Cost and Performance Comparison of a Membrane Bioreactor (MBR) Plant and a Bardenpho Plant for Wastewater Treatment

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Abstract

Rotorua District Council constructed a membrane bioreactor (MBR) system to work in parallel with the existing Bardenpho system, to improve the overall capacity and removal efficiency of the wastewater treatment plant. The aim of this research was to examine and compare the performance of the MBR and the Bardenpho systems, as well as the operating costs and the net present value. The MBR showed a higher removal rate of TKN, TSS, COD, DRP and organic nitrogen than the Bardenpho system, with average removal efficiencies of 94.52%, 99.05%, 94.04%, 67.43% and 92.75% respectively. These removal rates were shown to be 0.62%, 15.47%, 8.51%, 5.72% and14.74 % better than the Bardenpho system. Furthermore, the MBR removal rate remained stable over a range of flows, while the Bardenpho removal rate fluctuated with a change in flow. The total energy consumption of the treated wastewater in the MBR and the Bardenpho systems were 0.53 kWh/m^3 and 0.25 kWh/m^3 respectively. Major contributors to the total operational costs were the disposal cost of sludge, the cost of ethanol and aeration. In the MBR, the cost percentage of sludge disposal was 22.11%, and the total aeration cost was 20.25%, while the cost of disposal of sludge was 45.15% and aeration 7.22% in the Bardenpho. The capital cost to construct the MBR and the Bardenpho systems was calculated to be similar when the cost of land is not taken into account, however, the size of the Bardenpho required to treat the same volume of wastewater is found to be 2.2 times bigger than a MBR system. It was concluded that from a cost point of view, the Bardenpho system is more economical; however, the low cost of running the Bardenpho comes with lower removal efficiencies in comparison to the MBR.

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Chapter 1: Introduction

1.1 Background

In Rotorua the wastewater collection system receives an average flow of approximately 19.2 million liters a day (ML/d) produced by a population of approximately 68,000 people. Future (2051) flows are estimated to be 24.0 ML/d for a predicted population of 75,000. The majority of the waste comes from residential and commercial properties, with a small amount coming from the industrial areas of the city. This wastewater is treated, before disposal, using both an advanced activated sludge system and a recently installed membrane bioreactor (MBR) system.

Untreated wastewater generally contains pathogenic microorganisms, nutrients, and toxins that can affect humans and the environment. The wastewater is characterised in terms of its physical, chemical and biological composition. The important constituents of concern in wastewater treatment are:

- Nutrients (nitrogen and phosphorus)
- Biodegradable organics
- Suspended solids
- Pathogens
- Heavy metals

Nitrogen (N) and phosphorus (P) are essential nutrients for cell growth. Disposal of these nutrients in waterways can lead to algal blooms, invasive weeds and loss of aquatic diversity and amenity value. Biodegradable organics are mainly measured as five day biochemical oxygen demand (BOD_5) or chemical oxygen demand (COD), originating in the wastewater as proteins, carbohydrates and fats. High levels of biodegradable organics if discharged can reduce oxygen levels in waterways and develop septic conditions. Suspended solids when discharged in an aquatic environment can lead to the development of sludge deposits and anaerobic conditions. Exposure to pathogens in an area in which wastewater is disposed can lead to a variety of diseases such as gastroenteritis, diarrhea, and skin infections. When wastewater sources are from commercial and industrial activities, heavy metals can be present if not treated. Heavy metals in the effluent will affect the reuse capability of the wastewater.

In Rotorua the raw wastewater concentrations entering the plant has an average level of 48.00 mg/l of total Kjeldahl nitrogen (TKN), 204.02 mg/l of biodegradable organics (BOD_5), 455.25 mg/l of chemical oxygen demand (COD) and 383.59 mg/l of total suspended solids (TSS) (Table 1).

Table 1: Rotorua raw wastewater characteristics

Parameters	**Average Level (mg/L)**
Total Phosphorus	6.05
Total Kjeldahl nitrogen (org N + NH_4^+)	48.00
Biodegradable organics (BOD_5)	204.02
Chemical oxygen demand (COD)	455.25
Total Suspended Solids	383.59
pH	7.18

Data was taken from RDC WWTP Laboratory result sheet from the 11/01/2012 until 28/12/2012

1.2 Wastewater regulations

Wastewater regulations are in place to reduce the impact of wastewater on public health and the environment. The objective of the regulations is to reduce the impact of degradable organics, total suspended solids (TSS), nutrients such as phosphorus and nitrogen and pathogenic organisms on the environment (Langdon, 2011).

Various bodies who work to protect surface water against trophication have published water quality standards for phosphorus and nitrogen compounds. The American Environmental Protection Agency (EPA) published nutrient quality criteria for disposal of nutrients in streams and rivers, marked as total phosphorus ranges from 0.01 to 0.076 mg/l and total nitrogen from 0.12 to 2.18 mg/l, dependent on the environment (Environmental Protection Agency, 2000). The European guidelines recommend treated effluent levels for total phosphorus of 1 to 2 mg/l and total nitrogen of 10 to 15 mg/l.

In New Zealand the level of wastewater treatment and disposal is controlled by the Resource Management Act 1991 (RMA). Under the RMA it is necessary to obtain resource consent before it is permitted to discharge wastewater. Wastewater treatment plants are required to go through a consent application process. During the consent process the regional authority will determine the level of treatment and disposal case by case. These levels will then be enforced by law (New Zealand Legislation, 2012). Current wastewater legislation has become more restrictive and requires improvements in treatment technologies that are capable of producing higher effluent quality and have a minimum impact on the environment

Under the RMA, the Rotorua District Council (RDC) is required to apply for discharge consent from the Bay of Plenty Regional Council (BOPRC). The current treated effluent discharge resource consent allows RDC to discharge the secondary treated effluent on to the land in the Whakarewarewa forest using irrigation sprinklers. The consent specifies the exact location of discharge, quantity and rates, forest operational requirements, monitoring requirements, nutrient limits (Table 2) and emergency situations.

Table 2: Rotorua District Council discharge nutrient limits

Constituent	Discharge nutrient mass in effluent (kg/day)	Discharge nutrient concentration in effluent (mg/l)[1]
Total phosphorus	8.33	0.34
Total nitrogen	83.33	3.47

[1] Considering future flow rate of 24 ML/d

The Whakarewarewa forest (Figure 1) acts as an additional treatment stage with nutrients taken up by the trees, and effluent volumes being reduced by evapotranspiration and plant and soil adsorption. Furthermore, organic content is degraded by the soil bacteria, salts are removed by ion exchange and particles by filtering action. The residue flow then finds its way to the Rotorua Lakes mainly via the Waipa stream.

Figure 1: Whakarewarewa forest irrigation

The Rotorua Lakes are important natural assets and maintaining their quality is very important. Therefore the level of treatment at the wastewater treatment plant and the disposal method needs to take into account that the effluent will eventually find its way into the Lakes.

To make sure compliance with the consent nutrient condition is achieved, RDC monitors the Waipa stream for any signs of contamination. Concentration values of total phosphorus and total nitrogen is monitored together with flow monitoring. The measurement of flow and concentration will then be calculated to make sure consent limits are met.

1.3 Rotorua wastewater treatment plant

In the late 1960's and early 1970's the urban area of Rotorua district was reticulated. The reticulation scheme was designed to transport the sewage to a central wastewater facility where treatment took place through an activated sludge process. The effluent was discharged to the Puarenga stream which flows out to the Rotorua Lake. Due to growing concern over the deterioration of the Rotorua Lake water quality in 1991, the treatment plant was upgraded to a Bardenpho process and irrigation as a means of disposal in the Whakarewarewa forest was started. At the same time a discharge resource consent was obtained from the regional council. Later it appeared that the system had insufficient treatment quality to comply with consent regulations leading to a second upgrade in 2005. Changes included improvements to the returned activated sludge (RAS) pumps, extension of the Bardenpho process, upgrading of the aeration system and methanol dosing. Despite these changes, findings showed the nutrient levels discharged into the lake were higher than stated in the resource consent and the capacity of this plant was insufficient. In 2009 a capacity study took place, recommending that in order to meet the population growth and the required levels of treatment quality the plant needed further upgrading. Therefore in 2012, Rotorua District Council (RDC) decided to install a submerged membrane bioreactor (MBR) system to supplement the existing activated sludge process (Figure 2).

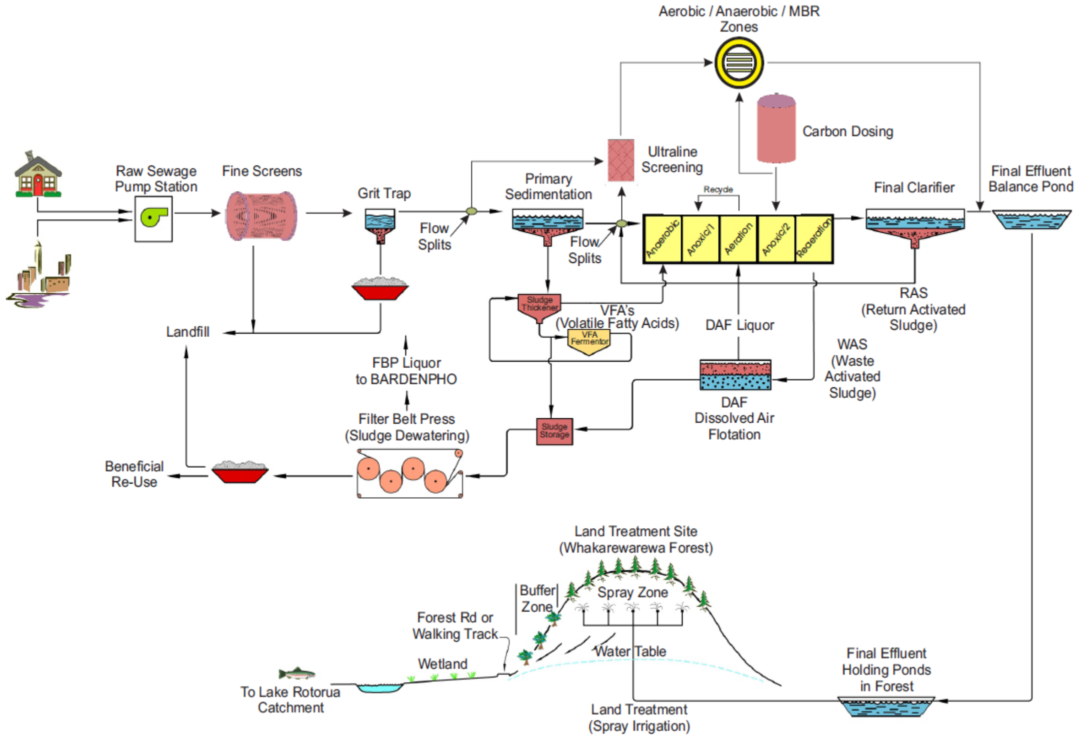


Figure 2: Process diagram of Rotorua wastewater treatment system

1.4 Research Problem

During peak flows and high rainfall events the wastewater inflow coming into the Rotorua wastewater plant will be treated by both the activated sludge process and the MBR, which will run at full capacities. However, when the inflow rates are lower the option to divert the influent to either of the Bardenpho activated sludge process or MBR is available. The impact of this choice of flow diversion between the Bardenpho and MBR on the overall performance and running costs of the plant will be investigated and compared in this study.

Chapter 2: Literature review

2.1 Introduction

The purpose of this chapter is to review the literature concerning membrane bioreactor (MBR) and the traditional wastewater treatment. At the end of this chapter the research questions for the present study will be outlined.

2.2 Background

Disposal of wastewater can cause major problems, impacting on the environment and human life. Wastewater composition and quality are dependent on the type of users and water consumption rates. For example wastewater generated by an area high in industrial activities will be different from an area with high domestic use. Table 3 shows typical contaminants and their concentration range in untreated domestic wastewater (Crites & Tchobanoglous, 1998; Metcalf & Eddy, 2003).

Table 3: Typical raw wastewater composition

Constituent	**Typical Concentration Range (mg/L)**
Total phosphorus	4-15
Total nitrogen	20-85
Chemical oxygen demand (COD)	250-1000
Biodegradable organics (BOD_5)	110-400
Total suspended solids	100-350

Typical data on the total quantities of waste discharged per person per day from individual residences are presented in Table 4.

Table 4: Typical wastewater concentrations from individual residences

Constituent	Concentration (mg/L)
Total phosphorus	17.3
Total Kjeldahl nitrogen	70.4
Chemical oxygen demand (COD)	1050
Biodegradable organics (BOD_5)	450
Total suspended solids	503

Concentration rate based of consumption of 189 l/person/day

2.3 Conventional Wastewater Treatment

Conventional large scale wastewater treatments normally include a combination of physical, chemical, and biological processes and operations to remove solids, organic matter and nutrients from wastewater. The treatment is normally divided into different stages of treatment. These stages can be preliminary, primary, secondary, and tertiary, ordered by increasing treatment level. A typical wastewater treatment process diagram is shown in Figure 3 (Langdon, 2011).

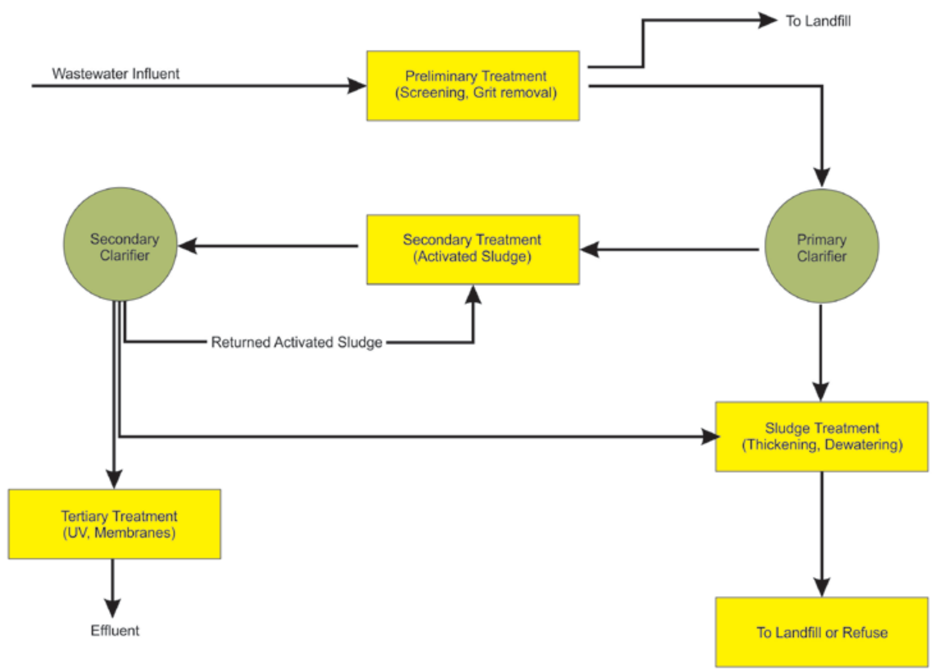


Figure 3: Typical wastewater treatment process diagram (Langdon, 2011)

2.3.1 Preliminary treatment

The main goal of the preliminary treatment is to protect treatment plant systems by removing components of wastewater that can cause damage to the plant facilities or interfere with the treatment process. Typically the preliminary treatment system will be designed to remove or reduce the following:

- Coarse solids and other large materials often found in raw wastewater, for example pieces of wood, plastic, cloth, paper.
- Large inorganic solids such as gravel, sand, glass and metal.
- Fat, oil and grease.

Preliminary treatment can include:

1. Screening: this process is usually mechanically operated to intercept floating and suspended debris. The screening elements can include parallel bars, rods or wires, grating, wire mesh or a perforated plate.
2. Grit removal: where there is centrifugal and/or gravitational separation of the inorganic grit particles. Grit chambers are generally located after the screen and before the primary treatment.
3. Removal of fat, oil and grease by skimming (Environmental Protection Agency, 1995; Metcalf & Eddy, 2003).

Treatment plants often include a flow equalisation mechanism to operate efficiently and reduce the cost and the size of the downstream treatment facilities. Flow equalisation is applied to the wastewater inflow in order to balance the flow and quality to achieve a flow pattern as uniform as possible (Crites & Tchobanoglous, 1998; Metcalf & Eddy, 2003). The flow equalisation mechanism normally includes a tank that is sized based on a mass balance analysis. The size of the tank is designed in such a way to allow withdrawal of wastewater during low influent rates and storage of excess wastewater during high influent rates.

2.3.2 Primary treatment

The main goal of the primary treatment is the removal of organic and inorganic solids by sedimentation, and the removal of materials that will float (scum) by skimming. Primary treatment is used as an initial step for the further processes to follow. Typical primary treatment will remove 25 to 50% of the biological oxygen demand (BOD_5), 50 to 70% of the total suspended solids (TSS), and 65% of the oil and grease (Metcalf & Eddy, 2003). Some organic nitrogen, organic phosphorus, and heavy metals associated with solids are also removed during primary sedimentation but colloidal and dissolved constituents are not affected.

The most commonly used method in wastewater primary treatment is sedimentation e.g. coagulation, flocculation and flotation, all of which normally take place in a clarifier.

2.3.3 Secondary treatment

The main goal of the secondary treatment is to remove the residual inorganic, organic and suspended solids remaining after the primary treatment (Giridhar, 2011). Normally secondary treatment uses aerobic biological treatment processes. Aerobic biological treatment is performed in the presence of oxygen, an electron donor (organic matter, proteins, carbohydrates, fats), and aerobic microorganisms that metabolise the organic matter in the wastewater. This biological process produces more microorganisms and inorganic end-products which are determined by the electron acceptor (O_2 CO_2, sulphur and oxidized N_2). Secondary treatment differs from primary treatment in that oxygen is supplied to the microorganisms and the rate at which organisms metabolize the organic matter (Giridhar, 2011).

The microorganisms, biomass, must be separated from the treated wastewater, normally by sedimentation, to produce a clarified secondary effluent. The sedimentation tanks (secondary clarifiers) used in secondary treatment, operate in the same basic manner as the primary clarifiers. The biomass removed during secondary sedimentation is normally combined with primary sludge for sludge processing (Metcalf & Eddy, 2003).

In a large scale operation an activated sludge process will be included as a secondary treatment.

2.3.3.1 Activated Sludge

In the activated sludge process, the microorganism’s growth reactor is an aeration tank containing a suspension of the wastewater and microorganisms, the mixed liquor. The liquor in the tank is aerated supplying oxygen to the microorganisms. Aeration devices commonly used include submerged diffusers that release compressed air and mechanical surface aerators that introduce air by agitating the liquid surface. Hydraulic retention time in the aeration tanks usually ranges from six to eight hours but can be higher with high organic content wastewaters. Following the aeration step, the microorganisms are separated from the liquid by sedimentation and the clarified liquid is secondary effluent. A portion of the biological sludge is recycled to the aeration basin to maintain a high mixed-liquor suspended solids (MLSS) level. The remainder is removed from the process and sent to sludge processing to maintain a relatively constant concentration of microorganisms in the system (Figure 4) (Giridhar, 2011; Metcalf & Eddy, 2003).

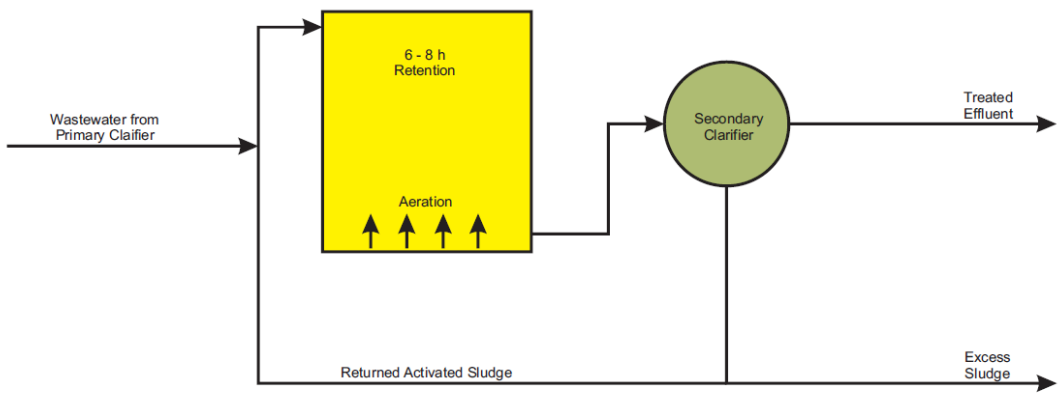


Figure 4: Typical activated sludge treatment process (Giridhar, 2011)

Some modern secondary treatment systems incorporate anaerobic, anoxic and aerobic stages in their design (Figure 5). These different stages are used for an improved removal of nitrogen and phosphorus. Nitrogen is removed through the processes of denitrification and nitrification, while phosphorus is more effectively removed by alternating anaerobic and aerobic conditions (Grady, Daigger, Love, & Filipe, 2011).

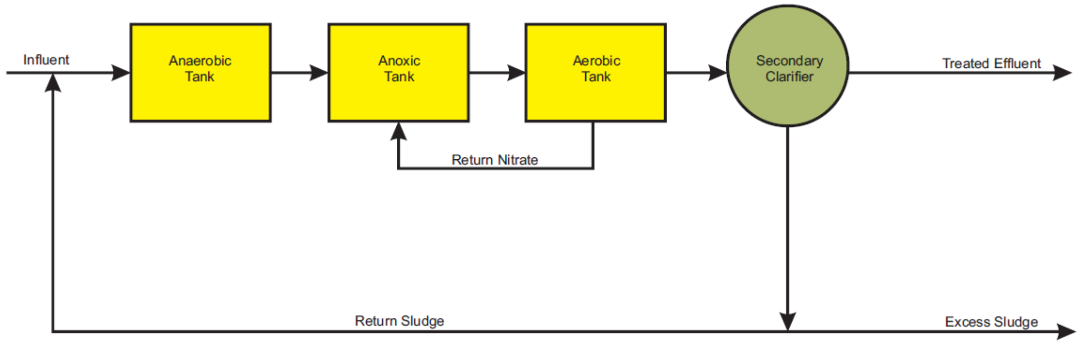


Figure 5: Multi stage secondary treatment

The common biological treatment processes is that they are divided into tanks containing different biochemical environments to remove nitrogen and/or phosphorus. Such an example for this process is the Bardenpho system, biological nitrogen and phosphorus removal includes an activated sludge system with five tanks in series, in which the first tank is anaerobic to encourage the growth of phosphorus storing bacteria followed by primary anoxic tank, an aeration tank, secondary anoxic tank and re-aeration tank. Removal of nitrate will take place by nitrification and denitrification systems in the aerobic and anoxic tanks, with a recycle stream to convert ammonia to nitrate and nitrate to nitrogen gas. The re-aeration tank ensure that the mixed liquor passed on to the final clarifier is well aerated and do not suffer oxygen depletion during retention in the clarifier to prevent release of phosphorus into the final effluent (Grady et al., 2011).

2.3.4 Tertiary treatment

Tertiary wastewater treatment is used when specific wastewater constituents which cannot be removed by secondary treatment are required to be removed. The purpose of tertiary treatment is to provide a final treatment stage to raise the effluent quality before it is discharged to the receiving environment, for example, the sea, rivers, lakes, or the ground (Langdon, 2011). More than one tertiary treatment process may be used. Tertiary treatment can include disinfection (UV, O_3), multistage biological removal, membrane separation and chemical removal of nutrients and heavy metals (Metcalf & Eddy, 2003).

2.4 Wastewater treatment around New Zealand

In New Zealand an increased understanding of the impact of inappropriate treatment has led to legislation which, in turn, has driven many different Councils to upgrade their level of treatment. For example, Hamilton City Council which disposes their treated effluent into the Waikato River upgraded their old plant in 2002. The upgrade valued at 28 million New Zealand dollars, included a new screen, an additional secondary treatment, digesters, ultraviolet disinfection, new congeneration plant and new dewatering system. Further upgrade work is planned to comply with the new resource consent and the increased capacities of the plant (Hamilton, 2012).

To be able to comply with their resource consent, Gisborne City Council upgraded their wastewater plant in December 2010. The new upgraded system, valued at 39.5 million New Zealand dollars, includes a biological trickling system which enables an advanced treatment to remove the solids and nutrients (Gisborne, 2012).

Taupo District Council also upgraded their existing plant built in 1972. The upgrade in 2010, included an additional (third) primary sedimentation tank, a trickling filter tower, a second thickener, a third digester and UV disinfection. A further future upgrade is planned and will include an additional trickling filter tower, a second secondary clarifier and tertiary filtration (Taupo, 2012).

Rosdale wastewater treatment plant in Auckland is amongst the largest in New Zealand, built in 1951. The regional authorities gave the city council 10 years to apply for resource consent to continue operation. Upgrade work started in 1995 and was completed in 2008. Part of the upgrade work included an activated sludge system, dewatering facility, rotating screen and UV treatment (North-Shore, 2012).

Table 5 shows a comparison of the level of treatment, the final effluent qualities and the type of disposal in different wastewater treatment plants around New Zealand (NZ).

Table 5: Sample of treatment plants around NZ

Plant location	Average daily volume (m3)	Method of treatment	Constituent	Average influent concentration (mg/l)	Average effluent concentration (mg/l)
Hamilton[1]	40000	Activated sludge and UV	COD	665	50.1
			Total-P	9.0	2.8
			Total-N	52.1	12.5
			NH_4-N	28.1	1.9
			TSS	-	15.2
Gisborne[2]	17634	Biological trickling filter	TSS	275.7	180.2
			BOD_5	283.1	91.7
			COD	502.9	264.6
			Total-P	5.1	4.3
			Total-N	17	13
Rotorua	19200	Activated sludge and MBR	COD	449.5	41.9
			Total-P	6.0	2.3
			TKN	48.6	3.2
			TSS	387.3	21.5
			TOXN	-	2.32
Wellington Hutt city[3]	56000	Secondary treatment and UV	BOD_5	-	12.4
			Tot-P	-	2.9
			TOXN	-	1.9
Tauranga[4]	2700	Activated sludge and UV	Tot-P	736	-
			TKN	-	7
			NH_4-N	57	18
			BOD_5	42	17
			TSS	312	9
			COD	361	19

[1] (D. Hight, personal communication, February 15, 2013)

[2] (R. Long , personal communication, February 12, 2013)

[3] (C. Low, personal communication, February 18, 2013)

[4] (R.Wiersma , personal communication, February 11, 2013)|

2.4.1 Cultural aspects

The final step in the wastewater treatment is the return of the treated wastewater (effluent) into the environment. Generally there are two main environmental recipients: one is surface water which can include rivers, lakes, and the sea and the second is ground water (Langdon, 2011). In New Zealand indirect disposal to ground water or surface water via land is a favored option because the land provides a certain degree of additional treatment and better meets the Maori cultural values. Land disposal systems can include wetlands and forest spray irrigation which can, in addition to the disposal, provide amenity and commercial value.

In most areas around New Zealand the level of treatment and means of disposal also needs to satisfy the cultural beliefs of the local Maori. The RMA requires that local authorities (regional, city and district councils) objectives, policies and methods "achieve integrated management of the effects of natural and physical resources" of the region or district (New Zealand Legislation, 2012). The Maori perspective on the management of natural resources and the disposal of human waste into natural resources is a matter of great significance. When looking at options for effluent disposal, protection of water bodies and sacred places to the Maori should be considered. A common way to protect cultural beliefs is to ensure the effluent comes into contact with land before entering a water body. This land contact arrangement can be included as a final treatment stage where treated human wastewater contacts Papatuanuku (earth mother) before being discharged into a natural water body. This stage is rather providing a spiritual cleansing than an additional treatment stage. A number of case histories suggest a range of Papatuanuku arrangements to spiritually treat the wastewater before discharging to water (Bradley, 2012, August). Some of the Papatuanuku types used in parts around New Zealand include wetlands, rock passages, riparian strips, overland flow, rapid infiltration and enhanced treatments (Table 6).

Table 6: Papatuanuku types installed around New Zealand

Papatuanuku type	Location
Wetland	Cambridge, Palmerston North, Tauranga, Whangarei
Rock Passage	Hastings, Te Awamutu, Taumaranui, Waikato
Riparian Zones	Te Puke
Overland Flow	Oamaru
Gravel Beds	Ngaruawahia, Huntly, Otorohanga
Rapid Infiltration Beds	Motueka, Takaka, Cambridge

(Bradley, 2012, August)

2.5 Biological-Wastewater Treatment

2.5.1 Introduction

Large scale wastewater treatment plants normally include some form of biological treatment to treat the waste. In the biological treatment of wastewater, organic and inorganic matter is converted by microorganisms into products that are easily removed from the water or made into less-harmful substances and in the process increase the total biomass. In this way, the removal of nitrogen, phosphorus, organic matter and solids is possible. The conversion of harmful substances by biological bacteria is possible in the treatment process. The type of bacteria that accumulates in a process will be determined by the specific conditions, such as presence of oxygen, pH, and temperature. These conditions will be created according to the compounds that are required to be removed (Grady et al., 2011).

2.5.2 Microbiology

Microorganisms play a big part in the removal of nutrients and organic matter. This removal process is part of the microorganisms' metabolism and includes all of the biochemical reactions that occur in the cell. Energy is required for the biochemical reactions to occur in the cell. Energy is released in the cell when oxidizing organic or inorganic matter, the energy produced is stored by organic compounds e.g. adenosine triphosphate (ATP) and then used for cell synthesis in the breakdown of organic or inorganic matter. In addition to the organic matter, nutrients such as nitrogen for protein and phosphorus for ATP are important in the growth of bacteria (Crites & Tchobanoglous, 1998).

Bacteria grow by binary fission where the original cell becomes two organisms and the two become four and so on. The time required for each cycle growth is dependent on the environmental conditions, environmental limitations such as the availability of nutrients, organic matter, and pH. The growth rate of each population of microorganisms will normally follow a curve pattern (Figure 6).

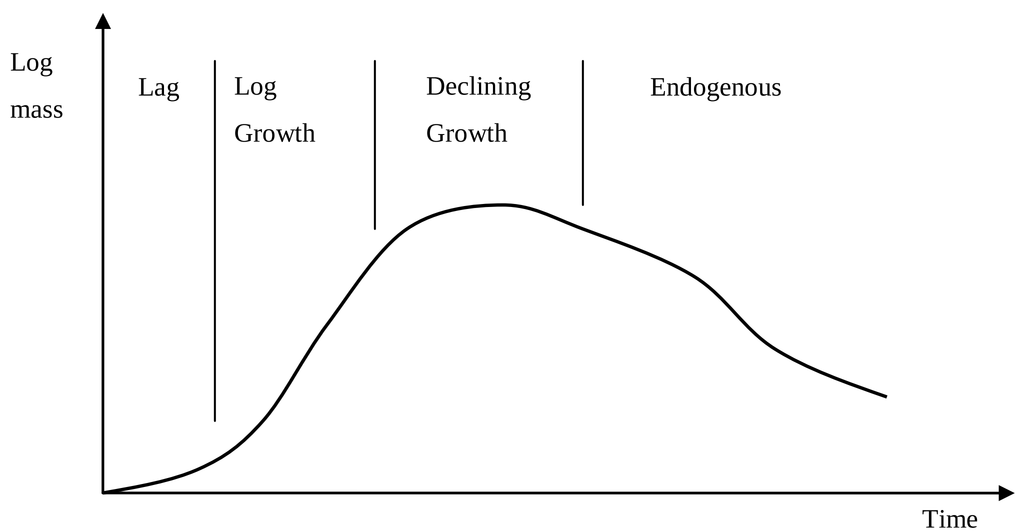


Figure 6: Single population of microorganism's growth curve

The biomass growth curve normally has four phases: lag, log growth, declining growth and endogenous. In the lag phase the microorganisms adjust to the new environment and the effluent. In some cases when the microorganisms are already acclimatised to effluent the lag phase can be very short or not exist. Such cases can be achieved during continuous operation of a wastewater treatment plant when nutrients are continuously fed to the tank and bacteria biomass is recycled.

When the microorganisms have adjusted, the log growth period begins and the organisms' biomass increases at a rate determined by their generation time and ability to metabolise the degradable organic material. As the biomass increases, the substrate is consumed. Eventually most of the substrate will be consumed and the declining growth phase will occur. During the wastewater treatment process when an external source of carbon and nutrients are added to the system and removal of biomass cell has taken place, the growth curve will be constant.

During the endogenous stage the organisms' death rate exceeds the production of new cells. The final stage is particularly important to determine the volume of mass of microorganisms (sludge) created in the process.

During wastewater treatment processes the biological population is mixed, with each particular microorganism in the system having its own curve. The position and shape of each growth curve will be dependent on the food available and on environmental conditions, such as temperature, oxygen level and pH. The bacteria has an important role in the wastewater treatment, many other different microorganisms can take part in the stabilisation of the organic waste (Crites & Tchobanoglous, 1998).

2.5.3 Biological nitrogen removal

Most of the available nitrogen in municipal wastewater is in the form of organic or ammonia nitrogen. During the wastewater treatment, about 20 percent of the total nitrogen settles out during primary sedimentation. The biological treatment, a major portion of the organic nitrogen is converted to ammonia nitrogen which is removed by the nitrification/denitrification processes (Crites & Tchobanoglous, 1998).

2.5.3.1 Nitrification

Nitrification is a microbiological two-step oxidation. In the first stage ammonium (NH_4^+) is oxidized to nitrite NO_2 (Equation 1) by Nitrosomonas bacteria. In the second faster stage the nitrite will be converted into nitrate NO3 by Nitrobacter (Equation 2).

(1) $2NH_4^{\ +} + 3O_2 \rightarrow 2NO_2^{-} + 2H_2O + 4H^+$

(2) $2NO_2^{-} + O_2 \rightarrow 2NO_3^{-}$

Both Nitrosomonas and Nitrobacter are autotrophic. Oxidation of ammonia and nitrite provides the energy that is needed for the cell synthesis. These bacteria are obligate aerobes. As autotrophic ammonia-oxidizing bacteria are generally characterized by low growth rates and yields, nitrification is commonly a rate-limiting step in the biological nitrogen removal process. Therefore normally the nitrification tank (aerobic tank) will allow higher retention times and maintain an adequate level of nitrifiers (Metcalf & Eddy, 2003).

2.5.3.2 Denitrification

Denitrification is applied when complete nitrogen removal is required. Denitrification is the dissimilative reduction of nitrate to molecular nitrogen. Four steps are involved in biological denitrification and the following intermediates are produced (Equation 3).

(3) $NO_3^{-} \rightarrow NO_2^{-} \rightarrow NO \rightarrow N_2O \rightarrow N2$

The bacteria responsible for the denitrification process are widespread among heterotrophic and autotrophic bacteria. The heterotrophic and autotrophic bacteria can shift between oxygen respiration and nitrogen respiration. Either inorganic or organic waste compounds can serve as substrate for denitrification where the electron donor is oxidized and nitrate is reduced in the absence of dissolved oxygen (DO) or limited DO concentrations.

An important design parameter for denitrification processes is the amount of degradable organic matter that is needed to provide a sufficient amount of electron donors for nitrate removal. Often an external carbon source, an easily degradable substrate such as ethanol or methanol, is used to increase denitrification rates, rather than a substrate such as raw waste water that is harder to degrade (Metcalf & Eddy, 2003).

2.5.3.3 Phosphorus biological removal

Biological phosphate removal from wastewater can be achieved by enhanced storage in the biomass as polyphosphate. The key to enhanced biological phosphorus removal (EBPR) is the activity of certain bacteria that take up phosphorus under aerobic or anoxic conditions, accumulating it within the cell. Alternating between anaerobic and aerobic conditions promote the growth of the polyphosphate organisms, which are able to store more phosphorus than they need for growth (Powell, 2009). Galil, Malachi and Sheindorf (2008) showed that by adding an anaerobic cell at the beginning of the biological treatment, improved removal of total phosphorus by 38%.

The removal of phosphorus can be achieved with the use of chemicals such as lime, ferric chloride and aluminum sulphate (Alum). At the right pH, the phosphorus will combine with the chemical which then can be removed by sedimentation (Metcalf & Eddy, 2003). This kind of chemical intervention is undesirable for various reasons, including effluent salinification, higher sludge mass and volume, high cost and impact on the effluent quality.

2.5.3.4 Biodegradable organics removal

The removal of dissolved and particulate carbonaceous material and the stabilisation of organic matter found in wastewater are accomplished biologically, using a variety of microorganisms, mainly bacteria. The main substrate, organic carbon, is converted into simple end products (CO_2) and additional biomass (Equation 4).

(4) $\text{Organic} - C + O_2 \rightarrow \text{Biomass} + CO_2 + \text{energy}$

The relative yields of biomass and CO_2 depend on the environmental conditions and the microorganisms' level, for example at a slow growth rate, microorganisms utilise available substrates mainly for maintenance purposes and will not consume high levels of organic matter. In the MBR, high levels of biodegradable organics removal will occur due to the higher number of microorganisms and increased retention time, this leads to better degradation in a given time span (Judd, 2011).

2.6 Membrane bioreactor technology

2.6.1 Introduction

The use of membrane bioreactors (MBR) for wastewater treatment was introduced in the late 1960's to early 1970's. (Judd, 2011; Leiknes, 2010). Since then MBR treatment has been developed and an increasing number of MBR treatment plants are being built (Judd, 2007; Kraume, Bracklow, Vocks, & Drews, 2005; Z. Wang et al., 2008).

The MBR includes a combination of biological treatment and membrane separation. The membrane acts like a filter, rejecting anything bigger than the membrane pores to provide a clarified final product in the permeate.

A basic diagram of a common MBR system is presented in Figure 7.

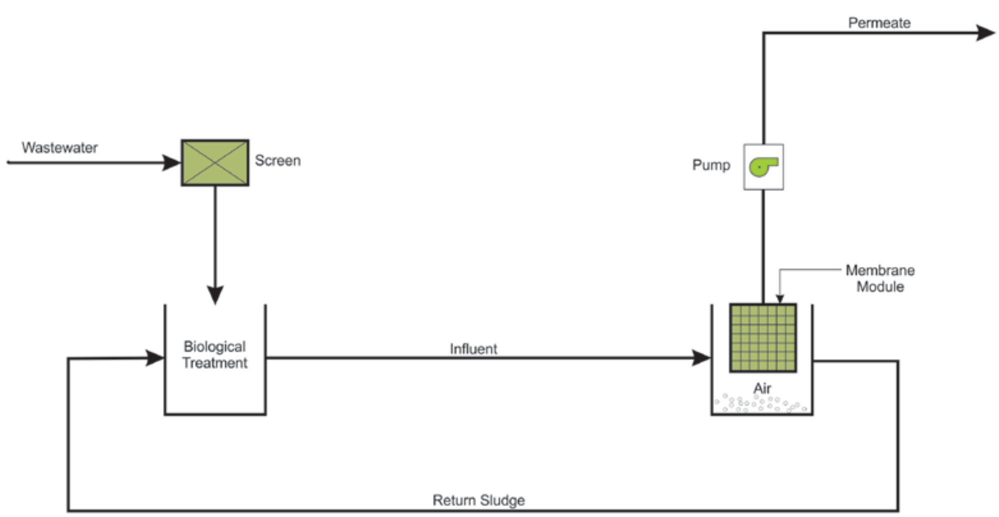
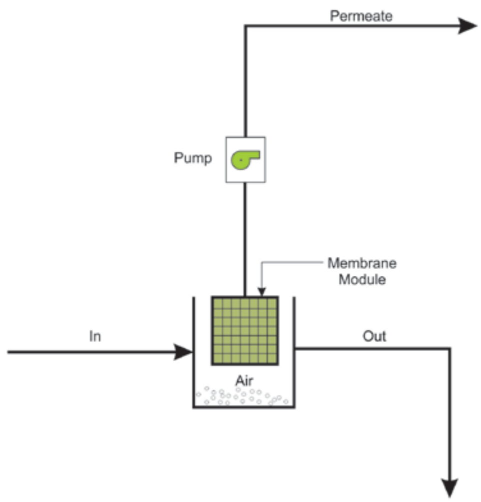
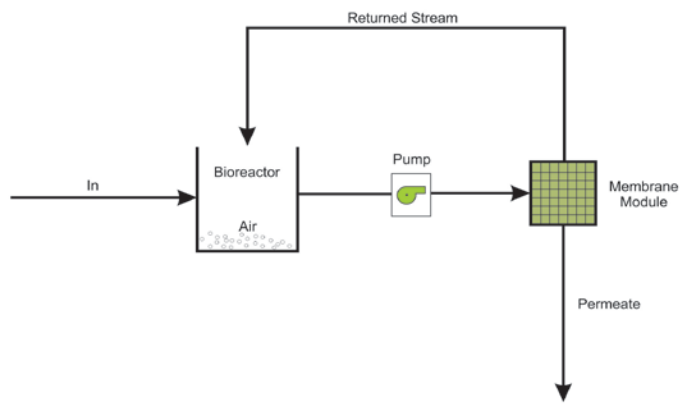


Figure 7: Basic process diagram of a typical MBR system

Generally there are two types of MBR configurations, the first where the membrane filtration occurs within the bioreactor (immersed MBR, Fig 8a) and the second where the filtration occurs externally through recirculation (sidestream MBR, Fig 8b).



(a)



(b)

Figure 8: Types of MBR configuration: (a) Submerged (b) Side stream

In sidestream membrane (sMBR), the feed enters the bioreactor from the treatment plant and then is pumped around a recirculation loop that contains a membrane unit. A portion of the flow will go through the membrane, while the rest is recycled to the bioreactor. Normally the sMBR arrangement is considered to be more costly due to the higher pumping costs (Judd, 2011; Leiknes, 2010).

2.6.2 Membrane Bioreactor biological process

The biological principles in the MBR reactor are similar to those of the conventional treatments. Removal of nutrients from wastewater occurs during cell growth and metabolism, nitrification and denitrification, and enhanced biological phosphorus removal.

The MBR process allows accumulation of slow growing microorganisms, high biomass concentration (Kraume & Drews, 2010), high amounts of predators such as protozoa and metazoan (Ghyoot & Verstraete, 1999), lower sludge production (Ghyoot & Verstraete, 1999), very high sludge volume index (N. I. Galil & Jacob, 2009) and better nitrification rates (Kraume & Drews, 2010; Munz, Gualtiero, Salvadori, Claudia, & Claudio, 2008). However, research by Ghyoot and Verstraete (1999) suggested grazing of the protozoa and metazoan on the nitrifying bacteria can possibly lead to a significant reduction of the nitrifying capacities of the MBR plant.

Sludge age can impact on the rate of the nitrification and denitrification in the MBR. Research by Murat Hocaoglu, Insel, Ubay Cokgor and Orhon (2011) testing the effect of sludge age on simultaneous nitrification and denitrification in a membrane bioreactor treating black water showed increasing sludge age from 20 days to 60 days increased the removal efficiency of the total nitrogen from 49% to 73% respectively. The study was done under low dissolved oxygen concentration of 0.1 mg/l to 0.2 mg/l with influent rich in nitrogen and low in COD/nitrogen ratio. The results indicate a reduction of 40% in nitrification rate and almost 100% denitrification rate. However, a study done by Galil and Jacob (2009) using a pilot plant MBR with three stage anaerobic anoxic and aerobic system showed different results, where a decrease in sludge age from 29 days to 16 days increased the removal efficiencies of total nitrogen from 89% to 91%.

Similar to the activated sludge processes in section 2.3.3.1, to improve treatment quality the biological section in the MBR can contain different tanks, where each tank can have different conditions (Figure 9).

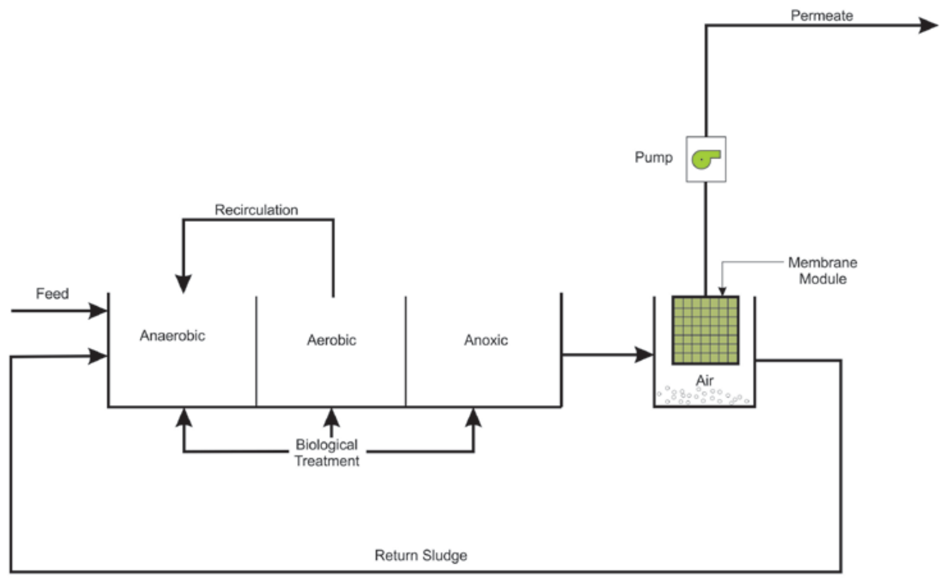


Figure 9: Typical large scale MBR process diagrams

The most common arrangement for biological nitrogen removal in wastewater treatment is the pre-denitrification MBR (Kraume & Drews, 2010). This system consists of an anaerobic and/or anoxic tank followed by an aeration tank where nitrification occurs. Nitrate produced in the aeration tank is recycled into the anoxic tank for denitrification. The organic matter in the wastewater provides the electron donor that is needed for denitrification. The multi stage cell configuration (Figure 9) provides an additional nutrient removal loop and alternating nitrification and denitrification for improved treatment. Different recirculation rates can be used to optimise performance (Abegglen, Ospelt, & Siegrist, 2008; Ersu, Ong, Arslankaya, & Brown, 2008).

Ersu et al. (2008) used a lab-scale MBR with five different recirculation configurations, recirculating mixed liquor and permeate flow rate, to examine the effects of the type of recirculation on nutrients and carbonaceous removals. The recirculation configurations included:

1. The influent fed to the bottom of the anaerobic cell along with the mixed liquor from the anoxic cell.
2. The influent fed to the bottom of the anaerobic cell along with the mixed liquor from the anoxic cell and permeate was recirculated into the anoxic cell.
3. The influent fed to the bottom of the anaerobic cell and the mixed liquor was recirculated into the anoxic cell.
4. The influent fed to the bottom of the combined anaerobic / anoxic cell along with the mixed liquor from the anoxic cell.
5. The influent fed to the bottom of the combined anaerobic / anoxic cell along with the mixed liquor from the anoxic cell and the permeate was recirculated into the anoxic cell.

Of the five configurations tested by Ersu at al. (2008), the configuration with 100% mixed liquor recirculation into the anaerobic cell and 100% permeate recirculation into the anoxic cell (configuration 2), had the highest percentage removal, with an average 92.3% soluble chemical oxygen demand, 75.6% total nitrogen and 62.4% total phosphorus. The highest removal of total phosphorus was when the 300% mixed liquor recirculation into the anaerobic cell and 100% permeate recirculation into the anoxic cell (configuration 2), with a removal rate of 88.1%. Under the same configuration (configuration 2) the highest total nitrogen removal was obtained for the 200% mixed liquor and 300% permeate recirculation with a removal rate of 90.3%.

Bracklow et al. (2007) found no major differences in the efficiency between the four types of recirculation schemes tested on a laboratory scale MBR systems. The MBR configuration was similar in all the recirculation schemes and included first anaerobic and aerobic cell then anoxic followed by the membrane cell. The recirculation schemas included the following arrangements:

1. Recirculation of 200% from the anoxic cell into the anaerobic cell and 310% from the membrane cell into the aerobic cell.
2. Recirculation of 200% from the membrane cell into the anaerobic cell and 310% from the membrane cell into the aerobic cell.
3. Recirculation of 200% from the membrane cell into the anaerobic cell and 310% from the membrane cell into the anoxic cell.
4. Recirculation of 310% from the membrane cell into the anaerobic cell.

Of the four recirculation tested by Bracklow et al. (2007) showed removal rates of biodegradable organic ranging from 99.6% to 97.9%, total nitrogen of 89% to 92.1% and total phosphorus of 97.4% to 99.4%.

An alternative to the pre-denitrification configuration is the post-denitrification configuration shown in Figure 10 (Kraume & Drews, 2010). In this configuration the order for nitrification and denitrification is seemingly more logical; nitrate produced in the aerobic cell is passed into the anoxic cell to be converted to nitrogen gas. Since most organic matter is already being used in the aerobic zone usually a dosage of external carbon-source is applied to the anoxic tank in order to achieve higher denitrification rates.

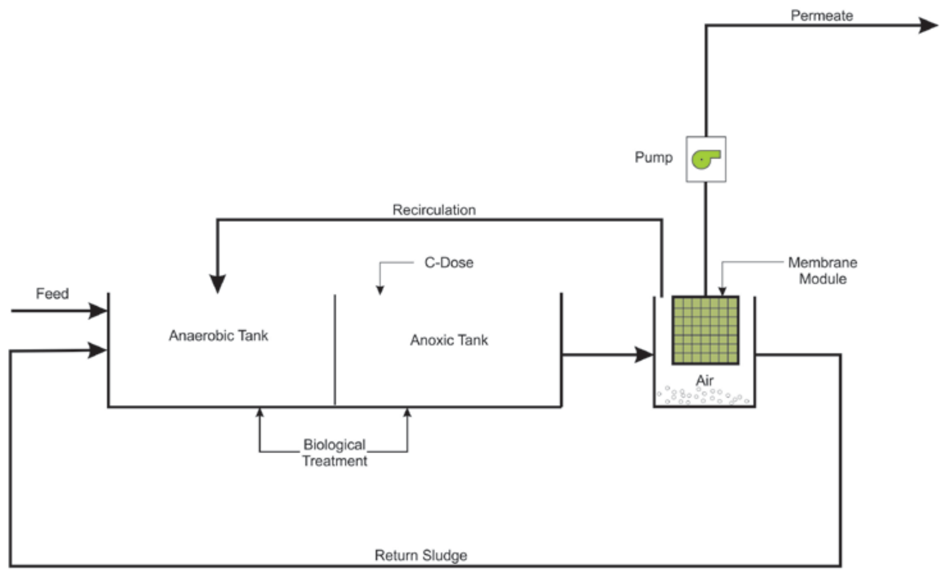


Figure 10: Post – denitrification MBR configuration

2.6.3 MBR Hydraulic and Solid retention times

The advantages of the MBR over the CAS process are mainly associated with the possibility of separating the solids retention time (SRT) from the hydraulic retention time (HRT) (Pollice, Laera, Saturno, & Giordano, 2008). Similar to any biological treatment process, the treated water in the MBR is separated from the biomass. During conventional treatment this separation will normally occur in the second clarifier by sedimentation. For this sedimentation to take place the solids must be retained in the clarifier for an appropriate length of time to allow the particles to settle. However, in the MBR systems the separation is made by membrane filtration, which is not dependent on gravity, and therefore does not require a long HRT. The SRT or the age of sludge in the MBR is not directly dependent on the HRT and can be controlled separately (Judd, 2011; Leiknes, 2010). Due to SRT and HRT not being linked the separation of the liquid and solids during MBR process is superior (Pollice et al., 2008).

Due to the SRT playing a crucial role in the performance of the MBR treatment plant and the fact it can be altered to suit treatment requirements, many researchers have investigated the effects of the SRT on the performance of the MBR to remove nutrients and biodegradable organics (Han, Bae, Jang, & Tak, 2005; Hong, Aryal, Vigneswaran, Johir, & Kandasamy, 2012; Pollice et al., 2008; Van den Broeck et al., 2012; Zubair, Jinwoo, Byung-Ran, Kyung-Guen, & Kyu-Hong, 2007).

Wastewater plants with longer solids retention time have been found to produce less sludge volumes (Ghyoot & Verstraete, 1999). In those plants, sludge volumes will be reduced due to the growth of organisms higher in the food chain (protozoa and metazoan) which feed on bacteria (Ghyoot & Verstraete, 1999), and complete mineralization of influent organic matter (Brindle & Stephenson, 1995).

The SRT time was found to impact membrane fouling, which is one of the major issues in the operation of the MBR system. Fouling reduces membrane permeability and increases pressure loss across the membrane leading to a reduction in the membrane flux. There are several types of membrane fouling which will be discussed in more detail in section 2.8.4.5.

Longer SRT of 60 days was found to overall have less membrane biological fouling than 20 days, however, the membrane cake resistance fouling type was found to decrease when increasing the SRT (Han et al., 2005; Zubair et al., 2007). Hong et al. (2012) also found that the pressure across the membrane, called transmembrane pressure (TMP), is increased when using lower SRT and HRT. An SRT of 20 days was shown to have the highest TMP levels, while 100 days was shown to have the lowest TMP levels (Hong et al., 2012; Zubair et al., 2007) The increased TMP led to membrane flux being reduced and fouling to start on lower flux conditions (Han et al., 2005; Van den Broeck et al., 2012).

Biodegradable organics removal using the MBR system has been shown to have similar performance in different ranges of SRT from 20 days to 100 days. This is due to the microorganisms staying within the aeration tank in contact with the influent for a longer time period (Han et al., 2005; Pollice et al., 2008). At SRT's of 10 days, the biodegradable organics removal rate was shown to be slightly lower (Van den Broeck et al., 2012).

In several studies nitrogen removal has been shown to be highest with a SRT of 60-70 days due to the higher concentration of nitrifying and the denitrifying bacteria, but when SRT was adjusted to above and below 60-70 days, the removal efficiency decreased because of lower growth rate of specific bacteria. (Han et al., 2005; Pollice et al., 2008; Van den Broeck et al., 2012). Phosphorus removal was poor at longer SRT since excess sludge was reduced (Han et al., 2005; Kraume et al., 2005).

2.6.4 Membrane Technology

In the membrane bioreactor systems the membrane replaces the second clarifier existing in the conventional treatment. The membrane will allow some physical or chemical components to pass through it, while others will be rejected. The level of selectivity depends on the membrane pore size and the operation conditions (Environmental Protection Agency, 2005; Judd, 2011; Leiknes, 2010).

2.6.4.1 Membrane materials

There are a number of different types of membrane material that are used in membrane filtration. The membrane materials should demonstrate thermal stability over a wide range of temperatures, chemical stability over a range of pH, and good mechanical strength. A reasonable degree of flexibility is required for cleaning when air scouring is used. The membranes used for wastewater treatment usually contain a very thin surface layer supported on a thick micro-porous layer. The thin skin layer is the selective layer to perform separation, while the micro-porous substrate mainly provides the mechanical strength (Judd, 2011).

The two membrane materials used for wastewater treatment are organic, polymeric (Figure 11a) or inorganic, metal or ceramic (Figure 11b). The most common inorganic membrane material is ceramic. It is considered expensive in comparison to polymeric membranes (Leiknes, 2010). Therefore at present the commercial membranes for wastewater treatment are mainly fabricated from polymeric materials.

(a) (b)

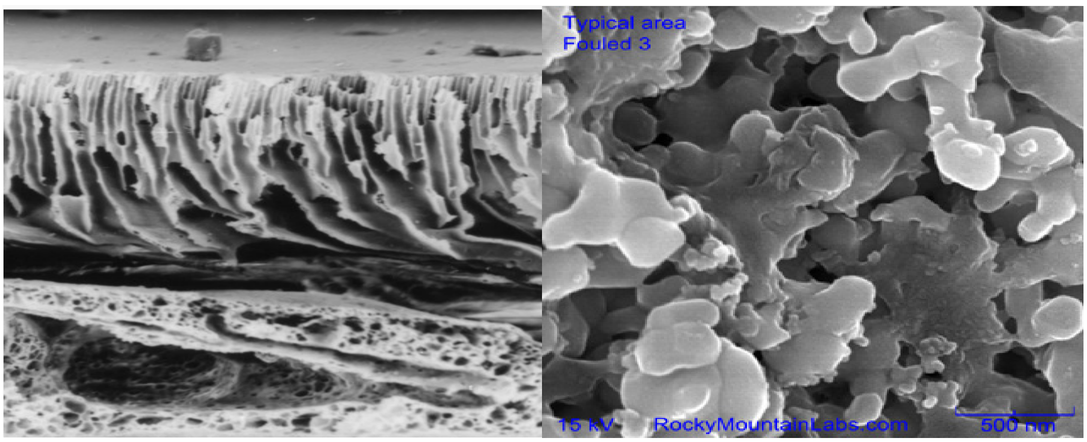


Figure 11: MBR most common material: (a) Polymeric (b) Ceramic

There are a variety of polymers that are suitable for making membranes. For the purposes of wastewater treatment, a limited set of materials, including polyvinylidenefluoride (PVDF), polysulfone (PS), polyethylsulfone (PES), polyethylene (PE) and polypropylene (PP) are used. The polymers are relatively easy to work with and can be made to suit different sizes, shapes and physical strengths, allowing the membrane to be manufactured to meet specific applications (Peng et al., 2012).

2.6.4.2 Membrane selectively

There are various types of membrane separation processes for wastewater treatment; the selectivity of a membrane will be determined by pore size. The largest pore membrane typically used in MBR is microfiltration (MF) while the most selective membrane used is reverse osmosis (RO) which can reject charged ions e.g. sodium, chloride.

The four key membrane separation processes in which the treatment quality is determined are RO, nanofiltration (NF), ultrafiltration (UF) and MF.

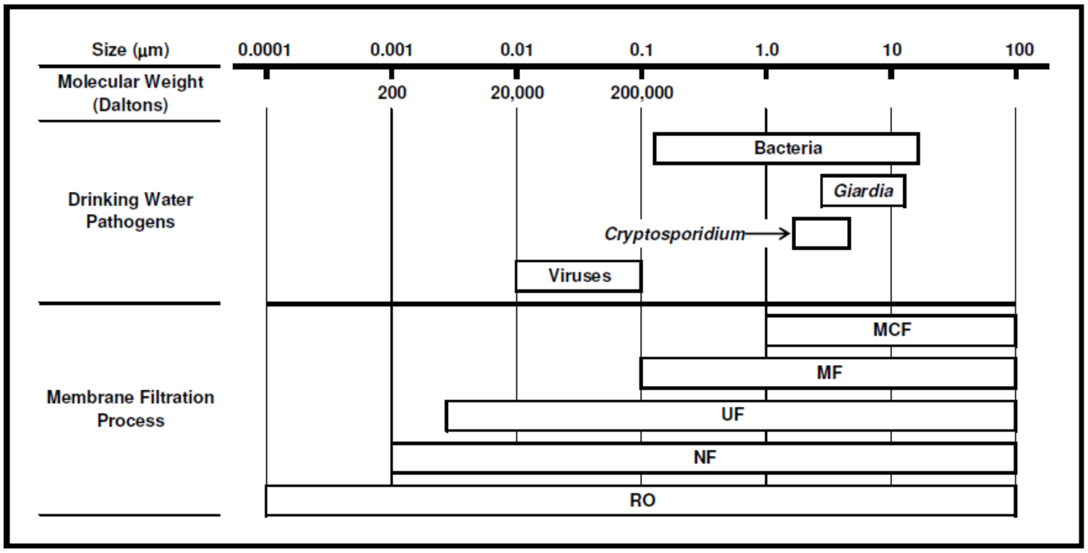


Figure 12: Different membrane filtration processes and their ability to remove pathogens (Environmental Protection Agency, 2005).

The ability of each membrane type to remove different types of pathogens is dependent on the pore size of the membrane. Figure 12 shows the approximate size of viruses, cryptosporidium, giardia and bacteria as well as the different membrane filtration process and their ability to remove these pathogens. As well as the removal of pathogens, membrane filtration can also remove particles, solids, ions and molecules.

Microfiltration has the largest pore size of the four main membrane types. In the microfiltration systems the filtration process removes micron-sized particles such as particles and bacteria from fluids. MF membranes have pore sizes ranging from 0.1 to 5 micrometers (µm). Because the pores are relatively large comparing to the other membranes, it can operate under low pressures and therefore low energy.

In ultrafiltration (UF) systems the membranes have a porous structure ranging from 0.1 to 100 μm. UF membranes are used as a barrier to viruses, endotoxins and proteins.

In nanofiltration (NF), the membranes are of porous filter media with an average pore diameter of 0.001 to 0.01 μm. NF membranes can filter particles up to and including some salts and sugars, however it is unable to remove most aqueous salts and metallic ions.

Reverse osmosis has a pore size range of 0.0001 to 0.001 μm and it is used on a large scale for the desalination and purification of water as it filters out everything but water molecules. The reverse osmosis concept is based on a natural phenomenon discovered as early as 1748. Diffusion of fluid (usually water) through a semipermeable membrane from a solution with a low solute concentration to a solution with a higher solute concentration occurs until fluid concentration on both sides of the membrane reaches equilibrium. The pressure difference between the two sides of the membrane is equal to the osmotic pressure of the solution. The chemical potential gradient across the membrane acts as the driving force for the water transport. By applying a pressure in excess of the osmotic pressure, pure water flows from the high solute concentration side through a membrane to the low solute concentration side and thus, the separation of water from the solution is achieved. In effect the “pores” are transient channels in the dense film.

The selection of the membrane type and size will be determined by the required performance. Normally the membrane filtration sizes that are used in the MBR process include the MF and UF (Judd, 2011; Leiknes, 2010).

2.6.4.3 Membrane configuration

The geometry and the way the membrane is mounted and oriented in relation to the flow of the water is known as the membrane configuration (Judd, 2011). The way in which the membrane elements are housed in containers is described as modules. The membrane configuration and the modules are crucial for the overall process performance.

Ideally the membrane configuration should have a high packing density, allow some degree of turbulence, allow low operating cost, facilitates cleaning and permits modularisation (Judd, 2011).

There are six main membrane configurations:

1. Flat sheet (FS)
2. Hollow fiber (HF)
3. Multi-tubular (MT)
4. Capillary tube (CT)
5. Filter cartridge (FC) and
6. Spiral wound (SW).

Only the multi-tubular (Figure 13a), hollow fiber (Figure 13b) and the flat sheet (Figure 13c) configurations are suitable for wastewater treatment using the MBR technologies because those arrangements of membrane permit turbulence and effective cleaning.

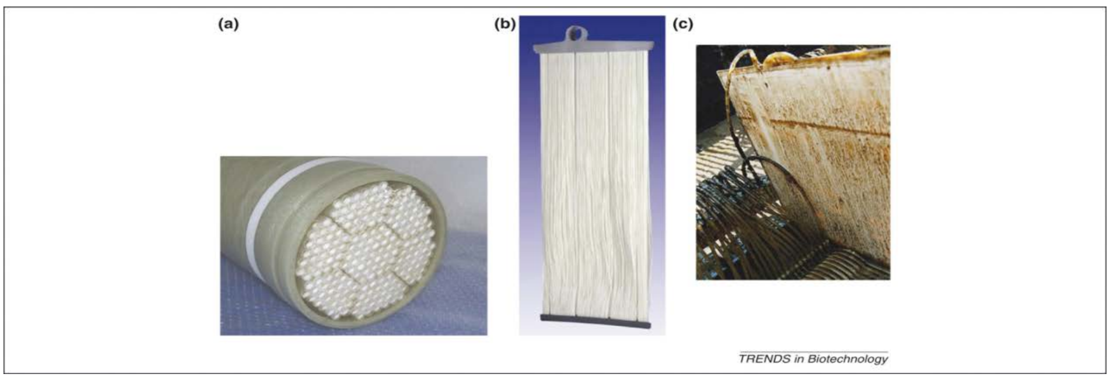


Figure 13: Membrane configuration in the MBR technology

The membrane module and design are dependent on the type of membrane used, for example, the hollow fiber membranes are commonly installed in housing units and are designed to be submerged in fluid. The membrane normally will be connected to the suction side of the permeate pump/s, permeate is extracted by applying a vacuum to the inside of the fiber.

The flat sheet and hollow fiber modules normally operate with flow passing from the outside to the inside, whereas the multi-tubular module operates from inside to outside. The inside to outside arrangement limits the packing density; high packing density will reduce the internal gaps and increase the risk of clogging (Judd, 2011).

For wastewater application, the polymeric hollow fiber membrane is considered to be most superior due to its low cost and ease of fabrication, therefore most membrane separation research today focuses on this type of membrane (Peng et al., 2012; Verrecht, Judd, Guglielmi, Brepols, & Mulder, 2008).

2.6.4.3 Membrane process operation

Pressure is normally used to push or pull water through the membrane, while the membrane selectivity and separation mechanism is different from one process to another. The basic separation mechanism includes two main elements; purified permeate and a concentrated rejected mixture (Figure 14).

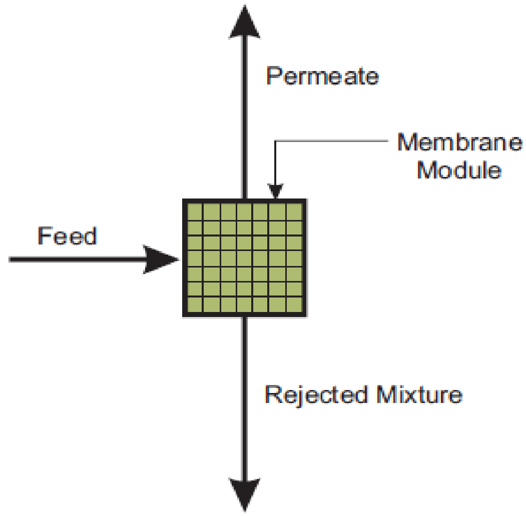


Figure 14: MBR basic separation mechanism (Judd, 2011)

The rejection of particles places a fundamental constraint on all membrane processes. The rejected constituents tend to accumulate at the membrane surface which leads to a reduction of water flow through the membrane, a phenomena referred to as fouling. Fouling can take place through a number of physiochemical and biological mechanisms. Section 2.6.4.5 will discuss in more details the types of fouling and its affects.

There are two ways which the membrane process can be operated, “dead end” and “cross flow”. In the dead end arrangement all the flow goes through the membrane. The material removed from the feed water accumulates on the membrane surface and increases pressure across the membrane. The dead end configuration normally causes a rapid membrane resistance increase, which requires a frequent cleaning regime e.g. back-flush. Normally this operation will include aeration which reduces the sludge accumulation on the membrane surface. In the cross flow operation, a portion of the feed water is going across the membrane at a certain velocity that cleans the membrane surface, while the rest of the flow is converted into permeate product. Normally the side stream configuration requires pumps to maintain sufficient cross flow velocities across the membrane to clean up and prevent build up on the membrane surface (Judd, 2011; Leiknes, 2010).

2.6.4.4 Flux and Transmembrane Pressure

The membrane filtration systems are normally characterized by the system flux. The flux (J) is the quantity of material passing through a unit area of membrane per unit time. Common units are liters per m^2 per hour (LMH) and m/day. MBRs generally operate at fluxes between 10 to 150 LMH (Judd, 2011). The flux through the membrane relates directly to the transmembrane pressure (TMP) over the membrane (equation 4) (Field, Wu, Howell, & Gupta, 1995).

$$J = TMP / (\mu(Rm + Rf)) \tag{4}$$

Where μ is viscosity, Rm is clean membrane hydraulic resistance and Rf represents the resistance that occurs from membrane fouling.

The TMP is the difference between the pressure on the outside of the membrane and the applied pressure inside the membrane. The TMP value can vary depending on the level of the membrane fouling.

Higher flux causes the transmembrane pressure to increase more rapidly and more severe membrane fouling is a risk. The TMP has a direct connection to the membrane fouling; it was found that as soon as the TMP level increased, fouling was observed (Field et al., 1995; J. Zhang, Chua, Zhou, & Fane, 2006). To sustain higher flux operation there are several requirements: increased air scouring, shorter filtration cycles, longer backwash duration and higher backwash flux (Hirani, DeCarolis, Adham, & Jacangelo, 2010).

Since the flux and TMP are related to each other, each of them can be fixed during operation. Normally the flux value is fixed, and then the level of TMP is determined, when the TMP reaches a certain value, cleaning will be used to maintain the required flux (Defrance & Jaffrin, 1999; Judd, 2011; J. Zhang et al., 2006).

2.6.4.4.1 Critical flux

To be able to reduce the risk of fouling and the operational requirement, a critical flux should be determined. A critical flux or sustainable flux is when a certain flux is used for which the pressure across the membrane is low and does not increase with time (Field et al., 1995; S. Wang, Guillen, & Hoek, 2005). At times of critical flux, membrane cleaning is not required and fouling rates are small or negligible (Field et al., 1995).

A common way to determine the critical flux is by the flux-stepping method (Le-Clech, Chen, & Fane, 2006). The idea behind this method is to run different flux levels for a fixed time, observe the TMP at each flux rate and look for stability. When the observed TMP is not stable at each flux step this point is referred to as the critical flux (Le-Clech et al., 2006).

Colloidal particles have been identified as playing a major part in controlling the critical flux, while organic solutes and MLSS have been found to have no influence on the critical flux (Fan, Zhou, & Husain, 2006).

Operating at sub-critical flux can increase the life of the membrane. However, there are a number of studies suggesting that fouling in the MBRs can occur when operating below the critical flux (Field et al., 1995; Ognier, Wisniewski, & Grasmick, 2004). Wang et al. (2005) found no cell accumulation below critical flux, however, small fractions of cell deposit were observed regardless of the flux.

2.6.4.5 Membrane Problems

Membrane fouling and the high cost of running the system are major drawbacks of the MBR process (Meng, Chae, Drews, Kraume, & Shin, 2009). Fouling affects the membrane operation leading to a decline in flux over time and high filtration pressure across the membrane. To deal with the fouling problem, a combination of different cleaning methods such as use of chemicals and physical cleaning are used which increases the operating cost of the membranes.

2.6.4.5.1 Fouling

Membrane fouling is a major issue in the MBR operation. Fouling is the accumulation of unwanted deposits on a membrane surface and within the membrane pores. These deposits reduce the effective membrane surface area, increase TMP and consequently increase energy consumption. The fouling process in the MBR is determined by three basic factors: (1) the nature of the feed – biomass, (2) the membrane properties and (3) the hydrodynamics experienced by the membrane (Meng et al., 2009; J. Zhang et al., 2006).

Fouling in MBR systems can be divided into three stages (Figure 15) (Meng et al., 2009; J. Zhang et al., 2006). The first stage is characterised by a short term fouling where the system adjusts to the set operating parameters e.g. flux, mixed liquor suspended solids (MLSS) and pore blocking. Stage two is a stage of long term operation with relatively low fouling rates until there is a sudden and sharp increase in TMP. The sharp increase in TMP level can be attributed to an increase in fouling (Fane & Chang, 2002). Zhang at al. (2006) showed a relationship between the TMP jump and a sudden change of the biofilm or cake layer structure and the release of extracellular polymeric substances (EPS) from the dead microorganisms. The EPS has been found to be closely related to an increase of TMP. During subcritical flux the generation of EPS inside the biocake was found to be the main reason for TMP increase (Hwang et al., 2008). When working in higher flux or supercritical flux, the TMP was observed to increase very rapidly in one stage only, rather than the normally observed three stages (Hwang et al., 2008).

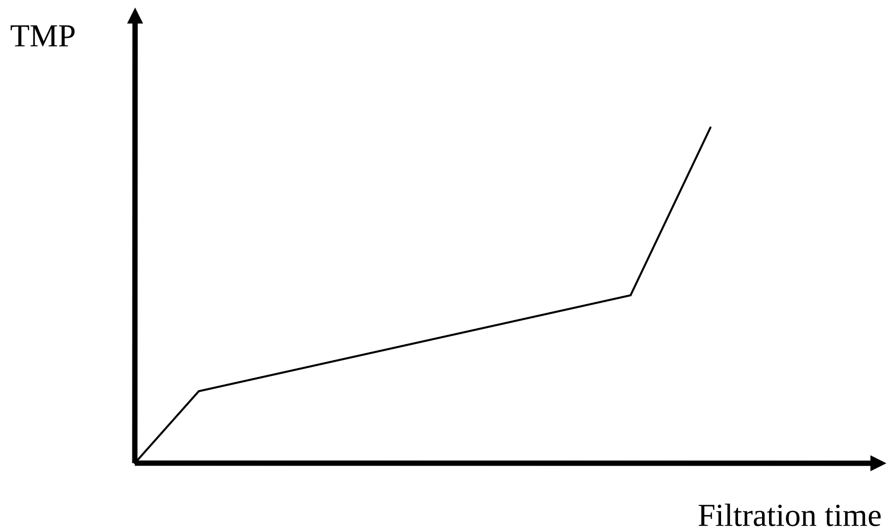


Figure 15: Fouling stages

Fouling mechanisms can be divided into four main groups: complete blocking, standard blocking, intermediate blocking and cake filtration. The fouling can occur either within the membrane or on the membrane surface. Each group of fouling mechanisms can fall into one or more of the following groups: biological, precipitative, adsorptive and cake formation fouling (Meng et al., 2009).

Biological fouling

Aerobic and anaerobic living materials such as bacteria, fungus and algae can cause biological fouling, with bacteria identified as the major cause (Komlenic, 2010).The organisms can colonize on the surface of the membrane and grow into massive quantities that effectively block flow through the membrane surface (S. Wang et al., 2005) (Figure 16). Biofouling contributes to more than 45% of all membrane fouling (Komlenic, 2010). This type of fouling is problematic in the MBR system because the sizes of the microbiological flocs are larger than the MF and UF membrane pores, therefore creating membrane blockage. An important factor that contributes to the biological fouling is the extracellular polymeric substances (EPSs) which are released by the bacteria under limited oxygen conditions or when the bacteria die (Meng et al., 2009). The ESPs hold the wastewater flocs more tightly on the membrane which clog the membrane pores and increase the difficulty in cleaning the membrane (Hwang et al., 2008; Jianfeng, Fenglin, Yaozhong, Fook-Sin, & Hwee, 2008; Kim, Chen, & Yuan, 2006; Ramesh, Lee, & Lai, 2006).

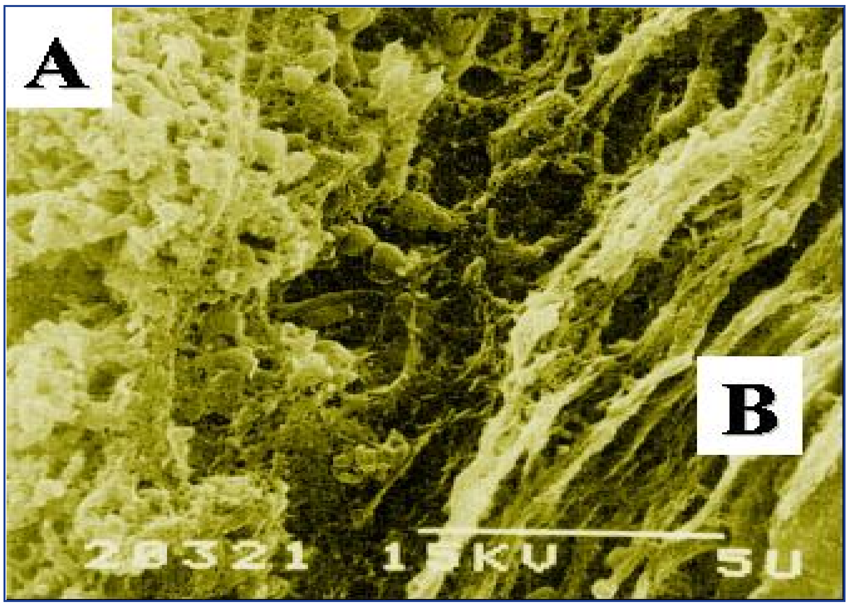


Figure 16: Biological fouled membrane (A=biological growth, B=membrane surface)

Precipitative fouling

Precipitative fouling (Figure 17) will occur when the compounds in the feed water are concentrated beyond their solubility and form flakes. Common precipitates include calcium carbonate, magnesium sulphate and calcium chloride. Choo & Lee (1996) found $MgNH_4PO_4$ $6H_2O$ to be a major foulant. Evidence of deposition of calcium carbonate when treating high calcium rich wastewater was observed by Mooseok & Jeonghwan (2009), resulting in significant surface fouling with minimum membrane pore blockage. Fouling created by calcium cannot be removed by aeration, but the use of citric acid and sodium hypochlorite have been found to be very effective (Mooseok & Jeonghwan, 2009).

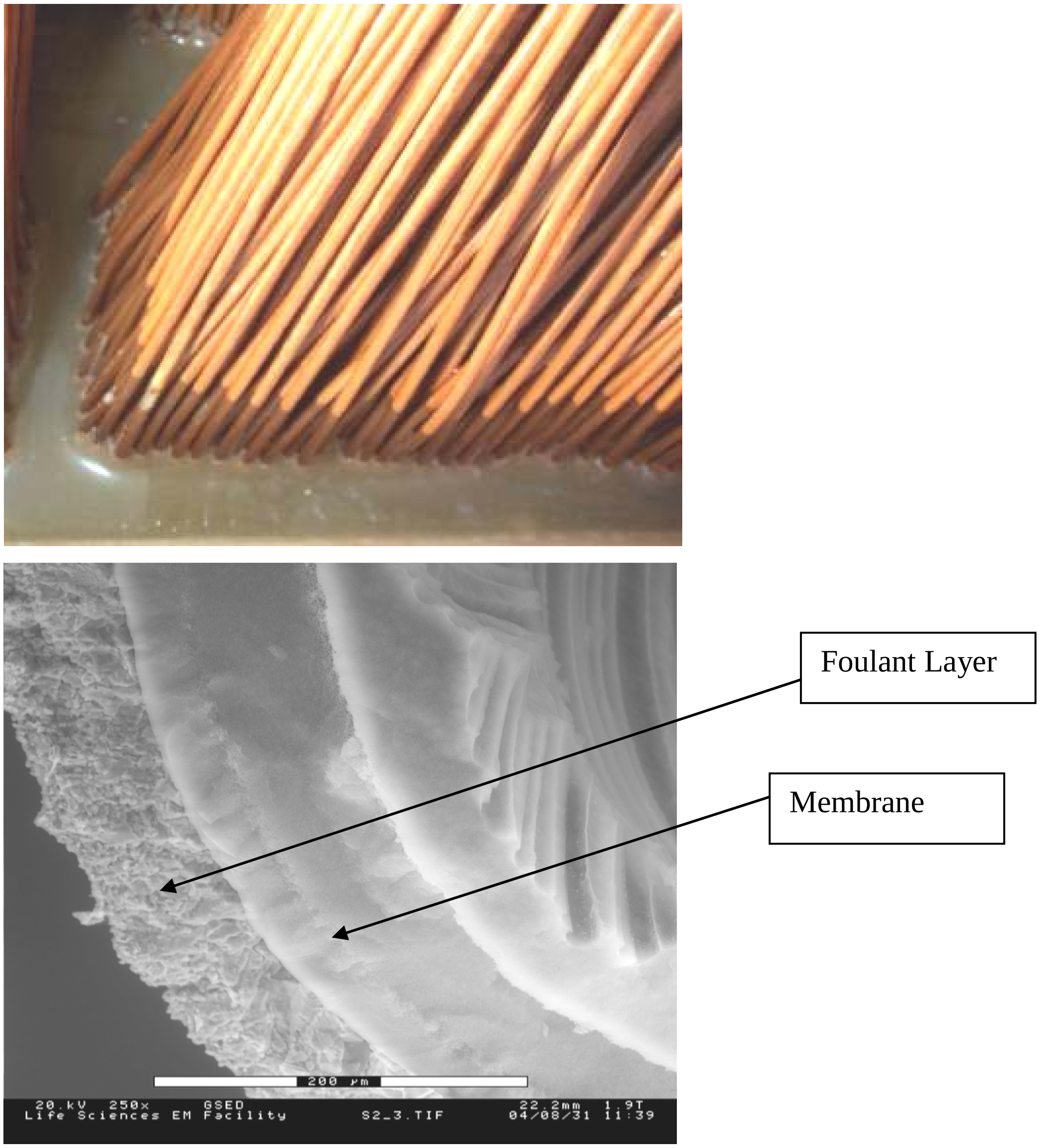


Figure 17: Precipitative fouling

Adsorptive fouling

Adsorptive fouling occurs when compounds attach to the surface of the membrane. Examples of adsorptive foulants include materials such as oil, polymers, cationic surfactants, and hydrocarbons. Absorptive fouling is very hard to clean; therefore pretreatment measures should take place to avoid this type of fouling. Adsorptive fouling can occur when wastewater contains fatty acids. It has been found that adsorptive fouling caused by fatty acids increases as the pH decreases. Wastewater above pH 7-8 showed almost no fatty acid adsorption (Brinck, Jonsson, Jonsson, & Lindau, 2000).

Cake formation

The cake formation layer (Figure 18) is considered to be the main cause for fouling. Lee, Ahn and Lee (2001) and Lin et al. (2010) found that out of the total filtration resistance, cake resistance can be in a range of 80-98%. Since cake layer formation plays an important role in membrane fouling, there has been much research on this issue (Chu & Li, 2005; Di Bella et al., 2007; Lin et al., 2010). Chu and Li. (2005) found that the cake layer is not uniformly distributed on the entire membrane fiber surface. Some of the area was covered with a sludge cake that could not be removed by aeration, and partly by a thin sludge layer that can be easily cleaned by the aeration bubbles. The EPS released by the bacteria leads to difficulties in scouring the membrane by air, by holding the flocs together more tightly. Cake formation in anaerobic conditions was found to be different from that in aerobic conditions. In the anaerobic MBR the cake formation was found to be more irreversible and dependent on many factors, such as biogas, permeate flow rate, MLSS and COD levels (Lin et al., 2010). Di Bella et al. (2007) found that during the first stages of the filtration process, or start up, most of the irreversible fouling occurs and cannot be cleaned by physical cleaning such as aeration or relaxation methods.

Figure 18: Cake formation

2.6.4.5.2 High Cost

An overview of the MBR market reported that the investment cost of a MBR plant increases exponentially with decreasing plant size (Judd, 2007, 2011). When the MBR was first introduced into the market the processes were considered to be very expensive (Leiknes, 2010). However, today with new developments the capital cost of the membrane has dropped significantly. In the last 15 years the operation and maintenance of the MBR have improved, reducing the running cost of the plants. Efficient operation of the MBR will increase the life of the membrane, therefore reducing the total cost of the membrane (Judd, 2011; Leiknes, 2010).

Cleaning the membrane is the main contributor to the high running costs of an MBR. Of the cleaning methods, membrane aeration, typically used in the submerged membrane configuration for scouring, has the highest energy demands (Gil et al., 2010). Zhang et al. (2003) in their analysis found that the aeration required to run the MBR comprised almost 50% of the total plant energy requirements and had an energy demand of two to four times higher than the conventional activated process treatment. Others found the membrane aeration rate to be more than 70% of the total energy cost (Cornal, Wagner, & Krause, 2003; Verrecht, Maere, Nopens, Brepols, & Judd, 2010). Other power consumers are the aeration system within the aerobic cells and the recirculation pumps. Lower flux will have a significant impact on the aeration energy consumption, in some cases the lower flux can reduce the aeration energy by 45% (Verrecht et al., 2008).

Gil et al. (2010) investigated the energy costs using two different flux rates of 19 LMH and 25 LMH entering a pilot plant, and found a total consumption of 6.06 kWh/m^3 for 19 LMH flux and 4.88 kWh/m^3 for 25 LMH flux. Verrecht et al. (2010) compared the average total energy demands of a hybrid MBR plant and a MBR plant designed to work at a maximum flow of 59,580 m^3/day. Using an average flow of 20,851 m^3/day, the energy consumption was found to be 0.7 kWh/m^3 and 1.07 kWh/m^3 respectively. Zhang et al. (2003) found some types MBR systems were more cost effective than others. The lowest energy membrane investigated in this study was shown to have a total power consumption of 24 kWh/m^3.

A net present value calculation done by Verrecht et al. (2010) found that a change of operation parameters, such as sludge retention time, will increase the operation cost due to a greater aeration demand, while changing the hydraulic retention time does not have any significant impact on the costs. The cost of sludge treatment can be very high, sometimes higher than aeration costs (Yoon, Kim, & Yeom, 2004). In an MBR system the sludge volumes are related to the aeration rates used; when higher aeration is used, less sludge is likely to be produced. Therefore in those cases it will be more economical to increase aeration to reduce the overall power consumption costs (Yoon et al., 2004).

Operating in higher flux will require higher aeration demand, however it is predicted by Verrecht et al. (2010) that the high aeration cost in this case will be offset by reduction of the capital value and membrane replacement costs, when the life of the membrane is extended and therefore decrease the total cost.

A comparison between a conventional activated sludge system using an MBR as a tertiary filtration with a classic MBR by Cote, Masini and Mourato (2004) found the classic MBR to be less expensive.

2.6.4.6 Membrane cleaning and preservation

Membrane fouling requires frequent cleaning, and in the worst case, membrane replacement. In addition to the cleaning, pretreatment prior to membrane processing is also required. To clean the membrane there are two cleaning mechanisms; physical and chemical.

2.6.4.6.1 Physical cleaning

Physical cleaning is normally achieved by backflushing, where the flow is reversed using the permeate water to flush the membrane. The second physical method used is relaxation, which involves ceasing permeation, then scouring the membrane with air bubbles. These two techniques can be used in combination i.e. back flushing and air (Judd, 2011). Qaisrani and Samhaber (2011) found that the combination of back flushing and bubbling to be very effective for reducing membrane fouling, half an hour of a combined cleaning cycle removed 98.5 % of the membrane fouling.

For submerged membranes, aeration plays an important role in preventing fouling. High aeration rates will reduce the attachment of sludge to the membrane (Qaisrani & Samhaber, 2011), and also influences the biomass characteristics (Meng et al., 2009). Too high aeration intensity can break the sludge flocs which may lead to a production of EPS, resulting in an increase in the risk of fouling, as well as creating membrane pore blocking (Fan & Zhou, 2007).

During a backwash cycle, the direction of flow is reversed for a period ranging from about 30 seconds to 3 minutes (Qaisrani & Samhaber, 2011; Zsirai, Buzatu, Aerts, & Judd, 2012). The force and direction of the flow dislodges the contaminants at the membrane surface and washes accumulated solids. An increased backflush flux was found to be slightly more effective than increased backflush duration (Zsirai et al., 2012). Backwashing is conducted periodically according to manufacturer specifications and site specific considerations (Judd, 2011; Leiknes, 2010).

In general, a backwash cycle is triggered when a performance-based benchmark is exceeded, such as a threshold for operating time, increase in TMP, and/or flux decline. Ideally, the backwash process restores the TMP to its baseline level; however, most membranes exhibit a gradual increase in the TMP that is observed after each backwash, indicating that the accumulation of foulants cannot be removed by the backwash process alone. These foulants are addressed through chemical cleaning.

Siembida, Cornel, Krause and Zimmermann (2010) found mechanical membrane cleaning using granular material, reduced the fouling rate. Using granular material could lead to longer treatment operation with no chemical cleaning, and higher flux. Mechanical declogging was found to be more effective than chemically enhanced backflush in recovering membrane permeability (Zsirai et al., 2012).

2.6.4.6.2 Chemical cleaning

Chemical cleaning is another means of controlling membrane fouling, particularly those foulants such as inorganic scaling and some forms of organic and biofouling that are not removed via the backwash process. Chemical cleaning can be performed in situ or ex situ. The goal of chemical cleaning is to attempt to restore the TMP of the system to its baseline level (Judd, 2011).

There are a variety of different chemicals that may be used for membrane cleaning, and each is generally targeted to remove a specific form of fouling. For example, citric acid is commonly used to dissolve inorganic scaling, and other acids may be used for this purpose as well. Strong bases such as caustic soda, are typically employed to dissolve organic material. Detergents and surfactants may also be used to remove organic and particulate foulants, particularly those that are difficult to dissolve. Chemical cleaning may also utilize concentrated disinfectants such as a strong chlorine solution to control biofouling. A combination of different chemicals was found to be more efficient when more serious and multiple fouling occurred (Wei, Huang, Ben Aim, Yamamoto, & Amy, 2011). Chemicals that are used in the MBR system can include sodium hypochlorite, hydrogen chloride, and sodium hydroxide (Porcelli & Judd, 2010; Wei et al., 2011). Sodium hypochlorite was found to be the most efficient cleaning reagent (Raffin, Germain, & Judd, 2011).

Wei et al. (2011) found that the effluent quality before and after chemical cleaning showed changes only within the first hour after the cleaning, constituents such as total nitrogen decreased and total phosphorus increased. One hour after cleaning the effluent quality returned to the pre-cleaning levels.

2.6.4.6.3 Cleaning operation

In some systems a combination of chemical cleaning and backflushing is also an option; the cleaning chemical is added to the backflush water to produce a chemically enhanced backflush. However, it has been shown that when using chemically enhanced backflush there still is a small decline of the membrane permeability (Zsirai et al., 2012).

Physical cleaning has several advantages over the chemical cleaning for a number of reasons. Physical cleaning is generally a faster process than chemical cleaning, lasting about two minutes. It demands no chemicals therefore produces no chemicals in the waste and also is less likely to cause membrane degradation. On the other hand physical cleaning is less effective than the chemical cleaning (Judd, 2011).

During the operating cycle of the membrane, using both physical and chemical cleaning, the membrane permeability will not be fully returned to its initial state (Judd, 2011; Meng et al., 2009; Zsirai et al., 2012). The remaining residual will have an impact on the membrane permeability, a mechanism called irrecoverable fouling, which will determine the life of the membrane (Meng et al., 2009).

The increased resistance or the permeability decline of the membrane will affect the flux, therefore the flux rate is normally used to determine between physical cleaning (backflushing or relaxation). Normally physical cleaning is used first due to the low cost and low impact on the membrane. Since the physical cleaning does not return the membrane permeability to its original condition, only a given number of backflush cycles can be performed before flux rate will be low and maximum pressure will be reached (Judd, 2011).When the point of maximum pressure is reached the membrane operation cannot be sustained, at this point chemical cleaning is required to return the pressure close to the original pressure value.

By determining the “right” flux rate a reduction of the membrane fouling can be achieved, a balance between the flux, physical and chemical cleaning should be obtained to optimise operation and reduce the running cost of the membranes (Judd, 2011).

2.7 Summary and Research Questions

The activated sludge process is known to have several deficiencies in comparison with the MBR, such as high sludge production, bigger footprint and reduced effluent quality (Gonzalez, Petrovic, & Barcelo, 2007; Judd, 2007, 2011; Leiknes, 2010; Visvanathan, Ben Aim, & Parameshwaran, 2000). However, the MBR also has drawbacks, such as greater process complexity and higher running costs. The MBR cost is mainly associated with membrane fouling and membrane cleanliness (Cornal et al., 2003; Gil et al., 2010; Judd, 2007; Verrecht et al., 2010).

There appears to be only a limited number of studies comparing the MBR wastewater treatment system with the Bardenpho activated sludge process. Therefore there is a need to investigate the operation of both the plants, focusing on treatment qualities in relation to the operating costs, to provide a better understanding of MBR systems in comparison to a Bardenpho activated sludge treatment process.

Based on the literature presented in this chapter, the research questions of this thesis are:

1. How does the effluent quality of the MBR system and a Bardenpho activated sludge process, operating under similar conditions and influent qualities, compare?
2. How do the operational costs of the MBR system and a Bardenpho activated sludge process, operating under similar inflow rates, compare?

Chapters 3 and 4 will describe the MBR and the Bardenpho activated sludge wastewater treatment plants and Chapter 5 will present the details of the methodology which was used to examine these questions.

Chapter 3: Rotorua's Conventional Treatment Plant Description

3.1 Introduction

The first sewerage scheme in Rotorua was constructed in 1891 covering the central business district. Between 1926 and 1969 the reticulation network was expanded to include other parts of the city. In 1973, the wastewater treatment plant (WWTP) was expanded with additional secondary treatment, comprising of a standard biological nutrient removal system, and some chemical stripping of phosphorus. At this time there was a substantial expansion of the reticulation network and increase in population.

In 1991, another significant upgrade of the WWTP was completed converting the standard biological removal system to the Bardenpho biological nutrient removal system, which was designed to further reduce nitrogen. Also at the same time the direct discharge of the effluent to the lake was replaced with a Land Treatment System (LTS). The LTS was constructed in the Whakarewarewa Forest using a slow rate irrigation of trees.

In 2005 the treatment plant was upgraded to suit an increased population of 75,000. New return activated sludge (RAS) pumps and methanol dosing were installed to further increase nitrogen removal.

Between the years 2006 and 2008, ethanol replaced methanol as it was found that the population growth rate of methanol-utilising bacteria was too slow for the available residence time. Flow-balancing was introduced where a portion of the peak inflow is stored and then fed in during periods of low-flow, to create a slightly more stable flow through the Bardenpho (Figure 19).

Figure 19: Rotorua Wastewater treatment plant before the MBR upgrade

In 2009 it was determined that the Bardenpho process alone could not produce the desirable removal rate of nutrients, therefore additional improvement was required. After investigation of the market and site limitations it was agreed to add the MBR process as part of the plant (Described in more detail in Chapter 4).

The construction of the MBR was completed at the beginning of 2012. Up to this time all the flow, averaging approximately 19 ML/d, was directed to the Bardenpho process. The performance of the Bardenpho during this time will be presented in the results chapter.

3.2 Process description

A description and illustration (Figure 20) of the Rotorua conventional treatment process is presented in this section.

At the beginning of the treatment process the raw wastewater entering the treatment plant initially passes through a three millimeter rotary drum screen and a Vortex Pista-grit trap to remove large material and grit. The large material and grit from both the screen and the grit trap is collected and sent to the landfill.

The raw wastewater is then pumped to three primary sedimentation tanks (each 1000m3) where primary sludge settles. The primary sludge is then thickened; a small fraction is partially fermented to produce volatile fatty-acids (VFA), while the rest is pumped to the sludge storage tanks. The VFA is pumped to the anaerobic cell of the Bardenpho as a carbon supplement.

The primary effluent exiting from the primary clarifier and the partially fermented sludge from the VFA system are pumped to the secondary treatment. The secondary treatment involves a five stage Bardenpho process: (1) anaerobic fermentation; (2) first anoxic; (3) aeration; (4) second anoxic; and (5) reaeration; with eighty percent recycle from stage 3 back to stage 2. To be able to maintain sufficient carbon required for the biological treatment, ethanol is added to the fourth stage of the Bardenpho.

Excessive flow coming into the Bardenpho is directed to an equalisation tank to ensure flow and contaminant level are balanced. An old clarifier is utilised for equalisation.

The effluent coming out of the Bardenpho is separated in two clarifiers and stored in ponds (each 6000m^3) at the treatment plant. To maintain a constant population of micro-organisms in the Bardenpho, a return activated sludge is circulated from the second clarifier back into the Bardenpho stage 1.

The waste activated sludge (WAS) from the Bardenpho process is directed to the dissolved air flotation (DAF) facility, which is used to separate the liquid from solids. The sludge coming out of the DAF is then pumped to the sludge storage tank and the liquid is returned to the Bardenpho stage 3.

Sludge from the sludge storage tank goes through a filter bed press which dewaters the sludge. The dewatered sludge is then re-used in worm farms, while the liquid is pumped back to the Bardenpho.

The final effluent stored in the plant ponds is pumped and discharged to the land treatment system (LST) through a slow irrigation system located at the Whakawerawera forest.

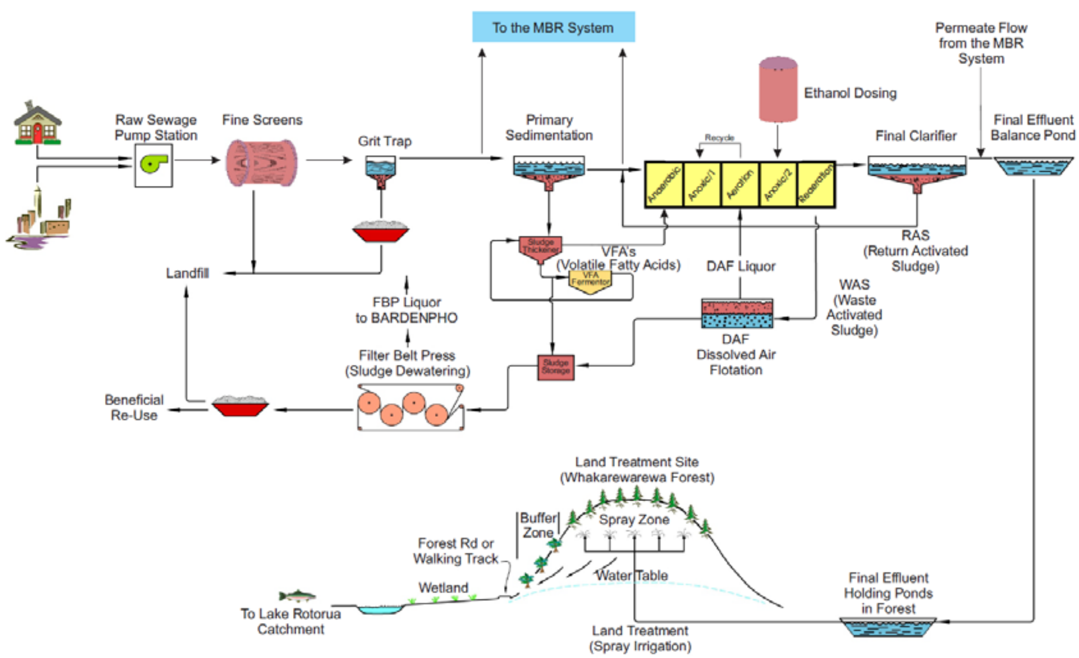


Figure 20: Rotorua conventional wastewater treatment plant – process diagram

3.3 Plant facilities

3.3.1 Introduction

This section explains in more detail the major components within the conventional treatment process. The relevant components will be used in the results chapter to evaluate the total running cost of the Bardenpho system.

3.3.2 Preliminary screen

The first element of treatment at the treatment plant is a rotating drum screen with a three millimeter wedge wire mesh (Figure 22). The screen is designed to work continuously and is common in both the MBR and the Bardenpho processes.

Figure 21: Rotorua treatment plant preliminary rotary drum screen

The screen collects on average approximately 202 tonnes a year of waste that is disposed in the landfill daily at $49 per tonne.

After screening, the wastewater continues toward the second treatment within the preliminary stage, the grit trap. Table 7 presents the wastewater composition before it enters the grit trap.

Table 7: Post screen wastewater characteristics

Parameters	**Average Level (mg/L)**
Alkalinity	201.09
Total Phosphorus	6.01
Ammonia	32.29
Total Kjeldahl nitrogen (org N + NH_4^+)	48.6
Biodegradable organics (BOD5)	197.83
Chemical oxygen demand (COD)	449.48
Total Suspended Solids	387.34
pH	7.17

Data was taken from RDC WWTP Laboratory from the 22/2/2012 until 20/12/2012

3.3.3 Grit trap

After the wastewater passes through the three millimeter screen it gravitates to two Vortex Pista grit trap chambers. The Vortex Pista type grit trap consists of a cylindrical tank in which the flow enters tangentially, creating a vortex flow pattern. The grit settles by gravity into the bottom grit trap hopper and is then removed by two grit pumps (Figure 22). The grit trap system removes approximately one tonne a week of fat and solids which are collected in a bin using a screw conveyor. The collected fat and solids are disposed of at the landfill daily, with a monthly cost of $ 400.

Figure 22: Rotorua treatment plant Vortex Pista grit trap

3.3.4 Primary clarifier pumps

The primary clarifier feed consists of three submergible NP3201.180~624, 21kW Flygt pumps (Figure 23). The primary pumps transfer preliminary treated wastewater to the three primary clarifiers. The pumps are submerged in a well, situated after the grit trap.

The primary feed pumps are programmed to operate according to the water level. The pumps will stop whenever the water level in the pump chamber is below pump start level and will be ramped as water level increases.

Figure 23: Conventional process primary clarifier feed pumps

3.3.5 Primary clarifier

Preliminary treated wastewater is sent to three 1000m^3 circular sedimentation tanks which are mechanically cleaned (Figure 24). The wastewater is introduced in the center of the tank in a pipe encased in concrete beneath the tank floor. At the center of the tank, the wastewater enters a circular well, designed to distribute the flow equally in all directions.

The sludge is removed by the scraper attached to the arms at the bottom of the tank, while a blade that is connected on the surface removes the scum. The clarified liquid or primary effluent is skimmed off over a weir at the clarifier edge.

Figure 24: Rotorua's primary clarifier

The primary clarifier has retention time of between two to three hours. This retention time was set to allow a minimum of suspended solids to be removed. The relatively high suspended solids effluent is used to feed the biological treatment, reducing the need to add higher amounts of carbon source (ethanol) into the biological treatment.

The typical performance of the Rotorua primary clarifier is summarised in Table 8.

Table 8: Rotorua primary clarifier performance

Constituents	Clarifier influent (mg/l)[1]	Clarifier effluent (mg/l)	% removal (%)
BOD5	197.92	100.00	49.47
COD	451.76	241.99	46.43
TKN	47.94	47.49	0.94
TP	6.02	5.54	7.97
TSS	383.50	119.27	68.90

Data was taken from RDC WWTP Laboratory from the 11/1/2012 until 20/12/2012

[1] The clarifier influent concentration represents values prior to the grit trap

3.3.6 Volatile fatty acid system

The sludge generated in the primary clarifier is transferred into a sludge thickener tank. The sludge thickener dewaters the sludge and the liquor is pumped via a set of 5.9kW pumps (Figure 25) to the anaerobic tank within the Bardenpho process.

The sludge out of the sludge thickener is pumped by a set of 11 kW pumps to the volatile fatty acid (VFA) fermenter tank for an anaerobic digestion of the sludge. There is a ten day sludge digestion period to allow the volatile fatty acid to be produced through the acidogenesis process. The volatile fatty acids are then recycled back into the anaerobic tank within the Bardenpho for reuse.

Figure 25: VFA system

Table 9 summaries the average VFA outlet characteristics going into the anaerobic tank within the Bardenpho process.

Table 9: VFA wastewater characteristics

Parameters	**Average Level (mg/L)**
VFA	72.35
Total Phosphorus	7.60
Ammonia	34.54
Total Kjeldahl nitrogen (org N + NH_4^+)	55.70
Organic nitrogen	21.24
Chemical oxygen demand (COD)	591.13
Total Suspended Solids (TSS)	265.40
pH	6.74

Data was taken from RDC WWTP Laboratory from the 16/2/2012 until 21/12/2012

3.3.7 Bardenpho feed pumps

The Bardenpho feed consists of four submergible NP3202.180~616, 22kW Flygt pumps (Figure 26). The pumps transfer the primary treated wastewater (Table 8) to the first stage within the Bardenpho process, the anaerobic tank.

Figure 26: Bardenpho feed pumps

3.3.8 Bardempho process

The Bardenpho system in Rotorua is a biological nutrient removal process which has five stages (Figure 27) and was built on a total area of 1773 m^2 (88.67 m by 20 m) with a total volume of 9044 m^3.

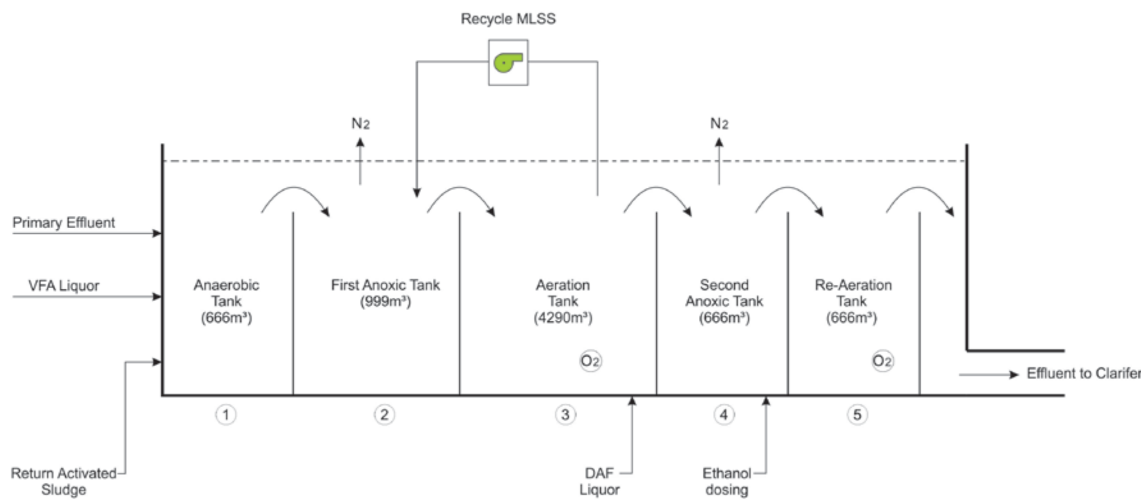


Figure 27: Rotorua five stage Bardenpho treatment system

The first stage in the Bardenpho is the anaerobic tank; the anaerobic tank feeds from three different sources which are mixed within the tank. The sources are:

1. Primary treated influent pumped by the Bardenpho feed pumps (section 3.3.7)
2. Activated sludge coming from the second clarifier pumped by the RAS pumps (section 3.3.9)
3. Volatile fatty acid produced by the VFC system (section 3.3.6)

The anaerobic tank plays an important role in:

1. The breakdown of complex particulate matter and the conversion of this particulate matter into dissolved compounds of lower molecular weight.

2. Assisting with the removal of phosphorus. The phosphorus removal rates are improved by using biological stress by transferring from anaerobic conditions to aerobic conditions.

3. Encouraging the growth of denitrifying bacteria by providing an environment without oxygen.

The second stage of the Bardenpho system is the first anoxic tank. The first anoxic tank feeds from the anaerobic tank and from the aerobic tank via the MLSS pumps. The mixed liquor from the aerobic tank contains nitrates (due to the nitrification process occuring within the aerobic tank) which are mixed with the bacteria produced in the anaerobic tank. The bacteria uses the organic matter in the influent (sourced from the primary and the VFA feed) to reduce the nitrate to gaseous nitrogen through the denitrification process. Approximately two thirds of the total nitrogen entering the Bardenpho system is removed at this stage.

The third stage of the Bardenpho process is the aerobic stage aerated by blowers and diffusers. Each blower has a capacity of 3230 Nm^3/h when operating with a discharge pressure of 55kPa. The diffusers are spread around the tank to create a uniform distribution of dissolved oxygen. The blowers will operate to maintain different levels of dissolved oxygen (DO) in three compartments within the aerobic tank. The range of the DO is 1.8 mg/l to 2 mg/l. In the aerobic tank ammonia is converted to nitrate in the nitrification process. During the nitrification process a source of carbon is needed which is supplied by a feed from the first anoxic tank and from the DAF liquor feed. Uptake of phosphorus by the bacteria also occurs in this tank.

The fourth stage of the Bardenpho process is the second anoxic stage, where the nitrates that were not recycled in the first anoxic tank from the aerobic tank continue on to the second tank to be removed by the denitrification process. In the second anoxic tank organic carbon is limited because of the denitrification and the nitrification processes occurring in the previous tanks consuming the carbon. Ethanol is injected into this tank to increase the level of carbon.

The final stage of the Bardenpho is the re-aeration tank. The re-aeration tank receives flow from the second anoxic tank and is aerated to a DO level of 1.5 mg/l to ensure retention of phosphorus by the biomass.

3.3.9 Recycled activated sludge pumps

The recycled activated sludge (RAS) pumps consist of three dry NP3201.180~627 Flygt pumps (Figure 28). The pumps transfer sludge from the second clarifier to the anaerobic tank within the Bardenpho process.

Figure 28: Bardenpho RAS pumps

3.3.10 Ethanol dosing

The wastewater feed entering the Rotorua treatment plant has relatively low carbon content. Carbon is an important component in the removal of nitrogen. To ensure the treatment process has enough carbon, an external carbon source is added in the form of ethanol to the Bardenpho second anoxic tank. The Bardenpho ethanol dosing system (Figure 29) shares the same ethanol tank with the MBR plant, but includes a separate dosing pump and injection system.

Figure 29: Bardenpho and MBR Ethanol tank

The ethanol is supplied through a dosing pump at an average flow rate range of 30 L/hr. The daily volume of ethanol is set manually according to the level of nitrate in the final effluent and will vary throughout the day.

3.3.11 Second clarifier

Liquor from the de-aeration tank is gravity fed to two 3180 m^3 clarifiers situated behind the Bardenpho process (Figure 30). The clarifier originally was designed to have a retention time of approximately seven hours; however with the MBR taking some of the load, the retention time is currently estimated to be nine to ten hours.

The operation and structure of the second clarifier is similar to the primary clarifier presented in section 3.3.5. However, the second clarifier is larger and allows longer retention times to permit enough time for the solids to settle, producing the final sludge which is pumped back to the Bardenpho first tank, the anaerobic tank. The liquid produced in the second clarifier is the Bardenpho final effluent.

Figure 30: Second clarifier

The final effluent composition is presented in Table 10.

Table 10: Bardenpho final effluent characteristics

Parameters	Average Level (mg/L)
Alkalinity	77.62
Total Nitrogen	4.96
Total Phosphorus	2.56
Ammonia	0.25
Organic nitrogen	3.05
Chemical oxygen demand (COD)	40.90
Total Suspended Solids (TSS)	27.24
pH	7.12

Data was taken from RDC WWTP Laboratory from the 29/10/2012 until 18/1/2013

3.3.12 Final effluent pond

The final effluent produced in the second clarifier is gravity fed to two 6000 m^3 ponds (Figure 31) situated approximately 100 meters away. From the ponds the final effluent is moved using four pumps to another set of ponds located in the Whakarewarewa forest where the effluent is used for irrigating 193 hectares of forestry.

The ponds at the plant have two days of average flow storage to allow sufficient capacity in case of a pump or a pipe failure. The ponds are emptied during low electricity tariff periods.

Figure 31: WWTP final effluent ponds

3.3.13 Dissolved air flotation

The waste from the Bardenpho de-aeration tank is pumped to the dissolved air flotation (DAF) system (Figure 32) using the WAS pumps. The Bardenpho WAS average concentration is 3000 mg/l and is removed by WAS pumps at an average flow of 925 m^3/d.

The WAS entering the DAF is supersaturated with pressurised air. The dissolved air is released as fine bubbles that float suspended material to the surface where it can be scraped off. The liquor coming out of the DAF is recycled back to the Bardenpho aeration tank. The DAF solids are transferred to the sludge storage tank and mixed with the WAS pumped from the MBR.

Figure 32: DAF arrangement

3.3.14 Filter Belt Press arrangement

Sludge from the DAF and VFA system are pumped to the sludge storage tank, mixed, and fed into the filter belt press (FBP) system (Figure 33).

Figure 33: Filter belt press system

The sludge feed on average contains 4 % solids, it is passed between two belts which are compressed together between sets of rollers, squeezing out the water which is then pumped back into the Bardenpho process. When the sludge passes through the final pair of rollers in the process, the belts holding the solids are separated and the solid layer is scraped off into a container. The total final dewatered sludge produced by both the MBR and the Bardenpho is 8523 m^3/year with 17% being solids. The de-watered sludge is composted in worm farms at a total cost of $750,000 a year.

3.4 Summary of the Bardenpho components

Some of the plant facilities described in section 3.3 are common to both the Bardenpho and MBR processes and are not part of the plant secondary treatment. To be able to compare the Bardenpho with the MBR process only the components that are relevant to each of the processes will be analysed. The Bardenpho major components are presented in Table 11. Each of the components' power rating will be used together with the operating hours to show the running costs of the Bardenpho process and the results presented in Chapter 6.

Table 11: Bardenpho proccess major components parameter summary

Description	No of items	Power rating[1]	Flow[2]	Pressure[2]
WAS pump	2	8.5 kW	178 m^3/h	8.6 m
Bardenpho feed pump	4	21 kW	774 m^3/h	7 m
Anaerobic tank mixer	3	2 - 5.9 kW	-	-
First anoxic tank mixer	3	2 kW	-	-
Aeration tank blower	5	40 kW	3230 m^3/h	0.993 bar
MLSS pump	2	7.4 kW	1991 m^3/h	0.65 m
Second anoxic mixer	3	2 kW	-	-
Second clarifier	2	0.37 kW	-	-
RAS pump	3	16.8 kW	756 m^3/h	5.2 m
DAF liquor pump	2	18.8 kW	52.5 m^3/h	8.5 m
Blower fan	3	1.1 kW	-	-

1 The power rating rate was taken from the manufacturer performance curve where available when the curve was not available the size of the motor was used

2 Best efficiency point at 50 Hz

3.5 Bardenpho capital cost estimate

3.5.1 Introduction

The aim of this section is to provide a capital cost estimate of the Rotorua Bardenpho treatment system so that it can be compared with the capital cost of the MBR system. The calculation of the Bardenpho capital cost was evaluated in two stagess. Initially the MBR capital costs were used to estimate the cost of the Bardenpho by using a correcting factor. Then these costs where compared against the Bardenpho costing from the wastewater treatment plant assent management plan to conclude the Bardenpho estimated capital cost.

3.5.2 Bardenpho capital cost estimate using the MBR costing

The MBR system was designed and constructed using existing infrastructure within the wastewater treatment plant. Therefore, the capital cost of the MBR as presented in Chapter 4, would be different if it was built without the use of existing infrastructure. To be able to compare the two processes it is important to evaluate the cost of the Bardenpho as if it was also built on existing infrastructure. In order to do this, the MBR cost components were used with some adjustments to give a comparable capital cost for the Bardenpho.

To calculate the estimated capital cost of the Bardenpho, the cost of the individual items required to build the Bardenpho system were calculated by taking the quoted cost for the relevant items from the MBR work and using the power law equation (equation 5) to account for scale (Bouman, Jesen, & Wake, 2005).

$$Cs/Cr = (S/Sr)^{(a)} \qquad (5)$$

Where:

Cs - represents the cost of the Bardenpho equipment.

Cr - represents the known cost of relevant MBR equipment.

S - represents the Bardenpho equipment size, for example tank volume or flow.

Sr – represents the known relevant equipment size within the MBR.

a – represents a scaling exponent of 2/3.

A summary of the estimated capital costs of the Bardenpho components, calculated using the power law equation are presented in Table 12.

Table 12: Bardenpho estimated capital cost using the power law equation

Item	Component	Cost ($)
1	Design work	1,276,470
2	Blowers and diffusers	929,065
3	Pumps and mixers	2,612,584
4	SCADA and control	338,073
5	Construction and Installation	10,303,272
	TOTAL CAPITAL COST	15,459,464

Note: The cost breakdown is presented in Appendix 1

The general description of the scope of work of each item included in Table 12 is presented below:

1. Design work: The cost of this item was estimated to be 9% of the total construction cost, item 2 to 5 (The Institution of Professional Engineers NZ, 2004)

2. Blowers and Diffusers: The cost estimate of these items included the supply of diffusers and blowers. The value mentioned in this item taking the relevant costs from the MBR and multiplies it by 2.83, which is the ratio of the aerobic Bardenpho tank and the MBR aerobic tank.

3. Pumps and mixers: The cost estimate of these items includes the supply, manufacture and delivery of MLSS pumps, RAS pumps, anoxic and de-aeration tanks mixers, WAS pumps, feed pumps and ethanol dosing pump. The pump value mentioned in this item taking the relevant costs from the MBR and multiplies it by 1.74, which is the ratio of the flow rates of the Bardenpho process and the MBR system. The mixer estimated cost in this item was calculated using the tanks ration of 2.82.

4. SCADA and control: The cost of these items were estimated using the MBR relevant cost and multiply it by 1.74, which is the ratio of the flow rates of the Bardenpho process and the MBR system.

5. Construction and installation: The cost of these items were estimated using the MBR relevant cost and multiply it by 1.74, which is the ratio of the flow rates of the Bardenpho process and the MBR system. The Bardenpho tanks estimated cost in this item was calculated using the ratio of Bardenpho process tank size with the MBR system tank size (2.82).

3.5.3 Bardenpho capital cost estimate using the Bardenpho asset value

For the purpose of budgeting and maintenance of the wastewater treatment plant, the Rotorua District Council (RDC) keeps records of the cost estimate of some of the plant components. These costs are reported annually in the asset management plan of the wastewater treatment plant. Only the relevant costs for the Bardenpho system were taken from the asset management plan and were used for calculating the capital cost of the Bardenpho (Table 13).

Table 13: The Bardenpho estimated capital cost using the RDC asset plan

Item	Component	Cost ($)
1	Primary Pump Station	513,557
2	Bardenpho	6,423,989
3	Second Clarifiers	2,766,524
4	Sludge Liquids Pump Station	127,384
5	WAS Pump Station	204,977
6	Blower Plant Room	520,857
7	Motor Control Centre	1,300,702
8	Ethanol Dosing Facility	636,900
9	Final Effluent Pump Station[1]	513,557
10	Design work	1,170,760
	TOTAL CAPITAL COST	**14,179,207**

Note: The cost breakdown is presented in Appendix 2

[1] Added to allow a “like with like” comparison with the MBR system

The general description of each item in Table 13 is presented below:

1. Primary pump station: The cost estimate of this item included the supply and construction of the primary pumping station wet well, pumps, spare parts and electrical equipment.

2. Bardenpho: The cost estimate of this item included the supply and construction of the Bardenpho structure and lining, mixers, diffusers, pumps, valves, electrical equipment and monitoring equipment.

3. Second clarifiers: The cost estimate of this item included the supply and construction of two clarifiers and lining, motors, pumps, flow meters, valves and electrical equipment.

4. Sludge liquids pump station: The cost estimate of this item included the supply and construction of the sludge liquids wet well, pumps and electrical equipment.

5. WAS pump station: The cost estimate of this item included the supply and construction of the waste activated sludge wet well, pumps and electrical equipment.

6. Blower plant room: The cost estimate of this item included the supply and the installation of the blowers and air compressors.

7. Motor control centre: The cost estimate of this item included the supply and construction of the control room building, cabinets, pumps drivers and starters, meters and control panels.

8. Ethanol dosing facility: The cost estimate of this item included the supply and the installation of the ethanol tank, pipework, dosing pumps and electrical equipment.

9. Final effluent pump station: The cost of the primary pump station (Item 1) was used.

10. Design work: The cost of this item was estimated to be 9% of the total construction cost, items 1 to 9 (The Institution of Professional Engineers NZ, 2004).

3.5.4 Summary

Two methods were used to evaluate the capital cost of the Bardenpho. The first method made use of the common components of the MBR and the Bardenpho, for example pumps, blowers and tanks, from the MBR contract and adjusting these component costs by a correcting factor using the power law equation. The second method made use of the relevant items and costs of the Bardenpho from the Rotorua District Council asset management plant figures. Table 14 presents the capital cost values obtained by using the two methods of calculation.

Table 14: The Bardenpho capital cost comparison

Method	Capital Cost ($)
Bardenpho estimated capital cost using the power law equation	15,459,464
The Bardenpho estimated capital cost using the RDC asset plan	14,179,207

The cost difference between the two methods is $ 1,280,257. In order to compare the Bardenpho capital cost with the MBR capital cost, the average cost of both methods of calculation was used (Table 15).

Table 15: The Bardenpho system capital cost estimate

Description	Capital cost ($)
Bardenpho treatment system	14,819,336

Chapter 4: Rotorua's Membrane Bioreactor Plant Description

4.1 Introduction

The MBR plant at the Rotorua wastewater treatment plant was added to provide an increased capacity and improve treatment qualities. The flow through the MBR was designed to be 7,300 m^3/day during normal operation allowing three hours of maximum flow of up to 11,000 m^3/day. Before the construction of the MBR in May 2012, all of the influent wastewater flow was directed to the Bardenpho.

The MBR was constructed in an unused clarifier with two concentric rings (Figure 34). The outer ring was separated into two zones, an anoxic zone (430 m^3) and an aerobic zone (900 m^3). The inner circle contains an additional anoxic zone (430m^3) and a sealed unit to accommodate the membranes (249m^3). After the membrane zone is the de-aeration zone (330m^3).

Figure 34: Existing clarifier was used to host the MBR system

Top view of the MBR with the different zones and their flows is presented in Figure 35.

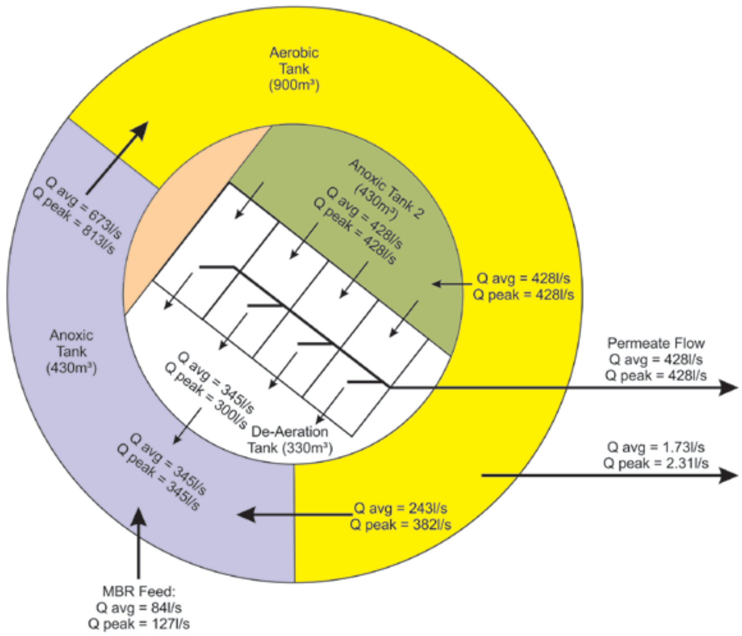


Figure 35: Rotorua MBR top view and flows

The MBR system (Figure 36) includes two feeds, a fine screen (light blue), biological reactors, recirculation pumps, the membranes, coarse and fine blowers (orange), chemical dosing arrangement (purple) and permeate pumps and tanks (dark blue).

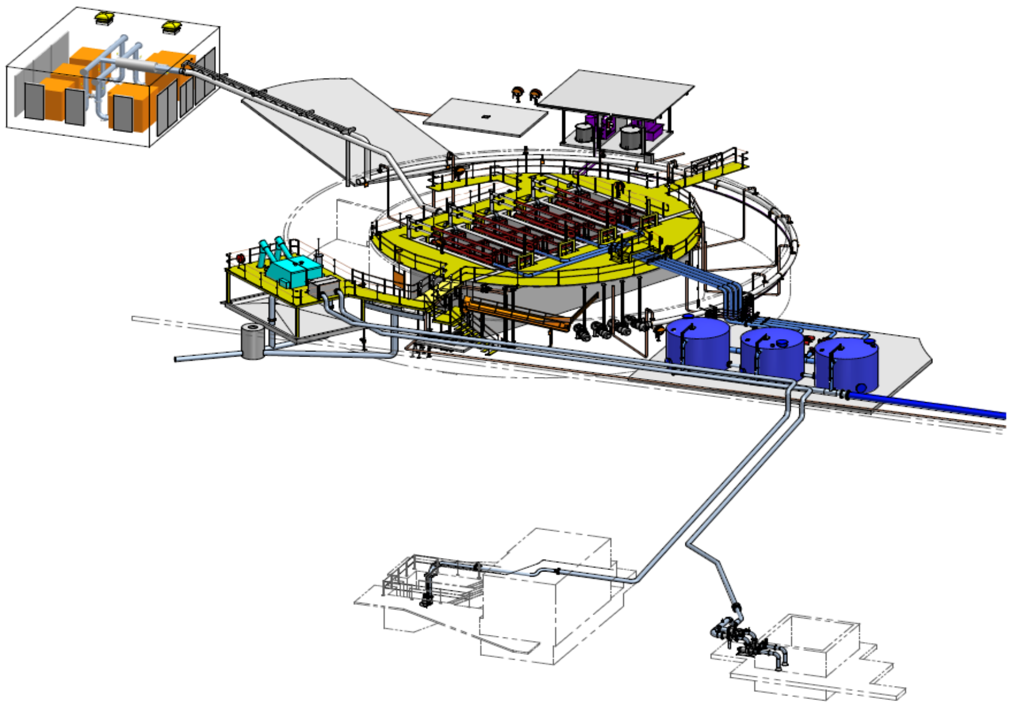


Figure 36: Rotorua MBR configuration (taken from the MBR construction drawings - RDC)

4.2 Rotorua Membrane Bioreactor process description

The membrane bioreactor process is illustrated on Figure 37. The influent coming to the plant is primarily passed through a 3 mm screen and a grit trap which are common systems to both the Bardenpho and the MBR processes. The first feed to the MBR is taken after the grit trap. This feed, called preliminary feed, is high in organic carbon and can be used to reduce ethanol dosing later in the process. The second feed to the MBR is sourced after the primary clarifier; this feed is called primary feed.

Both of the preliminary and primary feeds are pumped to the MBR process via separate pumps controlled by variable speed drivers. It is possible to feed the MBR from each of the feeds individually or combined whilst the remaining wastewater is treated in the Bardenpho process.

The two preliminary and primary wastewater streams undergo a secondary screening stage using a three mm screen. After passing through the screen the influent flows to the first anoxic tank. From the anoxic tank the inflow continues to the aerobic tank, which is fitted with bubble diffusers to create an environment rich in oxygen.

From the aerobic tank the mixed liquor is pumped to the second anoxic tank using the mixed liquor suspended solids (MLSS) pumps. From the second anoxic zone the liquid gravitates into the membranes tanks.

With the use of the permeate pumps the effluent (permeate) is forced out of the membranes delivering the treated effluent into three storage tanks. The permeate pumps can work in reverse using the effluent from the storage tanks to backwash the membranes. From these tanks the effluent gravitates to a receiving manhole where it joins the effluent coming from the Bardenpho process.

The sludge from the membrane zone drains into a chamber where it can be pumped to the de-aerated tank. Untreated liquid from the de-aerated tank passes to the first anoxic cell through an opening controlled by actuated penstock valves.

The line connecting the membrane outlet and the permeate pumps is also joined to a citric acid and sodium hypochlorite dosing system. These chemicals can be used to clean the membrane when required. To chemically clean the membrane using either the citric acid or the sodium hypochlorite, the permeate pumps are reversed, this way the chemicals are injected to the permeate line via a set of booster pumps and mixed with the permeate to clean the membranes.

An ethanol dosing system is connected to the second anoxic tank to provide supplementary carbon for the biological process. This ethanol dosing arrangement is common to both the MBR system and the Bardenpho.

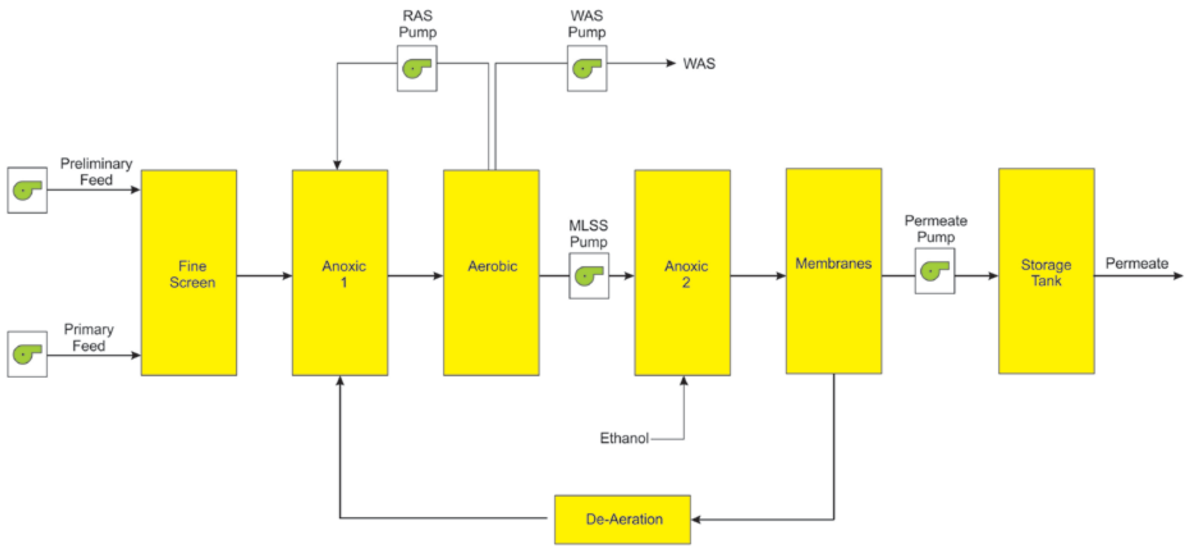


Figure 37: MBR process diagram

4.3 The MBR facilities

This section explains in more detail the major component within the MBR process. These components will be used in the results chapter to evaluate the total running cost of the MBR.

4.3.1 Preliminary Feed Pump

The preliminary pump is a single submergible NP3153.181 LT623, 9kW Flygt pump (Figure 38). The preliminary pump transfers preliminary treated water to the MBR screen. This feed increases the carbon matter required in the biological process, thereby reducing the amount of external carbon needed to be added later on in the process therefore reducing the operating costs of the treatment plant.

Figure 38: MBR Preliminary wastewater feed pump

The preliminary feed pump is programmed to normally operate according to the water level in the pump well and / or the water level in the first anoxic cell. The pump will generally stop whenever the water level in the wet well is below pump start level, the first anoxic cell water level exceeds a set high water level, no membrane tank is active or the MBR screen in not available.

The flow coming out of the preliminary pump is monitored by a magnetic flow meter and can be used and viewed in the operator’s control room. The set points that control the pump, as well the pump driving frequency can be changed to revise the preliminary feed flow rate going to the MBR system.

The preliminary feed flow during this work was set to maintain a flow rate of 50% of the total flow entering the MBR. The maximum flow entering the MBR from the preliminary feed was set not to exceed the maximum capacity of 65 l/s. The total flow entering the MBR can vary and is dependent on the total flow entering the wastewater treatment plant. When the pump is on auto mode the pump duty point changes automatically and is based on continuous monitoring of the flow entering the treatment plant, including the MBR.

4.3.2 Primary Feed Pump

The primary feed consists of two dry mounted NP3153.181 LT623, 9kW Flygt pumps (Figure 39). The primary pumps transfer primary treated water (same primary effluent pumped to the Bardenpho) to the MBR screen. The pumps are submerged in a well situated after the primary settling tanks.

Similar to the preliminary feed pump, the primary feed pumps are programmed to normally operate according to the water level in the holding tank and the water level in the first anoxic cell. The pumps will generally stop whenever the water level in the holding tank is below the pump start level, the first anoxic cell water level exceeds a high water level set point, no membrane tank is active or the MBR screen is not available.

The pumps are driven by a variable speed driver and can operate together or separately. Normally the pump will operate in a duty/assist configuration. The primary feed flow during this work was set to maintain 18 to 45% of total flow entering the MBR. The total flow entering the MBR was set not to exceed the maximum design capacity of the MBR of 11,000 m^3/day. When the pump is on auto mode, the pumps duty point will change automatically and will be based on continuous monitoring of the flow entering the treatment plant, including the MBR.

Figure 39: MBR Primary wastewater feed pump

4.3.3 MBR Screen

The screen used for theRotorua MBR is a rotary drum screen with 2 mm circular holes, manufactured by SAVI S.r.l. The type of the screen is Spiramatic and the model is VSA-1000-T with 1.1 kW motor (Figure 40). The maximum flow rate of the screen is 345 m^3/h (95.83 l/s), which is 75% of the maximum flow of the MBR plant.

The screen drum rotates around its own axle, conveying the screening waste into a hopper placed in the center of the drum. During this process particulates bigger than 2 mm are separated from the wastewater allowing filtered influent to enter the MBR. After separation the solid particles are dehydrated and discharged into a container and disposed into the landfill. The screen needs to be washed periodically to operate without blockage.

Figure 40: MBR screen

4.3.4 First anoxic tank

After screening, the wastewater is gravity fed to the 430 m^3 first anoxic tank. The anoxic tank provides limited oxygen for the denitrification process to occur (Figure 41). The tank includes an ITT Flygt SR4640.411 304SS; 3.3 kW mixer to keep the solids in suspension. Normally this mixer will operate continuously.

The first anoxic tank also receives recycled flow from the aerobic tank via the RAS pump (section 4.3.6) and from the de-aeration tank (section 4.3.14).

Figure 41: MBR first anoxic tank

4.3.5 Aerobic tank

The 900 m^3 aerobic tank is rich in oxygen to allow biological nutrient removal to occur (Figure 42). Oxygen is introduced to the tank via two fine bubble blowers (maximum flow rate of air 5593 Nm^3/hr). The blowers operate in a duty/assist configuration and controlled by VSD. Normally the blowers will operate to maintain set dissolved oxygen (DO) level of 2 mg/L measured by two probes within the aeration tank.

Figure 42: MBR aerobic tank

4.3.6 Return activated sludge pumps

The return activated sludge (RAS) pumps consist of two ITT Flygt PP4640.411 304SS; 2.5 kW submergible pumps (Figure 43). The RAS pumps recycle the oxygenated liquor from the aerobic tank into the first anoxic tank to increase the biological nutrient removal efficiency. The two RAS pumps are driven by VSDs to allow adjustment of the recycle rate as necessary for an optimum operation of the system.

The RAS pumps are normally programmed to operate continuously and will only stop when both the feeding pumps are off or when the level in the aerobic tank is below a set pump start level.

Figure 43: MBR RAS pumps

4.3.7 Waste activated sludge pumps

The waste activated sludge (WAS) pumps consist of two ITT NP3085.183 SH256 2.4kW submergible pumps. The WAS pumps are installed within the aerobic tank and are responsible for removing the excess sludge from the biological system to maintain a healthy and balanced biomass in the MBR. A WAS concentration of 5000 mg/l is removed by WAS pumps at an average flow of 180 m^3/d

The WAS pumps are fitted with VSD's to enable a change of the time the pumps operate. The operation of the WAS pumps will be determined by the desired volume of WAS to be removed.

4.3.8 Scum harvester

The scum harvester (Figure 44) is located at the end of the aerobic tank to collect and remove floatable solids (scum) from the surface of the tank. The scum harvester is programmed to operate 20 minutes and stop for 10 minutes.

Figure 44: MBR scum harvester

4.3.9 Mixed liquor suspended solid pumps

The mixed liquor suspended solid (MLSS) pumps consist of four ITT PP4640.411 304SS; 2.5kW submergible pumps (Figure 45). The MLSS pumps are installed within the aerobic tank and are responsible for transferring the mixed liquor from the aerobic tank into the second anoxic tank, as part of the biological nutrient removal.

Each of the MLSS pumps is fitted with VSD’s to allow the alteration of the rate of flow into the second anoxic tank as required. The pumps operate in a duty / duty / duty / standby configuration. The duty and standby roles normally are alternated during the operation of the pumps. Under normal operating conditions the MLSS pumps are designed to operate continuously at a fixed rate of 37000 m^3/d.

Figure 45: MBR MLSS pumps

Table 16 summarises the average MLSS characteristics going into the second anoxic tank within the MBR process.

Table 16: MBR MLSS characteristics

Parameters	Average Level (mg/L)
Alkalinity	156.50
Total Phosphorus	125.08
Ammonia	3.58
Chemical oxygen demand (COD)	5649
Total Suspended Solids (TSS)	5328
pH	6.73

Data was taken from RDC WWTP Laboratory from the 18/4/2012 until 19/12/2012

4.3.10 Second anoxic tank

The water from the aeration tank is pumped using the MLSS pumps to the second 430 m^3 anoxic tank. Similar to the first anoxic tank, the second anoxic tank has reduced oxygen levels to allow for the denitrification process to occur. The tank contains an ITT Flygt SR4640.411 304SS; 3.3 kW mixer keep the solids suspended.

4.3.11 Membranes and membrane tank

Mixed liquor is fed to the membrane system from the second anoxic tank. The Rotorua membrane system contains a series of GE ZeeWeed 500d hollow fibre membranes with a nominal pore size of 0.04 micron. The membranes are grouped into modules (Figure 46). Each module contains thousands of individual fibers which are connected to a common permeate manifold at the top of each module. Each module has a membrane surface area of 31.6 m^2.

Figure 46: MBR membrane module with a surface area of 31.6 m^2

The modules are grouped into cassettes with 48 modules per cassette (Figure 47). Three cassettes can normally fit into one train. In Rotorua, the membrane tank is 249 m^3 and was designed to accommodate four trains, each train containing only two cassettes with room to fit the third if required in the future. For four trains, the total footprint occupied by the membrane plant itself is 12.6 m x 13.5 m.

When all the membrane trains are operating, the membrane operates at a maximum TMP of 55 kPa and average flux of 23 LMH at 7300 m^3/d and a flux of 34.7 LMH during a peak flow of 11,000 m^3/d.

Figure 47: MBR cassette, total membrane area per cassette 226.8 m2

4.3.12 Membrane drain pump

The drain pumps located at the bottom of the membrane tank transfer the liquid from the membrane tank to the de-aeration tank using four ITT NP3102.185 MY463 3.1kW submergible pumps

4.3.13 Membrane Aeration system

Effective membrane air scouring is critical for continuous performance of the membrane filtration system. The aeration of the membrane is required for the following reasons:

1. Promotes liquid circulation past and around the fibers.
2. Provides mechanical movement of the fibers.
3. Reduces localized solids accumulation.
4. Ensures that the process tank is completely mixed.

In Rotorua each of the membrane modules is connected to a coarse bubble aeration supply (Figure 48) to scour the membrane surface.

The three coarse bubble MBR blowers are located in the “Blower Room” together with the aeration tank blower. The Robuschi blowers supply low pressure air into a common pipe to all of the membrane trains. The blowers are controlled by a VSD which allows different air flow rates, which is dependent on the number of active membrane cassettes, typically 3574 Nm^3/hr when all the cassettes are operating.

Figure 48: MBR coarse bubble aeration system

4.3.14 De-aeration tank

A portion of the flow entering the MBR tanks is recycled into the de-aeration tank via an overflow arrangement (weir) controlled by actuated penstock valves at the back of each membrane train.

The water from the $330m^3$ de-aeration tank is also recycled using an overflow arrangement to the first anoxic tank. The de-aeration tank provides an oxygen free environment and contains an ITT Flygt SR4640.411 304SS; 3.3 kW mixer.

4.3.15 Permeate pumps and tanks

The permeate pumps are Borger CL 390 SK 32, 9.2 kW positive displacement, rotary lobe pumps (Figure 49) which can draw a vacuum to pull permeate through the membranes in normal operation, or generate pressure to push permeate out through the membranes to remove fouling on the membrane surface during cleaning. A flow meter on the discharge of each pump is used to measure the permeate flow and the backwash flow.

Figure 49: The MBR permeate pumps and tanks

The permeate flow can be changed using a number of pumps changing the driver speed. Normally the permeate flow will vary according to liquid level in the de-aeration tank, as the level increases, the pumps will increase speed and produce more flow.

Table 17 summarises the average permeate characteristics.

Table 17: MBR Permeate characteristics

Parameters	Average Level (mg/L)
Alkalinity	74.11
Total nitrogen	4.72
Total Phosphorus	2.66
Ammonia	0.30
$cBOD_5$	0.40
TKN	1.27
Chemical oxygen demand (COD)	16.48
Total Suspended Solids (TSS)	0.99
Total organic carbon (TOC)	9.18
Organic nitrogen	0.97
E.coli	10.86[1] (cfu/100ml)
Nitrate	4.21
Turbidity	0.44 (NTU)
pH	7.2

Data was taken from RDC WWTP Laboratory from the 23/3/2012 until 20/12/2012

[1] Value affected by a few spices of high counts

4.3.16 Chemical dosing

Part of the membrane cleaning includes the use of chemicals to prevent membrane fouling. The dosing point of the chemicals is connected at the permeate pipes (Figure 50), and are used for cleaning the membranes when the permeate pumps are reversed.

Figure 50: MBR chemical injection points (citric acid on the right and hypo on the left)

4.3.16.1 Sodium hypochlorite dosing system

12.5% sodium hypochlorite (hypo) is used to remove the organic fouling that may affect the performance of the membranes. The sodium hypochlorite system (Figure 51) consists of three booster pumps, a hypo tank, feed lines and valves.

Figure 51: MBR sodium hypochlorite dosing system

4.3.16.2 Citric acid dosing system

50% citric acid is used to remove any inorganic fouling that may affect the performance of the membranes. The citric acid system (Figure 52) consists of three booster pumps, a hypo tank, feed lines and valves.

Figure 52: MBR Citric acid dosing system

4.3.17 Ethanol dosing

The MBR system shares the same ethanol tank as the Bardenpho system. The MBR requires ethanol dosing to supplement carbon within the biological tanks to increase cell growth and nutrient removal. Ethanol is supplied through an ITT NP3102.185 LT 413 7.5 kW dosing pump at a flow rate range of 3.5 L/hr. The volume of ethanol is set manually according to the level of nitrate in the permeate flow and will vary during different times of the day.

4.4 Summary of the MBR components

To be able to breakdown the running costs used to operate the MBR, Table 18 presents the power rating of the power consumers within the MBR process. The power rating of each item will be used, together with the operating hours, to show the running costs of the MBR process and its components.

Table 18: MBR components parameter summary

Description	No of items	Power rating[1]	Flow [2]	Pressure [2]
Preliminary feed pump	1	5.9 kW	244 m^3/h	4.1 m
Primary feed pump x 2	2	5.9 kW	244 m^3/h	4.1 m
MBR screen x 2	2	1.1 kW	345 m^3/h	-
De-aeration mixer	1	3.3 kW	-	-
Anoxic 1 mixer	1	3.3 kW	-	-
Anoxic 2 mixer	1	3.3 kW	-	-
Biological fine bubbles blower	2	88 kW (at 50 Hz) 45 kW (at 35 Hz)	4058 Nm^3/h	450 mbar
Membrane coarse bubbles blower	3	63 kW (at 50 Hz) 19 kW (at 19 Hz)	3574 Nm^3/h	360 mbar
RAS pump x 2	2	2.6 kW	756 m^3/h	0.5 m
MLSS pump x 4	4	2.4 kW	583 m^3/h	0.6 m
WAS pump x 2	3	1.8 kW	30 m^3/h	8.70 m
Permeate pump	4	9.2 kW	36 – 85 m^3/h	-7 – 80 m
MBR drain pump x4	4	2.1 kW	91 m^3/h	5.4 m
Blowers fan	11	0.4 – 2.2 kW	-	-
Scum conveyor	1	0.88 kW	-	-
Scum pump	1	1.5 kW	-	-
Chemical dosing	2	0.25 kW	-	-

[1] The power rating was taken from the manufacturer performance curve when available; when the curve was not available the size of the motor was used

[2] Best efficiency point at 50 Hz

4.5 MBR capital costs

Rotorua District Council (RDC) is the owner and is responsible for the operation of the MBR. From the initial design stage of the MBR project, RDC engineers took part in the MBR project management which included the design, purchase and construction of the MBR systems.

The major MBR components were tendered and purchased separately to make sure the preferred items were selected and comparative prices were achieved. The total capital cost of the MBR plant and the breakdown of the components are presented in Table 19.

Table 19: MBR capital cost

Item	Component	Cost ($)
1	Design work	1,070,239
2	Membrane package	2,051,000
3	Blowers and diffusers	436,854
4	Pumps and mixers	195,336
5	SCADA and control	194,722
6	Construction and Installation	4,571,827
	TOTAL CAPITAL COST	8,519,978

Note: SCADA = Supervisory control and data acquisition

The item breakdown is presented in Appendix 3

The general description of the scope of work of each item included in Table 8 is presented below:

1. Design work: The cost of this item included a feasibility study, preliminary design, detailed design, contract management and supervision.

2. Membrane package: The cost of this item included review of the design and functionality of the MBR work, membranes and supporting structure, permeate pumps, inlet fine screen, membrane diffusers, membrane spare parts, inspecting and monitoring of installation work, operator training and membrane guarantees.

3. Blowers and Diffusers: The cost of these items includes the supply of the fine diffusers installed in the aerobic tank, two fine blower and three coarse blowers.

4. Pumps and mixers: The cost of these items includes the supply, manufacture and delivery of MLSS pumps, RAS pumps, anoxic and de-aeration tanks mixers, WAS pumps, MBR feed pumps, ethanol dosing pump and the MBR drain pumps

5. SCADA and control: The cost of these items includes the upgrade user license, configuration of SCADA screens and reports, integrations with PLC, supply of PLC hardware and implementation, testing and commissioning

6. Construction and installation: The cost for this item included structural and civil work involved with modifying an existing old clarifier to suit MBR process, installation of membranes, pumps, mixers, screen, blowers, chemical dosing systems and pipe work, connection of the MBR plant with the existing plant and the installation and integration of monitoring equipment, power supply, control and electrical work.

4.6 Membrane operation parameters and cleaning regime

The MBR system is controlled automatically by the programmable logic controller (PLC). The PLC is programmed to perform various operations, such as a cleaning regime and alarms, according to set point and calculated parameters.

4.6.1 Membrane cleaning regime

During the operation of the membranes there are several cleaning modes that are crucial for good performance of the MBR system. Every cleaning mode has a set time of operation, normally controlled automatically by a set of operation points programmed in the PLC.

The cleaning modes information and the set points are described below and summarised in Table 20.

1 Backpulse – In the backpulse mode the permeate flow is reversed back through the membrane for a period of 30 seconds to allow cleaning of the membrane without the use of chemicals.

2 Relax – In the relax mode the permeate pumps are stopped, no flow through the membranes, the membranes are allowed to relax with no pressure or vacuum for a period of 60 seconds. During the relax mode, membrane aeration will occur to scour the membrane surface and remove accumulated solids. Similar to the backpulse mode, the relax mode also allows physical cleaning of the membrane without the use of chemicals.

3 Maintenance clean – In the maintenance mode the use of either sodium hypochlorite or citric acid is used to clean the membrane. Cleaning of the membrane using citric acid is programmed to occur every seven days for a time period of 60 minutes. Cleaning the membrane using sodium hypochlorite is programmed to occur every 3.5 days for a time period of 60 minutes.

During the maintenance clean the following steps will occur:

i. Aeration of the membrane tank

ii. Initial chemically enhanced backpulse

iii. Relaxation period

iv. Second chemically enhanced backpulse

v. Repeat of steps i to iv

vi. Backpulse with no chemicals

vii. Production

4 Recovery clean – The recovery mode will be controlled manually by the operator , during the recovery mode the operator will select to start recovery clean with the citric acid wait for two hours and start another recovery clean with the sodium hypochlorite. The time interval between cleaning cycles is 180 days for a cleaning duration of 12 hours one clean cycle.

During the recovery clean the following steps will occur:

i. Membrane train is turned off

ii. Aeration of the membrane tank

iii. Drain of membrane tank

iv. Manual clean of membrane with clean water

v. First backpulse with chemical (citric acid or sodium hypochlorite)

vi. Soak membrane in chemical

vii. Second backpulse with chemical (citric acid or sodium hypochlorite)

viii. Soak membrane in chemical

ix. Repetition of the last two steps according to the time set

x. Backpulse without chemicals to fill membrane tank

xi. Soak membrane in chemical, plus aerate membrane tank

xii. Drain of membrane tank

xiii. Backpulse without chemicals to fill membrane tank

xiv. Drain of membrane tank

xv. Restart MLSS to fill the membrane tank

Table 20: Membrane cleaning modes and set points summary

Description	Backpulse	Relaxation	Citric acid	Hypo	Recovery
Interval between cleaning cycle	12 min	12 min	7 day	3.5 day	180 day
Cleaning cycle duration	30 sec	60 sec	60 min	60 min	12 hr
Permeate volume per clean	35.8 L per second	-	6400 L per train	6400 L per train	48000 L per train
Chemical volume per clean	-	-	13.3 L per train	6.85 L per train	387 L per train

4.6.1.1 Membrane aeration

The membrane aeration is produced by three blowers and used to keep the membrane surface clean; course bubble aeration is used for this purpose. The membrane cassette installed in Rotorua can be aerated in two configurations, 10/10 and 10/30 aeration. With the 10/10 aeration each membrane train is aerated for ten seconds and then stopped for ten seconds while in the 10/30 aeration mode the membranes are aerated for ten seconds and then stopped for 30 seconds.

The net airflow in the 10/30 aeration mode is half of the airflow in the 10/10 mode. When the MBR plant is working with four membrane trains in service, the 10/30 mode is normally used for wastewater inflow rates up to 7300 m^3/d and changes to 10/10 if the inflow rate is higher than 7300 m^3/d (Table 21) or if there is an increase in the TMP level or in cases where chemical use is required.

Table 21: MBR aeration modes and airflow rates

Inflow rate	Aeration mode	Membrane airflow (Nm^3/h)[1]
Up to 7300 m^3/d	10/30	1787
Higher than 7300 m^3/d	10/10	3574

[1] Airflow level for four membrane trains with two cassettes in each train

4.6.2 Membrane operation

During the operation of the permeate pumps, permeate water is produced through the membranes. This production period is limited to 12 minutes and is dependent on the time set and the membrane TMP. Normally once that the production cycle is finished a backpulse or relax mode will follow. After the backpulse or the relax mode is completed the production of permeate will start again. This cycle of production and cleaning by backpulse and membrane relaxation is called the production cycle.

The production cycle will continue until a setpoint of high TMP is reached. Once the production cycle is broken, a clean using either citric acid or sodium hypochlorite will occur at the set timing.

4.7 MBR Flow control

The flow entering the MBR can vary quit often and is promotion to the total flow coming into the entire treatment plant. The MBR feeding pump are controlled by a variable speed drivers that is ramped up and down based on the flow coming in to the treatment plant to maintain a balanced flow and contamination entering the MBR treatment process.

Out of the total flow coming into the wastewater treatment plant a portion of the total flow will be treated by the Bardenpho while the other portion will be pumped to the

MBR. The flow entering the MBR plant will use the plant average flow and multiply it by a targeted proportion feed. Normally the MBR preliminary feed flow provides 50% of the total flow entering the MBR system, while the remaining 50% is from the primary effluent feed.

Table 22 presents the range of flows that can feed the MBR during different inflow rates coming into the wastewater treatment plant.

Table 22: The MBR inflow rate

Plant flow range (l/s)	**MBR targeted proportion feed (%)**	**MBR primary feed flow range (l/s)**	**MBR preliminary feed flow range (l/s)**
>450	18	81 - 127	42 - 65
300 – 450	20	60 - 90	30 - 45
160 – 300	35	56 - 105	28 - 52
0 – 160	40	0 - 64	0 - 32

Note: The total flow into the MBR cannot exceed the daily flow of 7300 m^3/d and nor a maximum flow of 11000 m^3/d longer than three hour in a day.

Chapter 5: Methodology

5.1 Introduction

The MBR process is known to have some advantages and deficiencies compared to the traditional activated sludge process. One of the major drawbacks of the MBR is the high running costs (Meng et al., 2009) which were reported to be two to four times higher than the conventional activated process treatment (S. Zhang et al., 2003). On the other hand it is reported that the MBR advantages are a smaller plant footprint and superior effluent quality (Gonzalez et al., 2007; Judd, 2007, 2011; Leiknes, 2010; Visvanathan et al., 2000).

This work analysed actual data from the MBR and the Bardenpho activated sludge processes operating side by side within the Rotorua treatment plant. Ten months of data for the MBR system (April 2012 to January 2013) and thirteen months of data for the Bardenpho system (January 2012 to January 2013) was used to provide information regarding the operational costs and the treatment quality of each treatment process. The data for this study was collected from three different sources and showed the inflow rates, power consumption and working hours of each component within the process and the major contaminants in the wastewater. The use of the MBR capital cost and the Bardenpho estimated capital cost was made to enable a full cost comparison between the two processes. The method used to collect the data will be explained in the following sections.

5.2 Flow monitoring

The flow entering the wastewater treatment plant is measured by six different magnetic flow meters. Three of the meters are located prior to the primary clarifier, monitoring the flow pumped to the clarifiers and the Bardenpho process. The three other flow meters are located at the MBR feeds, two measuring the flow entering the MBR after the primary clarifiers and one that measures the flow entering the MBR system from the preliminary feed (Figure 53).

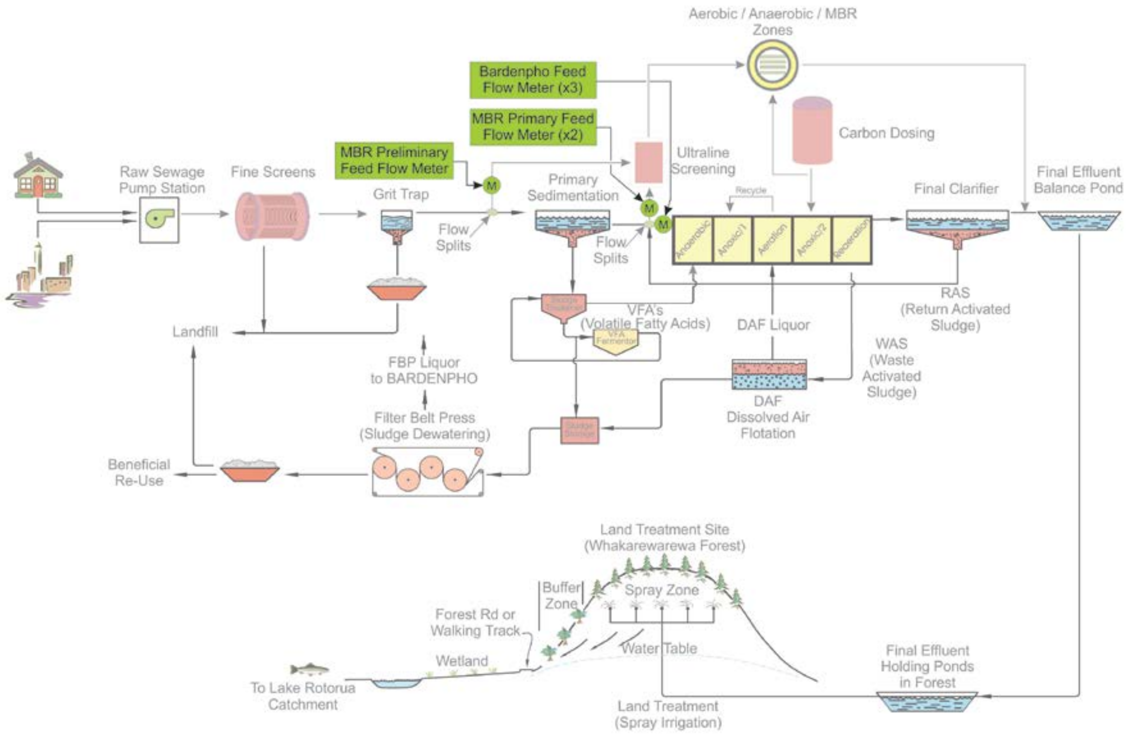


Figure 53: Rotorua wastewater treatment plant flow meter location

5.2.1 Flow data collection

The feed flow to the MBR and the Bardenpho is sent from the flow meter to the SCADA system situated in the operator's control room and can be viewed through the SCADA monitors. The flow data is displayed as the instantaneous flow entering each of the plants, as well as presenting the total daily (24hr) flow from the previous day.

The SCADA system is limited in storage capacity and cannot store the flow data for a long period of time. To be able to store the flow rates, flow rates from the SCADA system are sent to the "Proficy" system that collects and stores the data. The flow data sent from the SCADA system to the Proficy system only includes the total flow entering the entire treatment plant and does not have the breakdown of the flow.

Daily reading of the flow is also recorded by the treatment plant operators. The operators normally start the reading in the morning between 8:00 a.m. to 9:00 a.m. The operator will use a printed monthly table worksheet (Appendix 4) and will handwrite the flow figures presented on the SCADA system screen. The collected data includes the previous day's total flow, the instantaneous flow of the MBR preliminary (Raw) and primary feeds and the daily volume of permeate flow from each of the four MBR trains. Once the worksheet is completed for the month, it is stored in a box at the operation team room. Nine months of handwritten MBR flow data from the worksheets was then transferred onto an Excel spreadsheet for ease of analysis.

The flow recorded in the Proficy system was used to determine the flow rate coming into the Bardenpho. This was calculated by reducing the MBR total feed (taken from the operator worksheets) from the total flow coming into the entire treatment plant. Eleven months of the Bardenpho flow was used to analysed the Bardenpho system.

5.3 Working hours of the process components

To be able to monitor the power consumption of the MBR and the Bardenpho process, as well as to understand the division of power use between the major components in each system, the operating hours of the components within the systems and the power rating were needed. Once the power rating and the operating hours are known, the consumption over a period of time can be calculated by multiplying the power rating by the operating time.

5.3.1 Data collection of working hours

Each of the main components consuming power within both the MBR and the Bardenpho processes are connected to and controlled from a control room. The control room for the MBR system is situated in the main building, while the Bardenpho control room is located in a different building near the Bardenpho system.

In the relevant control room each of the components within the MBR and Bardenpho systems has a working hours meter that is read and recorded daily by the plant operators. Similar to the flow collecting procedure, the treatment plant operators will start recording the working hours in the beginning of the day, handwriting the plant's component working hours on a worksheet (Appendix 4). Due to the high number of power consumer items included in the system, the recording process may continue until the middle of the day and sometimes even towards the end of the day.

The handwritten recorded working hours of the MBR and the Bardenpho componentry was transferred from the operator's completed worksheets onto an Excel spreadsheet for ease of analysis.

5.4 Treatment quality

To be able to compare the performance of MBR treatment process with the performance of the Bardenpho treatment process and also to be able to analyse the cost of treating a specific contaminant, it is necessary to include the test results of the wastewater influent and the treated effluent (or permeate in the case of the MBR).

The Rotorua District Council Environmental Laboratory is located in the wastewater treatment plant and is responsible for monitoring both the MBR and Bartenpho plant performance. The Laboratory is certified by the international accreditation New Zealand (IANZ) and performs tests using the Standard Methods for the Analysis of Water and Wastewater (SMAWW) or United States Environmental Protection Agency (USEPA) methods.

5.4.1 Plant performance data collection

On average samples are taken by the treatment plant operators three times a week in the morning and delivered to the laboratory for analysis. The samples used in this study were taken from the point after the three millimeter screen (raw sewer), from a point after the primary clarifiers (primary effluent) and from the final pond area for the Bardenpho system (final effluent) or from the MBR permeate water (Figure 54).

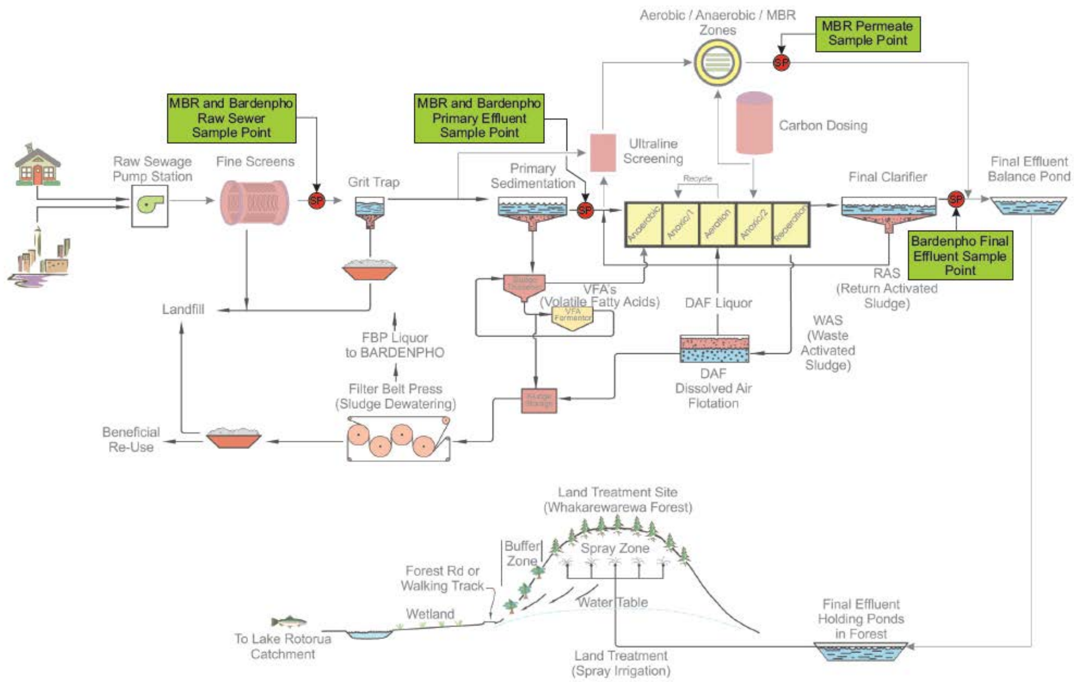


Figure 54: Rotorua wastewater treatment plant sampling location point

The laboratory monitoring consists of a series of tests to determine the operation and performance of the wastewater treatment plant. For the purpose of this study the tests which were used for the comparison of the MBR and the Bardenpho were:

- Total phosphorus (TP)
- Total Kjeldahl nitrogen (TKN)
- Chemical oxygen demand (COD)
- Total suspended Solids (TSS)
- Total oxygenaised nitrogen (TOXN)
- Dissolved reactive phosphorus (DRP)
- Ammonia

After the tests are performed, the data is stored in the laboratory filing system and once the data has been validated it is exported into an excel work sheet and is available to the wastewater treatment plant scientist for further analysis and reporting.

5.5 MBR and Bardenpho capital cost and net present value comparison

The capital cost of both the MBR and the Bradenpho wastewater system was used to allow a full cost comparison between the two systems. For the cost comparison the net present value for both systems was calculated for a plant life time of 30 years (Verrecht et al., 2010), and long term inflation rate of 7.0 %, taking into account the plants capital cost and operating expenditures during the plant lifetime.

5.5.1 MBR capital cost

The total capital cost of the MBR which was used for the net present value comparison is $8,519,978 (Section 4.5). The capital cost included all the costs required to design, build and commission the MBR (Table 19).

5.5.2 MBR operational cost

The MBR operational cost is a composition of the MBR energy demand, sludge disposal, ethanol costs, chemical usage for membrane cleaning, membrane modules replacement and the operation and maintenance costs.

5.5.2.1 Energy demand

The individual contribution to energy of each of the MBR process components, as described in Section 4.4, Table 18 and illustrated in Figure 55, was used together with the working hours of each component to calculate the cost of energy consumed in the process of treating the wastewater using the MBR system. The cost of the consumed energy is the power rating over time multiplied by the tariff rate of twenty New Zealand cents per kW an hour.

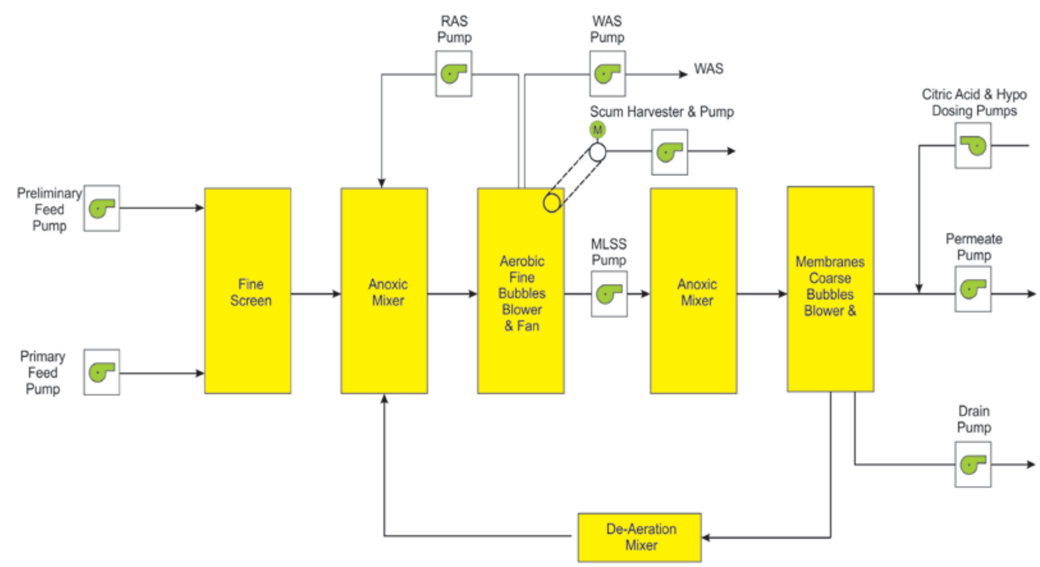


Figure 55: The MBR energy consumer items

5.5.3 Bardenpho capital cost

The total capital cost of the Bardenpho used for the net present value comparison is $14,819,336 (Section 3.5).

5.5.4 Bardenpho operational cost

The Bardenpho operational cost is a composition of energy demand, sludge disposal, ethanol dosing and the operation and maintenance costs.

5.5.4.1 Energy demand

The individual contribution to energy of the Bardenpho system relevant component as described in Section 3.4, Table 11 and illustrated in Figure 56, was used together with the working hour of each component to calculate the cost of energy consumed in the process of treating the wastewater using by the Bardenpho system. The cost of the consumed energy is the power rating over time will be multiplied by the tariff rate of twenty New Zealand cents per kW an hour.

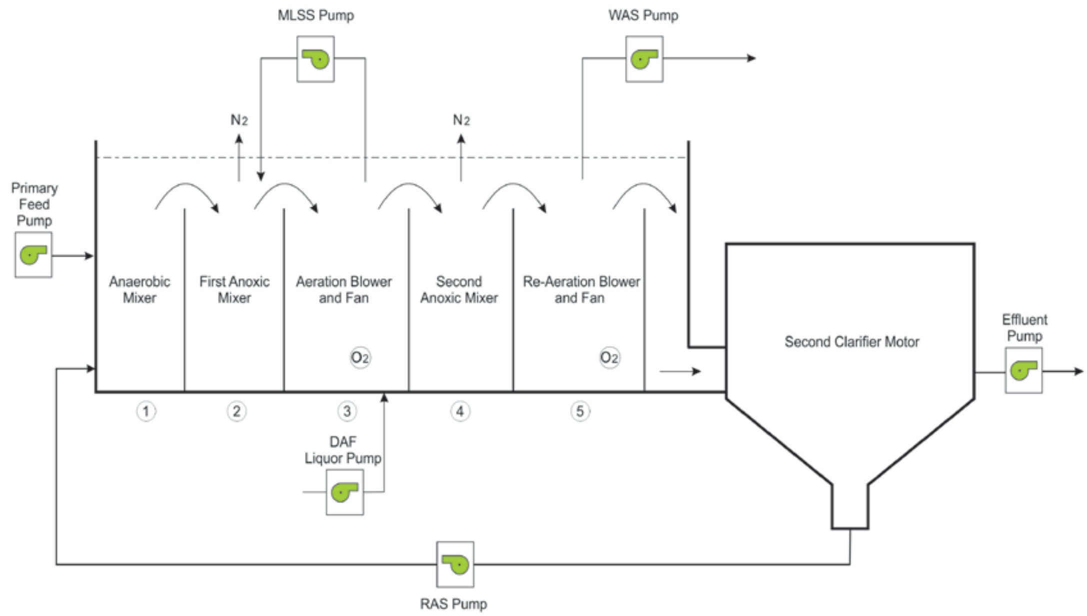


Figure 56: The Bardenpho energy consumer Items

Chapter 6: Results and Discussion

6.1 Introduction

The purpose of this study was to analyse and compare the performance and cost of the membrane bioreactor (MBR) and Bardenpho wastewater treatment systems, working side by side, under similar influent qualities and environmental conditions.

This chapter describes, discusses and compares the data collected over 13 months of operation for the Bardenpho activated sludge plant and 10 months of operation for MBR plant. The findings will be discussed in relation to similar studies completed in Europe, America, Australia and China.

6.2 Wastewater treatment performance

This section presents the performance results of the Bardenpho and MBR plants to address the first research question: "How does the effluent quality of the MBR system and a Bardenpho activated sludge process compare?"

6.2.1 Bardenpho

To evaluate the performance of the Bardenpho, the influent and effluent qualities of the Bardenpho wastewater treatment process were analysed during the time period 30/10/2012 to 31/1/2013. This time frame represents a period when the Bardenpho received a portion of the flow entering the entire wastewater treatment plant. During this time period the flow rates entering the Bardenpho system ranged from 7,466 to 16,496 m^3/d. Table 23 shows the mean and standard deviation for the five different constituents measured on the influent and effluent of the Bardenpho system.

Table 23: Average Bardenpho influent and effluent water characteristics from data collected between 30/10/12 to 31/1/13 at inflow rates of 7,466 to 16,496 m^3/d.

Constituent	Influent concentration (mg/l) [SD[1]]	Effluent concentration (mg/l) [SD[1]]	Percentage removal (%)
NH_4-N	36.11 [2.35]	0.22 [0.73]	99.38
TKN	48.77 [3.27]	2.98 [2.87]	93.90
Total-P	5.97 [0.55]	2.76 [1.95]	60.48
TSS	125.08 [22.43]	24.47 [34.75]	83.58
COD	252.52 [22.43]]	37.30 [36.08]	85.55
DRP	4.04 [0.45]	2.11 [1.64]	61.71
Organic-N	12.66 [2.48]	2.79 [2.48]	78.01
TOXN	-	2.35 [1.79]	-

[1] SD = standard deviation

The Bardenpho influent passes through a preliminary treatment that includes a 3 mm screen, a grit trap and primary sedimentation. During the preliminary treatment grit, suspended and floating debris are removed. The characteristics of the raw sewer are within the lower typical domestic wastewater concentration range of biodegradable organics, total phosphorus and total suspended solids, and midrange for total nitrogen (Crites & Tchobanoglous, 1998; Metcalf & Eddy, 2003). Through the primary sedimentation tanks approximately 50% of the biodegradable organics, 70% of suspended solids and less than 8% of nutrients are removed (Section 3.3.5). These removal ranges are due to a two to three hour retention time in the primary clarifier, and are within the higher range of typical removal rates mentioned in the literature (Crites & Tchobanoglous, 1998; Metcalf & Eddy, 2003).

Since the installation of the MBR system, reduced and better equalised flow has been entering the Bardenpho process. The reduced and equalised flow both contribute to a higher plant performance due to higher contact times between the wastewater and the bacteria, higher sedimentation time at the second clarifier and more uniform flow and concentration of contaminants.

6.2.1.1 Nitrogen removal

When the wastewater enters the Bardenpho it contains an average level of 48.77 mg/l of TKN, with a standard deviation of 3.27. The removal rate of the Bardenpho system during the monitored period was 99% for ammonia and 78% for organic nitrogen. This removal rate of ammonia occurs in the nitrification / denitrification processes taking place mainly within the Bardenpho aerobic tank and two anoxic tanks. This arrangement, together with a recirculation rate of 75% of the flow and dosage of ethanol, manages to remove the majority of the ammonia and convert it into nitrogen gas. The high removal rate of ammonia indicates the existence of a sufficient carbon source (achieved through ethanol dosing), appropriate anoxic conditions and aerobic conditions (achieved through aeration) and good contact time between the bacteria and the influent due to correct tank sizing and the equalisation of flow in the Bardenpho tanks.

The removal rate of the organic nitrogen is completed through two mechanisms; the first is the nitrification / denitrification which removes the organic nitrogen that was converted into ammonia in the anaerobic condition within the Bardenpho system and the second mechanism is through sedimentation, taking place in the second clarifier with an average retention time of 8 hours. Normally the organic nitrogen component makes up a large part of the total nitrogen in the effluent of a traditional activated sludge process (Sattayatewa, Pagilla, Pitt, Selock, & Bruton, 2009). In this study, the Bardenpho final effluent contained 52% organic nitrogen out of the total nitrogen.

Table 23 shows the Bardenpho removal rate of nitrogen for a range of flows from 7,466 to 16,496 m^3/d, with an average flow rate of 11,947 m^3/d. The MBR was sized to allow an annual daily flow of 16,700 m^3/d to enter the Bardenpho plant. To show Bardenpho performance in removing the TKN and organic nitrogen at different inflow rates, Figures 57 to 58 present the performance of the Bardenpho system during a flow range of 19,683 to 30,345 m^3/day.

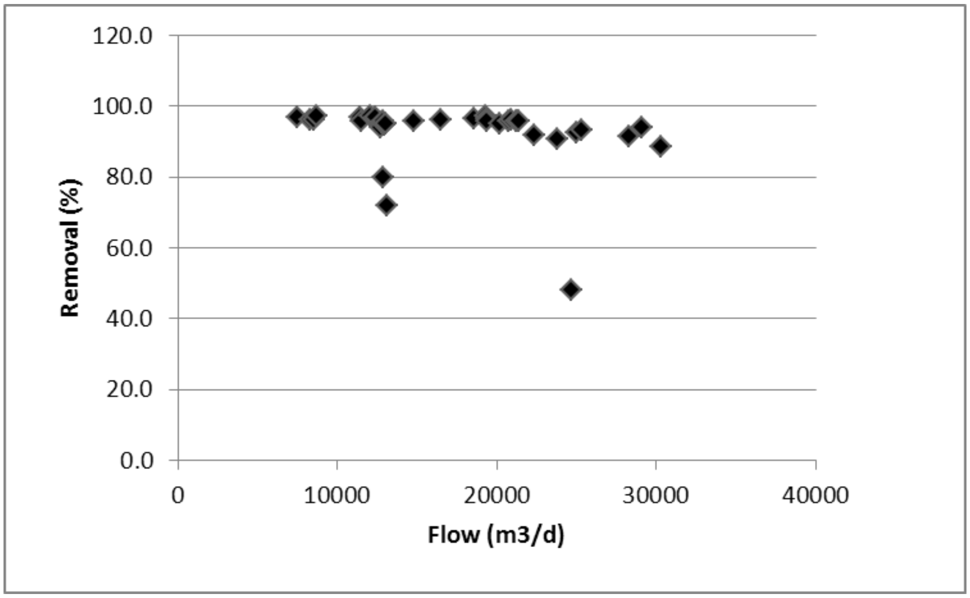


Figure 57: Flow dependent variation of the Bardenpho TKN removal rates (based on data collected from 31/1/12 to 31/1/13)

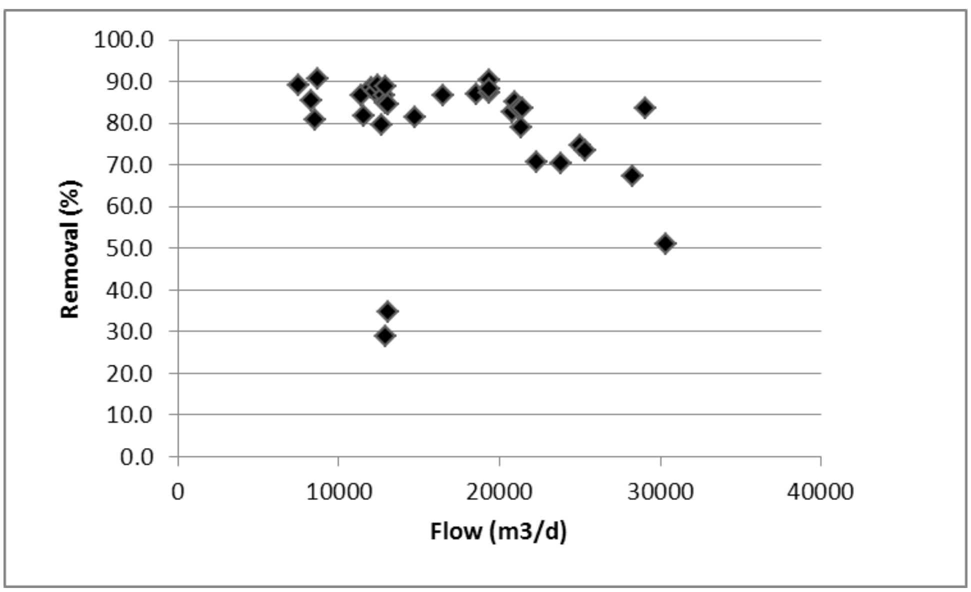


Figure 58: Flow dependent variation of the Bardenpho organic nitrogen removal rates (based on data collected from 31/1/12 to 31/1/13)

Figure 57 and 58 show that the Bardenpho percentage removal of TKN and organic nitrogen decreases when the flow entering the plant increases. This reduction of performance can be related to the reduced retention time during higher flows, which lead to shorter sedimentation time and shorter contact time between the nitrifying and denitrifying bacteria and the wastewater. It shows that up to the inflow rate of 20,000 m3/d the removal of TKN and organic nitrogen is relatively steady, while higher flow rates will reduce the percent removal.

6.2.1.2 Suspended solids

The Bardenpho influent contains average TSS levels of 125.08 mg/l, with a standard deviation of 22.43. The high standard deviation is caused by the inflow and infiltration (I/I) flow entering the sewer network during heavy rain (Crites & Tchobanoglous, 1998). During high I/I flow, the concentration level of solids will be diluted due to the mix of rain and/or ground water with wastewater.

While 70% of the total suspended solids entering the wastewater treatment plant are removed through the primary stage, Table 23 shows that a further 83.58% is removed by the Bardenpho. This removal rate is related to the performance of the second clarifier and its retention time, and the wasting volume out of the Bardenpho system.

Figure 59 presents the performance of the Bardenpho system in removing the suspended solids during a flow range of 19,683 to 29,082 m^3/day.

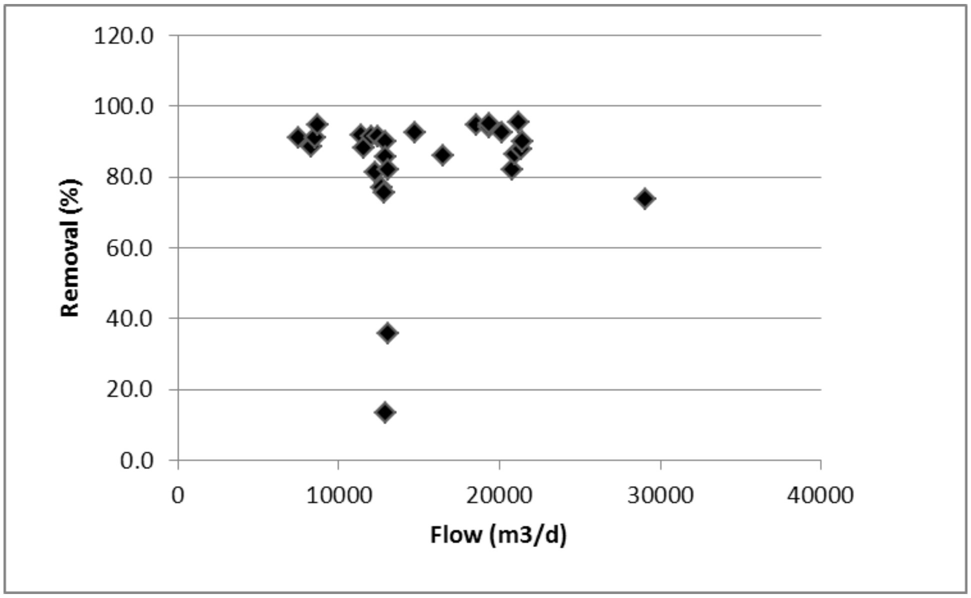


Figure 59: Flow dependent variation of the Bardenpho TSS removal rates (based on data collected from 31/1/12 to 31/1/13)

The Bardenpho system shows a relatively constant removal range between 78 to 90 % during flow up to 20,000 m^3/d, with only one sample at a flow of 29,082 m^3/d with 74% removal. Although there were few flows recorded at rates higher than 20,000 m^3/d, the removal of TSS is expected to drop with the increase of flow due to the reduction in sedimentation time. In Figure 59 there are two points during low flow rates with lower removal efficiencies, these points could be related to a laboratory error where a sample was not mixed well or data collection error.

6.2.1.3 Biodegradable organics

The Bardenpho influent contains average COD levels of 252.52 mg/l, with a standard deviation of 22.43. The biodegradable organics are used as an electron supplier required during the biological removal of nutrients. Within the Bardenpho first stages (anaerobic, first anoxic and aerobic zones), most of the biodegradable organic matter is utilised by the bacteria, therefore organic matter in the form of ethanol is added in the Bardenpho second anoxic tank to ensure continuous removal of nutrients. Measurement of COD level in the effluent indicates that 85.55% of the COD, in reference to the influent COD level, was removed during the Bardenpho biological and waste removal.

Figure 60 presents the performance of the Bardenpho system during a flow range of 19,683 to 21,399 m^3/day. The percentage removal of COD continually decreases when the flow entering the plant increases. This reduction of performance can be related to the reduced retention time during higher flows, which leads to shorter contact time between the bacteria and the wastewater and therefore smaller degradation of COD.

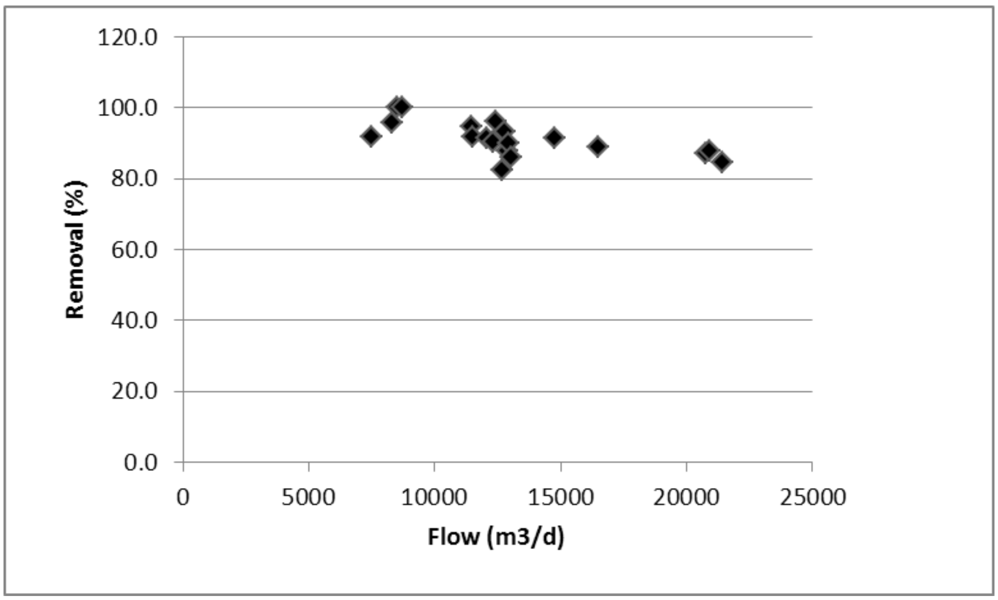


Figure 60: Flow dependent variation of the Bardenpho COD removal rates (based on data collected from 31/1/12 to 31/1/13)

6.2.1.4 Phosphorus

The Bardenpho influent contains average total phosphorus levels of 5.97 mg/l, with a standard deviation of 0.55, with most in the form of dissolved phosphorus with a concentration of 4.04 mg/l and a standard deviation of 0.45. The removal rate of the Bardenpho system during the monitored period was approximately 61% for both forms of phosphorus. This removal of phosphorus is achieved biologically by alternation of anaerobic and aerobic conditions through the Bardenpho process

(Powell, 2009) and the addition of the anaerobic tank (N. I Galil et al., 2008). The phosphorus is stored in the biomass as polyphosphate when taken by the bacteria under aerobic and anoxic condition, accumulating it within the cell. The removal efficiencies of phosphorus can be improved further (up to 90 to 95%) if a precipitate such as alum is used (Metcalf & Eddy, 2003). Once the phosphorus is taken by the bacteria it is removed at the Bardenpho aeration tank through wasting of the biomass. To maintain higher removal rates of phosphorus, it is crucial to ensure aerobic conditions to prevent the release of the attached phosphorus.

Figure 61 presents the performance of the Bardenpho system in removing phosphorus during a flow range of 19,683 to 30,345 m^3/day.

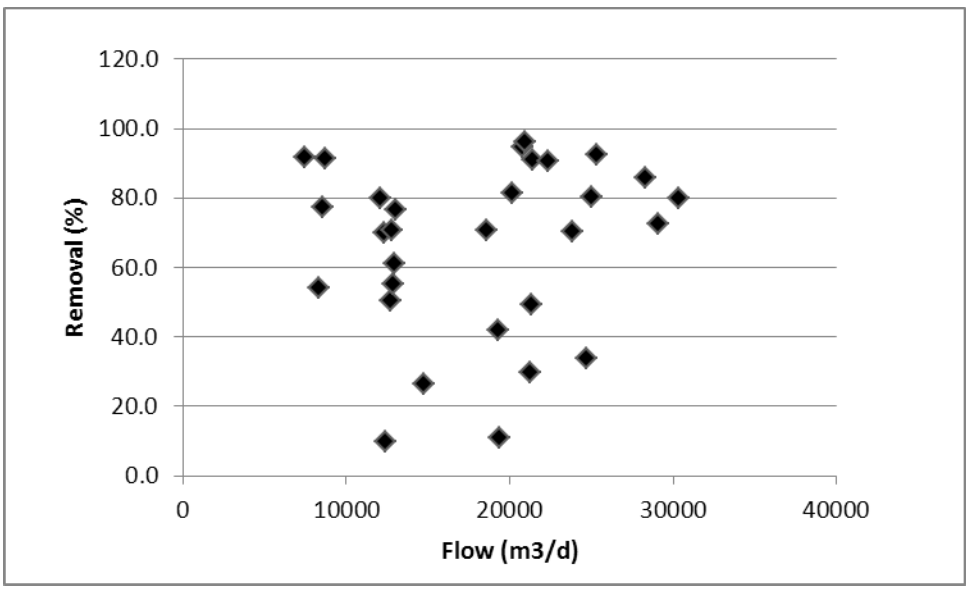


Figure 61: Flow dependent variation of the Bardenpho DRP removal rates (based on data collected from 31/1/12 to 31/1/13)

The percentage removal of DRP within the Bardenpho system varies greatly. There are many factors that can affect the DRP levels in the effluent. High levels of DRP could be related to low MLSS which can create reduced biomass and low aeration and dissolved oxygen levels which cause release of the phosphorus held by the bacteria. Low DRP levels can occur when low wasting volumes are removed from the aeration tank causing higher biomass volumes and age absorbing more phosphorus. Generally it is expected that during low inflow rates the removal of DRP will improve due to the higher hydraulic retention time within the process reactors (Grady et al., 2011).

6.2.2 Membrane bioreactor

The influent and effluent qualities of the MBR wastewater treatment process were analysed during 1/4/12 to 31/1/13. During this time period the flow rates entering the MBR system ranged from 981 to 6,970 m^3/day. Table 24 shows the average monitoring results and standard deviation of the seven different constituents measured on the influent and effluent of the MBR system.

Table 24: Average MBR influent and effluent water characteristics from data collected from 1/4/12 to 31/1/13

Constituent	Influent concentration (mg/l) [SD[1]]	Effluent concentration (mg/l) [SD[1]]	Percentage removal (%)
NH_4	33.98 [2.7]	0.54 [0.82]	98.41
TKN	48.96 [3.53]	2.59 [2.59]	94.52
TSS	252.82 [120.93]	1.09 [0.81]	99.05
COD	319.22 [38.99]]	16.82 [5.71]	94.06
DRP	3.70 [0.57]	1.37 [0.85]	67.43
Organic-N	14.67 [1.78]	1.00 [0.21]	92.75
TOXN	-	2.49 [1.39]	-

[1] SD = standard deviation

The MBR influent consists of two flow feeds, one feed is the same primary influent that enters the Bardenpho system and the second feed is wastewater which has passed through preliminary treatment only. The concentrations shown in Table 24 were calculated by adding the flow from each source multiplied by the concentration of each constituent, divided by the total flow. The MBR constituent removal efficiencies for TKN, TSS, COD and organic nitrogen were over 92%, with TSS particularly high at 99.05%. Only the DRP sat below 92%, with a removal rate of 67.43%.

6.2.2.1 Nitrogen removal

The MBR influent contains an average concentration of 48.96 mg/l of TKN with a standard deviation of 3.53 and 14.67 mg/l of organic nitrogen with a standard deviation of 1.78. The removal rate of TKN through the MBR system during the monitored period was 94.52%. These removal rates correlate with what was found by Galil and Jacob (2009) and but are lower than removal rates of 89 to 92.1% for TKN and 75.6% for organic nitrogen found in studies by Bracklow et al. (2007) and Ersu et al. (2008). The total nitrogen level in the MBR effluent, presented in Table 24, is 5.08 mg/l (made up of a combination of TKN and TOXN). This result is within the range of results found by Verrecht et al. (2008).

Similar to the Bardenpho process, the MBR removal rate of ammonia occurs through the nitrification / denitrification processes (Judd, 2011). The higher nitrogen removal rate results from the MBR higher biomass concentration (Kraume & Drews, 2010), higher sludge retention time (Leiknes, 2011; Van den Broeck, et al., 2012) and the filtration mechanism that is responsible for the removal of the organic nitrogen within the MBR system. Even though the MBR hydraulic retention time changes in relation to the inflow rate, the MBR solids retention time is fixed and set separately to 13 days, to allow high concentration of nitrifying and denitrifying bacteria and therefore higher removal efficiency. The MBR has a sludge concentration of 5000 mg/l (compared to 3000mg/l in the Bardenpho system) which also contributes to the higher removal of nitrogen.

The MBR system was designed to operate on an average daily flow of 7,300 m^3/d, however the flow entering the MBR was programmed to be proportional to the total flow coming into the entire treatment plant. During the time the MBR was monitored the flow entering the MBR ranged from 981 to 6,970 m^3/day with an average flow of

5,520 m^3/day. Figures 62 and 63 present the performance of the MBR TKN and organic nitrogen in removing as a function of flow.

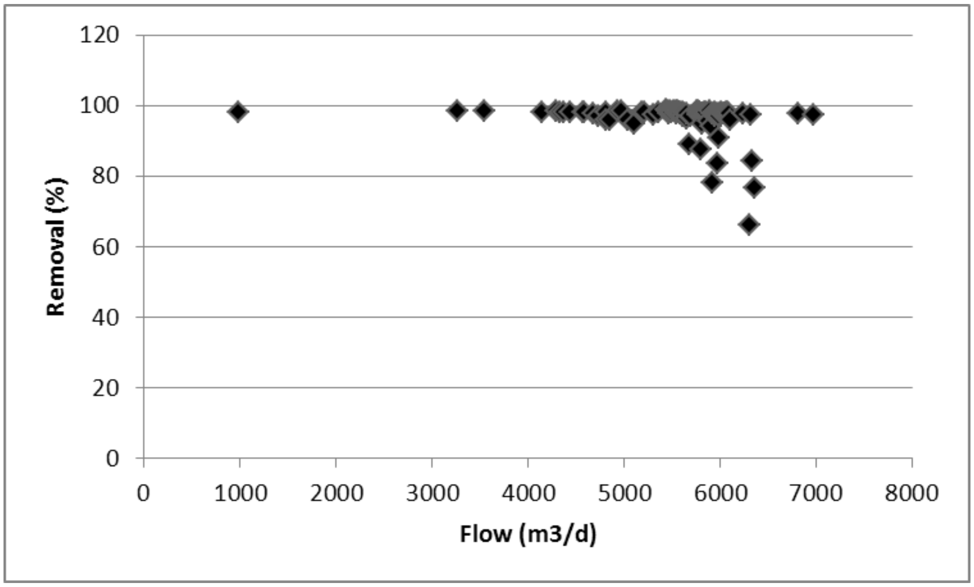


Figure 62: Flow dependent variation of the MBR TKN removal rate (based on data collected from 1/4/12 to 31/1/13)

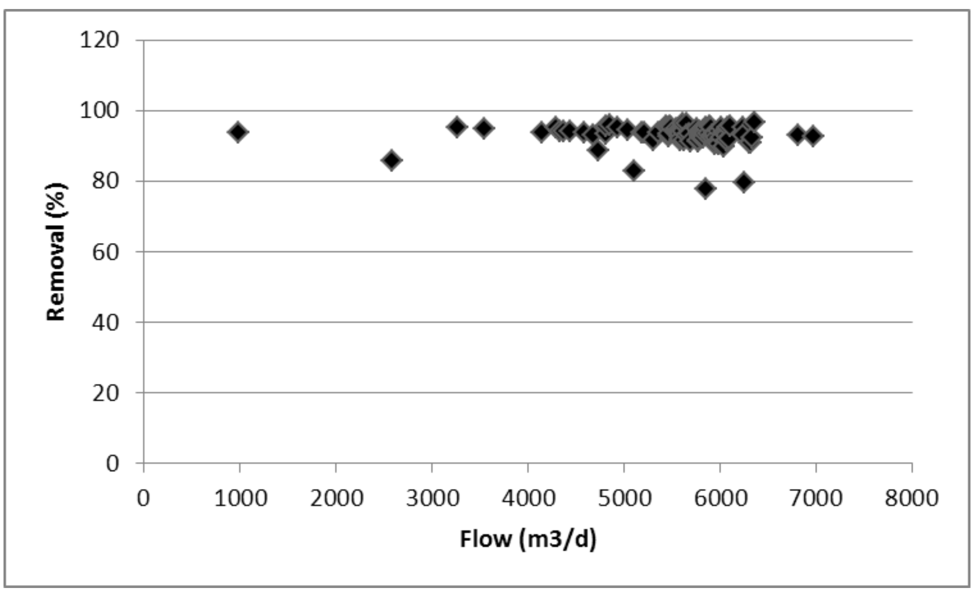


Figure 63: Flow dependent variation of the MBR Org-N removal rates (based on data collected from 1/4/12 to 31/1/13)

Figures 62 and 63 show that the percentage removal of TKN and organic nitrogen is steady and is not dependent on flow or the hydraulic retention time (Pollice, 2008). This performance can be related to the fixed contact time period of the biomass and the influent in the MBR system. Even though the hydraulic retention time changes in relation to the inflow rate, the biomass retention time can be set separately to allow high concentrations of nitrifying and denitrifying bacteria and therefore higher removal efficiency (Van den Broeck, et al., 2012). The results show that the MBR removal rate of nitrogen will be similar in low flow and high flow as long as the carbon source is not limiting and added to the system externally as ethanol. However the cost of running the membrane at a higher flow will increase (Hirani et al., 2010).

6.2.2.2 Suspended solids

The MBR influent contains TSS average levels of 252.82 mg/l, with a standard deviation of 120.93. The higher TSS levels entering the MBR compared to the Bardenpho result from the direct preliminary feed entering the MBR without going through the sedimentation stage. A removal rate of 99.05% was shown for TSS. The MBR removal rate of the TSS is associated with the separation ability of the membrane to filter the solids by physical methods, rather than relying on gravity as in the Bardenpho process. Solids bigger than 0.04 micron, the membrane pore size, will not pass through the membrane into the permeate tanks.

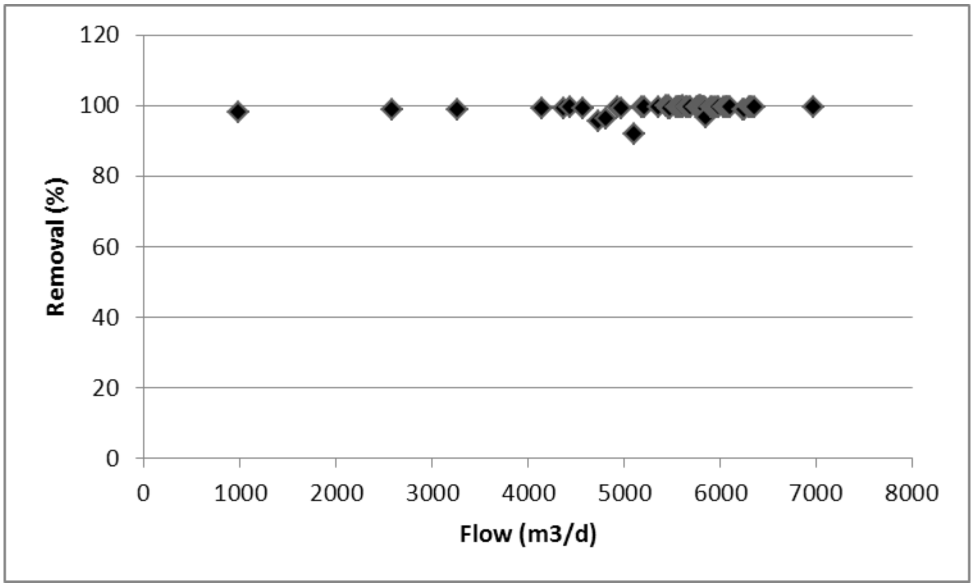


Figure 64: Flow dependent variation of the MBR TSS removal rates (based on data collected from 1/4/12 to 31/1/13)

Figure 64 shows a constant removal of TSS by the MBR system regardless of the inflow rate due to the membrane physical separation mechanism. The increase of flow will not change the removal efficiency of solids, but will increase the transmembrane pressure, the risk of fouling and the need for more regular cleaning (Hirani et al., 2010; Judd, 2011; Leiknes, 2010).

6.2.2.3 Biodegradable organics

The MBR influent contains average COD levels of 319.22 mg/l, with a standard deviation of 38.99. The removal of the biodegradable organics occurs through the biological process of bacteria metabolism. Organic matter in the form of ethanol is added to the process to ensure carbon source it not a limiting factor during the removal of nutrients.

The MBR removal rate results showed COD removal efficiencies of 94.06%, higher than the 92.3% in a study by Ersu et al. (2008) or the removal rate of 88% to 89.7% shown by Skouteris, Arnot, Feki, Jraou, & Sayadi (2012), but lower than 99.6 to 97.9% found by Bracklow et al. (2007). The MBR effluent COD concentration in this study of 16.82 mg/l is higher than the modeled COD level of 4.8 mg/l completed by Verrecht et al. (2008). The relatively high effluent COD concentration is related to the level of ethanol dosing. With careful refining of the ethanol dosage, the COD level could drop lower without affecting the removal performance of the nutrients.

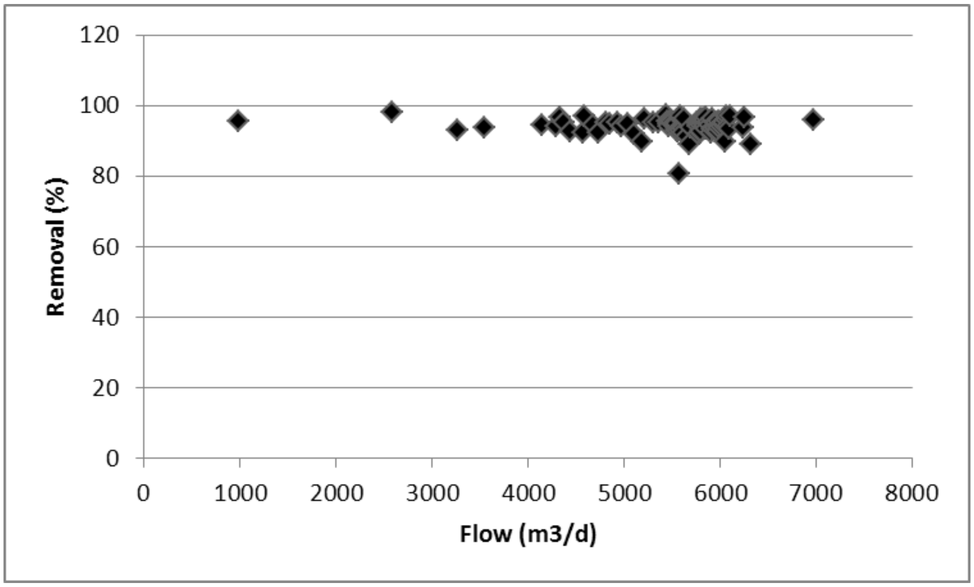


Figure 65: Flow dependent variation of the MBR COD removal rates (based on data collected from 1/4/12 to 31/1/13)

Figure 65 shows that the percentage removal of COD is relatively steady in proportion to the inflow rate. This steady performance is related to the fixed solid retention time and consistent influent nutrient concentration and flows.

6.2.2.4 Phosphorus

The MBR influent contains DRP average levels of 3.70 mg/l, with a standard deviation of 0.57. The DRP removal rate for the MBR system during the monitored period was 67.43 %. Similar to the Bardenpho, phosphorus is removed biologically by enhanced storage in the biomass as polyphosphate promoted by alternation of anaerobic and aerobic condition through the process. Within the MBR system there is the ability to use alum to increase the phosphorus removal rate (Judd, 2011). The removal rate showed in the results is 5% higher than what was achieved by Ersu et al.

(2008) when the MLSS and the permeate flow was circulated back to the anoxic tank, but less than to the 97.4 to 99.4% achieved by Bracklow et al. (2007).

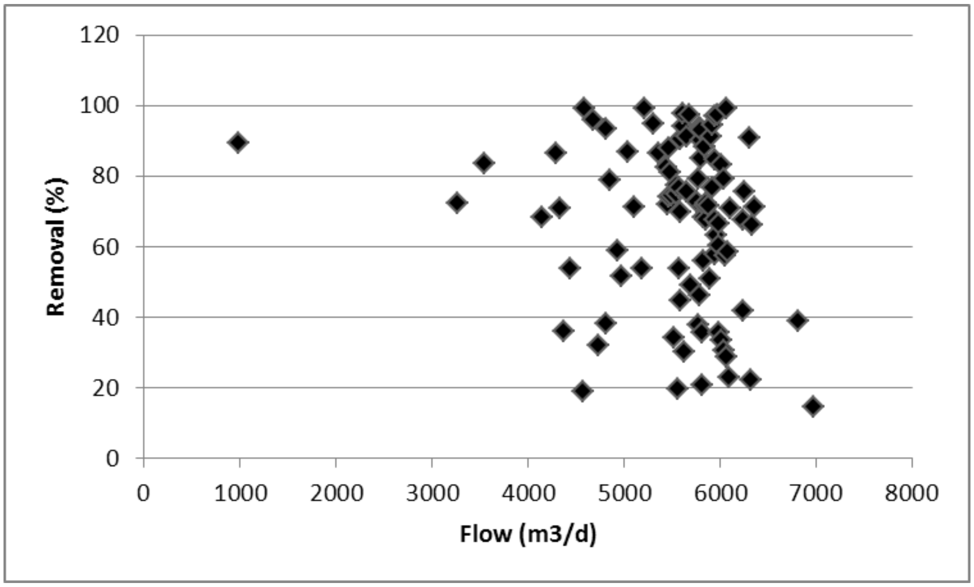


Figure 66: Flow dependent variation of the MBR DRP removal rates (based on data collected from 1/4/12 to 31/1/13)

Similar to the Bardenpho system, the percentage removal of DRP within the MBR system varies greatly (Figure 66). Again, there are many factors that can affect the DRP levels in the effluent. High levels of DRP could be related to low MLSS which can create reduced biomass and low aeration and dissolved oxygen levels which cause the release of the phosphorus held by the bacteria. Low DRP levels can occur when low wasting volumes are removed from the aeration tank causing higher biomass volumes and longer sludge age, which will increase the level of attachment of phosphorus.

6.2.3 Performance comparison

In order to answer the first research question, “How does the effluent quality of the MBR system and a Bardenpho activated sludge process compare?” a performance comparison between the MBR and the Bardenpho was done during 30/10/2012 to

31/1/2013 for the Bardenpho performance and 1/4/12 to 31/1/13 for the MBR. During these time periods, the Bardenpho and the MBR shared the flow entering the entire wastewater treatment plant.

The influent composition entering the Bardenpho and the MBR is not the same. The influent entering the MBR process is composed of 43% preliminary treated wastewater which has passed through a 3 mm screen and grit trap and 57% primary treated wastewater that has exited the sedimentation clarifier. The influent entering the Bardenpho is mostly primary treated wastewater exiting the sedimentation clarifier. For this reason, constituents such as TSS and COD which are reduced during the primary treatment show higher levels in the MBR influent. Table 25 compares seven different constituents measured on the influent and effluent of the MBR and the Bardenpho systems.

Table 25: Bardenpho and MBR influent comparison

Constituent	Bardenpho average concentration (mg/l)	MBR average concentration (mg/l)
NH_4-N	36.11	33.98
TKN	48.77	48.96
Total-P	5.97	5.69
TSS	125.08	252.82
COD	252.52	319.22
DRP	4.04	3.70
Organic-N	12.66	14.67

Table 25 shows that the $NH_{4,}$ TKN, TP and DRP concentrations in both the MBR and Bardenpho are close in value, however the average levels of TSS and COD are 2.02 and 1.26 times higher respectively in the MBR influent than in the Bardenpho influent. This difference in levels between the MBR and Bardenpho is linked to the source of the feed.

The ammonia average level entering the MBR is lower than the average ammonia level entering the Bardenpho, while the average concentration of organic nitrogen is higher in the MBR influent than in the Bardenpho influent. These values show that a portion of the organic nitrogen is converted into ammonia during the primary treatment. However, the total TKN concentration, which is the combination of organic nitrogen and ammonia, is similar for both the MBR and Bardenpho.

The MBR feed was deigned to have higher biodegradable organics and solids to reduce the ethanol cost necessary for the biological removal activities. However, the increase of solids concentration will increase the risk of membrane fouling and therefore the cost of membrane cleaning (Judd, 2011).

Table 26 compares five different effluent constituent concentrations in the Bardenpho and MBR systems.

Table 26: Bardenpho and MBR effluent comparison

Constituent	Bardenpho average concentration (mg/l)	MBR average concentration (mg/l)
TKN	2.98	2.59
TSS	24.47	1.09
COD	37.30	16.82
DRP	2.11	1.37
Organic-N	2.79	1.00

Table 26 shows that all the MBR effluent constituent concentrations are lower than in the Bardenpho effluent, therefore the MBR has better effluent quality, particularly for suspended solids, biodegradable organics and organic nitrogen. To be able to compare the removal efficiencies of both plants it is necessary to calculate the effluent constituents' level in reference to the influent concentration (Figure 67).

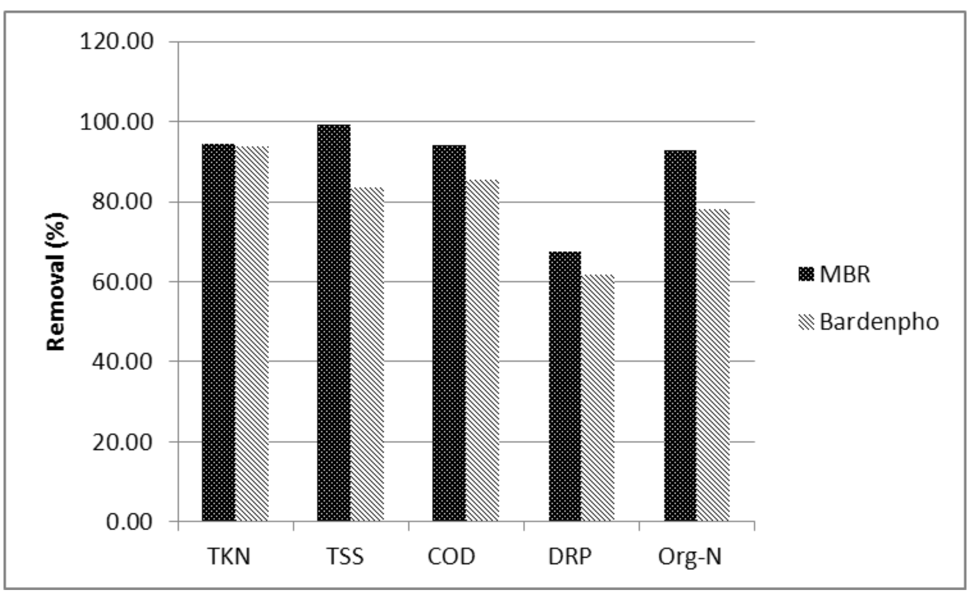


Figure 67: Comparison of Bardenpho and MBR constituent removal efficiencies

Figure 67 shows that the MBR overall removal performance, as expected, is better than the Bardenpho system (Gonzalez et al., 2007; Judd, 2011; Leiknes, 2010), with an increased removal of 0.62 % TKN, 15.47% TSS, 8.51 % COD, 5.72 % DRP and 14.74 % Org-N.

The MBR higher removal rate of the TSS and organic nitrogen is associated with the separation ability of the membrane to filter the solids by physical methods, rather than relying on gravity as in the Bardenpho process. The TSS removal through the Bardenpho system mainly occurs during the secondary clarifier, the clarifier retention time during the measured period was 8.8 hours which is a typical design time period (Metcalf & Eddy, 2003). Higher sludge retention time within the MBR system was found to mineralise the organic nitrogen and convert it into ammonia which was then removed biologically by nitrification / denitrification (Brindle & Stephenson, 1995).

The overall removal rate of ammonia and organic nitrogen (TKN) is 0.62% higher in the MBR than in the Bardenpho. This higher removal rate is due to the superior removal rate of the organic nitrogen in the MBR system in comparison to the Bardenpho system, rather than the removal of ammonia. The five stage Bardenpho showed 99.38% removal of ammonia during an average retention time of 12.5 hours. The MBR showed a slightly lower removal rate of ammonia of 98.41%, achieved by an average hydraulic retention time of 10.4 hours and solid retention time of 13 days.

The COD average concentration entering the MBR and the Bardenpho were 319.22 and 252.52 mg/l respectively. The MBR showed 8.51% better removal of COD than the Bardenpho system. Two main aspects affect the COD level in the effluent, the first is the dosing rate of ethanol and the second is the biological activity within the processes. The ethanol dosing in both the MBR and the Bardenpho is fixed to achieve optimal nutrient removal. The ethanol injected volumes in both the plants was set to be similar per cubic meter of wastewater entering the two systems, however the dosing was set according to the expected design flows which are not always met. The average flow monitored during the investigation period was lower than the design flow in the MBR and higher in the Bardenpho. The MBR SRT was set to 13 days, with a sludge concentration of 5000 mg/l, compared to 3000mg/l in the Bardenpho system. The longer the SRT is the longer the time the bacteria will stay in contact with the wastewater, which leads to better reduction of COD (Pollice et al., 2008; Van den Broeck et al., 2012). The Bardenpho effluent COD concentration is 37.30 mg/l, while the MBR COD effluent concentration is 16.82 mg/l (Table 26). These concentrations suggest that a better ethanol dose rate could lower the COD levels in the outlet of both plants, especially the Bardenpho.

6.3 Cost analysis

In order to address the second research question: “How do the operational costs of the MBR system and a Bardenpho activated sludge process compare?”, the power consumption and the operating hours of each component in the processes were used to calculate the total energy cost and the energy consumption and cost breakdown of both of the treatment systems.

6.3.1 Bardenpho operational costs

The operational costs of the Bardenpho system is a combination of sludge disposal, energy consumption costs and ethanol (operator hours and maintenance costs were not included in this analysis due to a lack of accurate records). The daily average operational cost of the Bardenpho was found to be \$3,898, treating an average flow of 17,233 m^3/d at \$0.22 per cubic meter of wastewater (Table 27).

Table 27: Bardenpho operational costs

Average flow (m^3/d)	**Bardenpho average operational cost (\$/d)**	**Bardenpho operating cost per cubic meter of wastewater (\$/m3)**
17,233	3,898	0.22

To conclude how each item within the Bardenpho system contributes to the total operational costs, six main cost consumer groups were analysed and presented in Figure 68:

1. The sludge disposal group is made up the disposal costs of the dewatered sludge, based on actual cost of \$88 per tonne.
2. The biological aeration group is made up of the energy consumption costs of the blowers and the associated fans.

3. The recycle pump group is made up of the energy consumption costs of the MLSS, RAS and the clarifiers.
4. The effluent pump group is made up of the energy costs of running the primary feed pumps, which deliver primary treated wastewater to the Bardenpho. Because the Bardenpho outflow does not include a pump exclusively for effluent, it was assumed the energy costs for the primary feed pump would be the same as an effluent pump.
5. The Bardenpho feed pump group is made up of the energy costs of running the primary pumps, DAF liquor pumps and storm water pumps.
6. The mixer group is made up of the energy costs of all the mixers existing within the Bardenpho system.
7. The WAS group is made up of the energy costs of the WAS pumps.
8. The ethanol dosing group is made up of the energy consumption costs of the dosing pumps and the ethanol cost of $1.27 per liter.

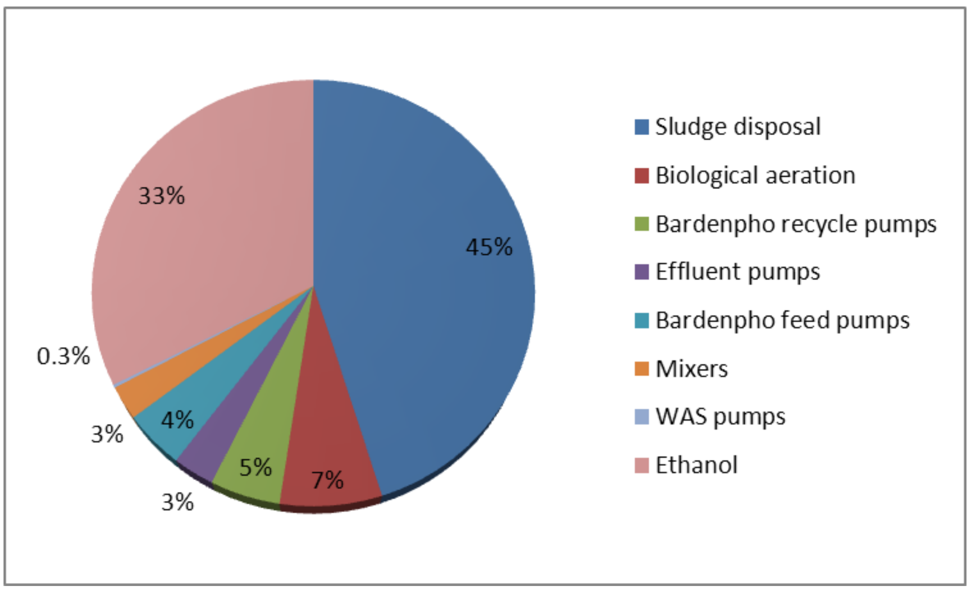


Figure 68: Breakdown of the Bardenpho system operational costs

Figure 68 shows the top two cost contributors in the Bardenpho system are the disposal of the dewatered sludge at 45%, and ethanol costs at 33% of the total operational costs. The remaining cost contributors represent the energy consumer groups which make up 22% of the total operational costs of the Bardenpho. To show how the energy consumption is distributed, Figure 69 presents the power consumption breakdown of the elements within the Bardenpho system.

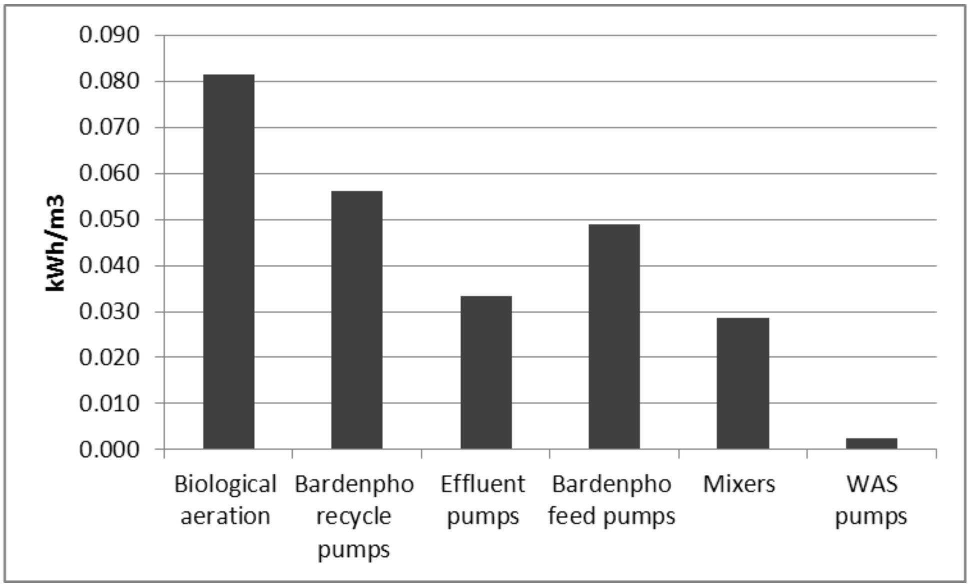


Figure 69: Breakdown of Bardenpho energy consumption components

Figure 69 shows that the aeration used in the Bardenpho consumes 0.081 kWh/m^3, 32.4% of the total energy used, while second highest consumer is the recirculation pumps with a consumption rate of 0.056 kWh/m^3, which is 22.3% of the total energy used. The energy consumption of the other components in the Bardenpho system were: feed pumps with 0.049 kWh/m^3 (19.5%), effluent pumps 0.033 kWh/m^3 (13.3%), mixer 0.029 kWh/m^3 (11.4%) and WAS pumps 0.003 kWh/m^3 (1%).

6.3.2 Membrane bioreactor operational costs

The operational costs of the MBR system is a combination of chemical costs and energy consumption (operator hours and maintenance costs were not included in this analysis due to a lack of accurate records). The daily average operational cost of the MBR was found to be $1,592, treating an average flow of 5,380 m^3/d at $0.29 per cubic meter of wastewater (Table 28).

Table 28: MBR operational cost

Average flow (m^3/d)	MBR average operational cost ($/d)	MBR operating cost per cubic meter of wastewater ($/m3)
5,380	1,592	0.29

To show how each item within the MBR system contributes to the total operational costs, thirteen main cost consumer groups were analysed and presented in Figure 69.

1. The sludge disposal group is made up of the disposal costs of the dewatered sludge, based on actual cost of $88 a tonne.
2. The biological aeration group is made up of the energy consumption costs of the biological fine bubble blowers and the associated fans.
3. The membrane aeration group is made up of the energy consumption costs of the membrane coarse bubble blowers and the associated fans.
4. The recycle pumps group is made up of the energy consumption cost of the MLSS, RAS and drain pumps.
5. The permeate pumps group is made up of the energy costs of running the permeate pump. The MBR permeate pumps can run in both directions, one to produce the permeate and the other to backwash the membrane. The cost recorded under this group includes both of the permeate pump functions.
6. The feed pumps group is made up of the energy cost of the preliminary and primary pumps.
7. The mixer group is made up of all the mixers existing within the MBR system.
8. The pretreatment group is made up of the fine screen pumps.
9. The chemical pumps group is made up of the energy costs of the citric acid and hypo pumps.
10. The scum collection group is made up of the energy cost of the scum conveyor and the scum pump.

11. The WAS group is made up of the energy cost of the WAS pumps.
12. The chemical group is made up of the costs of the citric acid and hypo.
13. The ethanol dosing group is made up of the energy consumption cost of the dosing pumps and the ethanol cost of $1.27 a liter.

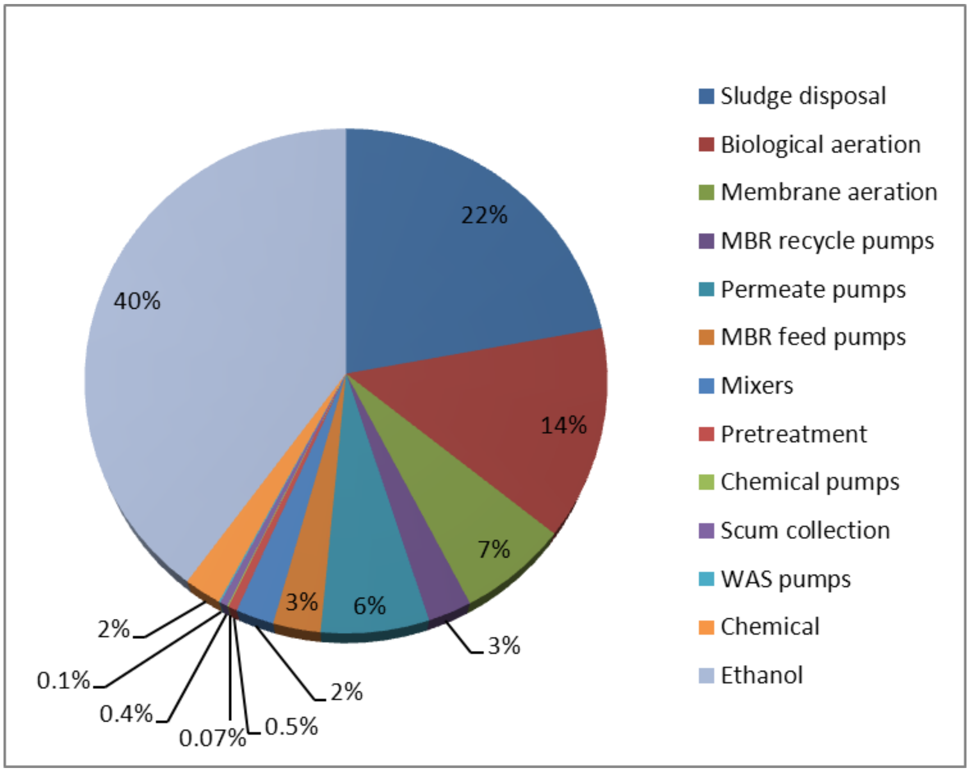


Figure 70: Breakdown of the MBR system operational costs

Figure 70 shows the top two cost contributors in the MBR system are ethanol costs at 40% and the disposal of the dewatered sludge at 22% of the total operational costs. The remaining cost contributors represent the energy consumer groups which make up 36% of the total operational costs of the MBR. Yoon at al. (2004) study showed a connection between aeration costs and production of sludge, the higher the aeration rate is, the lower the sludge volume produced. Yoon at al. (2004) also found that the cost of sludge disposal can be higher than the aeration cost.

The cleaning of the membrane includes membrane aeration, use of chemicals and the back wash by the permeate pumps. Assuming that the permeate pumps are reversed

5% of the operating time to backwash the membrane, the average energy demand percentage required to clean the membrane is 20 % of the total energy required to run the MBR. This percentage is 2.5 to 3.5 times lower than what was found by Cornel, Wagner and Krause (2003), Gil et al. (2010), Krzeminski, Jaap, ver der Greef and van Lier (2012) and Verrecht et al. (2010). In this study the membrane aeration cost was 18.61% of the total energy cost. Cornel, Wagner and Krause (2003) and Verrecht et al. (2010) found a membrane aeration rate as high as 70% of the total energy costs, and Gil et al. (2010) showed that membrane aeration was the highest energy contributor in the MBR system. The low membrane aeration percentage in this study could be related to the operation of the membrane at low flow rate (critical flow) (Verrecht et al., 2008).

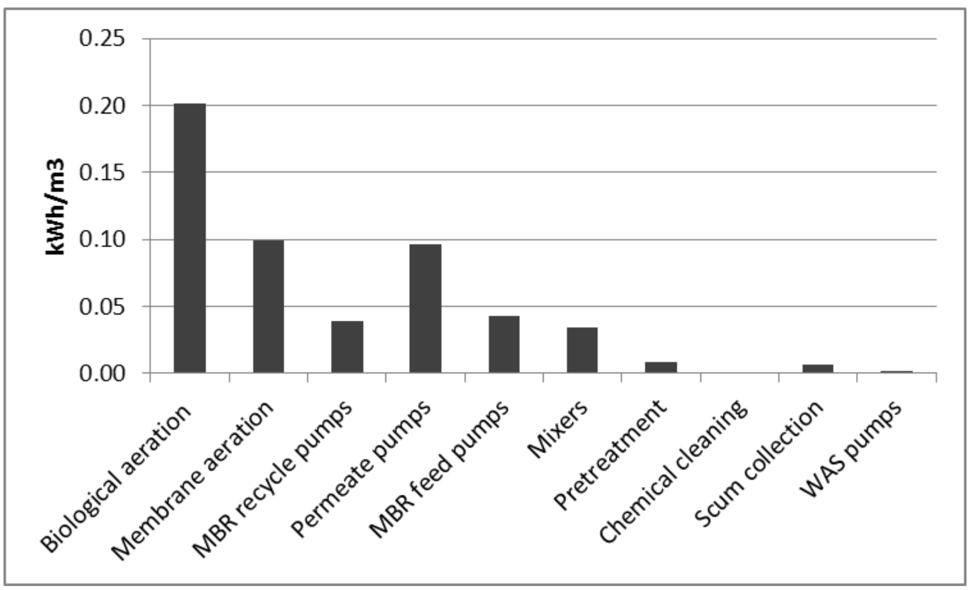


Figure 71: Breakdown of MBR energy consumption components

Figure 71 shows that out of the total energy consumption of 0.53 kWh/m^3, the membrane aeration consumes 0.1kWh/m^3 which is 18.61% and the biological aeration 0.2 kWh/m^3 which is 38.03% of the total energy. These findings are within the range of consumption found by Verrecht et al. (2008), which found biological aeration energy consumption of 0.12 to 0.2 kWh/m^3 and membrane aeration energy consumption of 0.1 to 0.21 2 kWh/m^3. The other energy consumers included the feed pumps with 8%, recycle pumps with 7.4%, mixer with 6.4%, pretreatment with 1.5%, scum collection 1.3%, WAS pumps 0.3% and chemical cleaning with 0.2% of the total energy costs.

The membrane aeration is linked to the flux value entering the membrane (Hirani et al., 2010). Figure 72 presents the membrane aeration energy demand proportional to the membrane flux. During a low flux, the energy consumption of the membrane aeration is higher and as the flux value gets closer to the design flux of 23 LMH, the energy consumption of the membrane aeration rate decreases.

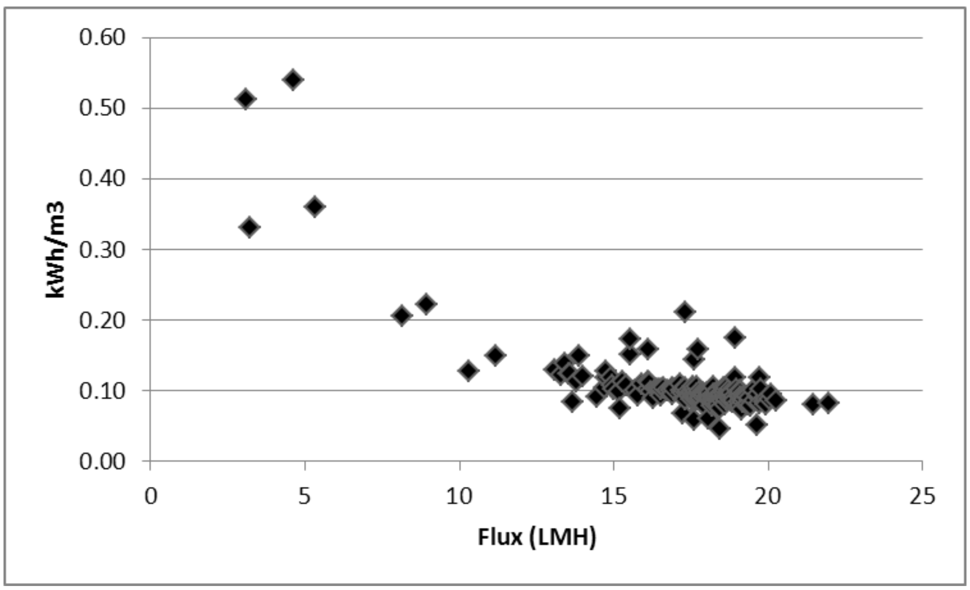


Figure 72: Membrane aeration energy demand proportional to membrane flux

The MBR average flux is 17.32 LMH, which is considered to be at the lower end of the range of 10 to 150 LMH (Judd, 2011). The average membrane aeration energy consumption at the flux rate of 17.32 LMH, was found to be 0.11kWh/m^3 (Figure 72). This energy consumption is within the range of 0.1 to 0.21 kWh/m^3 found by Verrecht et al. (2008), but lower than the rate of 0.5 to 0.7 kWh/m^3 found by Krzeminski et al. (2012). The low energy rate in this study could be related to the low flux under which the MBR is operated (Field et al., 1995; Hirani et al., 2010; S. Wang et al., 2005).

6.3.3 Operational cost comparison

In order to answer the second research question, an operational cost comparison between the MBR and the Bardenpho was carried out. Figure 73 presents the cost required to treat one cubic meter of wastewater through both of the systems. The results show that treating one cubic meter of wastewater through the MBR is 1.3 times more expensive than if it was treated using the Bardenpho system.

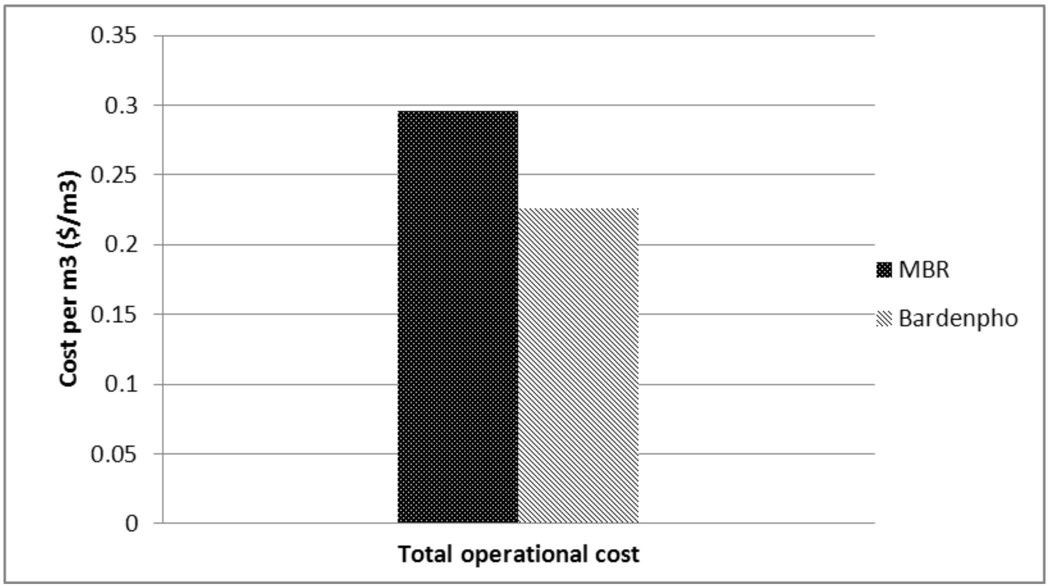


Figure 73: Bardenpho and MBR total operational cost comparison

The highest cost elements within the Bardenpho system were the sludge disposal, with 45% of the total operation cost and ethanol dosing with 33% of the total cost, with the remaining 22% being energy consumption. The MBR cost distribution shows that the highest costs are related to the ethanol dosing, with 40% of the total cost and sludge disposal with 22% of the total cost, with the remaining 38% power usage.

6.3.3.1 Sludge volumes

The sludge disposal cost was found to have the highest cost component within the Bardenpho system and second highest within the MBR total operational cost. The comparison of the sludge disposal cost between the Bardenpo and the MBR system show that the Bardenpho sludge disposal cost is 1.6 times higher than the disposal cost of the MBR system. This cost difference is related to the small volumes of sludge that the MBR produces in comparison to the activated sludge system (Gonzalez et al., 2007; Judd, 2011; Leiknes, 2010). The average wasting volume of the Bardenpho system is 925 m^3 per a day which is 5.36% of the total daily average volume of influent entering the Bardenpho system. The average wasting volumes of the MBR system is 180 m^3 per day which is 3.3% of the total daily average volume of influent entering the MBR system.

6.3.3.2 Ethanol

The use of ethanol, as an external source of carbon, is to make sure there is sufficient carbon source available in the treatment system for the bacteria to remove contaminants. The purchase cost of ethanol in this study was based on the actual cost of $1.27 a liter. The results show that the cost of ethanol is the highest operational component cost within the MBR system and second highest within the Bardenpho operational costs. The ethanol volumes used for the supplement of the biological treatment of both the Bardenpho and MBR system was based on many years of trialing the performance of the Bardenpho to remove nitrogen. Both the plants were set to have a similar fixed ethanol dose per a cubic meter of wastewater entering the systems. The ethanol dosing rate is set according to the design flows of the Bardenpho and the MBR, 16,700m^3/d and 7,300m^3/d respectively. The average flow monitored during the investigation period was lower than the design flow in the MBR and higher in the Bardenpho, therefore the costs reflected in the results show higher MBR ethanol cost per cubic meter of influent.

Figure 74 show the cost comparison of the common components in the MBR and the Bardenpho systems.

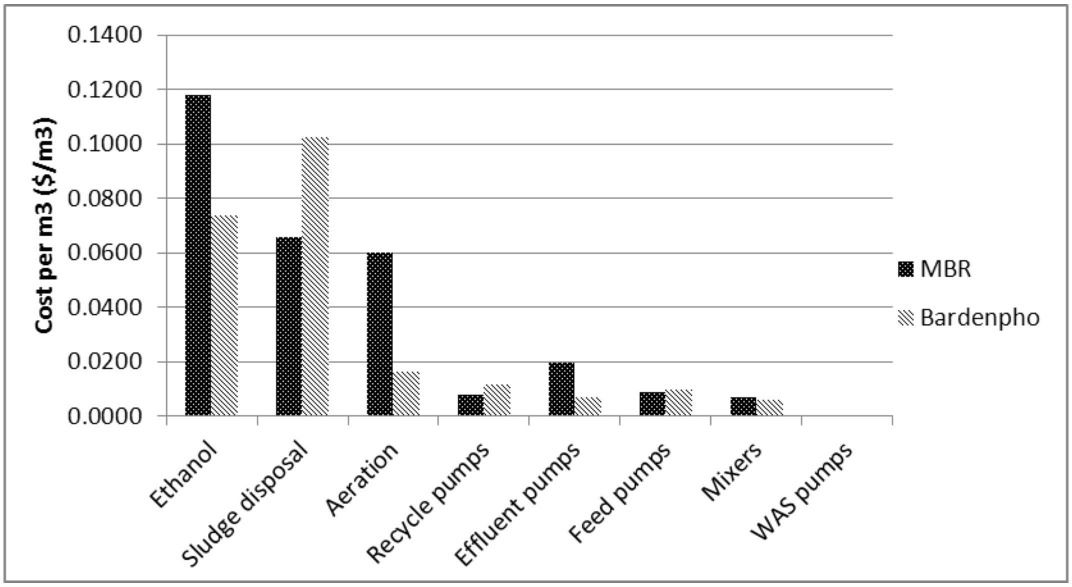


Figure 74: Bardenpho and MBR operational component cost comparison

Figure 74 shows that ethanol, aeration and final effluent pumps are 1.6, 3.6 and 2.8 times higher in cost respectively in the MBR than the Bardenpho per cubic meter of wastewater. However, the Bardenpho sludge disposal costs, recycling pumps and WAS pumps are 1.6, 1.4 and 1.6 times higher, respectively, than the MBR per cubic meter of wastewater. The feed pump and mixers show similar costs.

6.3.3.3 Energy consumption comparison

The power consumption cost in the Bardenpho system was 22% of the total cost of the plant, while in the MBR system the power cost portion was 38% of the total operation cost of the system. To show the differences in energy consumption between the Bardenpho and MBR processes, Figure 75 presents the energy consumption comparison of the common components in the Bardenpho and MBR systems.

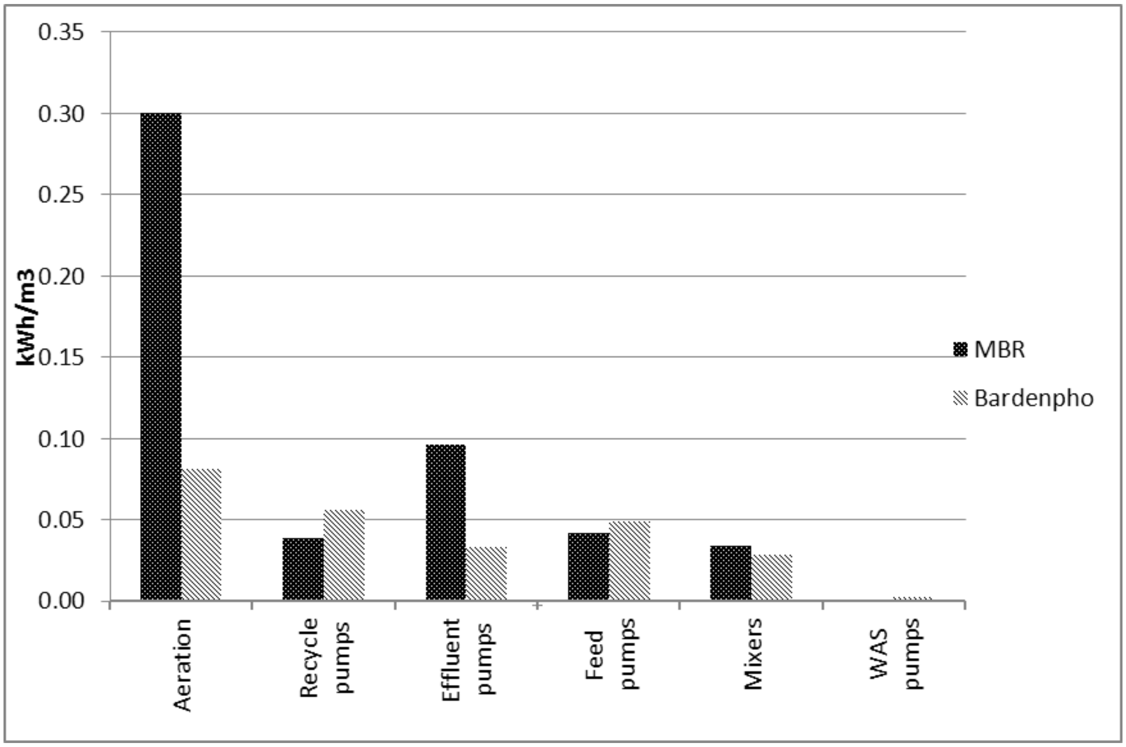


Figure 75: Bardenpho and MBR component energy comparison

Figure 75 shows that aeration, effluent pumps and mixers are 3.7, 2.9 and 1.2 times higher in power consumption, respectively in the MBR than the Bardenpho per cubic meter of wastewater. However, in the Bardenpho the recycling pumps, the feed pumps and the WAS pumps are 1.4, 1.2 and 1.7 times higher, respectively, than the MBR per cubic meter of wastewater.

The difference in aeration between the MBR and the Bardenpho systems is because of the additional aeration system existing in the MBR to clean the membrane in addition to the biological aeration. The MBR effluent pumps are used to clean the membrane by reversal of the flow as well as producing the MBR permeate, therefore resulting in a higher energy consumption for the MBR effluent pumps than the Bardenpho effluent pumps. The higher energy rate of the Bardenpho recirculation and feed pump is due to the different streams entering the Bardenpho system that do not exist in the MBR, such as recycled DAF liquor and overflow pumps. WAS pumps in the Bardenpho system had a higher energy rate than the MBR due to the higher volumes of sludge produced by the Bardenpho (Judd, 2011).

Figure 76 presents the overall energy consumption comparison between the MBR and Bardenpho. The energy consumption required to treat one cubic meter of wastewater in the MBR is 2.1 times higher than treating one cubic meter of wastewater through the Bardenpho system.

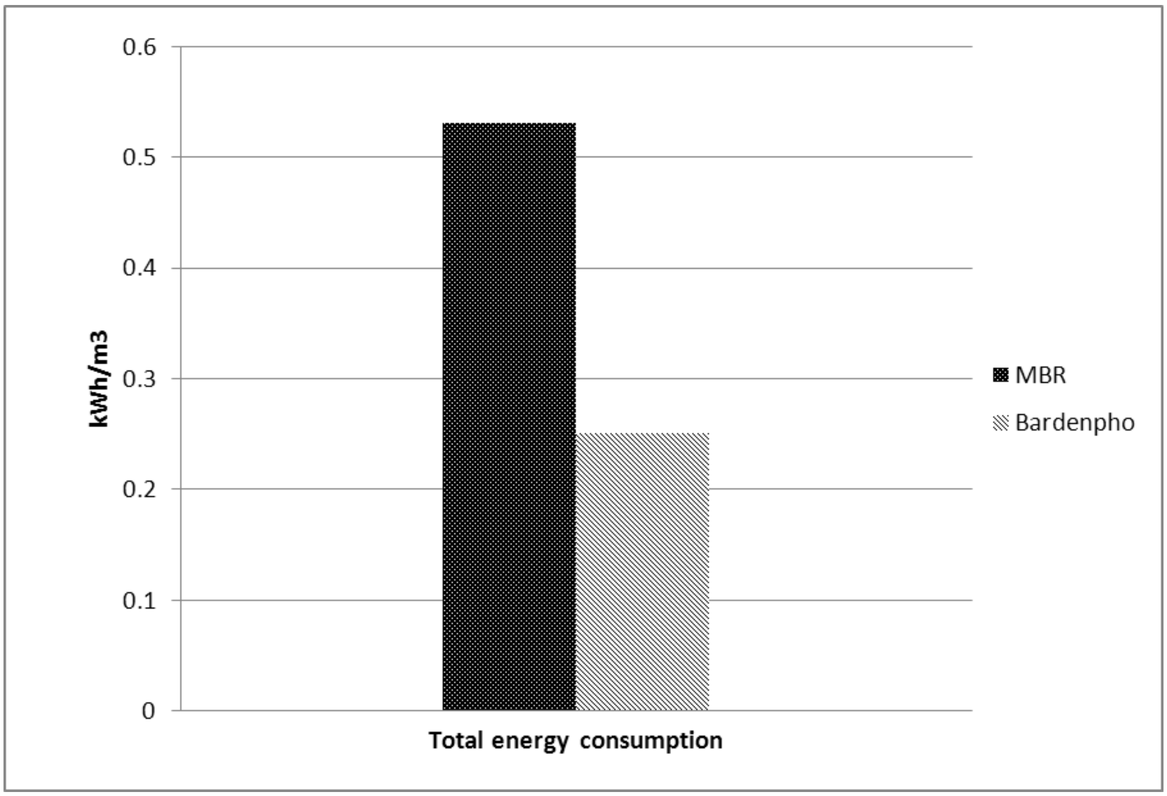


Figure 76: Bardenpho and MBR energy comparison

6.3.4 Net present value comparison

The net present value was calculated, taking into account the capital expenditures (capex) and operational expenditures (opex) during the plant's lifetime (equation 6).

$$NPV\,(i,N) = \sum_{t=o}^{N} \frac{(Capex)_t + (Opex)_t}{(1+i)^t} \tag{6}$$

In the NPV calculation, no cash inflow was used because the wastewater treatment plant is operated and owned by the council, therefore acts as a nonprofit service. Plant life time (N) of 30 years was used with a discount rate of 7 % (Table 29).

Table 29: Capex, opex and resulting NPV for the Bardenpho and MBR processes

	Unit	Bardenpho	MBR
Capacity	**m^3/d**	**7,300**	**7,300**
Total capex	**$**	**8,535,579[1]**	**8,519,978[1]**
Total opex	**$**	**490,785**	**744,880**
Energy, sludge disposal and chemicals	$	446,985[2]	581080
Operator cost	$	43,800	43,800
Membrane replacement	$	-	120,000
NPV[3]	**$**	**14,625,746**	**17,763,225**

[1] Cost of land is not included

[2] MBR costs are 1.3 times more than the Bardenpho costs (section 6.3.3)

[3]Time period 30 years and discount rate of 7 %.

The use of the MBR actual capital cost of $8,519,978 was used, as detailed in section 4.5. The annual operational cost of $581,080 (section 6.3.2) together with an estimated operator time of 1460 hours a year, with an hourly rate of $30 and a membrane life of 10 years was used to calculate the total opex.

The capital cost of the Bardenpho, as mentioned in section 3.5, represents the cost of the Bardenpho designed to treat an annual average flow of 16,700 m^3/d. To be able to compare between the MBR and Bardenpho systems, it is necessary to ensure both systems have a similar flow capacity. This was done using the power law equation (Bouman et al., 2005). The cost of the Bardenpho with a capacity of 7,300 m^3/d, was calculated by taking the cost of the 16,700 m^3/d Bardenpho estimated as $14,819,336 and dividing it by the Bardenpho and the MBR design flow ratio (16,700/7,300) and a scaling exponent of 2/3. The estimated cost capital cost for the Bardenpho plant, designed for an average flow of 7,300 m^3/d, was calculated to be $8,535,579. This capital cost and the annual operational cost of $1,422,770 (section 6.3.1) together with an estimated operator time of 1460 hours a year, with an hourly rate of $30, was used to calculate the Bardenpho NPV.

Table 29 shows that over a period of 30 years, the Bardenpho is $2,578,166 less expensive than the MBR. Figure 77 presents the NPV comparison over a time period of 100 years.

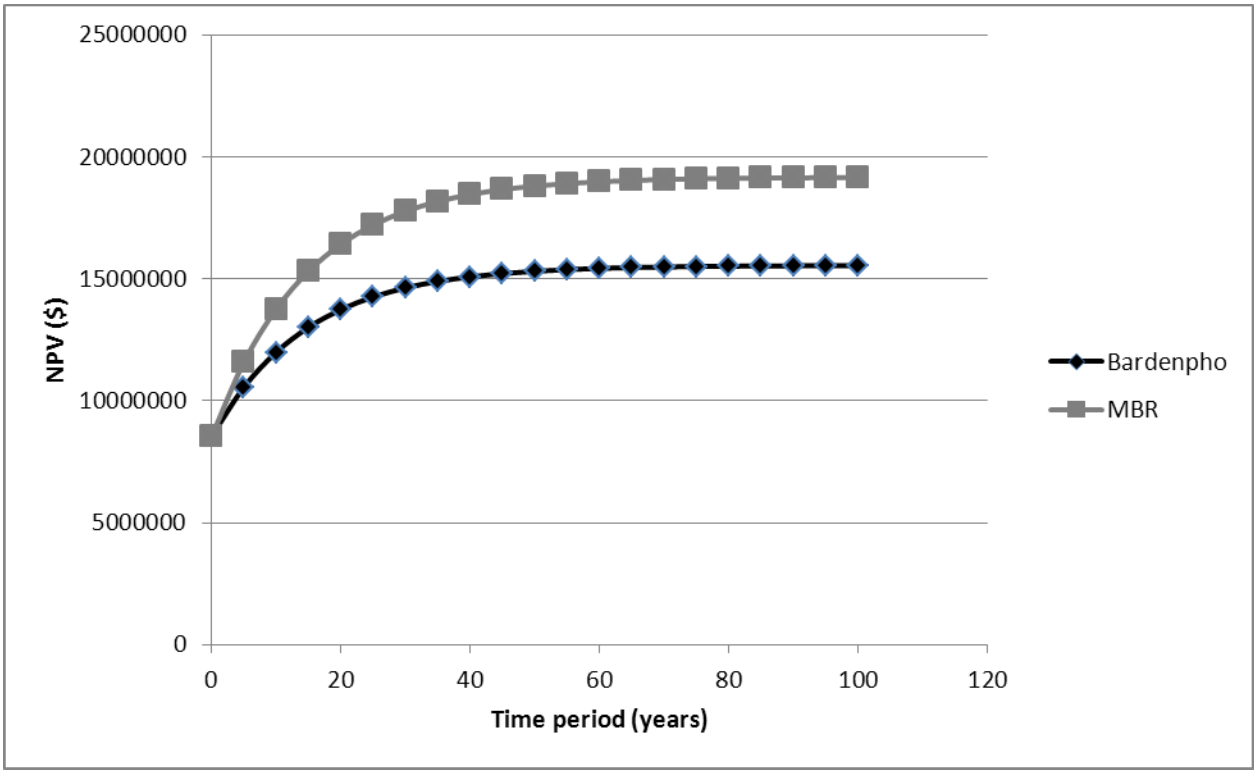


Figure 77: Bardenpho and MBR NPV comparison over a time period of 100 years

The capital cost of the Bardenpho and MBR plants, with the same flow capacity, is similar; however, the operational costs of the MBR are higher than the Bardenpho therefore the NPV cost difference increases with time as shown in Figure 77.

6.3.5 Cost per land area and footprint size comparison

As detailed in section 6.3.4, the capital costs of the Bardenpho and the MBR processes with the same flow rate of 7,300 m^3/d, are \$8,535,579 and \$8,519,978 respectively. These capital costs, together with the area occupied by the Bardenpho and the MBR, were used to evaluate the cost per square meter. To estimate the footprint of a Bardenpho system with a capacity of 7,300 m^3/d, a linear connection was used by dividing the MBR flow of 7,300 m^3/d with the Bardenpho flow of 16,700 m^3/d, multiplied by the actual 16,700 m^3/d Bardenpho footprint (Table 30).

Table 30: Bardenpho and MBR Capital cost and footprint area

Plant	Design flow[1] (m^3/d)	Capital cost[2] ($)	Footprint area (m^2)
Bardenpho	16,700	14,819,336	3668[3]
Bardenpho	7,300	8,535,579	1603[3]
MBR	7,300	8,519,978	732

[1]Average daily flow
[2]Does not include cost of land
[3]Area of clarifiers included

Table 30 shows that the area occupied by the Bardenpho system is 2.2 times larger than required for the MBR system. In the MBR the membrane replaces the second clarifier in the Bardenpho system, therefore the MBR does not require such a large area to settle out the solids (Environmental Protection Agency, 2005; Judd, 2011; Leiknes, 2010).

Table 31 presents the cost of the plants per square meter of land, calculated using the capital costs of the plants divided by the area each plant occupied.

Table 31: Bardenpho and MBR cost per square meter

Plant (flow m^3/d)	Cost per land area ($/$m^2$)
Bardenpho (7,300)	5,325
MBR (7,300)	11,639

Table 31 shows that the price per square meter of land of the MBR plant with a capacity of 7,300 m3/d, is $11,639, while the cost of the Bardenpho plant with a capacity of 7,300 m3/d, is $5,325 per square meter.

Chapter 7: Conclusion and Recommendations

7.1 Conclusion

The present study set out to compare the performance and costs of two wastewater treatment methods, a Membrane bioreactor (MBR) system and a Bardenpho activated sludge process. Rotorua District Council installed an MBR system, using ZeeWeed 500d hollow fibre 0.04 micron membranes with a total capacity of 7,300 m^3/d. The MBR was designed to operate alongside the existing Bardenpho process to increase the overall performance and capacity of the wastewater treatment plant. A portion of the preliminary and primary treated wastewater entering the entire treatment plant is diverted onto the MBR system, while the rest of the flow continues toward the Bardenpho system.

The results show the MBR system produces a higher quality effluent than the Bardenpho and the removal efficiencies remain at a high level regardless of a change of inflow. The Bardenpho system, despite showing lower effluent qualities in comparison to the MBR, was found to be very efficient in removal of nitrogen and biodegradable organics. The MBR treatment, although showing high removal rates of the constituents tested, comes with a cost. The operational cost comparison showed that sludge removal and ethanol dosing played a big part in the running cost of each plant, resulting in similar operational costs for the MBR and Bardenpho. However, when the cost of the ethanol and sludge disposal were not considered, and the energy consumption was compared, the cost differential between the MBR and Bardenpho increased, with the MBR energy costs being over double the Bardenpho energy costs. The capital cost comparison was found to be a similar value when the cost of land was not a factor; however, the Bardenpho footprint area was significantly larger than the MBR.

In conclusion, from a cost point of view, the Bardenpho system is more economical; however, the low cost of running the Bardenpho comes with lower removal efficiencies in comparison to the MBR.

7.2 Limitations

This study relied on large amounts of manual data collection over a long period of time, so the accuracy and consistency of this collection was a factor that was difficult to control for.

The power rating used to estimate the energy consumption of each component in the systems was taken from performance curves and not actual power measurements. Therefore more accurate power consumption could be determined if some method of measuring and recording power use was installed.

The Bardenpho NPV and cost per area calculations were based on an estimated capital cost using exponential factors and not actual costs. To be able to have a more accurate estimate of the capital costs of the Bardenpho at the same capacity as the MBR, the size of equipment and tanks would need to be determined and prices would then have had to be found through suppliers and contractors. The capital cost given in this study was intended to give a general idea in order to make a comparison between the two systems.

7.2 Recommendations

From this study it was shown that ethanol dosing has a big impact on the operational costs of the plants. It is recommended that with the optimization of ethanol dosing, there could a reduction in operational costs and improvement in effluent biodegradable organics concentration.

The flow diversion between the Bardenpho and MBR affects the overall performance and cost of the entire wastewater treatment plant. To be able to conclude how the choice of flow is affecting the performance and the cost an investigation should take place to find the optimal flow entering into the MBR and the Bardenpho systems.

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Appendices (attached in CD)

Appendix 1: Bardenpho capital cost breakdown using the power law equation

Appendix 2: Bardenpho capital cost breakdown using the Bardenpho asset value

Appendix 3: MBR capital cost item breakdown

Appendix 4: Plant's component working hours worksheet