UNIVERSITY OF CAPE TOWN
IYUNIVESITHI YASEKAPA • UNIVERSITEIT VAN KAAPSTAD

DEPARTMENT OF CIVIL ENGINEERING

Water Quality Engineering

MEng Thesis

Treating wastewater in a conventional activated sludge (CAS) system or a Membrane Bioreactor (MBR). A comparison of capital and operating costs.

By Delwin Smith

Declaration

I know the meaning of plagiarism and declare that all the work in the document, save for that which is properly acknowledged, is my own. This thesis/dissertation has been submitted to the Turnitin module (or equivalent similarity and originality checking software) and I confirm that my supervisor has seen my report and any concerns revealed by such have been resolved with my supervisor.

Signed by candidate

Signature

25 September 2019

Date

Abstract

More and more focus is going into the establishment of more sustainable approaches for wastewater treatment (WWT) in South Africa, as well as around the world. Governments are beginning to enforce more economical solutions for WWT, which will have less impact on costs as well as land area requirements.

Effective solid-liquid separation in biological wastewater treatment is an important step in the process as it has a major impact on effluent quality. Traditionally this has been achieved using Secondary settling tanks (SSTs) for liquid/solid separation in combination with a biological reactor (for biological degradation of organic matter). SSTs, however, require a large space, which becomes onerous on land requirements.

In an immersed membrane bioreactor (iMBR), solid-liquid separation takes place by the wastewater passing through membranes. As the WW flows through, at the same time solids are rejected by the membranes. These membranes are immersed in the bioreactor. iMBR thus eliminates the requirement for SSTs and are becoming more widely used to treat various types of wastewater, due to the decreasing cost of membranes and the resultant reduced plant footprint. MBR is thus becoming an attractive solution to clients due to its sustainable approach. As part of this investigation, 2 types of MBR technology were included, the Kubota FS MBR system and the Zeeweed HF MBR system. As the design of a CAS is sensitive to sludge settleability, various DSVI values were looked at as part of the CAS system. Each system was configured in an MLE and UCT process. In summary, the following systems were included in this investigation:

- CAS in an MLE configuration with DSVI of 100,150 and 200
- CAS in a UCT configuration with DSVI of 100,150 and 200
- iMBR using FS membranes in an MLE configuration
- iMBR using FS membranes in a UCT configuration
- iMBR using HF membranes in an MLE configuration
- iMBR using HF membranes in a UCT configuration

Each process configuration was designed and sized using the steady state models. Each configuration was then fully costed using actual construction prices from past and current projects. Costing of the MBR systems were done in conjunction with the membrane suppliers who also provided valuable design input.

The selection of design MLSS in an MBR and CAS has a significant impact on the reactor and SST size. The MLSS concentration also has an impact on the alpha factor which influences

aeration efficiency. As part of this investigation, an optimum MLSS concentration ($MLSS_{opt}$) cost optimization was done taking into account the effect on reactor size, SST area, membrane area, and aeration CAPEX and OPEX. This resulted in an $MLSS_{opt}$ of 5 500 mg/l and 6 000 mg/l for the CAS MLE and CAS UCT respectively, and 10 000 mg/l for the Zeeweed MBR and Kubota MBR system.

The CAS system had the lowest total cost (CAPEX+OPEX) of the 3 systems over a lifespan of 10 years, with the Zeeweed MBR having the 2nd lowest cost coming in at 61% higher than the CAS system. The Kubota MBR had the highest total cost with a 203% higher cost than the CAS system. In terms of land area requirement, the Kubota MBR required the least amount of land area, followed by the Zeeweed MBR which required 12% more land space. The CAS system required 127-514% more land space at the various DSVI values than the Kubota system. This was due to the additional SST area and a larger reactor requirement.

Acknowledgments

I would like to thank my supervisor, Professor George Ekama, for reviewing my work and always providing constructive feedback in due time.

I would like to thank iX Engineers (Pty) Ltd for the financial assistance and flexible study leave I was given to complete my Master's degree.

I would like to thank the membrane suppliers, Kubota and SEUZ Water for their valuable input.

I would like to thank my wife, Adele Smith, for her many sacrifices and tireless support over the last four years while I was completing courses and writing my thesis.

Lastly, I would like to give praise to my King and Saviour Jesus Christ for granting me the opportunity and ability to complete a Masters degree.

Contents

LIST OF FIGURES

LIST OF TABLES

LIST OF EQUATIONS

SECTION A: MBR LITERATURE REVIEW

1 Introduction

More and more focus is going into establishing more sustainable approaches for wastewater treatment (WWT) in South Africa as well as around the world. Governments are beginning to enforce more economical solutions for WWT, which will have less impact on costs as well as on land area requirements.

1.1 Implementing membranes for solid-liquid separation

Effective solid-liquid separation in biological wastewater treatment is an important step in the process as it has a major impact on effluent quality. Traditionally this has been achieved using Secondary settling tanks (SSTs) for liquid/solid separation in combination with a biological reactor (for biological degradation of organic matter). SSTs, however, require a large space, which can become onerous on land requirements.

Immersed membrane bioreactors (iMBRs) functions by solid-liquid separation taking place by the passing wastewater through membranes. These membranes are immersed in the aeration zone of the bioreactor. This thus eliminates the requirement of SSTs and are becoming more widely used to treat various types of wastewater due to the decreasing cost of membranes and the resultant reduced plant footprint. This, therefore, makes MBR a more sustainable option. Membrane technology has a high initial capital requirement, although this can be offset against the savings resulting from the omission of SSTs and a decrease in land area requirement. (Ramphao *et al.*, 2004)

Additional advantages of MBR over SSTS include the following:

- Insensitivity to sludge settleability and filamentous bulking; this is a major advantage as biological nutrient removal (BNR) systems have been proven to produce rather poor settling sludges (DSVI~150 ml/g) when aerobic mass fractions are low (<60%)
- SSTs are not required resulting in a wastewater treatment plant (WWTP) footprint reduction
- A higher biological reactor mixed liquor suspended solids (MLSS) concentration which can operate at 10 000 – 12 500 mgTSS/l (1 – 1.25%) resulting in a reduced reactor size compared to conventional BNR with settling tanks. Hence a further plant footprint

reduction. An increase in MLSS concentration also promotes the growth of ordinary autotrophic organisms (OAOs) thus enhancing ammonia removal. A further advantage is a reduction in sludge production due to the higher MLSS concentration.

- Production of potentially disinfected effluent as membranes can remove bacteria and viruses, given that the pore size is less than 0.01 µm
- Possible elimination for Waste Activated sludge (WAS) thickening, given that the Reactor concentration is operated at the high end.

It would thus be valuable to do an economical comparison between Conventional activated sludge (CAS) in combination with SSTs and MBR, in terms of Capital Expenditure (CAPEX), Operating Expenditure (OPEX) and land footprint requirement.

1.2 Objectives of this study

The main objective of this study was to evaluate the economical implication of treating Raw wastewater in a CAS BNR system with SSTs, against an MBR system where membranes are used for solid-liquid separation. The modelling done for MBR performed in this study was based on the equations derived from (Ramphao, M. Wentzel, M.C., Merritt, R. Ekama, G.A, Young, T. and Buckley, 2006) and from (Judd, 2011). The modelling of CAS BNR was based on the equations derived from (Ekama, GA. Wentzel, 2008). The modelling of the Secondary Settling tanks was based on the equations derived from (Takacs and Ekama, 2008). Two process configurations (MLE and UCT) were considered for each scenario. The costing for each system was done using actual construction prices from past and current projects. Costing of the MBR systems were done in conjunction with membrane suppliers.

2 Background to MBR

2.1 The use of membranes in an MBR

A membrane is a layer of semi-permeable material which separates substances when a driving force is applied across the membrane surface. Particles which are smaller than the membrane pore size will pass through the membrane, while particles larger than the membrane pore size will be rejected (remain behind). See Figure 2-1:

A membrane used in water and wastewater treatment allows constituents within the feed water to pass through it (thus called the permeate) while at the same time rejecting certain constituents from passing through, thus called the retentate. (See Figure 2-2:)

(Judd, S. Kim, B. Amy, 2008)

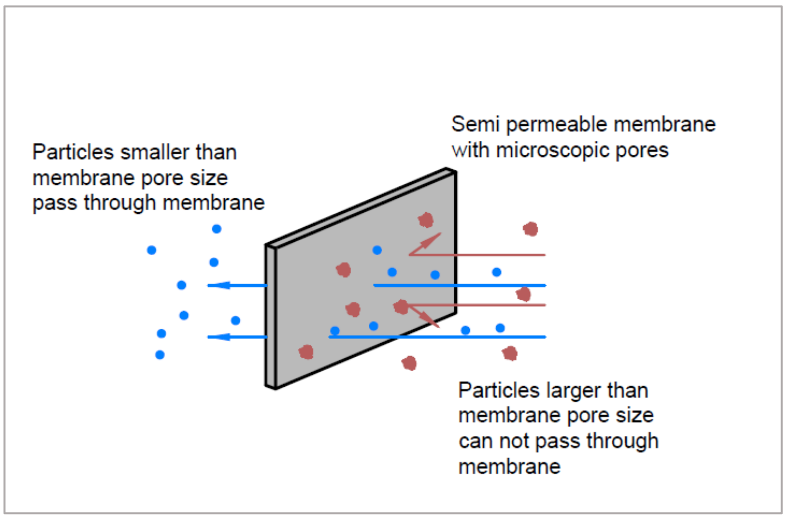

Figure 2-1: *Membrane rejection schematic*

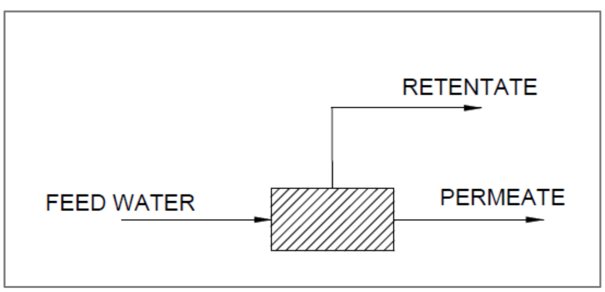

Figure 2-2*: Flow through the membrane*

Membranes used for water and wastewater treatment are typically classified as microfiltration (MF), ultrafiltration (UF), nanofiltration (NF) and reverse osmosis (RO) in an order of decreasing pore size (Bunani *et al.*, 2015).

Figure 2-3 below gives an indication of the various classifications of membranes, by looking at pore size (in µm, where 1 µm = 0.001 mm), and which type of substances in the fluid can be removed.

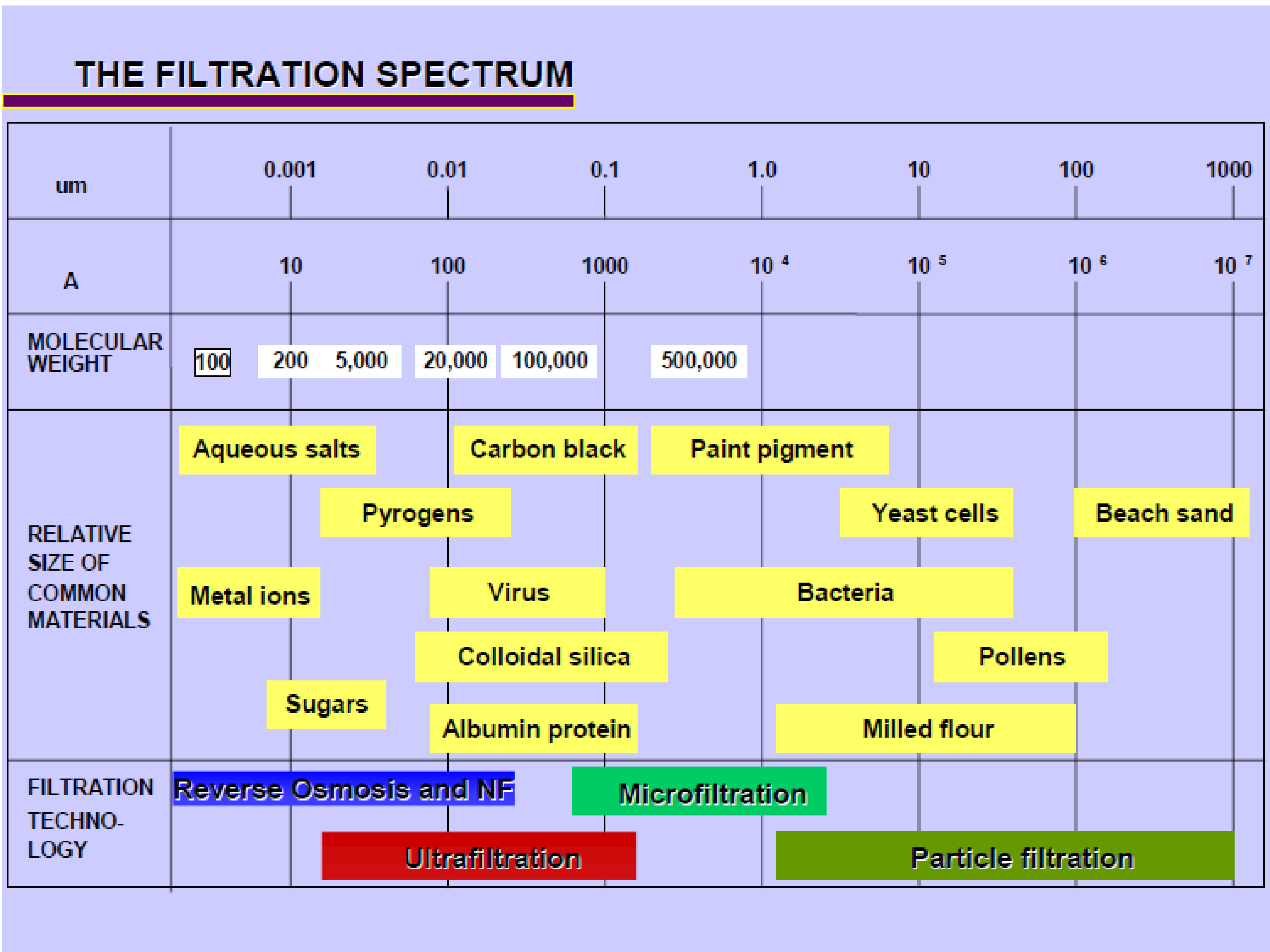

Figure 2-3: *Membrane classification* (Wilf, 2008)

2.2 Advantages of MBR over CAS with SSTs

Membrane Bioreactors (MBR) in wastewater treatment consist of a combination of biological treatment and membrane separation either by microfiltration (MF) or Ultrafiltration (UF). The advantages of MBR over conventional biological treatment in combination with Secondary settling tanks (for solid-liquid separation) which have been found in the literature are:

- Insensitivity to sludge settleability and filamentous bulking; this is a major advantage as biological nutrient removal (BNR) systems have been proven to produce rather poor settling sludges (DSVI~150 ml/g) when aerobic mass fractions are low (<60%)
- SSTs are not required resulting in a WWTP footprint reduction
- A higher biological reactor MLSS concentration which can operate at 10 000 – 12 500 mgTSS/l (1 – 1.25%) resulting in a reduced reactor size compared to conventional BNR with settling tanks. Hence a further plant footprint reduction. An increase in MLSS concentration also promotes the growth of ordinary autotrophic organisms (OAOs) thus enhancing ammonia removal. A further advantage is a reduction in sludge production due to the higher MLSS concentration.
- Production of potentially disinfected effluent as membranes can remove bacteria and viruses, given that the pore size is approximately 0.01 µm

(Ramphao *et al.*, 2004) (Judd, S *et al*, 2008)

Of these, it is the higher bioreactor concentration and omission of SSTs (resulting in a smaller WWTP footprint) and better quality treated water which renders the most significance to plant owners. In South Africa, most wastewater plants are owned and operated by local municipalities. Having said this, when compared to conventional activated sludge BNR with SSTs, it has been found that MBR has the following limitations:

- Potential for membrane fouling if not correctly maintained
- Higher initial capital equipment cost. Although this can be offset against the omission of SSTs and smaller reactor.
- Greater operating complexity. Operating requires diligently controlled process steps which if not carried out correctly could result in membrane fouling.
- Higher aeration requirement. (For biological process and membrane scouring)

Most of the points raised above relate directly or indirectly to membrane fouling. It does not come as a surprise that a huge amount of research has gone into membrane fouling, its cause, removal, and mitigation. (Judd et al., 2008)

2.3 Membrane fouling – main limitation of membranes

One major drawback of membranes is the need to control membrane fouling. According to the International Union of Pure and Applied Chemistry (IUPAC) Working Party on Membrane Nomenclatures, "membrane fouling is the process resulting in a loss of performance of a membrane due to the deposition of suspended or dissolved substances on its external surface, at its pore openings, or within its pores" (Union, Pure and Chemistry, 1996) Membrane fouling may result in a significant decrease in the membrane performance and its lifespan, thus leading to an increase in maintenance and operating costs. As membrane fouling decreases membrane permeability, it results in a reduction in permeate flux, which is defined as the volume (of water in this case) flowing through the membrane per unit area, per unit time. The SI unit used for permeate flux is thus $m^3/m^2.s$, or $\ell/m^2.h$, commonly referred to as "LMH". The performance of a membrane is thus measured by its permeate flux. During operation, evidence of membrane fouling is often seen in an increase in transmembrane pressure (TMP). TMP is the pressure required to push water through a membrane, or the average feed pressure minus the permeate pressure. A constant rise in TMP is often an indication of fouling. (Iorhemen, Hamza and Tay, 2016)

2.3.1 What causes fouling?

Membrane fouling is the main challenge to ensure reliable membrane performance. Fouling is a complex phenomenon involving various factors under various circumstances. Generally fouling is the accumulation of unwanted deposits on the membrane surface or inside the membrane pores. This then causes blocking of the membrane pores which then leads to a decrease in permeate flux and inevitably drastically reduces membrane lifetime. (Jiang, Li and Ladewig, 2017)

Various factors attribute to membrane fouling. These factors can be grouped into three main categories, namely: 1) membrane characteristics, 2) operating conditions and 3) feed and biomass characteristics. A summary of the various factors affecting membrane fouling is indicated in Figure 2-4

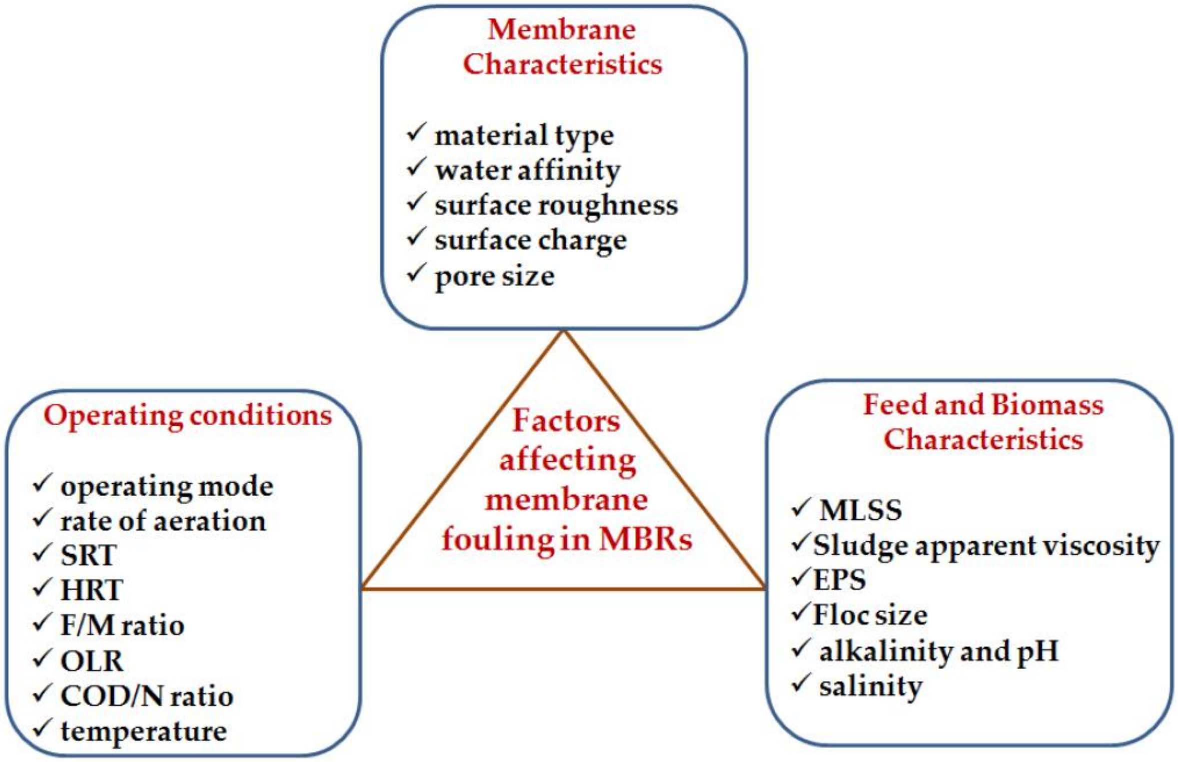

Figure 2-4: *Factors affecting membrane fouling*

To reduce fouling of the membranes in an MBR system, air is provided by the MBR aeration system which provides a crossflow velocity across the membrane surface. The air crossflow velocity provides effective scour of the membrane which is usually achieved via coarse bubble aeration. The aeration system in an MBR system thus provides air for effective scour of the membrane surface, as well as (full or in part) the oxygen required for growth of the biomass. Providing scour across the membranes helps maintain the flux through the membranes, by reducing the build-up of particles on the membrane surface.

To increase scour effectiveness, the total suspended solids (TSS) concentration in the reactor has to be in the order of 8000 – 20 000 mgTSS/l, which is much higher than conventional BNR systems with SSTs. On one hand, the higher concentration is advantageous in today's market as this results in a reduced reactor volume, on the other hand, it is disadvantageous as it reduces oxygen transfer rate, which increases energy consumption, especially for the biological air requirement in the aerobic zone of the bioreactor, which for optimum operating requirements needs MLSS to be kept below 8g/l (FBSA system).

MBR configurations – iMBR vs sMBR – immersed vs side stream

There are two main process configurations, specifically referring to how the membrane is integrated with the MBR: The membrane module is either located externally to the bio-reactor which is known as side stream MBR (sMBR) or is immersed directly into it, which is known as immersed MBR (iMBR). In an iMBR, the membranes are either immersed inside the aerobic tank or are immersed inside a separate membrane tank. The iMBR system where the membranes are located inside the aerobic tank makes use of flat sheet (e.g. Kubota, A3, Alfa Laval) The iMBR system where the membranes are located within a separate membrane tank uses hollow fibre membranes (e.g. Zenon, Koch, Siemens) Figure 2-5

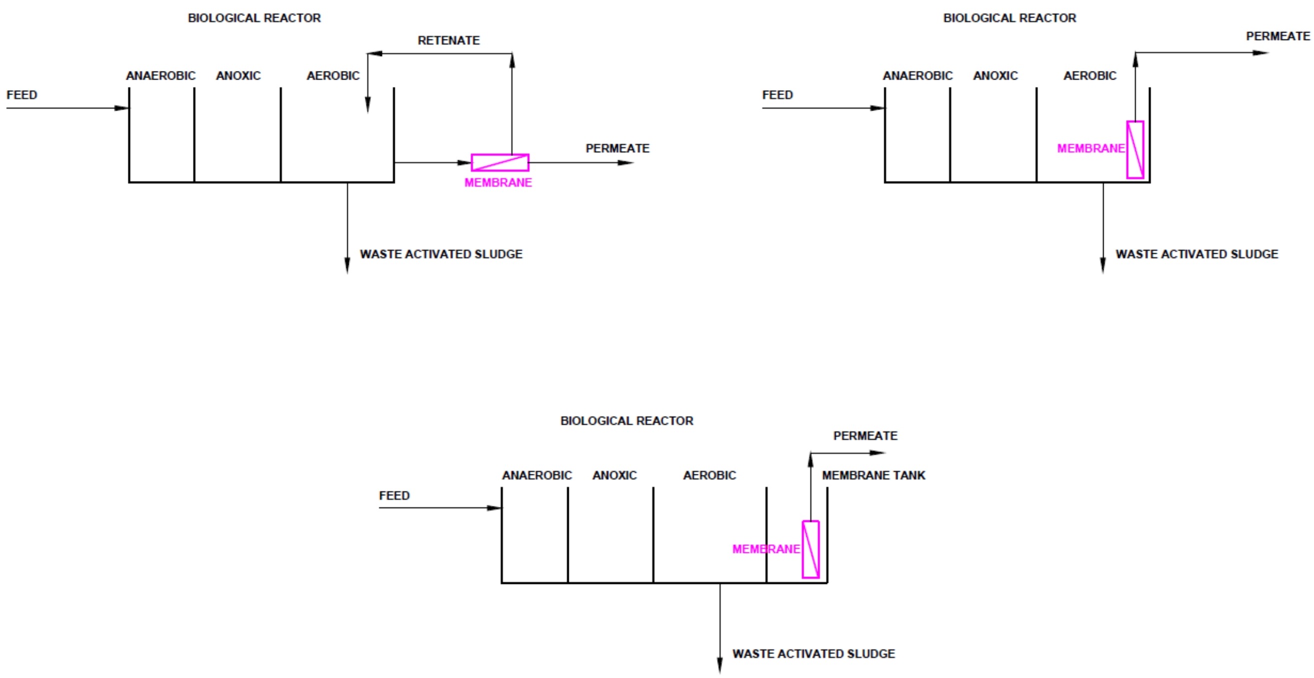

Figure 2-5: *MBR process configurations: (a) external or side-stream MBR (sMBR) –* ***top left*** *(b) submerged or immersed MBR (iMBR) where the membranes are located inside the aerobic tank –* ***top right*** *c) submerged or immersed MBR (iMBR) where the membranes are located inside a membrane tank –* ***bottom centre.***

sMBRs are hydraulically operated by predominantly using pumps (positive pressure) which feed water over the membrane in a cross-flow direction at high velocity. This has the advantage of providing high pressure scouring for fouling control but at the expense of high energy consumption. iMBR predominantly makes use of aeration to provide airlift (vacuum pressure) for hydraulic operation. The use of sMBRs requires significantly more energy to operate than iMBRs, as operation (including scouring) of the membrane is achieved through a pumped side stream crossflow requiring high pressures and flow rates. For iMBRs, operation (including scouring) is achieved by aeration which leads to lower energy consumption, however, permits the operation to lower fluxes. The above ignores the energy requirement for biological treatment (aeration). This makes iMBR a more attractive option as it requires less energy to operate (Judd et al., 2008) (Le-Clech, P. Jefferson, B. Judd, 2005)

MBR was commercialized in the 1970s as a side stream process, where the unit operation was housed externally from the biological reactor. The breakthrough arose in 1989 when Yamamoto and his co-workers came up with the idea to immerse the membrane within the biological reactor. This gave rise to the immersed bioreactor (iMBR). As immersed MBR does not require separate housing as with sMBR, the WWTP footprint is reduced. sMBR are still installed today but to a lesser extent than iMBR. (Singhania *et al.*, 2012) (Judd et al., 2008)

2.4 Membrane configurations used in MBR

There are three main commercial types of membrane configurations currently being used for MBR, these are flat sheet (FS), hollow fibre (HF) and multi-tube (MT). Figure 2-6

Figure 2-6: *MBR membrane types: flat sheet (FS, left), hollow fibre (HF, middle) and multi-tube (MT, right)*

Each type of membrane serves the same purpose, which is to allow water to flow through the membrane material but at the same time preventing certain particles within the water from passing through. What makes each type of membrane different is the water flow pattern through the membrane. Multi-tube (MT) operates by water passing from inside to outside the tube. ('lumen-side' to 'shell-side'), compared to hollow-fibre (HF) which operates with the flow generally passing from outside to inside, whereas flat sheet (FS) operates by allowing water to flow from one side of the sheet to the other. (Judd et al., 2008)

Although a complete review of all the available types of membranes is not possible, a review of the most available reveals that FS and HF are predominantly used for iMBR, with MT predominantly used for sMBR. The 2 most established FS and HF membrane products are the Kubota FS iMBR which boasts over 2,200 installations worldwide and the Suez (previously Zenon) HF iMBR which is being used in the largest MBR installations around the world. The latest Suez product for MBR at the time of this investigation is the Zeeweed 500D. These 2 membranes are shown above in Figure 2-6 (Judd et al., 2008)

These 2 membranes function differently in that the Kubota FS membranes are placed directly into the aeration tank of the biological reactor, whereas the Zeeweed FS membranes are housed in a separate tank/zone which will often form part of the reactor structure. These 2 types of membranes have different air scouring operating requirements. From pilot plants and full-scale plant data, it is reported that the scouring air demand (SAD) for FS membranes is approximately 3 times than that required for HF membranes at 0.75 to 0.9 $Nm^3/(m^2.h)$ compared to 0.15 to 0.3 $Nm^3/m^2.h$. This SAD, however, supplements the biological oxygen requirement in partial or in full. (Judd, 2011)

2.5 Wastewater treatment using CAS with SSTs

2.5.1 Modelling of CAS

Wastewater treatment models are used to simulate wastewater treatment processes. A model is a set of mathematical equations which can be used to explain, predict, decide or design. A model can be coded into a computer program for efficient use. (Gass, 1983) Wastewater treatment models have been developed in two main levels, steady state, and dynamic models. Steady state models are simpler in that they make use of constant flow and loads as an input. They are very useful for design as not all the system parameters need to be known such as kinetic and stoichiometric constants. Dynamic models are more complex than steady state models in that they use variable flows and loads with time being a factor. Steady state models are mainly used by designers, amongst other things, to determine things such as the bioreactor volume, expected sludge production volumes, existing plant capacities, predicted effluent quality, power requirements, and potential energy recovery. Dynamic models are mainly used to predict a specific system's response over time, that being of an existing or proposed system. Various dynamic models for the ASP (activated sludge process) have been developed over the years, which include amongst others ASM1, UCTOLD, and UCTPHO. (Henze et al, 2008) (Ekama & Wentzel, 2008)

Influent wastewater consists of organic and inorganic matter, with the organics consisting of biodegradable and unbiodegradable constituents. In order to predict each constituent's transformations in the bioreactor, it is also important to characterize the wastewater physically namely soluble and particulate.

Figure **2-7** below gives an indication of the different categories within wastewater (organic, inorganic, particulate, soluble), each constituents transformation and how that constituent contributes to the sludge.

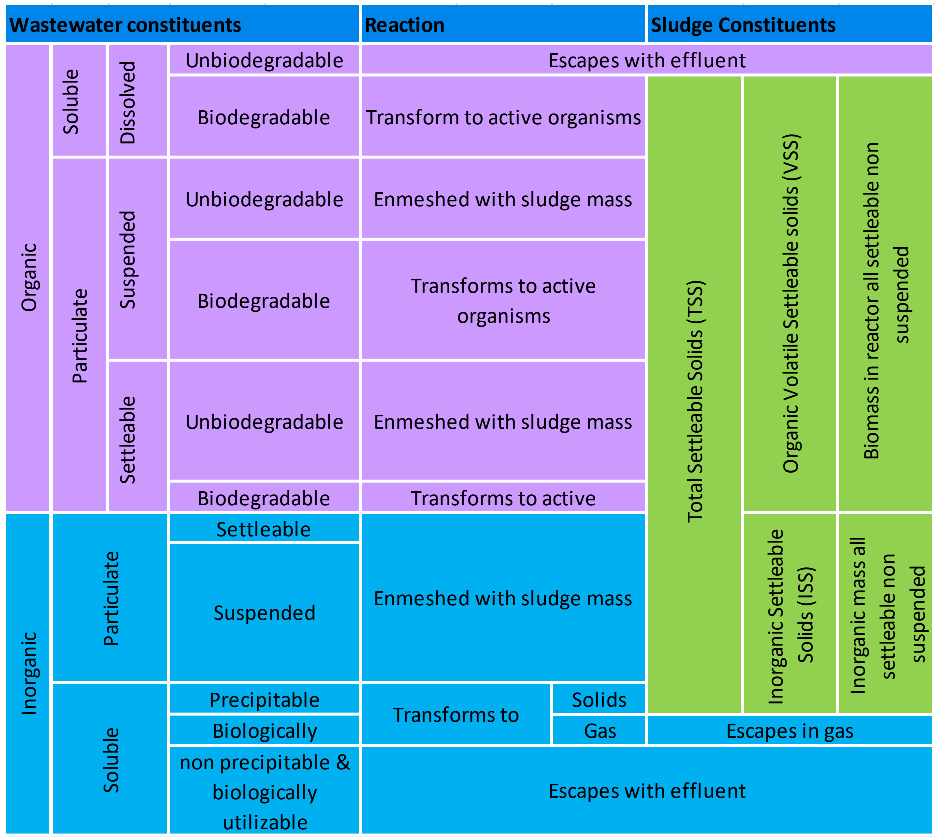

Figure 2-7: *Transformation reactions of organic and inorganic wastewater constituents from the particulate and soluble forms in the solid and liquid phases to the solid phase and sludge constituents, and gas and liquid phases escaping to the atmosphere and the effluent respectively.*

2.6 Membrane design considerations and modelling

Ramphao *et al* (2004) conducted a lab-scale study of a BNR AS, with an iMBR where the membranes were located inside the aerobic tank, in place of a conventional SST. The membranes used in the study was the Kubota FS where they concluded that the use of membranes for solid-liquid separation makes a significant difference in the design of BNR systems and the approach to the wastewater treatment in general. The design procedure for sizing an iMBR system where the membranes are located inside the aerobic tank is as follows: After the reactor zone mass fractions and recycle rations have been determined to ensure biological nutrient (N and P) removal, the PWWF along with the required sludge age to achieve the required aerobic MLSS concentration fixes the membrane surface requirement. This determines the volume requirement for the membrane cartridges. In addition to this, the membrane system also has a volume requirement for its aeration system (for scouring). If the membrane aeration demand meets the biological aeration demand, then the aerobic zone volume is determined from the membrane volume requirement. In the case where the membrane aeration demand cannot meet the biological aeration demand, additional aeration is then required to supplement the biological demand, and the aerobic zone must then be sized on the membrane volume requirement along with additional aeration devices. The aeration tank must thus be increased in order to meet the additional biological aeration demand. A detailed procedure for the design of an iMBR system is described in Ramphao *et al* (2004)

The design of an iMBR system where the membranes are located inside a separate (membrane) tank is as follows: The membrane area (A_m) required is fixed by the peak membrane flux ($j_{net, peak}$) and peak flow (Q_{pdwf}). The membrane packing density together with the membrane area fixes the membrane tank volume. The membrane air demand is determined based on the membrane area and is in the order of 0.15 - 0.3 $Nm^3/(m^2.h)$.

(Du Toit, 2006) conducted a lab-scale study to better understand the operating conditions and considerations of an MBR BNR system including the phenomena of increased sludge production and oxygen transfer in high concentration sludges. The study was conducted by running two parallel lab-scale MBR and CAS systems in order to monitor their performance and behavior. From the investigation, it was reported that the membranes produced an effluent of equal or superior quality to that produced by a conventional system using SSTs. Higher sludge productions of 0.311 (mgVSS/d)/(mgCOD/d) were observed in the MBR system, which is partly attributed to the retention of solids by the membranes. The steady state model closely predicted MBR system performance for COD and nitrification, however for de-nitrification, the theoretical D_{pp} was being underpredicted, requiring $K_{2'T}$ to be adjusted from 0.145 to 0.216 mgN/mgVSS/d at 20°C, in order to match the observed values. The BEPR predictions for aerobic P uptake were close to that observed from the systems, f_{XBGP} observed (0.376 mgP/mgVSS) was close to that determined theoretically of 0.38mgP/mgVSS. Aeration testing reported alpha values of 0.5-0.6 for 15000 mgTSS/l and 0.2-0.3 for 20 000 mgTSS/l. A comparison of laboratory CAS and MBR UCT systems showed that the biological kinetic rates associated with biological N removal and enhanced biological P removal in the steady state and dynamic simulation models were not decreased in MBR systems operating at high MLSS up to 20 gTSS/l, except the maximum specific growth rate of nitrifiers, which was about 20% lower in MBR systems at high MLSS concentration. (du Toit *et al.*, 2010, Parco *et al*., 2018). It was concluded that “the performance of membrane BNR systems can be simulated with current BNR activated simulation models…” (Parco *et al*., 2018)

2.7 Conclusions

MBRs have numerous advantages over CAS systems such as insensitivity to sludge settleability, smaller footprint due to the omission of SSTs, further footprint reduction due to a smaller reactor volume as a result of a higher reactor MLSS concentration. MBRs also have disadvantages such as the potential for membrane fouling if not correctly maintained, higher initial capital equipment cost, greater operating complexity, and a higher aeration requirement. (For biological process and membrane scouring). Even though MBR systems have been around for more than 50 years, the technology is still relatively new in South Africa with only a handful of MBRs in operation. With the current economic challenges facing the country, municipality's are under more pressure to implement services within a restricted budget. Municipalities thus need more sustainable approaches to WWT which costs less. It would, therefore, be valuable to do an economical comparison between conventional activated sludge (CAS) in combination with SSTs and MBR, in terms of Capital Expenditure (CAPEX) and Operating Expenditure (OPEX).

3 Designing an iMBR system

An MBR plant may consist of the following unit processes: Course and fine screening, grit removal; fats, oils, and grease removal; flow equalization; primary clarification; biological treatment; membrane separation; and disinfection. Although an MBR facility may include all of the above-mentioned unit processes, specific situations may require additional treatment in the case of targeted contaminant removal. Specific conditions may depict a shorter process train, land space constraints, budget constraints or low incoming flows. The information required to establish a biological design for an MBR is the same as for a CAS plant, which includes wastewater characteristics, flow rates, environmental conditions, and treatment objectives. (WEF, 2011)

The design of an immersed membrane bioreactor (iMBR) is generally based on a combination of empirical data and biokinetics/biochemical stoichiometry. Three interrelated design phases are required namely **1)** The membrane process along with its air demand **2)** the biological process along with its oxygen demand **3)** aeration systems (membrane aeration system supplements the biological system)

Two types of MBR configurations were considered as part of this investigation: **1)** iMBR using HF membranes where the membranes are located in a separate membrane tank and **2)** iMBR using FS membranes where the membranes are located inside the aerobic tank.

The membranes which formed part of this investigation was the Kubota FS (flat sheet) membranes and the Zeeweed HF (Hollow fibre) membranes.

3.1 Design of iMBR using HF membranes

In an MBR system with HF membranes, **the membranes are located in a separate membrane tank** following the aerobic tank. (Figure 3-2 and Figure 3-2)

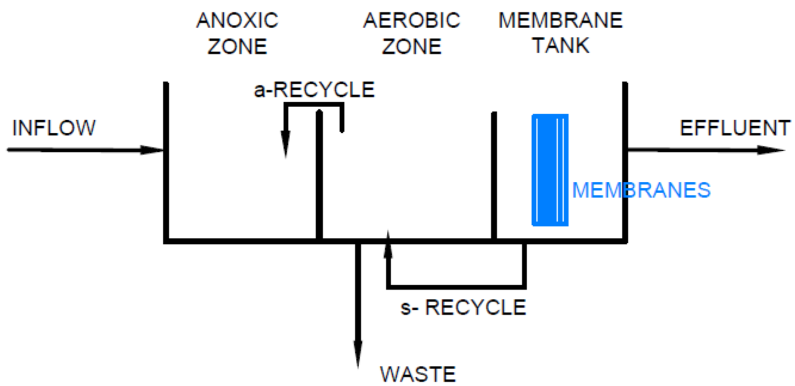

Figure 3-1 *– Process flow schematic of HF MBR in MLE configuration*

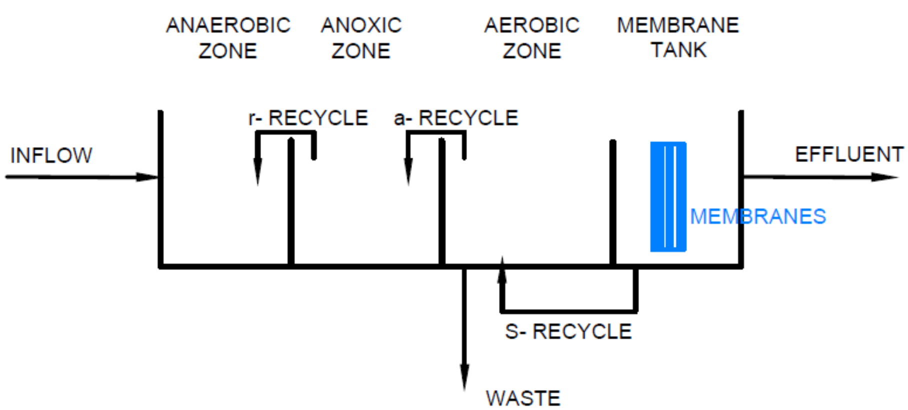

Figure 3-2 *– Process flow schematic of HF MBR in UCT configuration*

3.1.1 Determining the volume of the membrane tank

One of the key parameters in the design of the membranes is the **net flux (j_{net}),** which is defined as the quantity of liquid passing through a unit area of membrane per unit time, with SI units of $m^3/(m^2.h)$. The values are normally more accessible in non-SI units as liters per m^2 per hour (LMH). Fluxes are dependent on influent flow rate, permeability and temperature. The peak fluxes at Peak Dry Flow ($j_{net,\ pdf}$) and Peak Wet Flow ($j_{net,pwf}$), which are the maximum fluxes allowed during a limited time period and are normally taken as 140% and 170% respectively of j_{net}.(SUEZ, 2019) One then determines the membrane area for each flow scenario (Q_{adwf}, Q_{pdwf}, Q_{pwwf}) and selects the membrane area required. The required area of the membranes (A_m) is calculated by Equation 3-1

Equation 3-1: Minimum required membrane area

$$A_m = \frac{Q}{j_{net}}$$

Where:

A_m = Minimum membrane area required (m^2)

Q = Average Dry flow rate, Peak Dry flow rate or Peak Wet Flow rate (m^3/h)

j_{net} = flux at specific flow rate ($m^3/m^2.h$)

The greater the difference between Q_{ADWF} and the Peak flows, the greater the required membrane area, resulting in an increased membrane tank volume which results in a larger CAPEX cost. The larger membrane tank also increases the membrane aeration demand resulting in an increased energy requirement (Higher OPEX). It may thus be more economical to consider the installation of flow balancing (equalization) although this comes with its own challenges such as potential odour problems when the equalization tank is empty during low flows, and dealing with settled solids accumulating at the bottom of the tank.

Another key parameter in membrane design is the packing density (φ) of the membranes which is the area of membranes required per unit volume of membrane tank. (m^2/m^3). The packing density value is manufacturer-specific, which typically ranges between 40-300 m^2/m^3. The membrane area together with the packing density fixes the minimum required membrane tank volume ($V_{m,\ min}$), thus:

Equation 3-2: Minimum required volume of the membrane tank

$$V_{m,min} = \frac{A_m}{\varphi_{tank}}$$

Where:

$V_{m,min}$ = Minimum membrane tank volume (m^3)

A_m = Membrane area

φ_{tank} = Membrane packing density (m^2/m^3)

3.1.2 Aeration system design for HF membranes

The design of the aeration system is one of the most important differences between iMBR and CAS since iMBR has both membrane and biological oxygen requirements. It is thus required to determine how much of the membrane aeration contributes to the total oxygen requirement for biological breakdown. Dissolved Oxygen (DO) created by the membrane aeration system will be available for biological breakdown. It is assumed that all of the DO provided by the membrane aeration system contributes towards the oxygen required for biological breakdown. This is most likely an overestimation, and a more accurate representation requires the use of the International Water Association Activated Sludge Models under dynamic conditions. Equation 3-3 is used for calculating the membrane aeration rate:

Equation 3-3: Membrane aeration rate

$$Q_{A\prime m} = SAD_m . A_m$$

Where:

$Q_{A\prime m}$ = Membrane aeration rate (Nm^3/h)

SAD_m = Membrane aeration required for scouring as per manufacturer ($Nm^3/m^2.h$)

A_m = Membrane area (m^2)

It is then required to determine how much oxygen is to be supplemented by the biological aeration system, in order to achieve the Peak Oxygen demand for biological breakdown. For effective scour of the membranes, the membrane aeration system is normally in the form of course bubble aeration, due to the larger bubble surface area compared to fine bubble aeration. The biological aeration system is normally supplied in the form of fine bubble aeration which is more efficient for oxygen transfer. The fact that course bubble aeration is less efficient for oxygen transfer than fine bubble aeration, must be taken into account when determining how much of the membrane aeration system contributes to the Peak Biological oxygen demand.

3.1.3 Biological design

The biological design of a HF MBR system is very similar to that of a CAS system with the following minor differences. A relatively short retention time is required in the membrane tank to limit the concentration of the solids and to prevent membrane clogging. This demands that the recycle flow rate (the return activated sludge) from the membrane tank to the biological tank needs to be in the region of 3-5 times the Q_{adwf}. The return activated sludge (RAS) is also relatively high in DO, which makes the anoxic zone, to where it is usually returned to for denitrification, less efficient. It is thus usually returned to the aerobic zone.

3.2 Design of iMBR using FS membranes

For the design of an iMBR system with FS membranes, the membranes are located **within the aerobic tank** of the biological reactor as indicated in Figure 3-3 and Figure 3-4

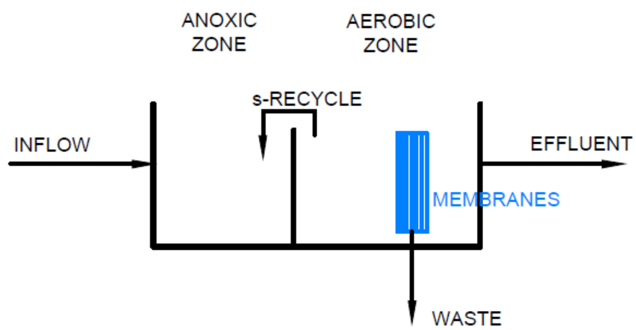

Figure 3-3 – *Process flow schematic of FS MBR in MLE configuration*

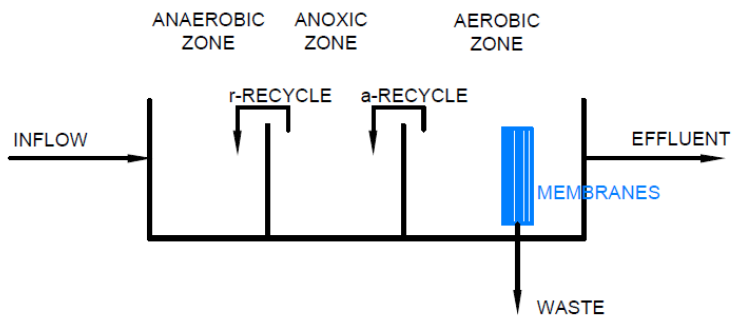

Figure 3-4 – *Process flow schematic of FS MBR in UCT configuration*

3.2.1 Determining Aerobic zone volume for membranes

Based on the research conducted by Ramphao et al (2004), for the design of an iMBR system using FS membranes, the volume of the aeration zone required by the membranes is governed by the PWWF and the maximum flux (m^3/m^2.day)

The aerobic zone volume requirement to accommodate the membrane $V_{aer,m}$ is thus give by :

Equation 3-4: Minimum Aerobic zone volume for membranes

$$V_{aer,m} = L \frac{f_q . Q_{ADWF}}{q_{max}}$$

Where:

L = Membrane Packing density = 0.0875 m³/m² for single-story membranes and 0.0625 m³/m² for double story membranes as per Kubota.

Q_{ADWF} is in Ml/d

q_{max} = maximum membrane flux m³/(m².d) which depends on the type and manufacturer of the membranes. In the case of flat sheet membranes, this is usually between 0.4 and 1.1 m³/m².d

f_q = $Q_{PWWF/ADWF}$ ratio

Thus from Equation 3-5 and Equation 3-4:

$$\frac{S_{ti}}{X_{taer}}.[A].f_{maer} \geq V_{aer} = \frac{0.1}{q_{max}}.f_q.Q_{ADWF} \quad \text{m}^3$$

In the above equation, there are two unknown values namely X_{taer} and the sludge age in the "A" term, which is the same as the Load Factor kgTSS in reactor per kgCOD/d applied to the reactor. As an MBR requires the aerobic TSS concentration to be in the range of 10 to 15 gTSS/l, the required concentration can be selected, and the sludge age calculated. This is then the sludge age required to ensure that the required amount of sludge is generated to ensure the required aerobic zone TSS concentration for those specific influent characteristics, membrane flux, influent flow rate, and system design parameters.

Since the required membrane surface area is directly dependent on the PWWF, the inclusion of flow equalization will thus reduce the required membrane surface area for the same ADWF, which in result reduces the volume of the aerobic zone. The introduction of flow balancing in a CAS system will however not affect the size of the biological reactor but will result in a decrease in SST size and peak oxygen demand.

3.2.2 Determining Aerobic zone volume for biological breakdown

First, it is necessary to calculate the volume of the aerobic zone required for biological treatment. This is done in the same way as for CAS using the Steady State ASM equations. This aerobic zone volume must be equal to or larger than the volume required to house the FS membranes. In the equation below, the aerobic zone sludge mass fraction (f_{aer}) is determined based on the anoxic and anaerobic mass fraction (f_{anx} & f_{ana}) requirements for N & P removal, the mass of solids (MX_t) in the reactor is determined by the wastewater characteristics, kinetic and stoichiometric constants. This leaves 2 unknowns, namely the sludge age and aerobic TSS concentration. Since the membranes require a specific TSS concentration, the sludge age can be calculated. This aerobic zone volume must be larger than the volume required for the membranes. If required, this aerobic volume increase is achieved by increasing the system sludge Age (Rs) which in effect increases the MXt. An increase in MXt will affect the entire reactor volume as f_{anx} & f_{ana} are fixed, by the N and P removal requirements.

Equation 3-5: Volume of the aerobic zone

$$V_{aer} = \frac{MX_t . f_{maer}}{X_{t,aer}}$$

Where:

V_{aer} = Aerobic volume (Ml)

MX_t = Mass of Total suspended solids in the reactor (kgTSS)

f_{maer} = Aerobic mass fraction

$X_{t,aer}$ = Total suspended solids concentration in Aerobic zone (mg/l)

This volume must be equal to or larger than the volume required for the membranes

3.2.3 Relationship between zone mass fractions and volume fractions

For all BNR systems with membranes, the concentrations of the TSS in the anaerobic, anoxic and aerobic zones as fractions of the average system TSS concentration are equal to the ratio of the sludge mass fraction and volume fraction of the zones (Parco et al. 2018) i.e.:

$$\frac{X_{tana}}{\bar{X}_t} = \frac{f_{mana}}{f_{vana}}; \frac{X_{tanx}}{\bar{X}_t} = \frac{f_{manx}}{f_{vanx}}; \frac{X_{taer}}{\bar{X}_t} = \frac{f_{maer}}{f_{vaer}}$$

Where:

f_m, f_v = zone sludge mass and volume fractions respectively

X_t = zone TSS concentration

$\bar{X}_t$ = average TSS reactor concentration, and

Subscripts ana, anx, aer are the anaerobic, anoxic and anaerobic zones respectively.

In BNR systems with membranes in the aerobic zone, the sludge mass distributes itself differently in the different zones of the system when compared with systems with SSTs. This is because the effluent is withdrawn via the membranes from the aerobic zone which concentrates the sludge in this zone relative to that in the other zones. However, the concentrated aerobic sludge is diluted by the less concentrated incoming sludge stream from the upstream zones. The higher the recycles from the downstream zones to upstream zones, the more uniformly the sludge mass is distributed around the system and the closer the sludge concentrations are in the different zones.

For a UCT system with membranes (Figure 3-3), the volume fractions (with respect to the total reactor volume) of the anaerobic, anoxic and aerobic zones (f_{vana}, f_{vanx} and f_{vaer}), and the anaerobic, anoxic and aerobic TSS concentrations (X_{tana}, X_{tanx} and X_{taer}), are related to the anaerobic and aerobic mass fractions (f_{mana}, and f_{maer}), recycle ratios (a & r) and system average TSS concentration $\bar{X}_t$, as follows (Parco *et al.*, 2018):

$$f_{vana} = \frac{f_{mana}(r+1)}{D\,r}$$

$$f_{vanx} = \frac{1 - f_{mana} - f_{maer}}{D}$$

$$f_{vaer} = \frac{a\,f_{maer}}{(a+1)D}$$

$$X_{tana} = \bar{X}_t \frac{r.D}{(r+1)};\ X_{tanx} = \bar{X}_t.D;\ X_{taer} = \bar{X}_t \frac{(a+1).D}{a}$$

Where $D = 1 + \frac{f_{mana}}{r} - \frac{f_{maer}}{a+1}$

The above equations also apply to the MLE ND (Figure 3-3) system, but the anaerobic mass fraction (f_{mana}) and the r recycle are set to 0.

3.2.4 Aeration system design for FS membranes

The design of the aeration system for FS membranes follows the same approach as for HF membranes, as some or all of the biological oxygen requirement is supplemented by the membrane aeration system. From the study conducted by Ramphao, 2005 and confirmed by Kubota, the oxygen required by the membrane system is:

$$OTR_{mbr} = 0.3 Q_{air} OTE_{mbr} \alpha V_{aer} 100\ x\ 24\ [kgO/d]$$

Where:

OTR_{mbr} = Oxygen transfer rate required for scouring of membrane system

0.3 = kg oxygen per Nm^3 of air

Q_{air} = 8.6 $Nm^3/h.m^3$ for single-story membranes and 12.9 $Nm^3/h.m^3$ for double story membranes

OTE_{mbr} = % Oxygen transfer efficiency. 5% for 3.5 water depths and 7% for 5m water depths.

α = alpha correction factor

V_{aer} = Aerobic zone volume required for membranes in Ml

This OTR required for effective scour supplies some or all of the biological oxygen demand in the aerobic zone. The biological oxygen demand for an MBR is determined the same way as for CAS BNR systems. The difference, if positive, is the requirement for biological treatment in surplus of the requirement for membrane scouring.

It is important to note that should the membrane oxygen demand not meet the peak biological oxygen demand, additional oxygen has to be supplied to supplement the latter. This additional oxygen cannot be supplied into the membrane section of the aerobic reactor, so additional aerobic reactor volume needs to be provided to enable the transfer of the oxygen deficit. This is usually done with fine bubble aeration. Therefore, the volume of the aerobic zone (and thus the biological reactor) is governed by either the volume requirement for the membranes or the biological oxygen demand, whichever one is greater volume. Generally, more concentrated wastewater will require additional fine bubble aeration. (Ramphao, M. et al, 2004)

3.3 Comparison of kinetic growth rates for biological N and P removal between MBR and CAS systems

As part of the WRC project K5/1537 and WRC consultancy K8/514, batch tests were performed on two identical parallel lab-scale UCT ND BEPR systems, with the objective to establish if there are any differences in biological kinetic growth rates in systems with higher VSS concentrations as is the case in MBR systems, compared to systems with lower VSS concentrations as is the case in conventional BNR system.

3.3.1 Requirement for Nitrification

Aerobic batch tests were performed on two identical parallel lab-scale UCT ND BEPR systems. It was found that the MBR system (with the higher VSS concentration up to 12gVSS/l) showed lower VSS specific ammonia utilization rates and hence also lower Autotrophic Nitrifier organism (ANO) maximum specific growth rates (μ_A) when compared with the CAS system. The reasons for the different organism behaviors with increasing concentrations are possibly due to 1) In MBR systems nitrifier organism loss via the effluent does not occur, retaining all ANOs including slow-growing ones. 2) At the higher VSS concentration, oxygen and ammonia transport limitations decrease the observed VSS specific ammonia utilization rate (SAUR) and μ_A (du Toit *et al.*, 2010)

3.3.2 Requirement for Denitrification

From anoxic-aerobic batch tests, the OHOVSS specific denitrification rate by OHOs (K_{2OHO}) utilizing slowly biodegradable organics (SBO) obtained at different MBR system VSS concentrations (3 to 12 gVSS/l) and different initial nitrate concentrations (10 to 90 mg NO_3-N/l), showed no effect to initial nitrate concentration. From the batch tests, the average K_{2OHO} was 0.264 mg NO3-N/(mg OHOVSS.d), which is very close to the value of 0.255 reported in the literature from Ekama and Wentzel (1999) (du Toit *et al.*, 2010)

3.3.3 The requirement for biological P removal

From anaerobic-anoxic-aerobic batch tests, it was found that the average PAOVSS specific anaerobic acetate uptake and P release rates and the aerobic P uptake rate obtained over different VSS concentration ranges were within the range as reported in the literature. It can, therefore, be confirmed that increasing VSS concentrations does not affect the rates.

3.3.4 Conclusion on Kinetic growth rates in MBR

The results from the WRC investigation (2010) and Parco et al (2006) show that the BNRAS steady state and kinetic models which have been developed for low VSS concentrations can be applied with reasonable confidence to MBR systems with higher VSS concentration, except for maximum specific growth rate of nitrifiers, which was observed to be significantly lower in MBR systems.

As the system sludge age is determined based on the pre-selected aerobic mass fraction which in effect fixes the aerobic volume requirement for the membranes, the lowest maximum specific growth rate of the nitrifiers at 20 C to ensure nitrification can be calculated. From (WRC, 2002) and (Ramphao, M. et al, 2004) it can be seen that:

$$\mu_{nm20} = \frac{b_{n20}(\theta_b)^{(T-20)} + {}^{S_f}\!/_{R_s}}{f_{maer}(\theta_\mu)^{(T-20)}} \quad /d$$

Where:

μ_{nm20} = maximum specific growth rate of the nitrifiers at 20°C

b_{n20} = endogenous mass loss rate of nitrifiers at 20°C = 0.04/d

θ_μ = temperature sensitivity coefficient for growth of nitrifiers = 1.123 (WRC, 1984)

θ_b = temperature sensitivity coefficient for endogenous mass loss of nitrifiers = 1.029

R_s = system sludge age (d)

S_f = factor of safety on nitrification = R_s/R_{sm}, where

R_{sm} = minimum sludge age required for nitrification.

Ramphao et al. (2006) concluded that because the SRT in MBR systems with FS membranes is likely to be long, to achieve the required high TSS concentration for effective membrane scour, achieving nitrification will unlikely be a problem.

3.4 Using the steady state theory to predict MBR performance

3.4.1 Denitrification

(Du Toit, 2006) reported that the steady state theory was under predicting the D_{pp} (denitrification potential) thus requiring $K_{2'T}$ to be adjusted from 0.145 to 0.216 mgN/mgVSS/d at 20C in order to match the observed system values

3.4.2 Sludge production

(Du Toit, 2006) and Ramphao et al (2004) reported higher sludge production in an MBR system of 0.311 and 0.32 (mgVSS/d)/(mgCOD/d). This increases in the production of sludge is partly attributed to the retention of all solids by the membranes. The increased sludge production is accommodated in steady state theory by increasing the unbiodegradable particulate COD fraction ($f_{S'up}$) to 0.200. Similarly, the unbiodegradable soluble fraction ($f_{S'us}$) must be decreased to account for the soluble COD retained by the membrane attributed to the finer pore size.

4 Cost optimization for selecting MLSS in an MBR system

One of the most important aspects of designing an MBR system is the selection of the design MLSS concentration in the membrane/aeration tank. The MLSS concentration has a direct effect on the reactor volume thus affecting CAPEX. A higher MLSS concentration results in a smaller reactor volume thus decreasing the CAPEX. On the contrary, a higher MLSS concentration has a negative effect on the aeration efficiency, flux, chemical cleaning frequency of the membranes and membrane replacement frequency which results in a higher cost. The above factors should be taken into account when the MLSS concentration is chosen.

4.1 MLSS concentration effect on Aeration in an MBR

The α -factor has a significant impact on the efficiency of the aeration system. The α -factor is dependent on the MLSS concentration in the bioreactor. Noting that MBRs are designed with a much higher MLSS concentration (>9000 mg/l), the higher concentration will have a greater effect on the aeration efficiency and ultimately, the operating costs.

The effect MLSS concentration has on the α-factor can be determined using the following equation from (Judd, 2011)

Equation 4-1: Alpha value

$$\alpha = e^{\omega \cdot X_t}$$

Where:

X_t = MLSS concentration

ω = Correction factor component = - 0.084 (Judd, 2011)

Further studies have been done to investigate the impact MLSS has on the α -factor. Figure 4-1 provides a summary of the proposed equations linking MLSS concentration to the α -factor:

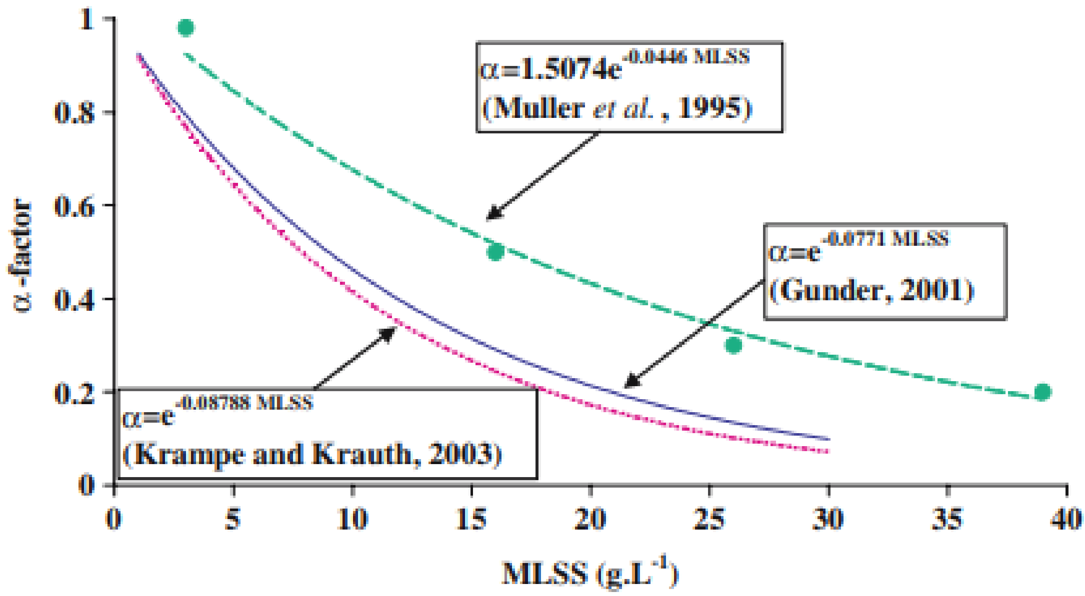

***Figure 4-1** – Alpha values and MLSS concentration* (Germain and Stephenson, 2014)

A pilot study at the Beverwijk WWTP, the Netherlands

A comparative MBR pilot plant was carried out at the Beverwijk WWTP between 2000 -2004 using 4 different types of membrane technologies: Kubota and Zenon being part of them. The MBR pilot was then run in parallel with the plant's existing CAS process.

There was a noticeable impact on the α -factor recorded, with the value decreasing from 0.78 – 0.79 for the CASP operation at the works, to **0.43 – 0.54** across all MBR technologies. The MLSS concentrations of the MBR systems ranged from 10.4 - 12 g/l. (Judd, 2011)

Using the above equations provides α -factor values within the range as observed from the pilot study at the Beverlijk WWTP thus providing confidence in the use of Equation 4-1.

A higher MLSS concentration results in a lower α -factor which decreases the Oxygenation rate (kgO/kWh). This results in a higher aeration power requirement, thus increasing CAPEX and OPEX. This must be offset against the savings occurred by a higher MLSS resulting in a smaller reactor volume.

The effect that MLSS has on the aeration CAPEX and OPEX must be taken into consideration when selecting a design MLSS. As part of this study, a cost optimization will be done, taking into account the effect the MLSS concentration has on the total cost of the project (CAPEX +OPEX). An optimized MLSS will be selected for the design which should result in the lowest Total cost (CAPEX+OPEX)

4.2 MLSS concentration effect on membrane flux

Membrane fouling has a direct impact on the operating flux of the membrane, with flux decreasing over time when fouling is present. While MLSS concentration provides a reasonable indication of fouling propensity, the relationship between MLSS and fouling is complex. The impact of increasing MLSS on membrane permeability can be either negative, positive or insignificant. The MLSS concentration did not appear to have a significant effect on membrane fouling between 8 and 12 g/l. Empirical relationships predicting flux from MLSS concentration have been proposed. However, these equations have limited use as they are generally obtained under very specific conditions. The lack of a clear correlation between MLSS concentration and any specific foulant characteristic indicates that MLSS concentration alone is a poor indicator of membrane fouling. (Judd, 2011) Some research has investigated the effect of increasing MLSS concentration has on membrane flux. Yigit *et al.,* (2008) reported a decrease in flux from 30 to 10 L/m^2.h with increasing MLSS from 4 600 to 12 600 mgTSS/l. The effect of MLSS on the membrane flux will be included this investigation's cost optimization calculation, in order to determine an optimal MLSS for sizing an MBR. This is further detailed in Section 9.

5 Optimizing the CAS MLE process

The MLE process is a modification from the conventional activated sludge system with the addition of an anoxic zone upstream of the aerobic zone. Nitrates produced from ammonia in the aerobic zone (Nitrification) are recycled back into the anoxic zone (a-recycle) and become the electron acceptor source for bacteria in the anoxic zone (denitrification), where nitrates are converted into nitrogen gas that escapes to the atmosphere. Sludge is returned from the clarifier (RAS) and the aerobic zone into the anoxic zone.

A schematic of a typical MLE system is shown in Figure 5-1

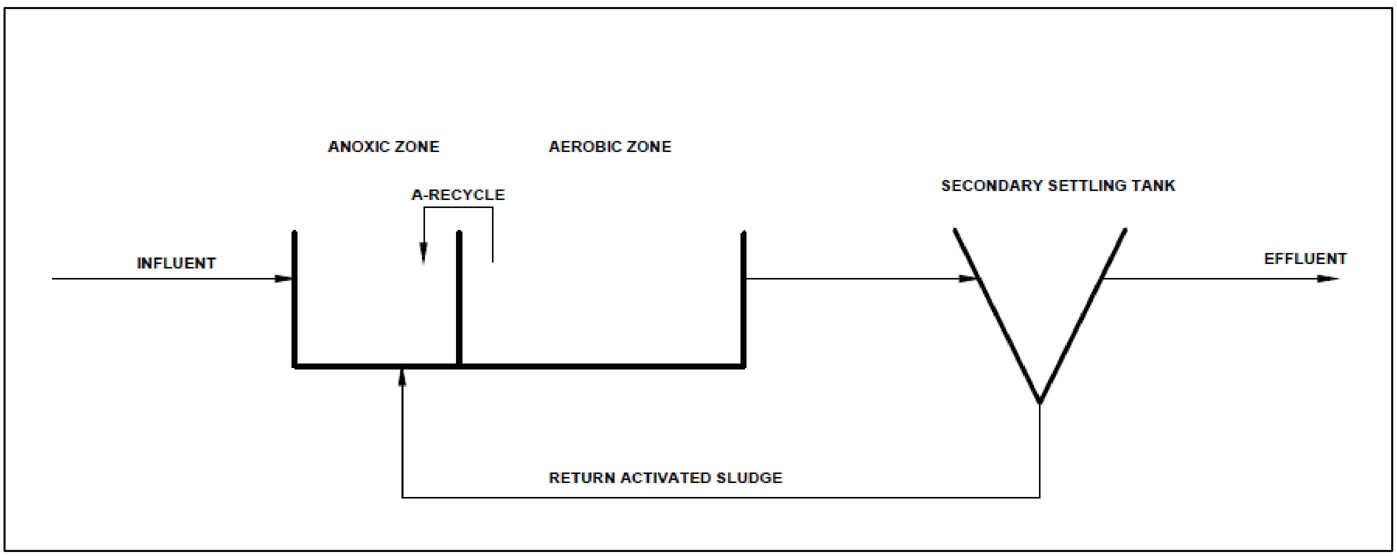

***Figure 5-1** – Typical schematic of MLE system*

5.1 Optimizing MLSS concentration in Reactor

The selection of a system MLSS concentration has a significant effect on the size of the bioreactor and SST. A higher MLSS results in a smaller reactor volume but larger SST area, compared with a lower MLSS concentration which will result in a larger reactor volume but smaller SST area. An optimum MLSS concentration is determined from a cost minimization calculation as part of this investigation.

5.1.1 Cost of Reactor

From the steady state activated sludge model, reactor volume is a function of the wastewater characteristics and sludge age and selection of MLSS concentration (X_t). (Henze M. et al., 2008; Ekama, 2011)

Equation 5-1: Volume of the reactor from steady state model

$$V_p = \frac{Q_{i,adwf}.S_{ti}}{X_t}\left(1 - f_{s'up} - f_{s'us}\right)\frac{Y_H R_s}{(1 + b_H R_s)}(1 + f b_H R_s + f_{iOHO}) + \left(\frac{f_{s'up}}{f_{cv}} + \frac{X_{loi}}{S_{ti}}\right) R_s \qquad m^3$$

Equation 5-1, therefore, indicates that an increase in Reactor concentration (X_t) results in a decrease in reactor volume (V_p), which in effect decreases the cost.

5.1.2 Cost of Secondary Settling tank

From the 1D idealized flux theory, the maximum overflow rate (q_{Amax}) should not be greater than the settling velocity of the SST feed concentration ($V_s\ of\ X_t$) reduced by the flux rating (f_j) to account for non-idealities. The SST area is therefore determined by Equation 5-2 (Ekama et al., 1997)

Equation 5-2: Minimum SST area required

$$A_{ST} = \frac{1000 f_q Q_{i,adwf}/24}{0.8 V_0 \exp(-n X_t)} \qquad \text{m}^2$$

Equation 5-2, therefore, indicates an increase in reactor TSS concentration results in an increase in SST area, which in effect increases the SST cost.

5.1.3 Cost of Aeration

The aeration cost is a significant contributor to the total cost of the system. Aeration OPEX is mainly made up of the energy cost to sustain the aeration system. The α -factor has the most significant impact on aeration efficiency when compared with the effect of the reduction factor for the saturation dissolved oxygen concentration in activated sludge relative to pure tap water (β) and temperature (T) as shown in Equation 5-3

Equation 5-3: Determination of actual oxygen requirement at site conditions

$$AOR_{site} = Q_{air,std}.C_{02,std}.SOTE_{std}.\{\}$$

and $\{\} = \alpha\theta^{(T-20)}\left[\left(\frac{P_{site}-p_{site}}{P_{std}-p_{std}}\right)\left(\frac{P_{site}+73.53hf-p_{site}}{P_{site}-p_{site}}\right)\left(\frac{51.6}{31.6+T}\right)\beta C_{Sstd} - C_L\right]/C_{Sstd}$

Where:

$Q_{air,std}$	=	Air flow at STP (m^3/h)
$C_{02,std}$	=	Oxygen content of air at STP (kgO/m^3)
$SOTE_{std}$	=	Oxygen transfer efficiency at STP (%)
α	=	(K_{La} of mixed liquor)/(K_{La} clean tap water)
$\theta^{(T-20)}$	=	effect of temperature on K_{La}
		$(1.024)^{T-20}$ for diffused aeration systems
P_{site}	=	atmospheric pressure on site (mm Hg)
p_{site}	=	saturated vapour pressure at site, for site temperature (mm Hg)
P_{std}	=	standard pressure
		=760 mmHg
p_{std}	=	standard saturated vapour pressure
		=17.51 mmHg @ 20ºC
73.53	=	conversion factor for m water to mm Hg
h	=	submersion depth for diffusers (m)
F	=	fraction of submerged depth (from surface) at which pressure corresponds to the average saturation concentration. Varies over the range 0.22 to 0.33 for a range of diffuser depths from 3.5 to 6.5m, accept = 0.325
T	=	temperature in C°
β	=	effect of impurities on C_s
		(C_s mixed liquor)/(C_s of clean tap water)
C_{Sstd}	=	saturated concentration of DO under standard conditions (mgO/l)

Several studies have observed an exponential relationship between the α-factor and MLSS concentration. (Figure 4-1) (Germain and Stephenson, 2014)

The following equation for determining α based on MLSS concentration is reported by Judd (2011)

$$\alpha = e^{-0.084.MLSS}$$

As can be seen from the equation above, the alpha factor decreases exponentially with increasing MLSS concentration. This results in a decrease in oxygen transfer with increasing MLSS concentration. As a result, a higher MLSS concentration will increase the aeration system power requirement, thus increasing the CAPEX and OPEX.

5.1.4 Total cost of AS-SST system

For the purpose of determining the optimum MLSS concentration, the total cost of the system is the sum of the reactor, SST and aeration cost. The MLSS concentration which results in the lowest total cost will be deemed the optimum MLSS concentration and is used further in the design such as determining biological reactor volume. (

Figure **5-2**) indicates the relationship between the costs of the various components (Reactor, SST and aeration) and MLSS concentration.

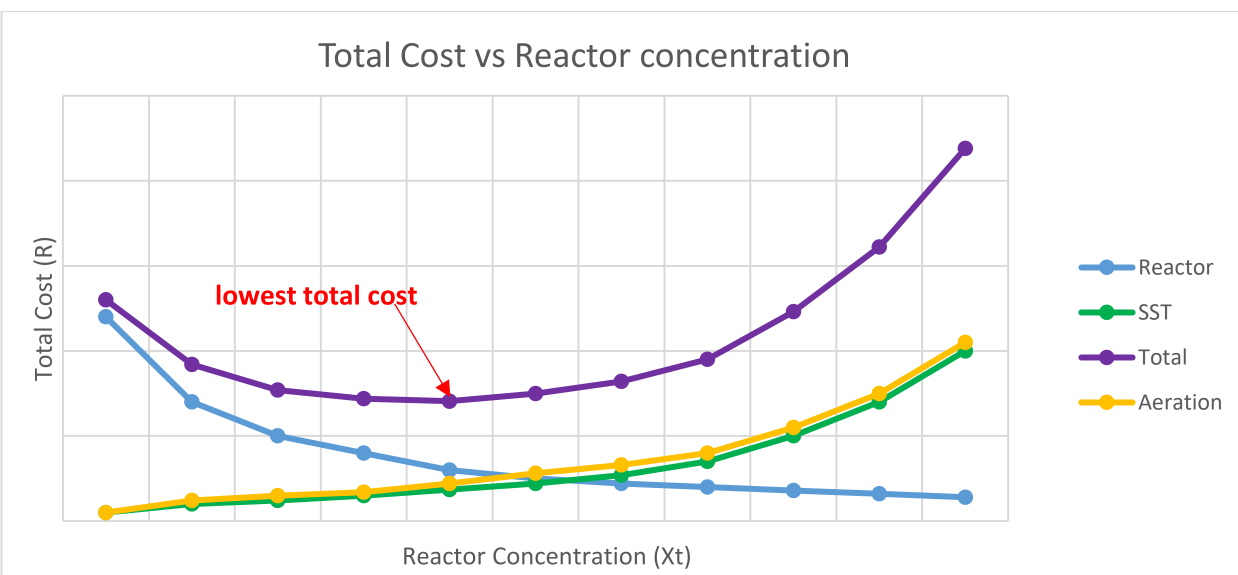

***Figure 5-2**: Total cost plotted against Reactor concentration*

6 Optimizing the UCT system

In a UCT configuration, nitrogen is removed in the same way as in an MLE configuration and equations based on similar denitrification principles apply (Henze *et al*., 2008, Ekama, 2011) except no influent readily biodegradable organics are available because these have been taken up by the phosphorus accumulating organisms (PAO), and the second slower rate of denitrification on slowly biodegradable organics is increased to 0.255 mgNO3-N/(mgOHOVSS.d) at 20°C. COD removal takes place similarly as in an MLE system as described in Section 5. However, as the UCT system includes also of an anaerobic zone in addition to an anoxic zone, the maximum unaerated fraction is shared between these 2 zones (anoxic and anaerobic). The VSS also now includes PAOs and their endogenous residue for the removal of P. The ISS now also includes the polyphosphate content of the PAOs, which significantly increase the ISS of the MLSS and decreases the VSS/TSS ratio of the activated sludge to below that of the MLE system.

A typical schematic of the UCT system is shown in Figure 6-1

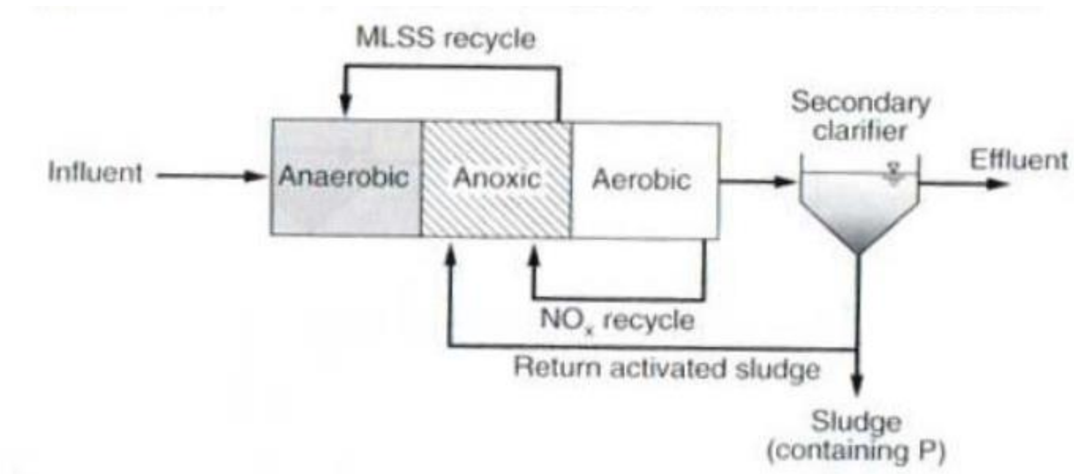

Figure 6-1 *– Typical process schematic for a UCT system*

6.1 Optimizing the Reactor MLSS concentration for a UCT system

The optimum TSS concentration for a UCT system which results in the lowest system cost is determined the same way as for an MLE system is described in Section 5.1, except that the MLSS now increases from 5 in the MLE to 8 constituents in the UCT system: (1) the OHO biomass (2) the OHO endogenous residue (3) the inert unbiodegradable organics from the influent (UPO) (4) the OHO ISS and (5) the ISS from the influent for the MLE plus (6) the PAO biomass, (7) the PAO endogenous residue and (8) the PAO ISS and polyphosphate (Ekama and Wentzel, 2004)

SECTION B: MBR AND CAS DESIGN APPLICATION

7 Input design parameters

The main objective of this study is to compare the CAPEX and OPEX of the CAS system with MBR technology. The following process scenarios were investigated and compared:

Option 1 - CAS in an MLE configuration

Option 2 – CAS in a UCT configuration

Option 3 - iMBR using FS membranes in an MLE configuration

Option 4 - iMBR using FS membranes in a UCT configuration

Option 5 - iMBR using HF membranes in an MLE configuration

Option 6 - iMBR using HF membranes in a UCT configuration

The flow rate (Q) used in all of the above scenarios was **20 Ml/d ADWF**

For the purpose of this study, settled wastewater was not included, due to the fact that the project's main objective was to compare CAS with MBR from an economical perspective. Limiting the study to raw wastewater only confined the scope of work to focus on the project's specific objectives.

7.1 Influent wastewater characteristics and system parameters

The influent characteristics used was based on typical domestic wastewater and are indicated in Table 7-1 and Table 7-2

Table 7-1: Influent wastewater characteristics

	COD	**VFA**	**TKN**	**FSA**	**TP**	**OP**	**TSS**	**ISS**
	mg/l	mg/l	mgN/l	mgN/l	mgP/l	mgO/l	mg/l	mg/l
Unfiltered (Raw)	800	0	60		18	0	400	50
0.45um membrane filt.			20	60	13.6	11.46		

Tests on AS Effluents	**COD**	**VFA**	**TKN**	**FSA**	**TP**	**OP**	**TSS**	**ISS**
	mg/l	mg/l	mgN/l	mgN/l	mgP/l	mgO/l	mg/l	mg/l
Raw WW AS system	45		1.36	0.28	12.04	12.04	0	0

Table 7-2: Additional design information

$f_{s'up}$	0.15*	As per WW characterization done
$f_{s'us}$	0.076*	As per WW characterization done
f_{cv}	1.481	mgCOD/mgVSS
f_c	0.518	mgC/mgVSS
f_n	0.1	mgN/mgVSS
f_p	0.025	mgP/mgVSS
T (min)	14	° C
T (max)	22	° C
f_i (PWWF factor)	2.5	

*These are the only characteristics that change between raw and settled wastewater. Their effect is to decrease the sludge production and increase the oxygen demand per kgCOD for settled wastewater.

The growth kinetic values used in this study are listed in Table 7-3. The values corrected for a temperature of **14°C** are also shown.

Table 7-3: Kinetic growth constants and their temperature correction coefficients used in the investigation

	MLE	**MLE (corrected for 14°C)**	**UCT**	**UCT (corrected for 14°C)**	θ
μ_{AMT}	0.5	0.249	0.5	0.249	1.123
b_{AT}	0.04	0.034	0.04	0.034	1.029
b_H	0.24	0.202	0.24	0.202	1.029
K_{nT}	1	0.5	1	0.5	1.123
Y_h	0.45	0.45	0.45	0.45	-
Y_G	-	-	0.45	0.45	-
f_{EH}	0.2	0.2	0.2	0.2	-
f_{EG}	-	1	0.25	0.25	-
f_{cv}	1.481	1.481	1.481	1.481	-
f_{XBGPBM}	-	-	0.03	0.03	-
f_{XBGP}	-	-	0.355	0.355	-
f_{iOHO}	-	-	0.15	0.15	-
f_{iPAOBM}	-	-	0.15	0.15	-

The parameters used for the aeration design are listed in Table 7-4. Here 22°C is used because this requires the highest oxygen demand.

Table 7-4: Aeration design parameters

	Unit	MLE	UCT
Altitude/elevation	m	1000	1000
T	C	22	22
DO	mgO/l	2	2
Water depth	m	3.75	3.75
$SOTE_{STD}$	%/m	6.72	6.72
P_{std}	mmHg	760	760
*P_{site}	mmHg	676	676
p_{std}	mmHg	17.51	17.51
**p_{site}	mmHg	18.79	18.79
α		Calculated	Calculated
β		0.9	0.9
***d_L		0.33	0.33

*$P_{site} = 754.4 - 0.07807.Alt$

**$p_{site} = 10^{(0.6979+0.02618\, T^{o}C)}$

***Peak/Average oxygen demand to Peak COD/Average Load damping factor (see Musvoto et al, 1992)

7.2 Sludge characteristics

A membrane's ability to separate solids from the liquid is insensitive to sludge settleability and filamentous bulking. This is a major advantage over SSTs, where design size is sensitive to sludge settleability. For the purpose of this study, it was therefore decided to include various DSVI values in the design of the SST, which deems a fairer comparison. The sludge settleability characteristics used in this investigation are shown in Table 7-5:

Table 7-5: Sludge characteristics

Parameter	Value	Unit	Equations
DSVI	100; 150 and 200	ml/g	
SSVI	67; 100 and 134	ml/g	SSVI =0.67 x DSVI
n	0.343; 0.435 and 0.526	l/gTSS or m^3/kgTSS	${V_o}/{n} = 67.9exp^{-0.016SSVI}$
V_0	7.98; 5.91 and 4.19	m/h	$n = 0.88 - 0.393log\left({V_o}/{n}\right)$

8 Determining optimum MLSS for CAS

8.1 Cost optimization

As mentioned in section 5.1, an optimum MLSS was determined which will result in the lowest total cost (CAPEX + OPEX). In order to achieve this, the total cost which consists of the Reactor, SST and Aeration costs were determined for a range of MLSS concentrations (1 – 15g/l). Cost functions were determined based on historical costing data.

8.1.1 Reactor CAPEX

A cost function was derived to calculate reactor CAPEX based on the reactor volume. The cost function was derived by plotting a range of reactor costs against their respective volumes and then deriving an equation from the graph. A total of 4 data sets were used to establish a graph from where the best fit trendline was established. An equation (cost function) from the trend line was then established. The costing information for the reactors used in establishing the cost function was taken from past wastewater projects between 2010-2018 curtesy of iX Engineers, who were the consulting engineers on the projects. These prices were then escalated to 2019 at 6% pa. The following cost function was derived:

Equation 8-1: Cost function for Reactor cost

Cost (R) = $26.35V^{0.663}$ x 1000, where V is reactor volume in m^3.

Reactor CAPEX includes the following:

- Rectangular, steel-reinforced concrete structure with a water depth of 4.50m and freeboard of 500mm (total wall height 5.00m). Structure semi-submerged below ground level.
- The structure includes steel-reinforced concrete walkways and platforms to support mixers. (No platforms in the aerobic zone)

Reactor costing was then determined using the above equation for a range of MLSS concentrations between 0 – 15000 mg/l. The mass of TSS in the Reactor was determined based on the parameters stated in **Section 7.1** and was based on the UCT steady state model equations in chapters 5 and 7 of Wentzel et al. (2008). The detailed calculations for this are not indicated. A summary of the results from the steady state equations to determine the mass in the reactor is shown in Table 8-1:

Table 8-1: Calculation results for Mass in Reactor used in cost minimization

	Unit	MLE	UCT
Q_{ADWF}	Ml/d	20	20
R_s	d	14	16
MX_t	kgTSS	71643.36	99167.49

8.1.2 SST cost function

A similar approach as described for the reactor cost function was followed for deriving a cost function equation to determine SST cost:

Equation 8-2: Cost function for SST cost

Cost (R) = 137.67Ø$^{0.957}$ x 1000, where Ø is SST diameter in m.

SST costing was then determined using the above equation for a range of MLSS concentrations between 0 – 15000 mg/l. The required SST area was calculated using the 1D Flux theory with the Sludge characteristics as stated in Section 7.2. The detailed calculations for these are not shown. A summary of the parameters used in the SST cost optimization exercise is shown in Table 8-2:

Table 8-2: SST design information for cost optimization

	Unit	MLE	UCT
Q_{ADWF}	Ml/d	20	20
f_q		2.5	2.5
SF*		1.25	1.25

*Factor of safety

8.1.3 Aeration CAPEX cost function

The method for determining a cost function to determine Aeration CAPEX was the same as for the reactor and SST cost function.

Equation 8-3: Cost function for blower cost

Cost (R) = 869.24(P)$^{0.34}$ x 1000, where P is blower *power in kW.

*Please note that the power is the installed power.

Aeration capital costs were based on bubble aeration which consists of the blowers. For the purpose of this investigation, the cost of the diffusers and piping were excluded due to the cost of these items being sensitive to the specific installation scenario.

As mentioned previously, the alpha factor is dependent on the MLSS concentration. An increase in MLSS concentration results in a decrease in the alpha value, thus resulting in a decrease in oxygen transfer efficiency.

Corresponding alpha values was plotted for a range of MLSS concentrations using Equation 8-4. Corresponding airflow rates to meet the system oxygen requirement (FO_t kgO/d) was then determined for each alpha value using Equation 8-5. The required blower size was then calculated Equation 8-6 from (*Metcalf & Eddy,* 2004) and then priced using Equation 8-3.

Equation 8-4: Alpha value based on MLSS concentration

$$\alpha = e^{-0.084.MLSS}$$

Equation 8-5: Required air flow rate

$$Q_{air,std} = \frac{AOR_{site}}{C_{02,20C}.SOTE_{STD}.\{\}}$$; {} as per Equation 5-3

Equation 8-6: Power equation

$$P_w = \frac{wRT_1}{29.7ne}\left[\left(\frac{p_2}{p_1}\right)^{0.283} - 1\right]$$

Where:

P_w	=	Power requirement of the blower (kW)
w	=	weight of air flow (kg/s)
R	=	Universal gas constant for air
		8.314kJ/k mol K
T_1		Absolute temperature (K)
p_1		Absolute pressure at inlet (atm)
p_2		Absolute pressure at outlet (atm)
n		(k-1)/k = 0.283 for air
29.7		Constant for SI units' conversion
e		Efficiency (usual range for compressors is 0.7 – 0.9) 0.8 used for this investigation

8.1.4 Determining Aeration OPEX

The most significant contributor to the OPEX of a wastewater treatment plant is energy (45%), followed by replacement parts, maintenance, and chemicals (20%), disposal (15%) personnel salaries (11%), and property (9%). The OPEX for this investigation consisted only of the electricity cost of the aeration system calculated over a project life cycle of 10 years. The electricity cost was based on Eskom's tariff structure for industrial usage which varies based on the time of day, day of the week and month of the year (increased rates in winter) as indicated in Figure 8-1

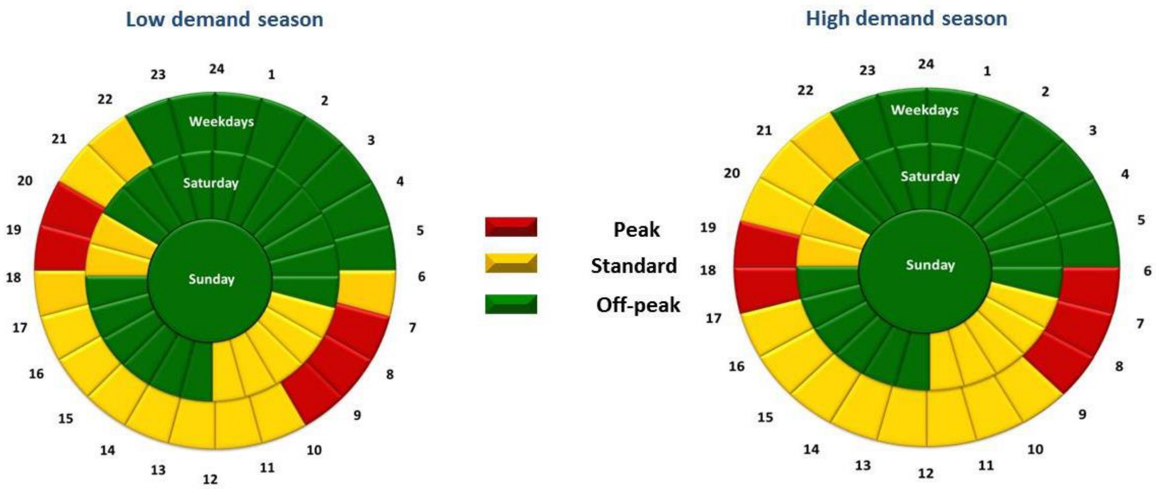

Figure 8-1 – Low and high demand seasons time of use periods

The power demand of the aeration system varies over time based on the required oxygen demand. The oxygen demand was determined based on the diurnal organic loading to the plant.

The aeration system parameters used are stated in Section 7.1

The following equations (one for each process configuration) was derived for determining OPEX based on MLSS concentration over a period of 10 years. The OPEX equations derived below are specific to this investigation's design inputs.

Equation 8-7: OPEX for CAS-MLE system

$$OPEX = 141.66MLSS^2 + 769.41(MLSS) + 19403\ x\ 1000 \text{ where MLSS is in g/l}$$

Equation 8-8: OPEX for CAS-UCT system

$$OPEX = 254.58MLSS^2 + 151.1(MLSS) + 37541\ x\ 1000 \text{ where MLSS is in g/l}$$

Equation 8-9: OPEX for FS MBR-MLE system

$$OPEX = 1226.2MLSS^2 - 26554(MLSS) + 184146\ x\ 1000 \text{ where MLSS is in g/l}$$

Equation 8-10: OPEX for FS MBR-UCT system

$$OPEX = 1215MLSS^2 - 20811(MLSS) + 165969\ x\ 1000 \text{ where MLSS is in g/l}$$

Equation 8-11: OPEX for HF MBR-MLE system

$$OPEX = 1819MLSS^2 - 12640(MLSS) + 60672\ x\ 1000 \text{ where MLSS is in g/l}$$

Equation 8-12: OPEX for HF MBR-UCT system

$$OPEX = 1819MLSS^2 - 12640(MLSS) + 60672\ x\ 1000 \text{ where MLSS is in g/l}$$

8.2 Cost optimization results

A summary of the system design used in the Aeration cost optimization exercise is shown in Table 8-3.

Table 8-3: Calculation results for aeration used in cost minimization exercise

	Unit	MLE	UCT
*a_L	kg/h	0.79	0.79
**d_L		0.33	0.33
$FO_{t,\ peak}$	kgO/d	6520	13482
OUR	kgO/h	271.68	561.75

*Peak to Average Load ratio

**Peak/Average oxygen demand to Peak COD/Average Load damping factor

The total cost which compromises reactor cost, SST cost and aeration Cost (CAPEX and OPEX) was then determined for each MLSS concentration. The results from the investigation were plotted and are shown below in Figure 8-2

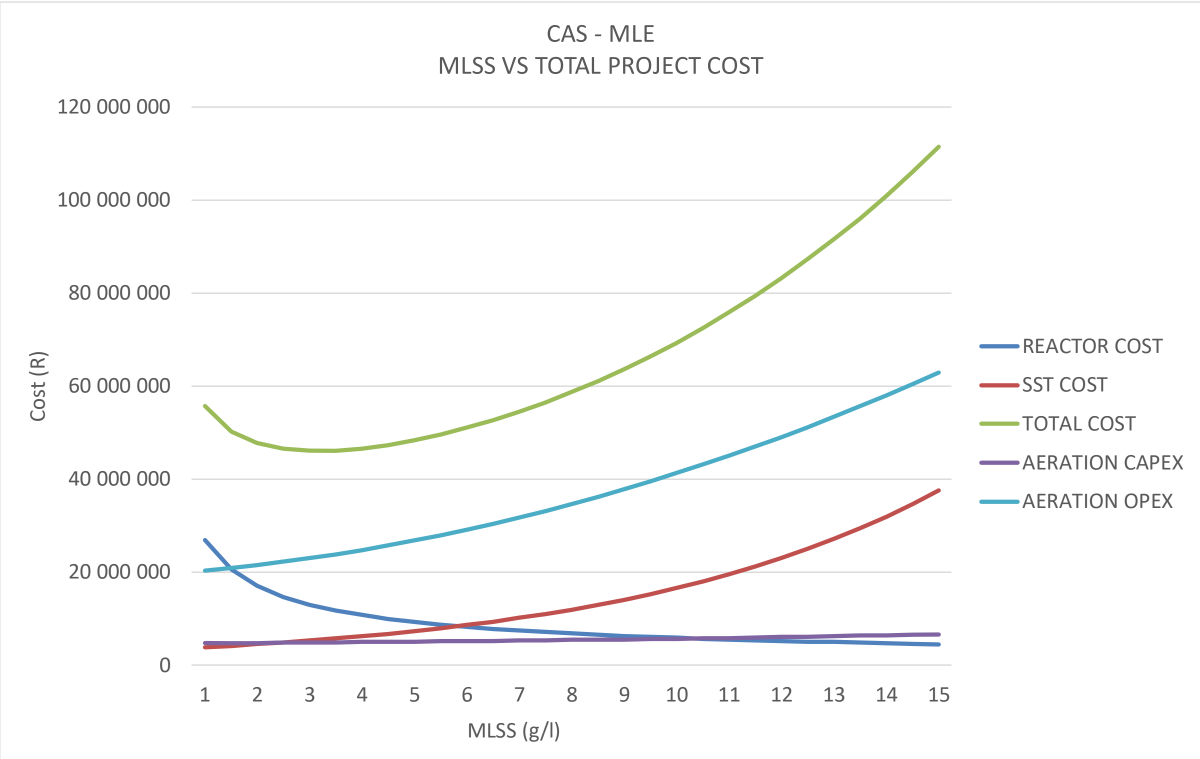

***Figure 8-2** – MLSS effect on Reactor, SST, Aeration CAPEX, Aeration OPEX and Total Cost in CAS MLE system*

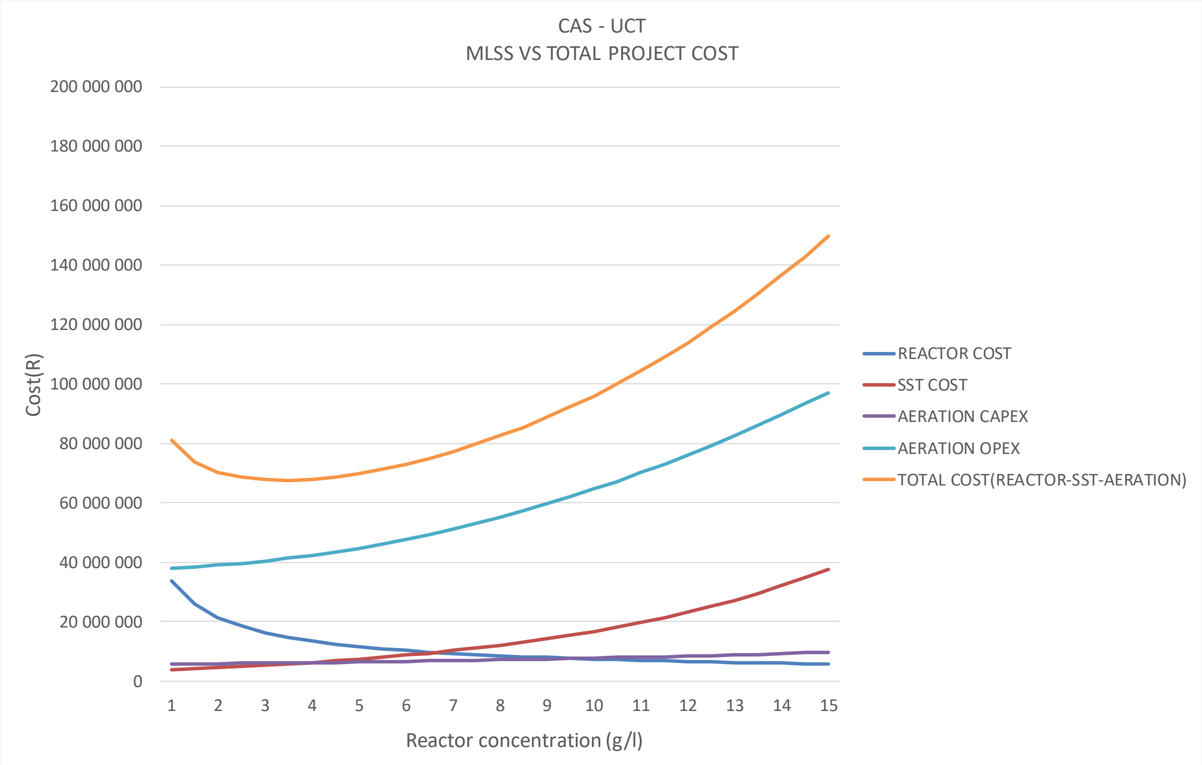

***Figure 8-3** – MLSS effect on Reactor, SST, Aeration CAPEX, Aeration OPEX and Total Cost in CAS UCT system*

The OPEX cost of the aeration system has a massive impact on the total project cost when compared to the cost of the reactor and SST. Aeration costs increase exponentially with increasing MLSS.

The most optimum reactor MLSS concentration was then determined where the total cost was the lowest. A summary of the results from this cost minimization is shown in Table 8-4:

Table 8-4: Summary of results from cost optimization

	Unit	MLE	UCT	
No. of reactor modules and vol/module	m³	2 x 11 516 m³	2 x 13 778 m³	V of 1000 - 16 000 m³
No. of SSTs and Ø	m	1 x 34 m	1 x 37.2 m	Ø of 10 - 40 m
Blower size	kW	137	295	
Reactor (cost)	(R/Million)	26	29.2	Equation 8-1
SST (cost)	(R/Million)	5.3	5.7	Equation 8-2
Aeration (CAPEX)	(R/Million)	4.8	6	Equation 8-3
Aeration (OPEX)	(R/Million)	22.9	41.2	Equation 8-7 Equation 8-8
Total cost	(R/Million)	59	82	
Optimum MLSS conc.	mg/l	3000	3 500	

The optimum MLSS concentration determined from the cost optimization model was 3000 mg/l for the MLE system and 3500 mg/l for the UCT system. This results in 2 Reactors of 11 516 m³/reactor for the MLE system and 2 x reactors of 13778 m³/reactor. However, this is not practical for construction due to the large land requirement. The low MLSS concentration is due to the very high impact of the aeration OPEX system cost on the total cost model. An increase in MLSS results in an exponential decrease in the α value which exponentially increases the aeration system capital and operational cost in the cost model. This can possibly be counterbalanced if land costs were included in the model, but this was not done in this investigation. Including the aeration system OPEX in the minimization calculation results in an MLSS which will result in the theoretical lowest total cost, though one must take into account the practicality of the design. It was therefore decided to exclude the aeration OPEX cost from the cost minimization calculation due to the sensitivity of the alpha value with changing MLSS. The results of the revised calculation which includes only the **Reactor-SST-Aeration CAPEX (Scenario 1)** are shown in Figure 8-4 and Figure 8-5. The results for the **Reactor-SST CAPEX (Scenario 2)** are indicated in Figure 8-6 and Figure 8-7.

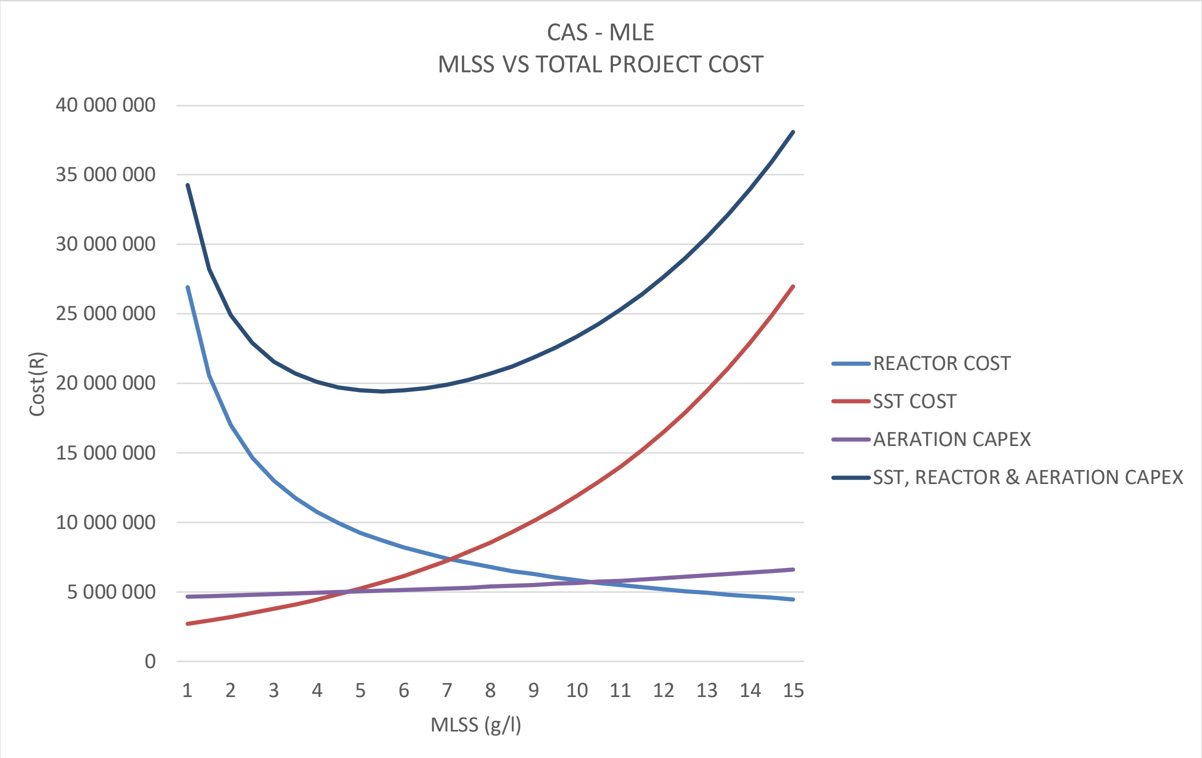

Figure 8-4 –** Scenario 1: Reactor MLSS concentration effect on Reactor, SST, Aeration CAPEX and Total cost for **MLE

Figure 8-5 *– Scenario 1: Reactor MLSS concentration effect on Reactor, SST, Aeration CAPEX and Total cost for UCT*

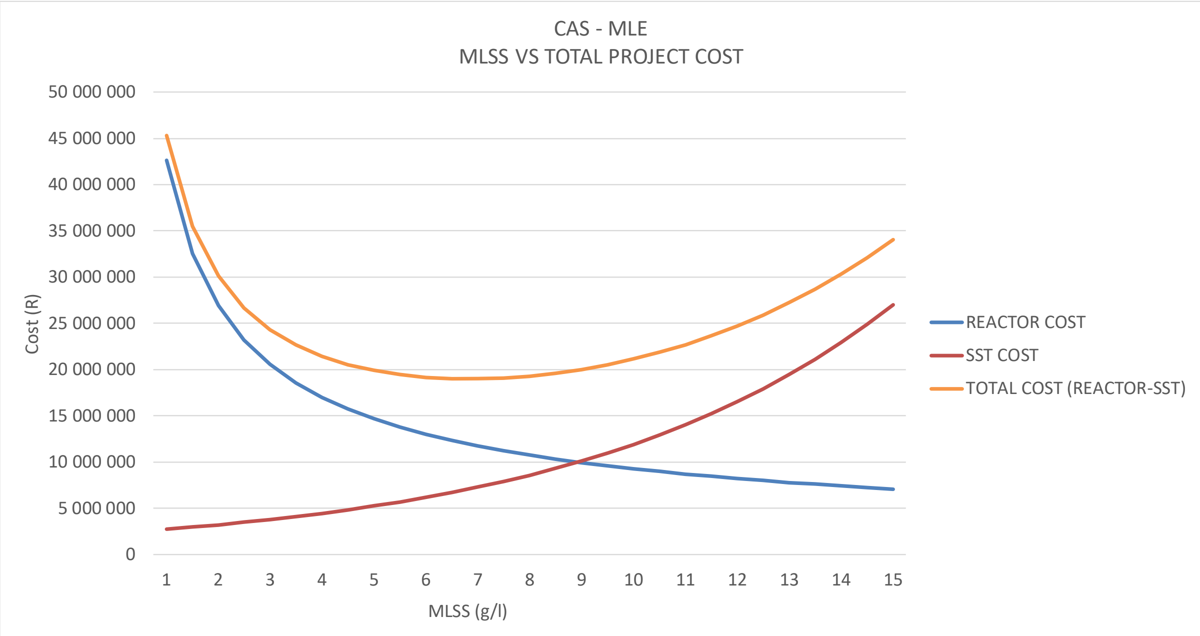

Figure 8-6 *– Scenario 2: Reactor MLSS concentration effect on Reactor, SST and Total cost for MLE*

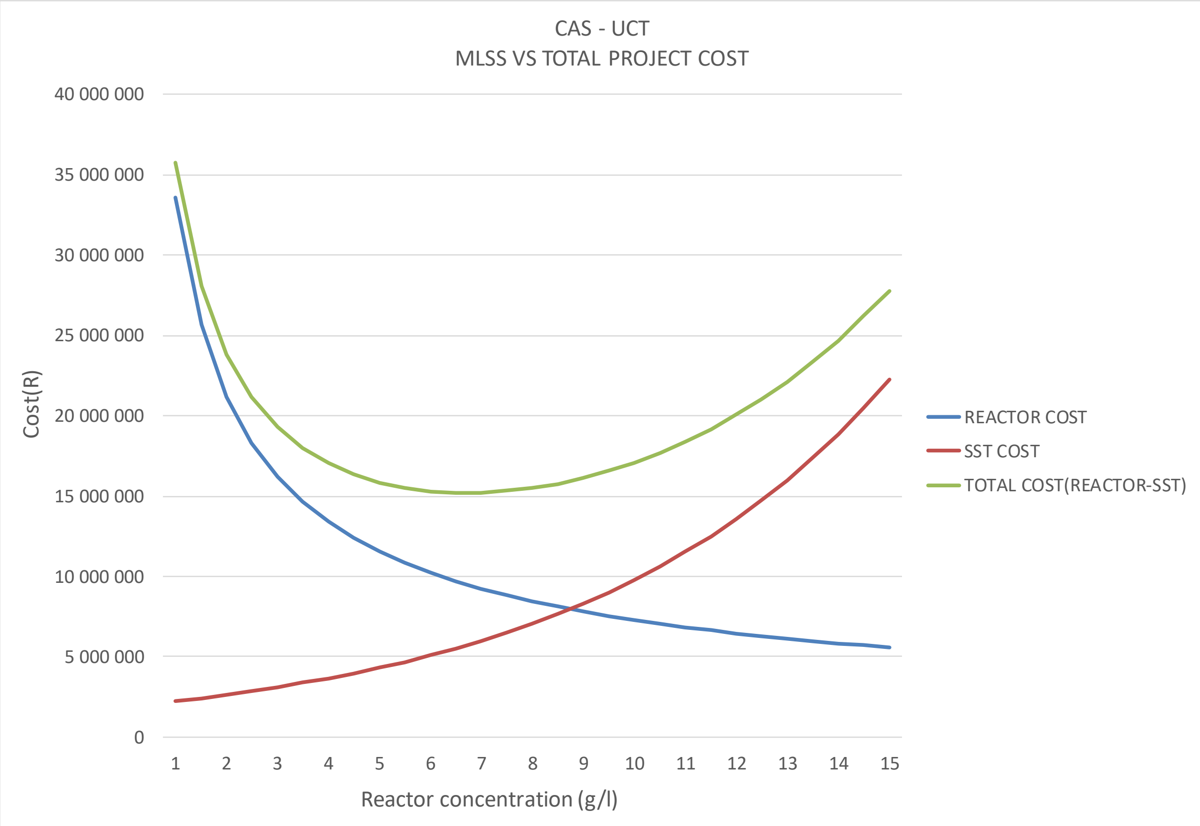

***Figure 8-7** – Scenario 2: Reactor MLSS concentration effect on Reactor, SST and Total cost for UCT*

A summary of the results from the revised cost optimization is shown in Table 8-5 :

Table 8-5: Summary of revised results from the cost optimization exercise

		SCENARIO 1 (REACTOR+SST+AERATION CAPEX)		**SCENARIO 2 (REACTOR+SST)**		
	Unit	**MLE**	**UCT**	**MLE**	**UCT**	**Parameter Range of validity**
Reactor volume per module	m³	6 282	8037	5758	7419	V of 1000 – 16 000 m³
No. of modules		2	2	2	2	
SST Ø	m	37	40.3	40.4	35.6	Ø of 10 – 40 m
No. of SSTs		2	2	2	3	
Aeration(blower) power	kW	169	364	n/a	n/a	
Reactor cost	(R/Million)	17.4	20.5	16.4	19.4	
SST (cost)	(R/Million)	11.3	12.3	12.4	16.5	
Aeration CAPEX cst	(R/Million)	5.1	6.6	n/a	n/a	
Total cost	(R/Million)	33.8	39	28.7	35.9	
Optimum MLSS conc.	g/l	**5.5**	**6**	**6**	**6.5**	

By excluding the Aeration OPEX from the cost minimization exercise, the optimum MLSS concentration increased from 3 000 mg/l and 3 500 mg/l in the MLE and UCT systems respectively, to 5 500 mg/l and 6 000 mg/l respectively in the 2 systems. By excluding the Aeration CAPEX, the optimum MLSS concentration further increased to 6 000 mg/l and 6 500 mg/l in the 2 systems respectively.

Based on the results from the cost minimization exercise, an MLSS concentration of **5.5 g/l** for the MLE system and **6 g/l** in the UCT system was used in the design of the CAS system which is further described in Section 11.

9 Determining optimum MLSS for MBR

A similar approach was followed to determine an optimum MLSS for an MBR system as was done for CAS, with the following differences: 1) The SST cost was replaced with the cost for the membranes 2) The aeration costing was done differently to CAS, due to the membrane aeration system contributing (partially or fully) to the biological aeration requirements.

It must be noted that MBRs require ultra-fine screening (<2mm) upstream of the bioreactor, this is not required for CAS systems. The cost for screening was not included in this study as this cost is insignificant compared to the membranes and aeration.

The permeate can either pass through the membranes via pumping or static hydraulic head. Pumping costs (i.e. electricity consumed by the pumps) in some membrane configurations are significant. For this study, permeate is allowed to pass through the membranes via the static hydraulic head. Hence, no pumping costs were included in this study.

It must be noted that the cost analysis tool used in this study was developed specifically for this research topic, and was confined to the research scope. The costing tool developed is not generic and other broader factors are also to be considered for engineering decisions.

9.1 Cost of the membranes

9.1.1 Kubota FS and Zeeweed HF system

As part of the cost optimization calculation, membranes were sized and priced for a range of MLSS concentrations. The following procedure was followed as shown in Figure 9-1

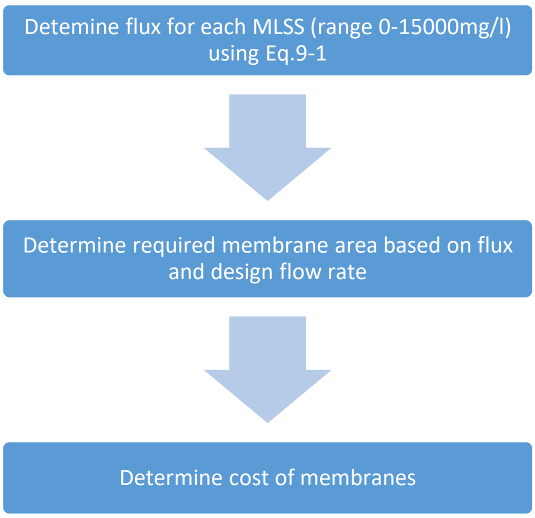

Figure 9-1: *Schematic methodology followed for membrane cost optimization exercise*

A pilot study was conducted by Yigit et al., 2008 where flux measurements were recorded for various MLSS concentrations. An equation was derived as part of the investigation relating Flux (LMH) to MLSS (mg/l):

Equation 9-1: Determination of critical flux based on MLSS concentration

$$q_{crit} = -0.0029MLSS + 46.254$$

Where:

q_{crit} = critical flux (LMH)

MLSS = Mixed liquor suspended solids concentration (mg/l)

It must be noted that the relationship between flux and MLSS concentration is a complex one. The values provided in this investigation were based on literature that involved experiments done under controlled environments. The purpose of this investigation is thus not to provide a design guide, but to illustrate the trends in the relationships between different parameters (Flux, MLSS, Aeration efficiency, etc.)

Fluxes were calculated for a range of MLSS concentrations (0-15000 mg/l) using Equation 9-1. The required membrane area was then calculated for each corresponding flux using Equation 3-1 as described in Section 3.1.1. The resultant effect of MLSS concentration on flux and membrane area is indicated below in Figure 9-2

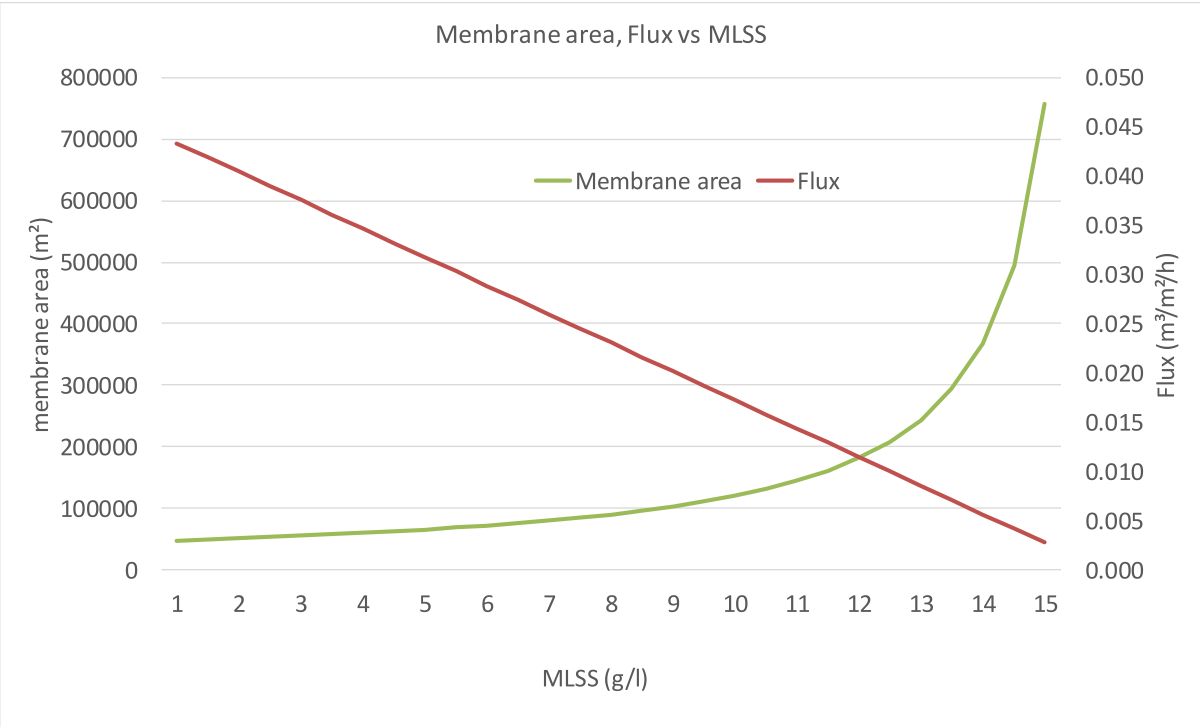

Figure 9-2: *Resulted effect of MLSS concentration on flux and membrane area for Kubota system at the design peak flow of 50Ml/d*

The membrane costing was based on the pricing data received from the membrane suppliers and are indicated in Table 9-1

Table 9-1: Membrane pricing data received from suppliers in 2019

	Kubota FS membrane	**Zeeweed HF membrane**
Q_{pwwf}	50 Ml/d	50 Ml/d
Membrane cost	R 900/m²	R 700/m²

The costing information was provided by Kubota and SUEZ Water Technologies in 2019.

9.2 Cost of the aeration system

9.2.1 Kubota FS and Zeewed HF system

The membrane aeration system contributes to the biological oxygen demand. For this investigation, it was assumed that all of the oxygen created by the membrane aeration was available to the biomass.

Aeration costs (CAPEX and OPEX) were determined for a range of MLSS concentrations as per the methodology described in Figure 9-3

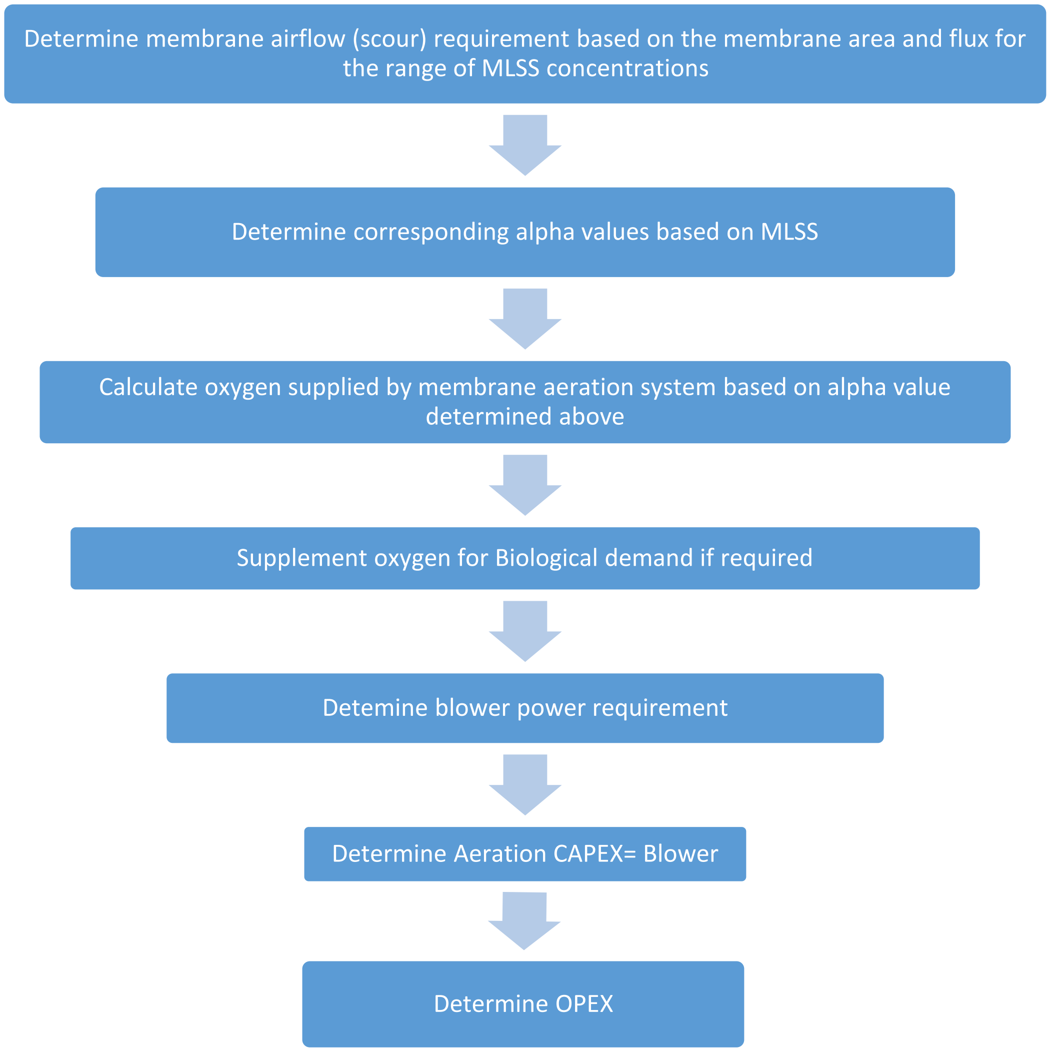

Figure 9-3: *Schematic methodology followed for aeration cost optimization exercise*

The membrane airflow requirement was determined as per the membrane manufacture guidelines for a range of MLSS concentrations. For the Kubota FS, an oxygen requirement of **8.6 $Nm^3/h.m^3$ aerobic zone volume or 0.76 $Nm^3/h.m^2$ membrane area** at STP was used (Ramphao *et al.*, 2013). For the Zeeweed HF membranes, the air scouring demand is **0.3 $Nm^3/h.m^2$ membrane area** as per Judd, 2011 and confirmed by SUEZ Water**.** The accuracy of these values for oxygen demand could be argued, however, the focus of the cost optimization was on the effect MLSS has on the Aeration CAPEX trend, and it will be seen later on in this report that the accuracy of the values used here for oxygen scour requirement have little effect on the optimal MLSS selection. The objective of this calculation is to merely select an MLSS which should result in the lowest total project cost. The detailed airflow calculations can be seen later in this report when each system is compared to one another.

The airflow required by the membranes was then determined for each corresponding membrane area. This airflow at STP was then transferred to the Actual Oxygen Rate at site conditions based on corresponding alpha values at MLSS concentrations to check if additional oxygen is required to meet the system biological demand. The power requirement of the blower was then determined using Equation 8-6. The effect of MLSS concentration on alpha and installed power for the Kubota MLE system is indicated below in Figure 9-6

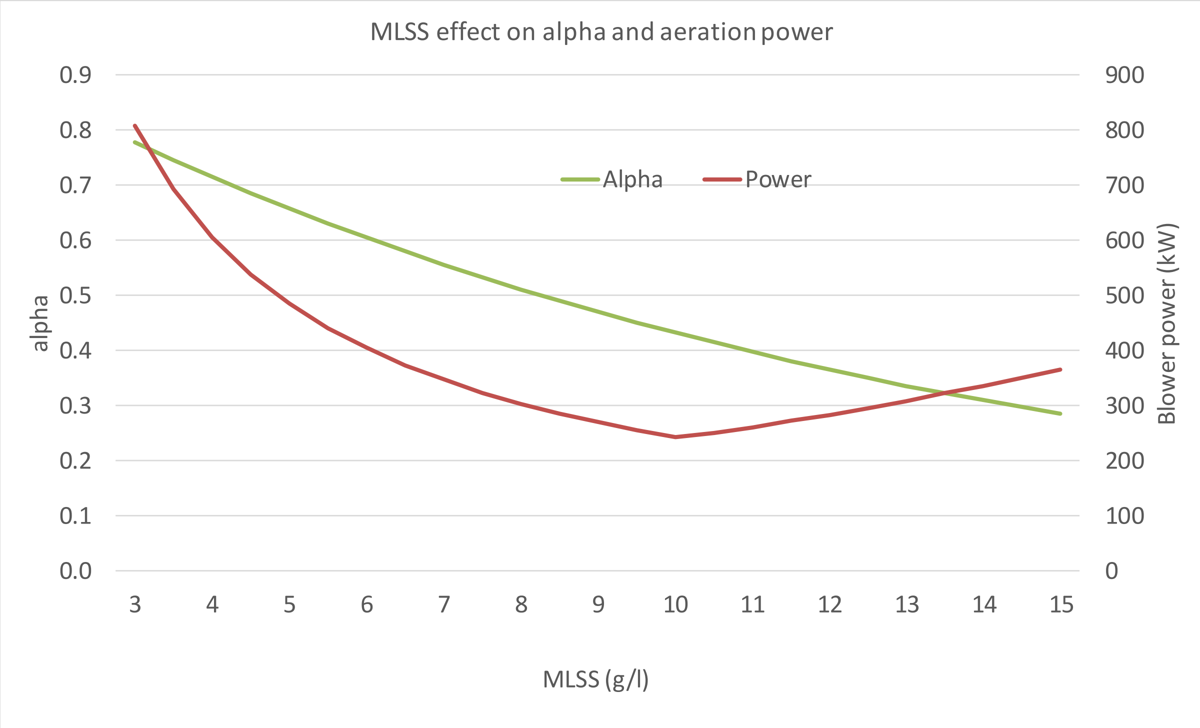

Figure 9-4: *MLSS effect on alpha and aeration power in Kubota MLE system*

It can be seen in the Kubota MLE system that an increase in MLSS concentration results in a decrease in power requirement which is due to a decrease in the aerobic volume. This is due to the membrane aeration demand being directly dependent on aerobic zone volume. A decrease in power is observed up until an MLSS concentration of ±10 g/l which at that point the membrane aeration does not meet the biological oxygen requirement and additional oxygen thus needs to be supplied.

9.3 Cost optimization results

The results from the cost optimization can be seen in **Figure 9-5** below:

***Figure 9-5** – MLSS concentration effect on Reactor, membrane and aeration cost in Kubota MLE system*

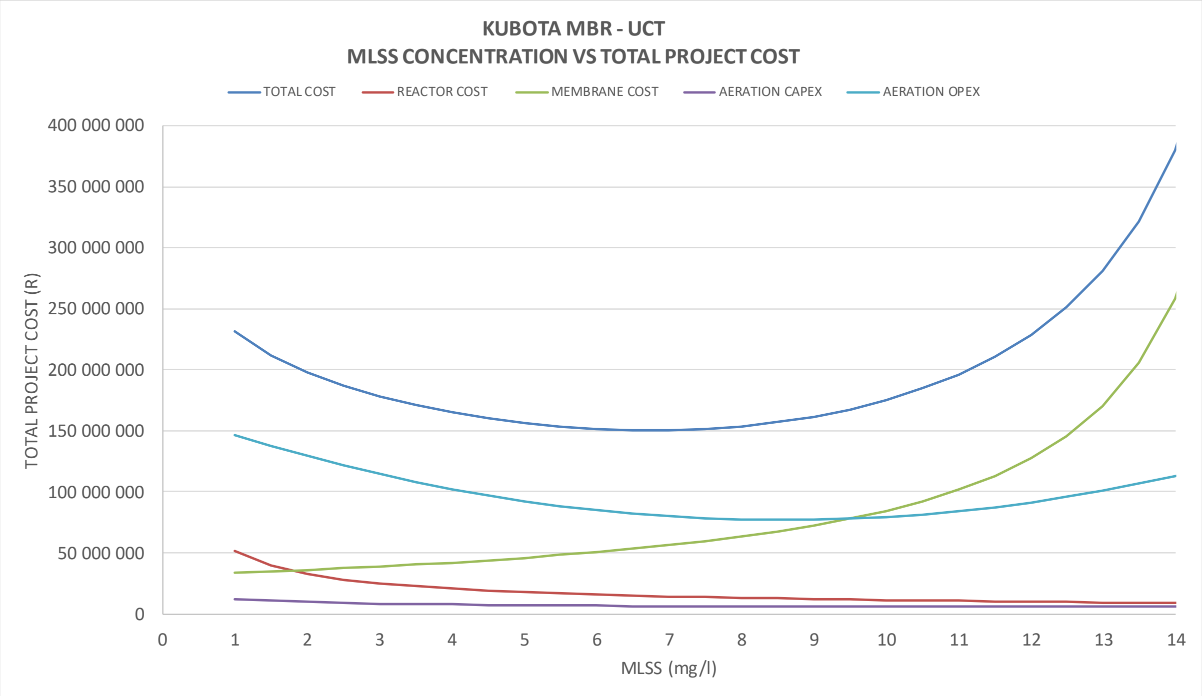

***Figure 9-6** – MLSS concentration effect on Reactor, membrane and aeration cost in Kubota UCT system*

The systems were priced using Equation 8-1, Equation 8-2 and Equation 8-3 which were also used in the CAS cost minimization calculation.

It can be seen from the results above that reactor cost decreases with increasing MLSS concentration. Membrane cost increase with MLSS concentration increase. The Aeration demand in a FS MBR system is dependent on the aerobic zone volume. The aeration costing thus decreases with increasing MLSS concentration up until ±10 g/l in the MLE system and ±6.5 g/l in the UCT system, an increase in aeration cost is then observed as MLSS concentration increases. This is due to the MBR aeration system needing to be supplemented in order to meet the biological oxygen demand. Additional oxygen is thus supplied.

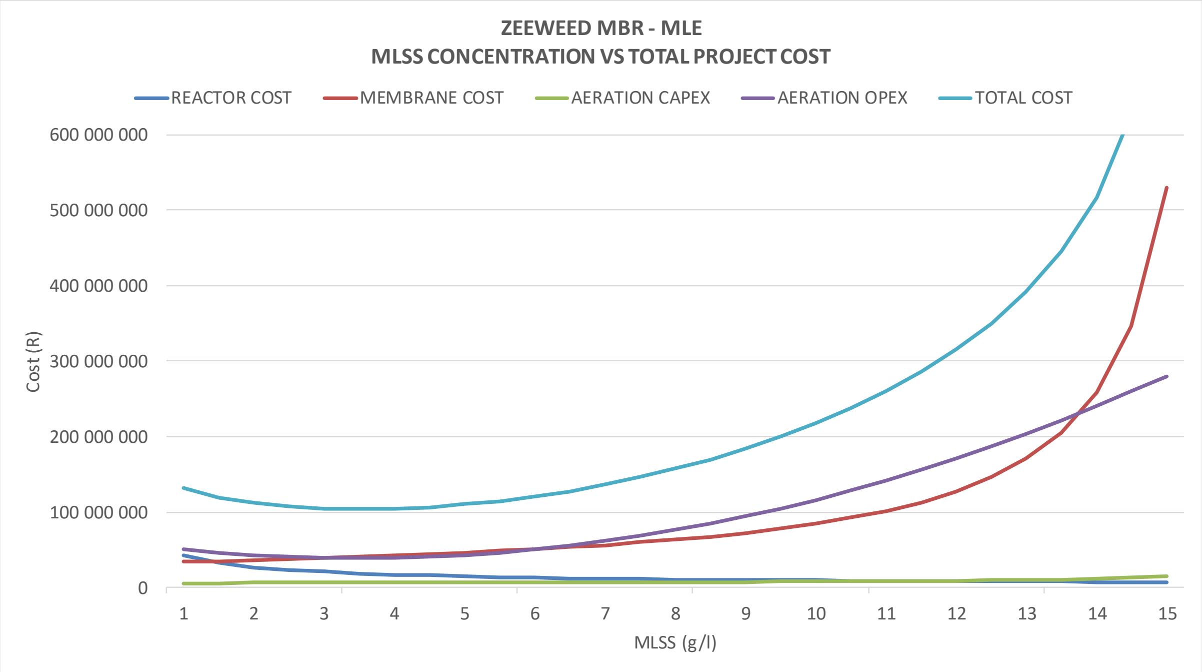

Figure 9-7 *– MLSS concentration effect on Reactor, membrane and aeration cost in Zeeweed MLE system*

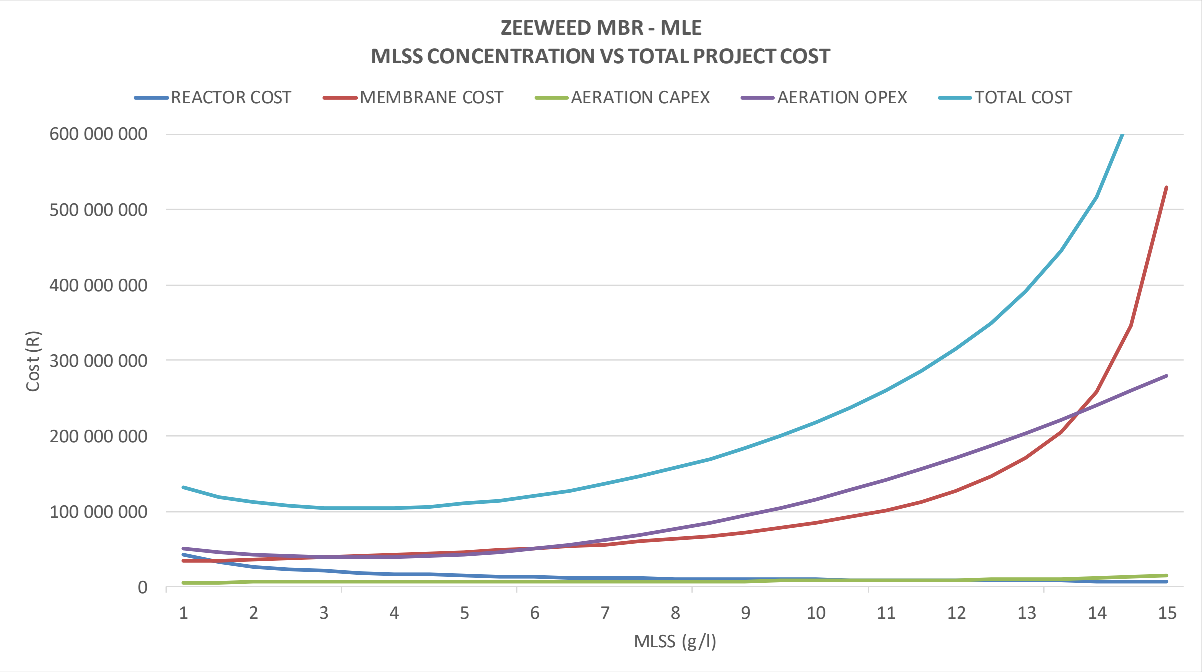

***Figure 9-8** – MLSS concentration effect on Reactor, membrane and aeration cost in Zeeweed UCT system*

From the results of the Zeeweed FS system cost minimization above it can be seen that an increase in MLSS concentration will result in an increase in membrane cost and aeration CAPEX and OPEX which out weights the saving from the decreasing reactor cost. The sudden increase in total cost in the Zeeweed MBR is due to the membrane aeration no longer supplying the full oxygen requirement for the biomass and additional fine bubble aeration is therefore required. For effective scour of the membranes, the MLSS concentration in the membrane tanks needs to be at least 10 000 mg/l, as recommended by SUEZ Water. Kubota membranes recommend an MLSS concentration of between 8000-12000 mg/l for effective membrane scour. The MLSS concentration which will result in the lowest total project cost over the life span will, therefore, be the **lowest MLSS concentration required to achieve effective scour of the membranes**. An MLSS concentration of 10 000 mg/l will, therefore, be used in the MBR design which is further described in Section 10

10 Results of MBR design

The wastewater characteristics used are indicated in Section 7.1. Wastewater characterization was done using block diagrams as shown in *Appendix A*.

10.1 Kubota FS MBR design

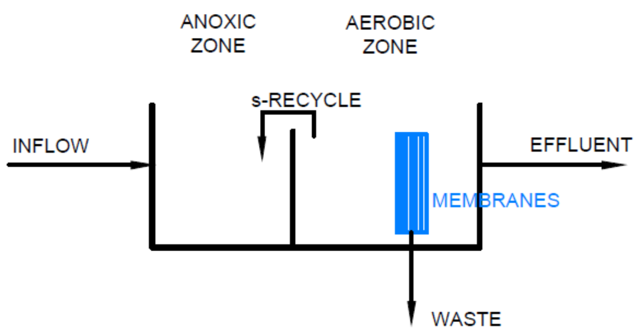

Figure 10-1 – *Typical Process flow schematic of FS MBR in MLE configuration*

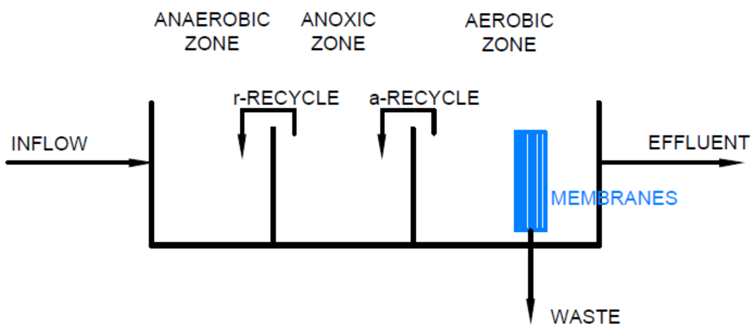

Figure 10-2 – *Typical Process flow schematic of FS MBR in UCT configuration*

The membranes considered in this study are the *Kubota FS* membranes by *Kubota*. The membrane design parameters used in this study have been confirmed by *Kubota (2019).*

Aerobic volume for membranes

As the membranes are located in the Aerobic zone, the Aerobic Zone volume required for the membranes ($V_{aer,m}$) was determined using Equation 10-1

Equation 10-1: Minimum aerobic zone volume for membranes

$$V_{aer,m} = \frac{L}{q_{max}} f_q . Q_{ADWF}$$

The calculation results for the minimum required aerobic zone volumes for the membranes are indicated in Table 10-1

Table 10-1: Calculation results for the minimum requirement for aerobic zone volume for membranes

	MLE	UCT	
Q_{ADWF}	20	20	Ml/d
f_q	2.5	2.5	
L	0.0875	0.0875	m³ reactor/m² membrane
q_{max}	1	1	m³ influent/m² membrane.d
$V_{aer,m}$	4.375	4.38	Ml

The values for the Membrane Packing density (L) and Flux (q_{max}) were provided by *Kubota*

Zone mass fractions and zone MLSS concentration

The system parameters selected for the determination of the zone mass fractions are shown in Table 10-2

Table 10-2: Zone mass fractions and recycle ratios used

	MLE	UCT
a	0	6
r	0	1
s	4	1
f_{mana}	0	0.13
f_{manx}	0.45	0.37
f_{maer}	0.55	0.5

In the MLE system, the mixed liquor recycle ratio (a) is set to zero because only one recycle (s) is required to return nitrate and sludge to the anoxic reactor.

With the recycle ratios chosen (Table 10-2) and Aerobic TSS concentration of **10 000 mg/l** as per the cost optimization results in Section 9. The sludge concentrations in the different zones (X_{tana}, X_{tanx}, and X_{taer}) can be calculated using the equations set out in (Maninna *et al.*, 2018) These are described previously in Section 3.2.3. The calculation results for the sludge zone concentrations are summarized in Table 10-3

Table 10-3: Calculation results for sludge zone concentrations

	MLE	UCT		
D	-	1.06		
E	1.11	-		
$\bar{X}_t$	8989	8097	mg/l	Based on X_{taer} of 10 000 as per cost optimization
X_{tana}	-	4286	mg/l	
X_{tanx}	8000	8571	mg/l	

Biological Aerobic zone volume and Sludge age selection

With the aerobic zone volume required for the membranes known, the aerobic zone for biological growth of the biomass can be determined. This was done with the UCT Steady State Model equations (Ekama, GA. Wentzel, 2008) This aerobic zone volume must be larger than the volume required for the membranes. With the Aerobic zone TSS concentration (X_{taer}) of 10 000 mg/l, Aerobic mass fraction (f_{maer}), wastewater characteristics and kinetic and stoichiometric values known, the sludge age was adjusted until the aerobic zone volume was greater/equal to the aerobic zone volume required for the membranes. The kinetic constants used are listed in 7.1. It has been proven that the use of the same kinetics constants as used in CAS, may be used in MBR. (du Toit *et al.*, 2010; Mannina *et al*., 2018). The results for the mass of TSS (MX_t) in the reactor and required Aerobic Volume are shown in Table 10-4

Table 10-4: Calculation results for Reactor mass and biological aerobic zone volume

	MLE	UCT	
Sludge Age (R_s)	18	16	days
MX_t	87776	95273	kgTSS
V_{aer}	4.83	4.76	Ml

The sludge ages for the MLE and UCT system were selected which will result in the Biological Aerobic zone being at least 4.375 ML, which is the volume required for the membranes.

Anoxic and/or Anaerobic zone volume

The Anaerobic and/or Anoxic zone volumes were then determined using the calculated zone TSS concentrations and calculated MX_t in the reactor. Along with the calculated Aerobic zone volume, the results for the Reactor volumes are indicated in Table 10-5:

Table 10-5: Calculation results for reactor zone volumes

	MLE	UCT	
V_{aer}	4.83	4.76	Ml
V_{ana}	-	2.89	Ml
V_{anx}	4.94	4.11	Ml
V_p	9.77	11.77	Ml

Ensuring Nitrification

With the system sludge age and aerobic mass fractions now known, the lowest maximum specific growth rate of the nitrifiers at 20°C to ensure nitrification was calculated based on the selected unaerated mass fractions of 0.45 and 0.5 for the MLE and UCT systems respectively.

Equation 10-2: Lowest maximum specific growth of the nitrifiers at 20°C

$$\mu_{nm20} = \frac{b_{n20}(\theta_b)^{(T-20)} + {}^{S_f}\!/_{R_s}}{f_{xt}(\theta_\mu)^{(T-20)}} \text{ /d}$$

The kinetic values used are indicated in Section 7.1.The calculated value for the lowest maximum specific growth rate is shown below in Table 10-6

Table 10-6: FS MBR MLE & UCT systems: Calculated results of lowest maximum specific growth rates

	MLE	UCT
μ_{nm20}	0.46	0.45

Since the values of 0.46/d and 0.45/d in the MLE and UCT systems respectively are lower than the design value of 0.5/d, the systems can be expected to nitrify.

Biological aeration demand

The biological oxygen demand for an MBR system is determined the same way as for a CAS system. The oxygen demand for the system is summarized in Table 10-7

Table 10-7: Oxygen demand of FS MBR- MLE and UCT systems

		MLE	UCT	
Carbonaceous oxygen demand OHO's	FO_{OHO}		8594	kgO/d
Carbonaceous oxygen demand POA's	FO_{POA}		560	kgO/d
Carbonaceous oxygen demand	FO_c	4237	9153	kgO/d
Nitrogenous oxygen demand	FO_n	2064	3396	kgO/d
Oxygen demand recovered by denitrification	FO_d	1130	1860	kgO/d
Total oxygen demand (FO_c + FO_n - FO_d)	FO_{tave}	5170	10690	kgO/d

The amplitude of the TOD influent (TOD_{peak}/TOD_{ave} - 1) is 0.8. The damping factor (a_L) used is 0.33. (Musvoto et al, 2002) The FO_{tpeak} (Peak total oxygen demand) was determined, with the results summarized in Table 10-8:

Table 10-8: Peak Oxygen demand of FS MBR- MLE and UCT systems

		MLE	UCT	
Peak total oxygen demand	FO_{tpeak}	6521	13482	kgO/d

Membrane aeration requirement

The aeration required for the membrane scouring is supplied via course bubble diffused aeration. From the suppliers of Kubota membranes, the required allowed airflow rates for effective scour of the membranes are is 180 Nm³/h per 200-panel unit at STP. This translates to **8.6 Nm³/h per m³ aerobic zone volume** (Q_{AIR}) for the single-story layout (3.5m deep). The oxygen transfer rate (OTR) for these airflow rates needs to be calculated to determine how much of the peak biological oxygen demand is supplied by the membrane aeration system.

The equation to calculate the oxygen supplied by the membrane system is indicated below.

Equation 10-3: Oxygen supplied by the membrane system

$OTR_{MBR} = 0.3\ Q_{AIR} OTE_{MBR} \alpha (V_{aer}) \times 1000 \times 24$ kgO/d (Ramphao *et al.*, 2013)

Kubota gives the oxygen transfer efficiency (OTE) at a water depth of 3.5m at around 5%. Alpha (α) was calculated based on the TSS concentration in the aerobic zone of 10 000 mg/l using Equation 10-4.

Equation 10-4: Alpha based on MLSS concentration

$\alpha = e^{-\omega . X_t}$

$\omega = 0.084$

The calculation results for the aeration design are summarized in Table 10-9:

Table 10-9: Oxygen supplied by the membrane system

		MLE	UCT	
MLSS concentration in the Aerobic zone		10000	10000	mg/l
$\alpha = e^{-\omega . X_t}$		0.43	0.43	Calculated
OTE_{MBR}		5.5	5.5	%/m
Q_{air}		8.6	8.6	Nm³/h
Aerobic zone volume		4.83	4.76	Ml
Oxygen transfer rate	OTR_{mbr}	7098	7004	kgO/d

This OTR required for effective scour supplies some or all of the biological oxygen demand in the aerobic zone.

In the MLE system, the membrane aeration system supplies all of the Peak biological oxygen demand. (7098 > 6521 kgO/d)

In the UCT system, the membrane aeration system does not supply the full biological demand. (7004<13482 kgO/d) Additional oxygen thus has to be supplied to supplement the biological demand deficit. This additional oxygen cannot be supplied into the membrane section of the aerobic reactor, so additional aerobic reactor volume needs to be provided to enable the transfer of the oxygen deficit. This is usually done with fine bubble aeration. Therefore, the volume of the aerobic zone (and thus the biological reactor) is governed, in this instance, by the **biological oxygen demand.**

The Aerobic zone volume thus needs to be increased in order to provide additional oxygen, hence the revised aerobic zone nominal hydraulic retention time is given by:

Equation 10-5: Revised aerobic zone nominal hydraulic retention time

$$R_{hn} = \frac{f_{vaer}V_R}{Q_{ADWF}} = \frac{Lf_q}{q_{max}} + \frac{MO_{tpeak} - OTR_{mbr}}{OTR_{FB}24} \text{ days}$$ (Ramphao *et al.*, 2013)

This additional aerobic zone volume increases the overall volume of the biological reactor because anaerobic and anoxic mass fractions remain unchanged at the specified design values and, hence the sludge age of the system must increase to generate sufficient sludge mass for the aerobic zone to operate at the required MLSS concentration. For the purpose of this evaluation, it is accepted that OTR_{FB} is 250 mgO/L.h aerobic zone volume. The adjusted reactor design based on the adjusted Nominal Hydraulic retention time (Equation 10-5) required for the UCT system is summarized in Table 10-10:

Table 10-10: Adjusted design for MBR - UCT

		UCT	
Nominal Hydraulic retention time	R_s	0.27	d
Sludge age	Rs	21	d
Mass of TSS in the reactor	MX_t	109097	kgTSS
Volume of Aerobic zone	V_{aer}	5.45	Ml
Volume of Anaerobic zone	V_{ana}	3.31	Ml
Volume of Anoxic zone	V_{anx}	4.71	Ml
Total volume of reactor	V_p	13.47	Ml

10.2 Zeeweed HF MBR design

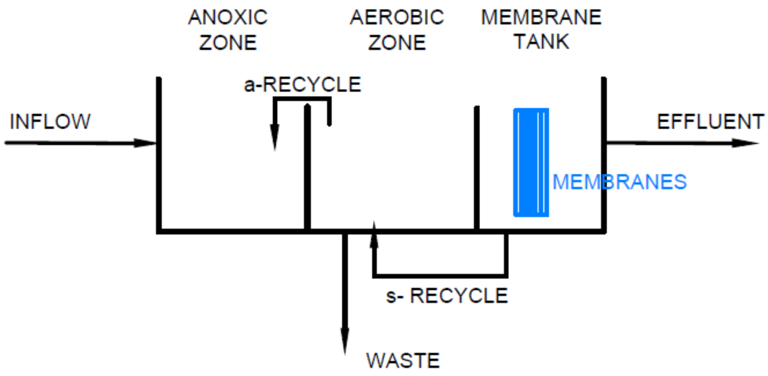

Figure 10-3 *– Process flow schematic of HF MBR in MLE configuration*

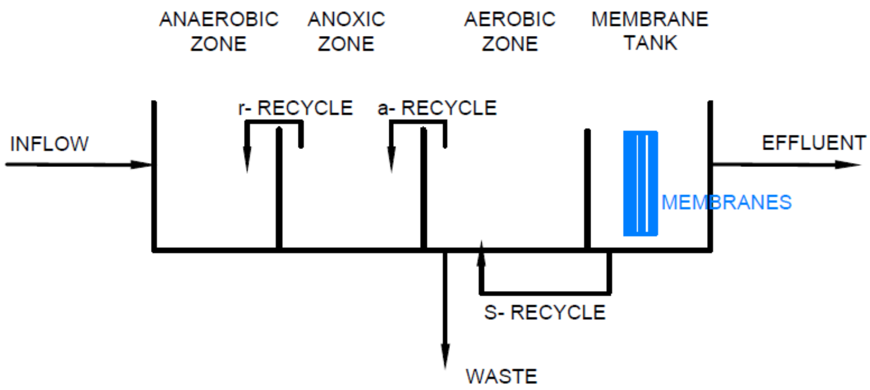

Figure 10-4 *– Process flow schematic of HF MBR in UCT configuration*

MEMBRANE DESIGN

The membranes considered in this study are the *Zeeweed 500D HF* membranes by *Suez Water Technologies*. The membrane design parameters used in this study have been received curtesy of *Suez Water Technologies* (SUEZ, 2019).

As mentioned previously, one of the key parameters in the design of the membranes is the **net flux (j_{net}).** The required membrane area (A_m) is calculated by dividing the relevant flow rate by the permissible flux. Membrane fluxes are provided for the different flow rates are shown in Table 10-11. During peak flows, the membranes can handle higher fluxes for short periods of time.

Table 10-11: Membrane fluxes (Confirmed by Seuz, 2019)

Q_{ADWF}	22.6 @10 g/l	LMH (normalized to 20°C)
Q_{PDWF}	35.2 @12 g/l	LMH (normalized to 20°C)
Q_{PWWF}	40.3 @12 g/l	LMH (normalized to 20°C)

The membrane area (Am) was then determined for each flow rate using

Equation 10-6, The calculated membrane area for each flow is shown in

Table 10-12

Equation 10-6: Minimum membrane area required

$$A_m = \frac{Q}{j_{net}}$$

Table 10-12: Membrane area required for each scenario

	MLE	UCT	
Q_{AWDF}	36874	36874	m²
Q_{PWDF}	47350	47350	m²
Q_{PWWF}	51697	51697	m²

Based on the above, the required membrane area for both the MLE and UCT systems is 51697 m².

The volume of the membrane tank

The membrane tank volume($V_{m,min}$) is determined by the membrane area (A_m) and membrane packing density (φ_{tank}). Membranes are installed as “cassettes” which consist of modules. These modules have a packing density (φ_{tank}) of **45 m²/m³,** therefore dividing the required membrane area by the packing density gives the required tank volume - Equation 10-7.

Equation 10-7: Minimum required membrane tank volume

$$V_{m,min} = \frac{A_m}{\varphi_{tank}}$$

The required membrane tank volumes are shown in Table 10-13

Table 10-13: Minimum membrane tank volume

	MLE	UCT	
$V_{m'min}$	1149	1149	m^3

Designing for Redundancy

It must be noted that for an MBR design for implementation, one would design for operation redundancy. This is achieved by dividing the membrane tank into a number of trains. This way a train can be taken out of operation for maintenance while the flow is diverted to other trains. This does not affect the membrane tank volume required, therefore this level of design will not be done as part of this investigation. The purpose of this study is to compare the various systems from an economic point of view. The procedure to determine the number of membrane cassettes is as follows: Membrane modules are 34.4 m^2 per module. There are 52 modules in a “cassette”. At 50 ML/d PWWF: 50 000 000 L/24 hrs/40.3 LMH = 51696 m^2 = 1503 modules = 29 cassettes. Hence 4 trains of 8 cassettes or 5 trains of 6 cassettes.

BIOLOGICAL DESIGN

Determination of Mass fractions and Sludge Age

In a HF MBR, the membranes are located in a separate membrane tank, therefore the selection of the sludge age is not based on the membrane requirement, as is for a FS MBR design where the aerobic zone size needs to cater for membranes. The sludge age in a HF MBR design is only based on the minimum requirement sludge age (SRT_m) to ensure nitrification. The minimum sludge age was calculated using Equation 10-8 (Ekama, GA. Wentzel, 2008):

Equation 10-8: Minimum required sludge age to ensure nitrification

$$RS_m \; or \; SRT_m = \frac{S_f}{(1 - f_{xt})\mu_{AmT} - b_{AT}S_f}$$

The maximum unaerated mass fractions (f_{xt}) were chosen and using the above equation results in the minimum sludge ages as summarized in Table 10-14

Table 10-14: Calculation results for minimum sludge age to nitrification

		MLE	UCT	
Maximum unaerated mass fraction	f_{xt}	0.38	0.5	0.38 and 0.5 chosen for this investigation
Temperature	T	14	14	
Minimum sludge age	R_{sm}	11.12	15.15	

Based on the above, the design sludge ages were therefore chosen as **12 days** for the MLE system and **16 days** for the UCT system.

For the MLE system, the aerobic mass fraction (f_{maer}) is therefore 1 minus the selected maximum unaerated mass fraction:

$1 – f_{xt} = 1 - 0.38 = 0.62$

For the UCT system the aerobic mass fraction (f_{maer}) is therefore 1 minus the selected maximum unaerated mass fraction:

$1 – f_{xt} = 1 - 0.5 = 0.5$

For the UCT system, the anaerobic mass fraction (f_{mana}) was selected as 0.12 . The anoxic mass fraction (f_{manx}) was therefore = $1 - f_{maer} - f_{mana} = 1 - 0.5 – 0.12 = 0.38$

Determination of MLSS in the Reactor

With the sludge age known, the mass of total suspended solids (MX_t) in the reactor can be calculated. This is done in the same way as for CAS. The equations used are based on the Steady State theory as per (Ekama, GA. Wentzel, 2008) As mentioned previously, the same kinetic constants used for CAS are deemed acceptable to use in the design of an MBR. The list of kinetic values used are indicated in Table 7-3 in Section 7.1. The calculation results for the MX_t for the 2 systems are indicated in Table 10-15

Table 10-15: Mass of total suspended solids in the reactor for the HF MBR MLE and UCT

		MLE	UCT	
Mass of total suspended solids in reactor	MX_t	71643	99167	kgTSS

Recycle ratios and Zone MLSS concentrations

A key constraint of MBR is the requirement for a relatively short retention time in the membrane tank to limit the concentration of solids and subsequent membrane clogging. This demands that the transfer rate between the membrane tank and biological tank (the RAS) is in the region of 3-5 times the treated water flow (Q_{ADWF}) – much higher than the equivalent RAS flow for CAS, which is usually around 1:1. The return flow is also relatively high in dissolved oxygen (DO), which makes the anoxic zone, to which concentrated sludge is returned for denitrification less efficient. It is thus usually returned to the aerobic zone. The sludge concentration in the membrane tank was selected to be 10 000 mg/l based on the minimum requirement for scouring by SUEZ, 2019.

The system parameters chosen are shown in Table 10-16

Table 10-16: recycle ratios selected

		MLE	UCT	
sludge concentration in the membrane tank	X_{mr}	10000	10000	mg/l
recycle ratio of sludge return from the membrane tank to the aerobic tank	r_{mr}	4	4	
recycle ratio of nitrate from aerobic zone to anoxic zone	r_{nit} or "a"	6	6	
recycle ratio from anoxic to anaerobic zone	r_p or "r"	0	1	

As the membranes are located in a separate membrane tank, the TSS concentrations are spread differently in the various zones compared to the Kubota MBR system. The sludge concentrations in the different zones (X_{tana}, X_{tanx}, and X_{taer}) can be calculated based on the recycle ratios. The equations used are as per (Judd, 2011):

Equation 10-9: Sludge concentration in the Aerobic zone

$$X_{aer} = X_{mr} \cdot \frac{r_{mr}}{1 + r_{mr}}$$

Equation 10-10: Sludge concentration in Anoxic zone

$$X_{anx} = X_{aer} \cdot \frac{r_{nit} \cdot}{1 + r_{nit}}$$

Equation 10-11: Sludge concentration in Anaerobic zone

$$X_{ana} = X_{anx} \frac{r_p \cdot}{1 + r_p}$$

The symbology used are as per (Judd, 2011) which is:

X_{mr} – sludge concentration in the membrane tank

r_{mr} – recycle ratio of sludge return from the membrane tank to the aerobic tank

r_{nit} - recycle ratio of nitrate from aerobic zone to anoxic zone

r_p - recycle ratio from anoxic to the anaerobic zone

The calculation results for the sludge zone concentrations are summarized in Table 10-17

Table 10-17: Sludge concentrations in different zones

	MLE	UCT		
X_{taer}	8000	8000	mg/l	Equation 10-9
X_{tanx}	6857	6857	mg/l	Equation 10-10
X_{tana}	-	3429	mg/l	Equation 10-11

The volume of Reactor zones

With mass fractions, zone sludge concentrations and mass of solids in the reactor known, the volumes of the different zones can be calculated using equations below. The results are summarized in Table 10-18

Equation 10-12: Volume of the Aerobic zone

$$V_{aer} = \frac{MX_t . f_{maer}}{X_{t,aer}}$$

Equation 10-13: Volume of the Aerobic zone

$$V_{anx} = \frac{MX_t . f_{manx}}{X_{t,anx}}$$

Equation 10-14: Volume of the Aerobic zone

$$V_{ana} = \frac{MX_t . f_{mana}}{X_{t,ana}}$$

Table 10-18: Reactor zone and total volumes

	MLE	**UCT**	
V_{aer}	5552	6198	m³
V_{anx}	3970	5496	m³
V_{ana}	-	3471	m³
V_p	9523	15164	m³

The Total reactor volume including the membrane tank volume is indicated in Table 10-19

Table 10-19: Reactor and membrane tank volume

	MLE	**UCT**	
V_t	10671	16313	m³

Aeration design

From the suppliers of the Zeeweed membranes, the minimum required airflow rate for effective scour (SAD_m) of the membranes is 0.3 Nm³/h per m² membrane area, based on the membrane area. The required membrane airflow rate is shown in Table 10-20:

Table 10-20: Required airflow for membranes

	MLE	UCT	
$Q_{A,m}$	15509	15509	Nm³/h

This airflow supplies part or the full biological oxygen demand. It is thus necessary to determine how much of the biological oxygen demand is being supplied by the membrane aeration system. This $Q_{A,m}$ is at STP and needs to be converted to Actual Oxygen Rate ($AOR_{site,\ mbr}$). This was done using Equation 10-15

Equation 10-15: Actual Oxygen Rate

$$AOR_{site} = Q_{air,std}.C_{02,std}.SOTE_{std}.\{\}$$

The $AOR_{site,mbr}$ was calculated using the aeration design parameters as indicated in Table 7-4. The results are indicated in Table 10-21

Table 10-21: Oxygen supplied by the membrane aeration system

	MLE	UCT	
$AOR_{site,mbr}$	7563	7563	kgO/d

Biological aeration demand

The biological oxygen demand for an MBR system is determined the same way as for a CAS system. The results are summarized in Table 10-22

Table 10-22: Oxygen demand of HF MBR- MLE and UCT systems

		MLE	UCT	
Carbonaceous oxygen demand OHO's	$FO_{C\ OHO}$	4236	8594	kgO/d
Carbonaceous oxygen demand POA's	$FO_{C\ POA}$	-	560	kgO/d
Carbonaceous oxygen demand	FO_c	4236	9153	kgO/d
Nitrogenous oxygen demand	FO_n	2064	3396	kgO/d
Oxygen demand recovered by denitrification	FO_d	1130	1860	kgO/d
Total oxygen demand (FO_c + FO_n - FO_d)	FO_{tave}	5170	10690	kgO/d

The amplitude of the TOD influent (TOD_{peak}/TOD_{ave} - 1) is 0.8 kg/h. The damping factor (d_L) used is 0.33. (Musvoto et al, 2002) The FO_{tpeak} (Peak total oxygen demand) was then as calculated and the results are summarized in Table 10-23:

Table 10-23: Peak Oxygen demand of FS MBR- MLE and UCT systems

		MLE	UCT	
Peak total oxygen demand	FO_{tpeak}	6520	13482	kgO/d

For the MLE system, the full FO_{tpeak} is provided by the membrane aeration system. (7563>6520). For the UCT system, the membrane aeration only provides part of the FO_{tpeak}.

The oxygen deficit is to be supplied by additional aeration located in the aerobic zone. This deficit to be supplied is the difference between the AOR $_{site,mbr}$ which is supplied by the membrane system, and the required FO_{tpeak}. The additional oxygen to be supplied = 13482 - 7563 = 5918.42 kgO/d

10.3 Performance comparison / Super summary of Kubota FS MBR and Zeeweed HF MBR design results:

A comparison of the results from the 2 MBR system designs are summarized in Table 10-24:

Table 10-24: Super summary of Kubota FS MBR and Zeeweed HF MBR design results

	Kubota FS MBR		**Zeeweed HF MBR**		
	MLE	**UCT**	**MLE**	**UCT**	
Membrane design					
Design ADWF	20	20	20	20	Ml/d
Design PWWF	50	50	50	50	Ml/d
Flux (at PWWF)	41.67	41.67	40.3	40.3	LMH
Membrane area required	50001	50001	51697	51697	m^2
Membrane packing density	12	12	45	45	m^2/m^3
Membrane tank TSS concentration	10000	10000	10000	10000	mg/l
Vol. of membrane tank	-	-	1.149	1.149	Ml
Biological design					
Temperature	14	14	14	14	T
Unaerated mass fraction	0.45	0.5	0.38	0.5	
Anaerobic mass fraction	0	0.13	-	0.12	
Anoxic mass fraction	0.45	0.37	0.38	0.38	
Aerobic mass fraction	0.55	0.5	0.62	0.5	
a-recycle ratio	0	6	6	6	
r-recycle ratio	0	1	0	1	
s-recycle ratio (RAS)	4	1	4	4	
MX_t	87776	95273	71643	99167	kgTSS
Adjusted MX_t		109097			kgTSS
Anaerobic TSS concentration	-	4286	-	3429	mg/l
Anoxic TSS concentration	8000	8571	6857	6857	mg/l
Aerobic TSS concentration	10000	10000	8000	8000	mg/l
Vol. of aerobic zone	4.83	5.45	5.55	6.20	Ml
Vol. of anoxic zone	4.94	4.71	3.97	5.50	Ml
Vol. of anaerobic zone	-	3.31	-	3.47	Ml
Total reactor volume (incl.membrane tank)	9.77	13.47	10.67	16.31	Ml
Aeration (membrane requirement)					
Membrane airflow req.	0.75	0.75	0.3	0.3	$Nm^3/m^2.h$
OTE	5.5	5.5	kgO/h	kgO/h	%/m
Alpha	0.43	0.43	0.43	0.43	Based on a X_t of 10g/l
Water depth	3.75	3.75	3.75	3.75	m
Membrane airflow			15509	15509	Nm^3/h
Oxygen available from membrane system	7098	7004	7563	7563	kgO/d
Aeration (membrane requirement)					
FO_c	4237	9153	4236	9153	kgO/d

FO_n	2064	3396	2064		kgO/d
FO_d	1130	1860			kgO/d
$FOT_{t,ave}$	5170	10690			kgO/d
FOT_{peak}	6521	13482	6520	13482	kgO/d
Additional oxygen req. to supplement membrane aeration to meet biological demand	NO	YES	NO	YES	kgO/d

11 Results from CAS design

11.1 Determining minimum sludge age for Nitrification

The design of the CAS system was based on the steady state model equations by (Ekama, GA. Wentzel, 2008)

The following procedure was followed for the CAS design. The inputs and results are indicated in Table 11-1.

1) The minimum sludge age to ensure nitrification was calculated based on a pre-selected unaerated mass fraction of 0.38 for the MLE system and 0.5 for the UCT system. The anaerobic mass fraction in the UCT system was pre-selected as 0.12.
2) The recycle ratios were pre-selected as a = 6 for the MLE and UCT system, and r = 1 for the UCT system.
3) With the sludge age known, the MX_t could then calculated. The kinetic constants used are listed in Section 7.1. The wastewater characteristics used are the same used for the MBR design.
4) The reactor zone volumes were then calculated based on the anoxic TSS concentration being the same as the aerobic zone concentration and the anaerobic zone TSS concentration being half that of the aerobic zone concentration. The aerobic zone concentration used was 5500 for the MLE configuration and 6000 mg/l for the UCT configuration as determined in the cost minimization exercise done in Section 8.
5) The Peak Aeration Demand was then calculated

The results for the CAS system design are summarized in Table 11-1

Table 11-1: Results from CAS design for MLE and UCT system

	MLE	UCT	
f_{xt}	0.38	0.5	
T	14	14	
R_{sm}	11.12	15.15	Minimum sludge age calculated based on the mass unaerated mass fraction
a-recycle	6	6	
r-recycle	1	1	
f_{ana}	-	0.12	
f_{anx}	0.38	0.38	Based on the selection of max unaerated mass fraction of 0.38 and 0.5
f_{aer}	0.62	0.5	
MX_t	71643	99167	kgTSS
$X_{t, aer}$	5500	6000	Xt based on cost optimization exercise
$X_{t, anx}$	5500	6000	Assumed the same as in Aerobic zone
$X_{t, ana}$	-	3000	Assumed half of Aerobic TSS concentration
V_{aer}	8177	8264	m^3
V_{anx}	5012	6281	m^3
V_{ana}	-	3967	m^3
V_p	13188	18511	m^3
FO_c	4236	9153	kgO/d
FO_n	2064	3396	kgO/d
FO_d	1130	1860	kgO/d
$FOT_{t,ave}$	5170	10690	kgO/d
FOT_{peak}	6520	13482	kgO/d

The SST was designed based on the 1D Flux theory by Takacs and Ekama, 2008)

The maximum rise rate was calculated using the sludge settling characteristics listed in Section 7.1 and Reactor TSS concentration. Together with the PWWF and Safety factor, the required SST area was then determined. The results are summarized in Table 11-2:

Table 11-2: Calculation results for SST design for MLE and UCT system

	MLE			**UCT**			
	DSVI 100	DSVI 150	DSVI 200	DSVI 100	DSVI 150	DSVI 200	DSVI 100
Q_{PWWF}	50	50	50	50	50	50	Ml/d
X_t	5.5	5.5	5.5	6	6	6	gTSS/l
$q_{A,MAX}$	1.21	0.53	0.23	1.01	0.43	0.18	m/h
S_f	25	25	25	25	25	25	%
A_{min}	2155	4809	11230	2558	5976	14608	m^2

A summary of the MBR system design compared to the CAS design results are summarized in Table 11-3

Table 11-3: Super summary/performance comparison for Kubota MBR, Zeeweed MBR, and CAS systems

	Kubota FS MBR		Zeeweed HF MBR		CAS		
	MLE	UCT	MLE	UCT	MLE	UCT	
Membrane design							
Design ADWF	20	20	20	20	20	20	Ml/d
Design PWWF	50	50	50	50	50	50	Ml/d
Flux (at PWWF)	41.67	41.67	40.3	40.3	-	-	LMH
Membrane area required	50001	50001	51697	51697	-	-	m^2
Membrane packing density	12	12	45	45	-	-	m^2/m^3
Membrane tank TSS concentration	10000	10000	10000	10000	-	-	mg/l
Vol. of membrane tank	4.375	4.38	1.149	1.149	-	-	Ml
Biological design							
Temperature	14	14	14	14	14	14	T
Unaerated mass fraction	0.45	0.5	0.38	0.5	0.38	0.5	
Anaerobic mass fraction	0	0.13		0.12	-	0.12	
Anoxic mass fraction	0.45	0.37	0.38	0.38	0.38	0.38	
Aerobic mass fraction	0.55	0.5	0.62	0.5	0.62	0.5	
a-recycle ratio	0	6	6	6	6	6	
r-recycle ratio	0	1	0	1	1	1	
s-recycle ratio (RAS)	4	1	4	4	1	1	
MX_t	87776	95273	71643	99167	71643	99167	kgTSS
Adjusted MX_t		109097					
Anaerobic TSS concentration	-	4286	-	3429	-	3000	mg/l
Anoxic TSS concentration	8000	8571	6857	6857	5500	6000	mg/l
Aerobic TSS concentration	10000	10000	8000	8000	5500	6000	mg/l
Vol. of aerobic zone	4.83	5.45	5.55	6.20	8.18	8.26	Ml
Vol. of anoxic zone	4.94	4.71	3.97	5.50	5.01	6.28	Ml
Vol. of anaerobic zone	-	3.31	-	3.47	-	6.28	Ml
Total reactor volume (incl.membrane tank)	9.77	13.47	10.67	16.31	13.19	18.51	Ml
Aeration (membrane requirement)							
Membrane airflow req.	0.75	0.75	0.3	0.3	-	-	$Nm^3/m^2.h$
OTE	5.5	5.5	6.5	6.5	-	-	%/m
Alpha value	0.43	0.43	0.43	0.43	-	-	Based on a X_t of 10g/l
Water depth	3.75	3.75	3.75	3.75	-	-	M
Membrane air flow required	41518	46912	15509	15509	-	-	Nm^3/h
Oxygen available from membrane system	7098	7004	7563	7563	-	-	kgO/d
Aeration (biological requirement)							
FO_c	4237	9153	4236	9153	4236	9153	kgO/d
FO_n	2064	3396	2064		2064		kgO/d
FO_d	1130	1860					kgO/d
$FOT_{t,ave}$	5170	10690					kgO/d
FOT_{peak}	6521	13482	6520	13482	6520	13482	kgO/d
Additional oxygen req. to supplement membrane aeration to meet biological demand	NO	YES	NO	YES	-	-	kgO/d

12 Costing and Discussions

Each system was costed using the cost functions derived in Section 8 and Section 9. The costing results are presented in this section.

12.1 CAPEX

REACTOR CAPEX

Based on the reactor volumes determined previously and calculating the reactor cost using Equation 8-1, the costing results for all of the systems are indicated in Table 12-1

Table 12-1: Reactor CAPEX results for all systems

	Kubota FS MBR		Zeeweed HF MBR		CAS		
	MLE	**UCT**	**MLE**	**UCT**	**MLE**	**UCT**	
Total reactor volume (incl.membrane tank)	9.77	13.47	10.67	16.31	13.19	18.51	Ml
Cost	R11 639 530	R14 408 702	R12 345 026	R16 356 632	R14 205 758	R17 786 511	

The Kubota MBR system had the lowest reactor CAPEX of all the systems at R11 639 530 and R14 408 702 respectively for the MLE and UCT configuration. The Zeeweed MBR had the 2nd lowest reactor cost which was 7% and 25% more than the Kubota systems respectively, for the MLE and UCT configurations. The CAS system had the highest reactor cost which was 22% and 54% more than the Kubota MBR respectively for the MLE and UCT system. DSVI does not affect reactor size/cost, therefore, the reactor cost was the same for the various DSVI values in the CAS systems.

MEMBRANE CAPEX

Based on the membrane areas determined previously and calculating the cost of the membranes as per Section 9.1, the costing results for all of the systems are indicated in Table 12-2

Table 12-2: Membrane CAPEX results for all systems

	Kubota FS MBR		Zeeweed HF MBR		CAS		
	MLE	**UCT**	**MLE**	**UCT**	**MLE**	**UCT**	
Membrane area required	50001	50001	51697	51697	-	-	m^2
Membrane cost	R45 001 333	R45 001 333	R36 188 004	R36 188 004	-	-	

The Zeeweed HF MBR had the lowest membrane cost of the systems at R36 188 004 for both the MLE and UCT configurations as the Q_{PWWF} were the same for both configurations. The Kubota FS MBR membrane cost was R45 001 333 which was 24% higher than the Zeeweed membranes.

SST CAPEX

Based on the SST areas determined in Section 11 previously and calculating the cost of the SSTs using Equation 8-2, the SST costing results for all of the systems are indicated in Table 12-3

Table 12-3: SST CAPEX results for all systems

	CAS MLE			CAS UCT			
	DSVI 100	DSVI 150	DSVI 200	DSVI 100	DSVI 150	DSVI 200	
SST area	2155	4809	11230	2558	5976	14608	m²
No of SSTs	3	4	9	3	5	12	
Ø	30	39	40	33	39	39	
Cost	R10 787 015	R18 402 534	R42 148 609	R11 709 680	R22 938 525	R55 538 438	

As can be seen from Table 12-3, the DSVI has a substantial impact on the SST area and cost. A DSVI increase to 150 and 200 (from 100) in the MLE system, resulted in a large cost increase of 71% and 291% respectively. In the UCT system, a DSVI increase from to 150 and 200, resulted in a substantial cost increase of 113% and 451% respectively.

AERATION CAPEX

Based on the aeration requirements determined previously and calculating the cost of the aeration system using Equation 8-3, the aeration costing results for all of the systems are indicated in Table 12-4

Table 12-4: Aeration CAPEX results for all systems

	Kubota FS MBR		**Zeeweed HF MBR**		**CAS**		
	MLE	**UCT**	**MLE**	**UCT**	**MLE**	**UCT**	
Power of blower	657	742	245	495	164	354	kW
Cost	R7 888 867	R8 223 361	R5 644 351	R7 166 429	R4 922 939	R6 392 768	

As seen in Table 12-4 the CAS system had the lowest aeration CAPEX of the systems at R4 922 939 and R6 392 768 for the MLE and UCT configurations respectively. The MBR system had a higher aeration CAPEX which was expected due to the additional scouring requirement of the membranes. The Zeeweed MBR had an aeration CAPEX of 13% and 45% more than the CAS system for the MLE and UCT systems respectively. Followed by the Kubota MBR which had an aeration CAPEX cost of 56% and 66% more than the CAS system for the MLE and UCT systems respectively. DSVI does not affect aeration requirement, therefore, the aeration cost was the same for the various DSVI values in the CAS systems.

TOTAL CAPEX

The total CAPEX of all the systems are indicated in Table 12-5 and below in Figure 12-1

Table 12-5: Total CAPEX results for all systems

	Kubota FS MBR		Zeeweed HF MBR		CAS					
	MLE	**UCT**	**MLE**	**UCT**	**MLE @ 100DSVI**	**MLE @ 150DSVI**	**MLE @ 200DSVI**	**UCT @ 100DSVI**	**UCT @ 150DSVI**	**UCT @ 200DSVI**
Reactor cost	R11 639 530	R14 408 702	R12 345 026	R16 356 632	R14 205 758	R14 205 758	R14 205 758	R17 786 511	R17 786 511	R17 786 511
Membrane Cost	R45 001 333	R45 001 333	R36 188 004	R36 188 004	-	-	-	-	-	-
SST Cost					R10 787 015	R18 402 534	R42 148 609	R11 709 680	R22 938 525	R55 538 438
Aeration Cost	R7 888 867	R8 223 361	R5 644 351	R7 166 429	R4 922 939	R4 922 939	R4 922 939	R6 392 768	R6 392 768	R6 392 768
Total	**R64 529 730**	**R67 633 396**	**R54 177 381**	**R59 711 064**	**R29 915 713**	**R37 531 231**	**R61 277 306**	**R35 888 959**	**R47 117 804**	**R79 717 717**

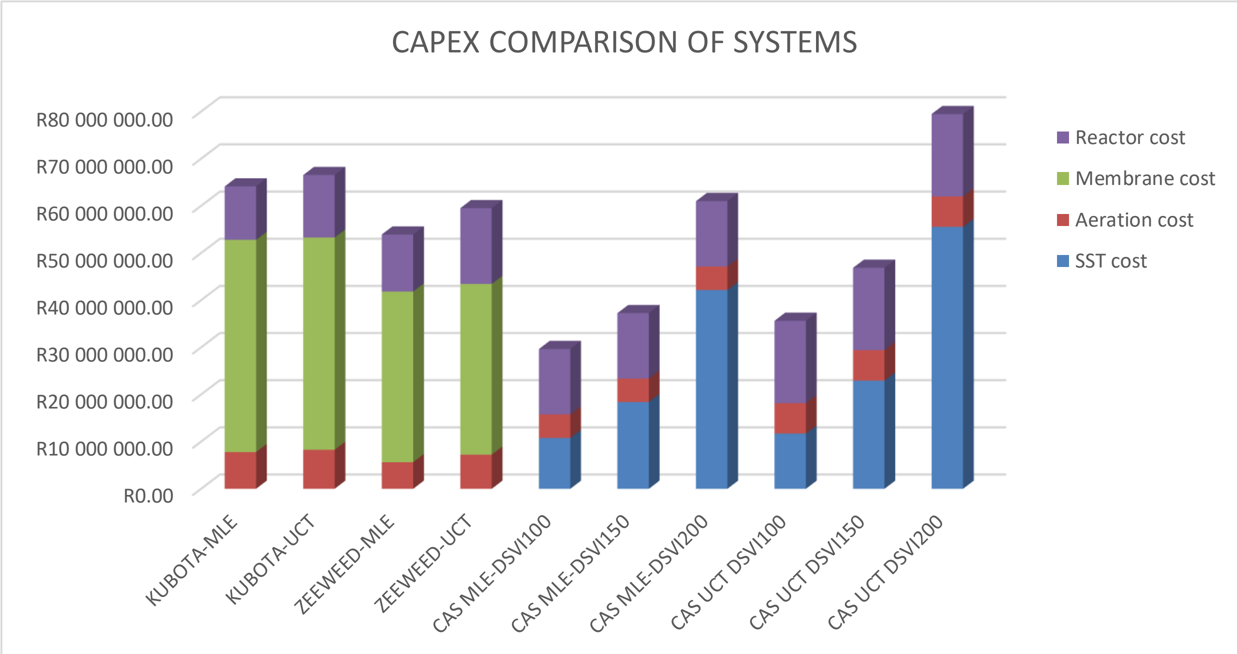

Figure 12-1 – *CAPEX comparison of all systems*

The systems which had the lowest CAPEX were the CAS MLE and UCT systems. The system which had the highest CAPEX was surprisingly the CAS systems with a DSVI of 200. These systems even had a higher CAPEX than the MBR systems due to the significant impact of poor sludge settleability on the cost of the SSTs. The membrane costs were by far the largest contributors to the CAPEX in the MBR systems. It must be noted that MBRs require ultra-fine screening (<2mm) upstream of the bioreactor, this is not required for CAS systems. The cost for screening was not included in this study as this cost is insignificant compared to the membranes and aeration.

Table 12-6 below compares the CAPEX of all the systems in relation to the lowest-ranked (lowest cost) system which was the CAS MLE @ DSVI 100

Table 12-6: Total CAPEX results for all systems

Ranking	System	% higher cost than the lowest
1	CAS MLE @ DSVI 100	
2	CAS UCT @ DSVI 100	20%
3	CAS MLE @ DSVI 150	26%
4	CAS UCT @ DSVI 150	58%
5	Zeeweed MBR MLE	82%
6	Zeeweed MBR UCT	101%
7	CAS MLE @ DSVI 200	106%
8	Kubota MBR MLE	116%
9	Kubota MBR UCT	129%
10	CAS UCT @ DSVI 200	168%

12.2 OPEX

The OPEX was determined for each system using the same equations derived for the cost minimization exercise done in Section 8.1.4 As previously mentioned, for this investigation, the OPEX consisted only of the aeration energy cost as this is the highest contributor to the OPEX on a wastewater treatment plant. The OPEX was determined over a lifecycle of 10 years. The results are indicated in Table 12-7 and Figure 12-2.

Table 12-7: OPEX results for all systems

	Kubota FS MBR		Zeeweed HF MBR		CAS	
	MLE	**UCT**	**MLE**	**UCT**	**MLE**	**UCT**
OPEX	R143 788 200	R162 467 559	R53 712 395	R108 403 023	R35 925 210	R77 467 305

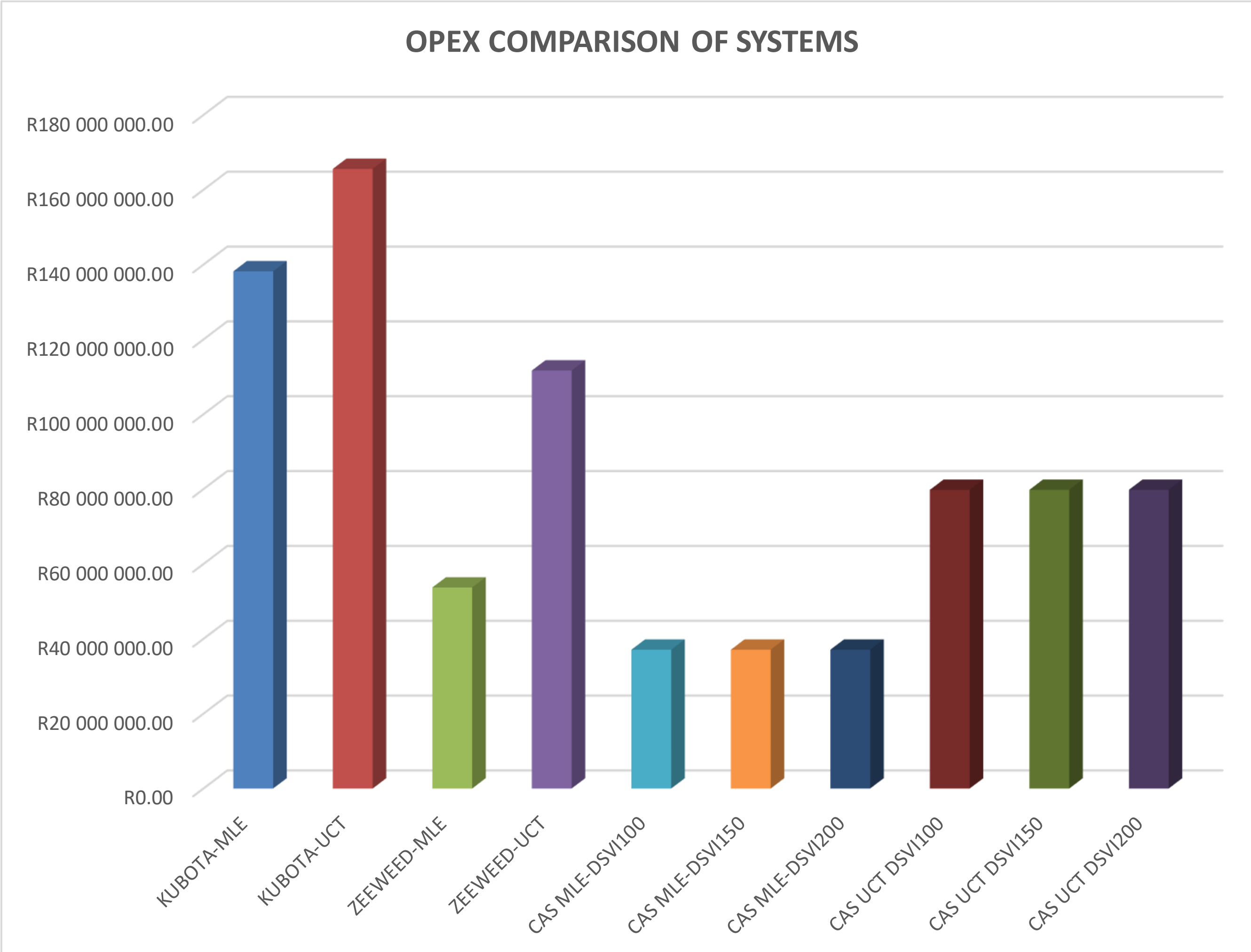

Figure 12-2 – *OPEX comparison of all systems*

The system which had the lowest OPEX was the CAS MLE system. The system with the 2nd lowest OPEX cost was Zeeweed MBR MLE which came in surprisingly lower than the CAS UCT system. The Kubota MBR systems had the highest OPEX costs of all systems. This was due to the higher aeration requirement for scouring (±2x) of the Kubota FS membranes compared to the Zeeweed.

Table 12-6 compares the OPEX of all the systems in relation to the lowest-ranked (lowest cost) system which was the CAS MLE.

Table 12-8: OPEX results for all systems

Ranking	System	% higher cost than the lowest
1	CAS MLE	
2	Zeeweed MBR MLE	45%
3	CAS UCT	115%
4	Zeeweed MBR UCT	201%
5	Kubota MBR MLE	273%
6	Kubota MBR UCT	346%

12.3 TOTAL COSTS

The total costs of all the systems are indicated in Table 12-9 and Figure 12-3

Table 12-9: Total cost results for all systems

	Kubota FS MBR		Zeeweed HF MBR		CAS					
	MLE	**UCT**	**MLE**	**UCT**	**MLE @ 100DSVI**	**MLE @ 150DSVI**	**MLE @ 200DSVI**	**UCT @ 100DSVI**	**UCT @ 150DSVI**	**APEX**
CAPEX	R64 529 730	R67 633 396	R54 177 381	R59 711 064	R29 915 713	R37 531 231	R61 277 306	R35 888 959	R47 117 804	R79 717 717
OPEX	R143 788 200	R162 467 559	R53 712 395	R108 403 023	R35 925 210	R35 925 210	R35 925 210	R77 467 305	R77 467 305	R77 467 305
Total cost	R208 317 929	R230 100 955	R107 889 776	R168 114 087	R65 840 923	R73 456 441	R97 202 516	R113 356 265	R124 585 110	R157 185 022

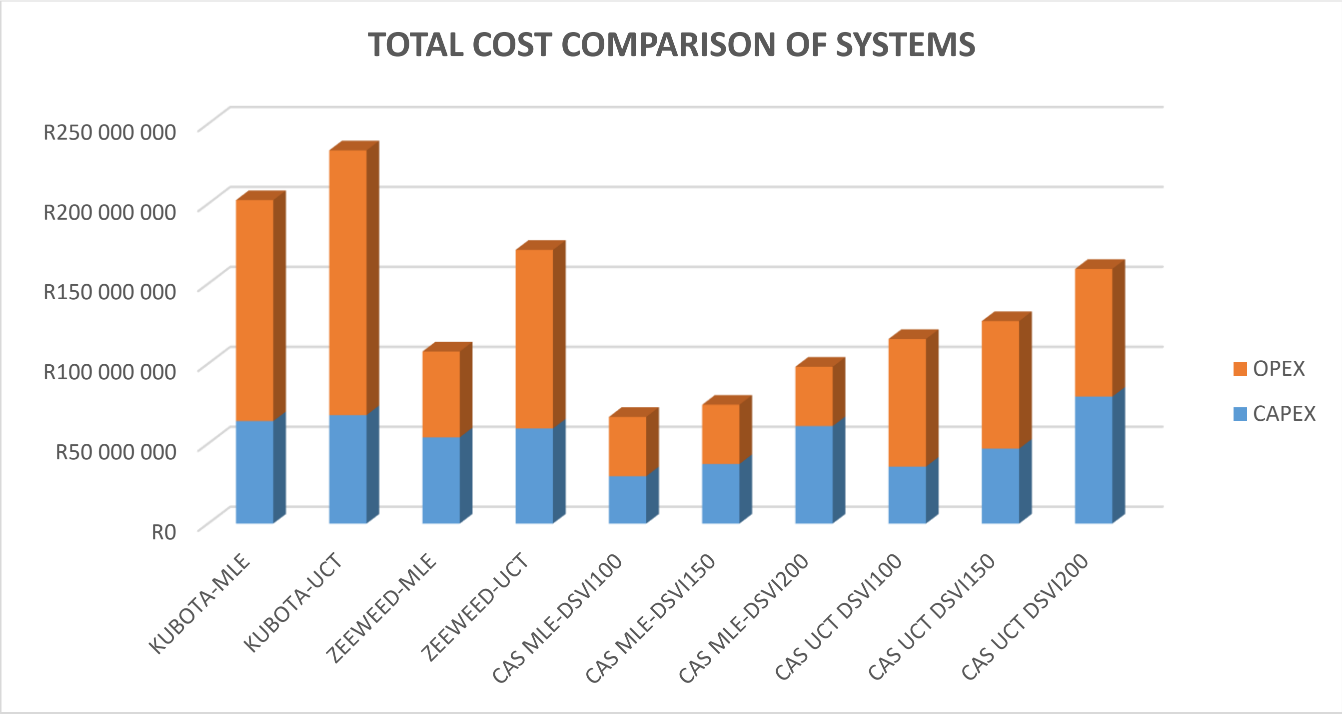

Figure 12-3 – *Total cost comparison of all systems*

The system which had the lowest total cost was the CAS MLE systems. The Zeeweed MBR MLE system had the 4th lowest total cost coming in at lower than the CAS UCT systems. This was unexpected. The systems with the highest costs were the Kubota MBR MLE and UCT systems which were 203% and 250% more than the system with the lowest total cost.

Table 12-6 compares the total cost (CAPEX + OPEX) of all the systems in relation to the lowest-ranked (lowest cost) system which was the CAS MLE DSVI 100.

Table 12-10: Total cost results for all systems

Ranking	System	% higher cost than the lowest
1	CAS MLE @ DSVI 100	
2	CAS MLE @ DSVI 150	11%
3	CAS MLE @ DSVI 200	47%
4	Zeeweed MBR MLE	61%
5	CAS UCT @ DSVI 100	73%
6	CAS UCT @ DSVI 150	90%
7	CAS UCT @ DSVI 200	139%
8	Zeeweed MBR UCT	157%
9	Kubota MBR MLE	203%
10	Kubota MBR UCT	250%

12.4 LAND AREA

The required land area was determined for each system based on the following:

- The area required for the reactor was based on the determined reactor volumes and a water depth of 4m. 40% was then added for access and pipelines.
- The area required for the SSTs was determined and 40% added for access and pipelines.
- Area for the blowers was not included

The required land areas for each system are indicated in Table 12-11 and Figure 12-4

Table 12-11: Land area results for all systems

	Kubota		**Zeeweed**		**CAS MLE**			**CAS UCT**			
	MLE	UCT	MLE	UCT	DSVI 100	DSVI 150	DSVI 200	DSVI 100	DSVI 150	DSVI 200	
Reactor vol.	9.77	13.47	10.67	16.31	13.19	13.19	13.19	18.51	18.5	18.51	Ml
Water depth	4	4	4	4	4	4	4	4	4	4	M
Reactor area	2441	3368	2668	4078	3297	3297	3297	4628	4628	4628	m^2
SST area					2155	4809	11230	2558	5976	14608	m^2
Total area (incl. access)	**3418**	**4716**	**3735**	**5710**	**7633**	**11349**	**20338**	**10060**	**14845**	**26931**	m^2

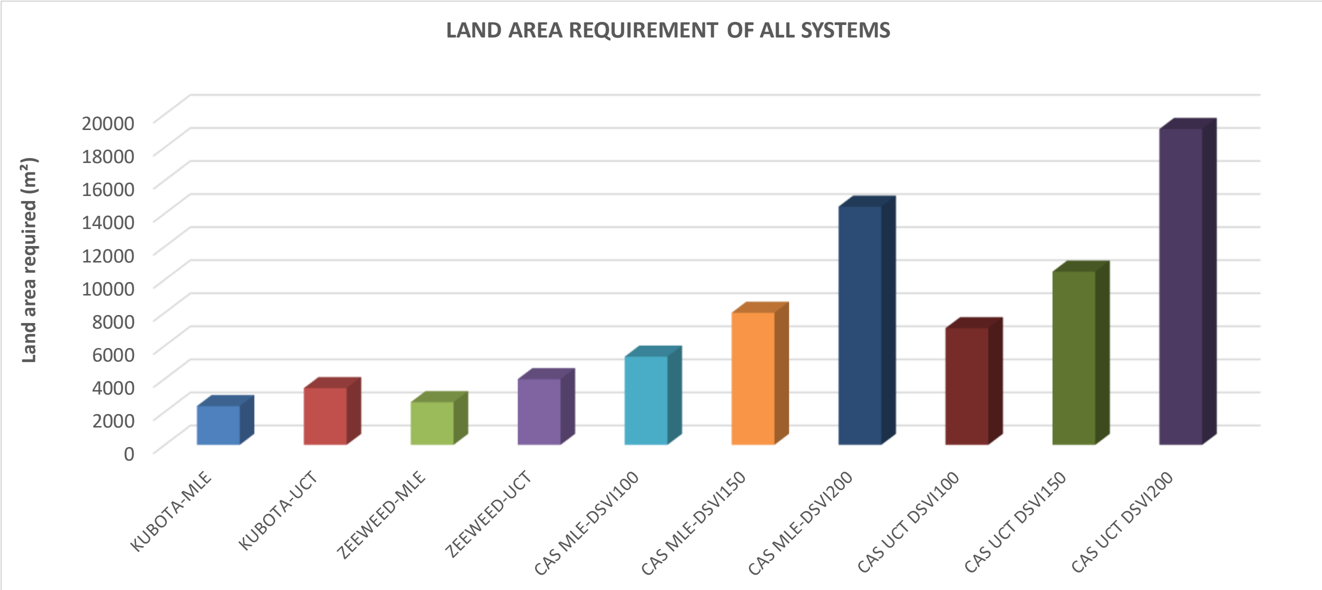

Figure 12-4 – *Land area requirement of all systems*

The systems with the least amount of land requirement are the MBR systems. This was expected. The CAS MLE systems have a 127-514% higher land requirement than the Kubota MBR MLE at various DSVI values. The CAS UCT systems have a 201-714% more land requirement than the Kubota MBR MLE system.

13 Conclusions and Recommendations

CAS systems have been successfully used to treat wastewater for more than 100 years. A drawback of CAS treatment is the large space required when compared to MBRs, where no SSTs are required and reactor volume requirement is smaller due to the increased MLSS concentration. Since its implementation, MBRs have been gaining much popularity around the world due to its smaller footprint requirement and better quality effluent produced. A disadvantage of MBR systems are the higher costs associated with the membranes and aeration system.

The main objective of this investigation was to economically compare the CAS system to an MBR in terms of CAPEX and OPEX, which could provide a more detailed extent of the cost difference between the 2 systems.

2 types of MBR technology was included in the investigation, the Kubota FS MBR system, and the Zeeweed HF MBR system. As the design of a CAS is sensitive to sludge settleability, various DSVI values were looked at as part of the CAS system. Each system was configured in an MLE and UCT process. In summary, the following systems were included in this investigation:

- CAS in an MLE configuration with DSVI of 100,150 and 200
- CAS in a UCT configuration with DSVI of 100,150 and 200
- iMBR using FS membranes in an MLE configuration
- iMBR using FS membranes in a UCT configuration
- iMBR using HF membranes in an MLE configuration
- iMBR using HF membranes in a UCT configuration

The following conclusions were drawn from this investigation:

- The Kubota system required the smallest reactor volume of all of the systems thus having the lowest reactor CAPEX at 7 - 54% lower than the other systems. This is due to the fact than in the Kubota MBR, the membranes are located in the aerobic zone. This eliminates the need for a separate membrane tank or SST, in the case of CAS.

- CAS systems require less oxygen than MBR due to the additional aeration requirement for scouring. The system which required the most oxygen and which had the highest aeration CAPEX was the Kubota MBR. This is due to the higher airflow requirement of the FS membranes compared to the HF. The FS membranes require 0.76 $Nm^3/m^2.h$ for the single-story arrangement and 1.13 $Nm^3/m^2.h$ for the double story. This is more than double that required by the HF membranes of 0.3 $Nm^3/m^2.h$

- The system with the lowest CAPEX (reactor, membranes, SSTS, and aeration) was the CAS MLE systems. This was expected. The Zeeweed MBR MLE system had a 7% lower CAPEX than the CAS UCT systems which was unexpected. The systems with the highest CAPEX was the Kubota FS MBR. This was mainly due to the higher cost of the membranes compared to the Zeeweed membranes (R900m^2 > R700m^2) and the higher airflow required by the FS membranes for scouring. In both MBR systems, the cost of the membranes was the main contributor to the CAPEX, comprising 60-70%.

- The system with the lowest OPEX was the CAS MLE systems. This was expected. The Zeeweed MBR had a lower OPEX than the CAS UCT which was unexpected. The systems with the highest OPEX were the Kubota systems which were mainly due to the high airflow requirement of the FS membranes.

- As the OPEX is such a huge contributor to the total cost, the total system cost followed the same ranking as the OPEX where the CAS MLE systems were the lowest, followed by the Zeeweed MBR MLE, followed by the CAS UCT, followed by the Kubota systems which had the highest total cost.

- The systems with the least amount of land requirement were the MBR systems. The CAS MLE systems have a 127-514% higher land requirement than the Kubota MBR MLE at various DSVI values. The CAS UCT systems have a 201-714% more land requirement than the Kubota MBR MLE system.

From the results of this investigation, CAS treatment has a lower CAPEX, OPEX and total cost than MBR. Even though only 2 MBR technologies were considered in this investigation, it can be stated with relative confidence that other MBR technologies would not have yield results which would be very different. It can be safely said that MBR is still considerably more expensive than CAS. As the majority of the MBR CAPEX consist of the cost of the membranes, the membrane cost would have to decrease by 40-50% to be equal to the CAPEX of CAS. The impact of including the cost of land in the comparison would have increased the total cost of the CAS. It was decided not to include this in this investigation to confine the scope.

14 Reference list

Bunani, S. *et al.* (2015) 'Application of reverse osmosis for reuse of secondary treated urban wastewater in agricultural irrigation', *Desalination*, 364, pp. 68–74. doi: 10.1016/j.desal.2014.07.030.

Ekama, GA. Wentzel, M. (2008) 'Organic matter removal', in *Biological Waste Water Treatment: Principles, Modelling and Design*. IWA Publishing, pp. 53–86.

Gass, S. I. (1983) 'What is a computer based mathematical model?', *Mathematical Modelling*, 4(5), pp. 467–472.

Germain, E. and Stephenson, T. (2014) 'Biomass Characteristics , Aeration and Oxygen Transfer in Membrane Bioreactors : Their Interrelations Explained by a Review of Aerobic Biological Processes Biomass characteristics , aeration and oxygen transfer in membrane bioreactors : Their interrelatio', (July). doi: 10.1007/s11157-005-2097-3.

Henze M. van Loosdrecht, M. Ekama, G.A. Brdanovic, D. (2008) *Biological Waste Water Treatment*.

Iorhemen, O. T., Hamza, R. A. and Tay, J. H. (2016) 'Membrane bioreactor (Mbr) technology for wastewater treatment and reclamation: Membrane fouling', *Membranes*, 6(2), pp. 13–16. doi: 10.3390/membranes6020033.

Jiang, S., Li, Y. and Ladewig, B. P. (2017) 'A review of reverse osmosis membrane fouling and control strategies', *Science of the Total Environment*. Elsevier B.V., 595, pp. 567–583. doi: 10.1016/j.scitotenv.2017.03.235.

Judd, S. Kim, B. Amy, G. (2008) 'Membrane Bio-reactors', in *Biological Waste Water Treatment: Principles, Modelling and Design*. IWA Publishing, pp. 335–360.

Judd, S. (2011) *The MBR Book*. 2nd edn. Edited by S. Judd and C. Judd. Elsevier.

Le-Clech, P. Jefferson, B. Judd, S. (2005) 'A comparison of submerged and sidestream tubular membrane bioreactor configurations', *Desalination*, 173(2), pp. 113–122. doi: 10.1016/j.

Parco, V., du Toit, G. and Ekama, G. (2018) 'Chapter 2 Advances in wastewater treatment', in Mannina, G. et al. (eds) *Biological nutrient removal activated sludge systems with membranes*. London: IWA Publishing.

Ramphao, M. Wentzel, M.C., Merritt, R. Ekama, G.A, Young, T. and Buckley, C. A. (2006) 'The impact of membrane solid liquid separation on the design of biological nutrient removal activated sludge systems', *Production*, (May), pp. 1–13.

'Ramphao(2005)ImpactofMBRB&B89(6)632-646.pdf' (no date).

Ramphao, M. C. *et al.* (2013) 'The Impact Membrane Bioreactor on thedesign of biological nutrient removal at Malmesbury WWTW', *Imesa*. doi: 10.1175/1520-0426(2001)018<0215:COUASA>2.0.CO;2.

Singhania, R. R. *et al.* (2012) 'Immersed membrane bioreactors: An overview with special emphasis on anaerobic bioprocesses', *Bioresource Technology*. Elsevier Ltd, 122, pp. 171–180. doi: 10.1016/j.biortech.2012.01.132.

SUEZ (2019) 'No Title'.

Takacs, I. and Ekama, G. A. (2008) 'Final settling', in *Biological Waste Water Treatment: Principles,* *Modelling and Design*, pp. 309–333.

Du Toit, G. (2006) *Design and Performance of Bnractivated Sludge Systems With Flat Sheet Membranes for Solid-Liquid Separation*, *Uct*.

du Toit, G. J. G. *et al.* (2010) *The Performance and Kinetics of Biological Nitrogen and Phosphorus Removal with Ultra-Filtration Membranes for Solid-Liquid Separation. Final WRC Report for Projects K8/814 and K5/1537*. Private Bag X03, Gezina 0031, Pretoria, South Africa.

Union, I., Pure, O. F. and Chemistry, A. (1996) 'Terminology for membranes and membrane processes (IUPAC Recommendation 1996)', *Journal of Membrane Science*, 120(2), pp. 149–159. doi: 10.1016/0376-7388(96)82861-4.

Waste water engineering: Treatment and Reuse (2004). Metcalf & Eddy.

WEF (2011) *Membrane Bioreactors*.

Wilf, M. (2008) 'Membrane Types and Factors Affecting Membrane Performance', in *Membrane types and factors affecting membrane performance*, pp. 1–92.

WRC (2002) *Theory, design and operation of nutrient removal activated sludge processes*, *Water SA*. Water Research Commission.

Yigit, N. O. *et al.* (2008) 'Membrane fouling in a pilot-scale submerged membrane bioreactor operated under various conditions', *Desalination*, 231(1–3), pp. 124–132. doi: 10.1016/j.desal.2007.11.041.

APPENDIX A :

Wastewater Characterization Block Diagrams

		COD - RAW WW		
			$f_{S'up}$	0,13
	TOTAL		$f_{S'us}$	0,06
	800		$f_{Sb's}$	0,15
			$f_{S'bs}$	0,07
VFA	**FBSO**	**USO**		
0	55	45	100 SOLUBLE	
	BPO	**UPO**		
	599	101	700 PARTICULATE	
	BIO	654 UNBIO	146 $S_{upi} = f_{s'up} . S_{ti}$	

		COD	TKN	TOC	TP
PARTICULATE	**Settleable**	46,88%	16,67%	36,11%	11,83%
	Non-settleable	40,63%	50,00%	43,13%	12,28%
SOLUBLE	**Dissolved**	12,50%	33,33%	20,76%	75,89%
	Total	100,00%	100,00%	100,00%	100,00%

(RAW)

		COD	TKN	TOC	TP
		mg/l	mg/l	mg/l	mg/l
BPO	Settleable	321	8	147	1,3
	Non-settleable	278	25	176	1,3
UPO	Settleable	54	2	16	0,8
	Non-settleable	47	5	19	0,9

Table: Subdivision of BPO and UPO into Settleable and Non-settleable concentrations

APPENDIX B :

Assessment of Ethics in Research Projects form

APPLICATION FORM

Please Note:

Any person planning to undertake research in the Faculty of Engineering and the Built Environment (EBE) at the University of Cape Town is required to complete this form before collecting or analysing data. The objective of submitting this application *prior* to embarking on research is to ensure that the highest ethical standards in research, conducted under the auspices of the EBE Faculty, are met. Please ensure that you have read, and understood the **EBE Ethics in Research Handbook** (available from the UCT EBE, Research Ethics website) prior to completing this application form: http://www.ebe.uct.ac.za/ebe/research/ethics1

APPLICANT'S DETAILS		
Name of principal researcher, student or external applicant		Delwin Smith
Department		Civil Engineering
Preferred email address of applicant:		Delwinwsmith@gmail.com
If Student	Your Degree: e.g., MSc, PhD, etc.	MEng
	Credit Value of Research: e.g., 60/120/180/360 etc.	60
	Name of Supervisor (if supervised):	Prof. George Ekama
If this is a researchcontract, indicate the source of funding/sponsorship		n/a
Project Title		An economical comparison of treating Waste water using Conventional Activated Sludge and MBR

I hereby undertake to carry out my research in such a way that:

- there is no apparent legal objection to the nature or the method of research; and
- the research will not compromise staff or students or the other responsibilities of the University;
- the stated objective will be achieved, and the findings will have a high degree of validity;
- limitations and alternative interpretations will be considered;
- the findings could be subject to peer review and publicly available; and
- I will comply with the conventions of copyright and avoid any practice that would constitute plagiarism.

SIGNED BY	Full name	Signature	Date
Principal Researcher/ Student/External applicant	Delwin Smith	Signature Removed	27 Feb 2019

APPLICATION APPROVED BY	Full name	Signature	Date
Supervisor (where applicable)	Prof. George Ekama	Signature Removed	04 Mar 2019
HOD (or delegated nominee) Final authority for all applicants who have answered NO to all questions in Section1; and for all Undergraduateresearch (Including Honours).	Dyllon Randall Click here to enter text.	Signature Removed	27.03.2019 Click here to enter a date.
Chair : Faculty EIR Committee For applicants other than undergraduate students who have answered YES to any of the above questions.			