An Evaluation of Pre-concentration Technologies for Domestic Sewage to Enhance the Performance of Anaerobic Digestion

by

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Abstract

This study compares the performance of three different pre-concentration technologies; woven fiber microfiltration (WFMF), tube settler (TSET) and conical membrane tank (CMT) that can apply to, concentrate the domestic sewage prior to the anaerobic treatment. The main goal of the pre-concentration is to concentrate as much as possible of the wastewater organic matters in a separate stream, which can later be used for energy recovery.

The pre-concentration performance was evaluated in terms of chemical oxygen demand (COD), suspended solid (TSS) concentration and the energy consumption. WFMF was able to concentrate 21 to 24.2 g COD/ m^3. d of COD, while CMT had 17.5 to 19.7 g COD/ m^3. d. TSET indicated that the lower COD pre-concentration performance with 0.005 m/h and 0.01 m/h loading rates as 1.8 and 2.6 g COD/ m^3. d. In terms of TSS accumulation, WFMF and CMT resulted in more than 90% while TSET had 63%. In terms of the effluent quality, the WFMF was able to remove 68% of COD while CMT has 77%. This could have a potential of reuse application of the permeate water for agricultural purpose. Thus, the WFMF was found to perform better among the three technologies in terms of domestic sewage pre-concentration.

Once found to be the best performing technology as WFMF 7.5 LMH flux, it was continued for pre-concentrating the domestic sewage. Concentrated domestic sewage was used as a feed water to the anaerobic membrane bioreactor (AnMBR) and the performance was evaluated in terms of biogas generation, methane content of the biogas and the removal efficiencies of TSS, BOD and COD. The AnMBR was able to generate biogas 28 mL/g COD with concentrated domestic sewage. 38 % of methane content was found to be in the biogas for concentrated domestic sewage at 3.2 kg COD/m^3.d loading rate. TSS, BOD, and COD removal efficiencies were 99, 67, and 71% for the AnMBR process, respectively.

Finally, this research proved that the capturing of solid fraction from the domestic sewage can lead to generate higher COD concentrations that can be effectively used in the anaerobic digestion process. Furthermore, this research discussed the common unit approach on different pre-concentration technology comparison and importance of the sludge cone volume for the mass balance approach. Moreover, this concept helps to reduce the anaerobic reactor volume by concentrating the domestic sewage.

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List of Abbreviations

AD	Anaerobic Digestion
AF	Anaerobic Filter
AnMBR	Anaerobic Membrane Bioreactor
AP	Aromatic Polyimide
APHA	American Public Health Association
ASBR	Anaerobic Sequencing Batch Reactor
ASM	Aerated Sewage Microfiltration
AWWA	American Water Works Association
BMP	Bio Methane Potential
BOD_5	Biochemical Oxygen Demand
CCM	Combined Coagulation Microfiltration
CE	Cellulose Acetate
CEPT	Chemically Enhanced Primary Treatment
CIP	Clean In Place
COD	Chemical Oxygen Demand
COP	Clean Out Of Place
CSTR	Continuously Stirred Tank Reactor
DAF	Dissolved Air Flotation
DO	Dissolved Oxygen
DSM	Direct Sewage Microfiltration
EC	Electrical Conductivity
ED	Electro Dialysis
EPA	Environmental Protection Agency
HRT	Hydraulic Retention Time
IMBR	Immersed Membrane Bioreactor
LMH	Liters/m^2/hour
MBR	Membrane Bioreactor
MF	Microfiltration
MGD	Million Gallons per Day
MLSS	Mixed Liquor Suspended Solids
MPN	Most Probable Number
NF	Nano Filtration
NH_4-N	Ammonia Nitrogen
N_{tot}	Total Nitrogen
NTU	Nephelometric Turbidity Units
O&M	Operation and Maintenance
OM	Organic Matter
OLR	Organic Loading Rate
PA	Aliphatic Polyimide
PAN	Polyacrylonitrile
PEI	Ployetherimide
PES	Polyethersulfone

pH	Power of Hydrogen
PI	Polyimide
PP	Polypropylene
PSF	Polysulphone
PTFE	Polytetrafluoroethylene
P_{tot}	Total phosphorous
PVC	Polyvinyl Chloride
PVDF	Polyvinylident Fluoride
RO	Reverse Osmosis
SEM	Scanning Electron Microscope
SRT	Solids Retention Time
T	Temperature
TDS	Total Dissolved Solids
TMP	Transmembrane Pressure
TP	Total Phosphorus
TSS	Total Suspended Solids
UASB	Up-flow Anaerobic Sludge Blanket
UF	Ultra-Filtration
VFA	Volatile Fatty Acids
WF-IMBR	Woven Fiber Immerged Membrane Bioreactor
WFCM	Woven Fiber Conical Membrane
WFIMMF	Woven Fiber Immersed Membrane Microfiltration
WFMF	Woven Fiber Micro Filtration

Chapter 1

Introduction

1.1 Background of the Study

Currently, one-sixth of mankind facing issues to access to any type of improved water system within 1 km radius for their household and improved excreta disposal facility (Davison et al., 2005). Lack of access to improved water could lead to water pollution or the water scarcity. Both the factors are affecting the available and usable water amount. Treatment of the domestic or industrial wastewater is expensive. Especially, for the goal of reuse for drinking, use for production and manufacturing. Discharging wastewater without proper treatment will lead to polluting the existing usable water bodies and partially treated wastewater also have the same results in the end. It is important to focus on the sanitation sector and find the most appropriate treatment technology for the situation. Treatment technologies can vary with the wastewater or excreta's physicochemical or biological characteristics. One technology may not apply to the all situations. Some technologies like anaerobic digestion (AD), have valuable by-products like methane gas. Focus on that kind of technology is a plus sign as it can treat the waste, with having valuable by-product.

Domestic sewage contains the detritus of our daily lives-faeces, leftover food particles, detergents and pharmaceuticals and many other contaminants. Domestic sewage, as per its starting point and arrangement can be divided into greywater, originating from sinks, shower, kitchen, laundry and blackwater, originating from the toilets. Considering the blackwater, it contains a high load of organic pollutants and pathogenic microorganisms even it generates in less quantity but represents the greatest contamination risk. In many countries, biological process is widely used to treat domestic sewage. In the biological treatment process, microorganisms are used to degrade the organic matter. The biological process can categorize mainly into two sections which namely aerobic and anaerobic process. The treatment efficiency depends upon the movement of an extensive variety of microorganisms, converting complex organic materials present in wastewater.

Over the past years, the domestic sewage has been treated using activated sludge process which is still the most popular process in wastewater treatment sector. The process is effective and also the simple as removing organic pollutant from the wastewater. But this comes with high energy consumption and the carbon footprint. To overcome this issue, the treatment sector currently moving to the anaerobic process which is economically viable. The sanitary engineers and decision makers are now progressively focusing about the anaerobic digestion of domestic sewage. Anaerobic digestion of wastewater treatment is being utilized effectively as a part of tropical, subtropical and temperate areas of the globe (Seghezzo et al., 1998). Anaerobic digestion is a procedure in which microorganisms extract energy and develop by metabolizing organic matter in a non-oxygen environment resulting in the generating of methane. Currently, biogas is considered as one of the main sources of non-renewable energy. Biogas can be easily converted into heat energy that is created from the burning process. Commonly, biogas is used for transportation, domestic use (heating and cooking), power generation (electricity) and industrial production process. Anaerobic procedures are considered as cost-effective and sustainable technology for wastewater treatment because of low biomass production, less energy requirement and diminish greenhouse gas outflow through the use of methane gas. The potential for biogas production and amount of organics in wastewater streams are quantified by the Chemical Oxygen

Demand (COD). During the anaerobic digestion process, the biodegradable COD present in the organic material is converted into methane and the other by-products.

Considering the current typical situation in sewage treatment, it is important to send the sludge generated from conventional treatment facility to an anaerobic digester to control the carbon footprint through capturing methane gas. Disposal of sewage sludge without treatment to landfill can lead to release 40-100 kg CO_2-eq/ (IE yr) due to release the methane gas (Diamantis et al., 2011) . Applying anaerobic practices directly to domestic wastewater could reverse those effects entirely and generate an excess of energy, but it is not currently possible with low concentrations of organics (Smith et al., 2014). In that case, anaerobic treatment plant can make a use of methane that produces electricity than consume it (McCarty et al., 2011). One of the major drawbacks of the anaerobic digestion is, efficiency decrement when it comes to the diluted phase. Domestic sewage is diluted due to the mixing of graywater. The efficiency of the anaerobic digestion shows the higher values when the wastewater is concentrated. Pre-concentration of the domestic sewage can lead to minimize the carbon footprint and the treatment cost. Moreover, it can help to maximize the water reuse potential, energy and nutrient recovery (Diamantis et al., 2011). Pre-concentration of domestic sewage produces an organically rich wastewater stream that is suitable for the anaerobic digestion process (Verstraete and Vlaeminck, 2011).

The main goal of fractionation or the pre-concentration is to concentrate as much as possible of the wastewater organic matters in a separate stream, which can later be used for energy recovery. Once it is pre-concentrated, due to its high organic load, the concentrate can easily be subjected to anaerobic digestion to extract energy and nutrient recovery. Presently, the pre-concentration process is conducted by the technologies such as chemically enhanced primary treatment (CEPT), dissolved air flotation (DAF), clarification tanks, etc. Moreover, the membrane technology is one of the advanced technology, which is widely used for both water and wastewater treatment processes. Thus this technology is widely applicable for the purpose of the pre-concentration process.

Nevertheless, very few studies are reported using the technology for pre-concentration of domestic sewage. It is important to develop pre-concentration technologies for domestic sewage and compare the efficiency of the anaerobic digestion. Thus, this study was focused on pre-concentrating domestic sewage with two different membrane configuration processes and the tube settler application. Performance evaluation and technology comparison were done and evaluated. The best performing technology was coupled with the lab-scale anaerobic reactor and evaluated the performance.

1.2 Objectives of the Study

The overall objective of this study is to develop different pre-concentration technologies for domestic sewage to improve anaerobic digestion efficiency in the final stage. To achieve this objective, two following objectives are proposed.

1. To study pre-concentration, efficiency of domestic sewage with woven fiber microfiltration, tube settler and conical membrane tank applications.
2. To evaluate the performance of anaerobic digestion, with best performing pre-concentration technology.

1.3 Scope of the Study

The research based on the bench scale experiments in order to achieve the objectives as mentioned above with following steps.

1. Three different laboratory scale, pre-concentration setups were fabricated and carried out experiments at AIT, EEM ambient laboratory.
2. AIT campus domestic sewage was used as a feed water for the experiment, which is a mix of graywater and blackwater.
3. Laboratory scale, anaerobic digestion system was fabricated and carried out experiments to evaluate the efficiency enhancement on pre-concentrated domestic sewage.

Chapter 2

Literature Review

2.1 Domestic Sewage

Domestic sewage flows and quality, mainly depend on the generating source, sub-streams, population density, habits and culture and also geographic and socio-economic variations. Domestic wastewater consist of different flows, which can be discharged separately (blackwater or graywater) or combined sewage. Backwater and graywater streams can be categorized into sub-streams. Table 2.1 presents the sub-streams categories of the domestic sewage.

Table 2.1 Domestic Sewage Sub-streams and their Sources (Friedler et al., 2013)

Stream	Sub-stream	Source
Blackwater	Yellow water	Urine
	Brown water	Faeces and toilet paper
	Beige water	Anal cleansing water
Greywater	Light greywater	Shower Bath tube Bathroom washbasin
	Dark greywater	Kitchen sink Dishwasher Washing machine, and laundry where applicable

2.1.1 Blackwater

Blackwater directly come from the toilets. It consist of flushing water with faeces, urine and wiping materials. Blackwater contains a high number of pathogenic microorganisms. The concentration of this waste stream is dependent on the amount of flushing water that use in the toilets. Mostly the conventional toilets about 10 L per flush is used. Also, Pour-flush toilets use 2-5 L per flush and modern vacuum toilets only use 1 L per flush (De Mes et al., 2003).

According to the **Table 2.1** blackwater can be divided it to main three sub-streams namely, yellow water, brown water and beige water. In conventional system, this brown and yellow water is mixed together and in urine diversion toilets can help to separate them and treat individually. For the treatment objective, it is important to consider the compounds that consist in the blackwater. **Table 2.2** summarize the compounds in the urine and faces.

Based on the **Table 2.2** it can be identified that the humanities generally produce 1,010–1,530 $g{\cdot}p^{-1}{\cdot}d^{-1}$ of urine with 94–96% water content. The remaining fraction is typically come from the nutrients and dissolved solids. Healthy human urine does not contain pathogenic microorganisms that will be transmitted through the environment. Mostly, the risks come with urine is due to contamination by faeces (Jönsson et al., 2004).

Table 2.2 Compounds in Urine and Faeces (Larsen et al., 2013)

Parameter	Unit	Urine	Faeces
		Range	Range
Wet mass	$g \cdot p^{-1} \cdot d^{-1}$	1,010–1,530	100–350
Dry mass			31–53
Water content	%	94–96	65–85
pH		5.0–7.2	
EC	$mS \cdot cm^{-1}$	8.7–31	
TSS	$g \cdot p^{-1} \cdot d^{-1}$		6–60
BOD_5		1.8–10	4.3–20
COD		5.0–24	2.6–63
N_{tot}		4–16	0.3–4.2
P_{tot}		0.8–2.0	0.3–0.8
K		1.0–4.9	0.2–1.3
Fecal coliforms	$cell \cdot p^{-1} \cdot d^{-1}$		10^8–10^{11}

Generally yellow water has a low organic load which is 5.0–24 $gCOD \cdot p^{-1} \cdot d^{-1}$. This organic load consists of various organic and amino acids, creatinine and carbohydrates. Those constituents can be degraded by anaerobic process (Udert et al., 2006). Moreover, urine is a rich source of nitrogen, phosphorus and potassium in domestic sewage. That r represents the possibility of agricultural application as a fertilizer after certain treatments.

Considering the brown water, human generate 100–350 $g \cdot p^{-1} \cdot d^{-1}$ of feaces with water content of 65–85%. Generally, those values are depending on the dietary habits, health, climate and many other factors. Feaces consist with relatively higher organic load which is a maximum of 63 $gCOD \cdot p^{-1} \cdot d^{-1}$. But, it has relatively lower in nutrients. This brown water consists of very high load of pathogens. As a indicator it can identify that fecal Coliforms around 10^8–10^{11} $cell \cdot p^{-1} \cdot d^{-1}$. Toilet paper also contributes to the high level of organic load in the brown water. Toilet paper contribution to the brown water TSS and COD are 11% and 8%, respectively (Friedler and Butler, 1996).

2.1.2 Greywater

Wastewater generates in-house activities such as washing, bathing does not contain or contaminated with excreta and therefore, graywater has less pathogens and some amounts of nutrients (N, P, K). Wastewater volumes and concentration are mainly dependent on water consumption. The components in greywater are based on water quality, quality and source. Basically, greywater is characterized in four categories based on its origin, namely: laundry, bathroom, kitchen and mixed origin. Grey water contains food particles, soaps, oil and grease, chemicals come from the costumes and pathogen. **Table 2.3** shows the grey water characteristics in different countries.

Base on the organic loading, grey water can be characterized into the low and high load. For high load grey water, it includes the more organically rich wastewater which is coming from the kitchen and laundry. Basically, graywater generation is 50-80% of domestic water usage and it has a large potential of treatment and reuse in household level (Al-Hamaiedeh and Bino, 2010)

Table 2.3 Domestic Wastewater Characteristics in Selected Countries (Morel, 2006)

Parameter	Costa Rica	Palestine	Israel	Nepal	Malaysia	Jordan
Flow (L/p/d)	107	≈ 50	≈ 100	72	≈ 225	≈ 30
pH		6.7-8.35	6.5-8.2	-	-	6.7-8.35
EC (µS/cm)	≈ 400	1585	1040-2721	-	-	475-1135
COD (mg/L)		1270	822	411	212	-
BOD (mg/L)	167	590	477	200	129	275–2287
COD/BOD	-	2.15	1.72	2.06	1.64	
TSS (mg/L)	-	1396	330	98	76	316
NH_4-N (mg/L)	-	3.8	1.6	13.3	13	-
PO_4-P (mg/L)	16	4.4	126	3.1	-	-
Fecal coli (cfu/100ml)	$1.5–4.6\times 10^8$	3.1×10^4	2.5×10^6	-	-	1.0×10^7

The light graywater and dark graywater characteristics are based on the people's life style, the different kind of appliances use in the household that consume chemicals such as washing machines. For the reuse purpose, it is important to identify the pollutants that include in the graywater. Dark graywater is rich with the higher pollutant load compared to light graywater which is coming from shower or bathroom washbasin (Friedler, 2004).

Considering the biodegradability of the graywater, COD/BOD_5 ratio is more important. For the light graywater which generates from the shower or bathtubs, represent COD/BOD_5 2–3.6 which is higher and it is indicated that light gray water has less biodegradability. This COD/BOD_5 ratio shows the level of biodegradation capacity of the wastewater and higher biodegradability wastewater indicating less than 2.0 or 2.5 of COD/BOD_5 ratio (Morel and Diener, 2006).

2.2 Source Separation and Pre-concentration of Wastewater

Domestic sewage consists of high energy value. This energy can be extracted and recovered from the wastewater. To achieve that objective, the domestic sewage characteristics need to bring to the certain level that could easily apply for the energy recovery process. **Figure 2.1** shows the benefits and the possibility of energy recovery from wastewater.

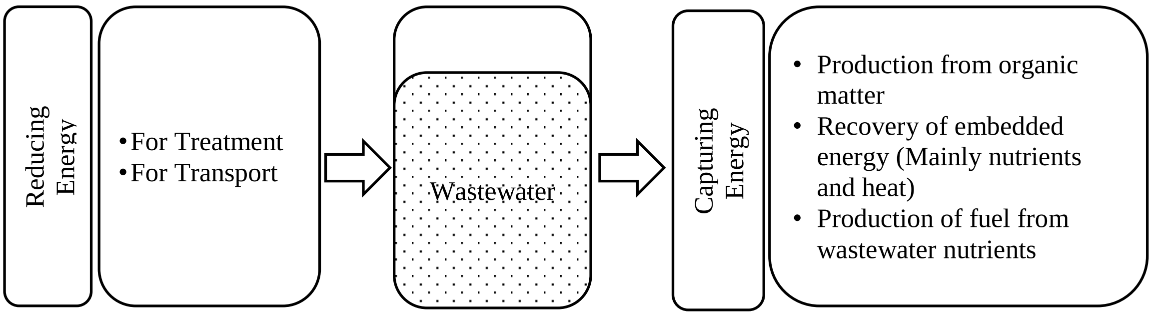


Figure 2.1 Energy capturing opportunity in wastewater (Larsen et al., 2013)

Urban wastewater treatment efficiency and productivity can be increased by solid and liquid faction separation in to the maximum level (Diamantis et al., 2011). This separation can be done at two points, namely at the source or with later fractionation. Pre-concentration or the fractionation can be identified as an alternative to separation in the source. Pre-concentration and fractionation leads to diminishing the carbon footprint, water and nutrient reuse or recovery as well as the reducing operational cost and increase the energy recovery.

For the objective of energy recovery, the pre-concentration or fractionation is helping to concentrate the organic portion of the wastewater in maximum portion. There are many technologies currently available for this purpose, such as chemically enhanced primary treatment (CEPT), Dissolved air flotation (DAF), bio flocculation and direct sewage filtration (Verstraete et al., 2013) . **Figure 2.2** illustrates the concept of pre-concentration for mixed municipal wastewater.

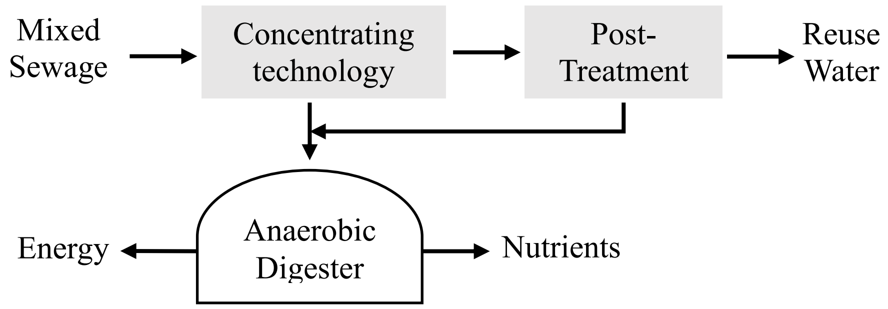
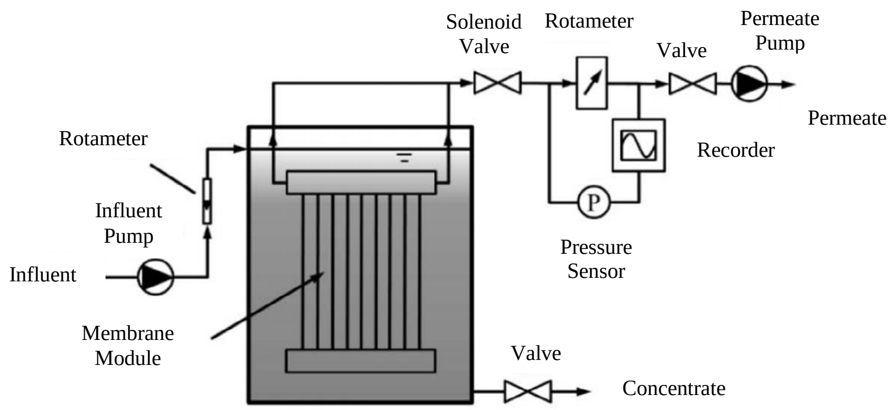


Figure 2.2 The concept of sewage pre-concentration (Larsen et al., 2013)

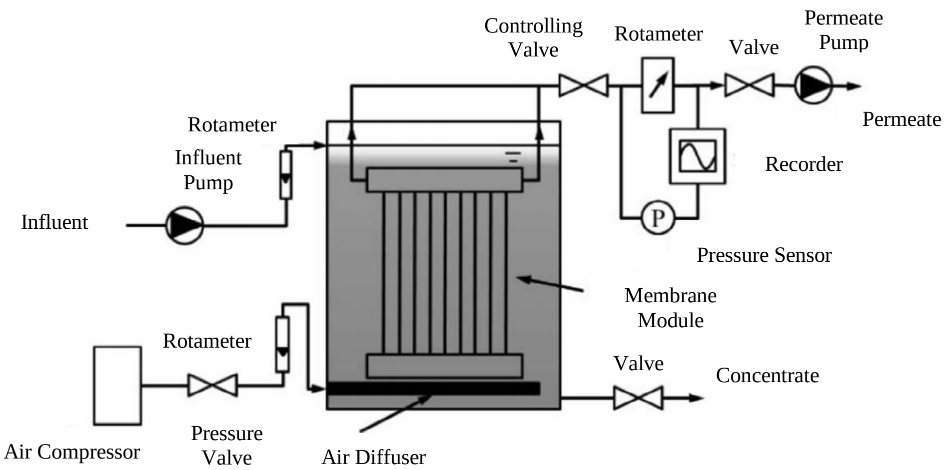
After pre-concentrate the mixed sewage, the concentrate can be transferred to anaerobic digestion for energy (biogas) and recovery of the nutrient due to its high organic load. On the other way the filtrate water can be reused after certain post-treatment depending on the application. This post treatment step is necessary if the water is released to sensitive water bodies. There is possibility to use an advanced treatment process such as membrane filtration, activated carbon for achieving the discharge or reuse water quality limits (Diamantis et al., 2010).

Considering treatment efficiency, large and small wastewater treatment facility can get the advantage of pre-concentration of the domestic sewage in terms of solid and liquid fractionation. After the solid fractionation, it is organically rich and can transfer it to the anaerobic digester for energy recovery. Filtrate water can be used in irrigation or water reclamation activity with proper post treatment. This kind of combination is more suitable for arid zones which there are not enough water to use in general activity (Diamantis et al., 2011)

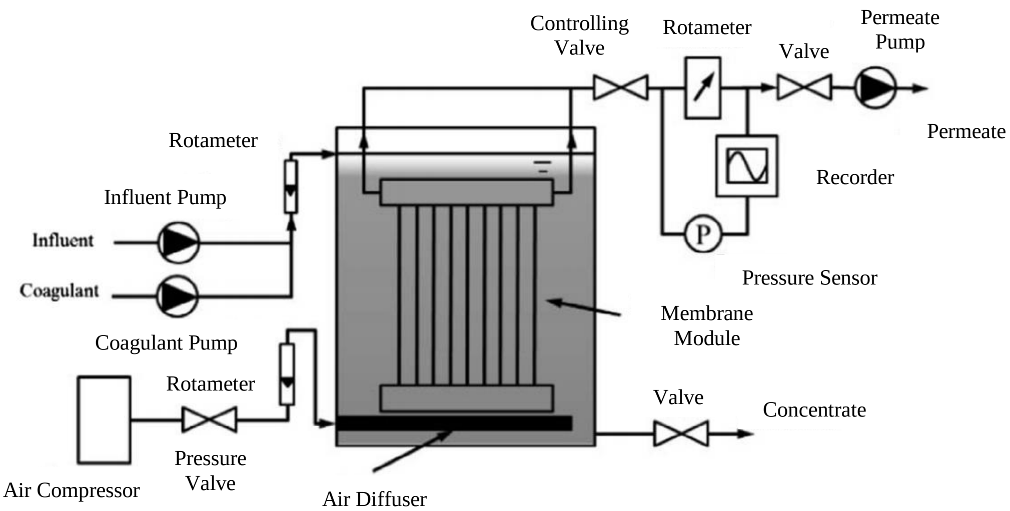
It has found that, combined coagulation microfiltration (CCM) is a suitable technology for sewage pre-concentration. With addition of coagulants enhance the concentration efficiency in sewage and reduce the fouling of the membrane. CCM can concentrate COD upto 16,000 mg/L and around 70% of total organic matter can be recovered. After concentrating the sewage, anaerobic biodegradability reached up to 56.5% (Jin et al., 2016). Schematic diagrams of the direct sewage microfiltration (DSM), continuous aerated sewage microfiltration (ASM) and combined coagulation microfiltration (CCM) shows in **Figure 2.3**.



(a) Direct sewage microfiltration (DSM)



(b) Continuous aerated sewage microfiltration (ASM)



(c) combined coagulation microfiltration (CCM)

Figure 2.3 Schematic diagrams of the (a) direct sewage microfiltration (DSM), (b) continuous aerated sewage microfiltration (ASM) and (c) combined coagulation microfiltration (CCM) (Jin et al., 2016)

2.2.1 Reasons behind the domestic sewage pre-concentration

In a conventional activated sludge process which based on aerobic biodegradation, does not focus on the energy recovery from the wastewater (Verstraete and Vlaeminck, 2011). Typically, activated sludge process needs energy to treat the COD around 35–40 kWh·IE^{-1}·y^{-1} and for the ammonia oxidation it need an extra 10–15 kWh·IE^{-1}·y^{-1}. Approximately 20 % of the energy is recovered by an anaerobic sludge digestion process (Larsen et al., 2013).

Based on the pre-concentration concept, it can generate the stream that organically rich which is the best condition to have anaerobic treatment with energy recovery. Also the domestic sewage, energy can be recovered in extremely high level and it helps to reduce the carbon footprint (Wett et al., 2007).

2.2.2 Fractionation/ pre-concentration technologies

There are many technologies in the market for separate the solid portion of the water. Most of the technologies currently use in large scale plants, but not for pre-concentrating purpose. Those technologies, mostly use for the removing solids, but not the interest with pre-concentration. Parallel plate decanters, dissolved air flotation (DAF) devices as well as various types of fine sieves can be used for this pre-concentration step. Also the chemically enhanced primary treatment (CEPT) and pre-concentration by flotation are the most common in the field. Using ultrafiltration membrane for the direct sewage filtration can be identified as an interesting process due to the support for the pre-concentration process by removing water out of the system and it produces the good quality water. There is a possibility to sewage up-concentration with a factor of 10 at constant flux of 60 LMH with TMP of 3 bar (Diamantis et al., 2011).

2.2.2.1 Chemically enhanced primary treatment (CEPT) and sedimentation

This is one of the widely used technology for increase particle separation with the addition of metal salts to raw sewage. Metal salts can destabilize the colloids and enhance the coagulation. Mostly the coagulant is an iron salt such as $FeCl_3$ or $FeSO_4$, other than that aluminum salts also can be used. Primary destabilized particles then flocculated to enhance the sedimentation process. This CEPT process helps to remove organic solids, phosphorus, heavy metals, bacteria and micro pollutants (Suarez et al., 2009). **Figure 2.4** shows the typical CEPT process flow.

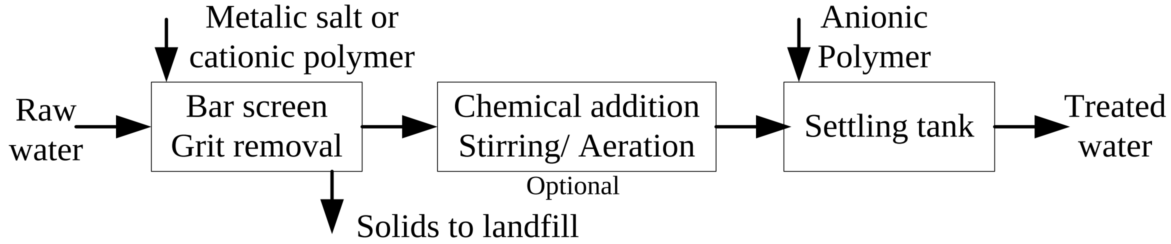


Figure 2.4 Typical CEPT process flow

2.2.2.2 Dissolved air flotation (DAF)

Dissolved air flotation (DAF) generally the small footprint treatment unit that can replace the clarification units either in the water treatment or wastewater treatment plants. It has lots of advantages over conventional sedimentation units. Generally it design for surface loading rate, 1-15 $m.h^{-1}$. DAF can achieve 70–90% COD removal efficiency with coagulation and flocculation process for domestic wastewater (Ødegaard, 2001).

Poly-aluminium chloride can use in DAF as a coagulant and it helps to two log microbial reduction and 90% of phosphorous reduction (Koivunen and Heinonen-Tanski, 2008). DAF can achieve more than 80% of particulate COD and turbidity removal By using polyelectrolytes as coagulants/flocculants (Mels et al., 2001) . **Figure 2.5** shows the schematic diagram of DAF system with recycle approach.

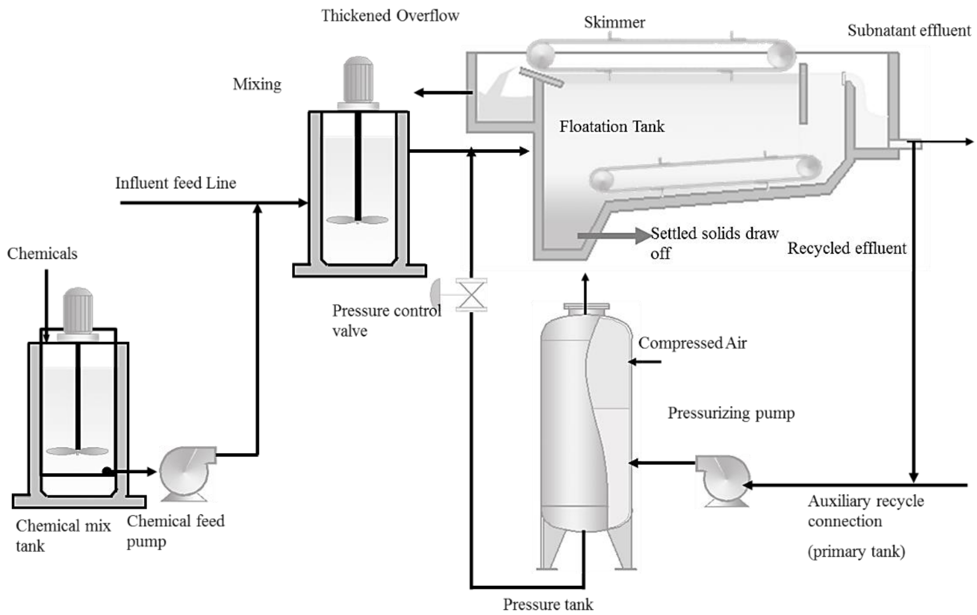


Figure 2.5 DAF with recycle approach

2.2.2.3 Advanced multi-compartment septic tanks

This is one of the common technology that use many places in the world. This technology is efficient with increasing the solid separation process from the wastewater. Sludge accumulates in the first compartment and need to be removed periodically. This compartment leads having an anaerobic condition and produce methane, which have to be controlled due to carbon footprint. In this kind of system, methane can release to the environment. **Figure 2.6** shows the schematic diagram of two compartment septic tank.

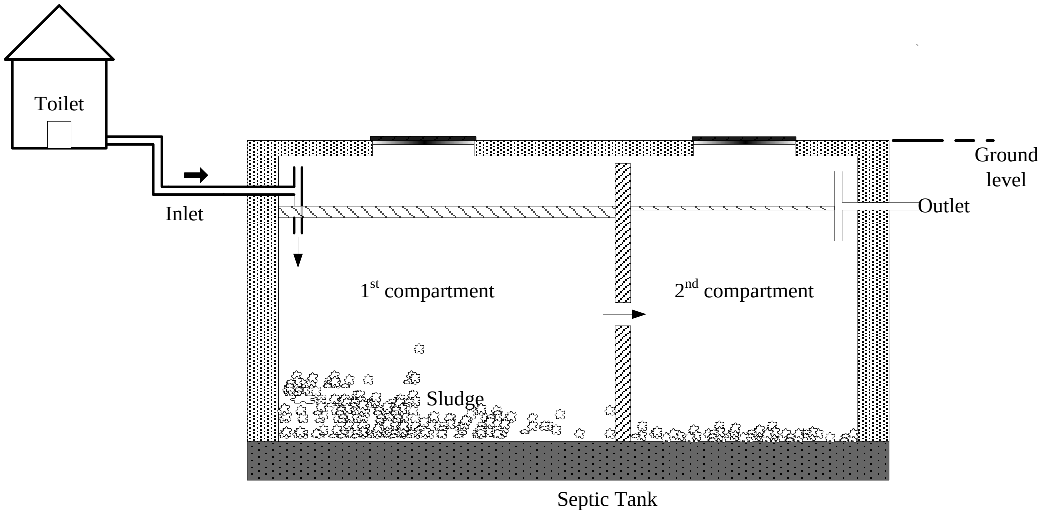
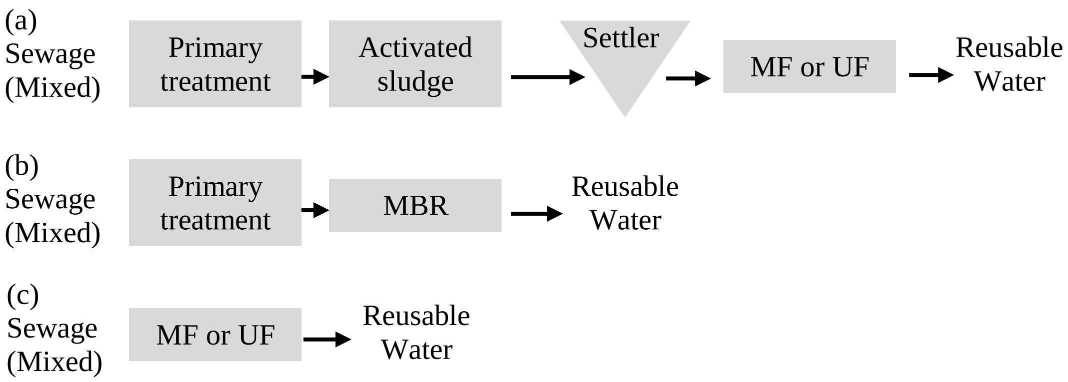


Figure 2.6 Schematic diagram of two compartment septic tank

2.2.2.4 Membrane filtration in sewage treatment

In the wastewater treatment process, there are many points that can include the membrane filtration process. Presently, membrane technology already developed to recover the water directly from the wastewater sources or septic tank. In this case, membrane filtration can act as a very good pre-concentration or fractionation technology that can lead to waste to energy option or water recovery. **Figure 2.7** illustrates the evolution of membrane technology for water recovery from sewage.



Legends
(a) Tertiary phase, (b) secondary phase, (c) primary phase.

Figure 2.7 Evolution of membrane technology for water recovery from sewage

With respect to the cost pre-concentration by using membrane technology has comparable to the conventional activated sludge process. The municipal wastewater coming through anaerobic digestion of organics and nutrient, water recovery is estimated the overall cost around €0.9/m^3. And that value is very competitive to the conventional system, but with lot of recoverable benefits (Verstraete et al., 2009).

2.3 Membrane Process

2.3.1 General characteristics of the membrane process

Membranes are selective barriers that allow specific entities to pass through while retaining others (Cheryan, 1998). Considering the wastewater treatment, the membrane filtration involves the separation of both particulate and dissolved organic matter from liquid. In the membrane field the influent to the membrane is known as feed. Moreover, liquid pass through the membrane is known as permeate and liquid retain in the feed side is known as concentrate or retentate. Permeate flux is defined as the permeate flow through the membrane. In general, membrane separation process involve in many fields and industries such as, chemical industry, pharmaceutical, power plant, water and wastewater treatment, textiles, food industry and many more.

Mainly, membrane process can be classified for four categories with respect to solid liquid separation namely: microfiltration (MF), ultrafiltration (UF), nanofiltration (NF) and reverse osmosis (RO). Membrane selectivity differentiates with pore size of the membrane, Molecular weight cutoff, hydrophilic and photophobic characteristics of the membrane. Microfiltration and ultrafiltration membrane is widely applied for wastewater treatment, especially for solid liquid separation. Typically, it is mainly used to separate suspended solids and colloidal particles by sieving mechanism. Selectivity of the membrane and the operating pressure is illustrated in **Figure 2.8**.

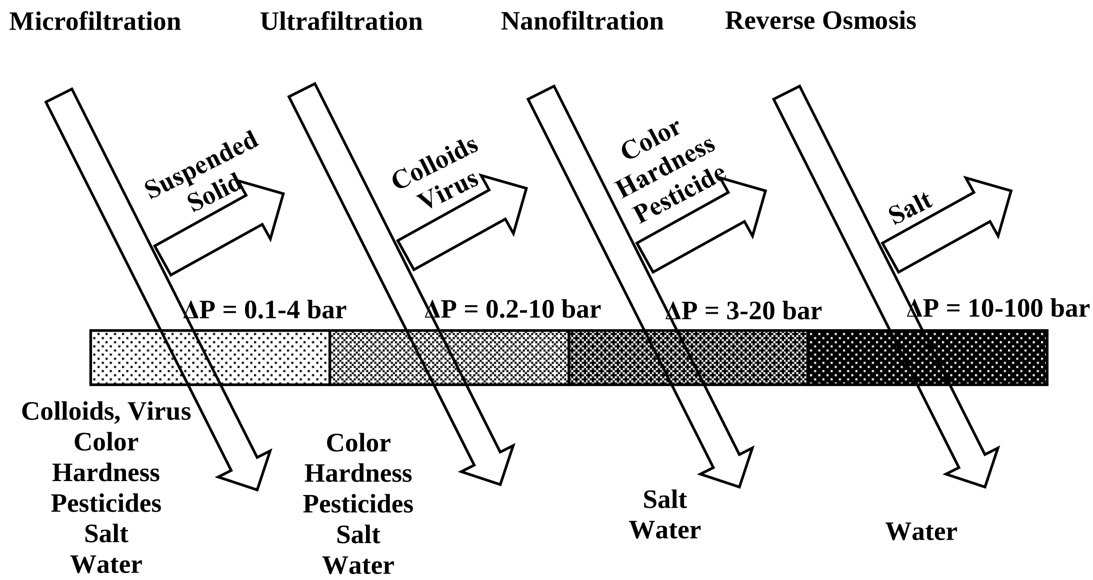


Figure 2.8 Membrane selectivity

For the wastewater treatment objective, MF and UF membrane process used to remove micron-sized particles such as suspended solids, colloids, microorganisms: for reducing the turbidity. Also, it has an ability to produce partially or fully disinfected effluent water. RO and NF membranes are semi-permeable membranes that can remove monovalent, divalent and trivalent ions also the other nutrients such as nitrogen and phosphorous. Moreover, RO and NF have an ability to remove some of trace organic matters and micro pollutants from the source water.

2.3.2 Pressure driven membrane operational configurations

Basically, membrane operation can be classified into two operational strategies such as dead-end and cross-flow filtration modes. The schematic diagrams of two membrane operational configurations are illustrated in **Figure 2.9**.

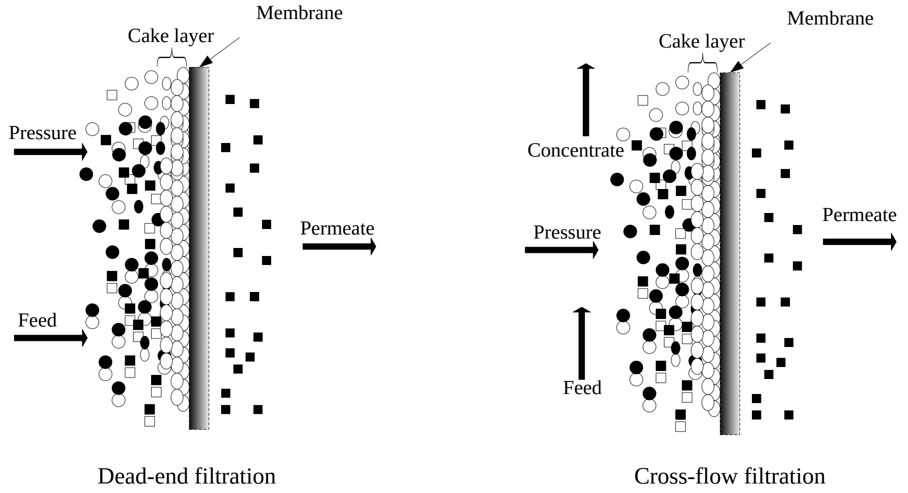


Figure 2.9 Membrane operational configurations

In dead-end configuration, feed water flow is perpendicular and it passes through the membrane. Particle that rejecting by membrane has accumulated on the membrane surface. Cross-flow operation feed water flow is tangential and rejected particles can be recirculated back to the system again as a concentrate. Cross-flow configuration can control the membrane fouling that generates from the cake layer by increasing the cross-flow velocity. **Figure 2.10** Shows the cake layer thickness and the flux changes with the time.

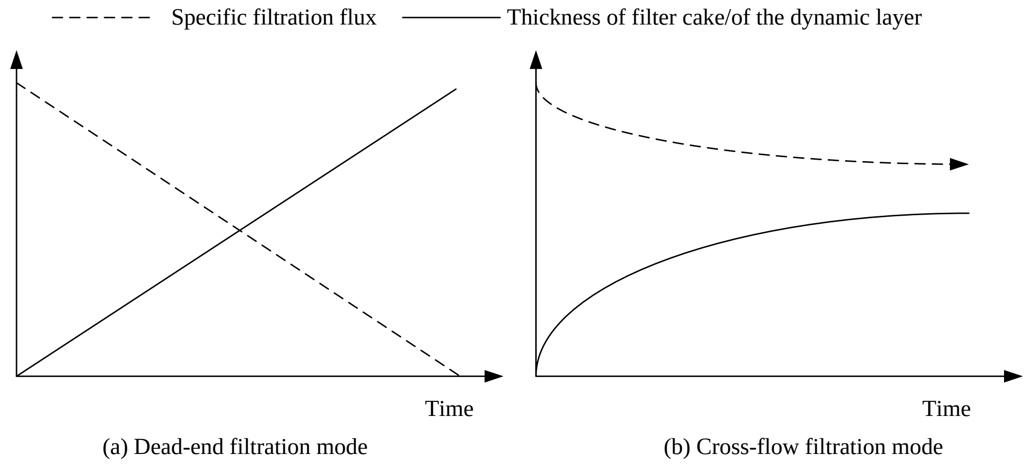


Figure 2.10 Cake layer thickness and the flux changes with the time in dead-end and cross-flow modes

Typically higher cross flow velocity helps to achieve higher flux. But the energy requirement for maintaining cross flow velocity, considerably higher than the dead-end operation. The general characteristic of membrane process is shown in **Table 2.4**.

2.3.3 Membrane materials

Membrane are manufactured with different kind of materials such as ceramics, metals, glass, polymers and many more. Membrane material, selecting for achieving effective separation with higher chemical, physical, thermal resistance and higher permeability. Depending on the polymer characteristics they can divide into two categories which are hydrophobic polymers and hydrophilic polymers. There are a number of hydrophobic polymer that use in the membrane manufacturing industry such as Polytetrafluoroethylene (PTFE), Poly vinylident fluoride (PVDF), Polypropylene (PP), Polysulphone (PSF), Poly ether sulfone (PES). Hydrophilic Polymers namely; Cellulose acitate (CA), Polyimide/Ployetherimide (PI/PEI), Aliphatic Polyimide (PA), Aromatic Polyimide (AP). Advantages and disadvantages of different membrane materials show in **Table 2.4**.

Table 2.4 Advantages and Disadvantages of Membrane Materials (Matsuo et al., 2006)

Material	Advantage	Disadvantage
PAN	Hydrophilic membrane provides higher resistance to membrane fouling. Good chemical and chlorine resistance	Relatively weaker than ceramic and PVDF
PVDF	Excellent resistance to physical and chemical deterioration	Hydrophobic membrane tends to provide low resistance to membrane fouling
PS	Higher chemical resistance and higher mechanical strength	Hydrophobic membrane tends to provide low resistance to membrane fouling
CA	Hydrophilic membrane provides higher resistance to membrane fouling	Lower chemical resistance and lower mechanical strength than PVDF or PS Easily attacked by bacteria
Ceramic	Excellent resistance to physical and chemical deterioration	High cost Wear against physical shock

2.3.4 Membrane module types

According to the membrane module configuration, the membrane can be characterized into four major groups, namely; Plate & Frame, Tubular, Spiral wound and Hollow Fiber. A qualitative comparison of those membrane configurations is presented in **Table 2.5**.

Table 2.5 Qualitative Comparison of Four Major Membrane Configurations (Visvanathan, 2016)

Characteristics	Tubular	Plate frame	Spiral-wound	Hollow fiber
Packing density	Low			Very high
Investment	High			Low
Fouling tendency	Low			Very high
Cleaning	Good			Poor
Operation cost	High			Low
Membrane replacement	Yes/No	Yes	Yes	yes

Figure 2.11 shows the main four membrane configurations currently in the market.

(a) Tubular (b)Plate & Frame

(c) Hollow Fiber (d) Spiral wound

Figure 2.11 Membrane module types

2.3.5 Membrane operational parameters

There are three main parameters, use in membrane operation, namely; Trans-membrane pressure (TMP), the permeate flux (J) and filtration resistance (R). The relationship between these operating parameters is given in the following equation.

$$J = \frac{\Delta P}{\mu R_t} \qquad \textbf{Equation 2.1}$$

$$R_t = R_m + R_c + R_f \qquad \textbf{Equation 2.2}$$

Where:

J	=	Permeate flux ($L/m^2.h$)
ΔP	=	The pressue drop accross the membrane as *Trans-membrane Pressure* TMP (kPa)
μ	=	Permeate viscosity (Pa.s)
R_t	=	Total membrane resistance (1/m)
R_m	=	Intrinsic membrane resistance (1/m)
R_c	=	Cake resistance (1/m)
R_f	=	Membrane resistance caused by adsorption of solute (1/m)

2.3.6 Process control strategy in membrane filtration

Typically, membrane operation based on flux or pressure constant mode. **Figure 2.12** shows the typical pattern of changing the flux and pressure with the time while one of them is constant.

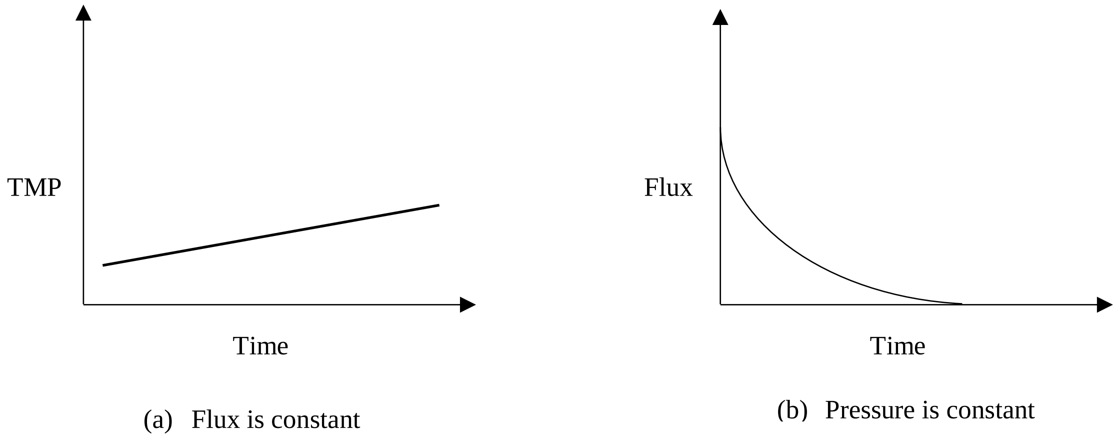


Figure 2.12 Typical pattern of changing the flux and pressure with the time

Generically, in water and wastewater treatment field, mostly use the (a) constant flux method. In this situation, it can help to fix the quantity of water production. In another way, membrane can clean when it reaches to certain pressure level.

2.3.7 Outside-in and in-side out operation

In submerge membrane process, there are two categories of operation mode. This mode depending on the feed water quality and the membrane material specification. **Figure 2.13** shows the flow pattern of both the modes.

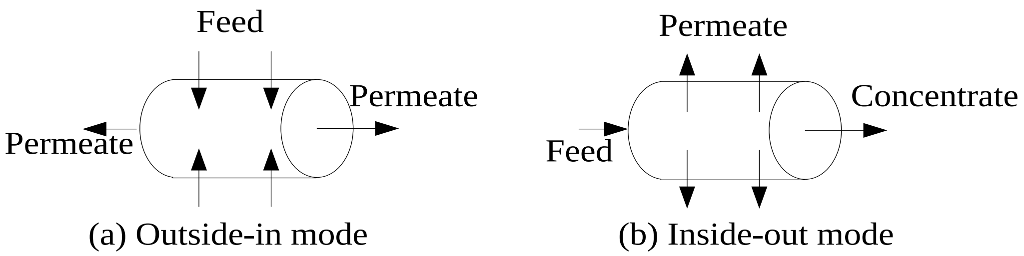


Figure 2.13 Outside-in and in-side out operation

2.3.8 Membrane fouling

The most important limitation in membrane technology is the fouling (Madaeni, 1999). Membrane Fouling can explain as irreversible deposition of material on the membrane, causing reduction of flux and rejection. The flux reduction is caused by pore clogging and/or by the cake layer formation on the membrane surface (Cabassud et al., 1991). As a result of membrane fouling, the resistance can increase and it leads to reduce the membrane flux. Membrane fouling can occur due to major three reasons; include pore narrowing, pore plugging, and gel/cake formation as shown in **Figure 2.14**.

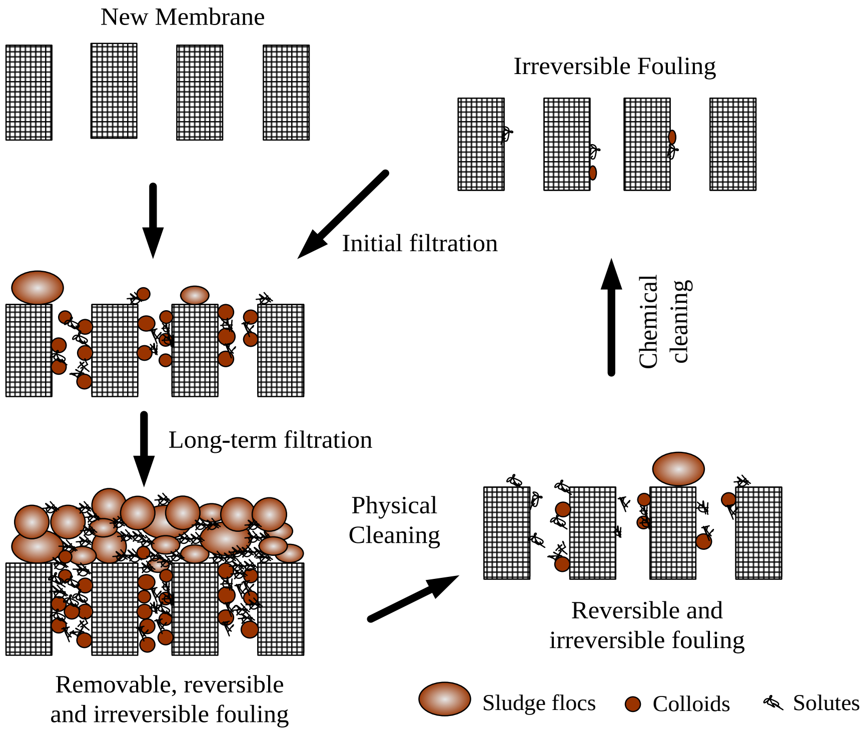


Figure 2.14 Membrane fouling and cleaning (Gkotsis et al., 2014)

Basically potential foulants categorize into four categories such as: organic, inorganic materials like minerals, microbial content like bacteria and the colloidal content like clay particles. Moreover, the extra polymeric substances that produce from microorganisms, have a major effect for the membrane fouling (Hu et al., 2013). Based on the removability of the foulants, membrane fouling can be classified into reversible fouling and irreversible fouling. Fouling that can be removed by physical cleaning method such as backwashing, defined as reversible fouling. Generally, irreversible fouling cannot be removed by simple physical cleaning. In this situation, chemical cleaning need to be done (Field, 2010). **Figure 2.15** shows the reversible and the irreversible flux with respect to operation time.

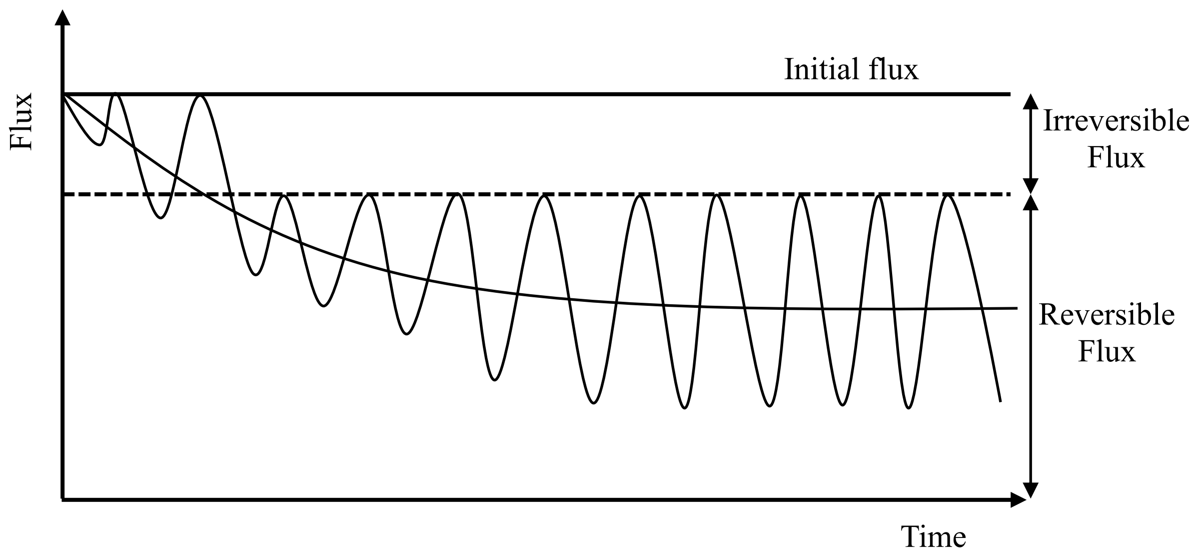


Figure 2.15 Reversible and irreversible flux (Visvanathan, 2016)

2.3.9 Membrane cleaning

Membrane cleaning needs to done for removing the reversible and irreversible fouling. Generally, particle fouling can be removed by physical cleaning method such as water jet application which is not using any chemical. Back flushing also can consider as one of the physical cleaning methods. Sometimes, depending on the material, it can use mechanical scrubbing like brush washing. Solar dying cleaning method can practice for the biofouling removal in woven fiber micro filtration flat sheet membranes (Vongsayalath, 2015).

When the membrane is hardly fouled with irreversible fouling, physical cleaning may not enough for the permeate flux recovery. In this situation, chemical cleaning is necessary for restoring the membrane flux. This cleaning, can be done with two different scenarios depending on the application.

- CIP - Clean In Place
- COP - Clean Out of Place

In the CIP method, membrane can be directly cleaned with chemical reagents without removing it from the system. Hence, in the COP method removing the membrane out of the system and clean separately with the chemicals. It is important to select the correct type of the chemical for this cleaning activity. Selecting a scenario depend upon several factors such as chemical concentration, cleaning time and intervals, the chemical resistance of the particular membrane and also depend on what type of fouling and degree of fouling. Mostly, low concentrated acid and alkali are used for the membrane cleaning depend upon the above mentioned factors.

2.4 Woven Fiber Microfiltration Membrane (WFMF)

2.4.1 Membrane material and the history

Membrane material for the application is, polyester woven fiber microfiltration (WFMF) fabric produced by Gelvenor, South Africa. The woven fiber material is robust and it can achieve a good turbidity rejection performance (Pillay and Jacobs, 2005). Originally, the material is a fabric and scanning electron microscope (SEM) image shows the material arrangement in **Figure 2.16**. Based on this structure, it is difficult to define the pore size of the woven fiber fabric. However, research work on the woven fiber material has shown that the it has an ability of removing particles down to 0.1 μm, particularly when it is pre-coated (Pillay, 2011).

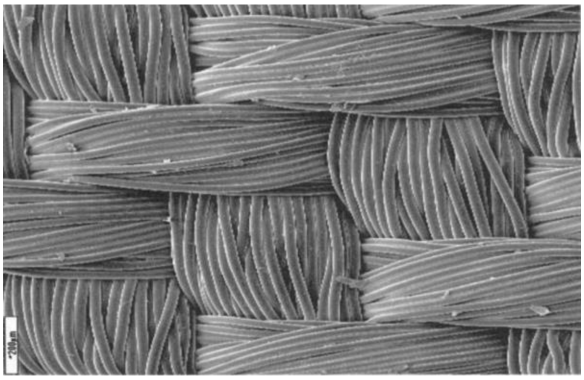


Figure 2.16 Scanning electron microscope (SEM) image of the woven fabric

2.4.2 Pore size distribution

But, recent research identified that woven fiber fabric has effective pore 1 to 3 microns. Pore size distribution of woven fabric is illustrated in **Figure 2.17**.

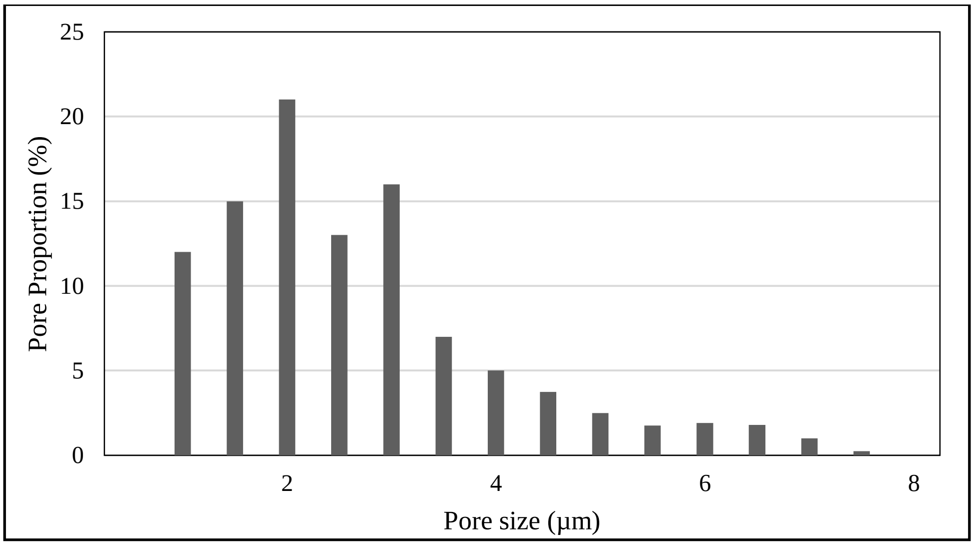


Figure 2.17 Pore size distribution of woven fabric (Kuhr et al., 2014)

2.4.3 Development of the module configuration

Woven fabric is available in two forms in the market as flat sheets or a tubular array namely as "curtain". **Figure 2.18** shows the image of the available configurations.

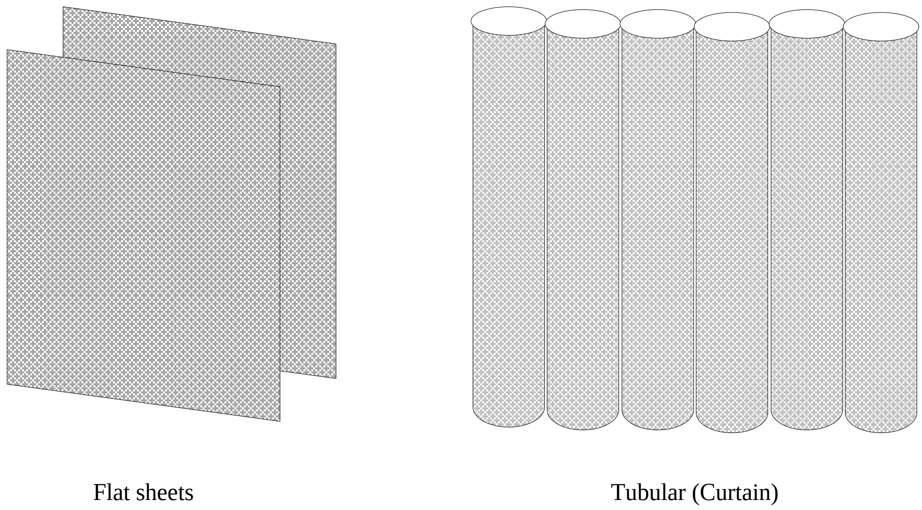


Figure 2.18 Available configuration of woven fabric in the market

Flat sheet modules has packing density than the tubular models. In that case technology developed with the flat sheet models. The first generation flat sheet model consists of a rectangular PVC frame onto which the woven fiber fabric is glued on both sides. Moreover, the spacer is introduced in between the membrane flat sheet to permeate to flow through to the outlet point. Also, the spacer helps to keep the membrane without touching each other while it has a suction pressure in operating phase (Pillay, 2010). **Figure 2.19** shows the first generation woven fiber flat sheet membrane module.

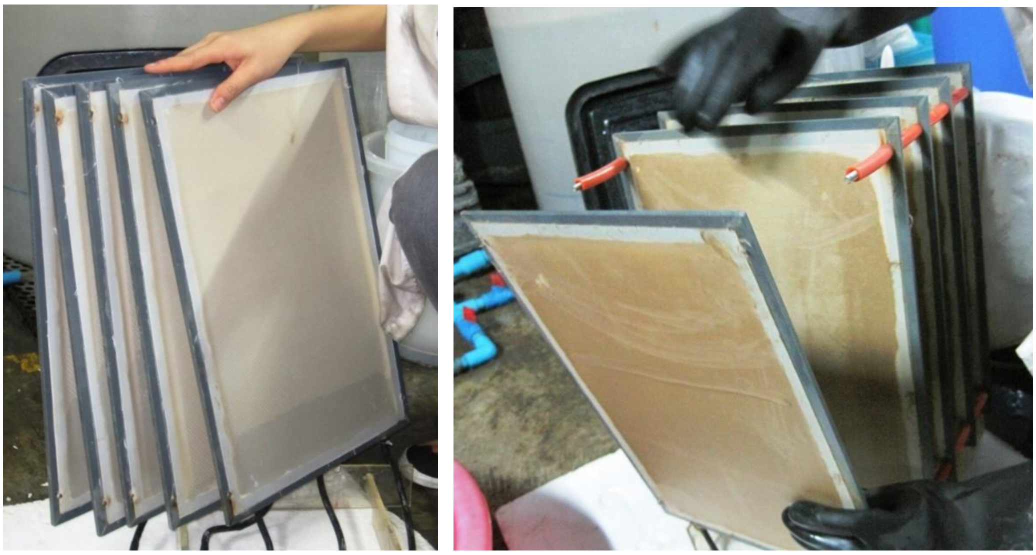


Figure 2.19 First generation woven fiber flat sheet membrane module (Thuy, 2010)

However, after operating the first generation flat sheet module, there were some issues related to the membrane flux, the glue that use for attaching the membrane and the permeate outlet port. The second stage flat sheet designed for overcoming the above mentioned issues. **Figure 2.20** shows the second stage membrane flat sheet design.

Figure 2.20 Second stage membrane flat sheet design (Vongsayalath, 2015)

2.4.4 Woven fiber membrane cleaning

Physical and chemical cleaning can be done depending the application and the fouling situation. Reversible fouling can be removed by physical cleaning methods. Relatively, physical cleaning methods for the woven fiber membranes are easy as it doesn’t use the chemicals. Researchers have found two effective physical cleaning method namely:

I. Spray brush method
II. Solar dry method

It has found that operating TMP more than 60 kPa is economically viable and operationally good. So that the membrane cleaning point consider as 60 kPa for the wastewater treatment practices in WFMF. The spray brush method use a simple method that cleaning the membrane using brush while spraying water. Solar drying method identified as an effective cleaning method for the biofouling removal in woven fiber flat sheet applications. The bio fouled later can be peeled off from the membrane once the membrane module keeps three days under the sunlight for solar drying. A simple brushing can remove the peels off once it is dry. **Figure 2.21** shows the pictorial view solar drying, cleaning performance.

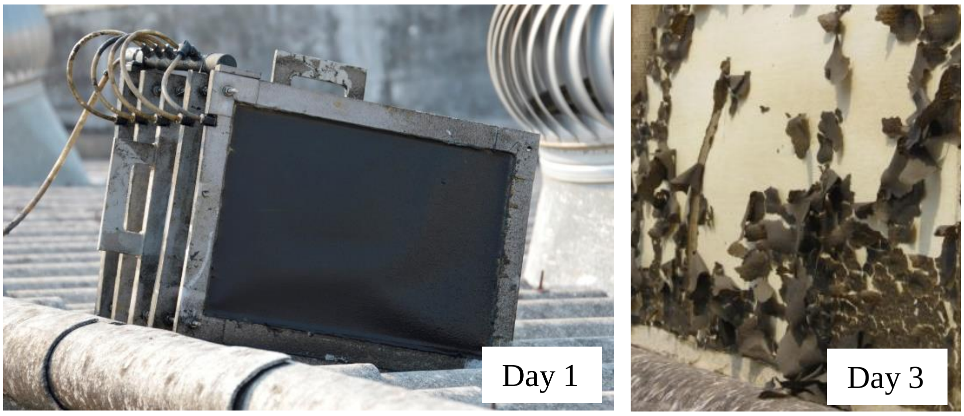


Figure 2.21 Pictorial view of solar drying cleaning performance

Chemical cleaning also can be done for the woven fiber micro filtration membranes. 0.03 % NaOCl solution can be used for this method. For the best cleaning performance, it need to dip the module in the solution for 8 hours. After a chemical dip, it needs to brush both the side of the membrane with clean water. Finally, tap water filtration can be used to evaluate the chemical cleaning performance (Thuy, 2010).

2.4.5 Current situation of the woven fiber membrane technology

Currently woven fiber membrane technology developed up to immersed membrane bioreactor (IMBR) applications. Woven fabric membrane are very attractive in IMBRs due to many reasons such as,

- Immersed membrane module can fabricate with based locally available woven fabric.
- WFMF membranes are robust, subjected to extreme physical and chemical conditions.
- Suitable for developing economies or lower operator skill.

The test that carried out at Veolia water reclamation plant in Durban, identified the WFMF-IMBR, can remove 100% of the MLSS in the activated sludge process. That kind of performance is level to the commercial grade IMBRs (Pillay and Cele, 2014). Considering the permeate flux, woven fiber immersed membrane microfiltration (WFIMMF) system shows the general performance characteristics same to the commercial IMBRs. Typically, commercial IMBR units can generate 20- 40 LMH flux with the MLSS of 12-15 g/L. WFIMMF shows the 10-15 LMH with MLSS of 9 g/L. For this study, air flow rates ranged from 0.3-2.5 L/min/module which is 5 -35 % of commercial IMBRs air flow supply (Cele et al., 2015).

2.5 Tube Settlers

Removing or separation of total suspended solids (TSS) from water is one of the major problem in water and wastewater treatment sector. One of the major technique is the sedimentation which consumes the major amount of the total capital expense of the treatment plant, in terms of the chemical use for coagulation. There are different kind of attempts have been taken to reduce the cost of sedimentation. One of the effort is direct filtration technique which is not applicable for the wastewater treatment or the high turbid water. Some techniques can reduce the size and the cost of sedimentation process. By using of high-rate sedimentation, reduces the hydraulic retention time in the settling tank by reducing the distance necessary for the particles to settle. These systems are generally tubes, parallel plates which are placed inclined at some angle to the horizontal.

Typically, the conventional rectangular settling tanks having hydraulic retention time of two hours or more. But sedimentation tanks incorporated with tube settler can achieve the detention times of 15 minutes or less. Tube settlers consist with multiple tube channels sloped at an angle between 45 to 6o and adjacent to each other. It helps to increase the settling area. Also, it provides the significantly less, particle settling depth than the conventional sedimentation tanks which helps to reduce the settling time (Metcalf and Eddy, 2003). Tube settlers help to remove the settleable fine floc and allows the larger floc to settle to the tank bottom in a more efficient way. Tube installed in a rectangular sedimentation tank, illustrates in **Figure 2.22**. Settling capacity can expand in the new or existing sedimentation basins, clarifiers by introducing the tube settlers. There are main three advantages of the tube settler.

1. Sedimentation basin can be designed much smaller because of increased flow capacity and the area.
2. Can be increased the flow capacity of the existing systems through the introducing of tube settlers.
3. The effluent quality can be improved by introducing the tube settlers to the sedimentation basin.

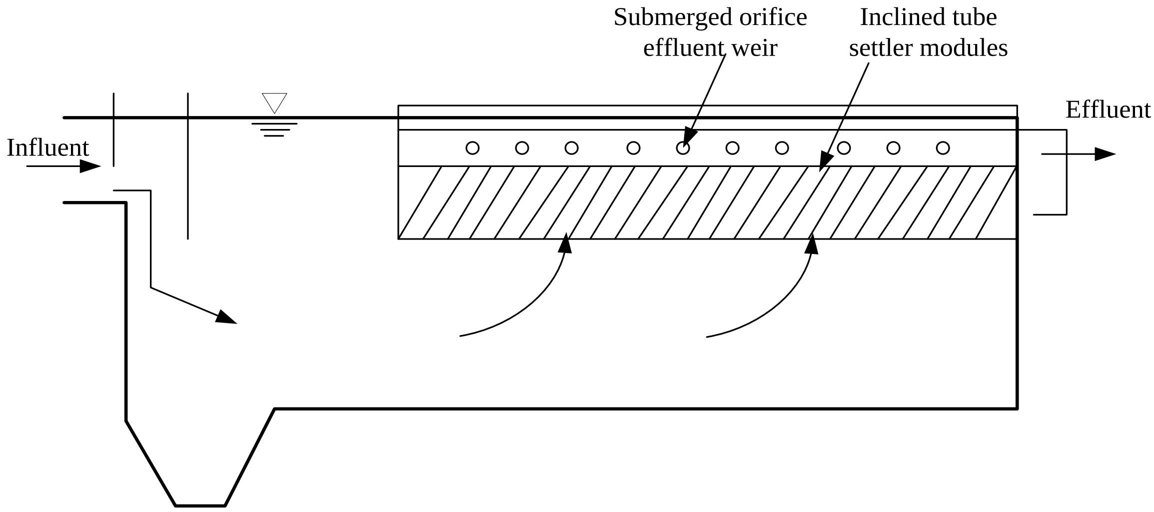


Figure 2.22 Tube installed in a rectangular sedimentation tank (Metcalf and Eddy, 2003)

There are many shapes and configurations that can use for the tube settler. Typically, the circular, hexagonal, diamond and square shaped tubes are using in tube settlers. There are two basic configurations of tube settlers, the "horizontal" and the "steeply inclined." Horizontal tubes have an angle of inclination, less than 7.5°, while steeply inclined tubes have any angle up to 60° (Fadel, 1985) .

Horizontal Tube Settlers settling angle is slightly inclined (50) in the direction of the flow. Also, requires frequent cleaning to wash down the accumulated sludge. This type of configuration has the lowest construction cost compared to the steeply inclined tube settlers. Due to this complex cleaning process, this type of installations is not advocated for large water treatment plants - limited only for small water treatment plants which have the capacity of 1-2 MGD. Inclined tube settlers has 45^0 - 60^0 of angle of inclination. At $\theta = 60°$ would effectively double the maximum fall distance for particle entering the tube (Visvanathan, 2015) .

Primarily, the efficiency of a sedimentation basin depends on particle settling distance and the overflow rate (Fadel and Baumann, 1990). The shape and configuration of each tube should be designed to give a low “Reynolds number” (<200) and laminar flow conditions, for increase the accumulation of the TSS through the tubes. Laminar flow is the most important design criteria for the optimal design and efficient operation of a tube settler system (Hendricks, 2010).

Tube settler design is based on following design criteria:

1. Flow (m^3/h): Required flow capacity through the basin
2. Area (m^2): tube settlers total area
3. Loading rate: Flow/Area

Typically, an overflow rate of 7.3 m^3/m^2.h is acceptable for the basin area covered by the tube settlers, when next unit is either dual or mixed media filters (Visvanathan, 2015). By applying to the wastewater treatment, tube settler can achieve the average removal of 97.6% TSS, 96.4% BOD_5, and 96.36% COD with 20 minutes HRT respectively (Faraji et al., 2013).

2.6 Water Reuse

Wastewater reuse is accepted as a principle in most developed and developing countries. In developing countries, generally, the wastewater reuse is mainly applied for agricultural activities. Generally, reuse of domestic wastewater occurs in regions where the water demand is high and the supply is low. In common, treated wastewater is reused for non-potable purposes such as toilet flushing, horticulture or the agricultural irrigation in certain conditions.

In developing countries, wastewater is often reused, without treatment. This practice comes with a huge health risk. For the better practice, secondary treatment of the wastewater is recommended. This process can be a combination of sub-processes such as sedimentation, filtration, chemical clarification, adsorption, membrane filtration, ion exchange, disinfection and many others. Microbiological safety is the most important factor when considering the health risk of reusing treated wastewater. **Table 2.6** highlights the microbiological criteria for different applications of wastewater reclamation.

Table 2.6 Microbiological Criteria for Different Applications of Wastewater Reclamation (Davis and Hirji, 2003)

Application	Fecal coliforms (per 100 mL)
Irrigation (restricted)	No standards recommended
Irrigation (unrestricted)	< 1000*
Aquaculture	< 1000* (measured in the fish ponds)
Landscape irrigation	< 200*
Groundwater recharge	23**
Non-potable urban use	3-1000**
Recreation	2.2-1000**
Drinking water	Must not be detectable*

* WHO standards
** USA-EPA standards

The major pathways of water reuse incorporate irrigation, surface water replenishment, industrial use and groundwater recharge. **Figure 2.23** shows the connection between the natural water cycle and the reuse options can explain through the engineered hydrologic cycle. The engineered systems coupled with recycling and reuse, wastewater reclamation plays a major role in the hydrologic cycle. The major pathways of water reuse are included irrigation, groundwater recharge, surface water replenishment and industrial use.

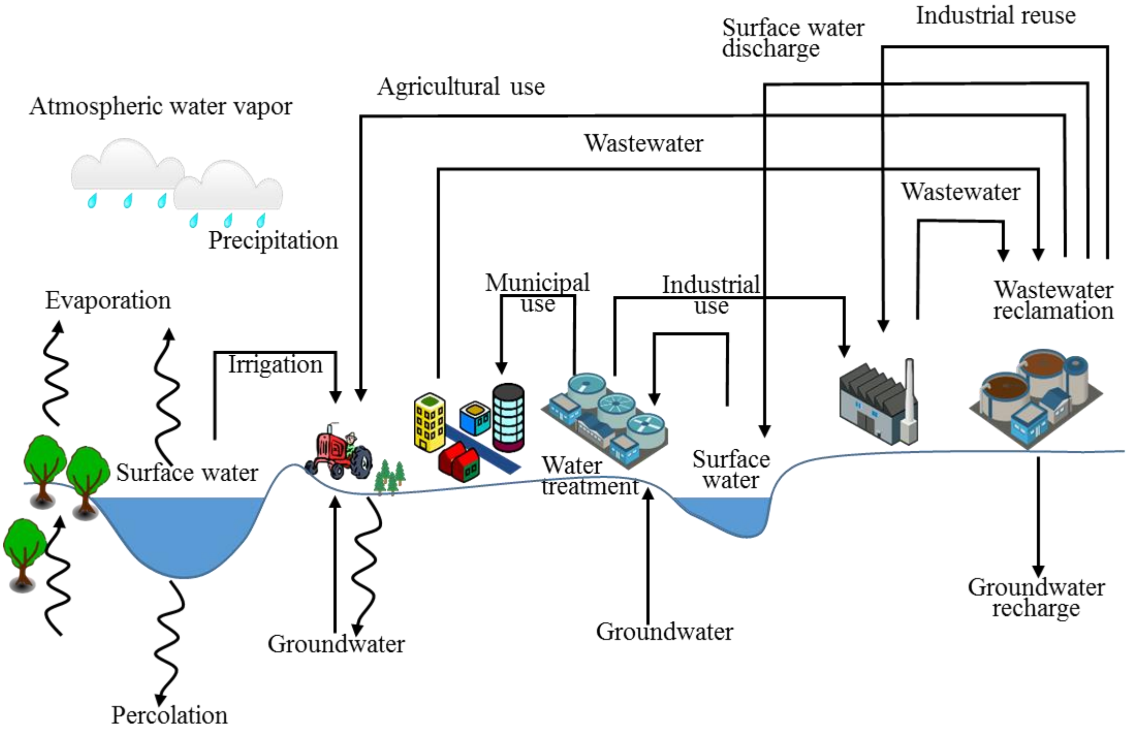


Figure 2.23 Water reuse facilities through hydrological cycle (Asano and Levine, 1996)

In some situations, wastewater is used for irrigation of crops. But this lead to accumulate of the toxic substances such as heavy metals in the soil. Wastewater should have a proper treatment, especially it reuse for crop irrigation. **Table 2.7** shows an overview of various national standards for wastewater reused for irrigation of food crops for human consumption.

Table 2.7 Overview of Various National Standards for Wastewater Reused for Irrigation of Food Crops for Human Consumption (Davis and Hirji, 2003)

Parameter	Unit	Saudi Arabia	Jordan	USA
BOD	mg/L	10	150	30
COD	mg/L	-	500	na
TSS	mg/L	10	200	30
Oil and grease	mg/L	absent	8	na
pH		6.0-8.1	na	6-9
Chlorine residual	mg/L	na	0.5	1

na = not available

Agriculture and landscape, Industrial recycling and reuse, Groundwater recharge, Recreational/environmental uses, Non-potable urban uses, Potable uses are the main categories of wastewater reuse. But this Municipal wastewater reusing options come with the potential issue or constraints. Ground and Surface water pollution, Public health concerns related to pathogens, Effect of water quality, Toxicity to aquatic life can be the major potential constraints. Other than that, compounds in reclaimed wastewater cause scaling, corrosion, biological growth, and fouling are the major issues when it comes to the industrial reuse. When the ground water recharge, organic chemicals in reclaimed wastewater and their toxicological effects can be the major issue. Due to the risk factors of using reuse wastewater, many organizations have the guideline to consume reuse water. For example, **Table 2.8** shows the EPA suggested guidelines for water reuse.

Table 2.8 EPA Suggested Guidelines for Water Reuse [a] (Asano et al., 2007)

Level of treatment	Types of reuse	Reclaimed water quality	Reclaimed water monitoring	Setback distances
Disinfected tertiary [b]	• Urban reuse [c] • Food crop irrigation • Recreational impoundments	• pH = 6–9 • ≤10 mg/L BOD_5 • ≤2 NTU • No detectable fecal coli/100 mL • ≥1 mg/L Cl_2 residual	• pH- weekly • BOD- weekly • Turbidity- continuous • Coliform- daily • Cl_2 residual- continuous	• 15 m to potable water supply wells [d]
Disinfected secondary		• pH = 6–9 • ≤30 mg/L BOD_5 • ≤30 mg/L TSS • ≤200 fecal coli/100 mL • ≥1 mg/L Cl_2 residual	• pH- weekly • BOD - weekly • TSS- daily • Coliform- daily • Cl_2 residual- continuous	• 90 m to potable water supply wells • 30 m to areas accessible to the public

[a] Adapted from U.S. EPA (2004).
[b] Filtration of secondary effluent.
[c] Uses include landscape irrigation, vehicle washing, toilet flushing, use in fire protection and commercial air conditioners.
[d] Setback increases to 150 m if impoundment bottom is not sealed.

2.7 Anaerobic Digestion (AD)

Conversion of the organic materials and the pollutants by anaerobic process is a well-established technology for waste and wastewater treatment. Bio gas, a mixture of methane and carbon dioxide, which is the end product of the anaerobic process. Biogas consider as the renewable energy source. AD technology is a simple process that consumes low energy and can be applied to a wide range of wastewater types. One of the major concern is in the environmental field is greenhouse gas emission. AD can potentially diminish the CO_2 emission by using the biogas as a renewable energy.

2.7.1 Principle of the AD process

In simply, the AD is a process that anaerobic microorganisms take energy from the organic matter and grow by metabolizing it in an oxygen free phase resulting the production of biogas. Basically, the complete anaerobic process can be categorized in to four major phases namely;

1. Hydrolysis: conversion of non-soluble biopolymers to soluble organic phase
2. Acidogenesis: conversion of soluble organic compounds to volatile fatty acids (VFA) and carbon dioxide (CO_2)
3. Acetogenesis: conversion of volatile fatty acids to acetate and hydrogen (H_2)
4. Methanogenesis: conversion of acetate and CO_2 and H_2 to methane gas (CH_4)

A stepwise presentation of organic matter degradation by anaerobic digestion is shown on **Figure 2.24**.

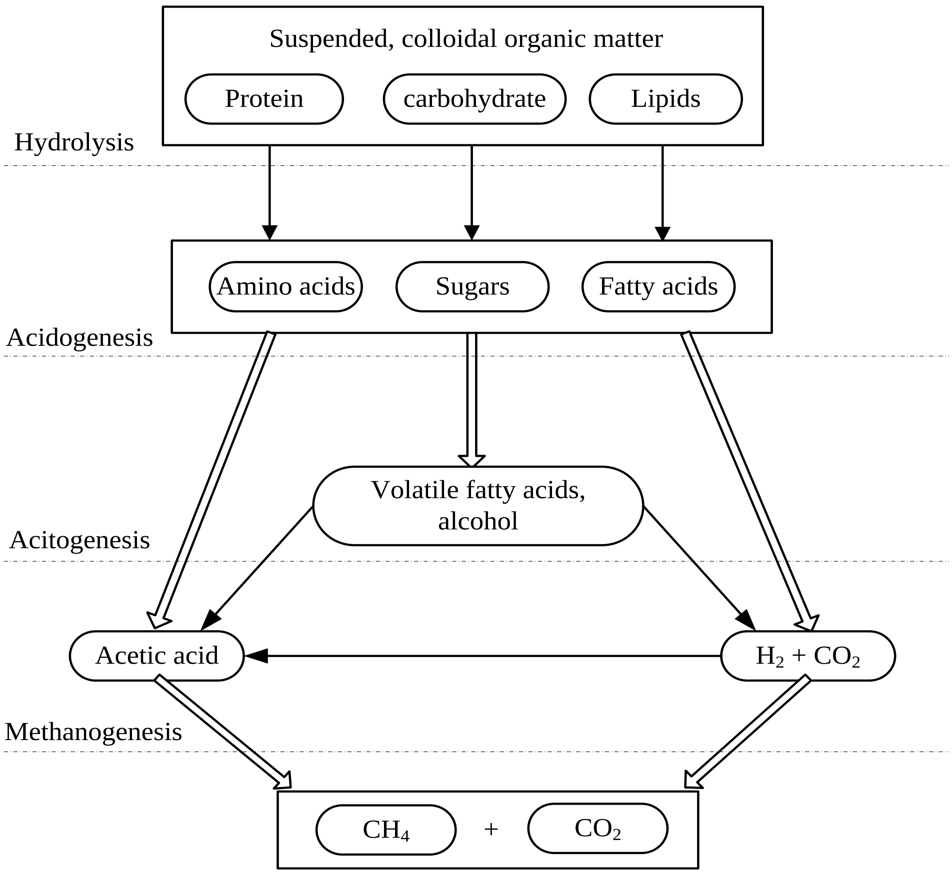


Figure 2.24 A stepwise presentation of the anaerobic digestion (De Mes et al., 2003)

The first stage, hydrolysis bacteria transform the suspended organic matters into soluble organics. The proteins, carbohydrates and fats are converted to amino acids, sugar (monosaccharides) and the fatty acids. In acidogenesis phase, the acidogenic bacteria convert the products of the hydrolysis reaction into volatile acids, ketones, alcohols, hydrogen and the carbon dioxide. Acetogenesis is the third reaction that VFA converted in the acetic acid, hydrogen, carbon dioxide. The final step is methanogenesis that hydrogen and acetic acid convert into methane gas and carbon dioxide. The reactions involve in the complete process presents in **Table 2.9**.

Table 2.9 Reactions Involve in the AD Process (Ostrem et al., 2004)

AD Process	Reactions
Hydrolysis	$C_6H_{10}O_4 + 2H_2O \rightarrow C_6H_{12}O_6 + 2H_2$
Acidogenesis	$C_6H_{12}O_6 \leftrightarrow 2CH_3CH_2OH + 2CO_2$ $C_6H_{12}O_6 + 2H_2 \leftrightarrow 2CH_3CH_2COOH + 2H_2O$ $C_6H_{12}O_6 \rightarrow 3CH_3COOH$
Acetogenesis	$CH_3CH_2COO^- + 3H_2O \leftrightarrow CH_3COO^- + H^+ + HCO_3^- + 3H_2$ $C_6H_{12}O_6 + 2H_2O \leftrightarrow 2CH_3COOH + 2CO_2 + 4H_2$ $CH_3CH_2OH + 2H_2O \leftrightarrow CH_3COO^- + 2H_2 + H^+$
Methanogenesis	$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O$ $2C_2H_5OH + CO_2 \rightarrow CH_4 + 2CH_3COOH$ $CH_3COOH \rightarrow CH_4 + CO_2$

2.7.2 Advantages of anaerobic digestion

2.7.2.1 Energy consideration

AD can be explained as an energy producing process rather than an energy consuming process. In aerobic process, it needs the aeration for the degradation process, hence the anaerobic process don't need any aeration for conversion of the organic matter. Energy balance conversion for high strength waster at 20 ºC, explain in the **Table 2.10** for anaerobic and aerobic process.

Table 2.10 Comparison of Energy Balance of Aerobic and Anaerobic Processes (George et al., 2003)

Energy	Value, kJ/d	
	Anaerobic	Aerobic
Aeration		-1.9×10^6
Methane produced	12.5×10^6	
Increase wastewater temperature to 20 ºC	-2.1×10^6	
Net Energy, kJ/d	10.4×10^6	-1.9×10^6

Note- Oxygen required = 0.8kg/kg COD removed
Aeration efficiency= 1.52 kg/ kWh and 3600 kJ= 1 kWh
Methane production= 0.35 m^3/ kg COD removed
Energy content of methane= 35,846 kJ/m^3 (at 0 ºC and 1 atm)
(Condition- wastewater flowrate= 100 m^3/d, Wastewater strength= 10 kg COD/m^3 and T=20 ºC)

2.7.2.2 Lower biomass yield and fewer nutrient requirement

Based on the anaerobic process energetics, it is producing lower biomass by a factor of 6 to 8 times. In that case, less sludge volume can be results and it is one of the advantage for the disposal of the sludge. Moreover, an anaerobic process nutrient addition is less because of the lower biomass content.

2.7.2.3 Higher volumetric loadings

Compared to the aerobic process, AD can achieve higher volumetric loading rates. In that case, reactor volume can be reduced with the plant footprint. Typically, anaerobic process can use organic loading rate of 3.2 to 32 Kg COD/m^3.d while aerobic process uses 0.5 to 3.2 kg COD/m^3.d. Generally, it needs the 1500-2000 mg/L COD concentration to produce enough quantity of methane for rising up the wastewater temperature without additional fuel. If the COD 1300 mg/L or less, aerobic treatment will be the better option (Metcalf and Eddy, 2003). Low temperature and the low strength organic loading rate are the main bottlenecks for direct anaerobic digestion of sewage. This will course the energy waste in anaerobic process and the leads ineffective situation (Jin et al., 2016). In this situation, pre-concentration of the domestic sewage is more important step to do the anaerobic digestion. **Equation 2.3** shows the relationship between the OLR and the reactor volume. OLR is an important control parameter in AD systems.

$$\mathrm{OLR} = \frac{Q \times S}{V} \qquad \textbf{Equation 2.3}$$

Where S = Substrate concentration ($kg_{substrate}$ in terms of TVS)
Q = Influent flowrate (m^3/d)
V = Reactor volume (m^3)
OLR = organic loading rate ($kg_{substrate}/m^3$.d)

2.7.3 Factors affecting anaerobic wastewater treatment

Anaerobic digestion is strongly influenced by environmental factors as it's a biological process. pH, Temperature, alkalinity and the toxicity are the main controlling factors in the AD process.

2.7.3.1 Temperature

There are three major temperature ranges that anaerobic process can work, namely; psychrophilic (10-20 ºC), mesophilic (20-40 ºC), or thermophilic (50-60 ºC). Typically, the microbial growth and organic conversion processes are slower in low temperature conditions. Considering the operation, psychrophilic digestion needs a much higher retention time and large reactor volumes. Compared to psychrophilic digestion, Mesophilic digestion requires lower reactor volume. Thermophilic digestion is especially suitable when the wastewater is discharged at a high temperature (De Mes et al., 2003). Generally, every 10 °C increment in temperature can be double the anaerobic microbial activity within the range of optimum temperature (Chaikasem, 2015).

2.7.3.2 pH and alkalinity

The optimum pH of acetogens and acidogens steps are 5.5-7.2 respectively. Also, methanogens reaction needs a pH range of 6.8-7.8 and it has a narrow operating pH range (Visvanathan and Abeynayaka, 2012). Alkalinity is one of the major factors that effect for the AD process. It has identified that the 2000 to 3000 mg/L as $CaCO_3$ alkalinity need for the maintaining pH high has phase carbon dioxide concentration (Metcalf and Eddy, 2003). If the influent wastewater or a degradation process cannot supply this range of alkalinity, then it need to externally to the correct amount of alkalinity.

2.7.3.3 Nutrient

To support the new biomass synthesis, anaerobic microorganisms need macronutrients such as nitrogen and the phosphorus. Also the micronutrient such as some trace elements. Moreover, it needs to maintain the C: N: P ratio as 100:5:1 for the new biomass synthesis. If the source water doesn't contain enough nutrient, it must be a need to add those additionally. **Table 2.11** shows the minimum requirement of trace elements and toxic concentration for the anaerobic process.

Table 2.11 Minimum Requirement of Trace Elements and Toxic Concentration for the Anaerobic Process (Chaikasem, 2015)

Substances	Minimum requirement of trace element (mg/L)	Inhibition concentration start (mg/L)		Toxicity (mg/L)
		Free ions	As carbonate	
Cr	0.005-50	28-300	530	3 (Cr^{+3}), 500 (Cr^{+6})
Fe	1-10	-	1,750	-
Ni	0.005-0.5	10-300	-	30-1,000
Cu	-	40-300	170	170-300
Mg	-	1,000-2,400	-	3,000
Zn	-	400	160	250-600
Cd	-	70-600	180	20-600
Pb	0.02-200	9-340	-	340
Co	0.003-0.06	-	-	-
Mo	0.005-0.05	-	-	-
Mn	0.005-50	1,500	-	-
Na	-	3,500-30,000	-	60,000
K	-	2,500-5,000	-	12,000
Ca	-	2,500-7,000	-	8,000

2.7.4 The technology use in anaerobic digestion

Mainly, AD treatment technologies can be divided into 'low rate' and 'high rate' systems. In 'low rate' systems, long hydraulic retention times are applied, and 'high rate' systems, in which hydraulic retention time is relatively short. Waste streams with slurries and solid waste, generally used low rate systems. Generally, high rate systems are used with wastewater streams. The SRT should be much higher than the HRT in high rate systems (De Mes et al., 2003). The common anaerobic reactor classification show in **Table 2.12**.

Table 2.12 Anaerobic Reactor Classification

Low rate anaerobic reactors	High rate anaerobic reactors		
	Suspended growth	Attached growth	Others
• Continuous stirring tank reactor (CSTR) • Anaerobic ponds • Septic tanks	• Up-flow anaerobic sludge blanket (UASB) • Anaerobic sequencing batch reactor (ASBR)	• Anaerobic filter (AF) • Fluidized/Expended bed reactor	• Anaerobic membrane bioreactor (An-MBR)

Continuously Stirred Tank Reactor (CSTR) is the most common form in low-rate systems. In this system, the feed is introduced to the reactor and apply stirred mixing. In the same time, equal quality of effluent is removed from the anaerobic reactor. Schematic diagram of the CSTR system is shown in **Figure 2.25**.

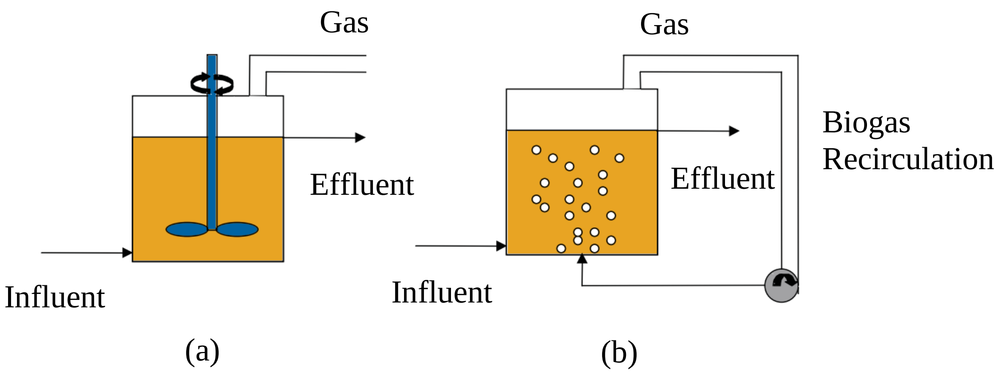


Figure 2.25 Schematic diagram of a CSTR system, (a) mechanically stirred and (b) stirred by biogas recirculation (De Mes et al., 2003)

Considering the high rate systems, the most common configuration is, up flow anaerobic sludge bed (UASB). Generally, the USAB is not attractive to treat concentrated slurries. But the USAB are much efficient in treating diluted and concentrated wastewater. UASB is a single phase high rate systems which all the anaerobic process steps take place at the same time. Schematic diagram of the UASB reactor configuration is shown in **Figure 2.26**.

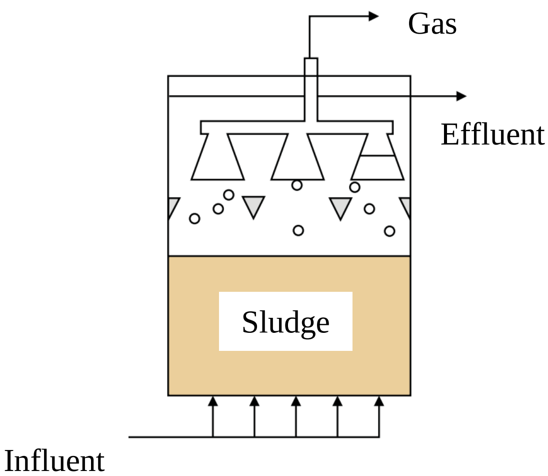
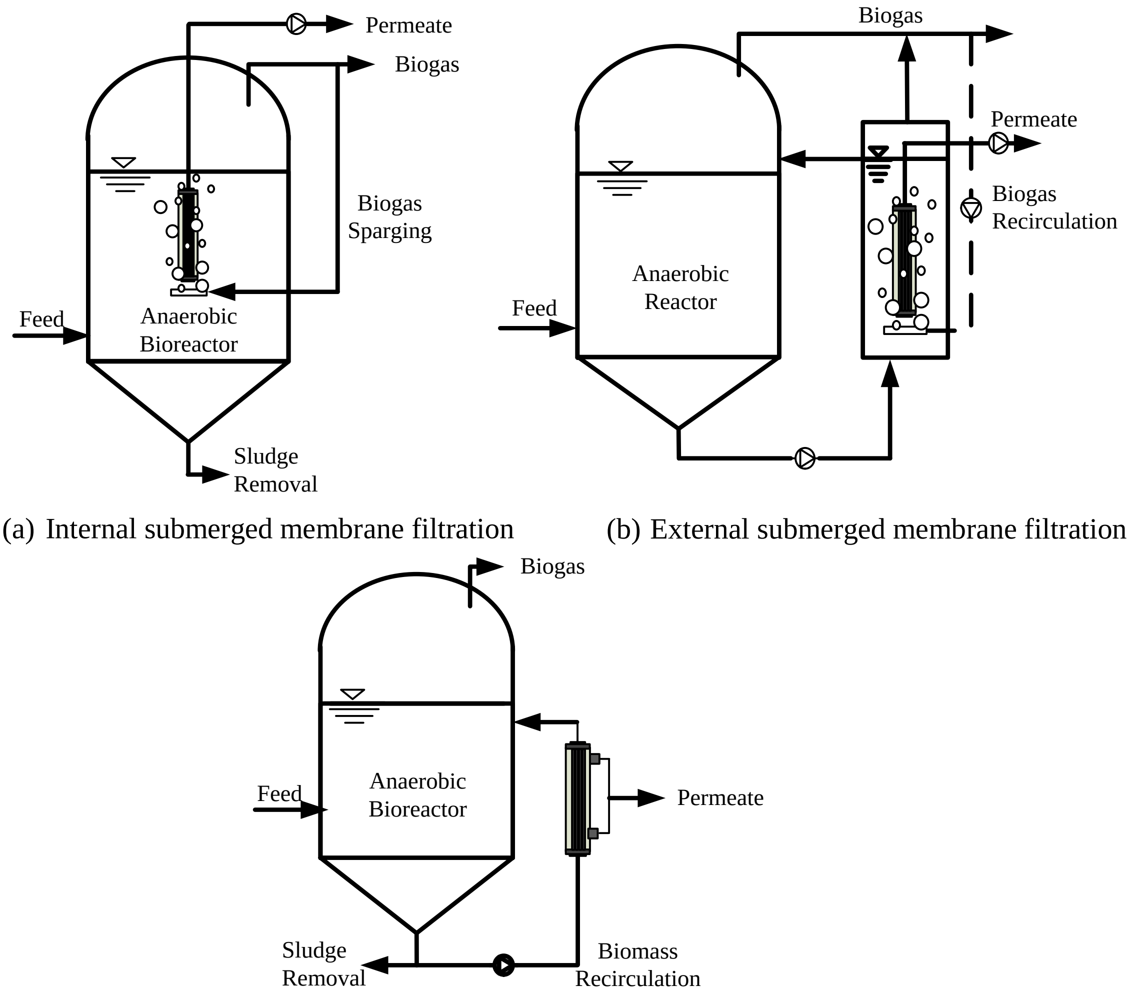


Figure 2.26 Schematic diagram of the UASB reactor configuration

2.7.5 Anaerobic membrane bioreactor (AnMBR)

The AnMBR concept of was developed in the 1980s. Initially, the AnMBR application has been limited due to membrane fouling, energy consumption of the membrane systems. But in recent years, MBR technology has been successful with large-scale membrane filtration systems. Those are well-developed for the operation and have the effective process design. Also, the operation and maintenance procedures developed to couple with the biological process. That kind of process helped to improve the potential of AnMBR as an energy recovery process rather than a treatment application (Liao et al., 2006).

The anaerobic membrane bioreactor is a combined system of the anaerobic bioreactor and the MF membrane filtration or low-pressure UF membrane system. The main objective to introduce the membrane is to retain TSS, including MLSS and inert solids. By introducing the membrane, solid retention time completely can be separated from the hydraulic retention time. In that case, separating solids do not depend on the wastewater characteristics, sludge properties, and the biological process. As shown in **Figure 2.27**, the membrane filtration system can be coupled with anaerobic bioreactors in main three different configurations which are, the internal submerged membrane filtration (a), external submerged membrane filtration (b) and the external crossflow membrane filtration (c).



(c) External crossflow membrane filtration

Figure 2.27 Different AnMBR system configurations (Liao et al., 2006)

There are main two different configurations that use in AnMBR process, which are submerged membrane filtration and the cross-flow pressurized membrane modules. AnMBR technology has been introduced for the treatment of a various kind of wastewaters and high solid content wastes, which include pulp and paper wastewater, food processing wastewater, etc. **Table 2.13** presents the results of the treatment of high strength wastewater of different sources by AnMBRs.

Table 2.13 Process Conditions and Treatment Performances of AnMBR used in High Strength Wastewater Treatment

	(Anderson et al., 1996)	**(Choo and Lee, 1996)**	**(Xie et al., 2010)**	**(Van Zyl et al.,**	**(Zayen et al., 2010)**
Wastewater	**Brewery**	**Distillery**	**Kraft evaporator condensate**	**industrial WW**	**Landfill leachate**
WW COD (g/L)	80 to 90	22.6	10	19.1	41
Temperature (°C)	35 to 37	53 to 55	36 to 38	37	37
OLR (kg	Above 30	1.5	22.5	Up to 25	6.27
HRT (day)	2.5 to 4.2	15	-	1.3	7
MLSS	Up to 51	-	8 to 12	36	-
COD removal (%)	99%	97%	93% to	96.8%	90.7%
Gas production (m^3 CH_4/kg COD)	0.28	0.26	0.35	-	0.48

Recently, there are some published research works related to the AnMBRs for municipal wastewater treatment. **Table 2.15** summarizes the reported results of the applications of AnMBR in municipal wastewater treatment.

Table 2.15 Treatment Performances and the Process Conditions of AnMBR used in Low Strength Municipal Wastewater Treatment

	(Lew et al., 2009)	**(Martinez-Sosa et al., 2011)**	**(Lin et al., 2011)**
WW COD (g/L)	0.54	0.6	0.342 - 0.527
Temperature (°C)	25	20	30
Reactor type	Complete mix	Complete mix	UASB
Reactor volume (m^3)	0.18	0.35	0.06
Membrane location	Side stream	Submerged	Submerged
Module type	Hollow fiber	Flat sheet	Flat sheet
Membrane area (m^2)	4	3.5	0.6
OLR (kg COD/m^3/day)	2.16	0.4-0.9	1
HRT (day)	0.25	1.5 - 0.67	0.42
COD removal	88%	84 - 94	90
Gas yield (m^3/kg COD)	-	0.24	0.24
Membrane flux (LMH)	7.5	7	12

Although it is feasible to treat domestic wastewater by using AnMBR in terms of theoretical energy and mass balance. But the full-scale applications is still being limited due to the efficiency of the energy recovery under low influent COD conditions. It is important to increase the influent COD, to achieve better performance in anaerobic condition. AnMBR process can produce better treatment and the energy recovery through high strength domestic sewage.

2.8 Research Gap

There are different kind of technologies available for the fractionation or the pre-concentration process. In wastewater treatment sector, those technologies are used for removing the solid fraction of the water. Presently, very few studies reported using a pre -concentration approach to the domestic sewage for enhancing the performance of the anaerobic digestion. Still, there are many research gaps related to the pre-concentration technologies. Membrane technology is one of the promising technology for the pre-concentration purposes. But there are many gaps to full fill for this particular application, such as membrane material, configuration, operational parameters, fouling tendency, membrane cleaning, energy consumption, etc.

Chapter 3

Methodology

3.1 Overall Experimental Framework

This study compared the performance of three different pre-concentration technologies; woven fiber microfiltration (WFMF), tube settler (TSET) and conical membrane tank (CMT) that can apply to concentrate the domestic sewage prior to the anaerobic treatment. The main goal was to concentrate as much as possible of the wastewater organic matters in a separate stream, which can later be used for energy recovery. Pre-concentration, performance was evaluated by in terms of chemical oxygen demand (COD), suspended solid (TSS) concentration and the energy consumption. Finally, the best performing technology was coupled with the anaerobic membrane bioreactor (AnMBR) and evaluated the overall performance. **Figure 3.1** represents the experimental framework.

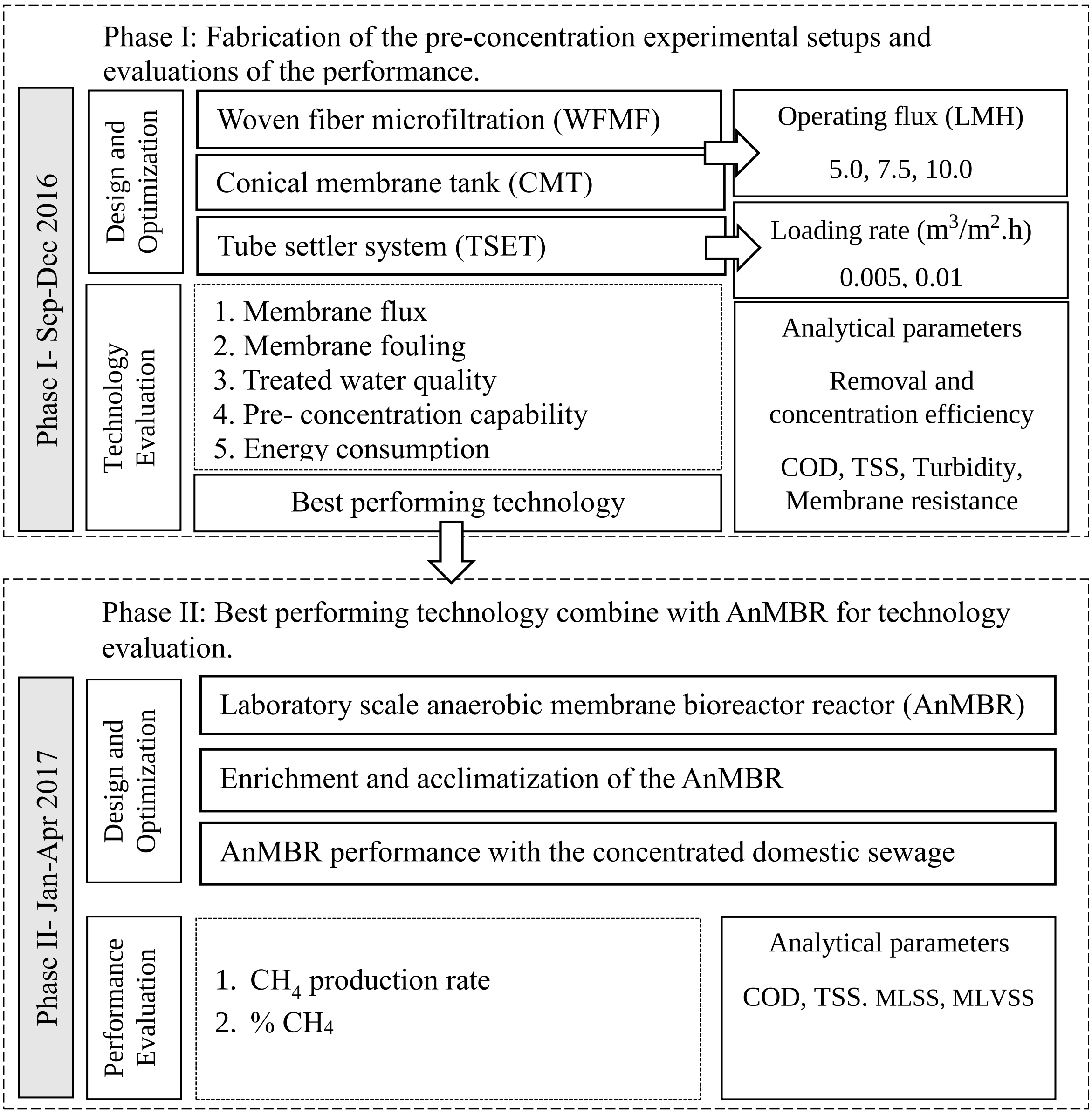


Figure 3.1 Overall experimental framework for the study

3.2 Type of Feed Water

The experimental tests were conducted using the AIT campus domestic sewage. Pre-concentration, performance were tested with WFMF, CMT and TSET technologies.

Table 3.1 Water Quality Characteristics of Domestic Sewage

Parameters	**Unit**	**Current Study**	**Domestic sewage** (Morel., 2006)	**AIT domestic sewage** (Tuan., 2008)
pH	-		6.7 – 8.35	7 – 7.65
Temperature	˚C	28.5 ± 0.5	25 - 30	24 - 32
TS	mg/L	487 ± 45	-	198 - 315
TSS	mg/L	80 ± 30	100 - 145	100 - 120
COD	mg/L	136 ± 50	200 - 400	150 - 200
BOD_5	mg/L	65 ± 8	150 – 200	100-120
DO	mg/L	< 1.66		

AIT domestic sewage can be identified as a combined sewage which was a mix of blackwater and graywater. This wastewater stream shows considerable amount of BOD in solid form to have a great possibility to treat with anaerobic digestion. To achieve that target, it was most important to have a pre-concentrating step before.

In phase I, laboratory scale woven fiber microfiltration (WFMF), tube settler (TSET) and conical membrane tank (CMT) setups were fabricated and tested in terms of domestic sewage pre-concentration ability.

3.3 Woven Fiber Microfiltration (WFMF) Flat Sheet Membrane System

WFMF system was designed with five woven fiber flat sheet membrane modules which corresponds to surface area of 1 m^2. Membrane module consisted with membrane sheets, spacers and flow channel. Overall the flat sheet WFMF membrane module had a 3-layer structure join together with the adhesive. WFMF system was operated dead-end outside-in configuration and membrane module operated on negative pressure. Rotary type peristaltic pump was used a permeate pump (Master flex-Model 77200-60). Suction pressure was created by changing the peristaltic pump speed.

The system was operated in the submerged mode as the membrane use negative pressure as a driven force. Pressure transducer (Trafag-model-579225-010) and data logger (Model-EL-USB-4) was used to record pressure data. Permeate pump was connected to a timer (Model-Omron twin timer) and operated with intermitted running intervals (OFF- 1 min, ON- 6 min). The level controller system was introduced to the membrane tank to maintain the water level inside the tank by connecting with the feed pump. Electrical control system was designed to have an automated operation. Schematic diagram of the system installation is shown in the **Figure 3.2**. Laboratory WFMF setup presented in Figure A-4 in **Appendix A**.

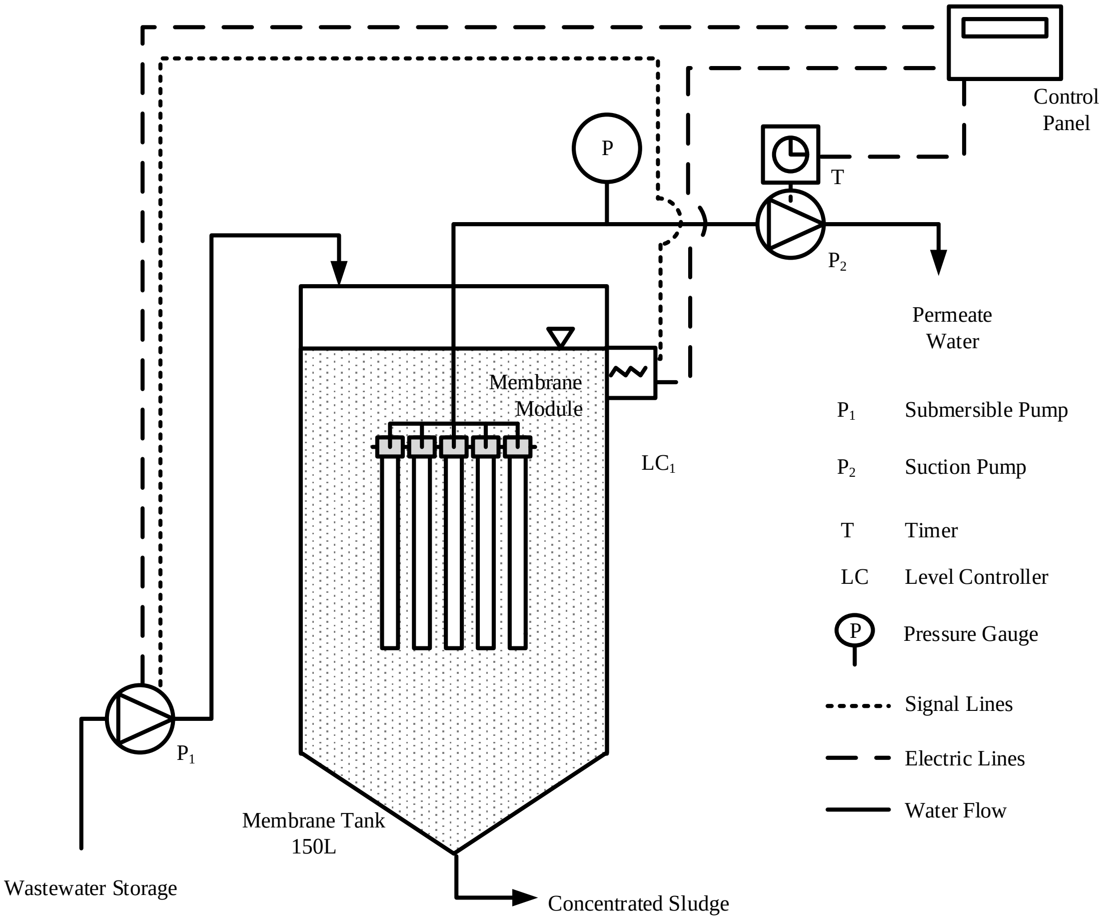


Figure 3.2 Schematic diagram of the WFMF Setup

Moreover, the membrane tank's bottom was fabricated to have a conical shape, thus it can help to settle down the accumulated sludge, and remove. Membrane tank was fabricated to have a total volume of 130 L and 23.5 L of sludge cone volume with PVC materials. As the membrane module was able to remove solid free water as permeate, domestic sewage concentrations increased and settled to the conical bottom of the tank. Energy consumption of the system was measured with a voltmeter. Characteristics of the membrane module shows in **Table 3.2.** Design criteria of the WFMF membrane module and the membrane tank, present in **Figure A-1, A-2 and A-3 in Appendix A.**

Table 3.2 Characteristics of the Woven Fiber Flat Sheet Membrane Module

Descriptions	**Characteristics**
Membrane material	Woven fiber
Type of membrane	Dead-end mode, outside-in, flat sheet
Pore size(μm)	1 - 3
Membrane dimension (cm) Length Width	 29.7 21.0
Each module consist of	2 sides
No. of flat sheet per module	5
Spacer type	Mesh PVC
Total membrane area (m^2)	1
Flux (Max for PWF)	50
Max operating pressure (kPa)	-60

3.3.1 Operation procedure of the WFMF system

Firstly, WFMF membrane was tested for pure water flux and calculated the initial membrane resistance.

3.3.1.1 Pure water flux

Tap water was used for this test. The purpose of this experiment was to find the maximum membrane performance and initial resistance of the WFMF membrane. Flux and transmembrane Pressure (TMP) data were recorded, while membrane runs with pure water. Calculation procedure of membrane flux and the transmembrane pressure shows in **Equation 3.1 and 3.2.**

$$\text{Membrane flux (LMH)} = Q/A \qquad \textbf{(Equation 3.1)}$$

Q = Volume (L)/ Time (h)
A = Membrane area (m^2)

$$TMP = P - (H+h)\,\rho\, g \qquad \textbf{(Equation 3.2)}$$

TMP = Trans Membrane Pressure (kPa)
P = Pressure gauge reading (kPa)
ρ = Density of water (1000 kg/m^3)
g = Gravity of acceleration (9.81 m/s^2)
H = Height from surface to three way socket for the pressure gauge (m)
h = Distance between ground level and water surface (m)

3.3.1.2 Membrane Resistance

Once the membrane tested for pure water flux, the graph was plotted by using TMP vs Flux. Based on the slope of the graph, membrane resistance was calculated by using **Equation 3.3 and 3.4.**

$$J = \Delta P/(\mu \times R_t) \qquad \textbf{(Equation 3.3)}$$

J = Permeate flux ($L/m^2.h$)
ΔP = Transmembrane pressure (kPa)
μ = Viscosity of the liquid (Pa.s)
R_t = Total resistance (1/m)

$$R_t = R_m + R_p + R_c \qquad \textbf{(Equation 3.4)}$$

Where,

R_t A total resistance to the flux
R_m Resistance due to membrane
R_p Resistance due to particle deposition
R_c Resistance due to colloidal deposition

3.3.2 Domestic sewage pre-concentration with WFMF

To evaluate the WFMF performance on pre-concentrating the domestic sewage, system was tested with three different flux. Concentrated sewage was collected from the bottom of the membrane tank and evaluated by analyzing chemical oxygen demand (COD), suspended solid (TSS) concentration of the sludge and the energy consumption that need to concentrate the one gram of COD with the technology. **Table 3.3** summarizes the operating conditions for the experiment.

Table 3.3 Overall Testing Procedure for the WFMF System

Membrane flux (LMH)	No. of test runs per each flux	Time (d)	Performance evaluation
5.0	3	7	• Flux vs TMP • Pre-concentration efficiency • Membrane fouling • Membrane cleaning • Permeate water quality
7.5			
10.0			

3.3.3 WFMF membrane cleaning and fouling analysis

Once the TMP of the system reached -60 kPa, the membrane modules were cleaned due to the higher energy consumption after the limit. In this case, -60 kPa was selected as the cleaning point. Cleaning procedure consisted with two major steps namely; physical and chemical cleaning. Chemical cleaning consisted with acid base cleaning. After every step, pure water flux experiments were conducted. Fouling analysis was followed to evaluate the cleaning performance.

3.3.3.1 Physical cleaning

Physical cleaning procedure follows:

- Stopped the operation.
- Disconnected the module from the system.
- Sprayed water to remove particles on the surface of the membrane.
- Cleaned the membrane module using a hand brush while spraying water.

3.3.3.2 Chemical cleaning

Base cleaning was carried out to remove the organic fouling of the membrane. 0.5 M NaoH solution and 0.03% NaOCl were used to clean the organic fouling of the membrane. Membrane module was immersed in 8 h in the solution and cleaned with the tap water. Once the base cleaning was carried out, membrane cleaned with the 0.01 % of HCl solution. The objective was to do the acid cleaning, is remove the inorganic fouling in the membrane. The membrane was immersed in the solution for 8 h, prior to tap water cleaning and use.

3.3.3.3 Cleaning performance evaluation

Before operating with the wastewater, the membrane was tested for the initial membrane resistance (R_m). Once the membrane clogged, cleaning procedure was carried out. Cleaning performance with the each steps was evaluated by the percentage recovery to the R_m. **Figure 3.3** illustrated cleaning performance determination procedure.

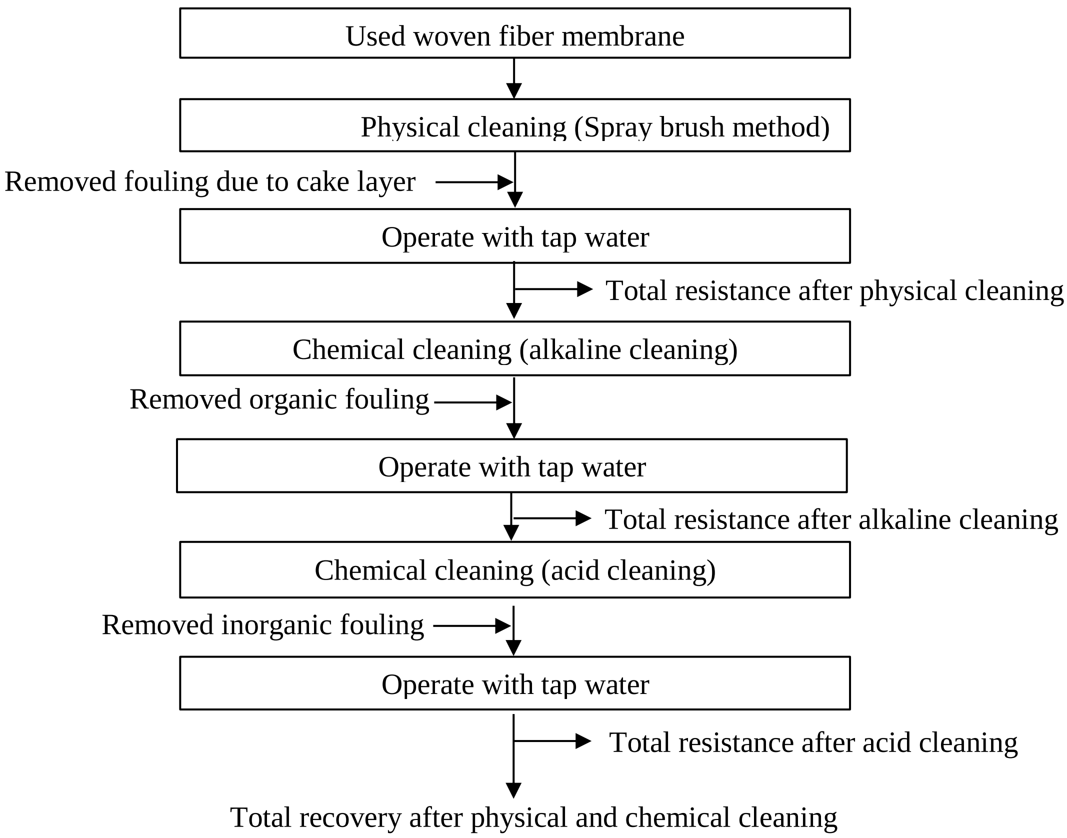


Figure 3.3 Cleaning performance determination procedure

3.4 Conical Membrane Tank (CMT)

3.4.1 Configuration of the CMT

The CMT system was designed with Polytetrafluoroethylene (PTFE) hollow fiber (HF) membrane of 0.1 µm pore size and the surface area of 0.1 m^2. Specially, the membrane tank shape was considered when designing. CMT system was operated dead-end outside-in configuration and membrane module operated on negative pressure. Rotary type peristaltic pump was used a permeate pump (Master flex-Model 77200-60). Suction pressure was created by changing the peristaltic pump speed. The system was operated in the submerged mode as the membrane use negative pressure as a driven force. Pressure transducer (Trafag-model-579225-010) and data logger (Model-EL-USB-4) was used to record pressure data. Permeate and feed pump was connected to a timer (Model- Omron twin timer) and operated with intermitted running intervals (OFF- 1 min, ON- 6 min). As the membrane module was able to remove solid free water as permeate, domestic sewage concentrations increased and settled to the conical bottom of the tank. Electrical control system was designed to have an automated operation. The cone-shaped membrane tank was designed to have a 23 L of total volume and the 0.9 L of sludge cone volume. Energy consumption of the system was measured with a volt meter. Schematic diagram of the system installation is shown in **Figure 3.4**. Laboratory scale CMT setup presented in **Figure A-5** in **Appendix A**.

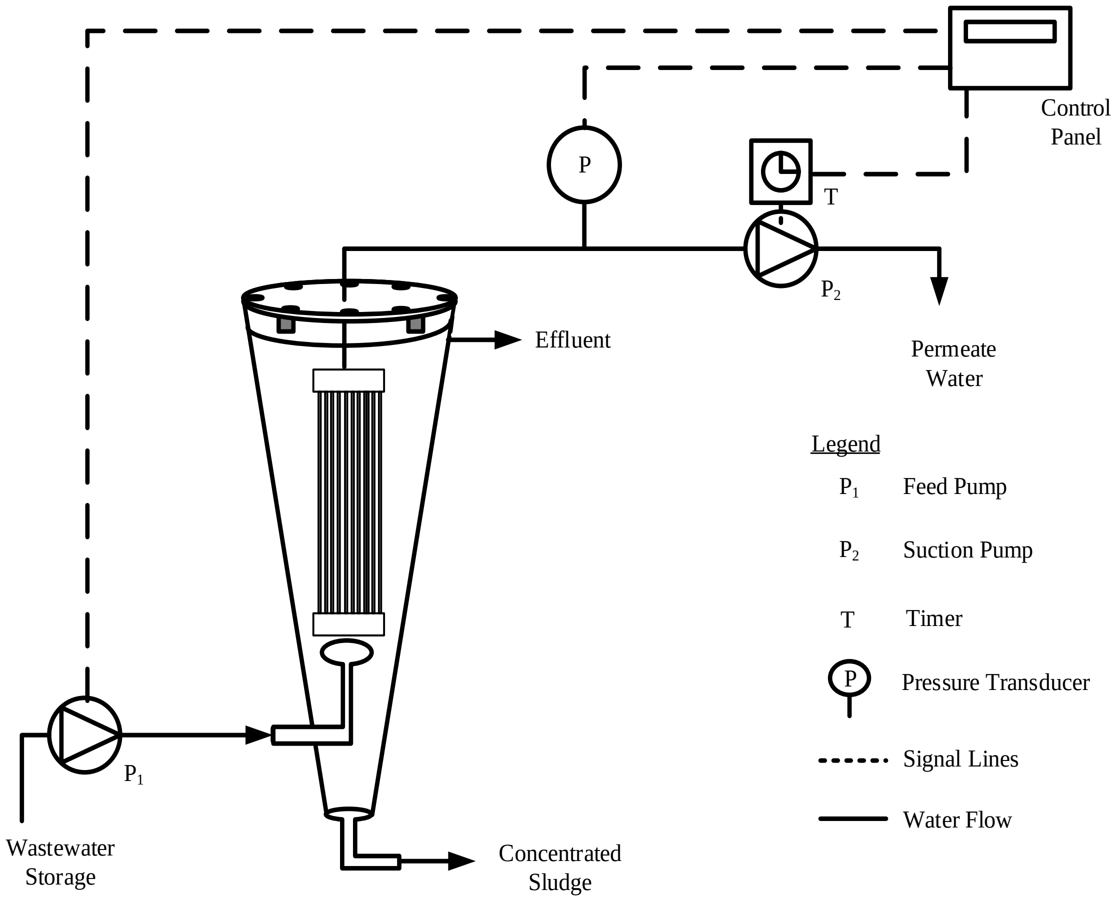


Figure 3.4 Schematic diagram of the conical membrane tank

The conical tank design was helped to solid, liquid separation. Settling of solids occurs in the upper area of the conical tank and sludge falls through the bottom of the tank, where it can remove for the further treatment. This system was a combination of sedimentation principle and the membrane filtration. Surface area of the settling was the main factor of sedimentation rate. **Equation 3.4, 3.5 and 3.6** were followed as a design principle for the CMT.

$$V_0 = \frac{h_0}{t_0} \quad \textbf{(Equation 3.4)}$$

Where V_0 = Settling velocity of the particle
h_0 = Depth of the tank
t_0 = Detention time

$$t_0 = \frac{V}{Q} = \frac{l \times w \times h_0}{Q} \quad \textbf{(Equation 3.5)}$$

Where t_0 = Detention time
V = Volume of the tank
Q = Flow rate

$$V_0 = \frac{Q}{L \times w} = \frac{Q}{As} \quad \textbf{(Equation 3.6)}$$

Where
Q = Flow rate
A_s = Surface area

Based on Equation **3.4, 3.5 and 3.6**, surface area has a major impact on particle settling. In this experiment, conical tank was used. In membrane tank, when it comes to the deep end, the surface area was reduced. In this case, the tank bottom area was reduced and settling velocity of the particles increased. This phenomena helped to particles to settle and accumulate in the bottom of the conical tank. Pre-concentrated sludge was removed from the bottom of the conical tank. The PTFE hollow fiber membrane was used in CMT system and the specification of the membrane is presented in **Table 3.4.**

Table 3.4 Specification of the PTFE Hollow Fiber Membrane

Description	Specification
Manufacturer	Sumitomo, Japan
Material	PTFE
Membrane configuration	Hollow fiber membrane
Membrane area	0.1 m^2 / module
Flux (PWF)	12-42 $L/m^2.h$
Pore size	0.1 µm
Tube diameter	0.8 mm
TMP (filtration)	< 60 kPa

3.4.2 Domestic sewage pre-concentration with CMT

Firstly, the PTFE hollow fiber membrane was tested for pure water flux and calculated the initial membrane resistance following the method that indicated in section 3.3.1.1 and 3.3.1.2. Once the initial tests were finished, domestic sewage was introduced to the system and evaluated the pre-concentration performance.

To evaluate the CMT performance on pre-concentrating the domestic sewage, system was tested with three different flux. Concentrated sewage was collected from the bottom of the conical tank and evaluated by analyzing chemical oxygen demand (COD), suspended solid (TSS) concentration of the sludge and the energy consumption that need to concentrate the one gram of COD with the technology. **Table 3.5** summarizes the operating conditions for the experiment.

Table 3.5 Overall Testing Procedure for the WFMF System

Membrane flux (LMH)	No. of test runs per each flux	Time (d)	Performance evaluation
5.0	3	7	• Flux vs TMP • Pre-concentration efficiency • Membrane fouling • Membrane cleaning • Permeate water quality
7.5			
10.0			

3.4.3 PTFE hollow fiber membrane cleaning and fouling analysis

Cleaning performance evaluated by analyzing the % recovery of each step. **Figure 3.5** illustrates the cleaning procedure of the membrane.

Physical cleaning- Rinse with tap water

↓

Test for pure water flux

↓

Immerse the membrane in the 1,500 mg/L NaClO + 2 wt% NaOH for 6 hours

↓

Rinse the membrane with tap water

↓

Test for pure water flux

↓

Immerse the membrane in 1 wt% H_2SO_4 for 6 hours

↓

Rinse the membrane with tap water

↓

Test for pure water flux

Figure 3.5 PTFE hollow fiber membrane cleaning procedure

Once the TMP of the system reached -60 kPa, the membrane modules were cleaned due to the high energy consumption and flux reduction. In this case, -60 kPa was selected as the cleaning point. Cleaning procedure consisted with two major steps, namely; physical and chemical cleaning. Chemical cleaning consisted with acid base cleaning. After every step, pure water flux experiments were conducted and did the fouling analysis and evaluate the cleaning performance. Chemical preparation method for the PTFE membrane cleaning, presented in **Appendix B**.

3.5 Tube Settler Application

3.5.1 Configuration of the tube settler

Laboratory scale tube setter tank was designed to concentrate the solid fraction of the domestic sewage. Typically, tube settler systems are an inexpensive solution for wastewater plants to increase treatment capacity, reduce new installation footprints, improve effluent water quality, and decrease operating costs. Tank bottom has slope to accumulate the settled solid and can remove it from the bottom of the tank. Tube settler was designed with PVC material with having tube area of 3.2 m^2. Tube area calculation highlighted in **Appendix B**. The total volume of the tube settler was 72 L with 18 L of sludge cone volume. Angle of inclination of the tube was adjusted to have a 60° to have a higher settling capacity. Feed pump was connected to the system and operated with intermitted mode (ON-5 min, OFF-1 min) for protection of the feed pump. Energy consumption of the system was measured with a volt meter. Graphical design criteria of the tube settlers illustrated in **Figure A-6** to **A-9** in **Appendix A**.

3.5.2 Domestic sewage pre-concentration with tube settler

Tube settler was operated with two scenarios and **Table 3.6** summarizes the operating conditions for the experiment.

Table 3.6 Operational Condition for Tube Settler

Loading rate ($m^3/m^2.h$)	No. of test runs	Time (d)	Performance evaluation
0.005 without coagulation	3	7	• Pre-concentration efficiency • Permeate water quality • Energy consumption
0.01 with coagulation			

For 0.005 $m^3/m^2.h$ loading rate system was operated without coagulant dosing as the loading rate was less. Moreover, jar test was conducted to finding out the optimum coagulant dose to the tube settler. Schematic diagram of the system installation is shown in the **Figure 3.6.**

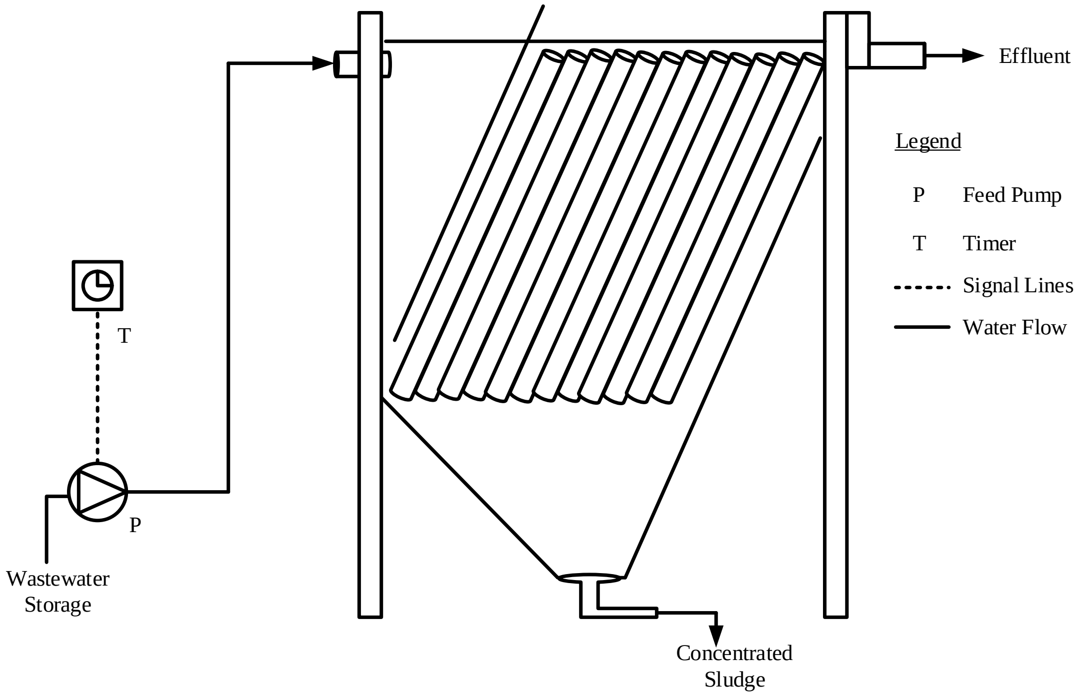


Figure 3.6 Schematic diagram of the tube settler application

For the 0.01 m^3/m^2.h loading rate, tube settler was operated with coagulation and system configuration changed to have a dosing pump. Dosing pump operated to have 20 ppm of coagulant dose to the system. System configuration shows in **Figure 3.7.**

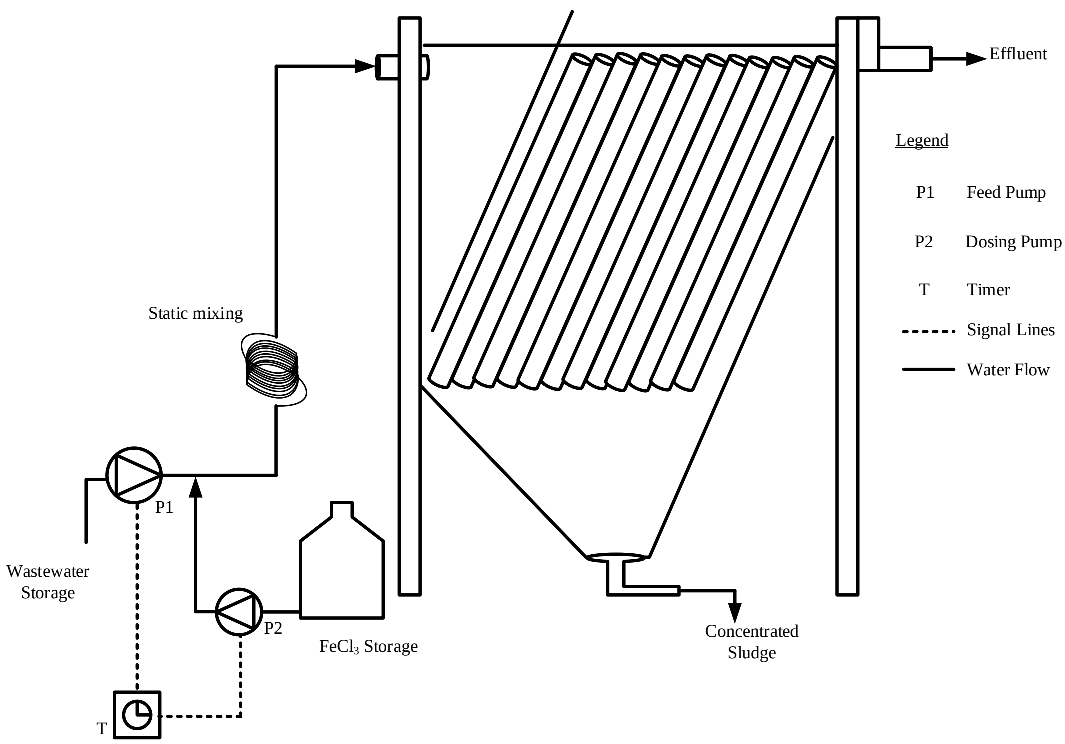


Figure 3.7 Schematic diagram of the tube settler with coagulation

3.6 Performance Evaluation of the Pre-concentration Technologies

3.6.1 In terms of COD pre-concentration

COD concentration of the concentrated sludge was taken to evaluate the system. Initial and final COD concentrations were measured and calculated the concentration factor. One experiment was carried out 7 days and tested for the COD concentrations. For the common comparison of all three technologies, g COD/m^3.d was taken to evaluate the system.

3.6.2 In terms of solid accumulation

All three pre-concentration systems were designed to capture more biodegradable solids in domestic wastewater streams. Solid accumulation considered as it represents the settleable COD which was targeting to capture. Membrane systems permeate and the tube settler's effluent were evaluated in terms of solid removal. Permeate and effluent quality were assessed in terms of TSS in stream.

3.6.3 In terms of energy usage

All the pre-concentration systems were evaluated with the energy that need to concentrate one gram of COD. To compare the technologies, kWh/g COD was taken as a common unit. Voltmeter was installed to each system and energy data recorded for every test.

Analytical parameters for pre-concentration evaluation

Table 3.7 Analytical Parameters and Methods to Evaluate Pre-concentration Performance

Parameters	Unit	Analytical methods	References
COD	mg/L	5220-C (Close reflux method)	APHA et al. (2005) 5510 B
TS	mg/L	Evaporation disk method	APHA et al. (2005) 2130 D
TSS	mg/L	Filtration/ Evaporation	APHA et al. (2005) 9221 B
Turbidity	NTU	Nephelometric method, Hach turbidity meter	APHA et al. (2005) 2540 D
Membrane resistance	1/m	Resistance in series model	(Choo and Lee, 1996)
TMP	kPa	Digital pressure gauge	-
DO	mg/L	DO meter	APHA et al. (2005) 2130 D
Energy consumption	kWh	Volt meter	-

3.7 Anaerobic Membrane Bioreactor (AnMBR)

The main objective of this phase was to investigate the performance of the anaerobic digestion with the concentrated domestic sewage. This operation was carried out as phase II, by using the concentrated sewage from the best pre-concentration technology. To achieve this objective, the single stage anaerobic membrane bioreactor operation was carried out by treating pre-concentrated domestic sewage.

The single stage AnMBR was designed to be an automated system. The system was constructed with a working volume of 6 L reactor using stainless steel, respectively. The system was operated in single stage with the ceramic membrane filtration system. Biomass recirculation was used in order to achieve a good mixing condition (5 min mixing and 1 min non-mixing). Initially the system was started with the synthetic glucose as feed water. Feed water was fed to the reactor by peristaltic pump (Masterflex L/S drives, 6-600 rpm) with intermittent feeding at a controlled feed flow rate by an automatic level sensor immersed in the reactor. When the reactor was filled up to the required level, the feed pump was stopped through a relay unit integrated with level sensor. The final biomass separation from effluent was carried out using a ceramic microfiltration membrane. The membrane module was operated in an external cross-flow configuration. Membrane filtration carried out with inside- out filtration mode. Furthermore, it was operated under suction mode to remove the constant flux. The filtration cycle was adjusted to increase suction pressure to obtain constant permeate flow rate. Schematic diagram of the system installation is shown in the **Figure 3.8.**

3.7.1 Operating conditions of the AnMBR

AnMBR system was operated in mesophilic conditions in order to evaluate the performances with the pre-concentrated sludge. **Table 3.8** presents the detail operating conditions of AnMBR operation. Design calculations were presented in appendix B.

Table 3.8 Operating Conditions of the AnMBR

Parameters	Unit	Overall
Temperature	°C	26-30
Influent COD	g/L	6-7
Loading rates	Kg COD/m^3.d	3.2
HRT	d	2.18
SRT	d	∞
Flow rate	L/d	2.74
Working volume	L	6
Biomass retention	-	Ceramic membrane filtration
Permeate flux	L/m^2.h	0.63

During the startup period, glucose solution was used as a feed water and used the same 3.2 kg COD/m^3.d loading rate. Once the system became stable, concentrated domestic sewage was introduced step by step to the reactor. **Figure 3.9** summarizes the procedure of introducing the real feed to the system. Loading rate, synthetic wastewater preparation and the design calculations were highlighted in **Appendix B**.

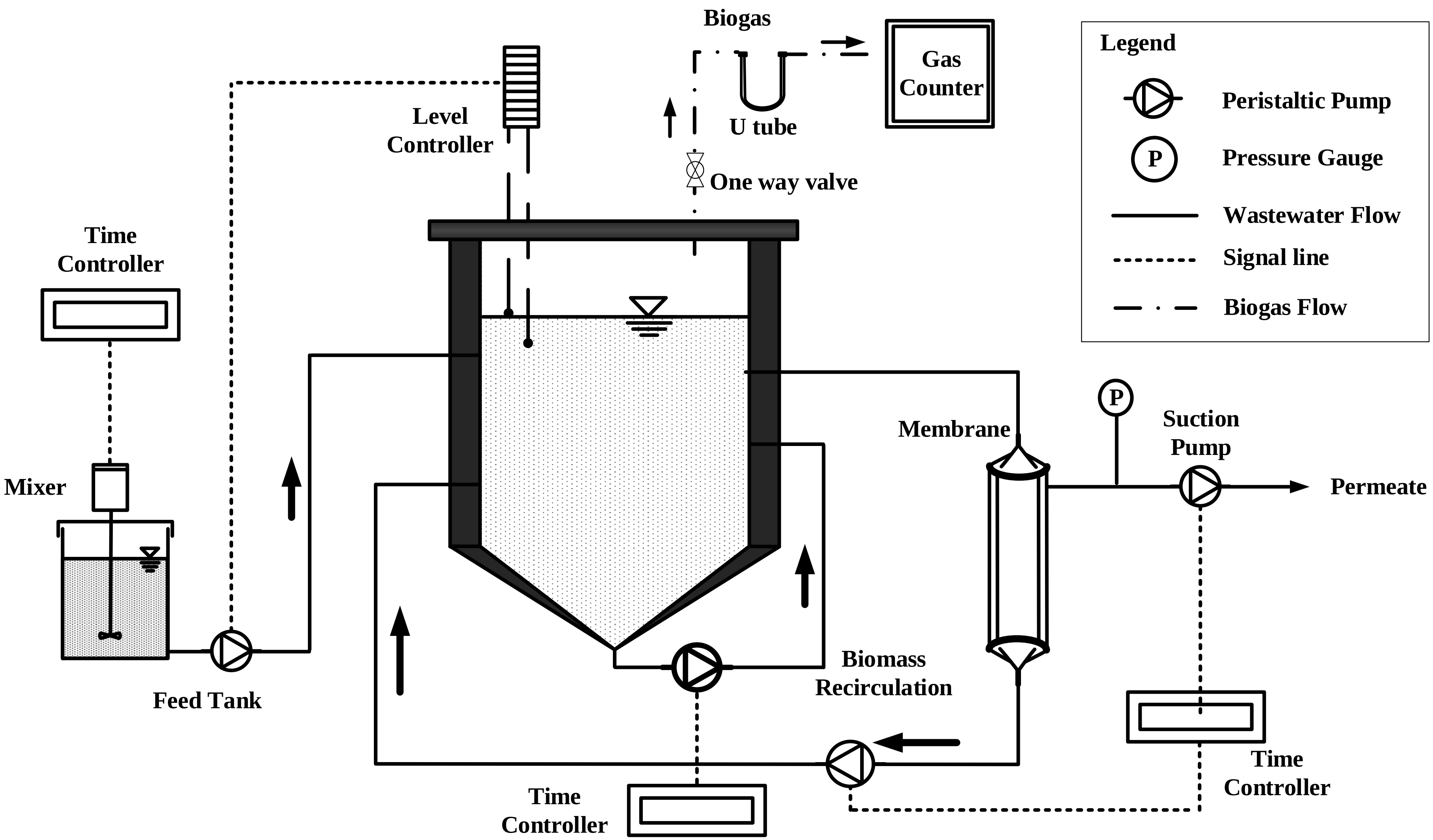
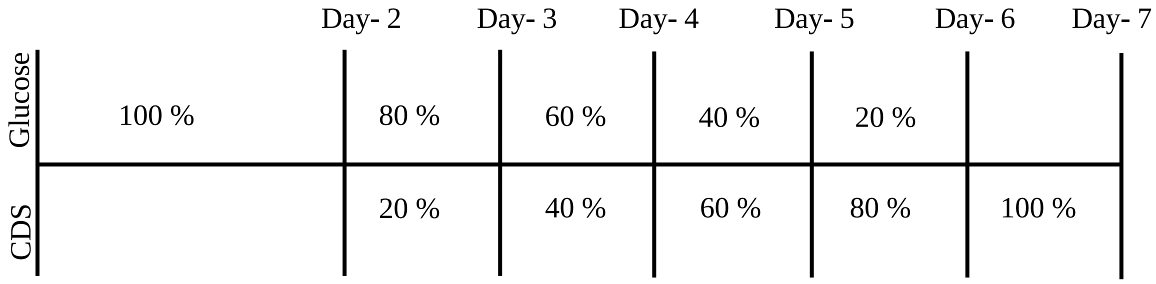


Figure 3.8 Schematic diagram of the anaerobic membrane bioreactor



CDS: Concentrate Domestic Sewage

Figure 3.9 Procedure of introducing the real feed to the AnMBR system

The mixing percentage was based on COD concentrations and after the 7th day, the system started to operate only this with the concentrated domestic sewage.

3.7.2 Evaluation of the anaerobic digestion process

To evaluate the process, biogas production, methane content, COD removal rate, and membrane performance were considered mainly. Analytical parameters and methods for the process shows in **Table 3.9**. Methane content in biogas is an important indicator of anaerobic wastewater treatment. Also, the composition of the biogas is a good indicator to assure the anaerobic condition of the process. The biogas composition in the reactor was analyzed by gas chromatography. Moreover, methane yield is one of the main indicators of anaerobic wastewater treatment. The biogas produced from anaerobic reactor were measured with gas counter and methane yield calculated as follows.

Methane yield (m^3CH_4/kg COD) = Volume of methane generated (m^3)/ kg COD **(Equation 3.7)**

3.7.3 Analytical parameters and method of analysis

Table 3.9 Analytical Parameters and Methods

Parameters	Unit	Analytical Methods	Equipments/Techniques	References
Biogas production	m^3CH_4/kg COD	Gas counter method	Measurement with gas meter	-
Methane content	%	Gas chromatography	Gas chromatography (Agilent 7890A) with TCD	APHA et al. (2005)
COD	mg/L	5220-C (Close reflux method)	Titration	APHA et al. 5510 B
TSS	mg/L	Filtration/ Evaporation	Filter/Oven	APHA et al. 9221 B
MLSS	mg/L	2540-D (Dry at 103-105°C)	Filter/Oven	APHA et al. (2005)
MLVSS	mg/L	2540-E (Ignite at 550°C)	Furnace	APHA et al. (2005)
pH	-	Glass Electrode	pH meter	-
Turbidity	NTU	Nephelometric Method	Turbidity meter	-
DO	mg/L	HACH DO meter (HQ-40d)	DO meter	-
TMP	kPa	Digital Pressure gauge	Pressure transducer	-

3.7.4 Biomass separation with ceramic membrane filtration

3.7.4.1 Ceramic membrane specifications

AnMBR was operated external cross flow mode with the ceramic membrane filtration. The ceramic membrane specifications are given in **Table 3.10.**

Table 3.10 Ceramic Membrane Specifications

Parameters	Values/Specifications
Membrane manufacturer	NGK Insulator, Japan
Membrane material	Ceramic
Membrane type	Microfiltration
Module configuration	Tubular (multi-channel)
Effective surface area	0.18 m^2
Pore size	0.1 µm
Maximum flux	87.5 $L/m^2.h$
Dimensions	Diameter-30 mm, Length-450 mm
Operating pressure range	20-90 kPa
Maximum operating temperature	300°C

3.7.4.2 Operating condition of the cross flow ceramic membrane system

Ceramic membrane was operated in cross-flow mode with intermitted time control to minimize the effect of high shear intensities on biological activity. The filtration cycle was adjusted to maintain required permeate flux. The ceramic membrane operating conditions are presented in **Table 3.11**.

Table 3.11 Ceramic membrane operating conditions

Descriptions	Characteristics
Temperature (˚C)	20-40
Maximum operating pressure (kPa)	-60
Filtration method	Cross flow with gravity settlement
Filtration cycle	2.5 min filtration and 15 min gravity settlement
Permeate flux ($L/m^2.h$)	0.63
Cross flow velocity (m/s)	0.073

3.7.4.3 Ceramic membrane cleaning procedure

When the suction pressure increased up to -60 kPa or constant flux not able to generate due to fouling, the membrane cleaning procure was carried out. When such scenario, the membrane was removed and cleaned with the tap water and then carried out the acid base cleaning. The membrane was soaked in cleaning solution, (a) 0.5 M NaOH for 15 minutes at 75°C to remove organic fouling and (b) a dilute (5 mL/L) mixture of nitric acid (HNO_3) at 58% and phosphoric acid (H_3PO_4) at 75% for 15 minutes at 50°C to remove inorganic fouling. In between every membrane cleaning step, the ceramic membrane was rinsed with deionized water.

Chapter 4

Results and Discussions

This chapter presents the findings from the experiments on domestic sewage pre-concentration technologies, namely; woven fiber microfiltration, conical membrane tank, and the tube settler as the first stage. The content of this chapter is divided into five sections. First four sections presented the results of the pre-concentration technologies and the performance evaluation. In the final section of this chapter presents the results and corresponding discussion for the performance of the anaerobic membrane bioreactor which uses the concentrated domestic sewage as an influent. Firstly, Pre-concentration, performance was evaluated in terms of chemical oxygen demand (COD), suspended solid (TSS) concentration and the energy consumption. Finally, AnMBR performance was evaluated with biogas production and methane percentage.

4.1 WFMF System Efficiency

4.1.1 Pure water flux performance and initial membrane resistance

This test was conducted to find the initial resistance of the membrane. To analyze the membrane fouling and cleaning performance, the pure water flux needs to be analyzed. In this test, membrane was operated with pure water and observed the TMP increment with different flux. The whole idea was to find the initial membrane resistance with the slope of the graph. **Figure 4.1** Illustrates the pure water flux experiment results.

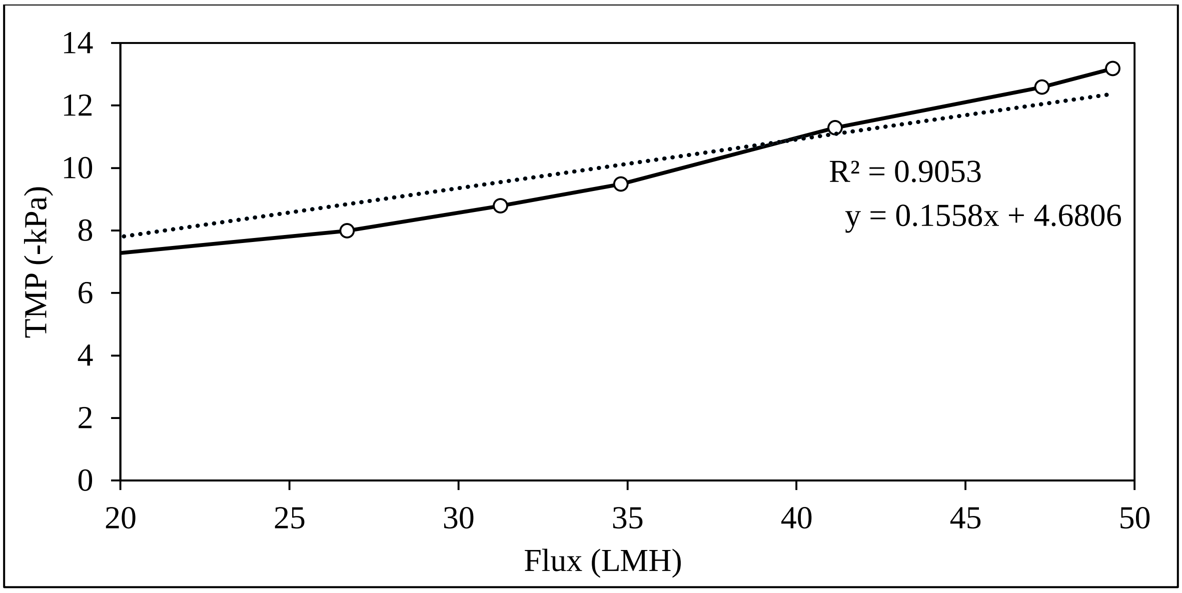


Figure 4.1 TMP vs pure water flux for woven fiber membrane

According to the theoretical explanation, this graph should start from the zero. But in practical situations, the peristaltic pump was used as a suction pump. In that case, the membrane flux increased manually by increasing the pump speed. Because of this reason, the minimum pump speed generated about 20 LMH flux by resulting -7.5 kPa as a tans-membrane pressure.

Based on the slope of the pure water flux graph, initial membrane resistance was calculated. Membrane resistance of WFMF membrane was found to be 1.47E+12 m^{-1}. Application of this membrane resistance value presented in **section 4.1.6**. Membrane resistance calculation method presented in **Appendix C.**

4.1.2 TMP increment with the different flux

In this experiment, WFMF membrane was tested with domestic sewage with three different membrane flux which was 5.0, 7.5 and 10.0 LMH. All the tests were done in triplicates and continued for 1 week time. Triplicate results for 5 LMH is illustrates in **Figure 4.2**.

4.1.2.1 5 LMH Flux- Compilation of the triplicates

Intermitted run - 6 min run/ 1 min stop Feed Water- Domestic sewage

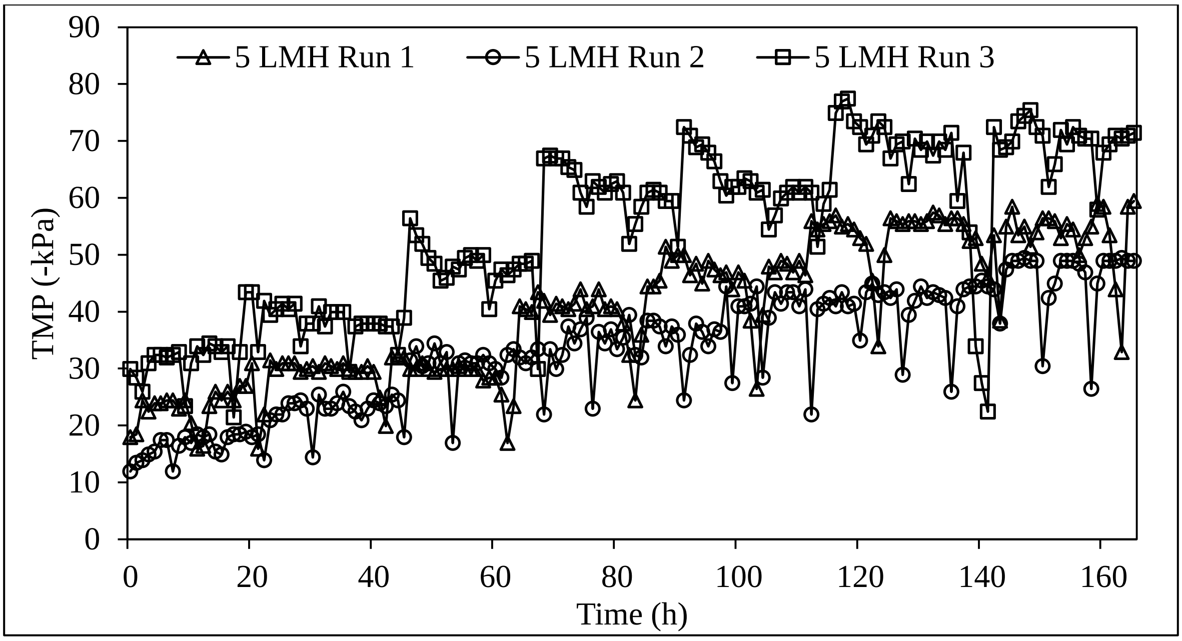


Figure 4.2 TMP vs Time for 5 LMH membrane flux- WFMF

It was observed that within 7 days of time, TMP was increased -18 kPa to -59 kPa with the constant flux of 5 LMH for the first test run. In the same way, TMP was increased -12 kPa to -49 kPa for the second test run. But in the third test run, it was observed that the higher TMP increment as -28 to -71 kPa, compared to the previous test.

Initially, only the 0.03% NaOCl solution was used to clean the membrane. NaOCl performed better for removing organic fouling of the membrane. But not the inorganic fouling. In this case inorganic fouling kept accumulating in the membrane and showed the higher TMP increment when it comes to the third run. In this case, it was identified that the acid cleaning method also need to introduce to the membrane for removing inorganic fouling of the membrane. By considering the three test runs, it was identified that the 5 LMH flux working well without reaching the maximum operating pressure of -60 kPa. This indicates that this system could be operated with higher filtration rate. This result suggests increasing the membrane flux by corporation with proper chemical cleaning. Then it has a possibility to operate with higher flux. Trans-membrane pressure data for all the test runs were presented in **Appendix C.**

4.1.2.2 7.5 LMH Flux- Compilation of the triplicates

Intermitted run - 6 min run/ 1 min stop Feed Water- Domestic sewage

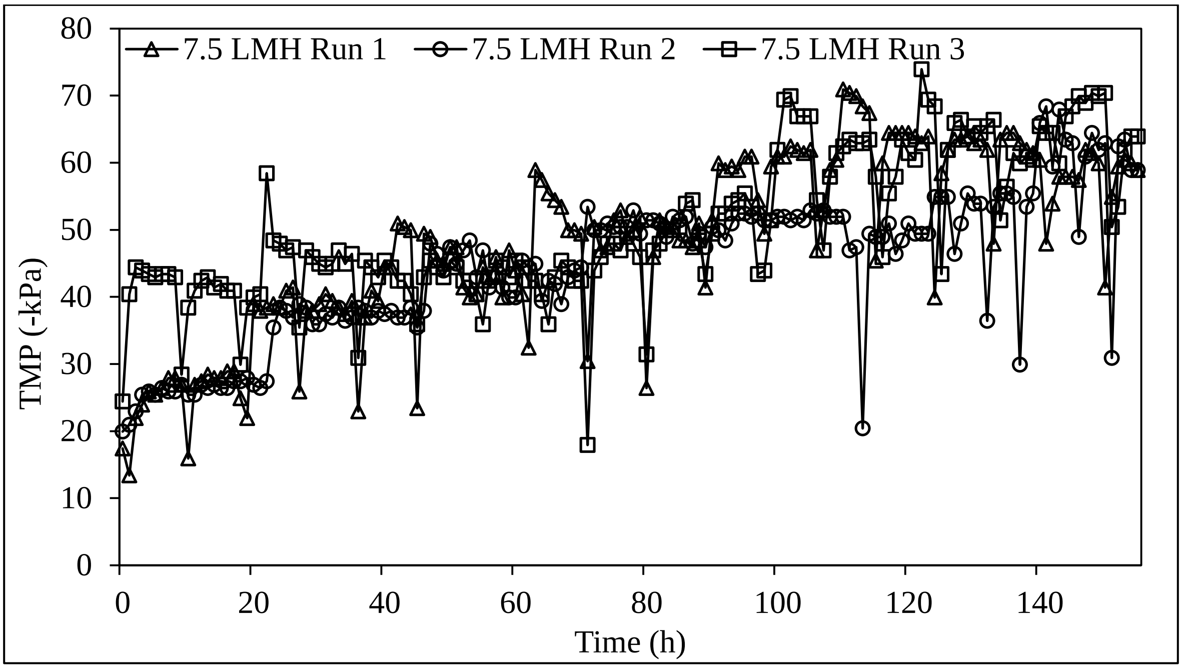


Figure 4.3 TMP vs Time for 7.5 LMH membrane flux- WFMF

For all the tests, TMP was fluctuating between -10 to -70 kPa with gradual increment. First two runs results were almost same compared to the third run. Before the initial two runs, physical and chemical cleaning was done. Before the third run, after doing the physical cleaning 76% recovery was achieved and decided to not to continue for the chemical cleaning as it is more than 70% recovery which is considerably better cleaning achievement. It was found that, if there is no chemical cleaning prior to the run, the pressure increment with the time is slightly higher. In that case it is always better to do physical and chemical cleaning if the membrane works with the domestic sewage.

TMP was gradually increased over the time. The tests were conducted with the intermitted runs and system pressure was dropped while the system off for one minute. This method helps to increase the operating time by controlling the TMP increment. Trans-membrane pressure data for all the test runs were presented in **Appendix C.**

4.1.2.3 10 LMH Flux- Compilation of the test results

10 LMH flux was unsustainable due to immediate membrane clogging. Complete test, supposed to conduct 1 week time and it only stood for 2 days. Within 48 hours, TMP was reached to cleaning point which was -60 kPa. Once the membrane clogged, it was not able to generate the constant flux of 10 LMH. Both the test results, trends were almost same which can conclude that the 10 LMH is not a sustainable flux at this point. **Figure 4.4** illustrates the TMP increment with time for 10 LMH flux.

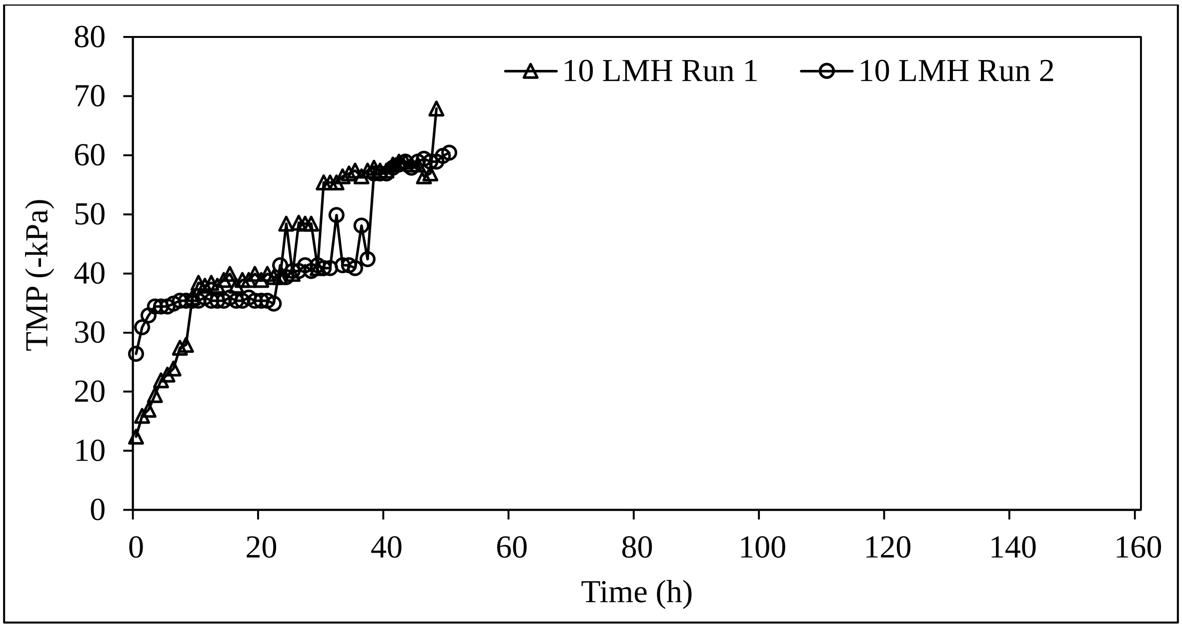


Figure 4.4 TMP vs Time for 10 LMH membrane flux-WFMF

After comparing the three different flux, the 7.5 LMH was found to be the best in terms of the operation. Trans-membrane pressure data for all the test runs were presented in **Appendix C.**

4.1.3 COD pre-concentration performance

COD pre-concentration, performance can be expressed in two ways, namely; the number of times that can concentrate from the initial level and system performance with the comparable common units. **Figure 4.5** shows the COD pre-concentration capability of the WFMF system.

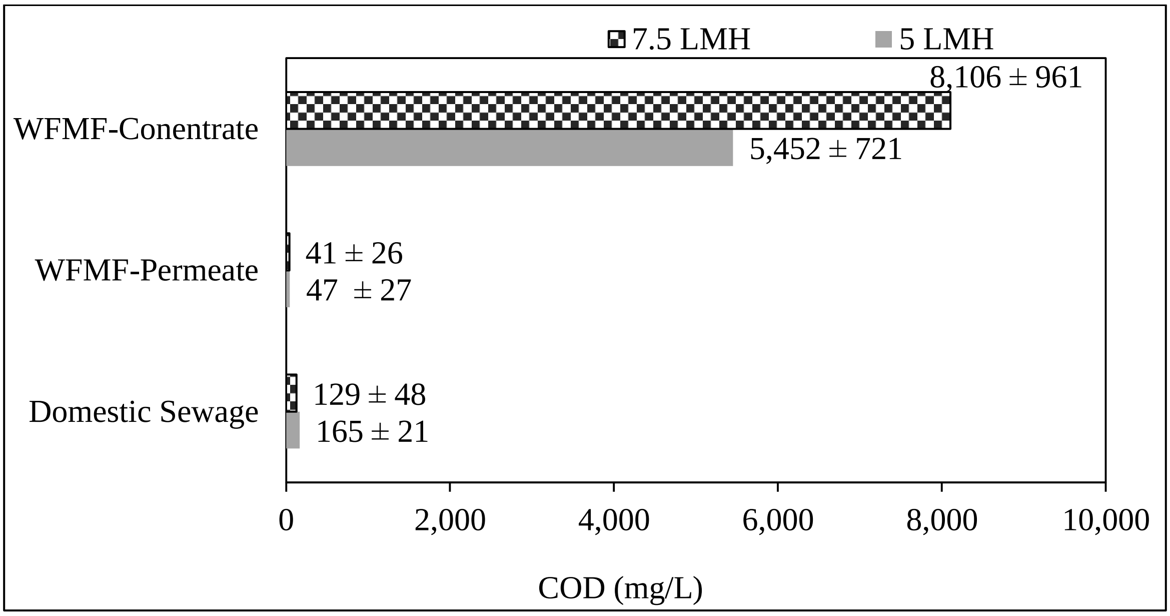


Figure 4.5 COD pre-concentration capability of the WFMF system

According to the **Figure 4.5** COD pre-concentration results, it can be identified that the WFMF system was able to concentrate domestic sewage COD within the one week of the operation time. When comparing the 5.0 and 7.5 LMH, it can be clearly seen the increment of the COD concentration as the system filtrate higher amount of domestic sewage in 7.5 LMH flux. With 5 LMH flux, WFMF system could filtrate 120 L/d. This number depends on the membrane area. For this experiment 1 m^2 membrane was used. In this case, the system could filter 0.84 m^3 for the one week of filtration time. But when it comes to the 7.5 LMH, the system could filtrate 1.26 m^3. Because of this reason more COD can accumulate in 7.5 LMH flux in the particular time period.

In this experiment, triplicate tests were conducted. Based on the data it is clearly identified that the system could pre-concentrate domestic sewage by 33 times for the 5 LMH flux. For the 7.5 LMH, it could concentrate COD up to 62 times. Based on this result, it can come to a conclusion that, more the filtration higher the COD concentration.

The permeate COD level is lower as the woven fiber membrane is filtering the domestic sewage. In this case, all the settleable COD is accumulating inside the system and only the soluble COD is leaving the system. For both the flux, COD concentration of permeate was very low. It can be seen that the approximately one-third to one-fourth of the COD, was leaving the system as permeate.

This COD pre-concentration performance depend on the system size and tank shape. For example, this system has a sludge cone, which has a volume of 23.5 L. In that case, all the suspended particles accumulated in this volume. During COD analysis, this complete sludge cone volume was considered. When sampling for the COD, complete sludge cone volume was mixed together and got the unique sample that can represent the complete sludge cone. Because of this reason, when the COD concentration calculated with mg/L, it represent the accumulated COD in sludge cone.

For the comparison purposes this kind of scale issues need to avoid and it is important to bring all the comparable factors to a common unit that can be compared. In that case, gCOD/m^3.d was selected as a common unit for the comparison and that can bring the system performance in a one place that can be compared. **Section 4.4** presents the common unit comparison in detail. Moreover, the complete data set for the COD analysis presented in **Appendix C**.

4.1.4 Solid accumulation of the WFMF system

In the WFMF system, membrane separates the total suspended solid fraction and remove permeate with very less suspended solid concentration. The pore size of the woven fiber membrane is 1-3 µm and almost all the suspended solid content could stop passing through the membrane. **Figure 4.6** shows the experimental results for the TSS accumulation for 5.0 LMH and 7.5 LMH for WFMF system.

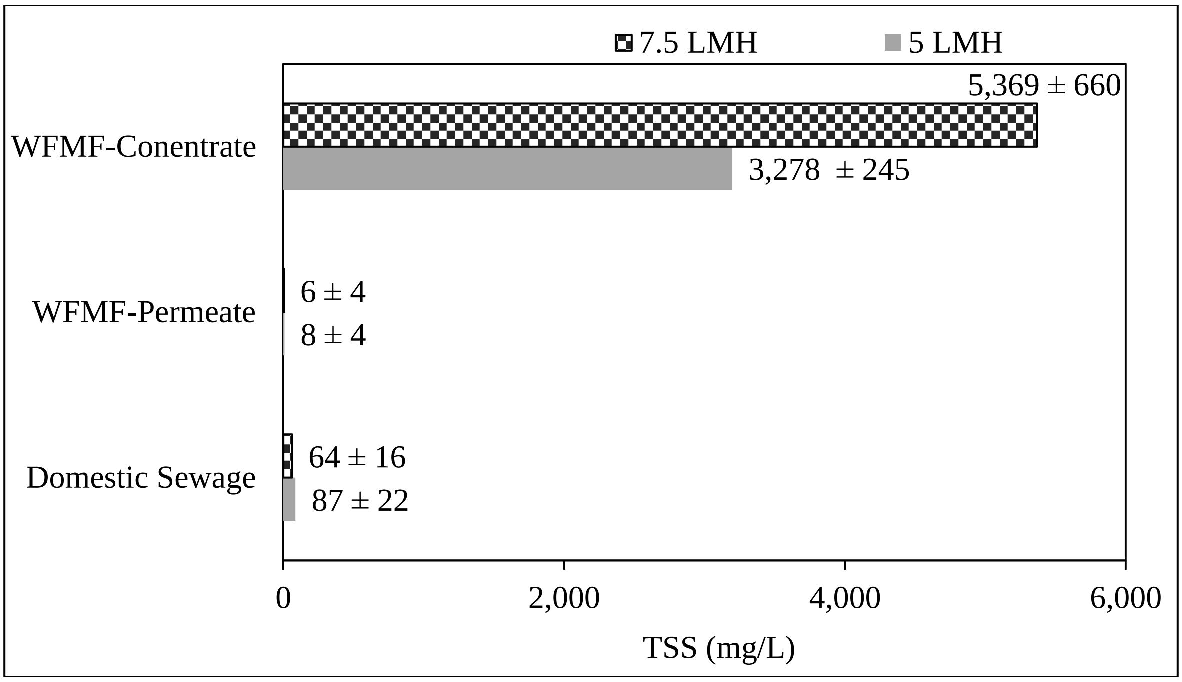


Figure 4.6 TSS accumulation for 5.0 LMH and 7.5 LMH- WFMF

Based on the TSS accumulation results, it can identify the WFMF system could accumulate or concentrate TSS for both the operating flux. As mentioned in the earlier in **section 4.1.3,** the 7.5 LMH flux is removing more water while the accumulation of the TSS than compared to the 5 LMH flux. More than 90% of TSS can be removed or hold with the WFMF system. In this case, WFMF system can generate almost solid free water as permeate which can have possibly using for secondary reuse application.

It has a positive trend of accumulating the TSS of the system, when the flux increased to 7.5 LMH from the 5 LMH. Permeate quality is still same in both the flux. For 5 LMH flux, the TSS accumulation was about 37 times from the initial concentration while the 7.5 LMH having 84 times. When the membrane system operates with the higher flux, the influent flow rate also becoming higher. In this case, more suspended solids come with the influent and immediately settle in the membrane tank.

Considering the COD and the TSS values, it was found that the COD has a positive relationship with TSS. It can be identified that the more TSS accumulation resulting the higher COD concentrations of the concentrate. The whole idea was to accumulate more solids in the membrane system which corresponds to the higher COD concentration. Based on these results it can make a statement that the pre-concentration of domestic sewage concept can be proved with WFMF membrane system.

4.1.5 Energy consumption of the WFMF

WFMF system consisted with the many of electronic and electrical devices, namely; permeate suction pump, feed pump, level control relay systems, electronic pressure gauge and data logger. Energy consumption for concentrating one gram of COD with the WFMF system is illustrated in **Figure 4.7** for 5.0 and 7.5 LMH flux.

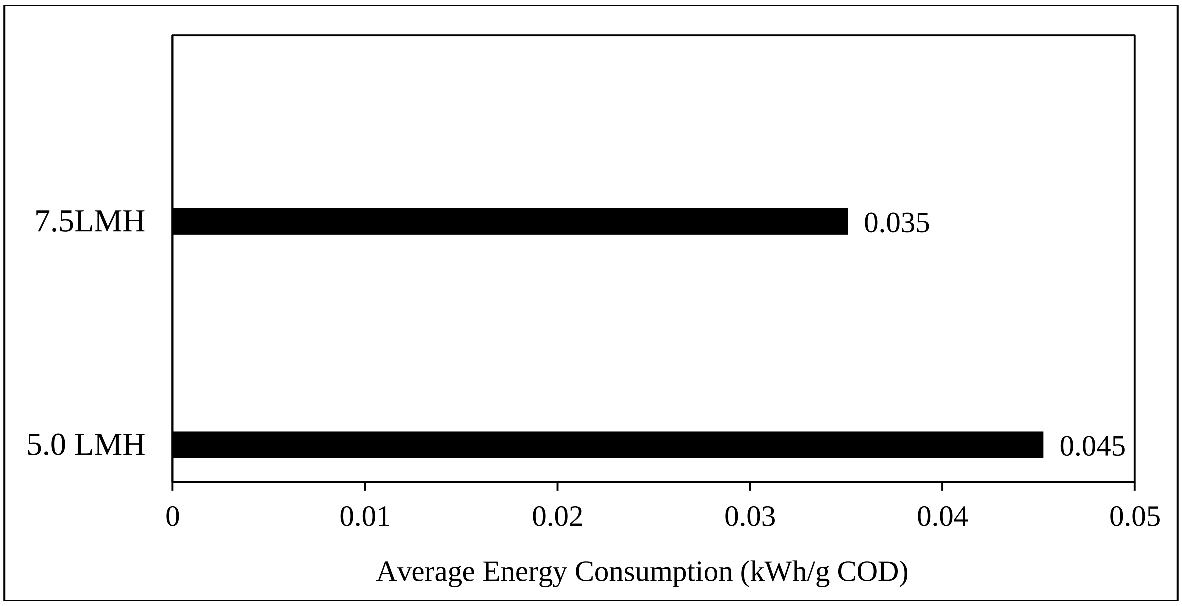


Figure 4.7 Average energy consumption for 5.0 and 7.5 LMH

Generally, when the system operates with higher flux, the energy consumption should be high as the pumps are working in high capacity. But in this case, when it considers the amount of COD, this scenario is coming in different ways. When the system works in higher flux, it can pre- concentrate more COD than it runs with lower flux. The energy consumption is lower compared to the low flux values when it considers per gram of COD.

The main reason for this is, any electrical equipment needs a fixed amount of energy to start and operate in minimum level. But the slight increase of pump increment does not make a big difference in this kind of laboratory-scale models. Other equipment like electronic pressure gauges, data loggers has constant energy consumption for any operating flux. Only permeate and feed pump speed is effected for the increment of the energy. Moreover, the energy consumption data for the complete operation highlighted in **Appendix C**.

4.1.6 Woven fiber membrane fouling and cleaning performance

As mentioned in section 4.1.1, initial membrane resistance was calculated with the pure water flux test. Once the membrane fouled, cleaning procedure was carried out step by step. Membrane resistance was calculated after physical, base and acid cleaning steps and calculated the percentage recovery of each cleaning step.

During the 5 LMH run, only physical cleaning was carried out to see the performance with the intension of no chemicals for cleaning. After the physical cleaning, pure water flux test, was carried out to evaluate the cleaning method.

Initial resistance of the WFMF membrane R_m = 2.44E+10 (1/m)
Resistance after physical cleaning = 5.16E+10 (1/m)

Recovery, as a percentage = $(R_m/ R_t)*100$
= (2.44E+10/5.16E+10)*100
= 47.3%

Only 47.3% recovered with physical cleaning, which is not acceptable. A proper cleaning method can recover at least more than 70%, which is near to the initial membrane resistance. According to the poor recovery with only physical cleaning, it was suggested to do the chemical cleaning afterward. As indicated in **Figure 4.2**, in the third 5 LMH run, it is identified that the initial starting pressure higher than the first and the second run. In the same way, pressure increment frequency also higher than previous runs. TMP was starting to increase -30 kPa to a maximum of -75 kPa. But average TMP was fluctuating around -60 kPa. In this step, chemical cleaning was carried out and evaluate the recovery. Chemical cleaning was done by mixture of 0.5 M NaOH and 0.03% NaOCl solution.

Initial resistance of the WFMF membrane R_m = 2.44E+10 (1/m)
Resistance after physical cleaning = 4.38E+10 (1/m)

Recovery, as a percentage = $(R_m/ R_t)*100$
= (2.44E+10/ 4.38E+10)*100
= 55.8%

Initially, only the higher amount of organic fouling was expected. But this point clearly shows the inorganic fouling also responsible for the membrane fouling in considerable level. In this test, the cleaning method could not achieve the 70% recovery, even with the base chemical cleaning. Typically, base chemicals use to remove the organic fouling. But in this case, inorganic fouling also playing a major role. In this case, acid cleaning was recommended with 0.01 HCl solution. During the 7.5 LMH testing period, Physical, base and acid cleaning were conducted. **Figure 4.8** illustrates the cleaning performance with respect to the pure water flux.

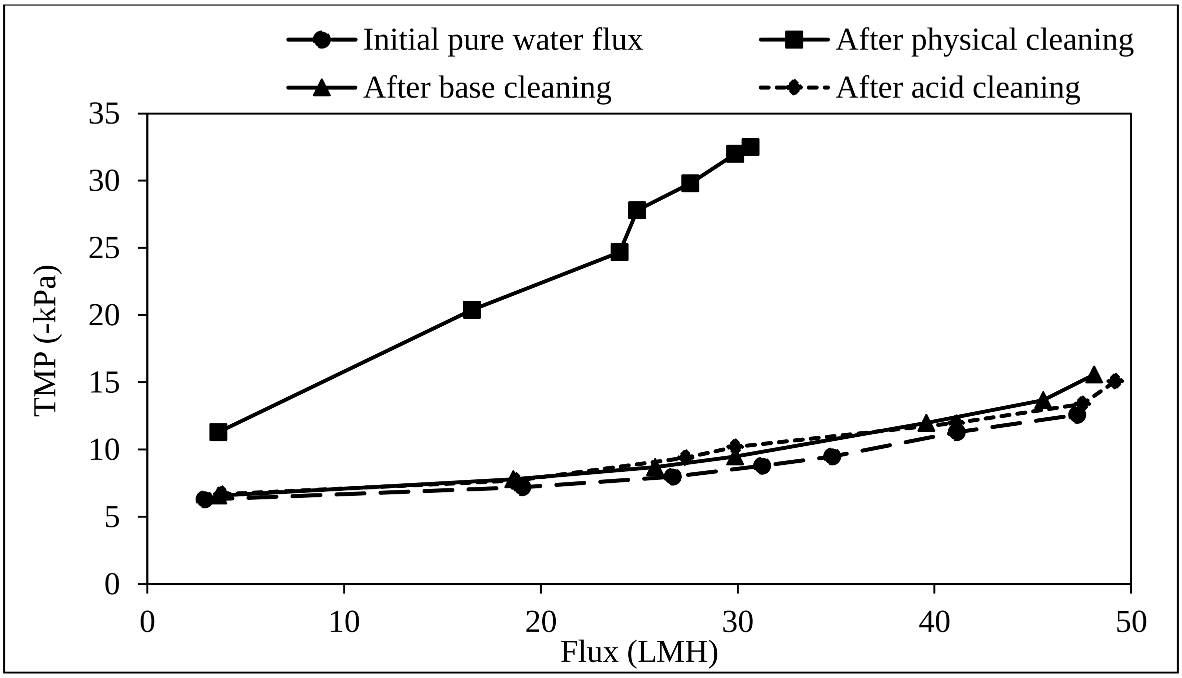


Figure 4.8 Cleaning performance of the WFMF membrane

Once the pure water flux done with all the steps, resistance was calculated for each step. Initially, only the membrane resistance was there while operating with the pure water. But when it operates with domestic sewage, resistance due to the cake layer, organic fouling and inorganic fouling plays a major role. Finally, the cleaning performance was evaluated at each step as a percentage recovery to the initial stage. **Table 4.1** shows the resistance values and the parentage recovery in each step. A pictorial view of each step of the cleaning presented in **Appendix C**.

Table 4.1 Percentage Recovery of the Cleaning Steps for Woven Fiber Membrane

	Membrane resistance	**After physical cleaning**	**After base cleaning**	**After acid cleaning**
Resistance (m^{-1})	1.47E+12	4.58E+12	1.64E+12	1.63E+12
Recovery (%)		32	89	90

Based on the cleaning performance data, it can be identified that the cake layer responsible for the 32% of the membrane fouling. This fouling can be removed with physically with the spray brush cleaning. The higher amount of fouling occurs due to the organic fouling which is responsible for the 57% of the fouling. Inorganic fouling was compared to less, at 1 %. As the domestic sewage was rich in organic and inorganic foulants, it was noted that woven fiber membrane system needs to accommodate with chemical cleaning. This statement can be proved with the membrane fouling potential and the cleaning performance of the membrane. Moreover, the detail calculation steps for the resistance, fouling analysis and cleaning performance were presented in **Appendix C**.

4.2 CMT System Efficiency

4.2.1 Pure water flux performance and initial membrane resistance

In this test membrane was operated with pure water and observed the TMP increment with different flux. **Figure 4.9** Illustrates the pure water flux experiment results of the PTFE hollow fiber membrane.

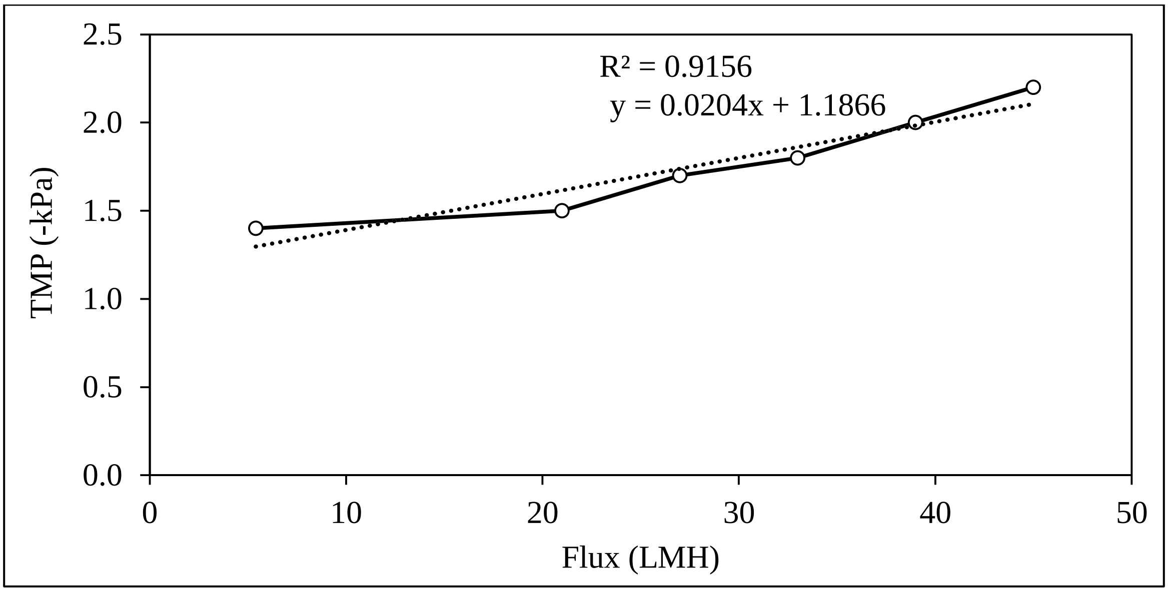


Figure 4.9 TMP vs pure water flux for the PTFE hollow fiber membrane

In this application, PTFE hollow fiber membrane was used. As mentioned in the section 4.1.1, the pure water flux test was conducted to find out the initial membrane performance.

The main objective of this test, was to find out the initial membrane resistance. To analyze the membrane fouling and cleaning performance, the pure water flux needs to analyze.

This particular membrane can generate considerably higher flux with pure water with lower TMP. Based on the slope of the pure water flux graph, initial membrane resistance was calculated. It was found to be the membrane resistance of this PTFE hollow fiber membrane is 2.47E+11 m^{-1}. Application of this membrane resistance value, presented in **section 4.2.6**. Membrane resistance calculation method was presented in **Appendix D.**

4.2.2 TMP increment with the different flux

In this experiment, PTFE-HF membrane was tested with domestic sewage with three different membrane flux which were 5.0, 7.5 and 10.0 LMH. All the tests were done in triplicates and continued for 1 week time. Triplicate results for 5 LMH is illustrated in **Figure 4.10**.

4.2.2.1 5 LMH Flux- Compilation of the triplicates

Intermitted run - 6 min run/ 1 min stop

Feed Water- Domestic sewage

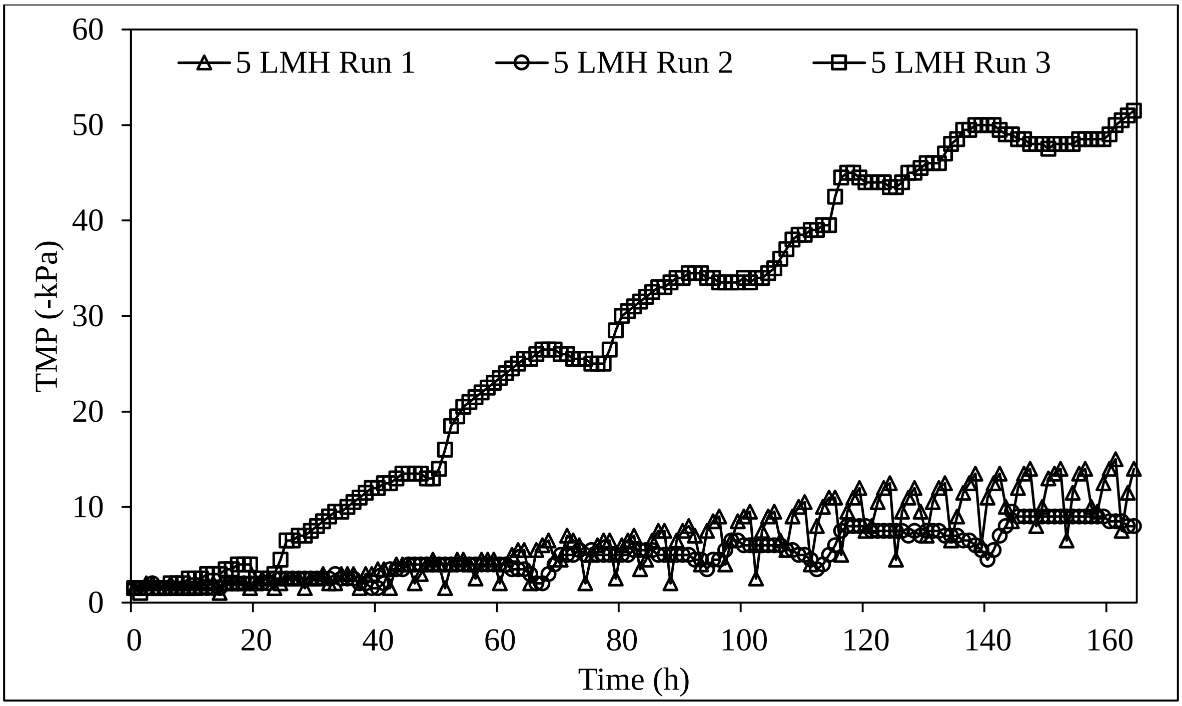


Figure 4.10 TMP vs Time for 5.0 LMH flux for PTFE hollow fiber membrane

Initial two runs results were almost same considering the TMP increment. In first two runs, TMP was increased -1 to -15 kPa within the 1 week time. It can be expressed as a lower pressure range with compared to the recommended operating pressure which is < -60 kPa.

In the initial stage, the membrane was trying to clean with only physical cleaning to see that there is a possibility of operation, without chemical cleaning. But the third test run results showed the sudden increment of TMP with the time, compared to the first two runs. Nearly one day, the membrane was operated like the initial two runs. But after that, there was a sudden TMP increment. That results were representing the accumulation of the organic and inorganic fouling of the membrane. In that situation, it was recommended to do the chemical cleaning after each run. In that case complete physical and chemical cleaning methods were conducted during the 7.5 LMH operation.

By analyzing the triplicate results, it was identified that the 5 LMH flux working well with the domestic sewage with the particular time frame. But this result shows that the membrane capacity is higher than 5 LMH and it can be achieved by cooperation with proper chemical cleaning. Trans-membrane pressure data for all the test runs were presented in **Appendix D.**

4.2.2.2 7.5 LMH Flux- Compilation of the triplicates

Intermitted run - 6 min run/ 1 min stop Feed Water- Domestic sewage

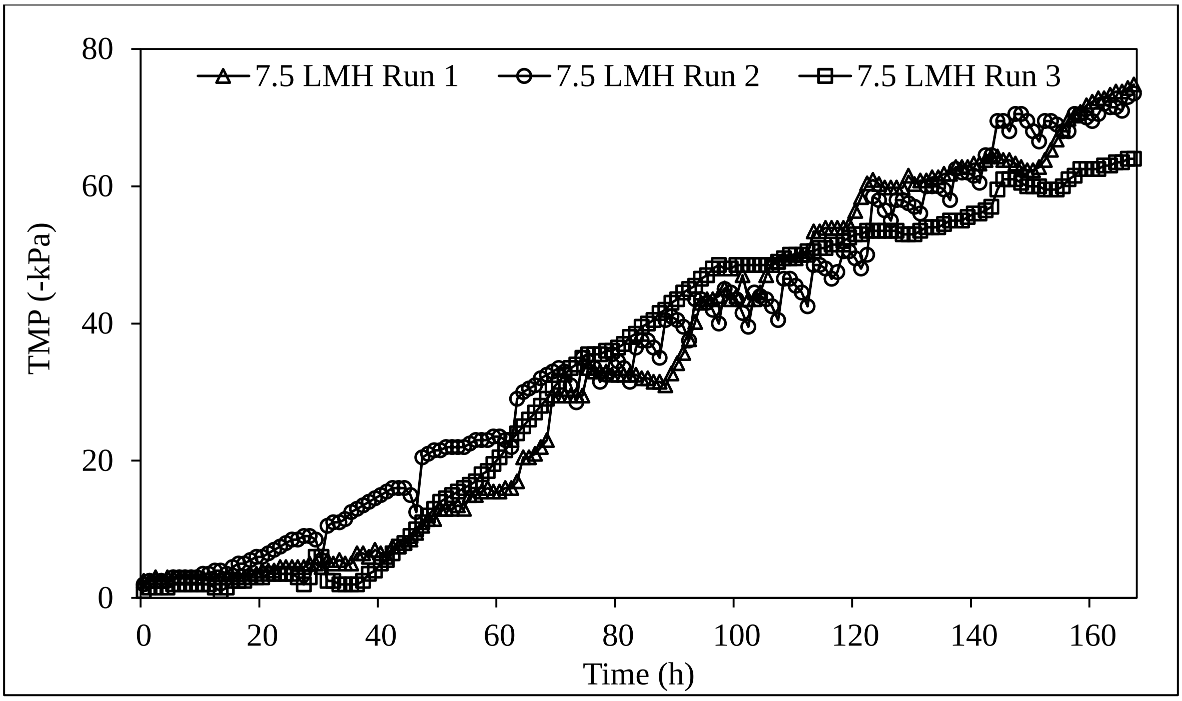


Figure 4.11 TMP vs time for 7.5 LMH flux for PTFE hollow fiber membrane

TMP increment shows the highest values compared to the 5 LMH flux. Considering the all triplicate tests, it has a sudden increment of the TMP after 1 day of operation. All the tests show the similar trend for the 7.5 LMH flux. TMP was gradually increased over one week of time. A test was conducted with the intermitted run and system pressure were dropped while the system off for one minute. This method helps to increase the operating time by controlling the TMP increment.

7.5 LMH flux performed well in terms of the 1 week time frame for this test. All the tests were done without chemical back flushing. Membrane cleaned chemically once finished the one week operation period. In that case, it can come to a conclusion that the 7.5 LMH performed better in terms of the pre-concentration objective. Trans-membrane pressure data for all the triplicate tests were presented in **Appendix D.**

4.2.2.3 10 LMH Flux- Compilation of the test results

The PTFE hollow fiber membrane was not able to work in 10 LMH flux. Tests were done in duplicates and both the time, same trend was reported. About one day period, the membrane was able to work in considerably lower TMP as less than -6kPa. But after a short time period, pressure was starting to increase gradually. After 60 hours of operating time it could reach up to more than -60 kPa. After this point, the operation was not successful due to the membrane not able to generate constant 10 LMH flux.

Intermitted run - 6 min run/ 1 min stop Feed Water- Domestic sewage

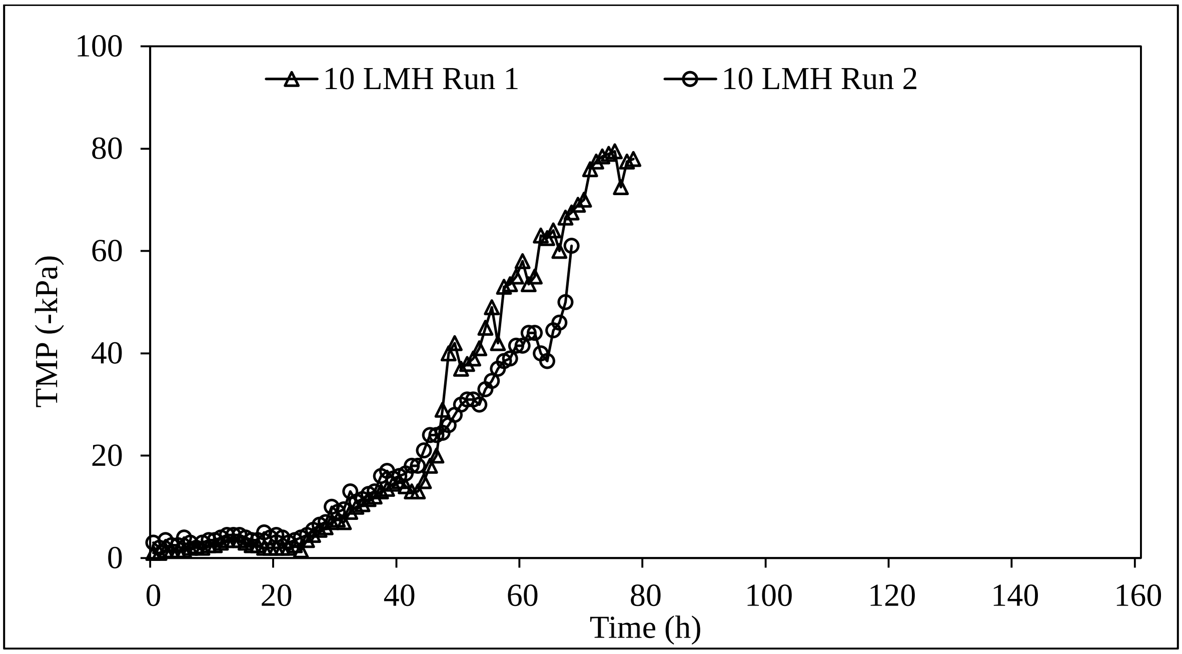


Figure 4.12 TMP vs Time for 10 LMH flux for PTFE hollow fiber membrane

After comparing three different flux, the 7.5 LMH was found to be the best in terms of the operation. Trans-membrane pressure data for all the test runs were presented in **Appendix D.**

4.2.3 COD pre-concentration performance

In conical membrane tank application, COD pre-concentration performance can be expressed in two ways same as to the WFMF system, namely; number of times that can concentrate from the initial level and system performance comparison with the common units. **Figure 4.13** shows the COD pre-concentration capability of the CMT system.

According to the **Figure 4.13** COD pre-concentration results, it can be identified that the CMT system was able to concentrate domestic sewage COD within the one week of the operation time. When comparing the 5.0 and 7.5 LMH, it can be clearly seen the increment of the COD concentration as the system filtrate higher amount of domestic sewage in 7.5 LMH flux. With 5 LMH flux, CMT system could filtrate 12.9 L/d. This number depends on the membrane area. For this experiment 0.1 m^2 membrane was used. In this case, for the one week filtration time, the system could filter 0.09 m^3. But when it comes to the 7.5 LMH, the system could filtrate 0.13 m^3. Because of this reason more COD can accumulate in 7.5 LMH flux in the particular time period.

In this experiment, triplicate tests were conducted. Based on the data it is clearly identified that the system could pre-concentrate domestic sewage by 84 times for the 5 LMH flux. For the 7.5 LMH, it could concentrate COD up to 140 times. Based on this result, it can come to a conclusion that, more the filtration higher the COD concentration. When operating with the 7.5 LMH, COD concentration capability was nearly two times higher than it works with 5 LMH. This trend can be explained in following manner.

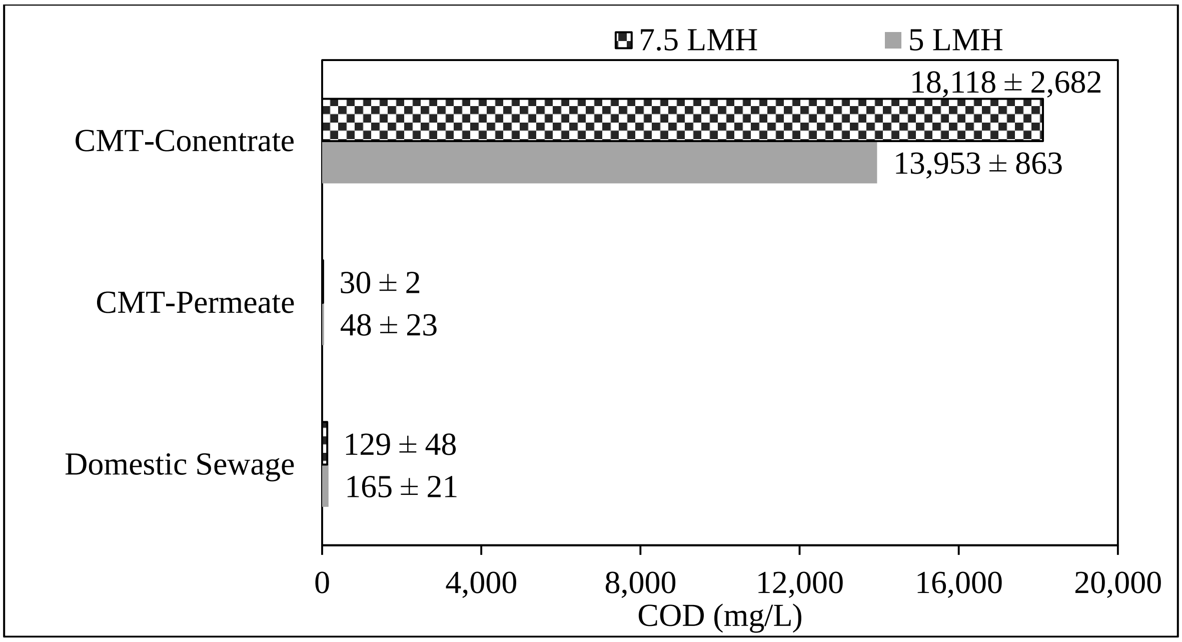


Figure 4.13 COD pre-concentration capability of the CMT system

In CMT system, COD accumulation is followed by two methods. First is, membrane filtration. When membrane separating the solid fraction, all the suspended COD can be accumulated in the system. The second thing is, conical shaped membrane tank helping to settle the solid particles in an efficient way than the regular rectangle or cylindrical shaped tank. In this case, the conical membrane tank also playing a major role on particle settling. When the flux was high, influent to the membrane tank also became high and suspended COD could settle in two different ways in the system.

The permeate COD level is lower as the PTFE-HF membrane is filtering the domestic sewage. In this case all the settleable COD is accumulating inside the system and only the soluble COD is leaving from the system. For both the flux, COD concentration of permeate was lower.

Moreover, this CMT system is a laboratory scale setup that can pre-concentrate the COD. This COD pre-concentration performance depend on the system size, tank shapes and it involves scale factors. For example, this system has comparatively small sludge cone, which has a volume of 0.9 L. In that case all the suspended particles accumulated in a small volume represent higher concentrations when it calculated for the mg/L. During COD analysis, this complete sludge cone volume was considered. When sampling for the COD, complete sludge cone volume was mixed together and got the unique sample that can represent the complete sludge cone. In that case, when the COD concentration calculated with mg/L, it represent the accumulated COD in sludge cone. For the comparison purposes, this kind of scale issues need to be avoided and it is important to bring all the comparable factors to a common unit that can be compared. In that case, g COD/m^3.d was selected as a common unit for the comparison and that can bring all the system performance in one place that can be compared. **Section 4.4** presented this common unit comparison in detail.

4.2.4 Solid accumulation of the CMT system

In the CMT system, the membrane separates the total suspended solid fraction and remove permeate with very less suspended solid concentration. Other than that, the conical shape of the tank also increasing the particle settling and accumulated more solid inside the system. The pore size of the PTFE-HF membrane is 0.1 µm and more than 90% the suspended solid content could stop passing through the membrane. **Figure 4.14** shows the experimental results for the TSS accumulation for 5.0 LMH and 7.5 LMH for CMT system.

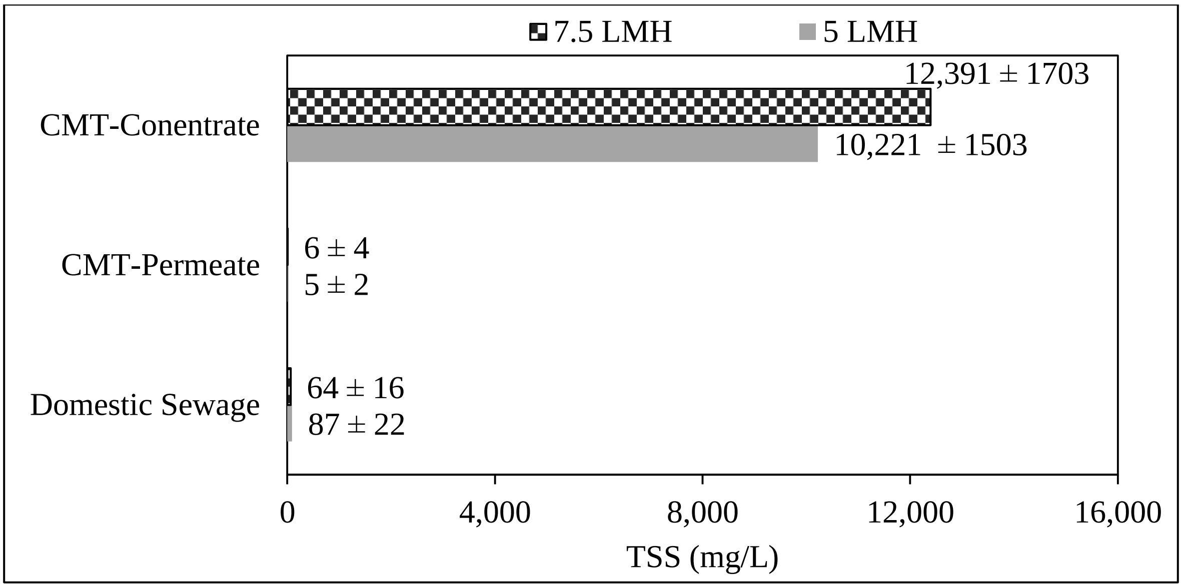


Figure 4.14 TSS accumulation for 5.0 LMH and 7.5 LMH for CMT system

Based on the TSS accumulation results, it can identify the CMT system could accumulate or concentrate TSS for both the operating flux in higher values. As mentioned in the earlier in **section 4.1.3,** the 7.5 LMH flux is removing more water while the accumulation of the TSS, than compared to the 5 LMH flux. More than 90% of TSS can be removed or hold with the CMT system. In this case, WFMF system can generate almost suspended solid free water as permeate which can have possibly using for secondary reuse application.

It has a positive trend of accumulating the TSS of the concentrate when the flux increased to 7.5 LMH from the 5 LMH. Permeate quality is still same the both flux. For 5 LMH flux, the TSS accumulation was about 117 times from the initial concentration while the 7.5 LMH having 193 times. When the membrane system operates in higher flux, the same time influent flow rate also high. In this case, the suspended solids come with the influent immediately settle in the membrane tank. This will represent the higher TSS accumulation ratio.

Conical shaped tank, enhancing the settlement of the suspended solid particles in an efficient way. In CMT system, influent was realizing to the system from the bottom. When the particles moved to the top, because of the conical shape of the tank, the settling velocity increases. Based on this reason, CMT system showing the higher TSS accumulation rates. In this case, the conical membrane tank also playing a major role on particle settling.

Considering the COD and the TSS values, it was found that the COD has a positive relationship with TSS accumulation. More the TSS accumulation, more the COD concentration. In this research, this is one of the major steps that expected. The whole idea was to accumulate more solids in the membrane system which corresponds to the higher COD concentration. Based on these results it can make a statement that the pre-concentration of domestic sewage concept can be proved with WFMF and CMT membrane systems.

4.2.5 Energy consumption of the CMT

CMT system consisted with the many of the electronic and electrical devices namely; permeate suction pump, feed pump, level control relay systems, electronic pressure gauge and data logger. Energy consumption for concentrating one gram of COD with the WFMF system, illustrates in **Figure 4.15** for 5.0 and 7.5 LMH flux.

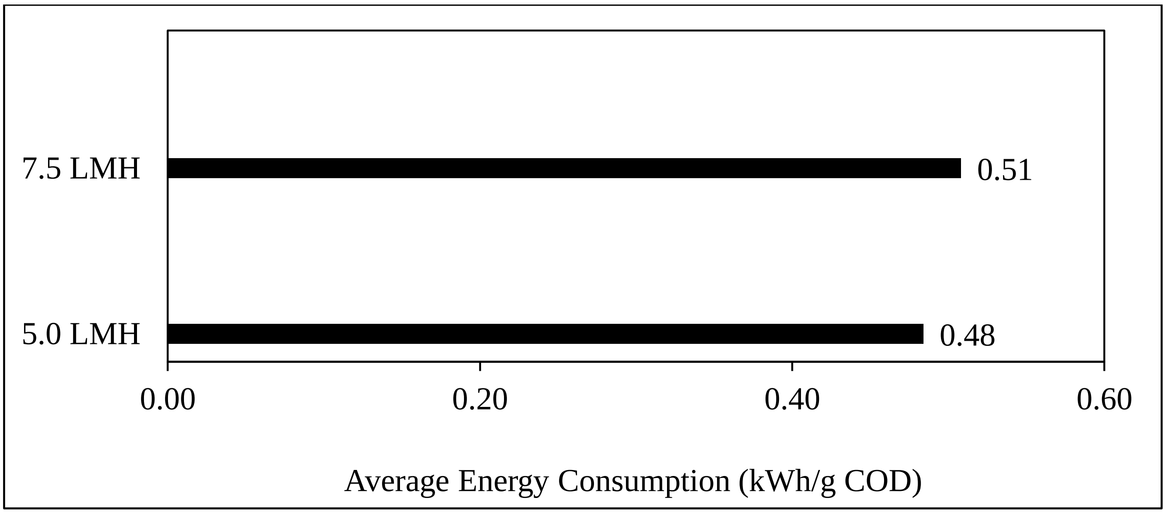


Figure 4.15 Average energy consumption for 5.0 and 7.5 LMH

Considering both the operating flux, there was a slight increment can be identified in 7.5 LMH flux. The main reason for this is, any electrical equipment needs a fixed amount of energy to start and operate in minimum level. But the slight increase of pump increment does not make a big difference in this kind of laboratory-scale models. Other types of equipment like electronic pressure gauges, data loggers have a constant energy consumption for any operating flux. Only permeate and feed pump speed is effected for the increment of the energy usage.

When the system operates with higher flux, the energy consumption is high as the pumps are working at high capacity. This is a typical scenario of the system's energy consumption. This CMT system was a laboratory-scale membrane system that consisted with two peristaltic pumps. This CMT system was smaller than the WFMF system. Especially the sludge cone volume. In that case accumulated COD load is lower in this system compared to the WFMF.

When considers the amount of COD, this scenario is coming in a different way. In this case, the energy consumption of the CMT is much higher than the WFMF system. This is because of the scale factor. When the system is on the smaller scale, energy consumption is much higher. Based on the evaluation of this system's energy consumption, it can come to a conclusion that the energy consumption comparison is not suitable for such kind of smaller scale setups. Moreover, the energy consumption data for the complete operation highlighted in **Appendix D**.

4.2.6 PTFE-HF membrane fouling and cleaning performance

As mentioned in section 4.2.1, initial membrane resistance was calculated with the pure water flux test. The initial resistance of the PTFE-HF membrane (R_m) found to be 2.47E+11 (1/m). Once the membrane fouled, cleaning procedure was carried out based on the recommended membrane cleaning procedure from the membrane supplier. Membrane resistance was calculated after physical, base and acid cleaning steps and calculated the percentage recovery of each cleaning steps.

During the 5 LMH run, only physical cleaning was carried out to see the performance with the focus of non-chemical cleaning. After the physical cleaning, pure water flux test, was carried out to evaluate the cleaning method. But based on the **Figure 4.8**, it can clearly see that the organic and inorganic fouling accumulated in the membrane and it leads to higher TMP values within 1 week of the time even with the lower operating flux. During the 7.5 LMH testing period, Physical, base and acid cleaning were conducted. **Figure 4.16** illustrates the overall results of the cleaning performance with respect to the pure water flux.

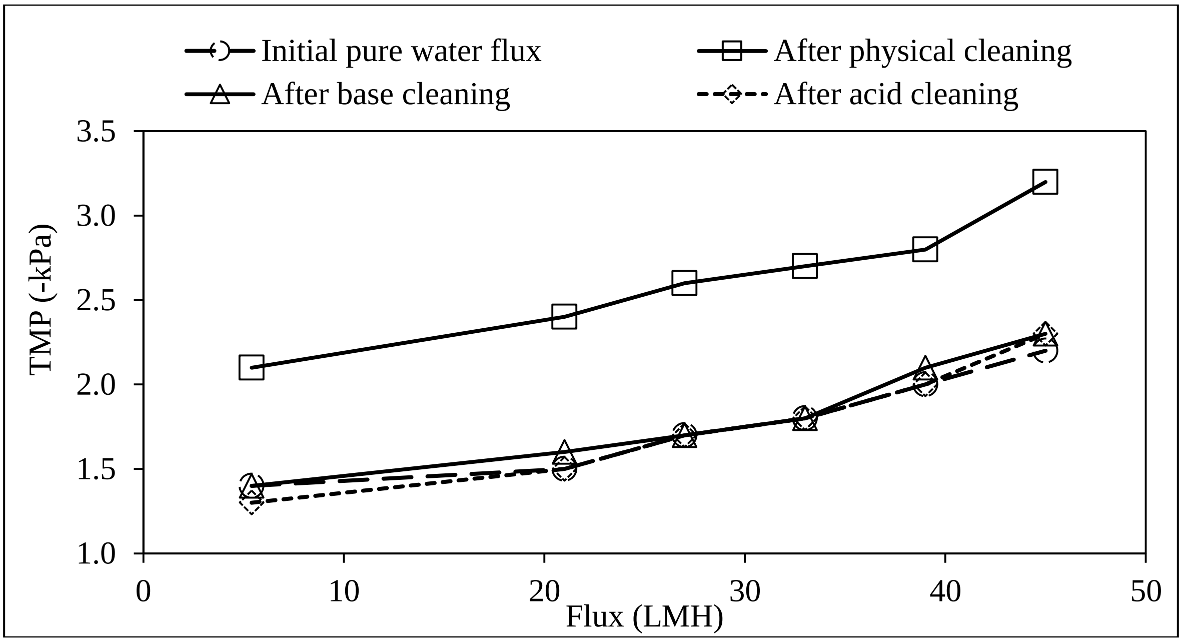


Figure 4.16 Cleaning performance of the PTFE-HF membrane

Once the pure water flux done with all the steps, resistance was calculated for each step. Initially, only the membrane resistance was there while operating with the pure water. But when it operates with domestic sewage, resistance due to the cake layer, organic fouling, and inorganic fouling plays a major role. Finally, the cleaning performance was evaluated at each step as a percentage recovery to the initial stage. **Table 4.2** shows the resistance values and the parentage recovery in each step. A pictorial view of each step of the cleaning was presented in **Appendix D**.

Table 4.2 Percentage Recovery of the Cleaning Steps for PTFE-HF Membrane

Test		Membrane resistance	After physical cleaning	After base cleaning	After acid cleaning
1	Resistance (m^{-1})	2.47E+11	3.66E+11	2.53E+11	2.51E+11
	Recovery (%)		68	98	99
2	Resistance (m^{-1})	2.47E+11	2.74E+11	2.55E+11	2.51E+11
	Recovery (%)		90	97	99

By analyzing the Test 1, it can see that the major fraction of the fouling occurs due to the cake layer which was 68% of the total resistance. This fouling can be removed physically with the water spray. Based on this step, nearly one-third of the membrane resistance was due to the organic fouling. Resistance due to inorganic fouling was very less as 1 % of the total.

By analyzing the Test 2, it can be identified that the cake layer responsible for the 90% of the membrane fouling. Also, inorganic and organic fouling responsible for only the 10%. But the chemical cleaning is necessary as the organic and inorganic fouling can be accumulated in the membrane and leads to higher operating pressure.

But this result can express the possibility of a longer run with the controlling particle accumulation on the membrane surface. This suggests the level of cake layer controlling mechanism to be coupled with the process for better performance. Moreover, the detail calculation steps for the resistance, fouling analysis and cleaning performance were presented in **Appendix D**.

4.3 Tube Settler System Efficiency

In this research, tube settler is the one which does not involve the membrane filtration like the other two systems. Typically, the loading rate is the main operational factor for the tube settlers. In this study, there were two scenarios tested with the tube settler, namely; 0.005 m^3/m^2.h loading rate and the 0.01 m^3/m^2.h loading rate with cooperating the coagulation. Tests were conducted in triplicate and evaluated the COD pre-concentration capability, TSS accumulation and the energy consumption of the system.

4.3.1 COD pre-concentration performance of the tube settler

Tube settler has a typical loading rate of 1-2 m^3/m^2.h with coagulation. In this research, initially planned to operate the system without using coagulants. As the first stage, the 0.005 m^3/m^2.h loading rate was selected as the starting point. Tube settler was tested with triplicate runs for the 0.005 m^3/m^2.h loading rate. Compared to the typical operating range, the 0.005 m^3/m^2.h was much lower. This scenario could be accepted due to non-coagulation prior to the tube settler. In this case the slow loading rate help to reduce the washing out the settleable solids fraction from the system. When considers the 0.005 m^3/m^2.h loading rate, it can pre-concentrate the COD by only 10 times with compared to its initial level. Based on the results, it can be identified that the nearly one-fourth of the COD is leaving as an effluent from the initial COD level. At this point, it can be concluded that the 0.005 m^3/m^2.h loading rate is not performing well. **Figure 4.17** illustrates the COD concentration for both the loading rates.

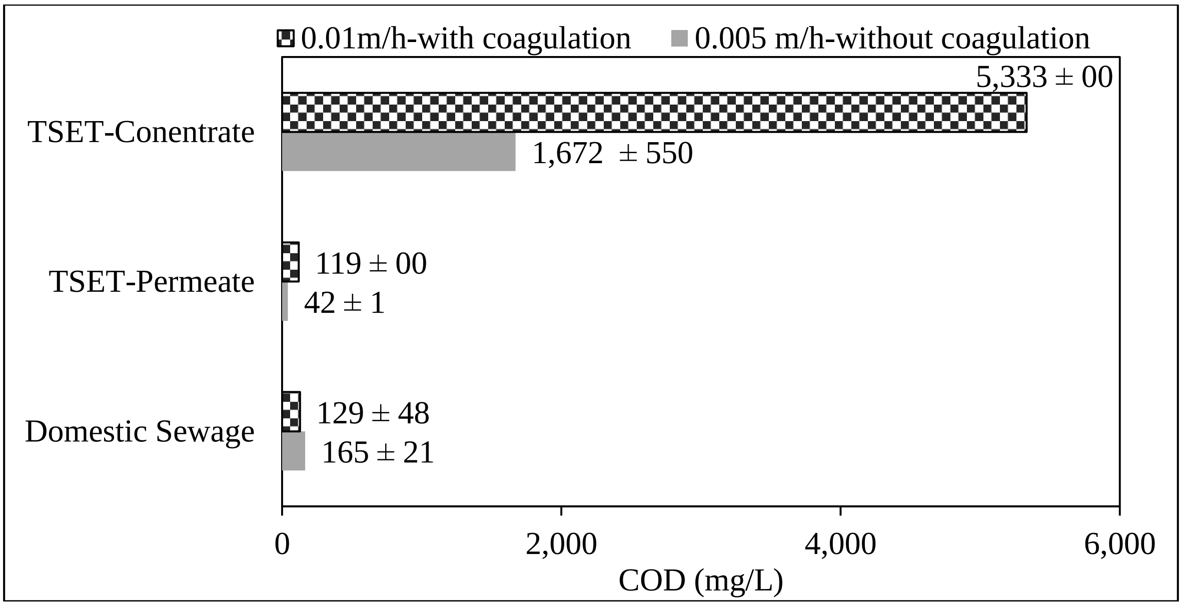


Figure 4.17 COD concentration for 0.005 m^3/m^2.h and 0.01 m^3/m^2.h loading rates

As the system did not perform well with the 0.005 m^3/m^2.h loading rate, it was suggested to increase the loading rate, with coagulation. In this case, $FeCl_3$ coagulation was introduced and increased the loading rate to 0.01 m^3/m^2.h which was twice a time of the previous test.

Considering the 0.01 m^3/m^2.h loading rate, it shows the capability of COD concentration by 45 times from the initial level. Compared to the previous test, it was achieved three times higher COD concentration. But the real issue was to appear higher COD concentrations in the permeate. Even with the coagulation, the loading rate seems to be higher due to coagulated solid partials washing out from the system and resulting in the higher COD in the permeate. A pictorial view of water sample was presented in **Appendix E**. Based on the COD concentrations, it can come to a conclusion that the system is not performing well in this condition. If only considering the COD concentration, this seems to be interesting. But when it considers the permeate quality, coagulant used, wastewater amount that treated, this is not attractive to pre-concentrate domestic sewage. Moreover, the tube settler compared with the other two systems namely; WFMF and CMT, with the common comparison units. Detail comparison presented in **section 4.4**. The COD analysis data were presented in **Appendix E**.

4.3.2 Solid accumulation of the tube settler

In the tube settler, the submerged inclined tubes help to separate the total suspended solid fraction from the domestic sewage. The whole idea was, to increase the COD concentration by capturing more solids into the system. **Figure 4.18** shows the experimental results for the TSS accumulation for 0.005 m^3/m^2.h and 0.01 m^3/m^2.h loading rates.

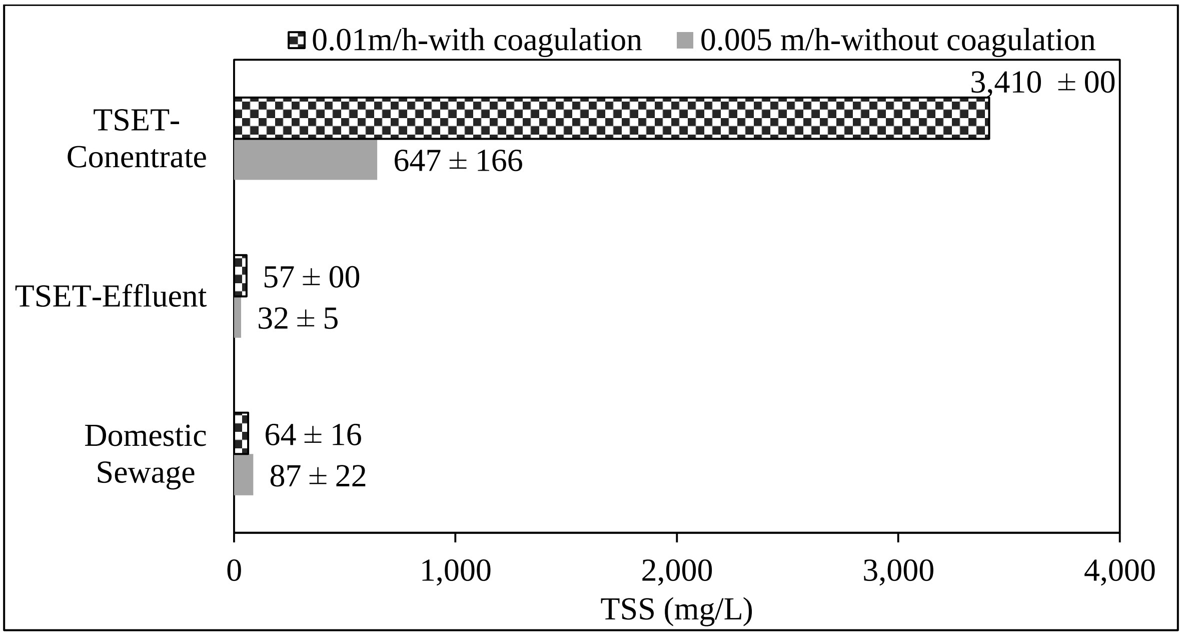


Figure 4.18 TSS concentration for 0.005 m^3/m^2.h and 0.01 m^3/m^2.h loading rates

When analyzing the 0.005 m^3/m^2.h loading rate, it can be seen that the TSS concentration only can increase by the factor of 7.4. But, when it comes to the 0.01 m^3/m^2.h loading rate with coagulation, the TSS concentration could increase by 53 times. At this point, it can clearly see the relation between the TSS accumulations with the COD concentrations. This two-factors have a positive relationship between them.

When considering the effluent of the tube settler, nearly half of the TSS is washed out from the system with the effluent in 0.005 m^3/m^2.h loading rate. This is because of the higher loading rate as there was no coagulation step prior to the tube settler. In the 0.01 m^3/m^2.h loading rate operated with incorporation with coagulation and expected higher removal of TSS in the effluent. But in this case, it was completely opposite and appeared higher TSS concentrations in the effluent. Still, the 0.01 m^3/m^2.h loading rate was high for the system even with the coagulation. It was identified that the coagulated solid particles were washed out from the system with the effluent. A pictorial view of the water samples was presented in **Appendix E**. In this situation, tube settler was not able to manage the studied loading rates with the focus of domestic sewage pre-concentration. TSS data for the complete operation, presented in **Appendix E**.

4.3.3 Energy consumption of the tube settler

The tube settler system consisted with only the feed pump, as an electric device which uses the electricity. When the system operated with coagulation, the extra dosing pump was used. Energy consumption for concentrating one gram of COD with the tube settler system, illustrates in **Figure 4.19** for 0.005 $m^3/m^2.h$ and 0.01 $m^3/m^2.h$ loading rates.

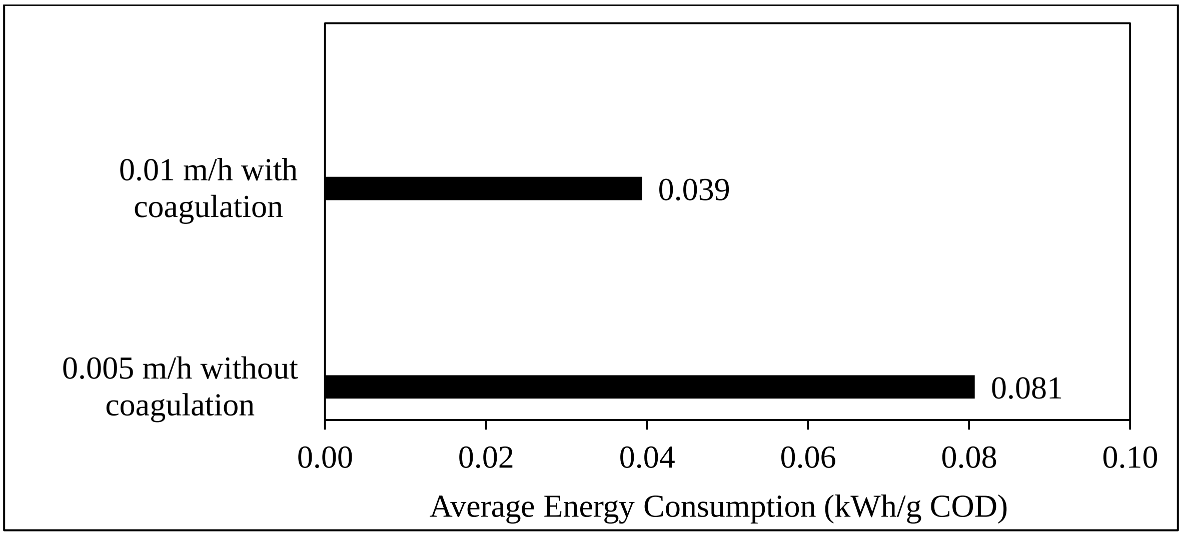


Figure 4.19 Average energy consumption for 0.005 $m^3/m^2.h$ and 0.01 $m^3/m^2.h$ loading rates

Considering both the loading rates, there was an increment of energy consumption can be identified in 0.005 $m^3/m^2.h$ loading rate. But when it comes to the 0.01 $m^3/m^2.h$ loading rate, there was a two-time increment of the energy usage than the previous test. Typically, when the loading rate is high, energy consumption also increases, due to the high efficiency of the pumps.

When considers the amount of COD that accumulated in the system, this scenario is coming in a different way. Based on the data, the energy consumption of the tube settler is much higher for the 0.005 $m^3/m^2.h$ loading rate even it was a slower loading and there was no dosing pump for the coagulation. But, 0.01 $m^3/m^2.h$ loading rate consumes less energy when it calculates for, per gram of COD. Even with the two pumps, this test showed the lower energy consumption to concentrate the COD, due to high accumulation of solids than the 0.005 $m^3/m^2.h$ loading rate. Based on the evaluation of energy consumption, it can come to a conclusion that the energy consumption comparison is not suitable for such kind of smaller scale setups. Moreover, the energy consumption data for the complete operation highlighted in **Appendix E**.

4.4 Overall Comparison of the Pre-concentration Technologies

In this section, the complete results of the pre-concentration technologies were summarized and compared in terms of the COD pre-concentration capability, solid accumulation and the energy consumption. Based on this evaluation, the best pre-concentration technology was finalized for coupling with the AnMBR process. When analyzing all the test results, there are two factors that need to be considered, namely: 1) mass balance approach, 2) comparison with the common unit. It is important to have a chance to explain this pre-concentration trend with a mass balance approach as it is important in real life application.

Also, the COD concentration depends on the system volume, sludge cone volume, filtration rate, filtration time. The data should bring in the same level that can be comparable. In this case, **g COD/m^3.d** was considered as a common unit for the comparison.

4.4.1 Mass balance approach in pre-concentration

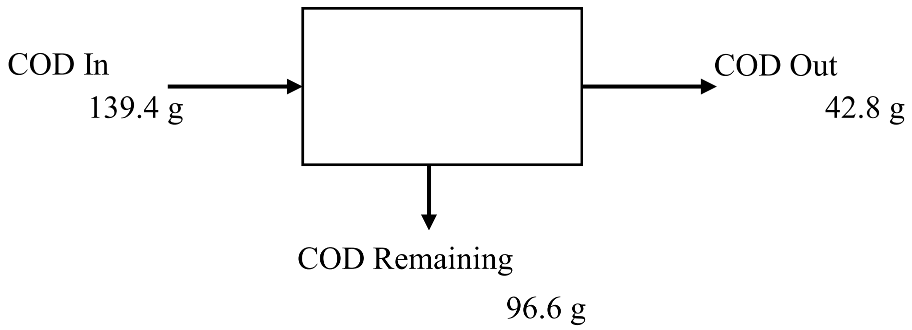
The three different systems had a different size of the tank volumes. Also, the sludge cone volumes were different from each other. In this mass balance approach, there were two scenarios that came to a consideration, namely; total system volume and the sludge cone volume. This research work based on the idea of capturing the solids to increase the COD in domestic sewage to enhance the anaerobic digestion. In this case, solid accumulation plays a big role in terms of COD concentration. These two scenarios projected different answers based on the volume of sludge cone or the complete system. It was important to find the best fit scenario for the mass balance approach. This can be explained with results related to the 5 LMH membrane flux for WFMF and CMT systems. Mass balance calculations for the 5LMH flux is presented in **Table 4.3**.

Table 4.3 COD Mass Balance Calculations for 5 LMH flux

	WFMF	CMT
Wastewater inflow to the system (L/d)	120	13
Amount of water that treated in a run (L)	120*7= 840	13*7= 91
Average COD of domestic sewage (mg/L)	166	166
COD load- In (mg)	840*166 = 139,440	91*166= 15,106
Permeate COD (mg/L)	51	47
COD out- with the permeate (mg)	840* 51 = 42,840	91* 47 = 4,277
COD remaining inside the system (mg)	139,440- 42,840 = 96,600	15,106- 4,277= 10,829
COD remaining inside the system (g)	**96.6**	**10.8**

This mass balance calculation can be express in graphically for better understanding. **Figure 4.20** illustrates the COD mass balance for WFMF and CMT system for 5 LMH flux. In this case, the real COD data for domestic sewage and the permeate, was used and theoretically calculated the accumulated COD in the system. At this point, the volume factor comes into the consideration.

For WFMF- 5 LMH Run



For CMT- 5 LMH Run

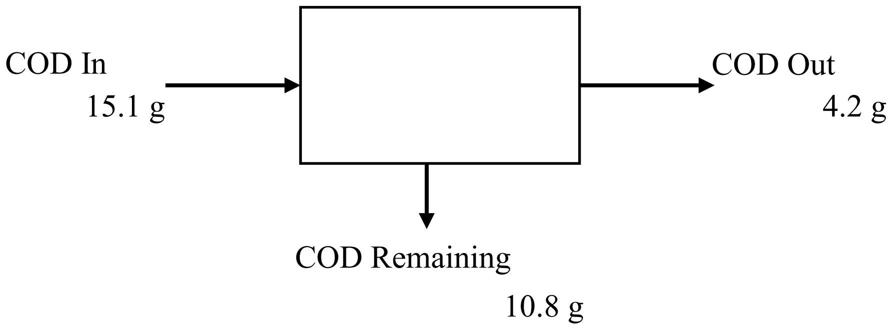


Figure 4.20 COD mass balance for WFMF and CMT system for 5 LMH flux

The two scenarios were considered to finding out the best fit mass balance approach for this study. In this case, complete system volume and the sludge cone volume were considered in terms of the COD concentration. **Table 4.4** and **4.5** presents the theoretical COD concentration for these two scenarios.

Table 4.4 Theoretical COD Concentrations when Considers the Full System Volume

	WFMF	**CMT**
Full tank volume (L)	130	23
COD concentration (mg/L)	96,600/130 = 743	10,829/23 = 470

Table 4.5 Theoretical COD Concentration when Considers only the Sludge Cone Volume

	WFMF	**CMT**
Sludge cone volume (L)	23.5	0.9
COD concentration (mg/L)	96,600/23.5 = 4,110	10,829/0.9 = 12,032

It can be seen that the COD concentration showed the huge difference when using this two volume. For the mass balance approach, it is important to find the suitable scenario. In this situation, theoretical COD concentrations and the practical COD values were compared.

Table 4.6 shows theoretical and practical COD concentrations of the WFMF and CMT systems for 5 LMH flux when considers the sludge cone volume.

Table 4.6 Theoretical and Practical COD Concentrations of the WFMF and CMT Systems for 5 LMH Flux when Considers the Sludge Cone Volume

	Theoretical COD (mg/L)	Practical COD (mg/L)
WFMF	4,110	6,047
CMT	12,032	13,953

It can be identified that when considers the sludge cone, the theoretical and practical COD concentrations were nearly same. It is important to use the sludge cone volume as it contains the major part of the COD as suspended solids. At his point, it can come to a conclusion that the considering the sludge cone volume for the mass balance approach will be a more appropriate than the full system volume. By using 5 LMH flux, it can calculate the theoretical COD concentrations for the 7.5 LMH with the mass balance approach. 7.5 LMH is 1.5 times higher the 5 LMH flux. So the flux increment factor can be considered as 1.5 for this case. **Table 4.7** presents the COD concentrations based on the flux increment factor.

Table 4.7 COD Concentrations based on the Concentration Factor

COD concentration (mg/L)	5 LMH Run COD (mg/L)	Flux increase factor	7.5 LMH Flux-Theoretical-Expected COD (mg/L)	7.5 LMH Flux-Practical COD (mg/L)
WFMF	6,047	1.5 X	9,070	7,900
CMT	13,953		20,929	17,900

By analyzing **Table 4.7**, it was identified that the concentration was increased with the flux. Practical COD values were nearly same to the theoretical calculations. This is one of the important findings of this research work to show that the mass balance approach can be accepted by considering the sludge cone volume rather than the full system volume.

4.4.2 System comparison with the common unit

As mentioned in the earlier, the system efficiency depends upon the scale factors such as sludge cone volume, total volume, etc. In this case, it is important to compare the systems with the common comparable unit. Based on the COD concentrations, domestic sewage that treated, time duration factors, the common unit was generated which can be compared. There were 3 different common units were used to evaluate the pre-concentration technologies, namely; (g COD/ m^3. d), % TSS accumulation, (kWh/ g COD). There were three major factors that considered in the evaluation, namely; COD, TSS, and the energy consumption.

The COD concentrating performance of the WFMF technology indicated 21 to 24.2 g COD/ m^3.d while CMT has 17.5 to 19.7 g COD/ m^3.d concentration capability. Tube settler application indicated the lower concentration capacity for the loading rate of 0.005 m^3/m^2.h, which was 1.8 g COD/ m^3. d. Moreover, even with the coagulation, tube settler could achieve only 2.6 g COD/ m^3.d for 0.01 m^3/m^2.h loading rate. **Table 4.8** summarizes the experimental results on pre-concentrating the domestic sewage.

Table 4.8 Performance Comparison of the Pre-concentration Technologies

	Membrane Flux				Loading Rate	
	5 LMH		7.5 LMH		0.005 $m^3/m^2.h$	0.01 $m^3/m^2.h$
	WFMF	CMT	WFMF	CMT	TSET	TSET
COD of domestic sewage (g/L)	0.14 ± 0.05					
COD of the concentrate (g/L)	6.0	14.0	7.9	17.9	1.8	5.3
Sludge cone volume (L)	23.5	0.9	23.5	0.9	18	18
Total COD in sludge cone (g)	142	13	186	16	33	96
Sewage treated per run (m^3)	0.8	0.1	1.3	0.1	2.7	5.4
Test duration (days)	7		7		7	
Concentrating ability (g COD/ m^3.d)	24.2	19.7	21.0	17.5	1.8	2.6
TSS in domestic sewage (g/L)	0.08 ± 0.03					
TSS in concentrate (g/L)	3.20	10.22	5.51	12.07	0.60	3.41
TSS in permeate/ effluent (g/L)	0.008	0.005	0.007	0.01	0.032	0.06
TSS accumulation %	90.8	94.3	89.6	92.5	63.2	38.4
Energy consumption (kWh/ g COD)	0.045	0.485	0.035	0.508	0.081	0.039

The CMT system shows the highest solid accumulation ratio, which is more than 92.5 % for 5.0 and 7.5 LMH flux. WFMF system also showed more than 89.6% of solid accumulation. Compared to the membrane systems TSET showed the lower solid accumulation percentage. TSET could only accumulate 63.2% of the TSS of 0.005 m^3/m^2.h loading rate. 0.01 m^3/m^2.h loading rate shows the lowest TSS accumulation due to washing out the particles even with the coagulation. TSET system's COD and TSS capture performance were lower among others. **Figure 4.21** shows the graphical variations for the comparison.

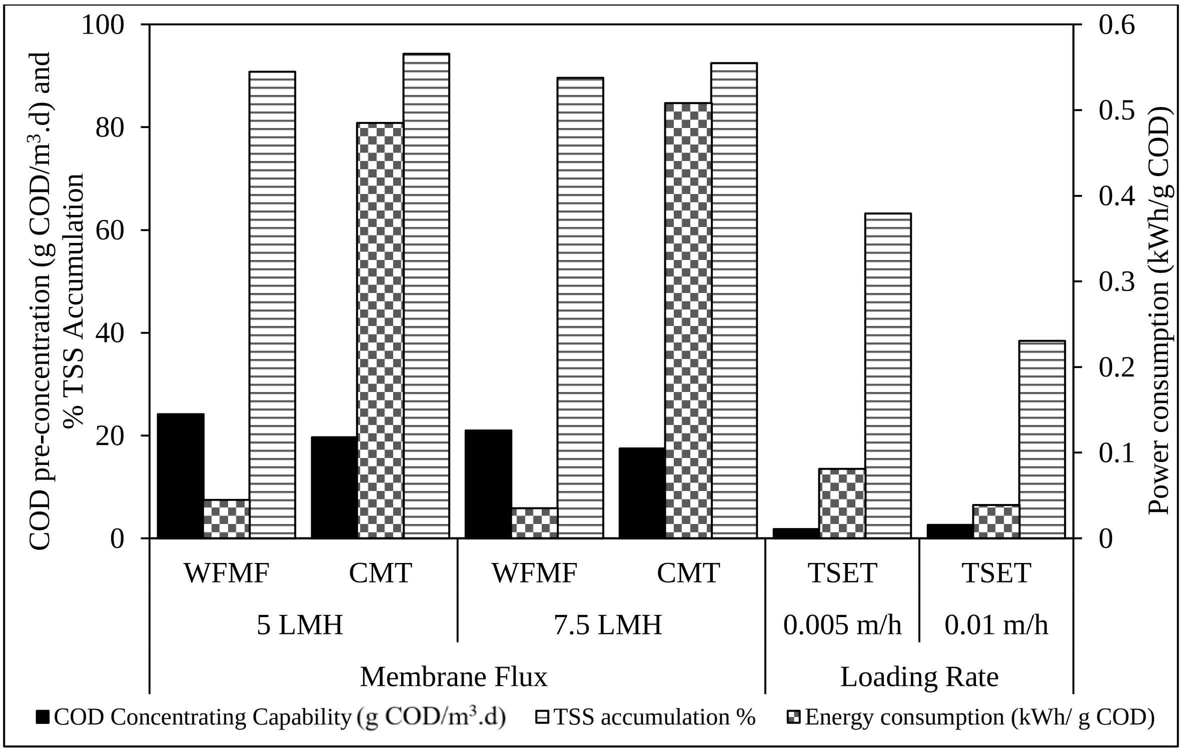


Figure 4.21 Overall comparison of the pre-concentration technologies

WFMF system showed the higher COD concentration ability and the nearly 10 times lower energy consumption compared to the CMT system. Thus, the performance of WFMF 7.5 LMH flux was the best among three technologies, in terms of its low energy consumption, higher COD concentration ability, and the higher TSS accumulation for domestic sewage pre-concentration.

4.5 Anaerobic Membrane Bioreactor

In this study, concentrated domestic sewage was used as a feed to the AnMBR system. The main objective of this experiment was to prove that the concentrated domestic sewage can be anaerobically digested and then find out the efficiency of the process by evaluating the biogas generation and methane content.

4.5.1 Biogas production of the AnMBR

AnMBR system was operated for two month period. After the acclimatization period, the concentrated domestic sewage was introduced to the system as the feed. Biogas production and pH variation of the system illustrate in **Figure 4.22**.

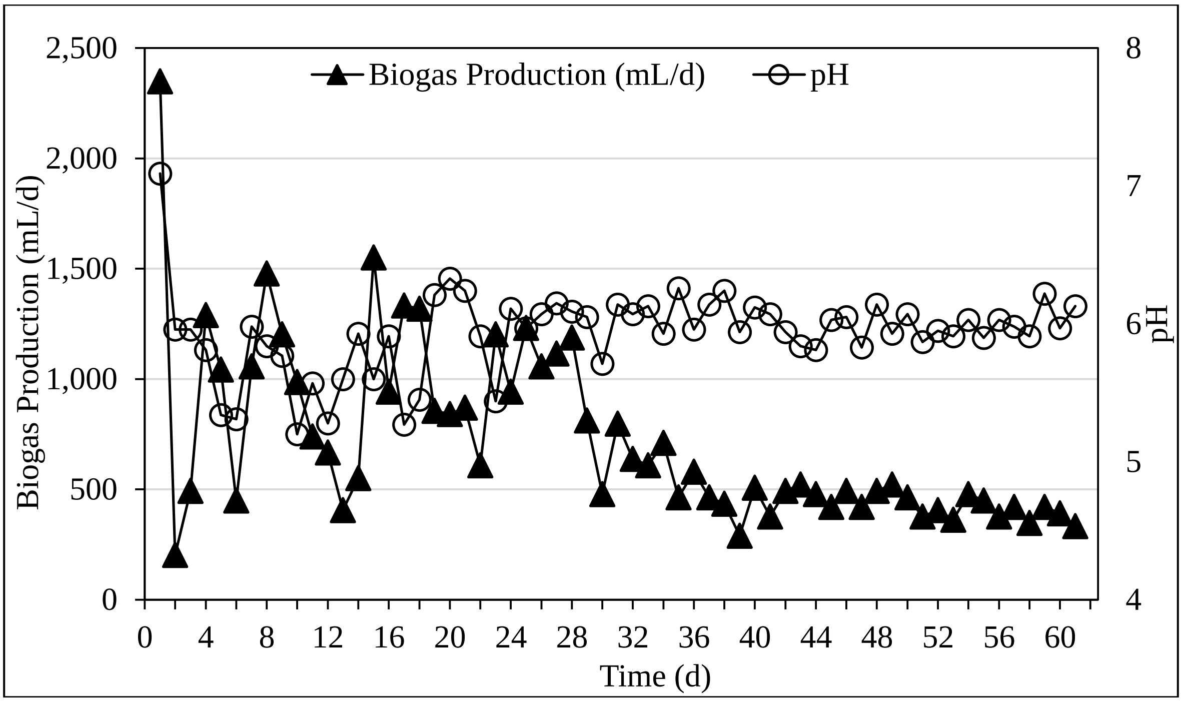


Figure 4.22 Biogas production and pH variation of the AnMBR

From day 1-7, the acclimatization period was continued and slowly introduced the concentrated domestic sewage to the reactor. From 8th day, the reactor was started to operate only with concentrated domestic sewage. Once started the feeding of domestic sewage, the biogas production was kept reducing with the time. In this situation, it suggested cross-checking the system performance with the glucose to confirm that the real feed has some issue of degradation in this contest. In that case, started to feed glucose solution to the reactor with the loading rate of 3.2 kg COD/m^3. d. At day 15, started the glucose feeding and it results in the sudden increment of biogas generation as 1.5 L/d. Based on that observation, it can be concluded that the concentrated domestic sewage is not easily degraded at the current operating condition.

From day 23, again started to introduce the concentrated domestic sewage to the system and continued only with real feed after day 27. At this point, again the biogas generation was reduced and the trend was almost same to the previous attempt. After that, the system continuously operated with concentrated domestic sewage to see the long-term operational performance. In this case, the AnMBR system continues with the concentrated domestic sewage over one month period. During this period, the system could generate 490 mL/d of biogas as an average. Biogas generation fluctuated the range of 544 ± 254 mL/d.

The biogas generation was lower as 0.028 L/g COD. This is much lower compared to the theoretical value. At standard conditions, one gram of COD could generate 0.35 L of methane. This study evaluated the performance of the AnMBR with the loading rate of 3.2 kg COD/m^3.d with 2.18 HRT. But this biogas production rate can be improved by optimizing the AnMBR process. Initially, the pH of the system was fluctuating the range of 5.7 ± 0.6. But after day 27 onwards, it can see that the pH was stable. It was observed that the pH increased up to 6. During this period, fluctuated the range of 6 ± 0.2. Biogas production and pH data for the complete experiment presented in **Appendix F**.

To address the low biogas generation issue, it is important to maintain the lower cross-flow velocity in the system. Higher cross-flow velocity can disrupt the bio-floss. Moreover, high cross-flow velocity generating in high shear intensity can negatively effect on biological activity. The loss in microbial activity was due to a reduction in bio-flock size, which in turn interrupted the syntrophic association between acidogenic and methanogenic bacteria. Therefore, it is important to maintain the lower cross-flow velocity in the system. Typically 1 to 5 m/s cross flow velocities maintains in AnMBR systems.

Generally, recirculation amounts depend upon the cross flow velocity. Even with the lower cross-flow velocity, recirculation rate can be large due to the smaller reactor size. This is one of the major fact that need to be addressed when the reactor size is small. If the recirculation rate is high, it increases the mechanical strain on the biological activity (Brockmann and Seyfried, 1997). Typically, the full-scale plants maintain the cross- flow velocity of 1-5 m/s. In this experiment, 0.073 m/s recirculation velocity was maintained. This is very low compared to the full- scale plant's condition. Lower cross-flow velocity is acceptable as the working volume of the bench scale setup was 6 L.

In this research, the main focus was to capture more organic solid fraction from the liquid waste. Basically, it is a solid-liquid fractionation with the focus of COD pre-concentration. In this case, it was important to identify that there is not a considerable level of biodegradation while pre-concentrating. To determine that, DO level of the concentrated domestic sewage was measured. It was found that the DO of concentrated domestic sewage is less than 1.14 mg/L for all the measurements. WFMF systems HRT was calculated for 7.5 LH flux and found to be the 0.7 d. Considering the DO levels and the HRT values, it can be concluded that there is no significant reduction of COD due to biodegradation.

4.5.2 Methane content of the biogas

From day 36 onwards, methane content measurement was started. Methane content results were illustrated in **Figure 4.23** for the AnMBR operation. The results on day 36, showed the higher amount of nitrogen gas content of the biogas as 62%. This is due to the nitrogen sparging after fixing the level sensor clogging issue of the reactor. Nitrogen gas had to inject to the reactor that makes sure it has oxygen free environment once the reactor open for a short time. In this case, it shows only 11% of methane content in the biogas.

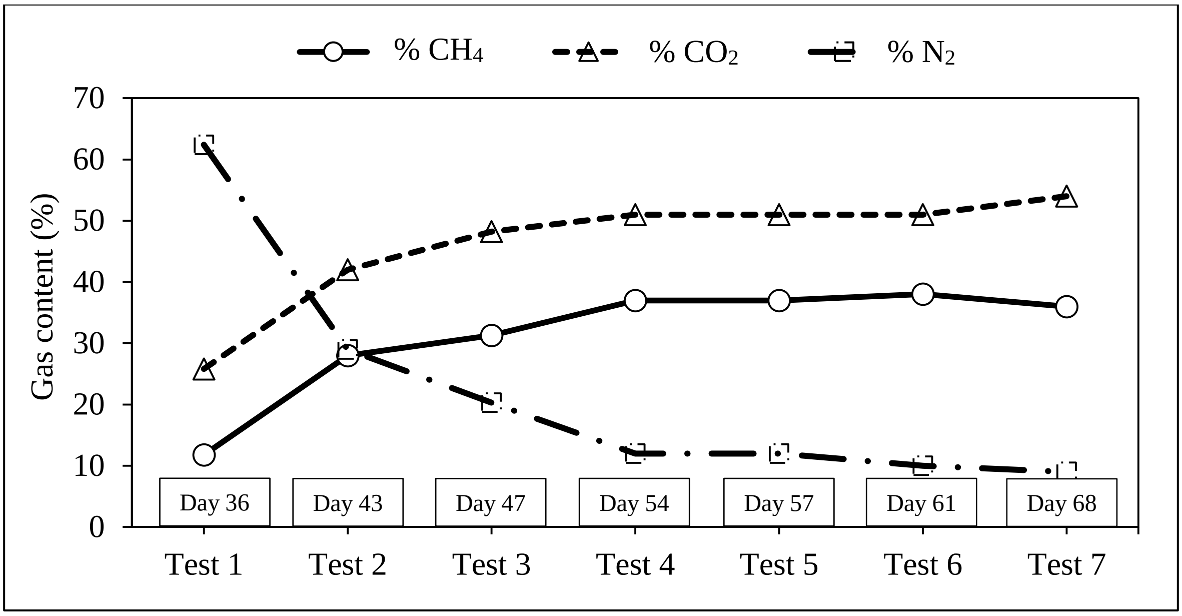


Figure 4.23 Gas composition of the biogas

It can be seen that the gradual increment of the methane content of the biogas over the study period and stable after, day 54. Methane content showed the stabilized value as 37% - 38% in the final stage of the study period. In the same way, the nitrogen was flushed out from the system and results in the lower nitrogen gas content of 9 % on the 68th day. CO_2 also increased with the time and it can see the same trend of increment with the methane and finally stable after, day 54. The content of the CO_2, stabilized at 51 - 54% at the final stage of the study period. Compilation of analytical data for methane content, presented in **Appendix F**.

4.5.3 Removal efficiencies of the AnMBR

COD, BOD and TSS removal rate were studied for this operation while it runs with the stable condition. Removal efficiencies present in **Figure 4.24**.

TSS removal of the AnMBR was nearly 100% as the permeate come through 0.1 µm ceramic membrane. This microfiltration helps to stop almost all the suspended solids in the system. During the operational period, the COD removal rate of the system reported the range of 71 ± 4.74 %, while BOD removal observed 67 ± 2.93 %. With the particular time period, this was the reported removal rate of 3.2 Kg COD/m^3. d loading rate with 2.18 d HRT in mesophilic condition.

Especially, COD and BOD removal rates can be increased by optimizing the AnMBR process. There are many factors that can optimize in the system such as mixing condition, mixing method, recirculation rate and time, membrane filtration method and frequency, operating temperature. In this bench scale setup, peristaltic pumps were used as mixing and cross-flow pump. Those pumps were rotary type pumps. It was reported that some pump types (rotary pumps) effect in the cell and the flocs, which resulting particle size decrement and increment of the soluble organics in anaerobic digestion systems. Rotary pump generating greater shear to the microbial flocs than a centrifugal pump the same operating conditions in membrane bioreactor operations. The study found that the COD removal efficiency, specific oxygen uptake rate, sludge yield can be affected with the pumping device even loading rates and the operating condition was the same (Kim et al., 2001). When optimizing the AnMBR process, it is important to consider these factors as it affects the biogas generation.

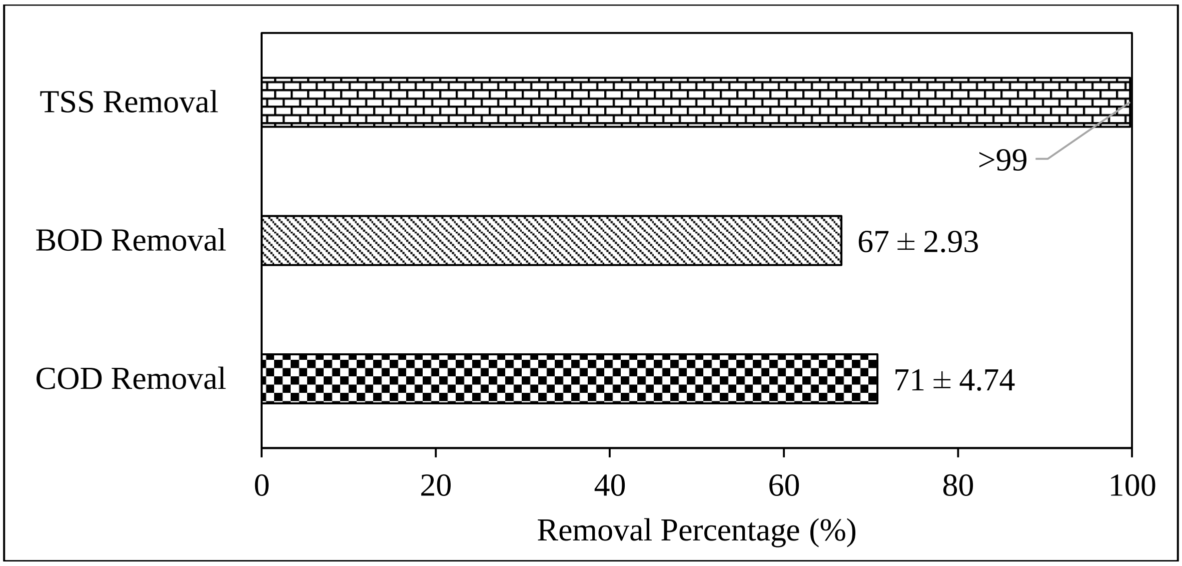


Figure 4.24 Removal efficiencies of the AnMBR

Moreover, when it needs to increase the biogas production, and the removal rates, it is important to focus on the shear effects of the system. The shear effect depends upon the configuration of the experimental setup. Including, pumps and valves. They can provide different amount of shear. The loss of biomass activity can also be a problem of a scale in laboratory bench scale modules. In practice, the shear effect of full-scale equipment can have a significant difference from lab scale pumps and valves (Dereli et al., 2012). In this case, it is better to practice the inside propeller mixing, rather than pump recirculation mixing conditions. This 3.2 Kg COD/m^3.d loading rate can be identified as a sustainable starting point for the anaerobic digestion. In that case, it is better not to reduce this loading rate further for optimization. It is always better to maintain this current loading rate and change the other factors to improve the removal rates. According to the current removal efficiencies, it was observed that the higher amount of COD and BOD is leaving with the permeate as dissolved form. This scenario is not recommended and it is important to further optimize the AnMBR process. Removal data presented in **Appendix F**.

4.5.4 Membrane performance of the AnMBR

In this setup, a tubular ceramic membrane was used to separate the biomass from the effluent water. In that case, the higher amount of biomass could maintain inside the reactor for higher performance. It was able to maintain the MLSS of 10,863 ± 250 mg/L in the anaerobic reactor. Also observed the MLVSS of 6,307 ± 666 mg/L. Data for the MLSS and MLVSS presented in Appendix F. Over the study period, the trans-membrane pressure was measured and illustrated in **Figure 4.25**.

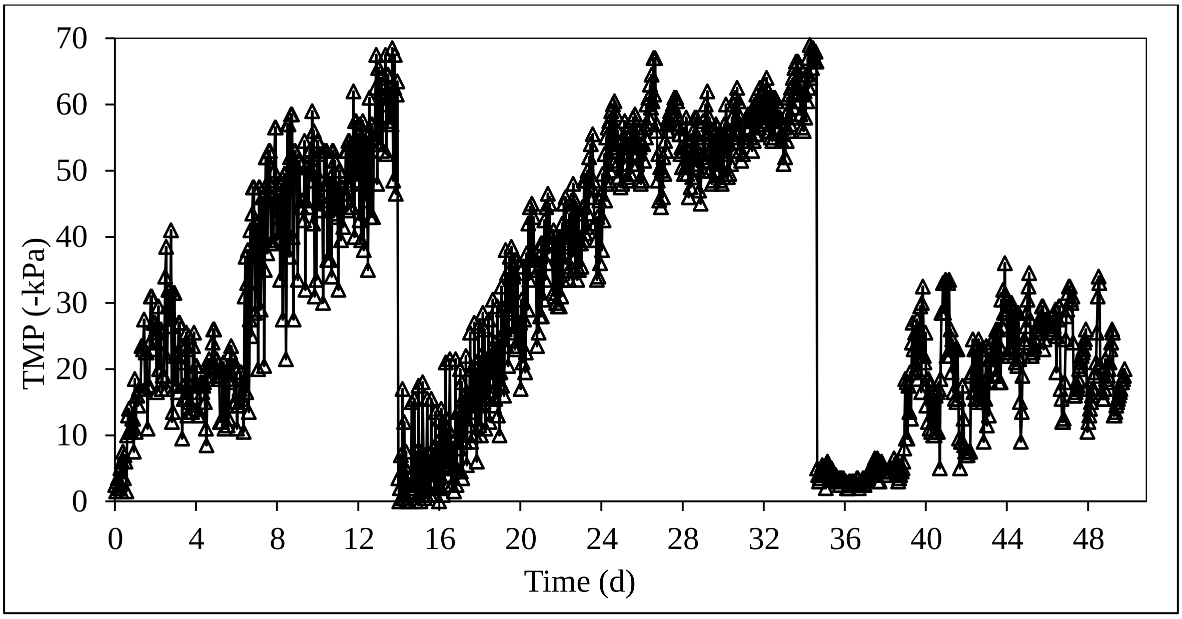


Figure 4.25 Transmembrane pressure variation vs time for the AnMBR

Figure 4.25 presents the compilation of the three continues filtration cycles of the ceramic membrane. The first cycle was able to operate Up to 14 days, till it needs a membrane cleaning. The second filtration cycle was able to operate 21 days. Third filtration cycle did not come to the cleaning point when the operation was stopped after a particular time. When analyzing the first and second filtration cycles, it can be identified that the second cycle was longer than the first cycle. Second cycle able to operate 7 days more, than the first one. This scenario has an explanation related to the filtration time intervals. During the first cycle, permeate pump was worked with the intermitted time mode. In this cycle, it was worked 10 minutes after the 30 minutes of relaxation time. In this cycle, it was identified that the membrane is clogging fast as the sedimentation happening inside the tubular membrane.

When it comes to the second cycle, it suggested reducing the relaxation time of the permeate pump to maintain the membrane clogging due to sedimentation inside the membrane. In this case, permeate pump setup to operate 5 minutes after 15 minutes of relaxation time. In this situation, the membrane was able to operate longer period as it was able to manage the clogging effect inside the membrane. When the TMP increase up to -70 kPa, the membrane was not able to maintain the constant filtration flux. Chemical cleaning was conducted once it reached to this cleaning point.

4.6 Potential Application of this Concept

Domestic sewage pre-concentration concept is more convenient with the decentralized approach. **Figure 4.26** illustrate the concept of the applicability.

This concept applies with decentralized wastewater treatment facility that handles less than 1000 m^3/d. As indicated in figure 5.2, the pre-concentration step can be done before and send the concentrate into the common anaerobic digestion reactor which can effectively utilize the organic fraction of the domestic sewage. Depending on the application and initial feed water quality, the permeate water can have a reuse potential in the nearby community. Mostly this can be an agricultural application. Depend upon the domestic sewage, sometimes permeate need further treatment. In that case, permeate can be released to the main sewer line to transfer it to the decentralized treatment plant. Finally, the biogas can be converted into electricity and use in the decentralized wastewater treatment plant or the nearby community.

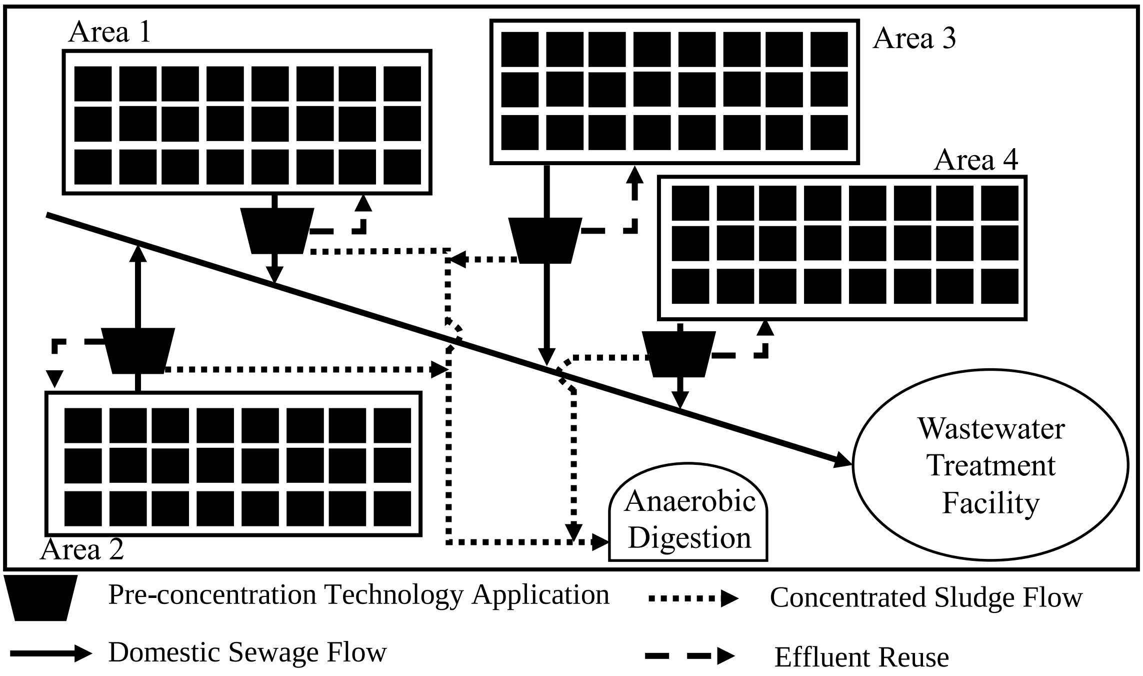


Figure 4.26 Concept of the applicability

As explained in section 4.5, the whole idea was to recover the energy from the domestic sewage while treating it. This research did not focus only on treating the COD. If the domestic sewage uses this concept, it needs to pre-concentrate because the initial COD is lower. If pre-concentration is not done, there will be three major drawbacks. Namely; need of larger reactor volume, lower biogas production, economically not attractive due to no methane extraction even with the anaerobic process. In such scenario, pre-concentration of the domestic sewage is attractive prior to the anaerobic process in terms of energy recovery. In this case, membrane permeate have a potential for agricultural purposes. Permeate water can bypass the anaerobic process and be used in secondary application. So, this water volume represents the reactor volume reduction. For example, in 7.5 LMH flux of WFMF system, can generate 1.26 m^3 of permeate while treating 1.4 m^3 of domestic sewage. That represent 90 % of volume reduction compared to the initial volume. In that case, the concentrate of the WFMF system has high COD concentration due to liquid fractionation. Thus, this concentrated domestic sewage can be use in an efficient way due to high COD concentration. The major point of this case is generating the biogas while treating the domestic sewage. Anaerobic treatment system can be optimized to have higher biogas generation due to the initial high concentrations after the pre-concentration. This type of technology is highly fit in where it has a possibility of water reuse application. As highlighted in figure 4.26, this type of system can be applied where it has a decentralized wastewater treatment plant with the capacity of 1,000 m^3/d. Pre-concentration technique with membrane systems is not attractive in household level. It is better to bring the concentrated domestic sewage to the common reactor for enough biogas production that can be converted into the energy.

4.7 Overall Picture of the Study

This research was compared three different domestic sewage pre-concentration technologies and studied the performance of an AnMBR with concentrated domestic sewage. What have been covered and other potentials illustrated in **Figure 4.27.**

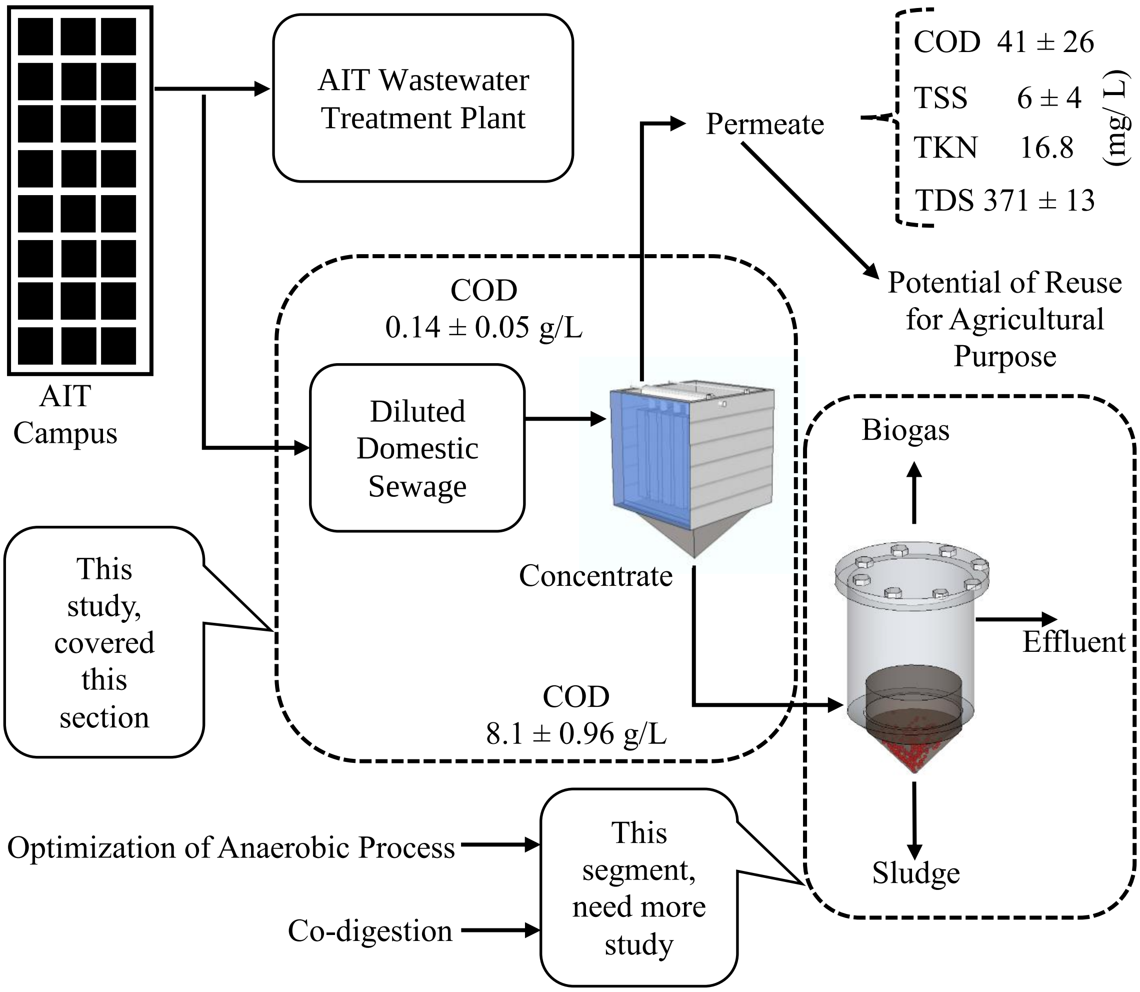


Figure 4.27 Overall picture of the study

The first segment of the study focused on concentrating the diluted domestic sewage for anaerobic digestion. By comparing three of the technologies, WFMF performed well in this contest and selected as a best among others. WFMF was able to concentrate domestic sewage up to 8,000 mg/L within a 1 week period. This depends upon the membrane area and the operating flux of the membrane. This study proved that the WFMF performance were high on pre-concentrating the domestic sewage. Other part of the study was to investigate the performance of the anaerobic process with concentrated domestic sewage. AnMBR system was able to generate 0.028 L/g COD of biogas with concentrated domestic sewage which is a comparatively lower biogas production. AnMBR process needs to be optimized to increase the biogas production. In this study, only one loading rate and HRT was observed in the AnMBR process. In that case, this segment needs more study to evaluating the AnMBR process with concentrated domestic sewage.

Moreover, there is a potential of reuse the permeate water. It can be seen that the permeate water is free from settleble COD. In this study, effluent water has COD of 41 ± 26 mg/L. In

Thailand, discharge limit of COD is 120 mg/L. In this situation, WFMF permeate is under the discharge limit. In terms of the TSS, WFMF permeate shows 6 ± 4 mg/L. This is very small compared to the discharging limits of TSS in Thailand as 50 mg/L. Microfiltration results the almost TSS free water that has a great potential of reuse. Considering the nutrient values, it also seems to be less. But microfiltration can not remove the dissolve ions in the water. If that is the case, the dissolved nutrients can be pass through the membrane as TDS. TDS of the permeate water was found to be 371 ± 13 mg/L for this domestic sewage. Also the TKN of the permeate found to be 16.8 mg/L. As mentioned earlier, dissolved nutrients can be pass through the membrane. For this feed water, it has a great potential of using for the agricultural activities as it contains the nutrients. Microbiology of the permeate water need to be studied before using it on any of reuse applications. The woven fiber membrane has 1-3 μm pore size. In that case, the permeate water may not be microbilogically safe. Depending on the reuse application, the required quality of the water can be differentiated. This part needs to be further investigated for better reuse application.

Chapter 5

Conclusions and Recommendations

The goals of this work were to assess performance of three different pre-concentration technologies; woven fiber microfiltration (WFMF), tube settler (TSET) and conical membrane tank (CMT) that can apply to, concentrate the domestic sewage prior to the anaerobic treatment. The performance was then evaluated based on the COD concentration ability, total suspended solid accumulation, and the energy consumption by each of these three pre-concentration membrane technologies. At the final stage, AnMBR was operated with concentrated domestic sewage to evaluate the anaerobic digestion performance. Based on the results observed, following conclusions and recommendations have been drawn.

5.1 Conclusions

According to the first objective, the pre-concentration performances of the technologies were investigated. The major conclusions of pre-concentration technology performance, are summarized below:

1. WFMF technology indicated 21 to 24.2 g COD/ m^3. d while CMT has 17.5 to 19.7 g COD/ m^3. d of COD concentration ability. Tube settler application indicated the lower concentration capacity for the loading rate of 0.005 m/h, which was 1.8 g COD/ m^3. d. Moreover, even with the coagulation, tube settler could achieve only 2.6 g COD/ m^3. d for 0.01 m/h loading rate. WFMF system showed the higher COD concentration ability and the nearly 10 times lower energy consumption compared to the CMT system. Based on the overall results, WFMF 7.5 LMH flux was the best among all scenarios, in terms of domestic sewage pre-concentration for anaerobic digestion.

2. Capturing solid fraction from the domestic sewage can leads to generate the higher COD concentrations that can be effectively used for the anaerobic digestion process.

3. Pre-concentration capability of the WFMF and CMT systems were better than the tube settler application. Microfiltration membranes can capture most of the suspended solid in the domestic sewage which mainly corresponds to the settleable COD of the sewage.

4. Only the physical cleaning methods are not suitable for the membrane cleaning while it operate with the domestic sewage at it contains organic and inorganic foulants. In this case, it need to be conducted the chemical cleaning procedures for better achievements.

5. There were many factors that effect for the capturing TSS from the domestic sewage. Particle settling and concentrations depend upon the system volume, sludge cone volume, tank shape, operating time, membrane area, and surface area. To compare the different technology performance, the comparable factors needed to bring it to the same level that can be compared. To make the common comparable unit, the amount of the COD, filtration time and volume was considered. **g COD/ m^3. d** was used as a common unit to evaluate the systems.

6. Evaluating energy consumption for bench scale process is acceptable. But not recommend to directly convert that type of energy data to full-scale operations. As the bench scale systems were small, the economy of the scale is higher in terms of the operation.

7. When membrane systems were used for pre-concentrate domestic sewage, it is better to operate with sustainable flux. It is better to operate membrane in lower filtration flux with lower TMP. To increase the permeate flow rate, membrane area can be increased rather than increasing the membrane flux. Higher membrane flux leads to clogging the membrane in a short time which has a negative economic impact.

5.2 Recommendations for Future Studies

Based on the overall experimental results, the following could important for further study in domestic sewage pre-concentration for anaerobic digestion.

1. Domestic sewage from AIT campus was used to use for this research. It was observed that the domestic sewage of AIT is diluted due to higher water consumption. It was found that the range of 65 ± 8 mg/L of BOD and COD of 140 ± 50 mg/L which is compared to lower than the typical domestic sewage characteristics. In that case, it is better to shift the source of domestic sewage which represent the typical concentrations. This study found that the observed systems could concentrate domestic sewage up to some level, even the feed water has lower BOD and COD concentrations. With the higher initial concentrations, the systems can concentrate more than the current values.

2. HRT of the WFMF system can be reduced by increasing the surface area of the membrane. In that case, it can be maintained less than 1 mg/L of DO level inside the system.

3. Two membrane system was studied in this research. Both the membrane systems operated without back flushing. The operation time of the membrane can be increased by introducing the intermitted backflush system. It can control the cake layer accumulation on the membrane surface.

4. Optimization of AnMBR system for concentrated domestic sewage is more important to achieving the higher biogas production. Below are the concern factors that can be considered in future works.

    I. Propeller mixing can be more attractive than the pump recirculation due to the negative effect on bio-flocs while pump circulation. Also, It is important to use sonar level sensors than the typical electrode type. Electrode type level sensors can be easily clogged inside and finally have an issue to control the exact working volume.

    II. It is important to operate the membrane separation, in external mode. Moreover, it is important to use bigger diameter tubular ceramic membrane as small diameter tubes can easily block with the biomass settlement. Moreover, Inline measurement of pH and temperature can be more attractive to observe and control the pH fluctuation in the anaerobic reactor.

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APPENDIX A

System Designs

I. WFMF system design

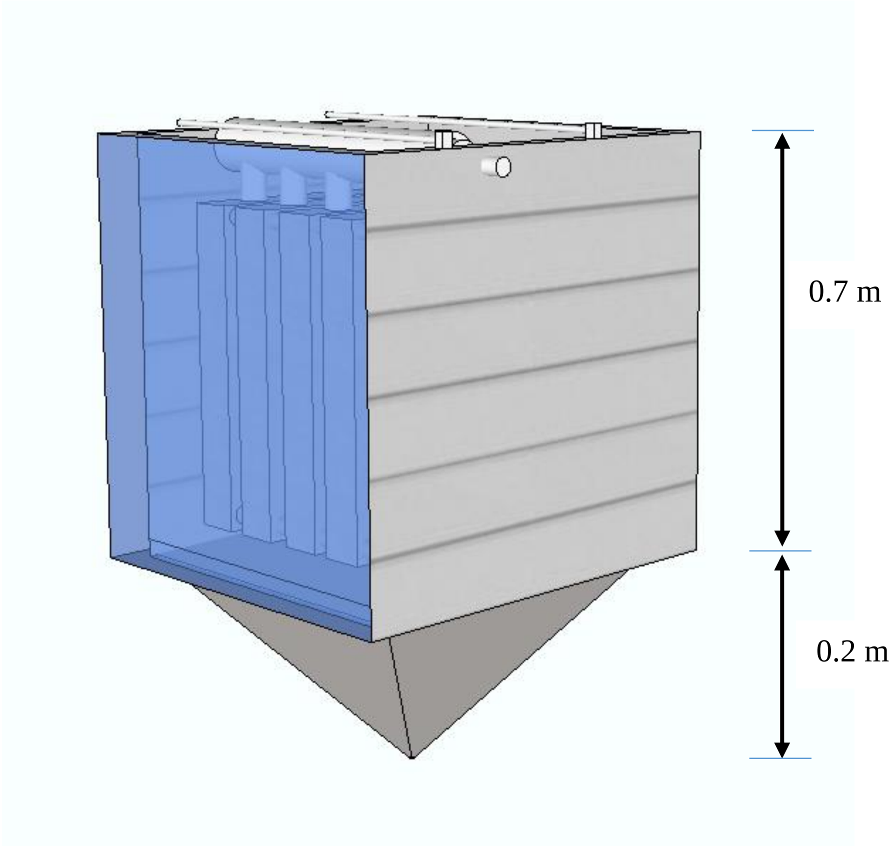


Figure A-1 WFMF membrane tank (side view)

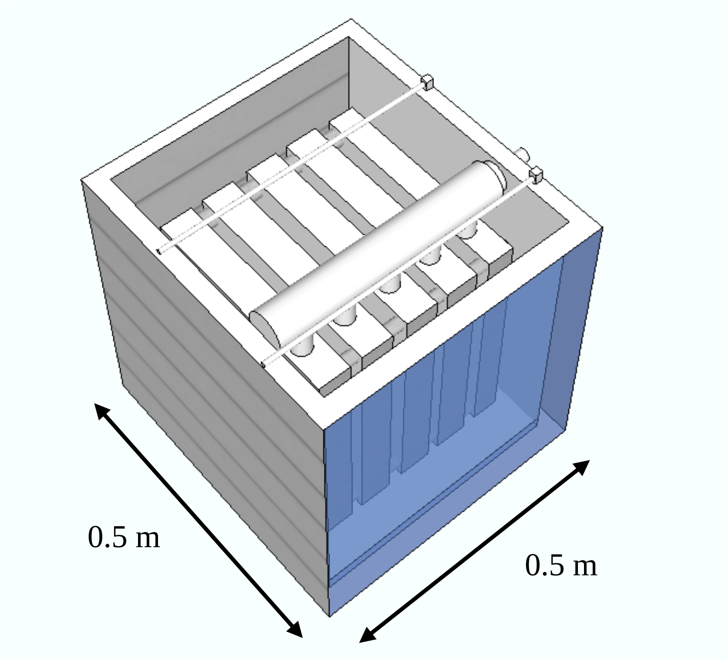


Figure A-2 WFMF membrane tank (top view)

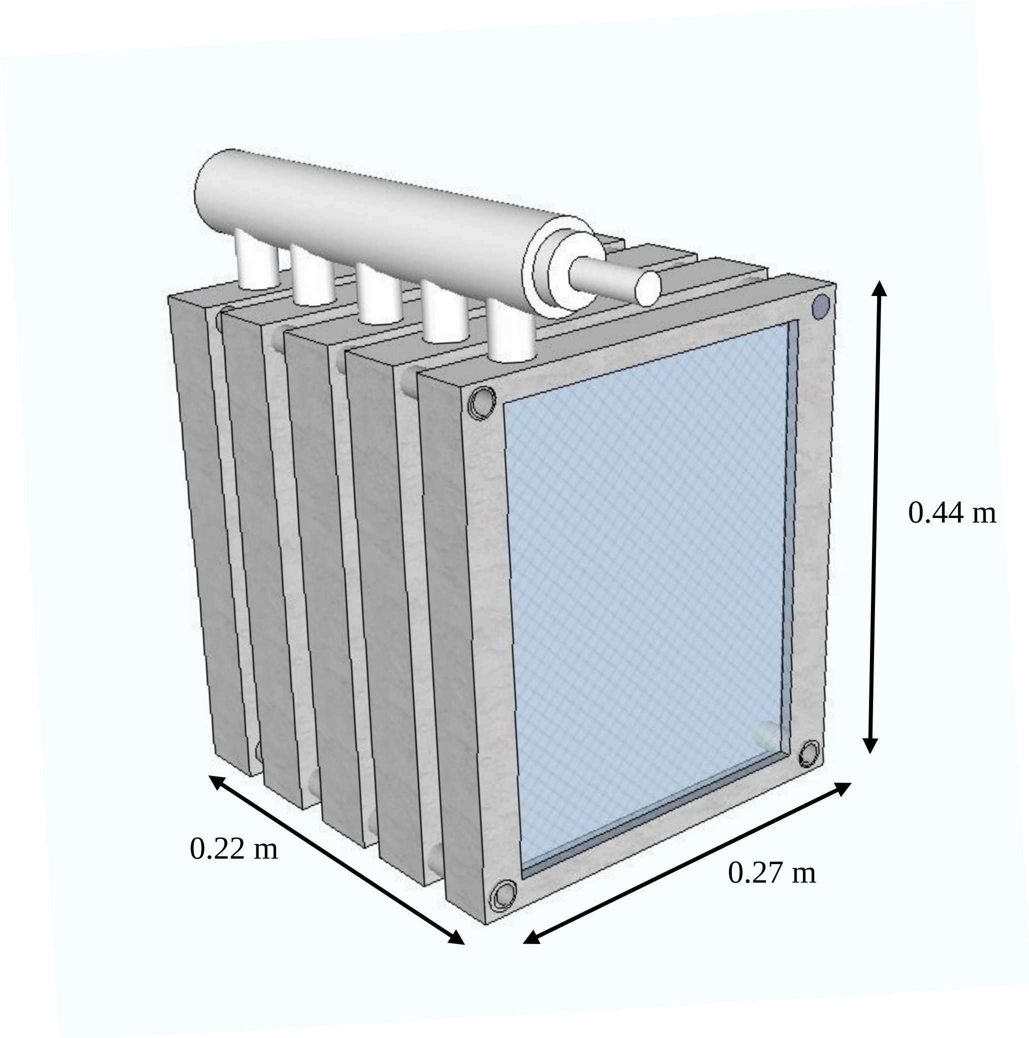


Figure A-3 3D view of the WFMF membrane module

Figure A-4 Laboratory scale WFMF system

II. CMT system design

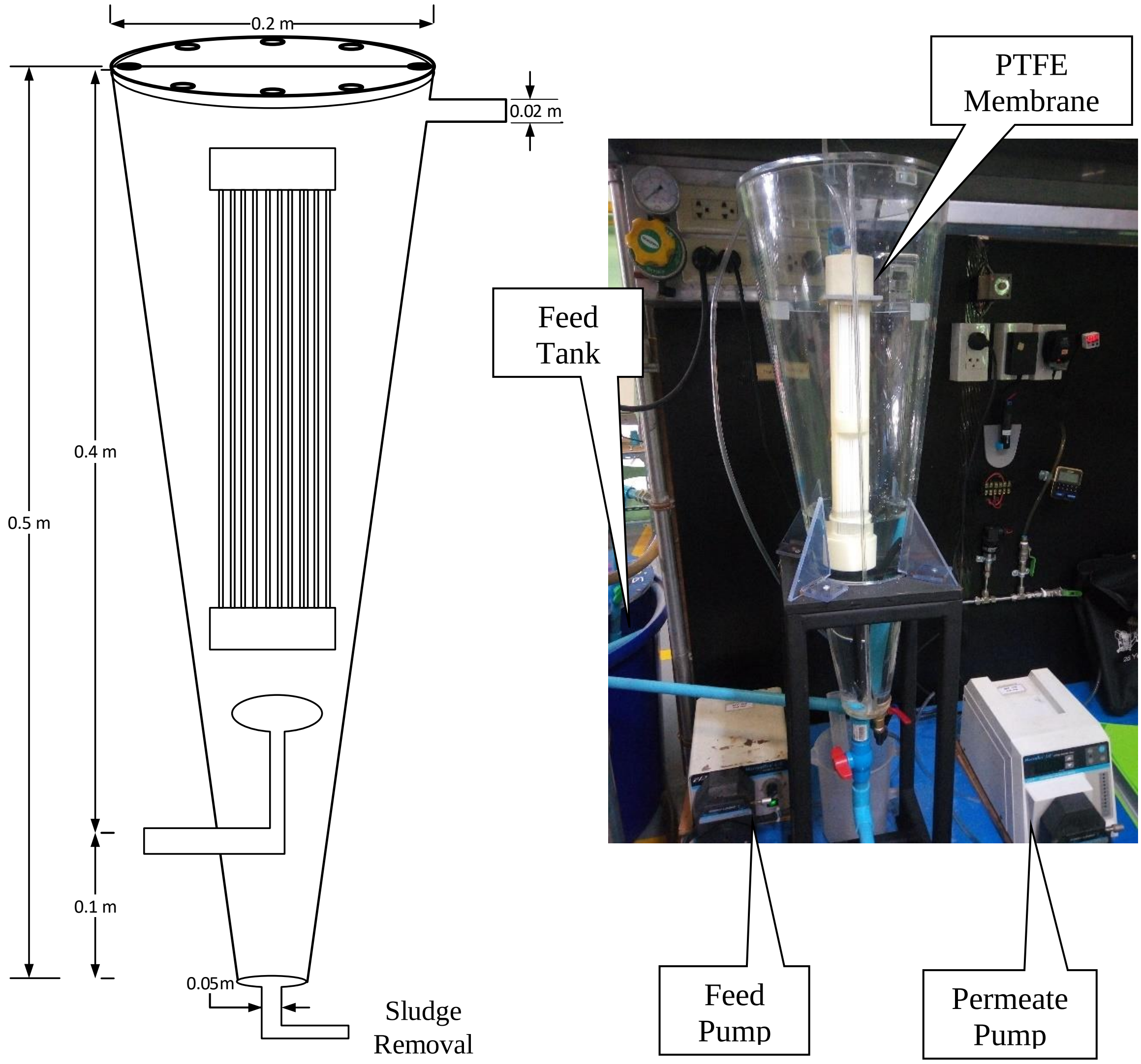


Figure A-5 CMT design configuration

III. Tube settler design

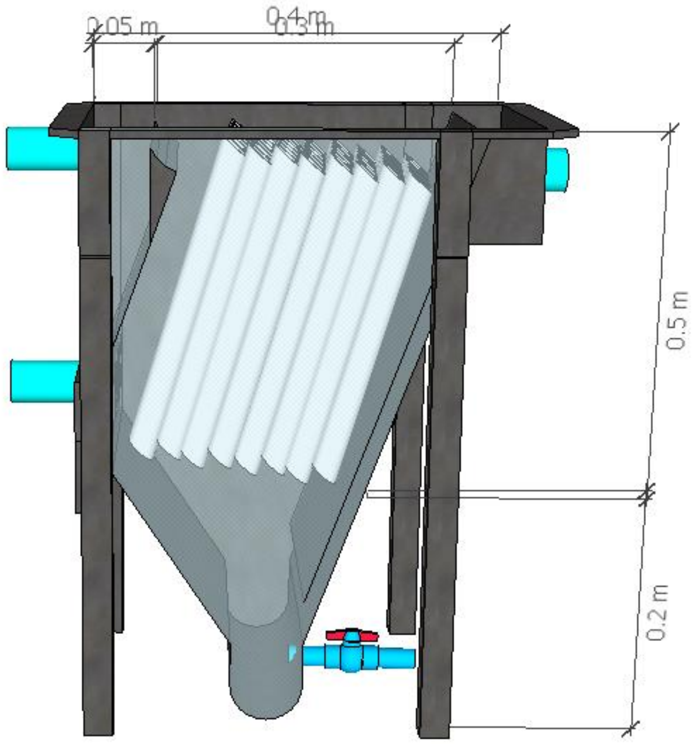


Figure A-6 Laboratory scale tube settler tank (Front view)

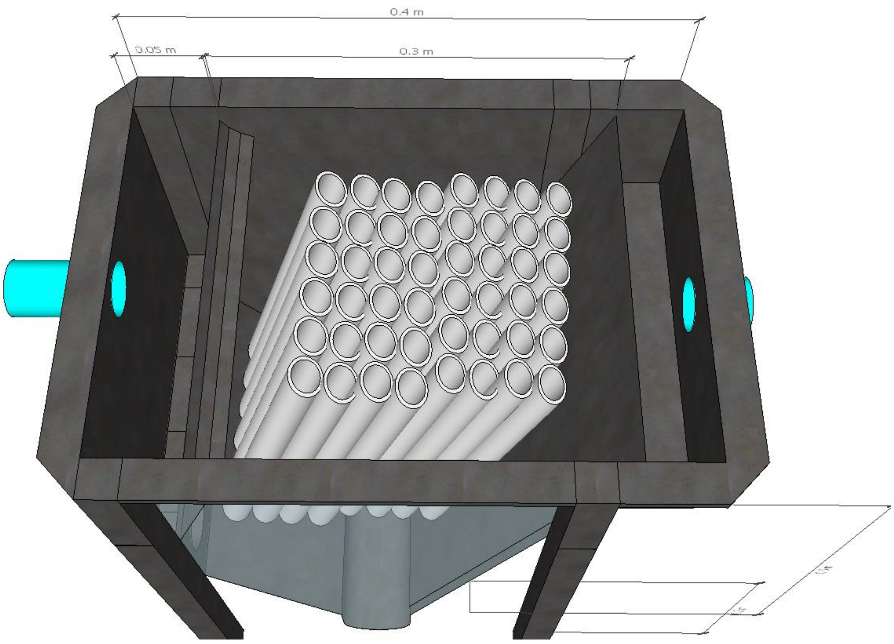


Figure A-7 Laboratory scale tube settler tank (top view)

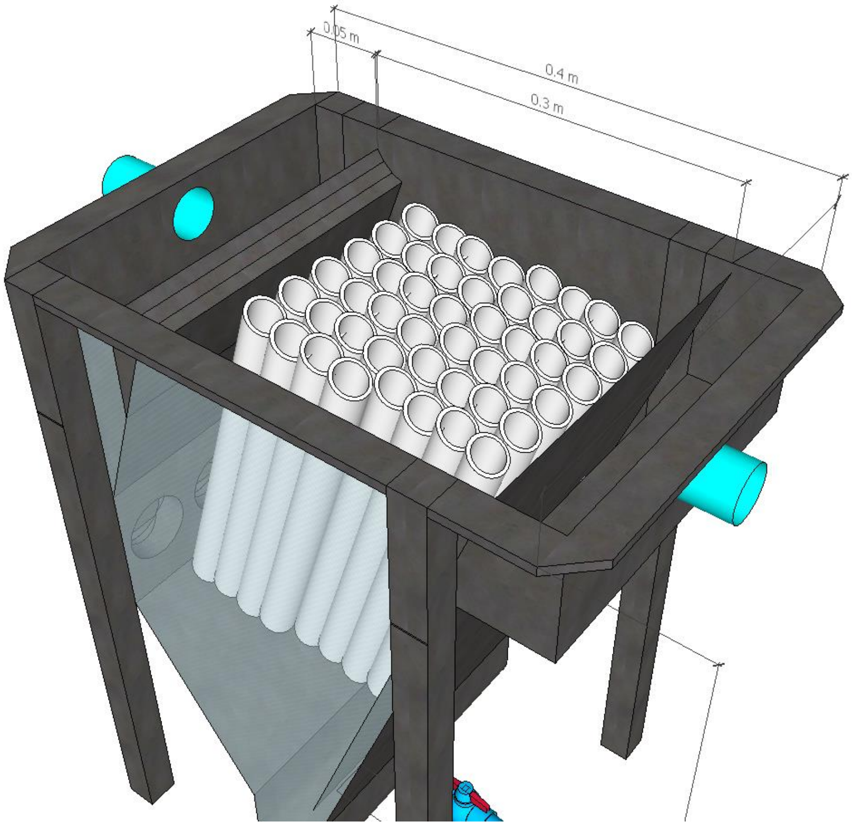


Figure A-8 Laboratory scale tube settler tank (side view)

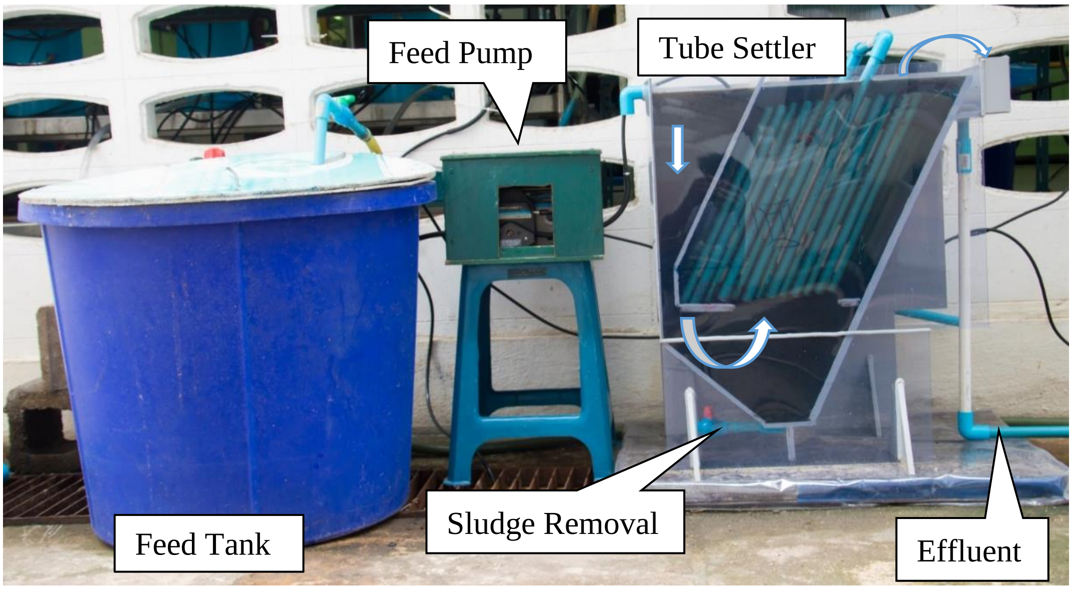


Figure A-9 Laboratory scale WFMF system

IV. Anaerobic digester design

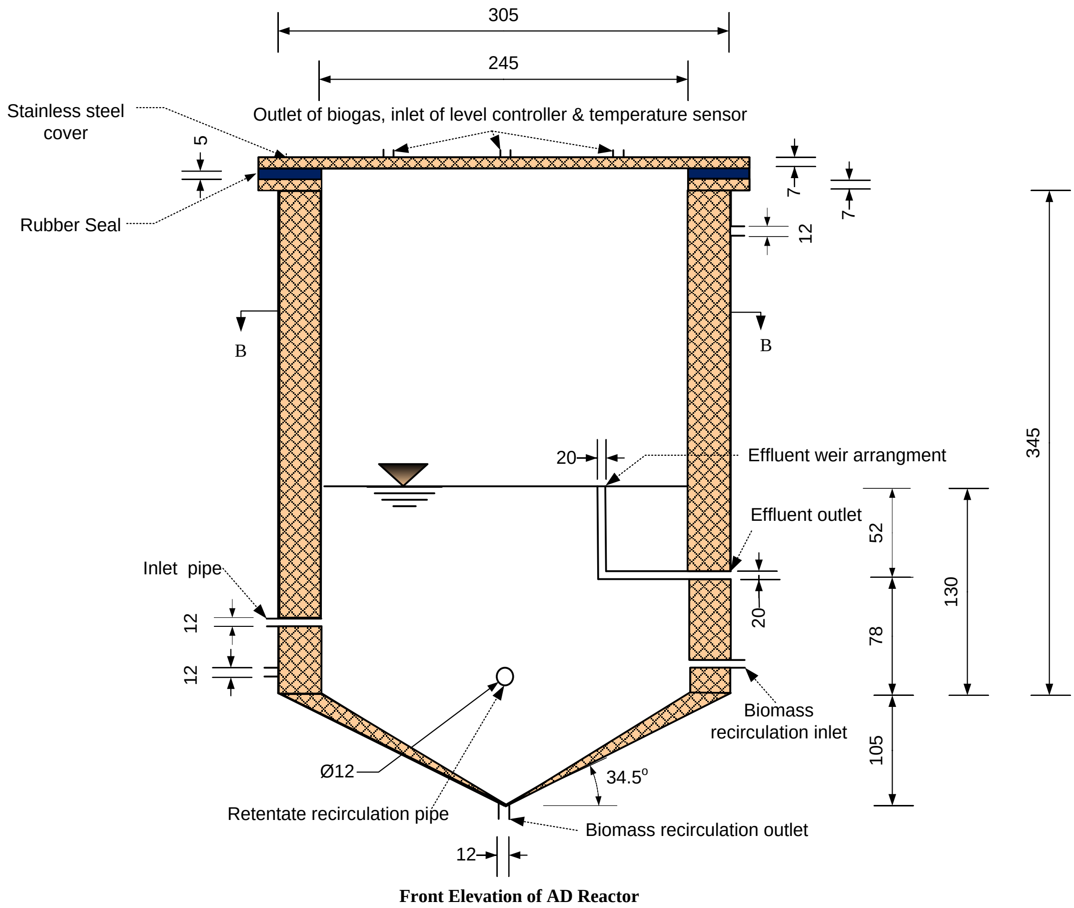
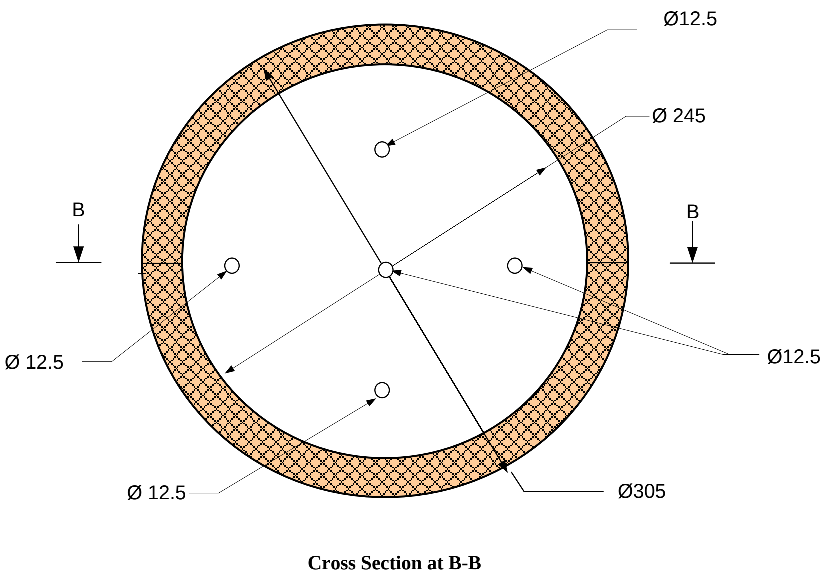
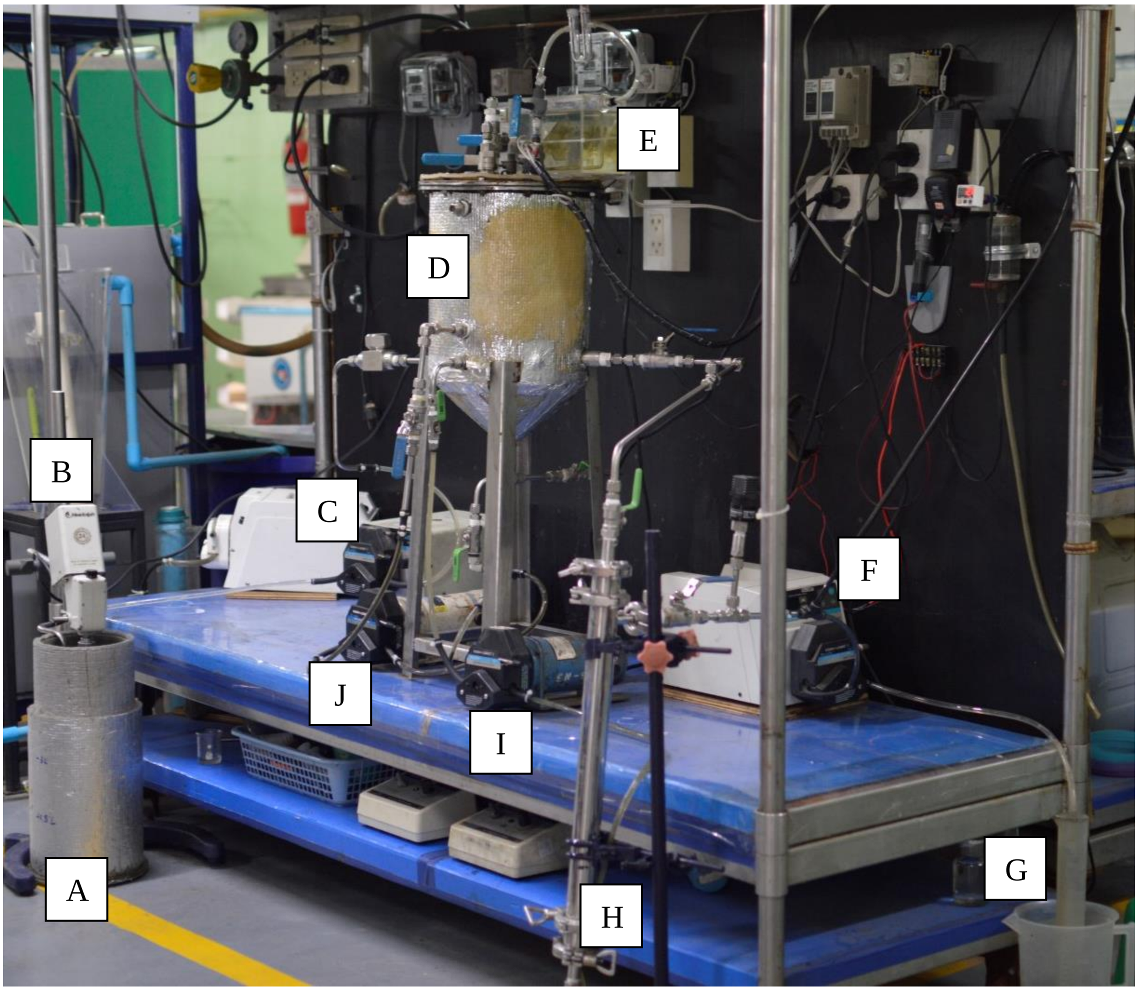


Figure A-10 Laboratory scale AD reactor configuration



*All dimension in millimeters (mm)

Figure A-11 Top view of the AD reactor (B-B cross section)



A- Feed Tank
B- Mixer
C- Feed Pump
D- Anaerobic reactor
E- Gas counter
F- Permeate Pump
G- Effluent
H- Ceramic membrane module
I- Cross-flow Pump
J- Mixing Pump

Figure A-12 Bench scale AnMBR system

APPENDIX B

Design Calculations and Chemical Preparation

I. Chemical preparation for PTFE membrane cleaning

To prepare 2 L of cleaning solution

Step 1

1,500 mg/L NaOCl → 1,500 mg/L NaClO = 0.15% NaOCl

From; $M_1V_1 = M_2V_2$

Stock Solution 12.5 % NaClO

Therefore;
$$12.5\,\%\ V_1 = (0.15\%) \times (2\ \text{L})$$
$$V_1 = 0.06\ \text{L}$$
$$= \mathbf{24}\ \text{mL}$$

2% NaOH → 100 mL Solution Require NaOH 2 g

Therefore; 2 L Solution Require NaOH $\frac{(2\ \text{g}) \times (2\ \text{L})}{0.1\ \text{L}} = \mathbf{40}\ \text{g}$

Step 2

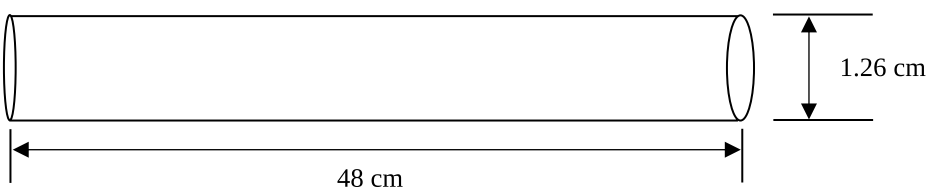
1% H_2SO_4

From; $M_1V_1 = M_2V_2$

Stock Solution 98 % H_2SO_4

Therefore;
$$98\,\%\ V_1 = (1\%) \times (2\ \text{L})$$
$$V_1 = 0.00204\ \text{L}$$
$$= \mathbf{20}\ \text{mL}$$

II. Tube settler- Area calculation



Surface area of a one tube = 2Πrh
= 2* 3.14*0.63*48
= 190 cm^2

No. of pipes = 169
Total area of the tubes = 160*190 cm^2
= 32110 cm^2
= 3.2 m^2

III. Design factors for AnMBR

Total volume of the reactor = 10 L
Working volume = 60% of the total volume
= 10 * 0.6
= **6 L**

Feed water (Pre- concentrated domestic sewage)

Average COD (Generate from WFMF 7.5 LMH) = 7,000 mg/L

Loading Rate = **3.2 Kg COD/m^3.d**

Typically, anaerobic digestion can use OLR of **3.2 to 32 Kg COD/m^3.d** while aerobic treatment uses 0.5 to 3.2 kg COD/m^3.d. Generally, it needs the 1,500-2,000 mg/L of COD concentration to generate enough quantity of CH_4 for increasing up the wastewater temperature except adding of fuel. (Metcalf and Eddy, 2003).

Organic loading rate of **3.2 Kg COD/m^3.d** was selected as it is the minimum start point in the range.

Amount of COD load that input to the reactor, to maintain the OLR of 3.2 Kg COD/m^3.d $= \frac{3.2\ Kg}{m^3\ .d} \times \frac{6\ L}{1} \times \frac{1m^3}{1000L}$

= 0.0192 Kg/d

= 19.2 g/d

Wastewater volume that need to input to the reactor $= \frac{19.2\ g\ COD/d}{7\ g\ COD/L}$

= 2.74 L/d

= 0.11 L/h

Membrane flux need to maintain to achieve 0.11 L/h of flow rate.

Membrane area = 0.18 m^2

Membrane flux = Q/A
= (0.11 L/h) / (0.18 m^2)
= **0.63 LMH**

HRT = V/Q
= (6 L) / (2.74 L/d)
= **2.18 d**

IV. Preparing synthetic wastewater

$$\underbrace{C_6H_{12}O_6}_{180\ g} + \underbrace{6O_2}_{192\ g} \rightarrow 6CO_2 + 6H_2O$$

1 mol of glucose = 192 gCOD
1.07 g/L glucose = 1 gCOD

Designed loading rate = 3.2 kg COD/m^3.d

COD need for 6 L working volume per day = 3.2*6 = 19.2 g COD

Glucose that needed for 6 L working volume per day = (1.07*3.2)* 6
= 20.5 g

Glucose was used as sole carbon source. NH_4HCO_3 and KH_2PO_4 were added as nutrient to maintain COD:N:P ratio at 100:5:1.

Elements	**C**	**N**	**P**
COD:N:P ratio	100	5	1
Chemical	$C_6H_{12}O_6$	NH_4HCO_3	KH_2PO_4
Ratio in terms of grams	3.2	0.16	0.032
Concentration (g/L)	(3.2*1.07)= 3.42	(79/14)*0.16= 0.9	(136/31)*0.032= 0.1

APPENDIX C

WFMF- Experimental Results and Calculations

I. Membrane resistance and cleaning performance of the woven fiber membrane

a) Calculation for the membrane resistance

Pure water flux test results in each cleaning step

Initial pure water flux		After physical cleaning		After base cleaning		After acid cleaning	
Flux (LMH)	TMP (-kPa)	Flux (LMH)	TMP (-kPa)	Flux (LMH)	TMP (-kPa)	Flux (LMH)	TMP (-kPa)
2.9	6.3	3.6	11.3	3.6	6.6	3.8	6.7
19.1	7.2	16.5	20.4	18.6	7.8	18.7	7.7
26.7	8.0	24.0	24.7	25.8	8.7	27.4	9.4
31.2	8.8	24.9	27.8	29.9	9.5	29.9	10.2
34.8	9.5	27.6	29.8	39.6	12.0	41.1	12.0
41.1	11.3	29.9	32.0	45.5	13.7	47.5	13.4
47.3	12.6	30.7	32.5	48.1	15.6	49.2	15.1

Water Viscosity at 30 degree (mPa.s) = 0.0007978
Pure water flux graph's equation → y = 0.1558x + 4.6806

Flux (LMH)	Flux ($m^3/m^2.s$)	Corresponding TMP value (Pa)	Initial resistance (1/m)
10	2.77778E-06	6238.6	2.81511E+12
20	5.55556E-06	7796.6	1.75907E+12
30	8.33333E-06	9354.6	1.40706E+12
40	1.11111E-05	10912.6	1.23105E+12
		Rm (Average)	**1.47E+12**

b) Total resistance after physical cleaning

After physical cleaning, pure water flux graph equation → y = 0.7849x + 7.866

Flux (LMH)	Corresponding TMP value (Pa)	After cleaning- Resistance (1/m)
10	15715	7.09E+12
20	23564	5.32E+12
30	31413	4.72E+12
40	39262	4.43E+12
	Average resistance	**4.58E+12**

c) Total resistance after base cleaning

After base cleaning, graph equation→ y = 0.1978x + 4.5778

Flux (LMH)	Corresponding TMP value (Pa)	After base cleaning- Resistance (1/m)
10	6556	2.96E+12
20	8534	1.93E+12
30	10512	1.58E+12
40	12490	1.41E+12
	Average resistance	**1.64E+12**

d) Total resistance after acid cleaning

After acid cleaning, graph equation→ y = 0.1792x + 5.0603

Flux (LMH)	Corresponding TMP value in Pa	After cleaning- Resistance (1/m)
10	6852	3.09E+12
20	8644	1.95E+12
30	10436	1.57E+12
40	12228	1.38E+12
	Average resistance	**1.63E+12**

II. Pictorial view of the membrane fouling and cleaning performance


Virgin membrane module


After operation- Fouled membrane


After physical cleaning

Base and acid cleaning

III. WFMF- TMP data for the 5 LMH flux (Triplicate test)

Time (h)	TMP (-kPa) 5 LMH		
	Run 1	Run 2	Run 3
0	17.91	11.91	29.91
1	18.41	13.41	28.41
2	24.41	13.91	25.91
3	22.41	14.91	30.91
4	23.91	15.41	32.41
5	23.91	17.41	32.41
6	24.41	17.41	31.91
7	24.41	11.91	32.41
8	22.91	16.41	32.91
9	24.41	17.91	23.41
10	20.41	16.91	30.91
11	15.91	18.41	33.91
12	16.41	17.91	32.41
13	23.41	18.41	34.41
14	25.91	15.41	33.91
15	24.41	14.91	32.91
16	25.91	17.91	33.91
17	24.41	18.41	21.41
18	26.91	18.41	32.91
19	26.91	18.91	43.41
20	30.91	17.91	43.41
21	15.91	18.41	32.91
22	21.91	13.91	41.91
23	31.41	20.91	39.41
24	29.91	21.91	40.41

Time (h)	TMP (-kPa) 5 LMH		
	Run 1	Run 2	Run 3
25	30.91	21.91	41.41
26	30.91	23.91	40.41
27	30.91	23.91	41.41
28	29.41	24.41	33.91
29	29.91	22.91	37.91
30	30.41	14.41	37.91
31	29.41	25.41	40.91
32	30.91	22.91	37.41
33	30.41	22.91	39.91
34	29.91	23.91	39.91
35	30.91	25.91	39.91
36	29.91	23.41	29.41
37	29.41	22.41	37.41
38	29.41	20.91	37.91
39	30.41	22.91	37.91
40	29.41	24.41	37.91
41	24.91	23.91	37.91
42	19.91	23.41	37.41
43	31.91	25.41	37.41
44	31.91	24.41	32.41
45	31.91	17.91	38.91
46	29.91	31.41	56.41
47	29.91	33.91	53.41
48	30.91	30.41	51.91
49	29.91	30.91	49.41

	TMP (-kPa)		
Time (h)	5 LMH		
	Run 1	Run 2	Run 3
50	29.41	34.41	48.41
51	29.91	30.91	45.41
52	29.91	32.91	45.91
53	29.91	16.91	47.91
54	30.41	30.91	47.41
55	29.91	31.41	49.41
56	29.91	30.91	49.91
57	29.91	30.91	48.91
58	27.91	32.41	49.91
59	28.41	30.91	40.41
60	28.41	29.91	45.41
61	25.41	28.41	47.41
62	16.91	32.41	46.41
63	23.41	33.41	47.41
64	40.91	31.91	48.41
65	40.41	30.91	48.41
66	39.91	31.91	48.91
67	43.41	33.41	29.91
68	41.91	21.91	66.91
69	39.41	33.41	67.41
70	41.41	29.91	66.91
71	40.91	32.41	66.91
72	40.41	37.41	65.41
73	41.41	34.41	64.91
74	43.91	36.91	60.91
75	40.41	38.91	58.41
76	40.91	22.91	62.91
77	43.91	36.41	61.91
78	40.41	34.41	60.91
79	40.91	36.91	62.41
80	40.41	33.41	62.91
81	37.91	35.41	60.91
82	32.41	39.41	51.91
83	24.41	32.41	55.41
84	35.91	31.91	58.41
85	44.41	38.41	60.91
86	44.41	38.41	61.41
87	45.41	37.41	60.91
88	51.41	33.91	59.41
89	48.91	37.41	59.41

	TMP (-kPa)		
Time (h)	5 LMH		
	Run 1	Run 2	Run 3
90	49.91	35.91	51.41
91	49.91	24.41	72.41
92	46.41	32.41	70.91
93	48.41	37.91	68.91
94	44.91	36.41	69.41
95	48.91	33.91	67.91
96	47.41	36.91	66.41
97	46.41	36.41	62.91
98	46.91	44.41	60.41
99	43.91	27.41	61.91
100	46.91	40.91	61.91
101	45.41	40.91	63.41
102	38.41	41.41	62.91
103	26.41	44.41	60.91
104	39.41	28.41	61.41
105	47.91	38.91	54.41
106	46.91	43.41	56.91
107	48.91	41.41	59.91
108	48.41	43.41	60.91
109	46.91	43.41	61.91
110	48.91	40.91	60.91
111	46.41	43.91	61.91
112	55.91	21.91	60.91
113	54.41	40.41	51.41
114	55.41	41.41	58.91
115	55.91	42.41	61.41
116	56.91	40.91	74.91
117	54.91	43.41	76.91
118	55.41	40.91	77.41
119	54.41	41.41	73.41
120	52.91	34.91	72.41
121	51.91	43.41	69.41
122	45.41	44.91	70.91
123	33.91	42.91	73.41
124	49.91	43.41	72.41
125	56.41	42.41	66.91
126	55.91	43.91	69.41
127	55.41	28.91	69.91
128	55.91	39.41	62.41
129	55.91	41.91	70.41

	TMP (-kPa)		
Time (h)	5 LMH		
	Run 1	Run 2	Run 3
130	55.41	44.41	68.41
131	55.91	42.41	69.91
132	57.41	43.41	67.41
133	56.91	42.91	69.91
134	55.41	42.41	68.41
135	56.41	25.91	71.41
136	56.41	40.91	59.41
137	55.41	43.91	67.91
138	52.41	44.41	53.91
139	52.91	44.41	33.91
140	48.41	45.41	27.41
141	45.91	44.41	22.41
142	53.41	43.91	72.41
143	38.41	37.91	68.41
144	54.91	47.41	68.91
145	58.41	48.91	69.91
146	53.41	48.91	73.41
147	54.91	49.41	74.41
148	51.41	48.91	75.41
149	53.91	48.91	72.41
150	56.41	30.41	70.91
151	56.41	42.41	61.91
152	55.91	44.91	65.91
153	52.91	48.91	71.91
154	55.41	48.91	69.41
155	54.41	48.91	72.41
156	49.91	48.41	70.91
157	52.91	46.91	70.41
158	54.91	26.41	70.41
159	58.91	44.91	57.91
160	58.41	48.91	67.91

IV. WFMF- TMP data for the 7.5 LMH flux (Triplicate test)

Time (h)	TMP (-kPa) 7.5 LMH		
	Run 1	Run 2	Run 3
0	17.4	19.9	24.4
1	13.4	20.9	40.4
2	21.9	22.9	44.4
3	23.9	25.4	43.9
4	25.9	25.9	43.4
5	25.4	25.4	42.9
6	26.4	26.4	43.4
7	27.9	25.9	43.4
8	27.9	25.9	42.9
9	26.9	26.9	28.4
10	15.9	25.4	38.4
11	26.9	25.4	40.9
12	27.4	26.9	42.4
13	28.4	26.4	42.9
14	27.9	26.9	41.4
15	27.9	26.4	41.9
16	28.9	26.4	40.9
17	28.9	27.4	40.9
18	24.9	27.4	29.9
19	21.9	27.9	38.4
20	38.9	26.9	39.9
21	37.9	26.4	40.4
22	38.4	27.4	58.4
23	38.9	35.4	48.4
24	38.4	38.4	47.9
25	40.9	37.9	46.9
26	41.4	36.9	47.4
27	25.9	38.9	35.4
28	37.9	38.4	46.9
29	37.9	35.9	45.9
30	38.9	35.9	44.9
31	40.4	37.4	44.4
32	39.4	36.9	44.9
33	38.4	38.4	46.9
34	37.4	36.4	44.9
35	39.4	36.9	46.4
36	22.9	38.4	30.9
37	36.9	37.9	45.4
38	40.9	36.9	44.4
39	39.4	37.9	42.9

Time (h)	TMP (-kPa) 7.5 LMH		
	Run 1	Run 2	Run 3
40	44.4	37.4	45.4
41	44.4	37.9	44.4
42	50.9	36.9	42.4
43	50.4	36.9	42.4
44	49.9	38.4	40.4
45	23.4	35.4	35.9
46	49.4	37.9	42.9
47	48.9	47.9	45.4
48	45.4	46.4	44.4
49	44.9	43.9	42.9
50	47.4	47.4	45.4
51	47.4	44.9	44.4
52	41.4	46.9	42.4
53	39.9	48.4	42.4
54	40.4	42.9	40.4
55	44.4	46.9	35.9
56	42.4	41.4	42.9
57	45.9	42.9	45.4
58	39.9	41.4	44.4
59	46.9	39.9	42.9
60	43.9	39.9	45.4
61	40.4	45.4	44.4
62	32.4	44.4	42.4
63	58.9	44.9	42.4
64	57.4	39.4	40.4
65	55.4	42.4	35.9
66	54.4	41.9	42.9
67	53.4	38.9	45.4
68	49.9	42.9	44.4
69	49.9	43.9	42.4
70	49.4	44.4	42.4
71	30.4	53.4	17.9
72	50.4	49.9	43.9
73	46.9	49.9	45.9
74	47.4	50.9	47.9
75	51.4	50.4	47.9
76	52.9	50.4	46.9
77	48.9	50.4	49.4
78	49.9	52.9	47.9
79	51.9	49.4	45.9

Time (h)	TMP (-kPa) 7.5 LMH Run 1	Run 2	Run 3
80	26.4	51.4	31.4
81	45.9	51.4	46.9
82	51.4	50.9	47.9
83	50.4	48.9	49.9
84	50.4	51.9	49.9
85	48.4	51.4	50.4
86	48.4	51.9	53.9
87	47.4	47.9	54.4
88	50.9	49.4	48.4
89	41.4	47.4	43.4
90	51.4	50.4	49.4
91	59.9	49.9	52.4
92	58.9	48.4	52.4
93	59.4	50.9	53.9
94	58.9	52.4	54.4
95	60.9	52.4	55.4
96	60.9	51.9	53.4
97	54.4	52.4	43.4
98	49.4	51.4	43.9
99	59.4	51.4	51.4
100	60.9	51.9	61.9
101	60.9	51.9	69.4
102	62.4	51.4	69.9
103	61.9	51.9	66.9
104	61.4	51.4	66.9
105	61.9	52.9	66.9
106	46.9	52.4	54.4
107	52.9	52.9	46.9
108	58.9	51.9	57.9
109	60.4	51.9	61.4
110	70.9	51.9	62.4
111	70.4	46.9	63.4
112	69.9	47.4	62.9
113	68.4	20.4	62.9
114	67.4	49.4	63.4
115	45.4	48.9	57.9
116	59.9	48.9	45.9
117	64.4	50.9	55.4
118	64.4	46.4	57.9
119	64.4	48.4	63.4
120	64.4	50.9	61.4
121	63.9	49.4	60.4

Time (h)	TMP (-kPa) 7.5 LMH Run 1	Run 2	Run 3
122	62.9	49.4	73.9
123	63.9	49.4	69.4
124	39.9	54.9	68.4
125	58.4	54.9	43.4
126	61.9	54.9	61.9
127	63.4	46.4	65.9
128	63.4	50.9	66.4
129	63.9	55.4	63.9
130	62.9	53.9	65.4
131	63.4	53.9	64.4
132	61.9	36.4	65.4
133	47.9	53.4	66.4
134	63.4	55.4	51.4
135	64.4	55.4	56.4
136	64.4	54.9	61.4
137	62.9	29.9	59.9
138	61.9	53.4	60.9
139	61.4	55.4	60.4
140	60.4	65.9	65.4
141	47.9	68.4	64.4
142	53.9	59.4	64.4
143	57.9	67.9	59.9
144	57.9	63.4	66.9
145	57.9	62.9	68.4
146	57.4	48.9	69.9
147	61.9	60.9	68.9
148	61.4	64.4	70.4
149	59.9	61.9	69.9
150	41.4	62.9	70.4
151	54.9	30.9	50.4
152	59.4	62.4	53.4
153	60.4	63.4	61.9
154	59.9	58.9	63.9
155	58.9	58.9	63.9

V. WFMF- TMP data for the 10 LMH flux

Time (h)	TMP (-kPa) 10 LMH	
	Run 1	Run 2
0	12.4	26.4
1	15.9	30.9
2	16.9	32.9
3	19.4	34.4
4	21.9	34.4
5	22.9	34.4
6	23.9	34.9
7	27.4	35.4
8	27.9	35.4
9	35.9	35.4
10	38.4	35.4
11	37.9	35.9
12	38.4	35.4
13	37.9	35.4
14	38.9	35.4
15	39.9	35.9
16	37.9	35.4
17	38.9	35.4
18	38.9	35.9
19	39.9	35.4
20	38.9	35.4
21	39.9	35.4
22	39.4	34.9
23	39.4	41.4
24	48.4	39.4
25	39.9	40.4
26	48.6	40.4
27	48.4	41.4
28	48.4	40.4
29	40.9	41.4
30	55.4	40.9
31	55.4	40.9
32	55.4	49.9
33	56.4	41.4
34	56.9	41.4
35	57.4	40.9
36	56.4	48.1
37	57.4	42.4
38	57.9	56.9
39	57.4	56.9
40	57.4	56.9
41	58.4	57.9
42	58.9	58.4
43	58.9	58.9
44	58.4	57.9
45	58.4	58.9
46	56.4	59.4
47	56.9	58.9
48	67.9	58.9

VI. COD analysis data for the 5.0 and 7.5 LMH flux for WFMF application

For the 5 LMH flux (Triplicate test results)

	COD (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	186	169	144	166	165 ± 21
WFMF-Permeate	74	19	59	51	47 ± 27
WFMF-Concentrate	4,730	5,922	6,173	5,608	5,452 ± 721

For the 7.5 LMH flux (Triplicate test results)

	COD (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	86	171	141	133	129 ± 48
WFMF-Permeate	14	67	25	35	41 ± 26
WFMF-Concentrate	7,144	7,484	9,067	7,898	8106 ± 961

Note: When analyzing, each test were done in triplicates.

VII. TSS analysis data for the 5.0 and 7.5 LMH flux for WFMF application

For the 5 LMH flux (Triplicate test results)

	TSS (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	71	64	109	81	87 ± 22
WFMF-Permeate	10	12	3	8	8 ± 4
WFMF-Concentrate	3,523	3,363	3,032	3,306	3278 ± 245

For the 7.5 LMH flux (Triplicate test results)

	TSS (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	48	72	80	67	64 ± 16
WFMF-Permeate	2	8	10	7	6 ± 4
WFMF-Concentrate	4,708	5,784	6,029	5,507	5,369 ± 660

Note: When analyzing, each test were done in triplicates.

VIII. Energy consumption for the 5.0 and 7.5 LMH flux for WFMF application

		Total energy consumption (kWh)	**Average energy consumption (kWh)/ Test**	**Total COD in sludge cone (g)**	**kWh/ g COD**
5.0 LMH	Test 1	7.05	6.43	142	0.045
	Test 2	6.23			
	Test 3	6.00			
7.5 LMH	Test 1	6.30	6.53	186	0.035
	Test 2	6.60			
	Test 3	6.68			

APPENDIX D

CMT- Experimental Results and Calculations

I. CMT- TMP data for the 5 LMH flux (Triplicate test)

Time (h)	TMP (-kPa) 5 LMH Run 1	Run 2	Run 3
0	1.5	1.5	1.5
1	1.5	1.5	1.0
2	2.0	1.5	1.5
3	2.0	2.0	1.5
4	1.5	1.5	1.5
5	1.5	1.5	1.5
6	2.0	1.5	2.0
7	1.5	1.5	2.0
8	1.5	1.5	2.0
9	1.5	1.5	2.5
10	1.5	1.5	2.5
11	2.0	1.5	2.5
12	2.0	1.5	3.0
13	2.0	1.5	3.0
14	1.0	1.5	3.0
15	2.0	2.0	3.5
16	2.0	2.0	3.5
17	2.0	2.0	4.0
18	2.0	2.0	4.0
19	1.5	2.0	4.0
20	2.0	2.0	2.5
21	2.5	2.0	2.5
22	2.5	2.0	2.5
23	1.5	2.5	3.0
24	2.0	2.5	4.5
25	2.5	2.5	6.5
26	2.5	2.5	6.5
27	2.5	2.5	7.0
28	1.5	2.5	7.0
29	2.5	2.5	7.5
30	2.5	2.5	8.0
31	3.0	2.5	8.5
32	2.0	2.5	9.0
33	2.0	3.0	9.5
34	3.0	2.5	9.5
35	3.0	2.5	10.0
36	3.0	2.5	10.5
37	1.5	2.0	11.0
38	3.0	2.0	11.5
39	3.0	1.5	12.0

Time (h)	TMP (-kPa) 5 LMH Run 1	Run 2	Run 3
40	3.5	1.5	12.0
41	3.5	2.0	12.5
42	1.5	3.5	12.5
43	4.0	3.5	13.0
44	4.0	3.5	13.5
45	4.0	4.0	13.5
46	2.0	4.0	13.5
47	3.0	4.0	13.5
48	4.0	4.0	13.0
49	4.5	4.0	13.0
50	4.0	4.0	14.0
51	1.5	4.0	16.0
52	4.0	4.0	18.5
53	4.5	4.0	19.5
54	4.5	4.0	20.5
55	4.0	4.0	21.0
56	2.5	4.0	21.5
57	4.5	4.0	22.0
58	4.5	4.0	22.5
59	4.5	4.0	23.0
60	2.0	4.0	23.5
61	4.0	4.0	24.0
62	5.0	3.5	24.5
63	5.5	3.5	25.0
64	5.5	3.5	25.5
65	2.0	3.0	25.5
66	5.5	2.0	26.0
67	6.0	2.0	26.5
68	6.5	3.0	26.5
69	4.5	4.0	26.5
70	4.5	5.0	26.0
71	7.0	5.0	26.0
72	6.5	5.0	25.5
73	6.0	5.5	25.5
74	2.0	5.0	25.5
75	5.0	5.5	25.0
76	6.0	5.0	25.0
77	6.5	5.0	25.0
78	6.5	5.0	26.5
79	2.5	5.0	28.5

Time (h)	TMP (-kPa) 5 LMH Run 1	Run 2	Run 3
80	6.0	5.0	30.0
81	6.5	5.0	30.5
82	7.0	5.5	31.0
83	3.5	5.5	31.5
84	4.5	5.5	32.0
85	6.5	5.5	32.5
86	7.5	5.0	33.0
87	7.5	5.0	33.0
88	2.0	5.0	33.5
89	6.5	5.0	34.0
90	7.5	5.0	34.0
91	8.0	5.0	34.5
92	7.0	4.5	34.5
93	4.0	4.5	34.5
94	7.5	3.5	34.0
95	8.5	4.5	34.0
96	9.0	4.5	33.5
97	4.0	5.5	33.5
98	6.5	6.5	33.5
99	8.5	6.5	33.5
100	9.0	6.0	34.0
101	9.5	6.0	33.5
102	2.5	6.0	34.0
103	7.5	6.0	34.0
104	9.0	6.0	34.5
105	9.5	6.0	35.0
106	6.5	6.0	36.0
107	5.5	5.5	37.0
108	9.0	5.5	38.0
109	10.0	5.0	38.5
110	10.5	5.0	38.5
111	4.0	4.5	39.0
112	8.0	3.5	39.0
113	10.0	4.0	39.5
114	11.0	5.0	39.5
115	11.0	6.0	42.5
116	5.0	7.5	44.5
117	9.5	8.0	45.0
118	11.0	8.0	45.0
119	12.0	8.0	44.5

Time (h)	TMP (-kPa) 5 LMH Run 1	Run 2	Run 3
120	7.5	8.0	44.0
121	8.0	7.5	44.0
122	10.5	7.5	44.0
123	12.0	7.5	44.0
124	12.5	7.5	43.5
125	4.5	7.5	43.5
126	9.5	7.5	44.0
127	11.0	7.0	45.0
128	12.0	7.5	45.0
129	9.5	7.0	45.5
130	7.0	7.5	46.0
131	10.5	7.5	46.0
132	12.0	7.5	46.0
133	12.5	7.0	47.0
134	6.5	7.0	48.0
135	9.0	7.0	48.5
136	11.5	6.5	49.5
137	12.5	6.5	49.5
138	13.5	6.0	50.0
139	6.0	5.5	50.0
140	11.0	4.5	50.0
141	12.5	5.5	50.0
142	13.5	7.0	49.5
143	10.0	8.0	49.0
144	8.5	9.5	49.0
145	12.0	9.0	48.5
146	13.5	9.0	48.5
147	14.0	9.0	48.0
148	8.0	9.0	48.0
149	10.0	9.0	48.0
150	13.0	9.0	47.5
151	13.5	9.0	48.0
152	14.0	9.0	48.0
153	6.5	9.0	48.0
154	11.5	9.0	48.0
155	13.5	9.0	48.5
156	14.0	9.0	48.5
157	10.0	9.0	48.5
158	9.5	9.0	48.5
159	12.5	9.0	48.5

II. CMT- TMP data for the 7.5 LMH flux (Triplicate test)

Time (h)	TMP (-kPa) 7.5 LMH Run 1	Run 2	Run 3
0	2.5	2.00	1.0
1	2.5	2.50	1.5
2	3.0	2.50	1.5
3	2.5	2.50	1.5
4	3.0	2.50	1.5
5	3.0	3.00	2.0
6	3.0	3.00	2.0
7	3.0	3.00	2.0
8	3.0	3.00	2.0
9	3.0	3.00	2.0
10	3.0	3.50	2.0
11	3.0	3.50	2.0
12	3.0	4.00	1.5
13	3.0	4.00	1.0
14	3.0	3.50	1.5
15	3.0	4.50	2.5
16	3.0	5.00	2.5
17	3.0	5.00	2.5
18	3.5	5.50	3.0
19	3.5	6.00	3.0
20	4.0	6.00	3.0
21	4.0	6.50	3.5
22	4.0	7.00	3.5
23	4.5	7.50	3.5
24	4.5	8.00	3.5
25	4.5	8.50	3.5
26	4.5	8.50	3.0
27	4.5	9.00	2.0
28	5.0	9.00	3.0
29	5.0	8.50	6.0
30	4.5	5.50	6.0
31	5.5	10.50	2.5
32	5.0	11.00	2.5
33	5.5	11.00	2.0
34	5.0	11.50	2.0
35	5.0	12.50	2.0
36	6.5	13.00	2.0
37	6.5	13.50	2.5
38	6.0	14.00	3.5
39	7.0	14.50	4.0

Time (h)	TMP (-kPa) 7.5 LMH Run 1	Run 2	Run 3
40	6.5	15.00	5.0
41	6.0	15.50	5.5
42	7.5	16.00	6.5
43	7.5	16.00	7.5
44	8.0	16.00	8.0
45	8.5	15.00	9.0
46	9.5	12.50	10.0
47	10.5	20.50	11.0
48	11.5	21.00	12.0
49	11.5	21.50	13.0
50	13.0	21.50	14.0
51	13.0	22.00	14.5
52	13.0	22.00	15.0
53	13.5	22.00	15.5
54	13.0	22.00	16.0
55	15.0	22.50	16.5
56	15.0	23.00	17.0
57	15.5	23.00	18.0
58	16.0	23.00	18.5
59	15.5	23.50	19.5
60	15.5	23.50	20.5
61	16.0	23.00	21.5
62	16.0	22.00	23.0
63	17.0	29.00	24.0
64	20.5	30.00	25.0
65	20.5	30.50	26.0
66	21.0	31.00	27.0
67	22.0	32.00	28.0
68	23.0	32.50	29.0
69	29.5	33.00	30.5
70	29.5	33.50	31.5
71	29.5	33.00	32.5
72	29.5	31.00	33.5
73	29.5	28.50	34.0
74	29.5	35.00	35.0
75	33.5	34.50	35.5
76	33.0	33.50	35.5
77	33.0	31.50	35.5
78	33.0	32.50	36.0
79	32.5	35.50	36.0

Time (h)	TMP (-kPa) 7.5 LMH		
	Run 1	Run 2	Run 3
80	32.5	34.50	36.5
81	32.5	33.50	37.0
82	32.5	31.50	38.0
83	32.5	36.50	38.5
84	32.0	37.50	39.5
85	32.0	37.50	40.0
86	31.5	36.50	40.5
87	31.5	35.00	41.5
88	31.0	40.50	42.0
89	32.5	41.00	43.0
90	34.0	40.50	43.5
91	35.5	39.50	44.5
92	37.5	37.50	45.0
93	40.0	43.50	45.5
94	43.0	43.50	46.5
95	43.5	43.00	47.0
96	43.5	42.00	48.0
97	44.0	40.00	48.5
98	45.0	45.00	48.0
99	43.5	44.50	48.0
100	44.0	43.50	48.5
101	47.0	41.50	48.5
102	43.5	39.50	48.5
103	43.5	44.50	48.5
104	44.5	44.00	48.5
105	47.0	43.50	48.5
106	49.0	42.50	48.5
107	49.0	40.50	49.0
108	49.5	46.50	49.5
109	49.5	46.50	50.0
110	49.5	45.50	50.0
111	50.0	44.50	50.0
112	50.0	42.50	50.5
113	53.5	48.50	50.5
114	53.5	48.50	51.0
115	54.0	48.00	51.0
116	54.0	46.50	51.5
117	54.0	47.50	51.5
118	54.0	50.50	52.0
119	54.4	50.50	52.5

Time (h)	TMP (-kPa) 7.5 LMH		
	Run 1	Run 2	Run 3
120	56.5	49.50	53.0
121	58.5	48.00	53.0
122	60.5	50.00	53.5
123	61.0	58.50	53.5
124	60.0	58.00	53.5
125	60.0	56.50	53.5
126	60.0	55.00	53.5
127	60.0	58.00	53.5
128	60.0	58.00	53.0
129	61.5	57.50	53.0
130	60.5	57.00	53.0
131	61.0	56.00	53.5
132	61.0	60.00	54.0
133	61.5	60.00	54.0
134	61.5	60.00	54.0
135	62.0	59.50	54.5
136	62.0	58.00	55.0
137	62.0	62.50	55.0
138	62.0	62.00	55.0
139	63.0	62.00	55.5
140	63.5	61.50	56.0
141	63.5	60.50	56.0
142	64.0	64.50	56.5
143	64.5	64.50	57.0
144	64.5	69.50	59.5
145	64.0	69.50	61.0
146	64.0	68.00	61.0
147	64.0	70.50	61.0
148	63.0	70.50	60.5
149	62.5	69.50	60.0
150	62.5	68.00	60.0
151	63.0	66.50	60.0
152	64.0	69.50	59.5
153	65.5	69.50	59.5
154	67.0	69.00	59.5
155	68.5	68.00	60.0
156	70.0	68.00	61.0
157	70.5	70.50	61.5
158	71.0	70.50	62.5
159	71.5	70.00	62.5

III. CMT- TMP data for the 10 LMH flux

Time (h)	TMP (-kPa) 10 LMH Run 1	TMP (-kPa) 10 LMH Run 2
0	1.0	3.0
1	1.0	3.0
2	1.0	2.0
3	1.5	3.5
4	1.5	2.5
5	1.5	2.5
6	1.5	4.0
7	2.0	3.0
8	2.3	2.0
9	2.0	3.0
10	2.5	3.5
11	2.5	3.5
12	3.0	4.0
13	3.5	4.5
14	3.5	4.5
15	3.5	4.5
16	3.0	4.0
17	2.5	3.5
18	2.5	3.5
19	2.0	5.0
20	2.0	4.0
21	2.0	4.5
22	2.0	4.0
23	2.0	3.0
24	2.5	3.5
25	1.5	4.0
26	3.5	4.5
27	4.5	5.5
28	5.5	6.5
39	13.5	17.0

Time (h)	TMP (-kPa) 10 LMH Run 1	TMP (-kPa) 10 LMH Run 2
40	15.0	16.0
41	14.0	16.5
42	13.0	18.0
43	13.0	18.0
44	15.0	21.0
45	18.0	24.0
46	20.0	24.0
47	29.0	24.5
48	40.0	26.0
49	42.0	28.0
50	37.0	30.0
51	37.9	31.0
52	39.0	31.0
53	41.0	30.0
54	45.0	33.0
55	49.0	34.5
56	42.0	37.0
57	53.0	38.5
58	53.5	39.0
59	55.0	41.5
60	58.0	41.5
61	53.5	44.0
62	55.0	44.0
63	63.0	40.0
64	62.5	38.5
65	64.0	44.5
66	60.0	46.0
67	66.5	50.0
68	67.5	61.0

IV. Membrane resistance and cleaning performance of the PTFE-HF membrane

a) Calculation for the membrane resistance

Pure water flux test results in each cleaning step

Initial pure water flux		**After physical cleaning**		**After base cleaning**		**After acid cleaning**	
Flux (LMH)	TMP (-kPa)	Flux (LMH)	TMP (-kPa)	Flux (LMH)	TMP (-kPa)	Flux (LMH)	TMP (-kPa)
5.4	1.4	5.4	2.1	5.4	1.4	5.4	1.3
21	1.5	21	2.4	21	1.6	21	1.5
27	1.7	27	2.6	27	1.7	27	1.7
33	1.8	33	2.7	33	1.8	33	1.8
39	2	39	2.8	39	2.1	39	2
45	2.2	45	3.2	45	2.3	45	2.3

Water Viscosity at 30 degree (mPa.s) = 0.0007978
Pure water flux graph's equation → y = 0.0204x + 1.1866

Flux (LMH)	**Flux ($m^3/m^2.s$)**	**Corresponding TMP value (Pa)**	**Initial resistance (1/m)**
10	2.77778E-06	1391	6.27E+11
20	5.55556E-06	1595	3.60E+11
30	8.33333E-06	1799	2.71E+11
40	1.11111E-05	2003	2.26E+11
50	1.38889E-05	2207	1.99E+11
60	1.66667E-05	2411	1.81E+11
		Rm (Average)	**2.47E+11**

b) Total resistance after physical cleaning

After physical cleaning, pure water flux graph equation → y = 0.0257x + 1.9048

Flux (LMH)	**Corresponding TMP value (Pa)**	**After cleaning- Resistance (1/m)**
10	2,162	9.75E+11
20	2,419	5.46E+11
30	2,676	4.02E+11
40	2,933	3.31E+11
50	3,190	2.88E+11
60	3,447	2.59E+11
	Average resistance	**2.74E+11**

c) Total resistance after base cleaning

After base cleaning, graph equation→ y = 0.0224x + 1.18

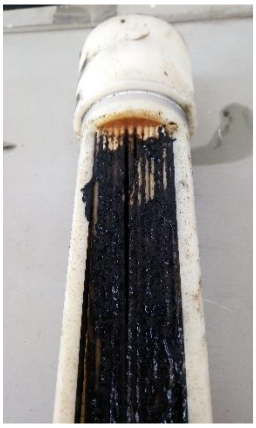
Flux (LMH)	Corresponding TMP value (Pa)	After base cleaning- Resistance (1/m)
10	1,404	6.34E+11
20	1,628	3.67E+11
30	1,852	2.79E+11
40	2,076	2.34E+11
50	2,300	2.08E+11
60	2,524	1.90E+11
	Average resistance	**2.55E+11**

d) Total resistance after acid cleaning

After acid cleaning, graph equation→ y = 0.0244x + 1.0735

Flux (LMH)	Corresponding TMP value in Pa	After cleaning- Resistance (1/m)
10	1,318	5.95E+11
20	1,562	3.52E+11
30	1,806	2.72E+11
40	2,050	2.31E+11
50	2,294	2.07E+11
60	2,538	1.91E+11
	Average resistance	**2.51E+11**

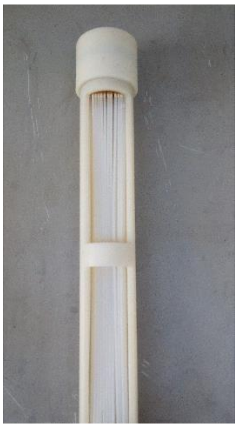
V. Pictorial view of the membrane fouling and cleaning performance



Fouled membrane

After physical cleaning

After base cleaning



After acid cleaning

VI. COD analysis data for the 5.0 and 7.5 LMH flux for CMT application

For the 5 LMH flux (Triplicate test results)

	COD (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	186	169	144	166	165 ± 21
CMT-Permeate	71	26	44	47	48 ± 23
CMT-Concentrate		13,090	14,816	13,953	13,953 ± 863

For the 7.5 LMH flux (Triplicate test results)

	COD (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	86	171	141	133	129 ± 48
CMT-Permeate	28	32	27	29	30 ± 2
CMT-Concentrate	15,435	17,400	20,800	17,878	18,118 ± 2,682

Note: When analyzing, each test were done in triplicates.

VII. TSS analysis data for the 5.0 and 7.5 LMH flux for WFMF application

For the 5 LMH flux (Triplicate test results)

	TSS (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	71	64	109	81	87 ± 22
CMT-Permeate		7	3	5	5 ± 2
CMT-Concentrate		8,717	11,724	10,221	10,221 ± 1503

For the 7.5 LMH flux (Triplicate test results)

	TSS (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	48	72	80	67	64 ± 16
CMT-Permeate	3	2	10	5	6 ± 4
CMT-Concentrate	11,412	10,689	14,094	12,065	12,391 ± 1703

Note: When analyzing, each test were done in triplicates.

VIII. Energy consumption for the 5.0 and 7.5 LMH flux for CMT application

		Total energy consumption (kWh)	Average energy consumption (kWh)/ Test	Total COD in sludge cone (g)	kWh/ g COD
5.0 LMH	Test 1	5.61	6.29	13	0.484
	Test 2	6.67			
	Test 3	6.60			
7.5 LMH	Test 1	6.93	8.13	16	0.508
	Test 2	7.48			
	Test 3	9.98			

APPENDIX E

Tube Settler- Experimental Data and Observations

I. COD analysis data for the 0.005 $m^3/m^2.h$ and 0.01 $m^3/m^2.h$ loading rates

For the 0.005 $m^3/m^2.h$ loading rate (Triplicate test results)

	COD (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	186	169	144	166	165 ± 21
TSET- Effluent	41	40	43	41	42 ± 1
TSET- Concentrate	1,122	2,222	2,143	1,829	1,672 ± 550

For the 0.01 $m^3/m^2.h$ loading rate

	COD (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	86	171	141	133	129 ± 48
TSET- Effluent	119	Remaining tests were did not continued due to unsustainable operation.			119 ± 00
TSET- Concentrate	5,333				5,333 ± 00

II. TSS analysis data for the 0.005 $m^3/m^2.h$ and 0.01 $m^3/m^2.h$ loading rates

For the 0.005 $m^3/m^2.h$ loading rate (Triplicate test results)

	TSS (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	71	64	109	81	87 ± 22
TSET- Effluent	27	37	33	32	32 ± 5
TSET- Concentrate	813	480	491	595	647 ± 166

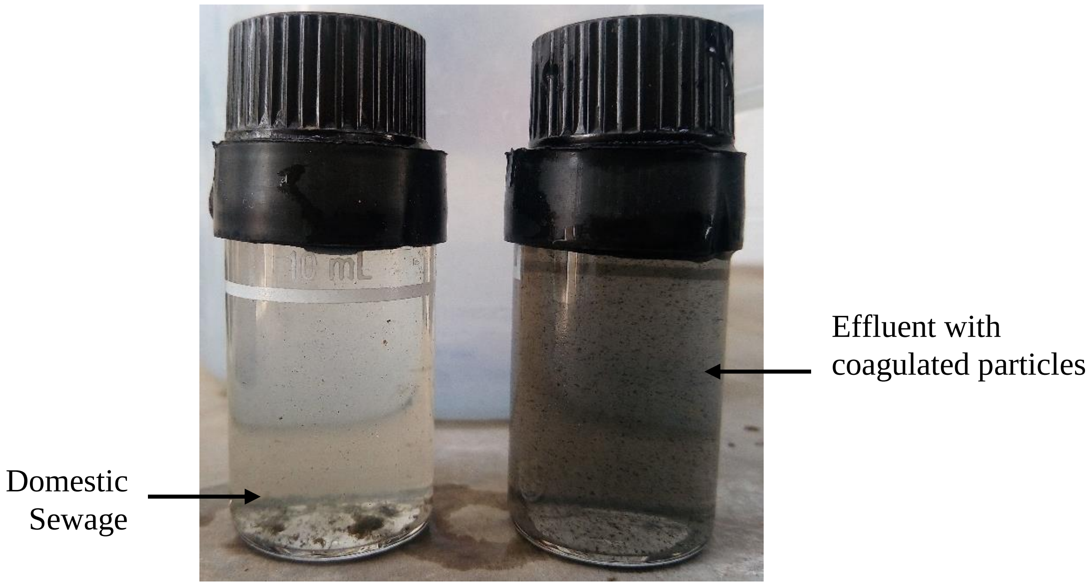
For the 0.01 $m^3/m^2.h$ loading rate

	TSS (mg/L)				
Sample	**Test-1**	**Test-2**	**Test-3**	**Average**	**Range**
Domestic Sewage	48	72	80	67	64 ± 16
TSET- Effluent	57	Remaining tests were did not continued due to unsustainable operation.			57 ± 00
TSET- Concentrate	3,410				3,410 ± 00

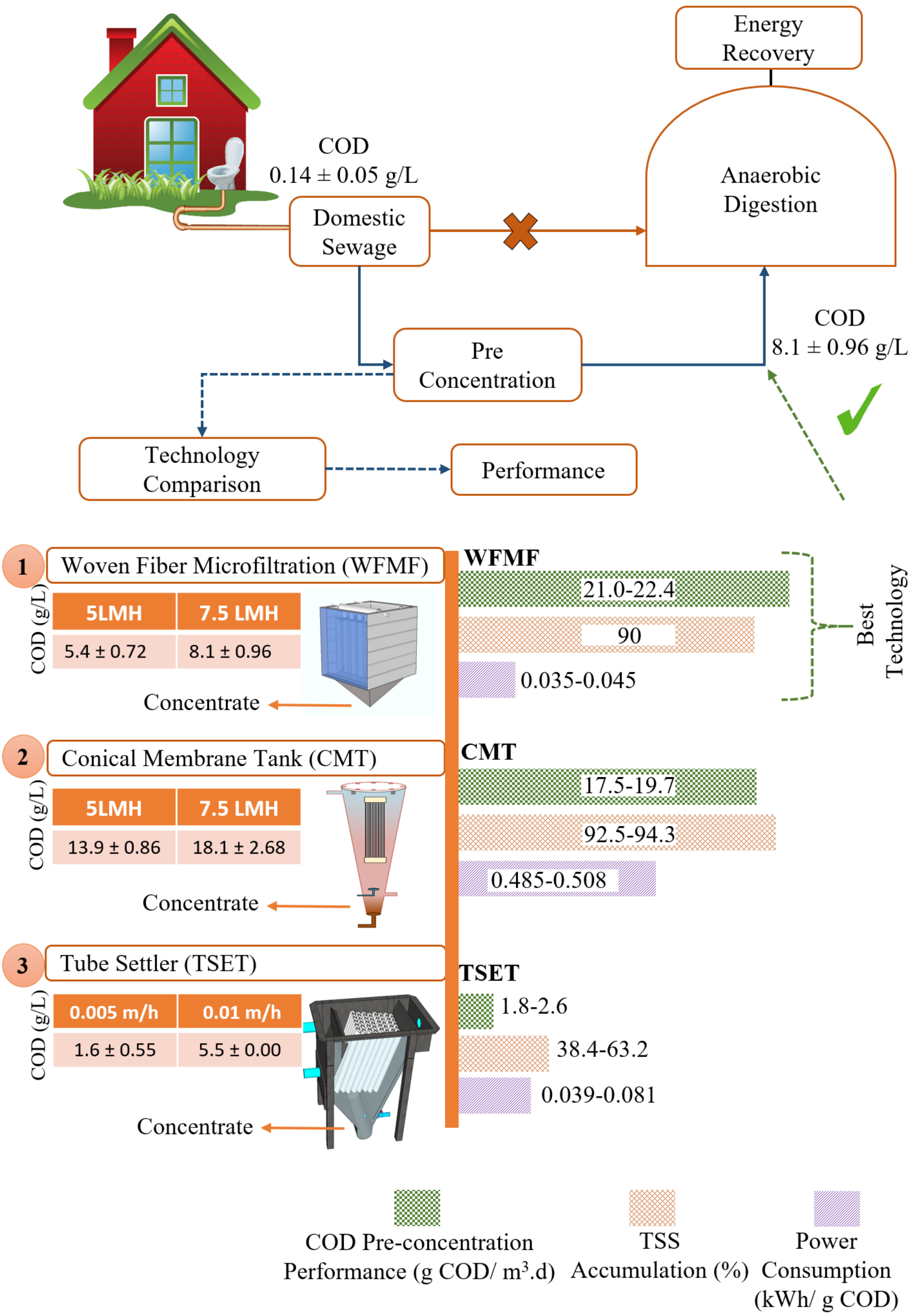
III. Energy consumption for the 0.005 $m^3/m^2.h$ and 0.01 $m^3/m^2.h$ loading rates

		Total energy consumption (kWh)	Average energy consumption (kWh)/ Test	Total COD in sludge cone (g)	kWh/ g COD
0.005 $m^3/m^2.h$ - Without Coagulation	Test 1	2.08	2.66	33	0.081
	Test 2	2.60			
	Test 3	3.31			
0.01 $m^3/m^2.h$ - With Coagulation	Test 1	3.78	3.78	96	0.039
	Test 2				
	Test 3				

IV. Pictorial view of the water sample for 0.01 $m^3/m^2.h$ loading rate (Issue related to solid particle washout)



Graphical Abstract



APPENDIX F

AnMBR- Experimental Data and Observations

I. Biogas production and the pH data for the AnMBR

Day	Biogas Production (mL)	pH
1	2,349	7.09
2	203	5.96
3	493	5.96
4	1,291	5.81
5	1,044	5.34
6	450	5.31
7	1,059	5.98
8	1,479	5.84
9	1,204	5.77
10	986	5.20
11	740	5.57
12	667	5.28
13	406	5.60
14	551	5.93
15	1,552	5.60
16	943	5.91
17	1,334	5.27
18	1,320	5.45
19	856	6.21
20	841	6.33
21	870	6.24
22	609	5.91
23	1,204	5.44
24	943	6.11
25	1,233	5.97
26	1,059	6.07
27	1,117	6.15
28	1,189	6.09
29	812	6.05
30	479	5.71
31	798	6.14
32	638	6.07
33	609	6.13
34	711	5.93
35	464	6.26
36	580	5.96
37	464	6.14
38	435	6.24
39	290	5.94
40	508	6.12

Day	Biogas Production (mL)	pH
41	377	6.07
42	493	5.94
43	522	5.84
44	479	5.81
45	421	6.03
46	493	6.05
47	421	5.83
48	493	6.14
49	522	5.93
50	464	6.07
51	377	5.87
52	406	5.95
53	363	5.91
54	479	6.03
55	450	5.90
56	377	6.03
57	421	5.98
58	348	5.91
59	421	6.22
60	392	5.97
61	334	6.13

II. Gas content of the biogas over the period

	Test 1	Test 2	Test 3	Test 4	Test 5	Test 6
% CH_4	12	28	31	37	37	38
% CO_2	26	42	48	51	51	51
% N_2	62	29	20	12	12	10

III. Removal rate of the AnMBR

	Test 1		Test 2		Test 3	
	CDS	Permeate	CDS	Permeate	CDS	Permeate
COD (mg/L)	6,120	2,080	6,660	1,633	6,060	1,813
BOD (mg/L)	3,500	1,220	4,200	1,341	4,250	1,383
TSS (mg/L)	4,332	11.67	5,100	10	4,791	13.33
COD Removal (%)	66.01		75.48		70.08	
BOD Removal (%)	65.14		68.07		67.46	
TSS Removal (%)	99.73		99.80		99.72	

CDS: Concentrated Domestic Sewage

IV. MLSS and MLVSS

	MLSS (mg/L)	MLVSS (mg/L)
Test 1	10,800	6,560
Test 2	10,500	6,780
Test 3	12,040	7,580
Average	11,113	6,973

	MLSS (mg/L)	MLVSS (mg/L)
Test 1	10,140	6,460
Test 2	10,120	5,560
Test 3	11,580	4,900
Average	10,613	5,640

Range	10,863 ± 250	6,307 ± 666