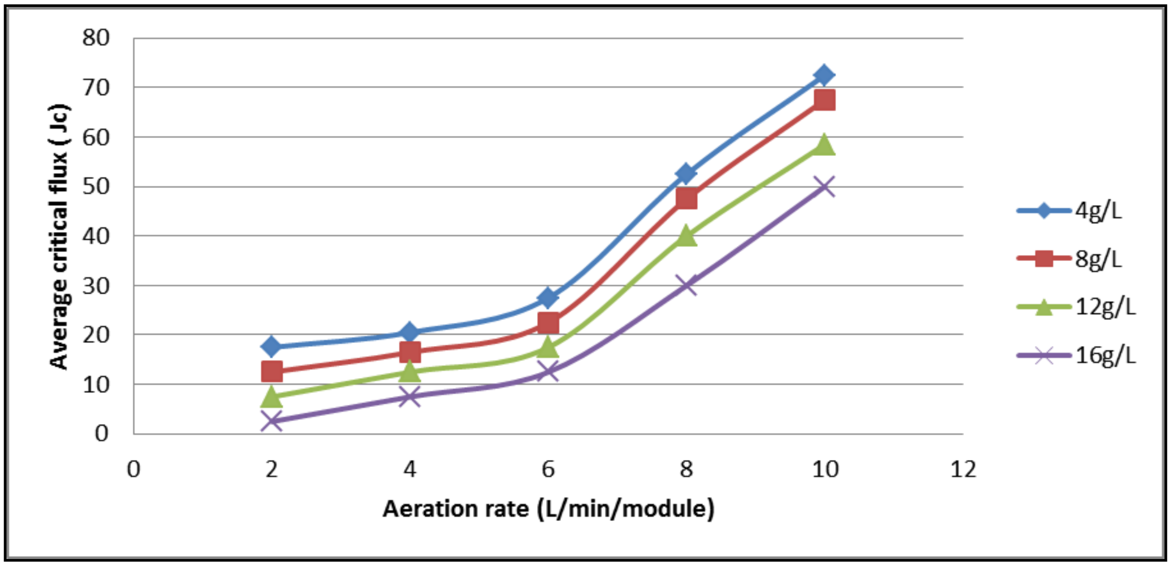

Figure C1. 13: the relationship between average critical flux and aeration rates at different concentration of activated sludge.

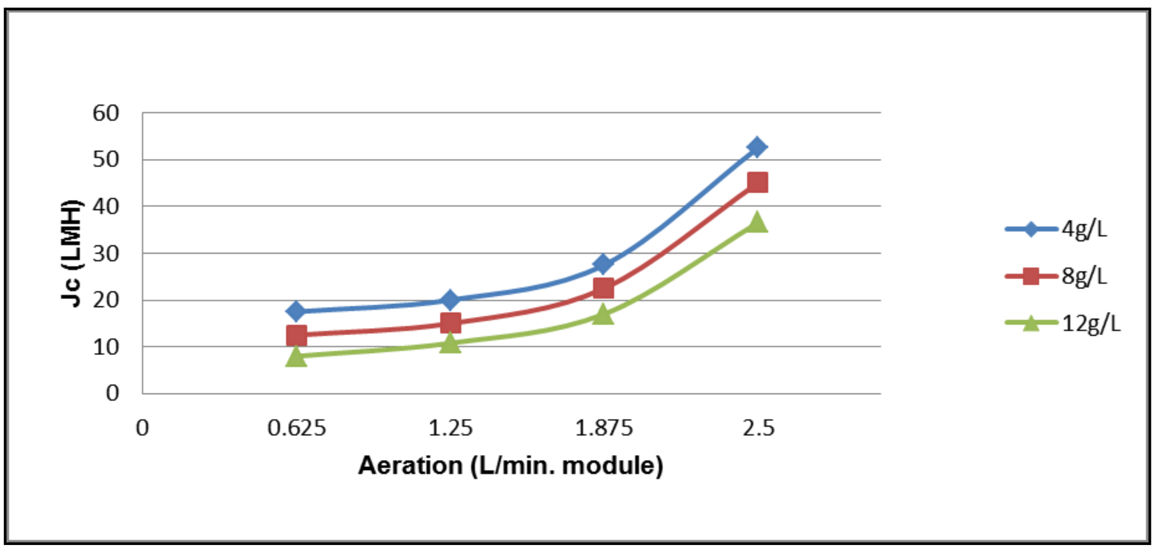

Figure C1.14: the relationship between average critical flux and aeration rates at different concentration of activated sludge.

APPENDIX D: DISSOLVED OXYGEN CURVE FOR WFM-IMBR PILOT SCALE OPERATION

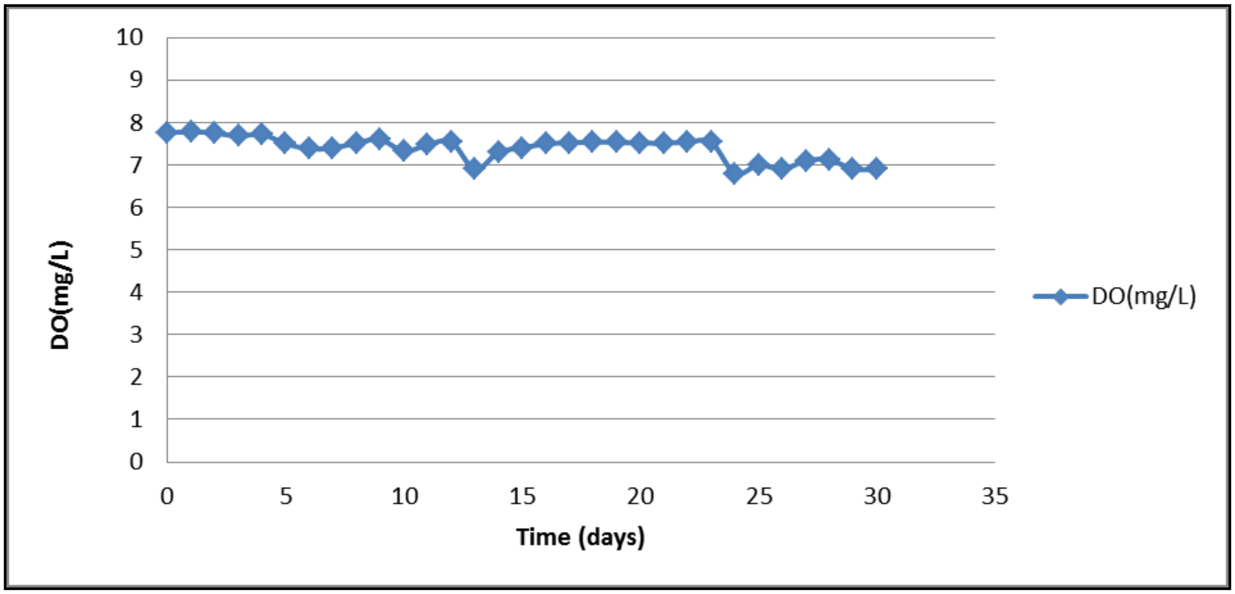

Figure D1.1: Dissolved Oxygen against time for pilot scale experimental trials

APPENDIX E: QUALITY OF RAW WASTEWATER AT DIFFERENT PHASES OF CONVENTIONAL WASTEWATER TREATMENT

Table E1. 1: Quality of raw wastewater and primary effluent at selected treatment plants in California

Quality parameters (mg/l, except as otherwise indicated)	**City of Davis**		**San Diego**		**Los Angeles County Joint Plant**	
	Raw wastewater	**Primary effluent**	**Raw wastewater**	**Primary effluent**	**Raw wastewater**	**Primary effluent**
Biochemical oxygen demand,BOD	112	73	184	134	-	204
Total organic carbon	63.8	40.6	64.8	52.3	-	-
Suspended solids	185	72	200	109	-	219
Total nitrogen	43.4	34.7	-	-	-	-
NH_3-N	35.6	26.2	21.0	20.0	-	39.5
NO-N	0	0	-	-	-	-
Org-N	7.8	8.5	-	-	-	14.9
Total phosphorus	-	7.5	-	10.2	-	11.2
Ortho-P	-	7.5	11.2		-	
pH (unit)	7.7	-	7.3	7.3	-	-
Cations:						
Ca	-	-	-	-	78.8	-
Mg	-	-	-	-	25.6	-
Na	-	-	-	-	357	359
K	-	-	-	-	19	19
Anions:						
SO_4	-		160		270	

Cl	-		120		397	
Electrical conductivity, dS/m	2.52	2.34			2.19	-
Total dissolved solids	-	-	829	821	1404	1406
Soluble sodium percentage, %	-		-		70.3	
Sodium adsorption ratio	-	-	-	-	8.85	6.8
Boron (B)	-	-	-	-	1.68	1.5
Alkalinity *($CaCO_3$)*	-	-	-		322	332
Hardness ($CaCO_3$)	-		-		265	

Source: Asano and Tchobanoglous (1987)

Table E1. 2: Quality of secondary effluent at selected wastewater treatment plants in California

Quality parameter (mg/l except as otherwise indicated)	Plant location			
	Trickling filters		Activated sludge	
	Chino Basin MWD (No. 1)	Chino Basin MWD (No. 2)	Santa Rosa Laguna	Montecito Sanitary District
Biochemical oxygen demand, BOD	21	8	-	11
Chemical oxygen demand	-	-	27	-
Suspended solids	18	26	-	13
Total nitrogen	-	-	-	-
NH_3-N	25	11	10	1.4
NO_3-N	0.7	19	8	5
Org-N	-	-	1.7	-
Total phosphorus	-	-	12.5	-
Ortho-P	-	-	3.4	-
pH (unit)	-	-	-	7.6
Cations:				
Ca	43	55	41	82
Mg	12	18	18	33
Na	83	102	94	-
K	17	20	11	-
Anions:				
HCO_3	293	192	165	-
SO_4	85	143	66	192
Cl	81	90	121	245

Electrical conductivity dS/m	-	-	-	1.39
Total dissolved solids	476	591	484	940
Sodium adsorption ratio	2.9	3.1	3.9	3.7
Boron (B)	0.7	0.6	0.6	0.7
Alkalinity ($CaCO_3$)	-	-	-	226
Total Hardness ($CaCO_3$)	156	200	175	265

Source: Asano and Tchobanoglous (1987)

Table E1. 3: Effluent quality data from selected advanced wastewater treatment plants in California

Quality parameter (mg/l except as otherwise indicated)	Plant location					
	Long Beach	**Los Coyotes**	**Pomona**	**Dublin San Ramon**	**City of Livermore**	**Simi Valley CSD**
Biochemical oxygen demand, BOD	5	9	4	2	3	4
Suspended solids	-	5	-	1	-	-
Total nitrogen	-	-	-	-	-	19
NH_3-N	3.3	13.6	11.4	0.1	1.0	16.6
NO_3-N	15.4	1.1	3	19.0	21.3	0.4
Org-N	2.2	2.5	1.3	0.2	2.6	2.3
Total phosphorus	-	-	-	-	-	-
Ortho-P	30.8	23.9	21.7	28.5	16.5	-
pH (unit)	-	-	-	6.8	7.1	-
Oil and grease	-	-	-	-	-	3.1
Total coliform bacteria, MPN/100 ml	-	-	-	2	4	-
Cations:						
Ca	54	65	58	-	-	-
Mg	17	18	14	-	-	-
Na	186	177	109	168	178	-
K	16	18	12	-	-	-
Anions:						
SO_4	212	181	123	-	-	202
Cl	155	184	105	147	178	110

Electrical conductivity, dS/m	1.35	1.44	1.02	1.27	1.25	-
Total dissolved solids	867	827	570	-	-	585
Soluble sodium, %	63.2	59.2	51.7	-	-	-
Sodium adsorption ratio	5.53	4.94	3.37	4.6	5.7	-
Boron (B)	0.95	0.95	0.66	-	1.33	0.6
Alkalinity ($CaCO_3$)	-	256	197	150	-	-
Total Hardness ($CaCO_3$)	212	242	206	254	184	-

[1]Advanced wastewater treatment in these plants follows high rate secondary treatment and includes addition of chemical coagulants (alum + polymer) as necessary followed by filtration through sand or activated carbon granular medium filters.

Source: Asano and Tchobanoglous (1987)

APPENDIX F: SUMMARY OF ESTABLISHED SMALL SCALE EXISTING IMBRS

GENERAL DESIGN OF WEISE MBR TECHNOLOGY

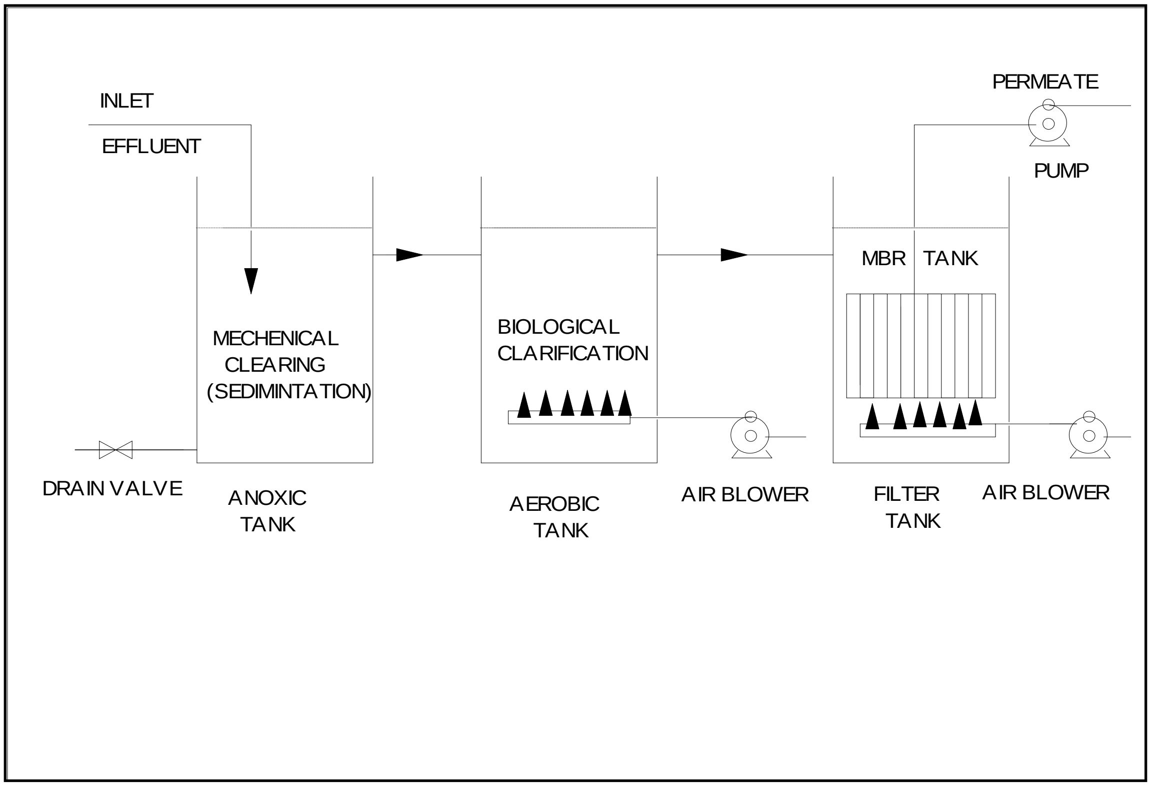

Figure F1.1: Process and instrumentation diagram of a general design of wiesa technology membrane bioreactors (microclear^{R}aquacell)

DESCRIPTION OF WIESA TECHNOLOGY

The product of Wiesa technology MBR is known as Microclear^{R}Aquacell. This product is one unit built in. This built in tank is divided into three compartments since the process is divided into three stages. Stages are mechanical clearing or sedimentation, biological clarification and membrane filtration respectively.

STAGE 1 (MECHENICAL CLEARING)

The first stage purification is performed in single tank (refer to fig. 1). This tank is divided into three zones. Solids settle out through sedimentation. The coarsely clarified water then flows over into the next stage (compartment or tank).

STAGE 2 (BIOLOGICAL CLARIFICATION)

In this stage, coarsely clarified water is aerated with air by means of a blower. This air provides various micro organisms' life support conditions depending on the amount of oxygen that penetrates into foam cubes. This causes substance to be degraded biological, without addition of chemicals.

STAGE 3 (MEMBRANE FILTRATION)

In this stage, effluent is filtered by means of membranes. Permeate (filtrate) is sucked by a pump. Aeration in this stage provides two services simultaneously, membrane scouring and oxygen supply for biological processes.

GENERAL DESIGN OF WEISE TECHNOLOGY MBR

Table F1. 1: The dimensions and operating filter parameters of the MBR

FILTER PARAMETERS		
DIMENSIONS		
Protective	207 x 207 x 497	415 x 207x 490
No. Of modules	2	2
Channel spacing	24	21
Filter area	5.5mm	6mm
Packing density	3.5 m^2	7 m^2
Membrane material	$160m^2/m^3$	$160m^2/m^3$
Weight	Polyether sulphate	Polyether sulphate
Connection	4.5 kg	10.5 kg
Filtration pressure	DN 25	DN 25
Back flush pressure	0.1.....0.15 bar	0.1.....0.15 bar
Mean flux	0.05 bar	0.05 bar
Maximum flux	15-30 LMH	15-30 LMH
Cut-off	50LMH	50LMH
Retention size	150 kDa	150 kDa
Application	Membrane bioreactor	Membrane bioreactor

THE QUALITY OUTPUT OF WATER

Table F1. 2: The permeate quality output of the Wiese technology MBR

PARAMETER	UNIT	LAYOUT	GUARANTEE
BOD_5	mg/L	< 5	
COD	mg/L	< 50	
TSS	mg/L	< 1	< 1
TKN	mg/L	< 5	-
Turbidity	NTU	< 1	< 1
pH value	-	6.5-8	-
E-Coli	x/100mL	0	0
Virus Retention	%	99.99%	99.99%

GENERAL SET UP CNFIGURATION OF WIESE TECHNOLOGY MBR

There are three set up configuration that are used by Wiese Technology.

- Internal set up with permeate extraction
- External set up with gravity filtration
- External set up with permeate extraction

INTERNAL SET UP WITH PRMEATE EXTRACTION

This is the general set up because is the same as commercial systems for example Kubota. The membrane pack is immersed inside the tank and output is sucked by a pump.

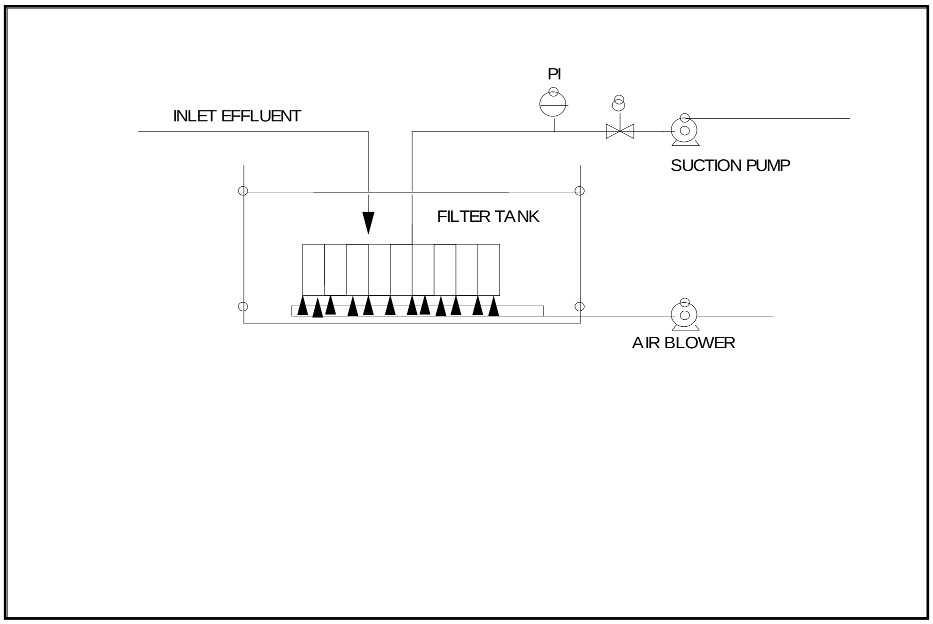

Figure F1.2: The process and instrumentation diagram of internal set up with permeate extraction membrane bioreactor.

EXTERNAL SET UP WITH GRAVITY FILTRATION

This set up is similar to the general Wiese technology MBR. The difference in this system is that permeate is collected by means of a head as a driving force therefore no pumping is required.

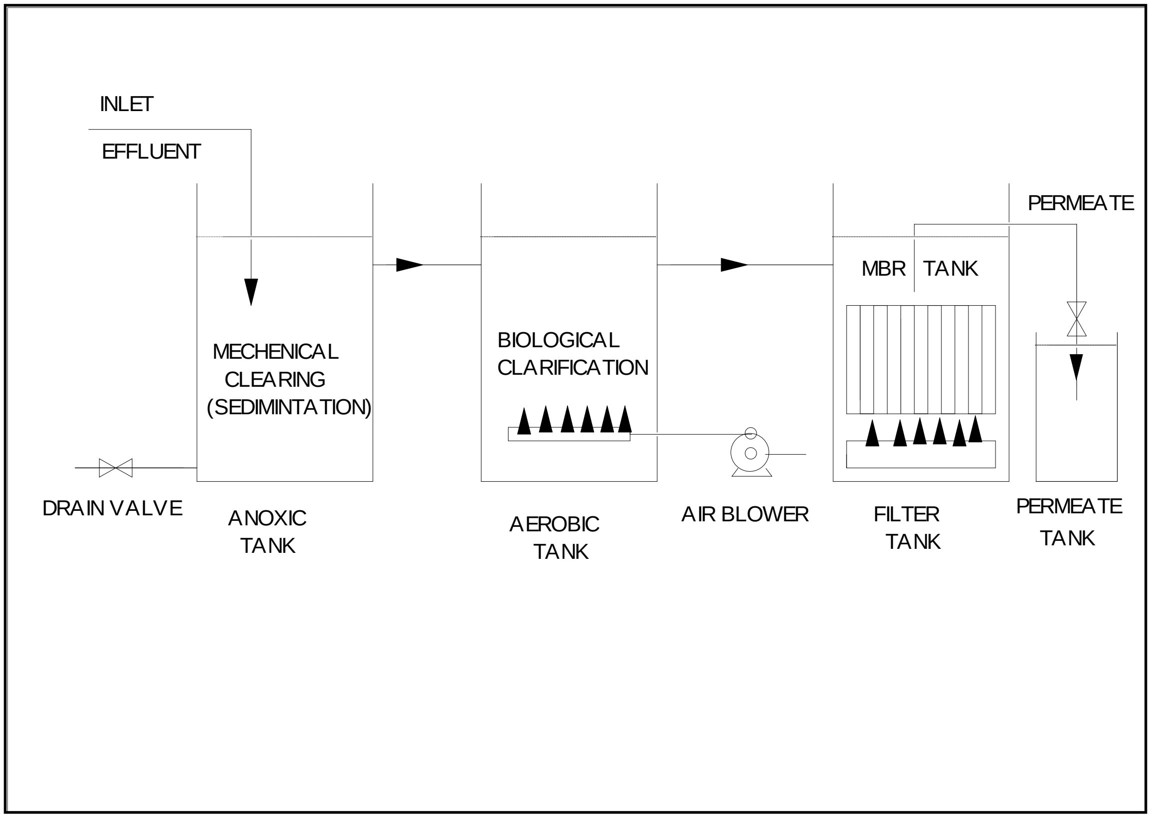

Figure F1.3: The process and instrumentation diagram of external set up with gravity filtration membrane bioreactor.

EXTERNAL SET UP WITH PERMEATE EXTRACTION

This set up is the same as external set up with gravity filtration but in this case permeate is sucked by a pump. This set up is flexible arrangement to any feed flow rate.

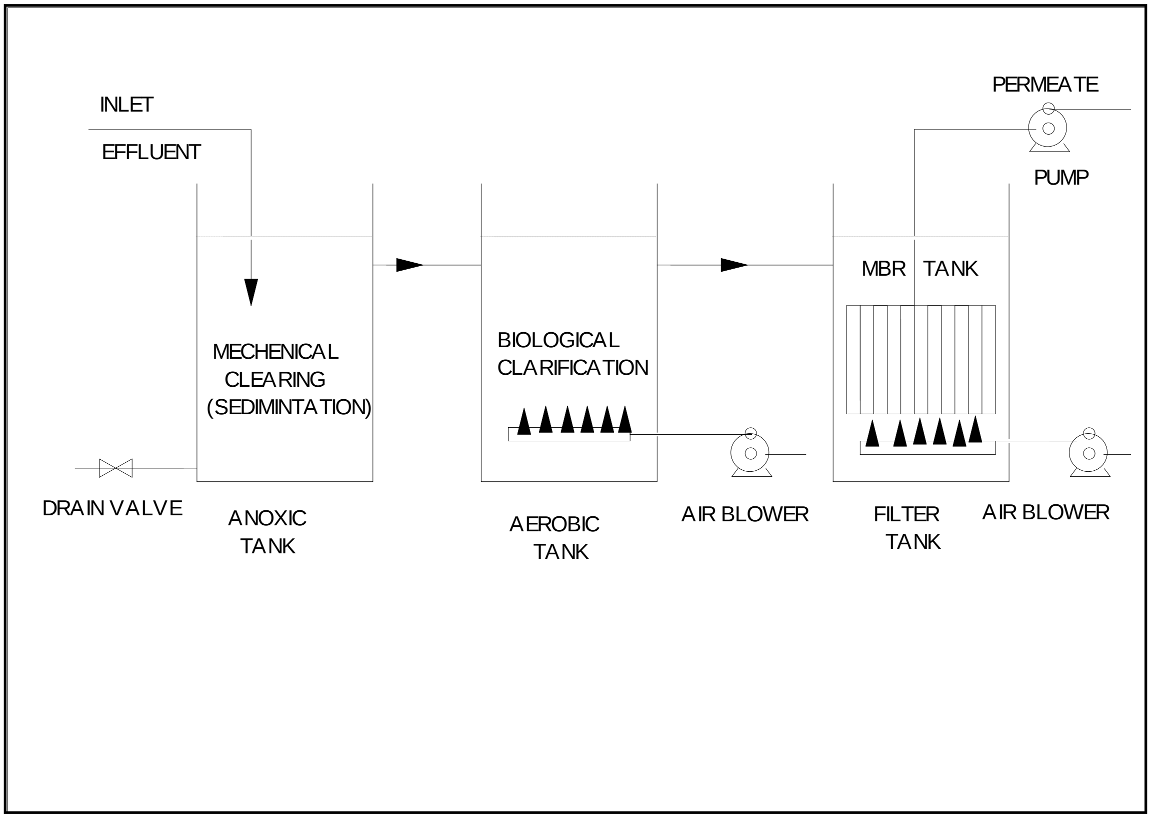

Figure F1.4: The process and instrumentation diagram of external set up with permeate extraction membrane bioreactor.

GENERAL DESIGN OF HUBER TECHNOLOGY MBR (MEMBRANE CLEAR BOX)

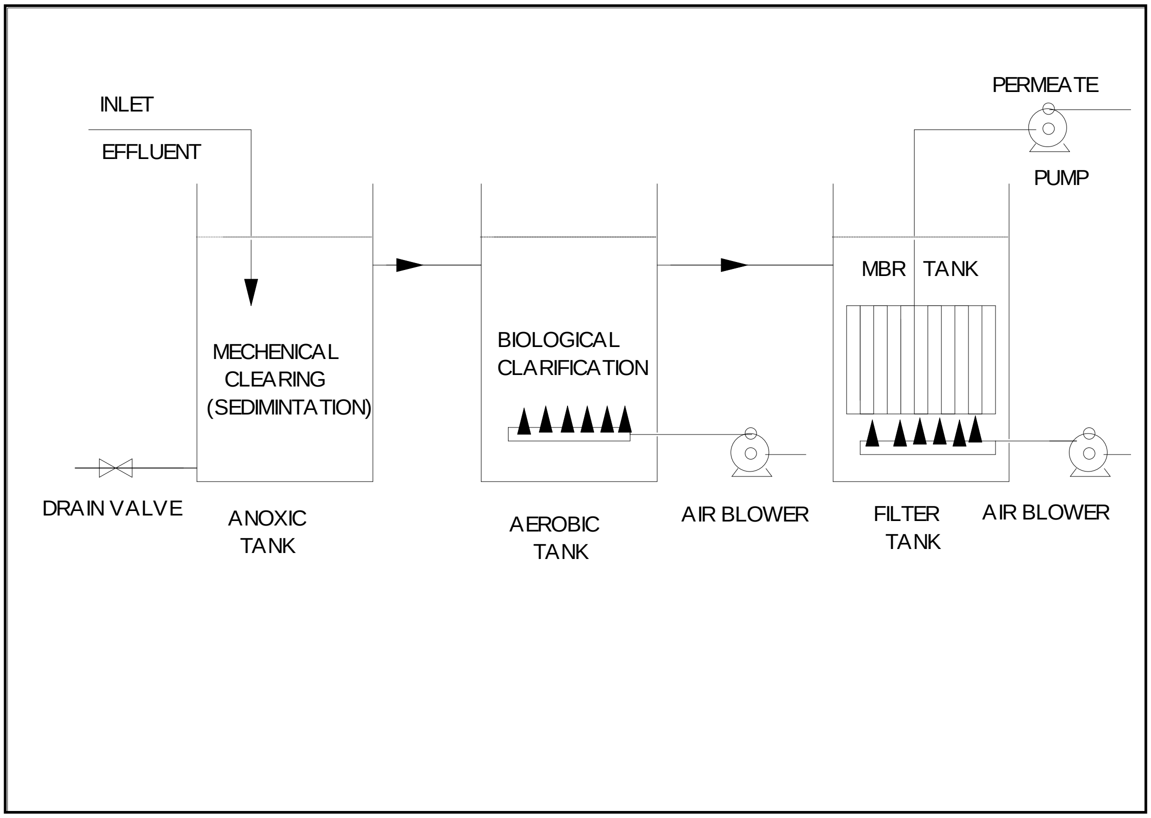

Figure F1.5: Process and instrumentation diagram of a general design of Huber technology membrane bioreactors (membrane clear box)

KEY OPERATING PARAMETERS

Table F1. 3: The representation of key operating parameters in Huber Tech. MBR

Operating parameter	value
Aeration flow rate	150L/m^2.min
Aeration pressure	250 mbar
Capacity	50-700 m^3.day

EFFLUENT PARAMETER

Table F1. 4: The permeate quality output of the Huber technology MBR

PARAMETER	LIMIT VALUE	GARANTEE
BOD_7	< 5mg/L	< 2.4 mg/L
Oxygen saturation	>50% (80-120%)	>50%
Total caliform bacteria	<100/mL	<1/mL
Faecal coliform	<10/mL	<1/mL
Pseudomonas aeroginosa	<1/mL	-

OVERVEIW SUMMURY OF HUBER TECHNOLOGY MBR

Huber technology applies two types of membrane technology, stationary modules for small scale and vibrating rotation modules (VRM) for large scale. VRM units excel with their high capacity of up to $100m^3/h$, low power consumption and long term stability.

The feature of these modules is their rotation that allows sequential high-intensity air scouring of their membrane surface.

For smaller application Huber technology MBR employs membrane clear box (small unit). These small units use stationary modules.

The design of the Huber technology MBR and Wiese technology MBR is the same. Both of them, their stages are fitted in one tank. The tank is divided into three compartments for mechanical separation, biological growth and membrane filtration.

WOVEN FIBRE MICRO FILTRATION MEMBRANE BIOREACTOR

According to our design here in southern Africa, we are using a single tank and we only use a single stage (membrane filtration). It is quite reasonable for the following reasons:

- To save energy for air supply into the system
- There is a suspension consistency of feed in our system
- Simple design
- Small investment
- Robust and reliable membranes employed

APPENDIX G: CHEMICAL OXYGEN DEMAND TEST METHOD

Apparatus required

i. COD digester block
ii. Stop watch
iii. COD reagents
iv. DR/890 Hach Calorimeter
v. Glass beaker
vi. Pipette
vii. Filter paper

Testing procedure

i. Filter the sample into 500 mL glass beaker.
ii. Shake the reagent 5 times.
iii. Pipette the 2 mL of the filtered solution into the reagents.
iv. After step (iii), Shake the reagent 10 times and put it into the digester block.
v. Repeat step (i) up to (iv) for all samples.
vi. Connect digester block into the power supply.
vii. Start the digester and burn the samples for 160 minutes at 120 ºC.
viii. After step (vii), stop the digester and switch off the power.
ix. Wait up until the temperature of the digester block is below 40 ºC.
x. Set the DR/890 calorimeter into COD mode and choose the COD range required.
xi. Zero the calorimeter.
xii. Put the samples into the calorimeter and press read.
xiii. Repeat step (xii) for all samples.

APPENDIX H: MIXED LIQOUR SUSPENDED SOLIDS STANDARD TEST METHOD

Apparatus required

i. Oven
ii. Analytical Balance
iii. Glass fibre filter papers
iv. Vacuum pump
v. Pocylene funnel
vi. Measuring cylinder

Testing procedure

i. Weigh the empty filter paper on analytical balance machine.
ii. Place the weighed filter paper on the pocylene funnel.
iii. Measure the desired volume of the sample.
iv. Filter the sample through the filter paper by using a vacuum pump.
v. Place the sample on the pre heated oven at 103ºC -105 ºC for 160 min.
vi. After step (v), cool the sample for 30 minute by placing the sample in decator.
vii. Weigh the sample on the analytical balance machine.
viii. Subtract results of (i) and (vii).
ix. Record the results in mg/L.
x. Repeat step (vii) to (ix) for all samples.

APPENDIX I: TURBIDITY TEST METHOD

Apparatus required

i. DR5000Hach Calorimeter
ii. Glass beaker

Testing procedure

i. Set the calorimeter into turbidity mode.
ii. Zero the calorimeter.
iii. Take a sample with 100 mL beaker.
iv. Pour into calorimeter test tube.
v. Wipe the test tube with clean cloth.
vi. Put the test tube into the calorimeter.
vii. Press read button.
viii. Repeat step (iii) to (vii) for all samples.

APPENDIX J: DISSOLVED OXYGEN TEST METHOD

Apparatus required

i. DR 6000 kit

Testing procedure

i. Insert the probe into the solution.
ii. Press read button.
iii. Wait until the reading stabilized.
iv. Write down the results in mg/L.
v. Repeat step (i) to (iv) for all samples.

APPENDIX K: AERATION-CRITICAL FLUX CURVES

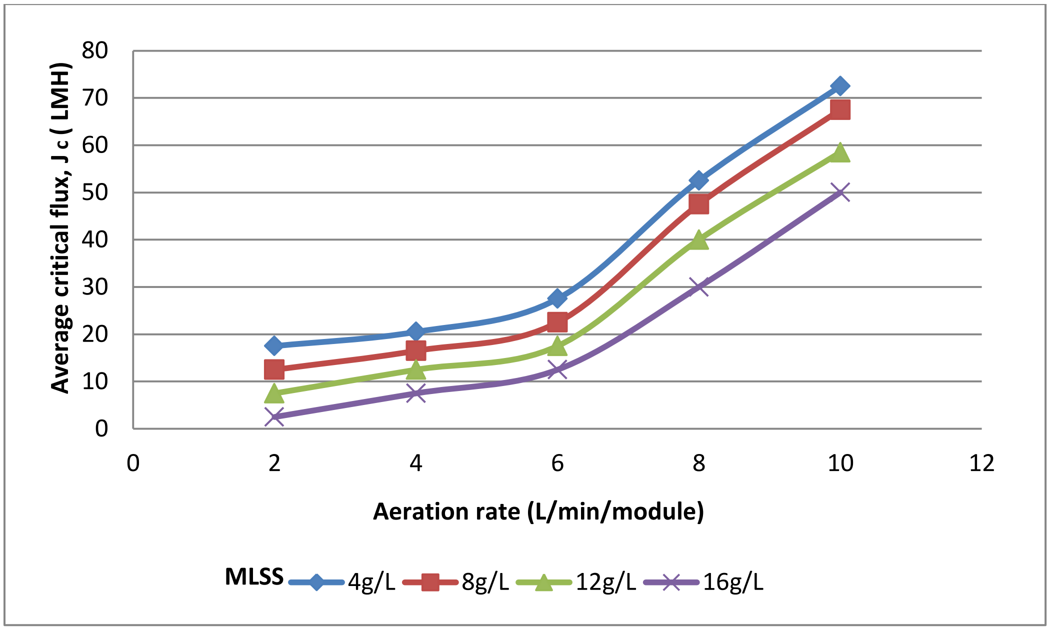

Figure K1:reltionship between aeration and critical flux at fixed sludge concentration.

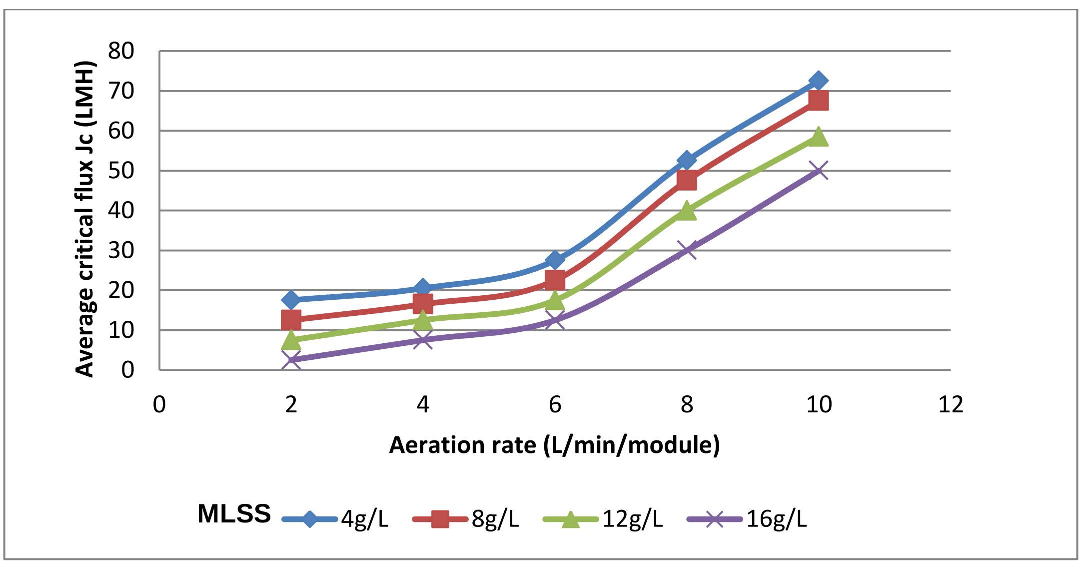

Figure K2:reltionship between aeration and critical flux at fixed sludge concentration. Repeatibility 2.

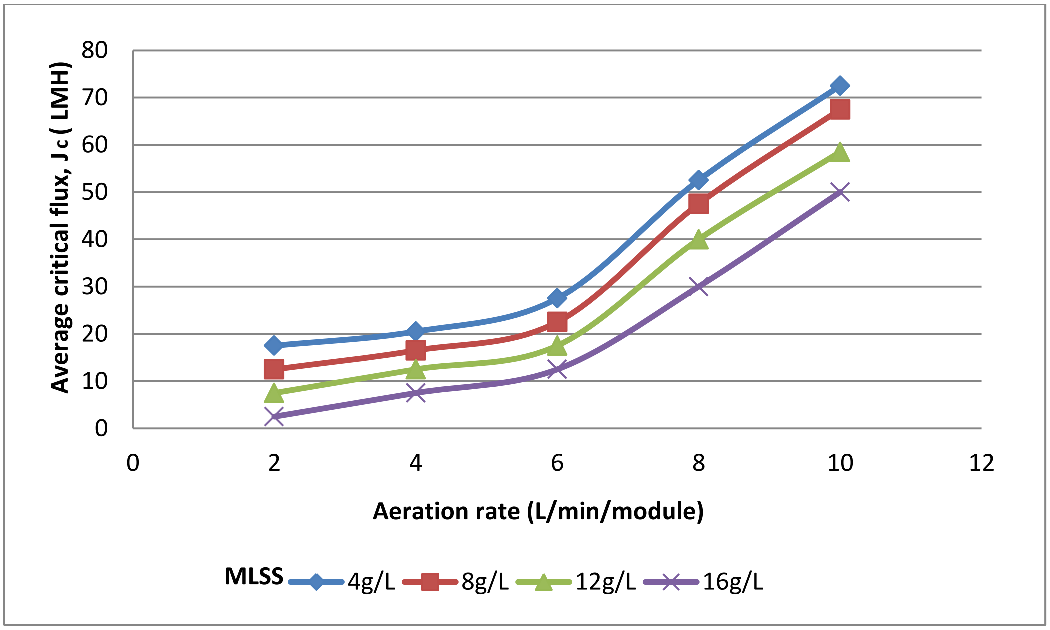

Figure K3:reltionship between aeration and critical flux at fixed sludge concentration. Repeatibility 3.

APPENDIX L: DETERMINATION OF THE BUBBLE SIZE DIAMETER

BUBBLE SIZE CHARACTIRIZATION ON BENCH SCALE EXPERIMENTATION

1. Introduction

The study of bubble size characterization was conducted prior to WF-IMBR pilot scale experimentation. The importance of this was to quantify the optimum bubble size for fouling combating and the bubble size that can travel between the normal spacing of flat sheet IMBR namely 5 mm, 6 mm 7 mm etc.

The unit of 0.5 m by 0.38 m by 1.19 m was designed in-house at Durban University of Technology, Chemical Engineering Laboratory(see figure L3.1).

Despite an increasing number of publication on the use of two-phase flow as flux enhancing strategy, there is still very little known about the manner in which two –phase flow enhances membrane flux. Slug flow has been shown to be the most effective two phase flow regime for bubbling inside tubular modules. For immersed flat sheet systems, the most effective two-phase flow regime has not yet been fully identified and thus the use of two-phase flow in this system has not yet been optimized.

Factors govern the two phase flow profile are gas feeding rate and diffuser pore size and geometry. Although numerous studies have been published on two phase flow profiles in bubble columns (Sokolichin & Eigenberger, 1994; Lapin & Lubert,1994a; Renade &Taylor, 2001). None of these studies has focused on the effect of the nozzle size and geometry in membrane systems particularly in membrane bioreactor applications. Some of the question need to be answered includes: 1) what is the effect of nozzle size on bubble size distribution? 2) Are small bubbles or large bubbles more effective for enhancing the flux? 3) What is the effect of air flow rate and nozzle size on bubble rise velocity? 4) What is the effect of nozzle geometry on bubble size distributions? 5) This chapter will attempt to develop a link between bubble size distribution and bubble rise velocity to flux enhancement.

2 Objectives

The specific objectives of this study are to:

- Study the effect of diffuser pore size diameter and geometry on bubble size distribution.
- Study the effect of airflow rate on bubble size distribution.
- Study the effect of air flow rate and diffuser pore size diameter on bubble rise velocity.

3 Experimental set up

A diagram of the experimental rig used for the experiments is presented in figure K 3.1. The process feed tank (40 cm x 20 cm x 100 cm) was open to atmosphere. Air was supplied to the system by means of a blower. The amount of air was measured using a rotameter. The walls of the MBR were constructed using Perspex. For this experimentation there was no need for suction of permeate therefore no filtering module was immersed inside the reactor. The pure tap water was used in this experimentation for the simplicity of observing the bubbles. The Hill-blow blower was used for aerating the tank (forming bubbles) on this experiment.

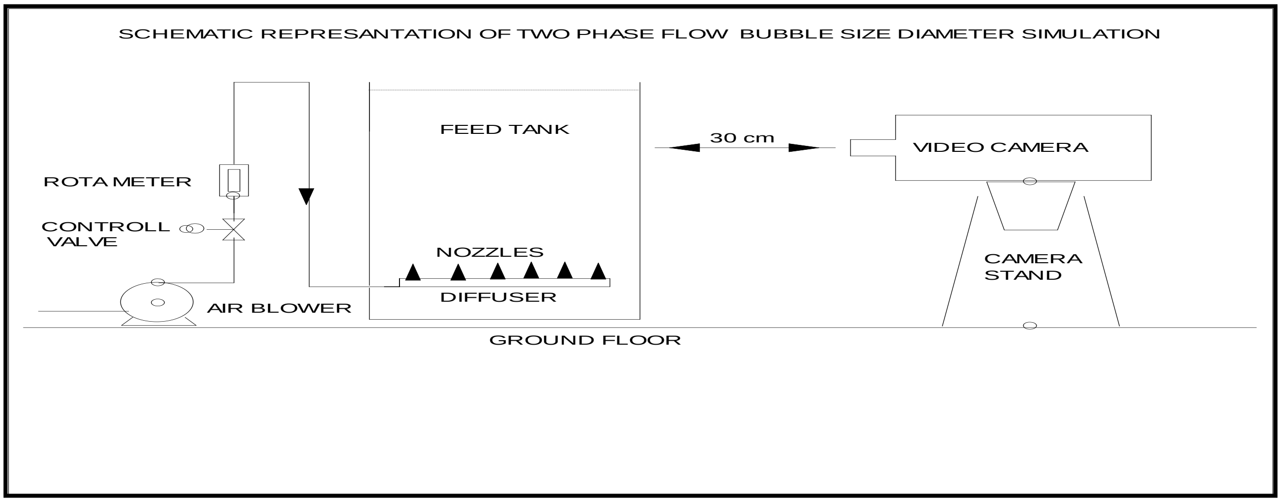

Figure L3.1: A diagram of the experimental rig used for bubble size diameter determination experiments.

To achieve the above objective three identical diffusers with same number of pores were designed. The diffusers were drilled 1 mm, 1.5 mm and 2 mm respectively pore size diameter. Each diffuser was drilled 10 pores. The canon digital camera, positioned about 30 cm from the feed tank was used to take digital images and to record short 120 seconds movies at bubbling pattern The desired diffuser was fitted to the tank. The tank was then filled with water. The air blower was switched on and the desired air flow rate selected between 8, 10 and 12 L/min. The first experimenting point of 8L/min was chosen because it was the first aeration where the bubbles are observed. The video moves were downloaded to a PC. The movies were then analyzed using AnalySIS8 software. For each movie analyzed, the software produced information such as total number of bubbles and mean bubble diameter.

4 Equipment descriptions

4.1 Process Tank

The tank was constructed with Perspex Plastic in order to make it transparent. Perspex made it easy to study the two-phase flow phenomenon. The dimensions of the tank were 40 cm by 20 cm by 100 cm respectively. The process tank was open to atmosphere and was filled with 60 L of tap water. However, no membrane was inserted to process tank because the interest of the study was to study the bubble distribution at varying conditions (see objectives of the study).

4.2 Diffuser

Diffuser in this study was made from PVC rod of 15 mm. The rod was drilled uniformly the cylindrical 15 holes of 0.5 mm, 1 mm, 1.5 mm and 2 mm respectively. The holes were 10mm equidistance. For cylindrical holes of 1.5 and 2 mm, corresponding square holes with equivalent open areas were made. The diffuser was coupled to the process tank 15 cm above the tank base (see figure K 3.1).

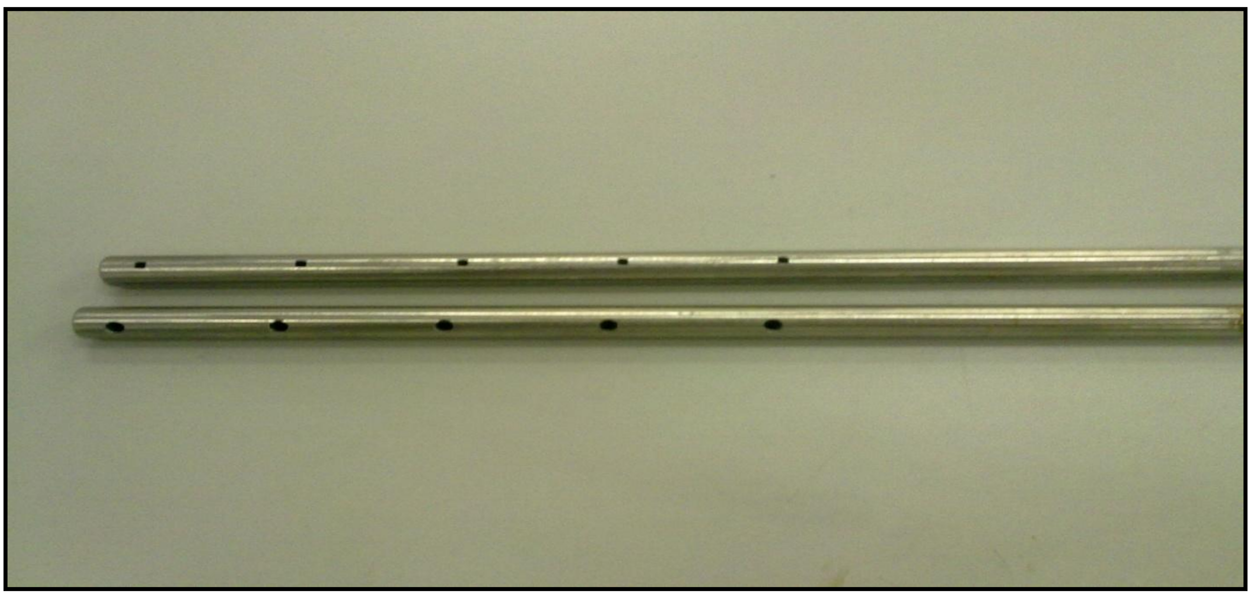

Figure L3.2: Schematic representation of the diffuser

4.3 The Air Blower

The Hillblow air blower was employed. It was connected to the diffuser passing through the globe valve and air flow meter (see figure K3.1). The air flow meter was used to manipulate the air flow rate.

4.4 Digital Camera

Kodak digital camera was used to take experimental videos. The camera was placed 30cm next to process tank. The videos of 2 minutes duration were taken at every given condition.

5 Experimental procedure

The process tank was filled up with tap water. The blower switched on. The startup procedure was followed (see appendix A1). The desired air flow rate was selected on the rotameter by regulating the globe vale. The air flow rate was varied from 8 L/min, 10 L/min and 13 L/min. The digital camera was place 30 cm to the process tank. The videos of 2 minute were taken at each given aeration rate and different height of a process.

The video moves were downloaded to a PC. The movies were then analyzed using AnalySIS8 software. For each movie analyzed, the software produced information such as total number of bubbles and mean bubble diameter.

5.1 Bubble size diameter determination method

For each experiment conducted, the IPG movies were stored on a computer. The analysis software was used for bubble size characterization. The analysis software produced the number of bubbles corresponding to its bubble size diameters. The number of bubbles versus bubble size diameter graphs was plotted for each aeration flow rate used.

5.2 Bubble rise velocity determination method

For each movie taken, the movie was broken into frames by Adobe premier 5.1, Adobe Photoshop software was used to make the frames more visible. The Quick trial software was used to predict the bubble rise velocity from the movie frames provide. The distance travelled by bubbles/ time was predicted by the software, it was assumed to be the bubble rise velocity at that aeration flow rate given.

6. Results

POPULATION BUBBLE SIZE DIAMETER CHARACTERISATION

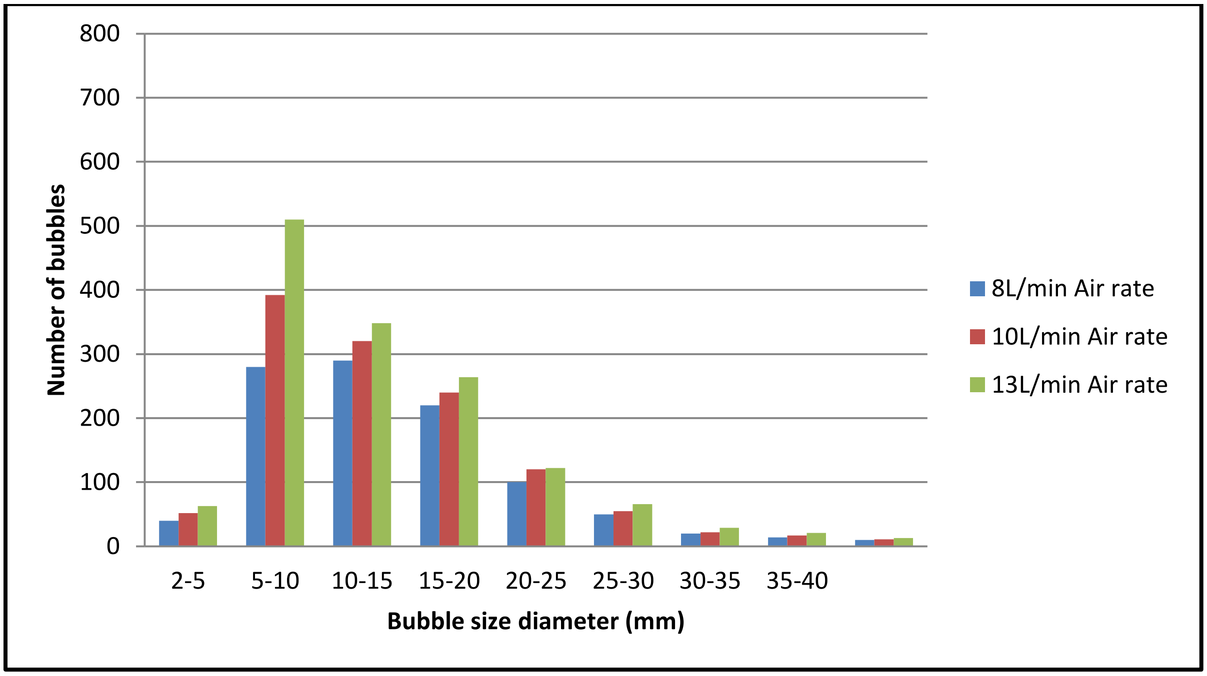

Figure L3.3: The relationship between number of bubbles and bubble size diameter for 1mm diameter nozzle

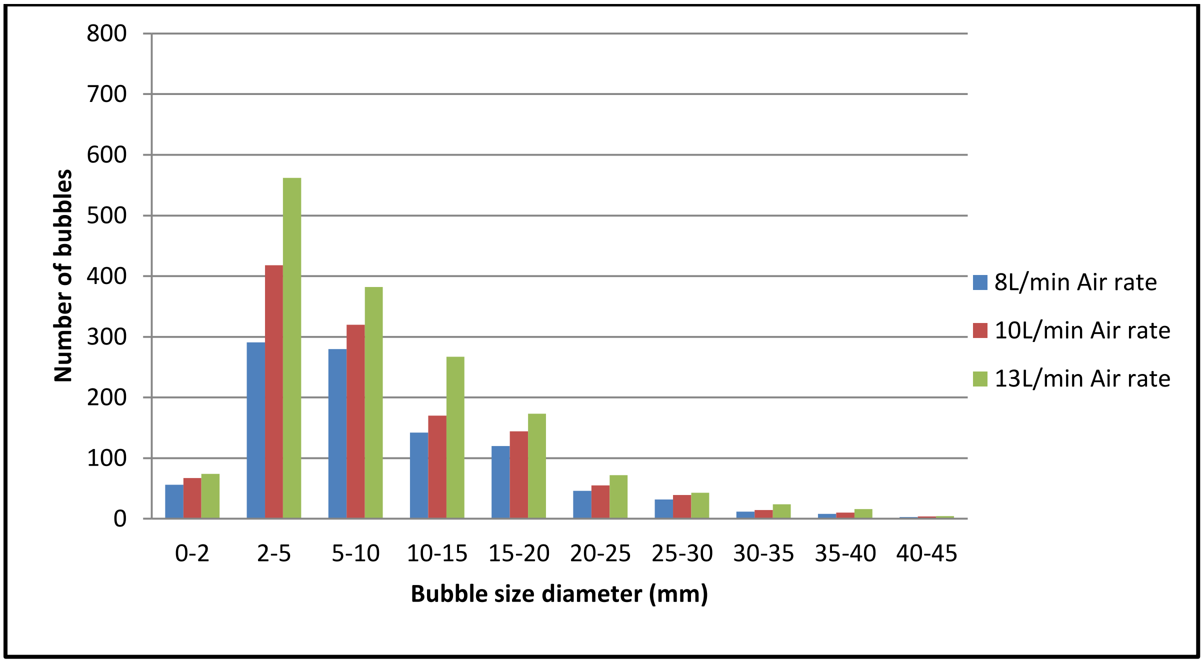

Figure L3.4: The relationship between number of bubbles and bubble size diameter for 1.5 mm diameter nozzle

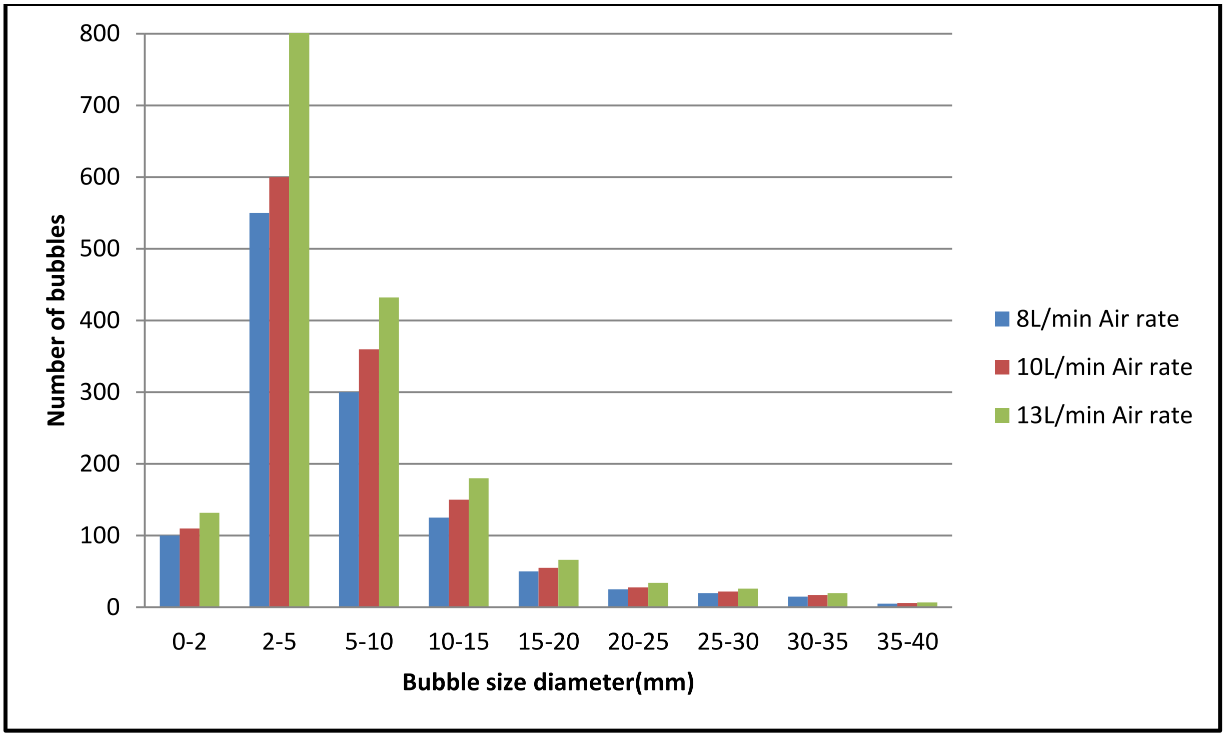

Figure L3.5: The relationship between number of bubbles and bubble size diameter for 2 mm diameter nozzle

7 Discussion of results

The analysis software was used in this study to develop the relationship between the number of bubbles and bubble size diameter. The software presented the following data

- bubble size diameter
- number of bubbles

The graphs were plotted based on the produced data by the analysis software; refer to figure L3.2, L3.3 and L3.4. From all plotted graphs presented (L3.2, L3.3 and L3.4) 2-5mm bubble size diameter was dominant.

It was observed that bigger bubble sizes were formed due to smaller bubble sizes colliding to each other. It was also observed that bubble size diameter slightly increase with decrease in pressure (decrease in head of the tank). The aeration flow rate was found to have effect on the number bubbles formed.

2mm diameter nozzles produced more 2-5 mm bubble size diameter compared to 1.5 and 1 mm diameter nozzles. The behavior of bubbles was observed in a manner that at low flow rate they rose almost in a straight line, while at high flow fate they tend to migrate to center of the column and move vigorously in a zigzag manner.

APPENDIX M: DETERMINATION OF THE BUBBLE SIZE DIAMETER AND BUUBLE RISE VELOCITY

DETERMINATION OF THE BUBBLE RISE VELOCITY

EXPERIMENTAL PROCEDURE

i. For accurate investigation, only bubbles with the same diameter rising in the central part of the Perspex tank were taken into consideration.
ii. Two software packages were used in this investigation, adobe premier 5.1 and Corel draw.
iii. The same video movies used for bubble size distribution analysis were used in this experiment.
iv. Adobe premier 5.1software was used to break down the video movies into frames of 60 seconds.
v. At this point the duration taken by bubbles to move from bottom of the frame to the end was known.
vi. The distance travelled by bubbles along a frame was evaluated by using Corel draw software.
vii. By knowing the time taken by bubbles to move from one point to another and the distance travelled by bubbles, the velocity of bubbles was calculated based on the following definition : $Velociy(mm/s) = \frac{Distacetravelled(mm)}{time(s)}$

RESULTS

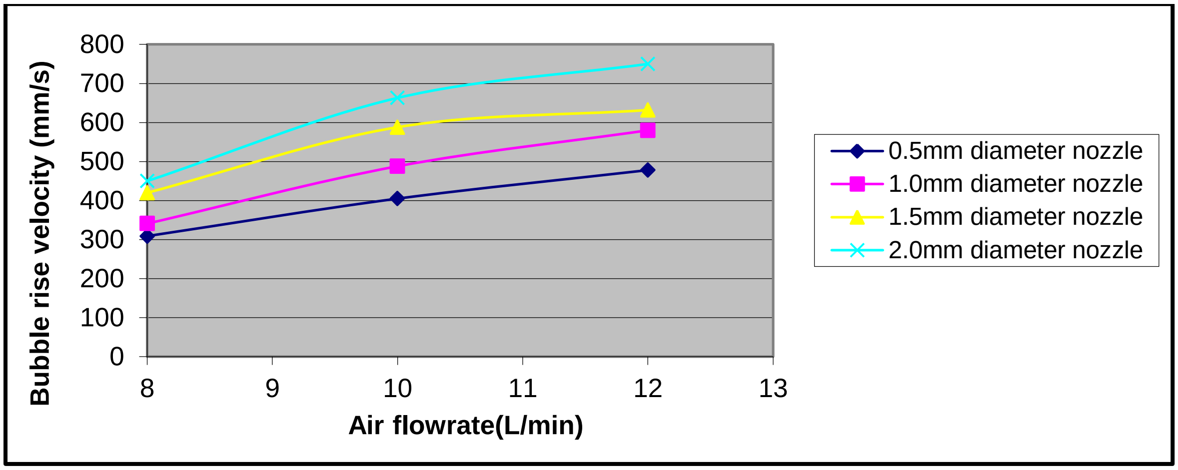

Figure M1.1: The relationship between bubble rise velocity and air flow rate in a water media under standard temperature and pressure.

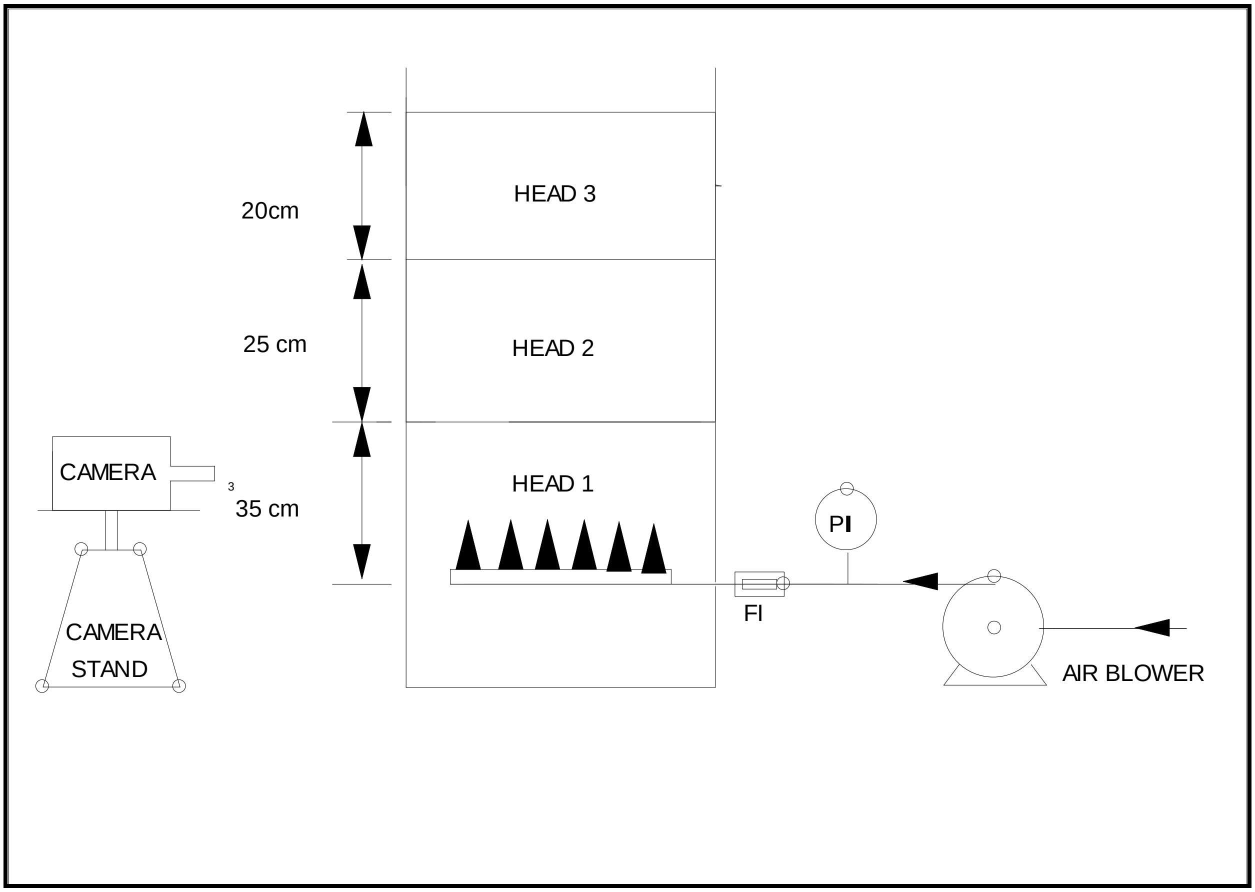

Figure M1.2: The equipment used for determination of bubble rise velocity.