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Department of Environmental and Geomatic Engineering

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Master of Science in Environmental and Geomatic Engineering

Design, Cost & Benefit Analysis of a Membrane Bioreactor

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Chapter 1

Introduction

Membrane bioreactor (MBR) technology, which is a combination of biological-activated sludge process and membrane filtration is widely spread, and accepted in recent years for the treatment of many types of wastewaters, while the conventional activated sludge (CAS) process cannot cope with poor sludge settleability and this does not allow to keep high MLSS concentration in the biological reactors. Complete solids removal, a significant disinfection capability, high rate and high efficiency organic removal and small footprint are common characteristics regardless the wastewater type to be treated or the commercial process used (Stephenson et al., 2001) MBR technology is also used in cases where demand on the quality of effluent exceeds the capability of CAS. Although MBR capital and operational costs exceed the costs of conventional processes, due to more stringent regulations, and lack of space prevent to add new treatment units the upgrade from CAS to MBR can be necessary when it comes improve treatment performances. Along with better understanding of emerging contaminants in wastewater, their biodegradability, and with their inclusion in new regulations, MBR may become a necessary upgrade of existing technology in order to fulfill the legal requirements in wastewater treatment plants (WWTPs).

However the development of MBRs has been limited by problems of membrane fouling during filtration of activated sludge. Membrane fouling is the most serious problem affecting system performance. Fouling phenomena on the membrane surface and within the pores reduce long term flux stability necessitating membrane cleaning which then add to the overall cost, as does membrane replacement in case where cleaning fails to produce adequate flux recovery (Gander et al., 2000).

Various techniques are used to reduce fouling. Basically fouling of membranes in MBR systems can be minimized by a reduction of flux, promotion of turbulence to limit the thickness of the boundary layer and/or periodical application of cleaning measures to remove the cake layer and foulants. Membrane cleaning commonly comprises intermittent physical cleaning (usually backwash) and periodic chemical cleaning. (Chang et al., 2002, Stephenson et al., 2001)

The idea for coupling the activated sludge process and membrane separation was firstly reported by research conducted at Rensselaer Polytechnic Institute, Troy, New York, and Dorr-Oliver, Inc. Milford, Connecticut, US (Jyoti et al., 2013). The first MBR installation (Membrane Sewage System-MST) was made by Dorr-Oliver, Inc., with flat sheet ultrafiltration plate and frame membrane. It did not gain much interest in North America but it had considerable success in Japan in the 1970s and 1980s. Before the 1990s, most of the installed MBRs were used for industrial water treatment. With the development of submerged membranes, firstly introduced by Yamamoto et al., the number of MBRs treating municipal wastewater increased while the MBR market is currently experiencing accelerated growth. The global MBR market doubled over the last 5-year period and in 2005 it has reached a market value of $217 million in 2005 with a projection for the year 2010 of $360 million (Judd, 2006).The MBR process can be configured in many different ways depending on project-specific nutrient removal objectives. Anoxic zones before or after the aerobic treatment may be used for denitrification, depending on the effluent nitrate and total nitrogen requirements. Anaerobic zones may be used to achieve enhanced biological phosphorus removal in any of its possible configurations.

Two basic MBR configurations are shown in Fig. 1 (modified from (Tchobanoglous et al., 2004). The first is a recirculated configuration with an external membrane unit (Fig. 1a). Mixed liquor is circulated outside of the reactor to the membrane module, where pressure drives the separation of water from the sludge. The concentrated sludge is then recycled back into the reactor. The second is a submerged configuration with the membrane module immersed in the activated sludge (Fig. 1b). A suction force is applied to draw the water through the membrane, while the sludge is retained on the membrane surface. A manifold at the base of the reactor diffuses compressed air within the reactor, providing oxygen to maintain aerobic conditions. The air bubbles also function to scour the membrane surface and clean the exterior of the membrane as they rise in the reactor. The submerged configuration is more commonly used than the recirculated configuration because it is less energy-intensive and provides a cleaning mechanism to reduce membrane fouling. Thus, more fouling models focus on the submerged configuration than on the external configuration.

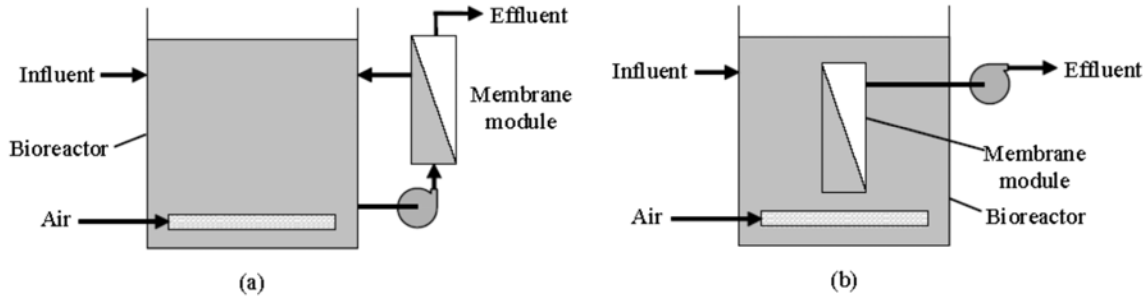


Figure 1. Schematic diagrams of basic membrane bioreactor configurations Yamamoto et al. 1989: (a) MBR with external membrane module and (b) MBR with immersed membrane module

Models that can accurately describe the MBR process are valuable for the design, prediction, and control of MBR systems. Complex models that are also practical for real applications can greatly assist in capitalizing on the benefits of MBR technology. The model which is used here is ASM1 model.

The application of ASMs are presumably meant for ASP operation in the ranges of conventional ASP operating parameters, for example, SRT range 3–15 days, HRT range of 3–5 h and MLSS range 1.5–4 gl^{-1} for completely mixed systems (Benedetti et al. 2008).

A recent study on design and operating experience with municipal MBRs in Europe has reported the ranges of various parameters (The HRT of 13 MBR plants have been reported to be in the range of 2.8–8.1 h, with most of the plant operating at HRT between 4–6 h. The MLSS of 11 MBR plants have been reported to be in the range of 7–13.5 g l−1, with most of the plant operating at MLSS higher than 10 $g.l^{-1}$. Further, the SRT values of seven plants have been reported to range between 15 and 40 days. Efforts have been made over the past 15 years toward appropriate application of ASMs for MBRs. While early trials, Chaize et al. 1991, used the very basic form of ASM1, using default parameter values, performing no systematic calibration or influent characterization, recent efforts have presented various aspects of systematic calibration of key and sensitive parameters along with emphasis on the influence of influent wastewater characterization in terms of various ASM-based fractions (Delrue et al., 2008, Spérandio et al. 2008). The early study of Chaize and Huyard, based on a laboratory-scale MBR fed with domestic wastewater, aimed to model effluent COD, TKN, and sludge production at two HRT values (namely, 8 and 2 h) and very high SRT (nearly 100 days). The MBR system was modeled with ASM1 using default values of parameters (Henze et al., 1987). The predicted effluent COD was reported to be slightly lower than that observed, and the predicted TKN was found quite close to the observed value. However, the major disagreement was reported on solids concentration. The model predicted a lower solids concentration than observed, and the solids concentration prediction was relatively better at higher HRT. The probable reason was thought to be the very high SRT (i.e., 100 days). These outcomes illustrate that a non-calibrated ASM1 is able to give a reasonable estimate of effluent COD and TKN, but is insufficient for very low HRT and very high SRT systems. Hence, this imposes care in the application of those models and in the investigation of appropriate parameter sets valid for these systems under variable operational conditions. This sets the scene for investigating the whereabouts of the encountered limitations.

The application of ASM1 moved toward better understanding of model parameters and, hence, a more systematic calibration, taking into account the nature of the MBR biology and specific operating conditions. The ASM1 application on side stream MBR by Jiang et al. stressed the importance of various sensitive biokinetic parameters and influent wastewater characterization. More recently, Delrue et al. commented that despite some difficulties, ASM1 is suitable for modeling MBR plants if influent characterization and systematic calibration of aeration can be considered.

The incorporation of storage phenomena (Gujer et al., 1999, Krishna et al., 1999) is a unique feature of ASM3 and might play a role in the case of MBRs on account of possibilities of low organic load conditions (Wintgens et al., 2003). Nevertheless, ASM1 has been shown to be sufficient where conditions are not favorable to storage phenomena (Delrue et al., 2008).In the aim of modeling MBRs over a large range of SRTs, Sperandio and Espinosa used ASM1 and ASM3 and commented that ASM models could provide satisfactory prediction of aerobic biological processes in submerged MBRs, although these could be improved for high SRT conditions. Studies so far are not conclusive as to whether ASM1 or ASM3 is better for MBRs

In a work done by Aileen et al. 2006 ASM1 was tested extensively against experimental and operational data for activated sludge systems. Main problems found with ASM1 were addressed in the development of ASM3. ASM2, ASM2d, and ASM3 were also validated against experimental data for conventional activated sludge systems, although less extensively. It has been suggested that ASMs may be suitable for characterizing biomass kinetics in an MBR system.

However, few studies have demonstrated the validity (or invalidity) of ASMs for modeling MBR systems.

Studies so far are not conclusive as to whether ASM1 or ASM3 is better for MBRs. It appears that the application of ASMs, in their original forms, often needs careful calibration of parameters, especially for sludge production and nitrification modeling. The issue of the significance of high SRT, which was a matter of further attention even in early MBR modeling studies (Chaize et al., 2008), remains a relevant point. It has been reasoned in recent research (Massé et al., 2004, Spérandio et al., 2008) that high SRT operation of MBRs is linked with corresponding influence on MBR specific sludge production and autotrophic biology. Throughout, it can generally be observed that all the recent efforts aiming at an accurate biological modeling of MBRs focus on MBR specificities (e.g., high SRT operation, membrane-retained microbial metabolites, etc.) and the corresponding parameter adjustment and modifications required in ASMs.

Chapter 2

Case study: Comodepur WwTP

Introduction

Lake Como in northern Italy is world renowned as a prime holiday location as well as being home to many celebrities. Considering Como as a touristic place puts a strong focus on maintaining the quality of the local environment and means that smart wastewater management is a particularly crucial problem in the region. An added challenge is the dramatic mountainous terrain; not the most ideal landscape for housing wastewater treatment stations. The rough nature and mountainous shape of the landscape and locating in hillside leads that the main sewage treatment plant is located just a few hundred meters from the historic city center.

A complex pump station management system oversees a total of 14 pumping stations which push 55,000 cubic meters of effluent per day to the treatment plant. Comodepur, the organization responsible for managing the wastewater system in the region, must ensure that the treated, purified water which is discharged directly back into Lake Como complies with stringent environmental standards.

Plant Description

Influent Characteristics

This plant has a treatment capacity of 55000 cubic meters per day and is designed to produce a high quality treatment plant effluent. First, the wastewater pumped to the preliminary treatment section.

The composition of the influent is shown in the following table:

Table 1. Composition of treated wastewater.

Type of urban wastewater	Discharge days / year	Flow 2002/03/04		Flow 2005/06/07		Flow 2008	
		mc/year	%	mc/year	%	mc/year	%
tributary total	365	17.443.100	100	15.133.294	100,00	18.996.460	100,00
bypass after primary	- - -	710.960	4	430.543	2,85	1.903.560	10,02
Total treated water	365	16.732.140	96	14.702.717	97,15	17.092.900	89,98
domestic wastewater	365	9.296.400	53	9.430.642	62,32	9.607.500	50,58
industrial wastewater	230	2.615.705	15	2.464.117	16,28	2.296.340	12,09
Rainwater	365	5.530.995	32	3.297.836	21.79	7.093.169	37.34

The plant treated also liquid wastes from the cleaning of sewers and drains from septic tanks in the amounts indicated in the following table.

Table 2. Composition of liquid wastes treated.

Waste from outside	2002 to 2004		Inflow 2005 to 2007		Inflow 2008	
	kg/year	%	kg/year	%	kg/year	%
Septic tank sludge	1.198.600	75	649.407	57,41	501.920	51
Cleaning wastes sewer	530.180	25	481.660	42,59	479.520	49

Data used for design contains years 2008, 2009 and 2010 of Comodepur Wastewater Treatment Plant. Since we have to design an MBR plant instead of current ASP, data related to biological section have been extracted. Following figures are the influent and effluent to the biological section of Comodepur WWTP 2008 data. Data of year 2009 and 2010 is available on APPENDIX 1.

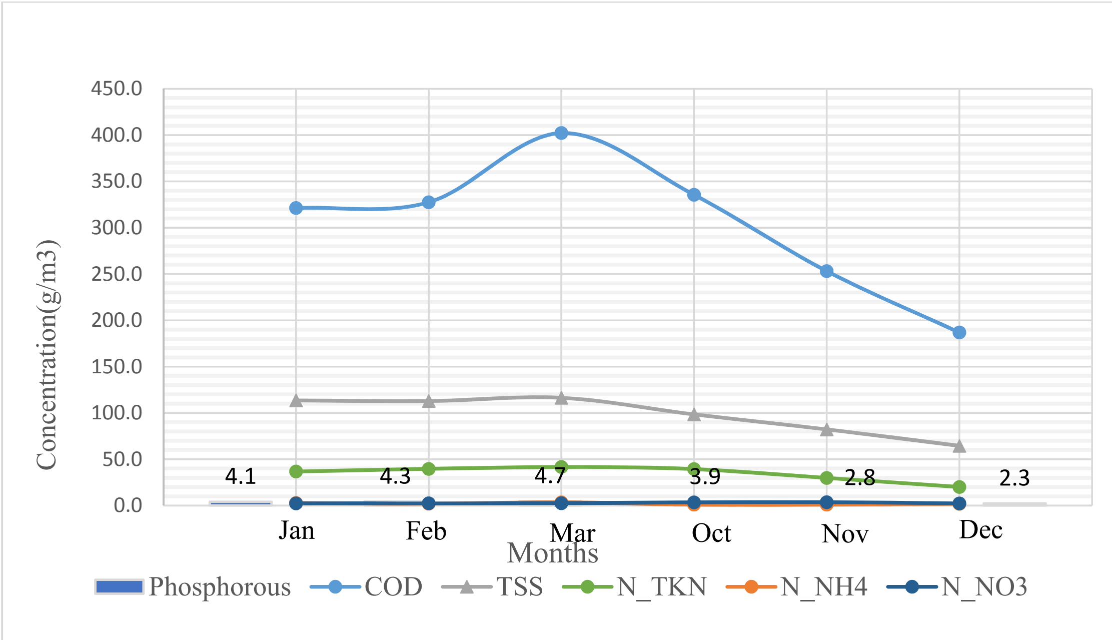


Figure 2. Six cold Monthly average influent Conc. of biological section, Comodepur WWTP Data 2008

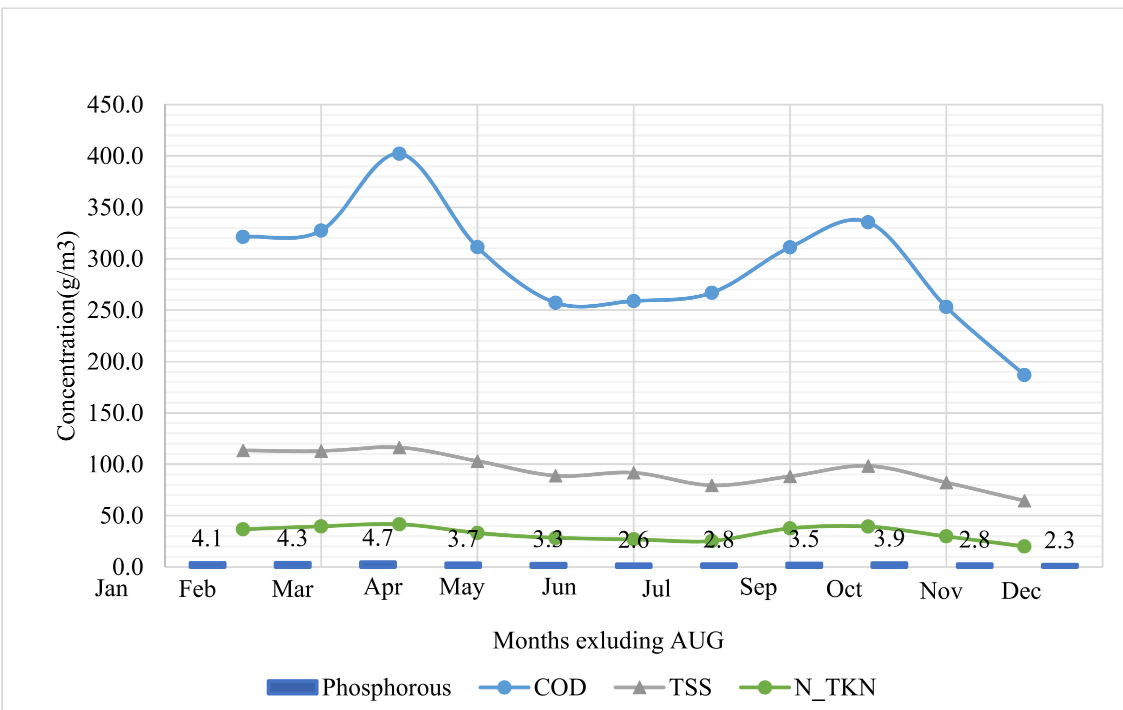


Figure 3. Monthly average influent Conc. of biological section, Comodepur WWTP Data 2008

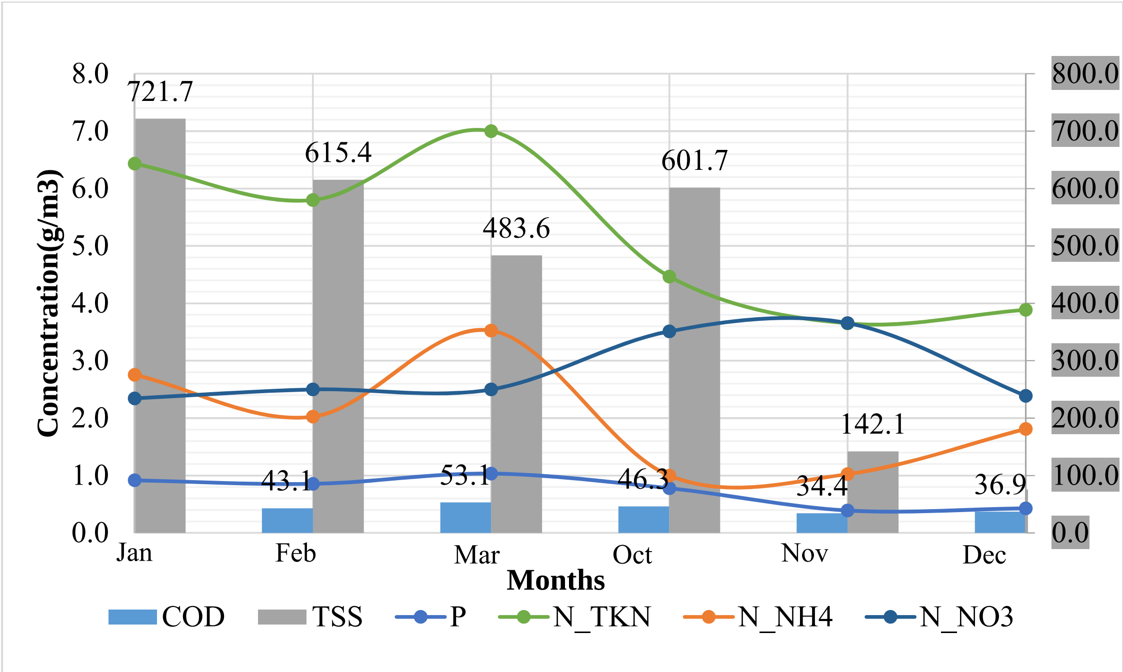


Figure 4. Six cold monthly average effluent Conc. of biological section, Comodepur WWTP Data, 2008

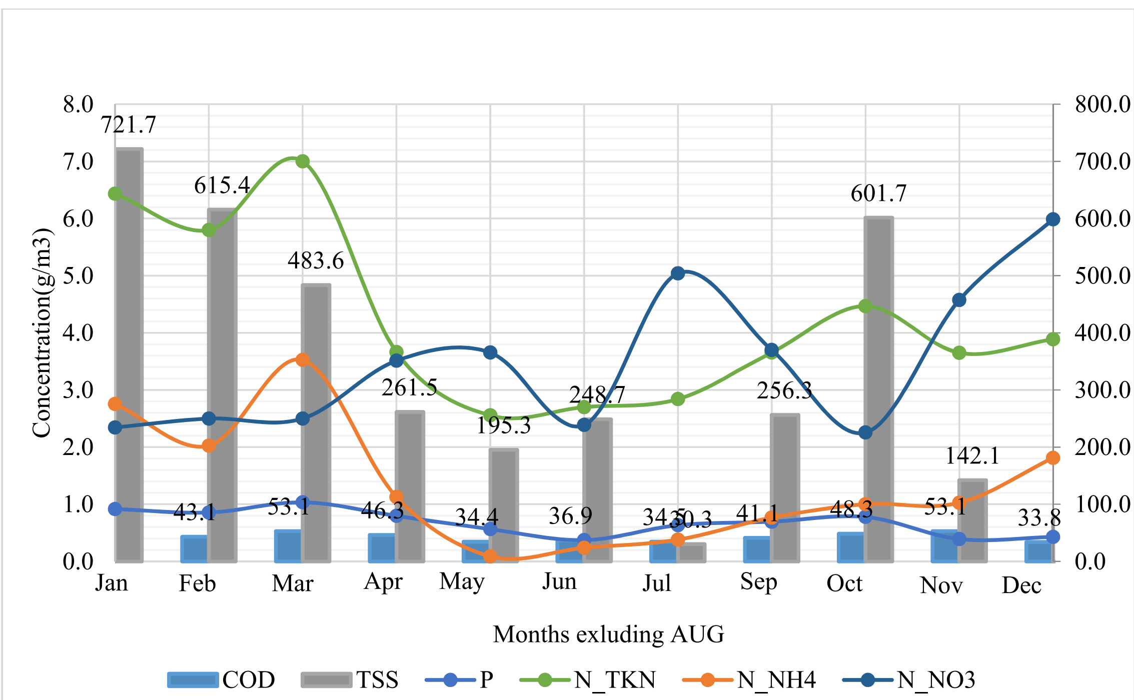


Figure 5. Monthly average effluent Conc. of biological section, Comodepur WWTP Data, 2008

The wastewater coming from the sewer main is sent to screens and, after pumping, is degritted. Then the sewage goes to the primary treatment. Contrary to the original design, primary flocculation with reactive flocculant (salts of iron or aluminum) are no longer added.

Primary effluent enters the biological section which is made of a pre-denitrification section, a nitrification section and a post-denitrification section. Recirculation of sludge contain denitrifying bacteria (heterotrophic), and of mixed liquor from the process of oxidation-nitrification containing nitrates get mixed with the incoming primary effluent that provides the carbon source.

Aeration for carbon and ammonia oxidation is performed by compressing air that is supplied to the oxidation tank through a fine bubble diffusion system.

Subsequently, the water runs into post-denitrification tanks. An external carbon source made of a hydro-alcoholic mixture is added to the mixed liquor to remove residual nitrates. Finally, after a post-aeration phase, the mixed liquor is sent to the final settling tanks, where the sludge is separated from the secondary effluent; sludge deposited on the bottom of the settling tank is pumped back to the pre-denitrification, and excess sludge is sent to the sludge treatment line.

After the biological treatment, the secondary effluent undergoes tertiary physico-chemical treatments (clari-flocculation) aimed at removing phosphorus, colloidal substances, and residual suspended solids still present. Polyelectrolyte flocculant is added to improve tertiary settling that is performed through lamellar packs. The tertiary sludge that settles on the bottom is sent to sludge treatment.

From clari-flocculation, tertiary waters are sent partly to cloth filtration and partly to the gravity filtration on a sand-bed for the removal of residual suspended solids.

After primary treatment, diurnal peak flow-rates are diverted to a biofiltration plant, made of three units in series: organic carbon removal, nitrification and post-denitrification. Effluent form the biofiltration unit are sent to the tertiary treatment.

In submerged biofilters, the microorganisms do not grow suspended in the water column, but form a bacterial film adhering to an inert surface. Periodic backwashing removes excess growth, and backwashing wastewater is sent to the conventional activated sludge process.

Finally, effluent water is disinfected by UV irradiation nan open channel and discharged to stream Cosia that ends up in Lake Como after few hundred meters.

Excess sludge from biological and tertiary treatment is sent to sludge thickening units, while the drained water is recirculated back to the head of the biological process.

The thickened sludge reduced further in volume by mechanical dewatering centrifuges. The dewatered sludge, containing about 20% of dry matter store in silos by means of conveyors or progressive cavity pumps, and the waters extracted from the sludge is recirculated to the pre-denitrification.

Downstream of the pre-treatment the wastewater is sent to the biological treatment phase (pre-denitrification).

The main treatments are shown in the following table.

Table 3. Main treatment

Preliminary treatment	Primary treatment	Biological Treatments	Treatments Tertiary	Sludge Treatment
Fine screening	tanks coagulation/ flocculation	Pre denitrification	tanks coagulation/ flocculation	Thickening
uplift		Oxidation Nitrification	Sedimentation lamellar pack	
Grit removal	Sedimentators lamellar packs	Post denitrification	Sand filtration	dehydration
		Sedimentation	Filtration fabric	
		Biofiltration	UV disinfection	

The following table shows the operating parameters of the system with reference to the flow rates and the analytical parameters of the influent and effluent.

Table 4. Operating Parameters.

Flow rates		**Unit**	**Design value**	**Average 2002 – 2004**	**Average 2005 – 2007**	**2008**
Average daily flow input	Qg	m^3/d	55.000	48.558	41.443	51.975
Average daily flow treated	Qg	m^3/d	- - -	46.303	40.210	46.787
Average flow rate (within 4 hours) input	Qm	m^3/h	2300	2.023	1.727	2.166
Average flow rate (about 24 hours) treated	Qm	m^3/h	- - -	1.929	1.675	1.949
Peak flow Qp project	Qp	m^3/h	3400	- - -	- - -	- - -

Table 5. Operating Parameters: analytical data
.

Analytical data	**Unit**	**2002/03/04**		**2005/2006/2007**		**2008**	
		Input	**Output**	**Input**	**Output**	**Input**	**Output**
COD	mg/l	371	53,67	435	46	419	39
SST	mg/l	122	15,33	155	11	160	7
Phosphorus	mg/l	3,91	0,59	4,74	0,72	4,70	0,64
Surfactants tot.	mg/l	19,75	1,65	22,4	1,57	20,4	1,20
$N\text{-}NO_2$	mg/l	0,16	0,13	0,18	0,16	0,19	0,05
$N\text{-}NO_3$	mg/l	0,78	5,08	0,27	4,88	0,36	4,95
$N\text{-}NH_4$	mg/l	27,32	1,31	25,91	1,45	25,9	0,79
N org	mg/l	13,43	3,76	14,90	3,39	22,84	2,55
TKN	mg/l	38,72	5,07	40,79	4,84	37,18	3,34
N total	mg/l	39,67	10,28	41,23	9,88	37,72	8,34

The following figure shows the block diagram of the system and the general plan indicating the location of the various compartments.

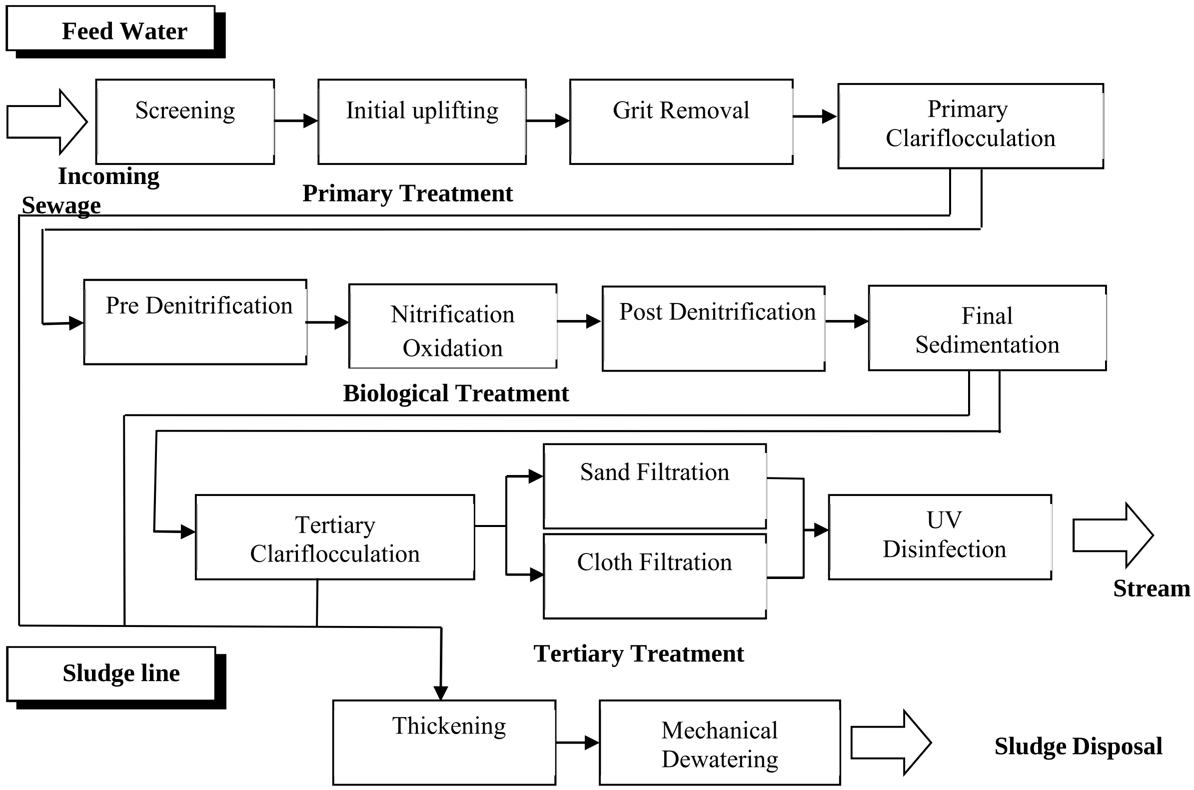


Figure 6. Comodepur Wastewater Treatment Plant Flow Diagram

Primary Treatment

Primary treatment is designed to remove gross, suspended and floating solids from raw sewage. It includes screening to trap solid objects and sedimentation by gravity to remove suspended solids. This level is sometimes referred to as “mechanical treatment”, although chemicals are often used to accelerate the sedimentation process. Primary treatment can reduce the BOD of the incoming wastewater by 20-30% and the total suspended solids by some 50-60%. Primary treatment is usually the first stage of wastewater treatment. Many advanced wastewater treatment plants in industrialized countries have started with primary treatment, and have then added other treatment stages as wastewater load has grown, as the need for treatment has increased, and as resources have become

Screening

Wastewater contains large solids and grit that can interfere with treatment processes or cause undue mechanical wear and increased maintenance on wastewater treatment equipment. To minimize potential problems, these materials require separate handling.

Preliminary treatment removes these constituents from the influent wastewater. Preliminary treatment consists of screening, grit removal, septage handling, odor control, and flow equalization.

Screening is the first unit operation used at wastewater treatment plants (WWTPs). Screening removes objects such as rags, paper, plastics, and metals to prevent damage and clogging of

Downstream equipment, piping, and appurtenances. Then the wastewater is uplifted to the grit removal section. The operation is conducted through three pumps (+ a reserve installed) vertical axis.

Grit Removal

Grit includes sand, gravel, cinder, or other heavy solid materials that are "heavier" (higher specific gravity) than the organic biodegradable solids in the wastewater. Grit also includes eggshells, bone chips, seeds, coffee grounds, and large organic particles, such as food waste. Removal of grit prevents unnecessary abrasion and wear of mechanical equipment, grit deposition in pipelines and channels, and accumulation of grit in anaerobic digesters and aeration basins. Grit removal facilities typically precede primary clarification, and follow screening. This prevents large solids from interfering with grit handling equipment. In secondary treatment plants without primary clarification, grit removal should precede aeration (Metcalf & Eddy, 1991).

Primary Clariflocculation:

Clariflocculation means the processes sum of coagulation, flocculation and sedimentation combined to obtain as principal objective the destabilization of a colloidal suspension and removal of the aggregates that are formed.

The clariflocculation together the effect of reduction of turbidity, allows to obtain a number of other important results, such as:

- Total Suspended Solids reduction (50 - 80%)

- Bacteria, viruses, and parasites concentration reduction (over 90% reduction of coliform);

- Inorganic micro polluting reduction (40 - 85% heavy metals reduction);

- Efficiency improvement filtration process;

- Residual organic fraction reduction (60 - 75% of BOD5)

- The possibility to realize simultaneously the phosphorous precipitation.

The coagulating agents most commonly used are the salts of iron and aluminum possibly in combination with anionic polyelectrolytes. Using aluminum chloride operating at pH levels below 6.5 and this avoids the use of lime while maintaining the low sludge production. Primary clarifloccilation has two lines, each of the two lines of primary clariflocculation presents characteristics listed in APPENDIX 1.

The water out of the flocculation are sent with a fixed rate to the pre-denitrification, the rest is sent through to the well spillway motorized lifting. The characteristics of the system has been described in APPENDIX 1

Biological Treatment

Biological treatment is an important and integral part of any wastewater treatment plant that treats wastewater from either municipality or industry having soluble organic impurities or a mix of the two types of wastewater sources. The obvious economic advantage, both in terms of capital investment and operating costs, of biological treatment over other treatment processes like chemical oxidation; thermal oxidation etc. cemented its place in any integrate wastewater treatment plant over a century.

Biological treatment removes the dissolved organic matter that come from primary treatment. This is achieved by microbes consuming the organic matter as food, and converting it to carbon dioxide, water, and energy for their own growth and reproduction. The biological process is then followed by additional settling tanks ("secondary sedimentation", see photo) to remove more of the suspended solids. About 85% of the suspended solids and BOD can be removed by a well running plant with secondary treatment. Secondary treatment technologies include the basic activated sludge process, the variants of pond and constructed wetland systems, trickling filters and other forms of treatment which use biological activity to break down organic matter.

Pre Denitrification

The tank of pre-denitrification is equipped with five submersible mixers and has the following geometrical characteristics:

Nitrification/Oxidation

Post Denitrification:

The post-denitrification tank is equipped with two submersible mixers and has the following geometrical characteristics:

Table 6. Post-denitrification tank characteristics

Length	mm	20500
Width	mm	16200
Liquid height	mm	7000
Working Volume	mc	2300

Post Aeration:

The post-aeration tank is equipped with a system of air blowing through membrane diffusers. The features are as follows:

Table 7. Post-Aeration tank characteristics

Length	mm	12500
Width	mm	3500
Liquid height	mm	7000
Working Volume	mc	306

Final Sedimentation:

The phase of biological sedimentation is carried out by four units of square type with round bottom and deck strength. To carry out the recycling of the sludge settled in pre-denitrification are installed three horizontal centrifugal pumps; for lifting the excess sludge to the thickeners are installed two horizontal centrifugal pumps; for the lifting of the supernatant, the sump is equipped with two submersible pumps (one installed).

The features are the following:

Table 8. Final Sedimentation tank characteristics

Height of cylindrical part	mm	22000
Average effective depth	mm	3500
Unit Surface Area	mq	484
Total Aera	mq	1936

*additional information is available in APPENDIX 1

Tertiary Treatment

Tertiary treatment is additional treatment beyond secondary. The purpose of tertiary treatment is to provide a final treatment stage to raise the effluent quality to the desired level. This advanced treatment can be accomplished by a variety of methods such as coagulation sedimentation, filtration, reverse osmosis, and extending secondary biological treatment to further stabilize oxygen-demanding substances or remove nutrients. Tertiary treatment can remove more than 99 percent of all the impurities from sewage, producing an effluent of almost drinking-water quality. The related technology can be very expensive, requiring a high level of technical know-how and well trained treatment plant operators, a steady energy supply, and chemicals and specific equipment which may not be readily available. An example of a typical tertiary treatment process is the modification of a conventional secondary treatment plant to remove additional phosphorus and nitrogen.

Disinfection, typically with chlorine, can be the final step before discharge of the effluent. However, some environmental authorities are concerned that chlorine residuals in the effluent can be a problem in their own right, and have moved away from this process. Disinfection is frequently built into treatment plant design, but not effectively practiced, because of the high cost of chlorine, or the reduced effectiveness of ultraviolet radiation where the water is not sufficiently clear or free of particles.

Coagulation Sedimentation

Chemical coagulation sedimentation is used to increase the removal of solids from effluent after primary and secondary treatment. Solids heavier than water settle out of wastewater by gravity. With the addition of specific chemicals, solids can become heavier than water and will settle. Alum, lime, or iron salts are chemicals added to the wastewater to remove phosphorus. With the chemicals, the smaller particles clump or 'floc' together into large masses. The larger masses of particles will settle out in the sedimentation tank reducing the concentration of phosphorus by more than 95%.

Tertiary Clariflocculation:

Each of the two lines clariflocculation presents the characteristics listed below:

Table 9. Tertiary Clariflocculation tank characteristics (Coagulation)

No. Tanks for line	n°	2
Tank Width	mm	3000
Length	mm	3000
Liquid Height	mm	5000
Liquid Volume per Tank/ Total	mc	45/90

Table 10. Tertiary Clariflocculation tank characteristics (Flocculation)

No. Tanks for line	n°	1
Tank Width	mm	6875
Length	mm	6600
Liquid Height	mm	5000
Liquid Volume	mc	226

Table 11. Lamella settling tank characteristics

No. Tanks for line	n°	1
Tank Width	mm	6875
Length	mm	16000
Liquid Height	mm	4300
Liquid Volume	mc	473
Lamellar Inclinatino	degree	55°
Lamellar Equivalent Surface	mq	1470
Lamellar type(material)		Parallel Corrugated sheets

Sand Filtration

A variety of filtration methods are available to ensure high quality water. Sand filtration, which consists of simply directing the flow of water through a sand bed, is used to remove residual suspended matter.

This section is consist of 7 units equipped with a blower for backwash (+ one standby unit) and a vertical axis pump (+ one standby unit).

Each unit has the characteristics listed below:

Table 12. Sand Filter characteristics

Width	mm	3600
Length	mm	3600
Surface	mq	13

Table 13. Filter Bed characteristics

Foundation Height (Gravel 10/15 mm)	mm	100
Bottom Height (Quartz 5/10 mm)	mm	200
Body Height (Sand 2/3 mm)	mm	1300

Fabric Filtration

Although there are a number of different methods of membrane filtration, the most mature is pressure driven membrane filtration. This relies on a liquid being forced through a filter membrane with a high surface area. Membrane filtration is designed to remove bacteria, viruses, pathogens, metals, and suspended solids.

This section consist of 3 cloth filtration units, each equipped with centrifugal vertical axis pump for backwashing and a system of nozzles to wash the cloth.

Each unit has the characteristics listed below:

Table 14. Cloth Filter characteristics

No. Of Filter Discs per Filter	n°	16
Disk Diameter	mm	2200
Filter Surface per Disc	m^2	7.8
Filter Surface per Filter	m^2	124.8
Dimenstion of Free Passage	μm	20
Dimension Length/Width/Height	mm	2286/.6580/.2429
Type of Fabric Filter	Pleated Polyester	

UV disinfection

The UV disinfection system is constituted by two channels in cement containing two rows of five modules each, and is equipped with a blower for the cleaning of the modules. Each channel of disinfection has the characteristics listed below:

Table 15. UV Disinfection System Characteristics

Channel Width	mm	1250
Length	mm	7000
Liquid Height	mm	1450
Liquid Volume	mc	12.7
No of Modules	n°	10
Lamps per Module	n°	40

Thickening

Thickening and dewatering are important components of biosolids management programs. The proper solids concentration is critical in stabilization processes. It optimizes the steps of conditioning, stabilization and dehydration by reducing the sizes of structures and operating costs.

Solids thickening reduces the volume entering subsequent solids processing steps. Thickening technologies include gravity, centrifugal thickening, gravity belt thickeners, and rotary drum thickeners.

The thickening stage is carried out by two units of circular type and bridge-wheel drive. To make the lifting of the thickened sludge dehydration three pumps are installed.

Table 16. Thickening Unit Characteristics

Inner Diameter	mm	16000
Height of Cylindrical Part	mm	3800
Average Effective Depth	mm	5150
Unit Surface Area	m^2	200
Total Area	m^2	400
Unit Volume	m^3	1030
Total Volume	m^3	2060

The mechanical dewatering of sludge is performed by means of two centrifuges (for emergency situations even with a filter press) having the following characteristics:

Table 17. Centrifuge Characteristics

Hydrulic Flow min/max	m^3/h	5/60
Solid Discharge	Kg SS/h	220/1200
Drum Rotation Speed	r.p.m	3250
Acceleration	g	2657
Drum Diameter	mm	450
Drum Length	mm	1910
Cylindrical Part Length	mm	1610
Conical Section Angle	Degree(°)	20
Installed Power	kW	37+15

Table 18. Filter Press System Characteristics

Fabric Width	mm	2500
Maximum Solid (with SS 2%)	Kg SS/h	500/400
Maximum Liquid (with SS 2%)	m^3/h	25/20

The dewatered sludge is sent to the storage silo by means of two mono pumps with the following features:

Table 19. Pump Characteristics

No. Of Units	n°	2
Flow	m^3/h	1.5/6
Prevalence	bar	24

Chapter 3

MBR configurations

Historical background

Vast number of trial-and-errors during the last two decades has led to the outcome of the current MBR technology with submerged or immersed membranes. Actually, most of misunderstandings, misguidance, and false estimations have been corrected and some of the most significant ones are as follows.

In early days, the importance of fine screens to eliminate debris were often underestimated due to the fact that membranes were taken to be compatible with non-biological debris. The ragging or clogging of membranes by fibrous materials, especially for hollow fiber modules, is unfavourable for the whole process of MBR. The debris rooted in hollow fiber bundles can be only picked by hand while mechanical surface scouring after having all membrane panels taken from the frame is normally and only method in flat sheet membranes. This problem has been widely solved when mechanical screens with 2-3 mm pore size or less was introduced. Nonetheless, this issue is still remaining because of the improper screens or glitches in installation (Stefanski, 2011).

Paying attention to plant footprint reduction, aeration basins were usually designed at to high MLSS like 20 g/L or 30 g/L. membrane scouring can be hampered by the high MLSS by slowing down upflow; moreover, biopolymer concentrations would be increased as well by hampering oxygen transfer. Besides, the slow mass transfer in and out of the membrane bundle could give rise to hollow fiber clogging. Furthermore, a great degree of membrane fouling was usually experienced in the plant designed for high MLSS. The design MLSS has fallen crucially since late 1990's; currently, the optimum MLSS in aeration basin is taken into account to be 8-12 g/L whereas some MBR are designed at some lover MLSS at 6 g/L.

It was thought that MBR would be compatible with high F/M ratio; solids settling in clarifier would not be a concern in MBR anymore as membrane rejects solids 100%. Consequently, the F/M ratio, controlled low in conventional activate sludge (CAS) for a good biosolids settling, was not taken into account as an important factor in MBR. This misperception was merged together with the desire of saving footprint, which resulted in overly compact aeration basins. The arising F/M ratio not only brought low oxygen transfer efficiency (OTE), but also increased membrane fouling rates by increasing biopolymer concentrations. Today, F/M ratio of MBR is taken into account to be 1/3 to ½ of the ACS as explained here.

MBR was considered tolerable to organic loading shock. In CAS, organic loading rate has to be fixed stably as much as possible to grasp favorable sludge settling. Since there is no clarifier existing, it was considered to be an obsolete concept for MBR. Therefore, it was believed that reducing or removing holding or equalization tank was probable. It is somewhat discussable, but greatly varying F/M expedites membrane fouling in many situations. If DO is not remained high enough during the high organic loading, membrane fouling can become even more important to be considered.

Biosolids production or sludge yield of MBR had usually been considered much lower than that in conventional activated sludge (CAS) process; maybe, it was based on some lab- or pilot-scale experiment carried out at the outset of MBR technology in early 1990's or earlier, where unrealistically high SRT such as 50-100days was deployed. Furthermore, the commercially motivated enthusiasm to show the advantage of MBR technology probably propelled this half-truth. However, field engineers soon figured out that there was small difference in obvious biosolids production from MBR and CAS by early 2000's. Although MBR produces slightly less biosolids because of its longer SRT (12-30 days for MBR vs 5-10 days for CAS), it does not lose biosolids through the effluent. Consequently, the apparent excess biosolids productions are not much different and often not recognizable. Fig. 7 illustrates the simulation data based on ASM1 assuming varying degrees of suspended solids loss from CAS, where apparent Yobs are about same in the common SRT ranges of CAS and MBR.

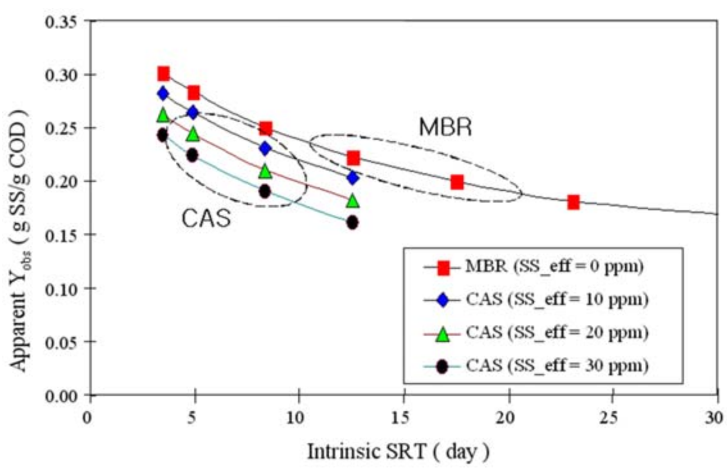


Figure 7. Comparison of observed sludge yield in CAS and MBR (unpublished data, Yoon 2003).

External/Sidestream

Submerged membrane bioreactor (sMBR) also called Cross flow or Sidestream MBR is an efficient technology for wastewater treatment that combines biological process and membrane filtration in one single stage. In the most usual configuration, submerged membrane hollow fibers are set in several panels. Air introduced from the bottom supply oxygen to the microorganisms but also to reduce fouling over the membrane fibers. The shear stress over the membrane surface is the main variable that contributes to fouling minimization (Martinez et al. 2010). Until immersed (or submerged) membranes were commercialized, MBR relied on cross flow filtrations using mostly tubular modules and some plate and frame modules. The external installation of membrane in sMBR system commonly provides a reliable performance and easy maintenance.

In External Membrane, the membrane modules are installed outside the reactor, Figure 8. In this system, the mixed liquor from the reactor is pumped into the external membrane module. External MBRs are also commercially used in industries as these require less membrane area compared to submerged MBRs and work better for high strength wastewater with poor filterability.

However, these MBRs consume more energy and need additional space and also require more frequent cleaning. In addition, the cost of building cross flow membrane system is high because of high pressure to hold membranes in them.

The capital and operating costs associated with the membrane component of an MBR system will be significantly affected by respectively, the value of the absolute permeate flux and the specific flux or permeability (Sutton et. al. 2006). High capital costs and the high operating costs for liquid circulation make cross flow membrane system prohibitively expensive for most of wastewater treatment applications especially for municipal wastewater.

Internal/Submerged

In the immersed or internal membrane MBR system (Figure 8), the membranes are directly submerged in the bioreactor mixed-liquor, preferably located in compartments or a separate tank coupled to the bioreactor to minimize membrane cleaning efforts. This configuration typically involves the use of polymeric membranes (Tajima, 1988; Yamamoto, 1989). The driving force across the membrane is typically achieved by creating negative pressure on the permeate side of the membrane. The membrane component of this configuration involves substantially more membrane area per unit volume relative to the membrane component of the external MBR configuration.

In iMBR, air scouring of immersed membrane reduce membrane fouling. The specific energy required per permeate volume is less than one tenth of that of crossflow side-stream filtration. The predominance of iMBR is undisputable as over 99% of the total installed membrane surface area in Europe in the period of 2003-2005 was immersed membranes (Lesjean, 2008).

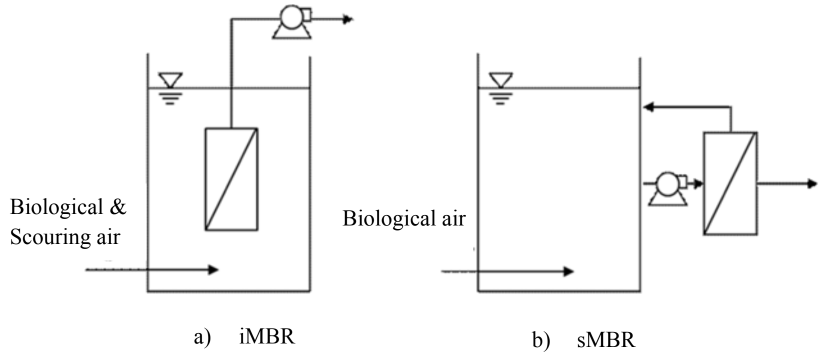


Figure 8. Schematics of the iMBR with immersed membranes and sMBR with side-stream cross flow membranes.

Membrane Configurations

Membrane module is the way the membrane is arranged into devices. Three membrane module configurations are predominant in the market: flat sheet, hollow fiber and tubular as shown below.

a) b) c)

Figure 9. Membrane Configuration: a) Flat Sheet. b) Hollow Fiber. c) Tubular

These membrane modules are designed and developed by industry manufacturers in order to achieve different characteristics based on different applications.

MBR Process

Fluxes and Membrane Area

Design fluxes of membrane is given by membrane manufacturers for typical municipal wastewaters with a set of conditions that need to be met to obtain the flux including MLSS ranges, temperatures, scouring air flow rate, F/M ratio, etc. Design fluxes again split depending on the limit in duration, e.g. monthly or weekly average flux, daily or hourly peak fluxes, etc. The number of membrane modules required to treat influent is determined in order not to violate any of the constraints given by the manufacture in terms of the flux and the durations at the water temperature in any given moment. But, determining required membrane area is not straightforward in real world due to the following reasons.

Actual wastewater flow rate pattern is quite complex because it varies depending on the time of the day, temporary weather conditions, season, types of the residence around the plant, existence of industrial sources, etc. In addition, the size and the operation strategy of the equalization tank also affect the actual flow rate to MBR. It is not only laborious to obtain rigorous information on the flow rate pattern, but also it is time consuming. As a consequence, MBR plant often has to be designed without having all required hydraulic information.

The rate of membrane fouling increases gradually at below the sustainable flux. But, it exponentially increases as it exceeds the sustainable flux. The sustainable flux itself is a rather fuzzy concept, which is dependent on biological condition. Since biological conditions tend to change depending on time, running membrane system under the manufacturers' guideline does not guarantee the successful operation. Therefore, the stability of the membrane system increases as the design flux is lowered below the guideline.

As a consequence of the varying flow rate and the varying sustainable flux, it is not completely certain how well membrane can perform during the peak time. Therefore, determining the membrane area is a matter of how much risk we would like to or must take due to the budget constraint.

Fig. 10 shows an example, where daily average flux varies between 15 LMH and 73 LMH while yearly average is 25 LMH. Assuming hourly flow variation is handled by equalization tanks, membrane system must be operated at above 40 LMH for more than 10 days in a row in April. If 50% more membrane modules are installed to handle the peak flow, yearly average flux and peak daily flux will decrease to 17 LMH and 50 LMH, respectively, but significant capital and operating costs will be required.

As discussed above, overlapping the membrane manufacturers' flux guidelines with the hydraulic profile and finding out the required membrane area in order not to violate the flux guideline is the standard procedure. But, given the many uncontrollable natural conditions, budget constraint, and the uncertainties in hydraulic data, design engineers also rely on the experiences obtained from the prior sites to estimate the number of membrane modules required.

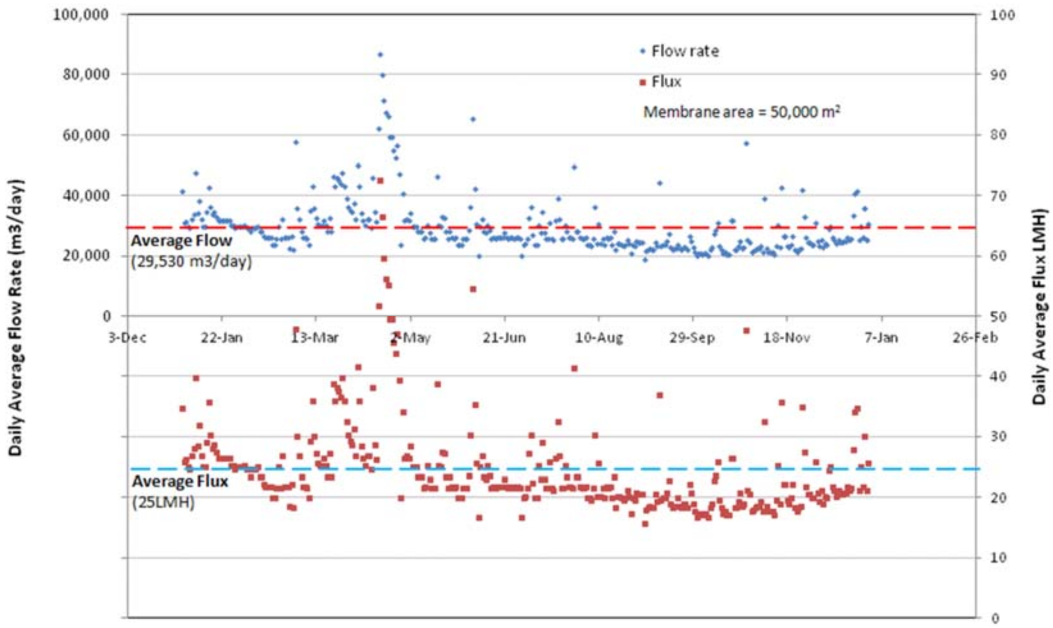


Figure 10. Daily average flux profile depending on daily average flow rate in a municipal WWTP, where total membrane area is 50,000 m2.

Fouling and Fouling Control

Although Membrane bioreactor is being increasingly applied for wastewater treatment replaced to the gravitational settling of the conventional activated sludge process (CASP) for the solid-liquid separation (Delgado et al., 2011, Meng et al., 2009 and Zahid and El-Shafai, 2011) and it has many advantages such as smaller footprint and better product quality, their wider application require better control of membrane fouling and permeate flux decline as foulant removal depends on chemical or physical cleaning. Because of the complete physical retention of bacterial flocs, the MBR has many advantages over conventional wastewater treatment processes, including reduced footprint, highly-improved effluent quality, higher biomass concentration and less sludge production (Le-Clech et al., 2006 and Zhu et al., 2011). In spite of their excellent retention characteristics, there are still problems that slow down use of membranes in these applications. Over time, membrane fouling and subsequent cleaning cause deterioration of membrane materials, resulting in compromised permeate quality and reduced membrane life spans. Chemical and biological fouling are major problems in membrane filtration due to reduced permeate flux, increased energy costs, and system downtime for maintenance. Biofouling in MBR systems is caused by the buildup of organic chemicals, microorganisms, and microbial communities at the membrane surface. Biofilms attached to a surface begin with cell adhesion, and progress to thick layers of extracellular polymeric substances (EPS), other organic chemicals, and a complex community of microbial cells that are often difficult to remove. Organic fouling due to the presence of natural organic matter (NOM) often leads to surface and internal pore fouling. More importantly, organic fouling progressively leads to biological fouling in so far as providing the organic nutrients for biofilm growth and sustenance. (Kim et al., 2012)

Biological performances/kinetics

COD removal and sludge yield

Simply due to the high number of microorganism in MBRs, the pollutants uptake rate can be increased. This leads to better degradation in a given time span or to smaller required reactor volumes. In comparison to the conventional activated sludge process (ASP) which typically achieves 95 percent, COD removal can be increased to 96 to 99 percent in MBRs (Kraume et al., 2005). COD and BOD5 removal are found to increase with MLSS concentration. Above 15 g/L COD removal becomes almost independent of biomass concentration at >96 percent (Drews et al., 2005). Like in the conventional ASP, sludge yield is decreased at higher SRT or biomass concentration. Little or no sludge is produced at sludge loading rates of 0.01 kgCOD/(kgMLSS d) (Stephenson et al., 2000). Because of the imposed biomass concentration limit, such low loading rates would result in enormous tank sizes or long HRTs in conventional ASP.

Nutrient removal

Nutrient removal is one of the main concerns in modern wastewater treatment especially in areas that are sensitive to eutrophication. Like in the conventional ASP, currently, the most common applied technology for N-removal from municipal wastewater is nitrification combined with denitrification. Besides phosphorus precipitation, enhanced biological phosphorus removal (EBPR) can be implemented which requires an additional anaerobic process step. Some characteristics of MBR technology render EBPR in combination with post-denitrification an attractive alternative that achieves very low nutrient effluent concentrations (Drews et al., 2005).

HF/ iMBR Performance

iMBR, combine fine screening with activated sludge biological process and advanced membrane filtration. They are used to treat municipal and industrial wastewater. Membrane modules are directly immersed in the activated sludge. The activated sludge is separated from the liquid as it passes through the membrane modules. Conventional sedimentation processes are not required as the small pores of an ultrafiltration membrane separate the suspended matter, bacteria, and viruses (pathogens) from the process liquid.

MBRs can be designed at higher MLSS concentrations because they are not affected by the limitations of gravity sedimentation for solid-liquid separation. SMBRs are typically designed for MLSS concentrations 8-12 g/L. Higher MLSS concentrations translate into advantages of High MLSS with Longer SRT same HRT, or Shorter HRT same SRT.

Membrane provides an absolute barrier and effluent quality is no longer a concern. Membranes increase the distance between reclamation and the risk of microbial disease, pathogens are removed by size exclusion.

All in all, considering the above advantages and disadvantages we can conclude hollow fiber iMBR system can be potentially a good choice to reach our purpose. In the next chapters we investigate a MBR system whether a good alternative or not.

Chapter 4

Methodology and Design

Membrane Design

Sustainable design flux

The sustainable flux is not a clearly defined concept, which is dependent on biological condition of feed flow. The stability of the membrane system increases as the design flux is lowered below the guideline. A summary of full scale data taken from case studies is given in different researches.

Flux, permeability, clean frequency and protocol, membrane aeration are key parameters regarding the membrane operation, and thus the maintenance of flow through the plant.

As we have varying sustainable flow rate, it may influence on the performance of membrane during peak time. Therefore, the membrane area should be determined based on the budget constraint and the level of risk we are going to take.

Flux, Permeability and Specific Aeration Demand

Extending the analysis to available full-scale data from Table 20 provides a more reasonable basis for an analysis and establishing appropriate operating conditions. A comparison of the key parameters of flux and SADp for the Kubota and Zenon mean full-scale plant data for municipal wastewater reveals similar mean fluxes but with lower aeration demand.

Table 20. Full scale data for wastewater treatment plant (The MBR book, Judd 2010)

		Flux, LMH		Permeability, LMH/bar		SAD_m $Nm^3/(m^2.h)$		SAD_p	
Config.		mun	Ind	mun	ind	mun	ind	mun	ind
FS	Mean	19.4	13.4	261	-	0.57	0.80	27.5	91.9
	%SD	21	17	66	-	67	93	56	98
	Data	12	5	8	0	10	5	10	5
HF	Mean	19.5	15.4	104	47	0.30	0.23	15.4	16.5
	%SD	39	33	65	87	35	36	41	59
	Data	14	9	12	7	11	6	11	6

Table 21. O & M Data (The MBR book, Judd 2006)

	Kubota	Mitsubishi Rayon	Zenon	NoritX-Flow
Membrane aeration capacity (Nm^3/h)	90-180	75-120	100(cycled)	140
Cycle (min)	8 on/2 $relax^a$	20h on/4h off^c	bflsh^f	20h on/4h off^3
• Normal	8.3-12.5	5-8	20^e	15-20, 37^i
• Peak	32.5-42	20.3-30.6	35^e	50
Biological aeration capacity (Nm^3/h)	160	160	100	140
F/M ratio	0.04-0.18	0.02-0.14	0.04-0.18	0.04-0.12
HRT (h)	10.2-15.4	15-22	7.6-12.3	15.2
SRT (day)	27-70	31-87	26-51	42-66
MLSS (g/L)	10.5-12	8.9-11.6	10.4-11.2	
Chemical cleaning reagents	NaOCI, 0.5% Oxalic acid 1%	NaOCI, 0.5% Followed by acid	NaOCI, 1% Followed by 0.3% citric acid	NaOCI, 0.5% Followed by Ultrasi
Derived data				
SAD_{m^a}(Nm^3/hm^3))	0.75	0.28-0.38	0.54^g	$0.33\text{-}0.6^h$
SAD_p(Nm^3 air/m^3 $permeate$)	60-90 normal 18-23 peak	48-56 normal 12-14 peak	27 15	30-40 normal 12-16 peak
Mean permeability,	200-250 vv/o r^a	200 $normal^d$	200-250	250 normal
LMH/bar	500-800 vv r^a 350 peak vv r^a	140-150 $peak^d$	320-350 after clean	75-200 peak
Permeability decline $\Delta k \Delta t$, LMH/(barh)	1.5^b	0.39^b	20^b	

a. Relaxation introduced mid-wat through phase I: permeability data refers to without (w/o) and with (w.) relaxation.
b. Refers to peak flux operation for the Zenon membrane this was 60 LMH.
c. Night-time relaxation introduced during phase Ⅱ . Along with backflushing at 20 LMH.
d. Assumed to be with relaxation
e. Refers to 500c module
f. Authors state "ratio of net to gross flux was 83-85%"; bflsh = backflushed.
g. Intermittent operation.
h. Night-time relaxation introduced during Phase Ⅱ.
i. With weekly maintenance clean.
j. Combination of sulphuric and phosphoric acid.

We focused on two well-known manufactures which are more commonly used in Italy, Kubota and Zenon manufactures. Based on the given table and Zenon and Kubota manufacture data 12 L/m3.h and 10 L/m3.h has been chosen conservatively as design flux for Zenon and Kubota.

Determine required SAD_m

Specific air demand based on membrane area (SAD_m) is defined as scouring air flow rate per membrane area. It is necessary to aerate the membrane unit to scout solids. Since the relationship between aeration and flux decline is not well understood, the membrane aeration value is not defined theoretically and in most cases the aeration value is recommended by manufactures (APPENDIX 2).

The key factor to energy demand in submerged systems is the specific aeration demand, the ratio of Q_A either to membrane area (SAD_m) or permeate volume (SAD_p)

$SAD_m = Q_A/A_m$

Where Q_A is membrane aeration rate (m^3/hr) and A_m is total membrane surface area (m^2)

$SAD_p = Q_A/ (J*A_m)$

It ranges 0.18-0.60 m3 air/m2 membrane area. SAD_m of 0.29 Nm3/m2.h has been chosen according to the manufacturer's design specifications for both Kubota 515 RW and Zenon ZW500D modules (APPENDIX 2). Based on the given value for SAD_m and the chosen design flux we can calculate the SAD_p for both modules.

For Kubota:

SADp= 1000*0.29 Nm^3/ (m^2.h)/ 12 L/ (m^2.h) = 24.17 m^3 Air/ m^3 permeate

For ZENON ZW500D:

SADp= 1000*0.29 Nm^3/ (m^2.h)/ 10 L/ (m^2.h) = 29 m^3 Air/ m^3 permeate

Determine membrane area required

Required membrane area, $A_m = \frac{2519\ m^3/hr}{12\ L/(m^2.hr)} = 209877.60\ m^2$

And since total membrane area per unit for Kubota 515 Panel RW is 580 m^2 the unit required for treatment is 362

Similarly for Zenon ZW500D:

Required membrane area, $A_m = \frac{2519\ m^3/hr}{10\ L/(m^2.hr)} = 251853.13\ m^2$

And total membrane area per unit of for Zenon ZW500D is 1651.2.

Determine required membrane air scouring capacity

The simple explanation for increased energy usage at MBR plants is membrane air scouring. Energy requirements are of primary interest that should be considered in an MBR system (Livingston, 2009) Aeration consumption has two forms: energy/air - to provide oxygen for biological nutrient removal, and energy - to provide scouring of the membrane to prevent clogging.

Air scour energy in an MBR system causes a high turbulent and surface contact to remove solids particles that attach to the surface of the membrane and to protect against membrane fouling. Membrane fouling can cause a lower production, less membrane life, and higher operational cost. The permeability decline associated with particle deposition on the membrane surface is a major problem in membrane filtration.

Moreover, unlike the circulation pump, air scouring does not produce extra shear stress that disrupts biological floc. Consequently, air sparging is used not only for immersed membranes, but also for the hollow fiber membranes housed in pressure vessel (Bérubé, 2006).

To determine the net air required for biological treatment, first we should calculate the M_O (Oxygen requirement for biological treatment) and M_m (Oxygen transferred by membrane aeration).

$$M_O(kg/h) = Q(S - S_e) - 1.42M_{X,Bio} + 4.33QNO_x - 2.86Q(NO_x - NO_e) = 529.97\ kg/h$$

For Kubota 515 Panel RW

$$M_m = Q_{A,m}\rho_A(SOTE_{coarse}y_{coarse})O_{A,m}\alpha\beta\Phi$$
$$= 24 * 60864.51\frac{m^3}{h} * 1.2\ \frac{kg}{m^3} * 0.02\frac{1}{m} * 2.3\ m * 0.232 * 0.43 * 0.95 * 0.83$$
$$= 6321.13\ kg\frac{O_2}{d}$$

So the net air flow required for biological aeration is:

$$Q_{A,b} = \frac{M_O - M_m}{\rho_A(SOTE_{fine} * y_{fine})O_{A,m}\alpha\beta\Phi} = \frac{\left(636.59 - \frac{6321.13}{24}\right)kg/hr}{1.2*0.05*5*0.232*0.51*0.95*0.83} = 13379.77\ Nm^3/h$$

And for Zenon ZW500D the oxygen transferred by membrane aeration is:

$$M_m = 7585.36\ kg\ O_2/d$$

$$Q_{A,b} = \frac{\left(636.59 - \frac{7585.36}{24}\right)kg/hr}{1.2 * 0.05 * 5 * 0.232 * 0.51 * 0.95 * 0.83} = 11491.30\ Nm^3/h$$

Submerged membrane bioreactors (MBRs) are often considered to be the best available technology for the treatment of wastewater, although immersed modules tend to suffer from membrane fouling and clogging especially in MBR due to the high membrane packing density and high MLSS. At the lower MLSS concentration, the MBR produced a higher water quality in comparison to the AS plant operating under the same conditions. The use of membranes was very efficient in removing COD content and pathogens in comparison with the gravity settling used in the AS plant. The MBR maintains higher MLSS concentrations than it is possible for an AS plant, which further improves the treated water quality for the same feed water concentration and HRT. This means that water productivity is higher for the MBR than for the AS system. Further improvement in the sustainable flux may take place if the fine-bubble diffuser is replaced by a coarse-bubble diffuser, as the larger bubbles are more capable of effective membrane scouring. In addition, it may be worth repeating the trials at the lower MLSS concentrations as a higher sustainable flux is expected to be achieved, further reducing the running cost as long as treated water quality is maintained.

Aeration plays a key role in mass transfer on membrane surface. It is crucial to design membrane module that allows efficient use of scouring air while maximizing mass transfer in the internal spaces of membrane module. Submerged membranes typically require coarse bubble aeration (air scouring) to remove foulants and sustain filtration capacity. So if it design or run inefficiently, coarse-bubble aeration systems can significantly impact the overall turndown capabilities of a system and drive up energy bills.

In addition, in the case of hollow fiber membrane, rising bubbles also increase random fiber movement that causes acceleration and deceleration of fibers in liquid, which greatly increases the anti-fouling effect.

As it has been mentioned to air scouring consumes in two parts, air fouling and nitrogen removal. In the following sections each part has been calculated.

Determine minimum membrane tank volume

Considering a conservative value for the membrane packing density in the membrane tank (ϕ_{tank}) based on ZENON and Kubota manufactures' data allows the minimum membrane tank volume ($V_{m,min}$ – Eq.(3.21)) to be determined. The larger the discrepancy existing between Q and Q_{peak} leads to the greater membrane area and more aeration demand and consequently the higher the CAPEX (capital cost) and OPEX (operational cost). It may therefore be more economical to install buffering capacity for flow equalization as it has been used in Comodepur WWTP.

$$V_{m,min}(m^3) = \frac{A_m}{\varphi_{tank}}$$

$$For\ Kubota(515\ Panel\ RW)2003\ \ V_{m,min}(m^3) = \frac{209878\ m^2}{115(1/m)} = 1825\ m^3$$

$$For\ ZENON\ ZW500D\ 2010\ \ V_{m,min}(m^3) = \frac{251853\ m^2}{253(1/m)} = 995.5\ m^3$$

Determine aerobic SRT

Since ammonia-oxidation kinetics are rate limiting compare to nitrite-oxidation kinetics in nitrification systems operated at temperatures below 28 °C, designs are based on saturation kinetics for ammonia oxidation as given below, assuming excess DO is available.

$$\mu_n = \left(\frac{\mu_{nm} N}{K_n + N}\right) - k_{dn}$$

Where μ_n= specific growth rate of nitrifying bacteria, g new cells/ g cells.d

μ_{nm}= maximum specific growth rate of nitrifying bacteria, g new cells/ g cells.d

N = Nitrogen concentration, g/m3

K_n= half velocity constant, substrate concentration at one-half the maximum specific substrate utilization rate, g/m3

k_{dn} = Endogenous decay coefficient for nitrifying organisms, g VSS/ g VSS.d

But for fully complete-mix activated sludge nitrification system, at temperature below 25°C with sufficient DO present, Nitrification rates are affected by the liquid DO concentration in activated sludge. To account for the effects of DO, the expression for the specific growth rate described above is modified as follows:

$$\mu_n = \left(\frac{\mu_{nm} N}{K_n + N}\right)\left(\frac{DO}{K_o + DO}\right) - k_{dn}$$

Where DO= dissolved oxygen concentration, g/m3

K_0= half-saturation coefficient for DO, g/m3

Other terms as defined previously.

Find $\mu_{n,m}$ at $T = 12^{\circ}C$

$$\mu_{n,m,12^{\circ}C} = \left(\frac{0.45\ g}{g.d}\right) * (1.07)^{12-20} = 0.26\ g/g.d$$

Find K_n at $T = 12^{\circ}C$

$$K_{n,12^{\circ}C} = 0.4 \frac{g}{m^3} * (1.053)^{12-20} = 0.26\ g/m^3$$

Find k_{dn} at $T = 12^{\circ}C$

$$k_{dn,12^{\circ}C} = 0.08 \frac{g}{g.d} * (1.04)^{12-20} = 0.06\ g/g.d$$

Substitution in formula and solve for μ_n

$$\mu_n = \left\{\frac{0.26\ g/g.d * 1.0\ g/m^3}{[(0.26 + 1.0) g/m^3]}\right\} * \left\{\frac{1.5\ g/m^3}{(0.5 + 1.5)\ g/m^3}\right\} - 0.06\ g/g.d = 0.097\ g/g.d$$

Determine the theoretical SRT and design SRT.

Find theoretical SRT:

$$SRT = \frac{1}{\mu_n}$$

$$SRT = \frac{1}{0.097\ g/g.d} = 10.32\ d$$

Determine the design SRT

$$FS = {}^{TKN_{peak}}\!/_{TKN_{average}} = 1.5$$

$$Design\ SRT = FS * theoritical\ SRT$$

$$Design\ SRT = 1.5 * 10.3 = 15.48\ d$$

Biomass Production Calculation

The design of the sludge-handling and disposal/reuse facility is based on the sludge production yield from published data from similar facilities. If the sludge-handling facilities are undersized, then the treatment process performance may be compromised and it can be accumulated. Similarly, the sludge capacity of the activated-sludge system will be exceeded and excess solids will exit in the secondary clarifier effluent, potentially violating discharge limits.

The formula for calculation of sludge production is given below.

$$P_{X,VSS} = \frac{QY(S_\circ - S)(1\ kg/10^3 g)}{1 + (k_d)SRT} + \frac{(f_d)QY(S_\circ - S)SRT(1\ kg/10^3 g)}{1 + (k_d)SRT} + \frac{QY_n(NO_x)(1\ kg/10^3 g)}{1 + (k_{dn})SRT} + Q(nbVSS)(1\ kg/10^3 g)$$

$$QY(S_\circ - S) = 60444 * 0.4 * (245.3 - 0.9) = 5909.63\ kg/d$$

$$(f_d)QY(S_\circ - S)SRT = 0.15 * 60444 * 0.4 * (245.3 - 0.9) * 15.48 = 1203.56\ kg/d$$

$$1 + (k_d)SRT = 1 + 0.088 * 15.48 = 2.36$$

$$1 + (k_{dn})SRT = 1 + 0.06 * 15.48 = 1.91$$

$$QY_n(NO_x) = 60444 * 0.26 * 19.04 = 300.41 \ kg/d$$

$$Q(nbVSS) = 60444 * 3/_{1000} = 179.29 \ kg/d$$

$$P_{X,bio} = \frac{5909.63 \ kg/d}{1.57} + \frac{1203.56 \ kg/d}{1.57} + \frac{300.41 \ kg/d}{1.91} = 3174.63 \, kg/d$$

$$P_{X,VSS} = \frac{5909.63 \ kg/d}{1.57} + \frac{1203.56 \ kg/d}{1.57} + \frac{300.41 \ kg/d}{1.91} + \ 179.29 \ kg/d$$
$$= 3353.92 \ \ kg/d$$

$$P_{X,TSS} \ = \frac{3174.63 \ kg/d}{0.80} + \ 179.29 \ kg/d + (60444 \, m^3/d)(18.3 \ g/m^3)(1 \, kg/10^3 g) =$$
$$= 5247.23 \ kg/d$$

$$NO_x = TKN - Ne - 0.12 \frac{P_{X,bio}}{Q} = 33.8 \, {}^{g}/_{m^3} - 8.49 \, {}^{g}/_{m^3} - 0.12 \frac{g \ N}{g \ VSS} * \frac{(3174.63 \ kg/d)(10^3 g/Kg)}{60444 \, {}^{m^3}/_{d}} = \ 19.04 \ {}^{g}/_{m^3}$$

Determine Aerobic Volume

$$Mass = P_x(SRT)$$

$$X_{VSS} * V = P_{X,VSS} * SRT$$

$$Mass \ of \ MLVSS \ \ = \ 3353.92 \ \frac{kg}{d} * 15.48 \ d = 51934.46 \ kg$$

$$Mass \ of \ MLSS \ = \ X_{TSS} * V = P_{X,TSS} * SRT = \ 5247.23 \ * 15.48 = \ 81251.85 \text{ kg}$$

$$@ \ MLSS = 8000$$

$$V = \frac{(81251.85 \text{ kg})({}^{10^3 g}/_{1 \ kg})}{8000 \ {}^{g}/_{m^3}} = 10156 \, m^3$$

Denitrification

Determine recirculation ratio

$$r_{int} = \frac{NO_x}{NO_e} - 1 = \frac{19.04}{4.55} - 1 = 3.18$$

Calculation of fraction of active biomass in anoxic zone

$$X_b = \left[\frac{Q\ (SRT)}{V}\right]\left[\frac{Y\ (S_0 - S)}{1 + (k_d)SRT}\right]\left[\frac{r_{int}}{r_{int} + 1}\right] = \frac{5909.63 * 15.48}{10156.5 * 2.36} * \frac{3.18}{1 + 3.18}$$
$$= 2908.02\ g/m^3$$

Calculation of nitrate load to anoxic zone

$$NO - loading\ in\ g/d = \ Qr_{int}(TKN - N_e - 0.12\ M_{X,bio}/Q)$$
$$= 60444 * 3.18 * 19.04/1000 = 3664.13\ \ kg/d$$

Choose anoxic HRT and determine F/Mb ratio and bCOD fraction

$$\tau = \frac{3\ h}{24\frac{h}{d}} = 0.13\ d$$

$V_{anox} = \ \tau * Q = 0.13$ d * 60445 m^3/d = 7555.6 m^3

$$F/_{M_b}\ in\ \ g\ BOD/(g\ TSS\ d) = \ \frac{Q * S_0}{V_{anox}X_b}$$

$$= \frac{60445\ \ m^3/d *\ \ 245.3\ \ g/m^3}{7555.6\ m^3 * 2908.02\ \ g/m^3} = 0.67\ \ g/g.d$$

$$Fraction\ of\ rbCOD = \ rbCOD/bCOD$$

$$= \ \frac{20.3}{245.3} * 100 = 8.27\ \%$$

Based on Judd book $SDNR_b = 0.21\ g/g.d$ @20^0C

$SDNR_{12} = 0.21\ (1.026)^{12-20} = 0.17\ g/g.d$

$$NO_r = V_{anox} * SDNR * (MLVSS, biomass)$$

$$= 7555.6\ m^3 * 0.17\ \ g/g.d * 2908.02\ \ g/m^3 = 3757.56\ \ kg/d$$

Therefore, $\tau = 3\ h$ is acceptable.

SRT, HRT and F/M ratio are interconnected. Longer SRT directly means longer HRT and lower F/M ratio for a system with a fixed reactor volume, fixed influent strength, and fixed MLSS. If reactor volume increases while MLSS is constant, both SRT and HRT increase and F/M ratio decrease.

Determine total SRT, HRT and total process volume

$$\theta_{x,process} = \frac{V_{aer}X_{aer} + V_{anox}X_{anox}}{Q_w X_{aer}} = \theta_{x,aer} + \theta_{x,anox}$$

$$= \frac{10156.48\, m^3 * 8000\, g/m^3 + 7555.6\, m^3 * 2908.02\ g/m^3}{655.90\ m^3/d * 8000\, g/m^3}$$

$$= 19.67\, d$$

$$V_{process}(m^3) = V_{anox} + V_{aer}$$

$$= 10156.48 + 7555.6 = 17712.07\, m^3$$

Design aeration system

Since iMBR has a membrane system in addition of biological section there is an important difference between aeration of CASP and iMBR system. There is a portion of oxygen from membrane aeration contributes in carbonaceous degradation and nutrient removal. As in all aerobic biological systems, biomass contained in the MBR requires oxygen to perform diverse chemical reactions. The right amount of oxygen needs to be provided to the micro-organisms and wastewater, in response to their three specific demands:

• Carbonaceous biochemical oxygen demand (BOD): conversion of the carbonaceous organic matter in wastewater to cell tissue and various gaseous end products,

• Nitrogenous BOD: ammoniacal nitrogen is oxidized to the intermediate product nitrite, which is then converted to nitrate; this process is nitrification,

• Inorganic chemical oxygen demand (COD): oxidation of reduced inorganic compounds within the wastewater.

Calculation of oxygen demand for biological treatment

$$M_O(kg/h) = Q(S - S_e) - 1.42M_{X,Bio} + 4.33QNO_x - 2.86Q(NO_x - NO_e) = M_m + M_b$$

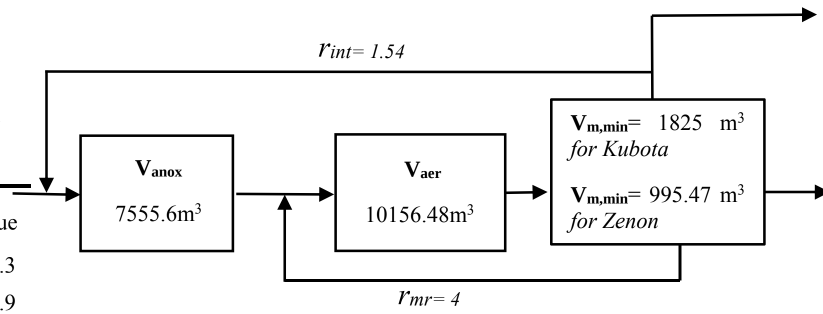
$$= (60445/(\,245.3 - 1.5)\, g/m^3\, kg/1000\, g - 1.42 * 4243.55\ kg\, VSS/d +\ 4.33 * 60445\, m^3/d * 15.14$$
$$- 2.86 * 60445\, m^3/d * (15.14 - 6.06)\, g/m^3)/24h = 529.97\ kg/h$$

Table below demonstrates the aeration system design and operating parameters, used in cost analysis. As it has obviously shown in table the efficiency of aeration system in biological system is much higher than membrane system because of less MLSS concentration, higher aerator depth y, and higher oxygen transfer efficiency SOTE in fine bubble system of compare to coarse bubble system MBR.

Table 22. Aeration system operating parameters and design calculations.

Parameter	Unit	Biology	Membrane
Diffuser type		fine bubble	coarse bubble
SOTEx	%/m	0.05	0.02
Air density, ρ_A	kg/m3	1.2	1.2
correction factor exponent, ω_x		0.084	0.084
mass % oxygen in air, $O_{A,m}$	%	23.2	23.2
mass transfer correction, solid, α	-	0.51	0.43
mass transfer correction, salinity, β	-	0.95	0.95
F fouling factor		0.9	0.9
mass transfer correction, temperature, Φ	-	0.83	0.83
Aerator depth, y_x	m	5	2.3
Point of Air released	m	0.5	0.5
Air flow rate, membrane tank, $Q_{A,m}$	Nm3/h	73037.41	
O_2 transferred by membrane aeration, Mm	kg/d	7585.36	
O_2 to be provided by biology aeration, Mo	kg/d	12719.36	-
Air flow rate, biotank, $Q_{A,b}$	Nm3/h	-	
Suspended solids correction factor, α		0.51	0.43

All in all we can summarize the design process parameters in the following diagram for both Kubota and Zenon manufactures.



Constituent	Unit	Value
BOD_5	g/m^3	153.3
COD	g/m^3	284.9
TSS	g/m3	92.5
TKN	g/m3	33.8

Constituent	Unit	Value*	Typical MBR effluent Quality
BOD_5	g/m^3	0	<5
COD	g/m^3	45	
TSS	g/m3	3.94	<5
TKN	g/m3	3.9	
Total N	g/m3	8.49	<10
Ammonia	g/m3	1	<1

*Assumed based on standard water effluent quality

Figure 11. Schematic overview of design MBR

Chapter 5

COST BENEFIT ANALYSIS

Introduction

A cost benefit analysis is a tool to determine the beneficial changes in plant design and operation. Two indexes can be used to quantify the impacts on cost and effluent quality. An objective economic index, net present value (NPV), calculated for a plant lifetime which integrates both investment, fixed and variable operating costs of a wastewater treatment plant (CAPEX and OPEX) over the plant life time and effluent quality index which shows the impact of an upgraded system. The development of the cost criterion may be specific to each particular case, especially to assess variable operating costs. Provided whole-life costs are considered, then the net present value (NPV) provides a reasonable indication of overall cost, NPV being a function of CAPEX, OPEX and the plant life and residual end-of-life value.

When designing a new wastewater treatment plant (WWTP) or when upgrading an existing one, different treatment alternatives and operating strategies may be evaluated with the help of cost index and effluent quality index (EQI) which present the potential pollution load to the receiving water body expressed in kg pollution per day.

Provided whole-life costs are considered, then the net present value (NPV) in units of USD/m3 treated water provides a reasonable indication of overall cost, NPV being a function of CAPEX, OPEX and the plant life and residual end-of-life value.

The methodology which has been used for Comodepur WWTP is based on the following figure associated with biokinetic Activated Sludge Model No. 1 (ASM1), using the default ASM1 values (Henze et al., 2000) for the biokinetic parameters.

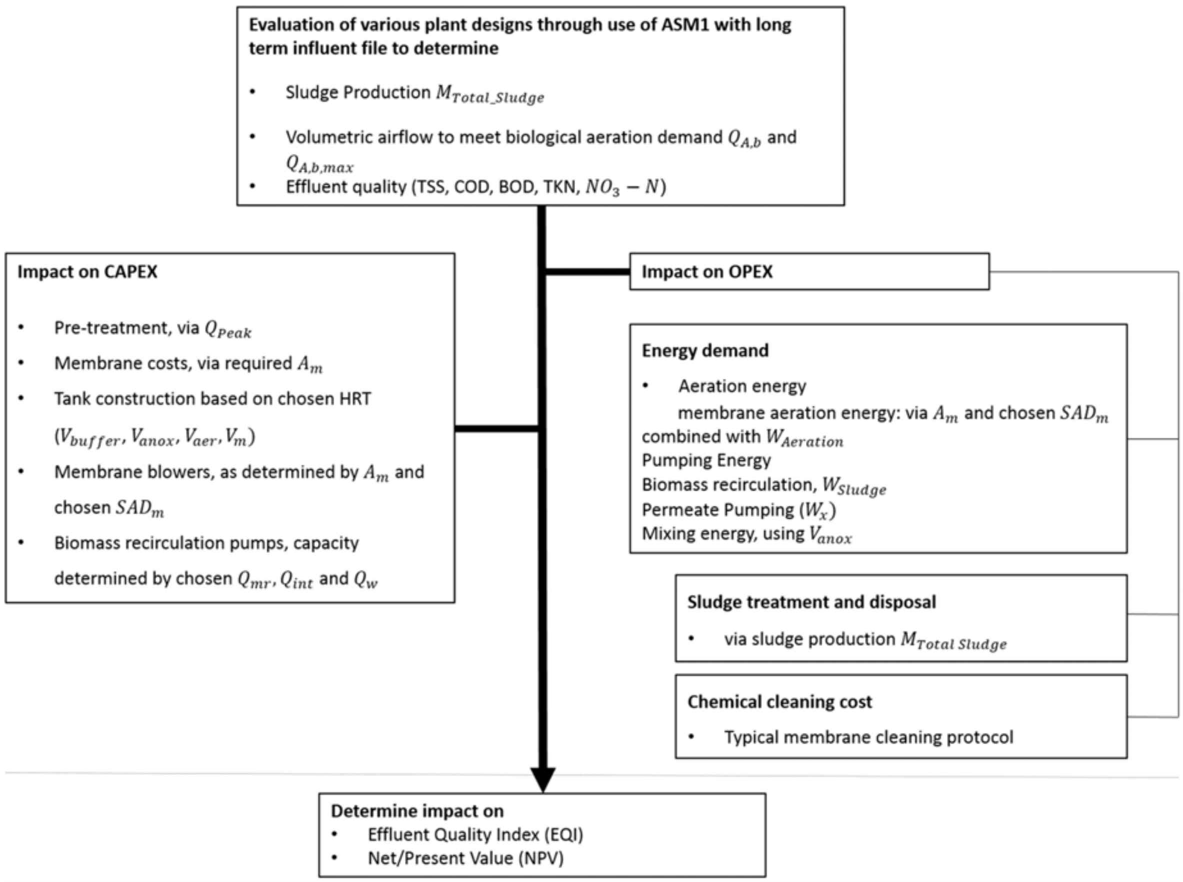


Figure 12. Schematic methodology for cost benefit analysis

Table 23. Equations for Determination of CAPEX, OPXEX, NPV and EQI.

Actual oxygen transfer rate, AOTR kg/d	$Q_A \rho_A O_{A,m} AOTE$
Actual oxygen transfer efficiency, AOTE, %	$SOTE_x y_x \frac{(\beta C^*_{T,av} - C)}{C^*_{20}} \Phi \alpha F_x$
Average dissolved oxygen saturation concentration for clean water at temp, T and tank depth h_x, $C_{T,av}$, gO_2/m^3	$0.5\ C^*_T (\frac{P_d}{P_{A,1}} + \frac{O_{out}}{O_{A,v}})$
Dissolved oxygen saturation concentration for clean water at temp, T and 1 atm, C^*_T, gO_2/m^3	$14.65 - 0.41T + 7.99 \cdot 10^{-3} T^2 - 7.78 \cdot 10^{-5} T^3$
Pressure at the bottom of aeration tank, P_d, Pa	$P_{A,1} + \rho_{sludge} g h_x$
O_2 in air leaving the surface of the aeration tank O_{out}, %	$\frac{O_{A,v}(1 - AOTE)}{1 - O_{A,v} AOTE}$
Blower power requirement, $W_{Aeration}$, kW	$\frac{P_{A,1} T_{K,1} \lambda}{2.73 \cdot 10^5 \xi (\lambda - 1)} \left[\left(\frac{P_{A,2}}{P_{A,1}} \right)^{1-1/\lambda} - 1 \right] Q_{A,x}$
Derived sludge pumping power requirement, W'_{sludge}, kWh/d	$\frac{E_q}{t_e - t_0} \int_{t_0}^{t_e} [Q_{int}(t) + Q_{mr}(t) + Q_w(t)] dt$
Power requirement for sludge pumping, W_{sludge}, kW	$\rho_{sludge} g Q_{sludge} \Delta h$
Power required for permeate pumping and backwashing, Wx, kWh/d	$\frac{1}{t_e - t_0} \int_{t_0}^{t_e} \frac{\Delta P_x Q_x}{\xi_p} dt$
Total sludge production, $M_{Total\ sludge}$, kg/d	$\frac{\left(M(TSS_{system})_{t_e} - M(TSS_{system})_{t_0} + \int_{t_0}^{t_e} TSS_w(t) Q_w(t) dt\right)}{t_e - t_0}$
Net present value, NPV, €	$\sum_{t=0}^{29} \frac{(CAPEX)_t + (OPEX)_t}{(1+i)^t}$
Effluent quality index, EQI, kg PU/d	$\frac{1}{t_e - t_0} \int_{t_0}^{t_e} [PU_{TSS}(t) + PU_{COD}(t) + PU_{BOD}(t) + PU_{TKN}(t) + PU_{NO}(t)] Q(t) dt$

The effect of MLSS on actual oxygen transfer rate (AOTR) has been considered through the α-factor. By calculating AOTE we are able to compute actual oxygen transfer rate which is a function of diffuser oxygen transfer efficiency SOTEx and fouling factor Fx (Krampe and Krauth, 2003 after Tchobanoglous et al., 2003; Germain et al., 2007; Henze, van Loosdrecht, Ekama, & Brdjanovic, 2008; Verrecht et al., 2008, 2010a; Strenstrom, & Rosso, 2008; Maere et al., 2009). By calculating the oxygen demand and the sludge production we are able to compute the EQI and CAPEX and OPEX and NPV during the evaluation period.

Capital Cost

Capital costs for MBR systems historically have tended to be higher than those for conventional systems with comparable throughput because of the initial costs of the membranes. In certain situations, however, in some cases, MBR systems can have lower or competitive capital costs compared with alternatives because MBRs have lower land requirements and use smaller tanks, which can reduce the costs for concrete.

Fleischer et al. (2005) reported on a cost comparison of technologies for a 12-MGD design in Loudoun County, Virginia. Because of a chemical oxygen demand limit, activated carbon adsorption was included with the MBR system. It was found that the capital cost for MBR plus granular activated carbon at $12/gallon treated was on the same order of magnitude as alternative processes, including multiple-point alum addition, high lime treatment, and post-secondary membrane filtration although MBR membrane price dropped to around 50 €/m2 from 400 €/m2 in the past decade.

Regarding evaluation of capital investment costs, pricing information (Table 2) was obtained from manufacturers or based on costs provided by end-users for similar items of equipment at full scale MBR plants (Brepols, 2010, Lesjean, 2008) (Table below). Minimum required membrane tank volume (Vm) is incorporated in Vaer. Each 10,000 m3 of membrane area is assumed to demand one membrane tank, and it's better to consider four or more membrane tanks to have flexibility in operation and cleaning. A conservative SADm is 0.29 Nm3/ (m2 h) and each membrane tank require one blower.

Cost of land, civil engineering, other electrical equipment and construction are excluded because they are all highly dependent to location.

Table 24. Parameter values and key assumption for determination of CAPEX and OPEX. (The MBR book, Judd 2010)

Parameter	Unit	Value	Parameter	Unit	Value
J_{net}	LMH	Zenon: 10 LMH Kubota: 12 LMH	Assumption for CAPEX calculation		
f_{anox}			Membrane Cost	€/m^2	50
θ_x	d	8.09	Civil works		
r_{mr}	-	4	Structural concrete	€/ m^2	400
r_{int}	-	3	Foundations	€/ m^2	171
Membrane packing density: aera per membrane tank, ϕ_{tank}	m^2/m^3	45	Assumption for OPEX calculation		
SAD_m	$Nm^3/(m^2h)$	0.29			
Blower inlet pressure, $P_{A,1}$	Pa	101,300	Mixing power demand	$kW/10^3$ m^3 V_{anox}	8
Blower outlet pressure, $P_{A,2}$	Pa	160,300	Energy cost	€/kWh	0.16
Blower inlet temperature, $T_{K,1}$	K	293	Sludge treatment cost	€/ton of DS	65
Blower efficiency, ξ	-	0.6	Citric acid 50%	€/ton	760
Specific heat capacity of air, λ	J/(kg.K)	1.4	NaOCl 14%	€/m^3	254
Aerator depth, y	m	5	Assumption for NPV calculation		
Total headloss in pipework, Δh	m	3			
Sludge pump efficiency, $\xi_{p,sludge}$	-	0.50	Membrane life	Year	Zenon:8 years Kubota:10 years
Permeate pump efficiency, ξ_P	-	0.75	Inflation	%	3
Transmembrane pressure, ΔP_m	Pa	35,000	Discount rate	%	6

Since Kubota and Zenon manufactures are two well-known membrane provider which are more commonly used in Italy, we have performed the cost analysis for both type of Membranes (Table below).

Table 25. Different MBR Manufactures' Technologies used

Technology tested	Point Loma, MWH*	Beverwijk, DHV**	Bedok/Ulu Pandan PUB	Pietramurata, Univ Trento†	Kloten Opfikon Eawag
Zenon	×	×	×	×	×
Kubota	×	×	×	×	×
MRE‡	×	×	×		×
Norit	×	×			
Huber	×	×			
Memcor	×			×	
Toray	×	×			
Asahi Kasei	×		×		
Puron	×				

*Adham et al. (2004); DeCarolis et al. (2009); Hirani et al. (2010)
** van der Roest et al. (2002); Lawrence et al. (2005).
†Guglielmi et al. (2007); Verrecht et al. (2008).
‡Mitsubishi Rayon Engineering

Although based on the full scale plant data for wastewater treatment (table below), the average net flux for iMBR is 19.5 LMH but we have conservatively considered 12 and 10 LMH for Kubota and Zenon membrane respectively.

Table 26. Full scale plant data for FS and HF MBR configurations**.**

		Flux, LMH		Permeability, LMH/bar		SAD_m $Nm^3/(m^2.h)$		SAD_p	
Config.		mun	Ind	mun	ind	mun	ind	mun	ind
FS	Mean	19.4	13.4	261	-	0.57	0.80	27.5	91.9
	%SD	21	17	66	-	67	93	56	98
	Data	12	5	8	0	10	5	10	5
HF	Mean	19.5	15.4	104	47	0.30	0.23	15.4	16.5
	%SD	39	33	65	87	35	36	41	59
	Data	14	9	12	7	11	6	11	6

Operational Costs

Operating costs for MBR systems are typically higher than conventional systems because of the higher energy costs used for air scouring to reduce membrane fouling. The air needed for the scouring is twice the amount needed to maintain aeration in a conventional activated sludge system. Since in membrane systems the sludge residence time is longer than those for conventional systems operating costs would be partially offset.

Operational costs were determined using the approach of the control strategy evaluation benchmark community (Copp et al., 2002), which was extended by Maere et al. (2009) for MBR applications. The opex analysis was limited to energy demand, sludge treatment and disposal, and chemical usage for membrane cleaning.

As illustrated in Figure below, the primary energy demands are related to aeration (66%), by far pumping is the second energy requirement. So, the key opportunities for energy reduction focus on aeration; however, all energy related elements should be considered. In order to provide the most cost effective and energy efficient system, it is important to look at opportunities related to design, operations, and equipment.

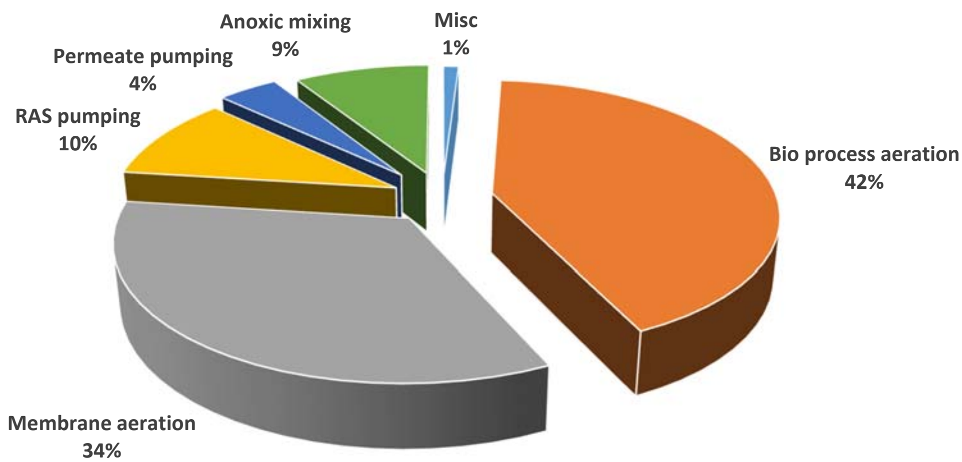


Figure 13. Operational energy demand participation.

Energy demand

The individual contributions to energy demand are described below, and a Italian- specific energy cost of 0.16 €/kWh used throughout.

There are some elements that should be considered to reduce the cost and energy requirement in designing a wastewater treatment plant. Various operational elements influence the overall energy efficiency of the MBR design. Currently the single largest energy cost is aeration – both for the biology and for the maintenance of the membranes. Hence, opportunities to reduce aeration have the potential to reduce the overall energy requirements significantly.

Aeration energy

Energy demand is a key cost factor when considering MBR technology. The main contributors to energy costs are sludge transfer, permeate production and, most significantly, aeration. Membrane aeration is normally achieved via coarse bubble aerators positioned beneath the membrane units. The influence of MLSS concentration (via the α-factor) and aerator type (fine and coarse bubble) on oxygen transfer was computed using the dedicated aeration model of Maere et al. (2009), combining several literature findings (Metcalf and Eddy, 2003; Henze et al.,2008; Verrecht et al., 2008; Krampe and Krauth, 2003; Germain et al., 2007; Stenstrom and Rosso, 2008). In order to achieve the optimum design parameters of MBR with which operational costs are minimized, aeration and sludge treatment costs should be estimated for various operational conditions. Generally sludge treatment cost and aeration cost were inversely proportional to each other, which means sludge treatment cost is minimized when aeration cost is maximized and vice versa. Therefore, there might exist an optimum point between the two extreme cases. However, sludge treatment cost turned out to overwhelm the aeration cost over the reasonable operational conditions. Therefore, sludge minimization was considered to be a key for the economical operation of MBR.

A value of 0.025 kWh.Nm-3 air has been specified for the aeration energy demand considering common practically measured values for blower outlet pressure (106300 Pa; for a typical aerator depth of 5 m and allowing for losses subjected to the pipework) and a blower efficiency ξ_B of 0.60 corresponding well with literature values (Verrecht et al., 2008) and data from blower manufacturers. The average total aeration energy in kWh.d-1 was obtained by summing blower power consumption for both membrane and biology blowers and integrating over the 365 day simulation period (Maere et al., 2009).

Filtration

One of the major obstacles, which stretch out more spread of membrane application is membrane fouling. Membrane fouling reduces the permeate production rate and increases the complexity of the membrane filtration operation. This is the most challenging issue for further membrane development and applications.

Filtration performances can be limited by membrane fouling and the aim of most studies about MBR process is to prevent or to limit fouling in order to enhance system performances. Consequently we should consider the parameters influencing fouling.

Permeate flux and transmembrane pressure (TMP) are the best indicators of membrane fouling. Membrane fouling leads to a significant increase in hydraulic resistance, results in permeate flux decline or TMP increase. In a system where the permeate flux is maintained by increasing TMP, the energy required to achieve filtration increases. As the number of filtration cycles increases, the irreversible fraction of membrane fouling also increases. In order to obtain the desired production rate, chemical cleaning is required for membrane to regain most of its permeability. The resultant makes the membrane process costly.

Membrane pore size, pore size distribution and pore geometry especially at the surface of the membrane has a considerable effect on fouling. The properties of the feed solution also significantly influence membrane fouling. Some of the important feed properties are solid (particle) concentration, particle properties, pH and ionic strength. Generally, an increase in the feed concentration results in a decline in the permeate flux (Vyas et al. 2000). Some other factors, such as: pH, ionic strength, and electric charges of particles, are also important. The effect of temperature on the permeate flux was investigated and found that at higher temperatures, the permeate flux increased, indicating a lower degree of fouling. Changing the feed temperature from 20°C to 40°C lead to an increase in the permeate flux up to 60% (Salahi et al. 2010).

Pumping energy

Sludge pumping requirements, for internal recirculation (Q_{int}, m^3/d), membrane recirculation (Q_{MR}, m^3/d) and wastage (Q_W, m^3/d), were determined from the expression of Maere et al. (2009), using a power requirement of 0.016 $kWh.m^{-3}$ of sludge pumped which was calculated from assuming a simple linear dependency of P_{Sludge} (Power required for sludge pumping) on sludge flow and assuming a total headloss Δh of 3m and a pump efficiency ξ_p of 50%. To calculate additional pumping energy for permeate pumping and backwashing, the expression provided by Maere et al. (2009) was applied.

Mixing energy

A typical constant mixing power requirement of 8 W per m3 of anoxic tank volume was used (Metcalf and Eddy, 2003), with no supplementary mechanical mixing required for the aerobic, membrane and buffer tanks.

Sludge production

Sludge production (Mtotal_sludge) (in kg/d) was calculated using the expressions of Copp et al. (2002), adapted for MBR use by Maere et al. (2009). Reported costs for sludge handling and disposal vary widely and location-specific, which accounts for chemicals, labour, treatment and disposal, 65€/ton has been considered as overall disposal cost of comodepur wastewater treatment plant based on costs for collection, thickening, digestion, dewatering, reuse.

Based on the formula given in table 22 and the assumptions given in table 23, parameters have been computed in the following table for both Zenon and Kubota membrane manufactures.

Table 27. Equations for Determination of CAPEX, OPXEX, NPV and EQI

Manufactures	ZENON ZW500D 2010	Kubota(515 Panel RW) 2003
Actual oxygen transfer rate, AOTR kg/d	16326.13724	19591.36469
Actual oxygen transfer efficiency, AOTE, %	0.963	0.963
Average dissolved oxygen saturation concentration for clean water at temp, T and tank depth h_x, $C_{T,av}$, gO_2/m^3	10.07	10.07
Dissolved oxygen saturation concentration for clean water at temp, T and 1 atm, C_T^*, gO_2/m^3	10.77	10.77
Pressure at the bottom of aeration tank, P_d, Pa	11.28	11.28
O_2 in air leaving the surface of the aeration tank O_{out}, %	9.24	9.24
Blower power requirement, $W_{Aeration}$, kW	5408.6	6490.4
Derived sludge pumping power requirement, W'_{sludge}, kWh/d	6326.7	6326.7
Power requirement for sludge pumping, W_{sludge}, kW	0.336	0.336
Power required for permeate pumping and backwashing, W_x, kWh/d	335.80	335.80
Total sludge production, $M_{Total\ sludge}$, kg/year	10,028,000.00	10,028,000.00
Net present value, NPV, €	15,987,246.26	14,165,383.28
Effluent quality index, EQI, kg PU/d	7478.05	7478.05

The Value of EQI for current Conventional ASP is 4822.04 which shows less potential pollutant load compare to those values for MBR system configurations.

Chapter 6

Conclusion

MBR have become an increasingly popular municipal wastewater treatment process alternative. The MBR technology is going to be a well-developed and mature technology and various authors denominate MBR as the best available technology for industrial and municipal, wastewater treatment (Kraume and Drews 2010, Lesjean, 2008). However, although MBR technology widely spread all over the world for more than a decade, MBRs are preferred over other treatment technologies mainly when certain criteria, i.e., high effluent quality, small footprint, easy retrofit and upgrade of old WWTP are the first priority. This is because of old problems which still remain unsolved despite continuous technological development. The retention of activated sludge and wastewater constituents cause a decrease in the membrane filtration performance. The efficiency of the filtration process in an MBR is governed by the activated sludge filterability, which is determined by the interactions between the biomass, the wastewater and the applied process conditions. Due to the interdependency of the mentioned factors and the dynamic nature of the feed and biomass, membrane fouling is a very complex phenomenon. Implemented strategies for prevention and removal of membrane fouling lead to an increase in the operational and maintenance costs of the treatment system. Consequently there is an optimum value for the diffuser lifetime as it has been mentioned in the previous chapter. In particular, the high energy requirements arisen from frequent membrane cleaning by air scouring remains a challenge in terms of energy consumption and overall cost efficiency of full-scale MBRs.

This study has investigated capital and operating cost of membrane bioreactors using membrane bioreactor (MBR) vs. conventional activated sludge (CAS) processes based on identical influent loading conditions considering new effluent water quality. The resulting capital and operating cost tables indicate that the system is most cost-effective. The purpose of this study was to analyze the performance and cost of the membrane bioreactor (MBR) designed for Comodepur wastewater treatment working under similar influent qualities and environmental conditions in comparison to the current existed CAS system considering new more stringent effluent quality.

Conceptual designs were completed for Comodepur wastewater treatment facilities using both Zenon and Kubota MBR technologies. The systems were designed to treat the same influent loading conditions as defined in Table below to reach the following mentioned effluent water quality (table 2).

Table 28. Characteristic of current WWTP

Parameter	influent Average	Effluent Average
Flow(mgd)	15.97	
BOD(mg/L)	153.3	-
TSS(mg/L)	92.5	264.3
COD(mg/L)	284.9	48.1
N_TKN	33.8	3.9

Table 29. Characteristic of MBR effluent.

Parameter	MBR Effluent Average
Flow(mgd)	15.97
Ammonia N(NH4-N)	1.00
Total N	8.49
COD	45.00
TP(phosphorus)	0.50

Based on the various results presented in this study, the following conclusions can be drawn.

In chapter 3, since Immersed (or submerged) membrane filtration was invented to save capital and operational costs and the specific energy required per permeate volume is less than one tenth of that of cross flow side-stream filtration which cause to significantly less energy cost compare to sMBR all in all considering the all the mentioned advantages and disadvantages we concluded that hollow fiber iMBR system can be potentially a good choice to reach our purpose.

In chapter 4, a new iMBR plant after biological section has been designed for the Comodepur wastewater treatment plant based on both Kubota and Zenon manufactures' data and the impact of both MBR on the functioning of full-scale MBR plants was analyzed. For the Kubota and Zenon cases which has been analyzed in this research, it can be concluded that both MBRs have some influence on the overall plant functioning, i.e., operation and performance and capital and operational costs. In addition, membrane selection does influence mainly operational strategies (filtration protocol, membrane cleaning and fluxes) considering different optimum lifetimes based on previous researches mentioned in chapter 4. The MBR plant layout has more general influence on plant functioning (operational flexibility and reliability, performance and O&M costs). Moreover, the activated sludge filterability was found to be independent of the membrane configuration but not of the MBR plant layout. Table below shows the design parameters used for MBR design calculation.

Table 30. Design Parameters

Parameter	Unit	Value
Design Flow rate	m^3/d	60444.8
SRT, $\theta_{process}$	days	19.67
Total tank volume	m^3	17712.1
O_2 demand	kg/h	633.7
Sludge wastage flow, Q_w	m^3/d	655.9
r_{int}		3.18
RAS(r_{mr})		4
Xm(typical value)	g/m^3	10000

In Chapter 5, the cost benefit analysis for both Kubota and Zenon membrane has been implemented and capital and operational costs calculated. Then both NPV which provides an indication for overall cost based on CAPEX and OPEX and plant lifetime and EQI which represent the potential pollution load based on the given methodology have been computed.

Results show that the amount of contingency built in to cope with changes in feed water flow has a large impact on NPV. Deviation from a constant daily flow increases NPV because it directly has impact on membrane area. Likewise, adding a buffer tank reduces NPV, since less membrane surface is required when average plant utilization increases. Membrane cost and lifetime is decisive in determining NPV: based on filterability and characteristics of material used in membranes an optimum value of 8 and 10 years have been considered for Zenon and Kubota membranes respectively. Operation at higher SRT increases the NPV, since the reduced costs for sludge treatment are offset by correspondingly higher aeration costs at higher MLSS level.

Table 31. EQI and NPV of Cost benefit analysis *for Zenon Membrane*

Parameter	Unit	Value
Design flow(ZENON)	m3/h	2,519
Effluent Quality Index (EQI)	kg PU/d	7478.044
TSS	mg/L	48.21386
COD	mg/L	166.5191
BOD	mg/L	76.6671
TKN	mg/L	20.91605
Total CAPEX	€	12597623
Total OPEX	€	556646
Net Present Value (NPV)	€	15987246

Table 32. EQI and NPV of Cost benefit analysis *for Kubota Membrane*

Parameter	Unit	Value
Design flow(Kubota)	m3/h	2,519
Effluent Quality Index (EQI)	kg PU/d	7478.044
TSS	mg/L	48.21386
COD	mg/L	166.5191
BOD	mg/L	76.6671
TKN	mg/L	20.91605
Total CAPEX	€	10494750
Total OPEX	€	556646
Net Present Value (NPV)	€	14165383

Both two MBR systems, Kubota and Zenon, demonstrated good membrane performance and both effluent quality comply with newly standards but there are some other factors in terms of cost, energy and maintenance affect the selection of best technology:

- Capital cost:

A cost analysis was performed to determine capital and operational costs of MBR Systems. Result in table 31 and 32 shows total capital cost for Kubota membrane is 17% less than the Zenon technology.

- Maintenance and cleaning:

From an operation and maintenance point of view, an MBR carries additional costs associated with the membrane system fouling and cleaning. Both systems appear to maintain reasonable fluxes by applying relaxation – intermittent physical cleaning attained simply by closing the permeate valve and allowing air to scour the membrane surface. Both Kubota and Zenon have also recently developed design modifications for increasing efficiency. In the case of Kubota, this is achieved by stacking the membrane modules (already employed by Mitsubishi Rayon in its MBRs). Comparing the two technologies shows the Kubota membrane requires less membrane cleaning during operation compare to Zenon membrane although we designed based on net flux of 12 and 10 LMH for Kubota and Zenon membranes. The Zenon membrane should be cleaned more frequently due to problems associated membrane fouling and foaming. The foaming is believed to have resulted from the membrane cleaning because chlorine was brought into direct contact with the MLSS.

- Energy consumption:

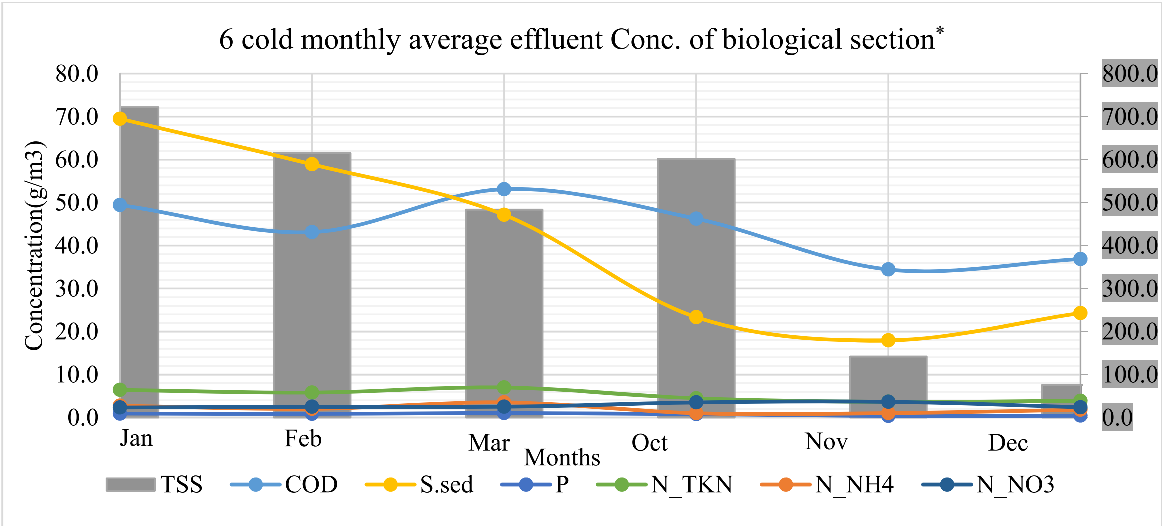
From the MBR plants investigated in many works, average unit energy consumption ranged from 1.43 kWh/m^3 to 4.24 kWh/m^3. energy consumption could reduce by implementing the new air scour strategies, or by operating their plant closer to design fluxes if the membrane performances allow. The main operating cost regarding energy consumption is membrane aeration. In designed MBR, the aeration cost for Kubota MBR is 5408.6 kW which is 20% less than the Zenon manufactures MBR.

All in all, Considering the above mentioned factors it's better to use Kubota membrane systems since it cost less and has less operational and fouling problem and also in terms of aeration cost it consume less aeration compare to Zenon manufacture system although Zenon recently has introduced intermittent aeration, which effectively halves the specific energy demand associated with aeration.

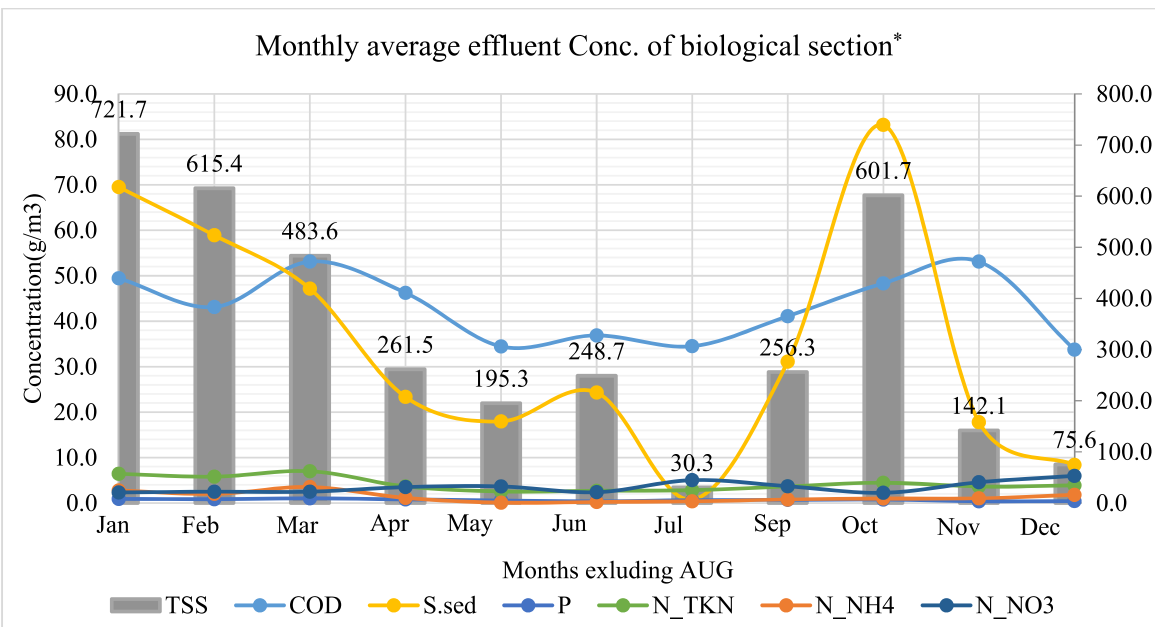
Recommendations for further research

Based on the conducted research described in this research cost and benefit aspect and the quality of water body considering optimum lifetime and an optimum value has been studied, but we can obtain a complete evaluation of two different scenarios for COMODEPUR by developing a mathematical model based on ASM1 model to simulate the behavior of the process, so that we can calibrate and validate the biological kinetic parameters to apply those parameters for an MBR biological process based on the reported technical literature.

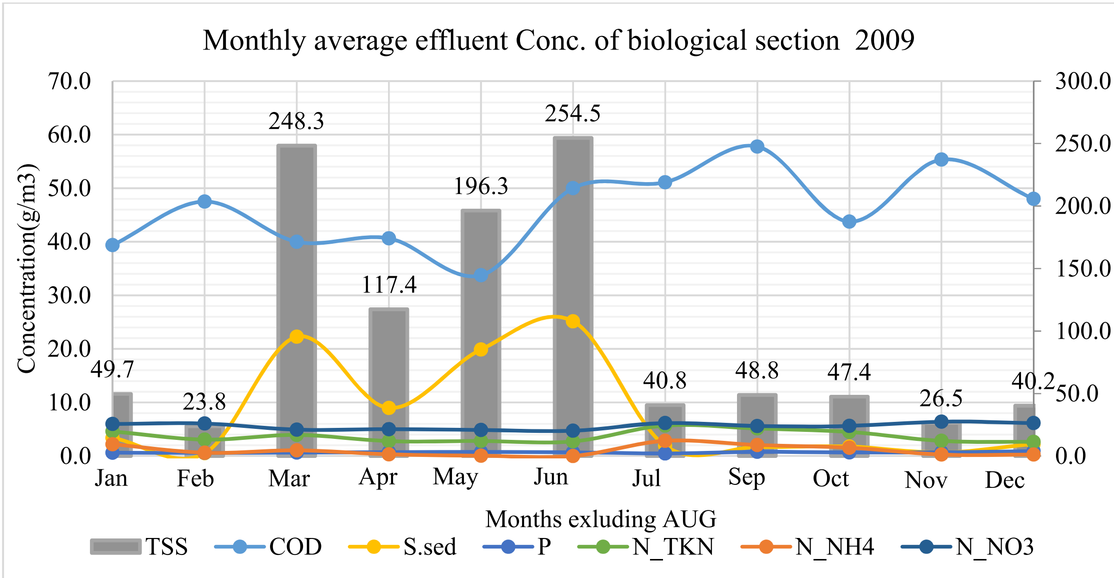
APPENDIX 1



* Comodepur WWTP Data, 2008



* Comodepur WWTP Data, 2008


Monthly average effluent Conc. of biological section 2009
Concentration(g/m3)
70.0
60.0
50.0
40.0
30.0
20.0
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0.0
300.0
250.0
200.0
150.0
100.0
50.0
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117.4
196.3
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Months exluding AUG
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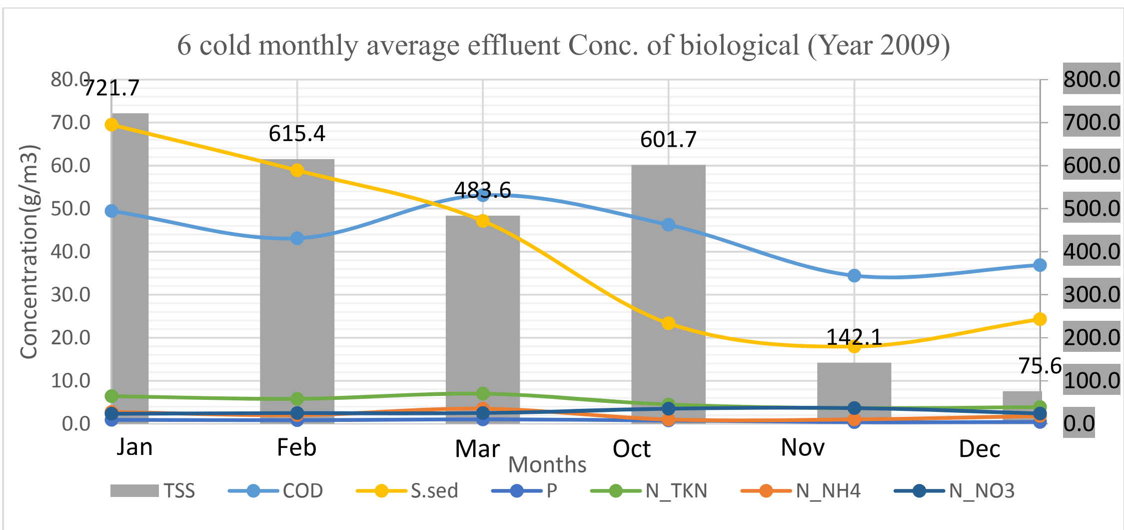
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Concentration(g/m3)
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40.0
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10.0
0.0
800.0
700.0
600.0
500.0
400.0
300.0
200.0
100.0
0.0
721.7
615.4
483.6
601.7
142.1
75.6
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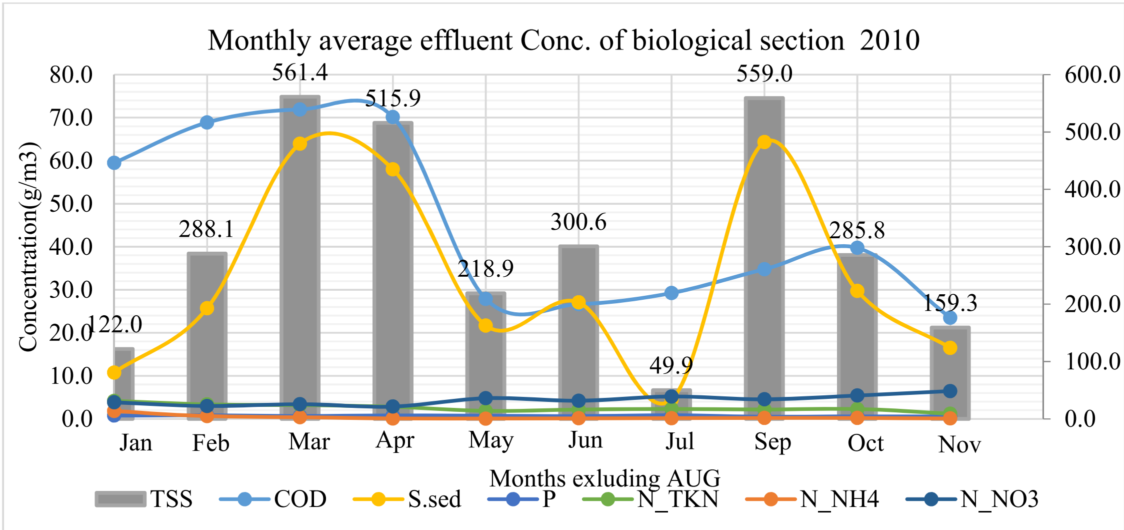
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Concentration(g/m3)
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10.0
0.0
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500.0
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300.0
200.0
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561.4
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218.9
300.6
49.9
559.0
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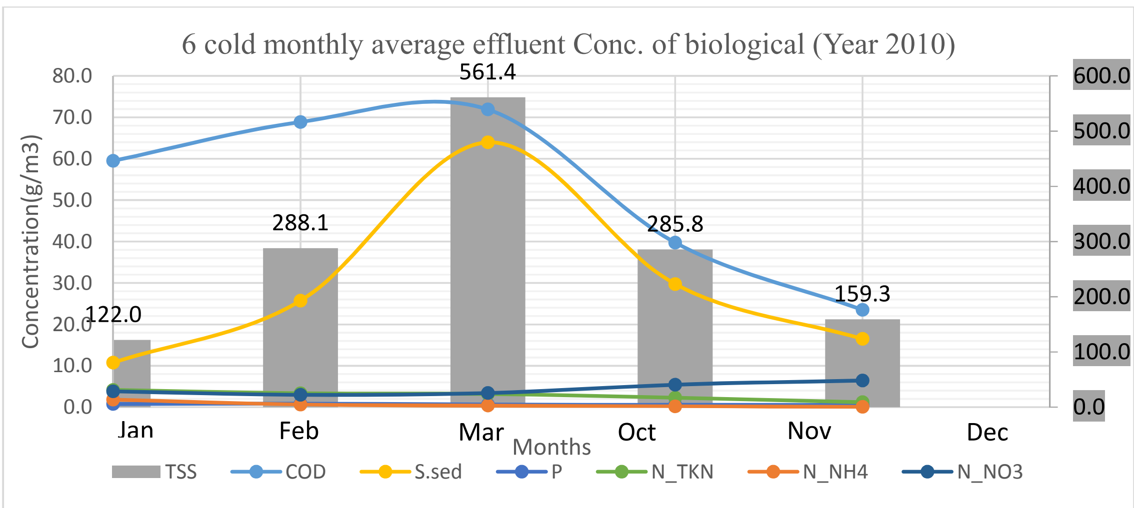
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Concentration(g/m3)
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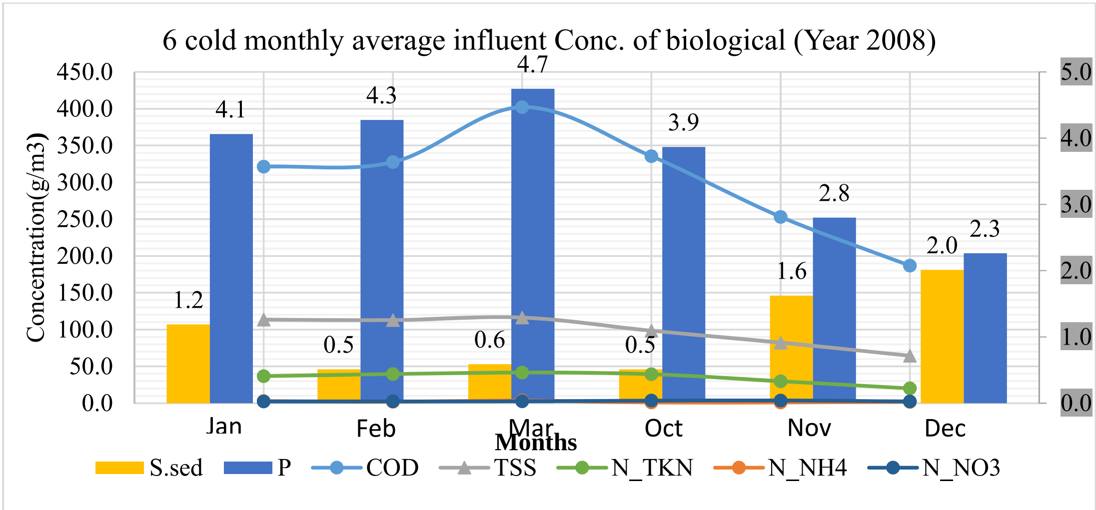
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Concentration(g/m3)
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250.0
200.0
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100.0
50.0
0.0
5.0
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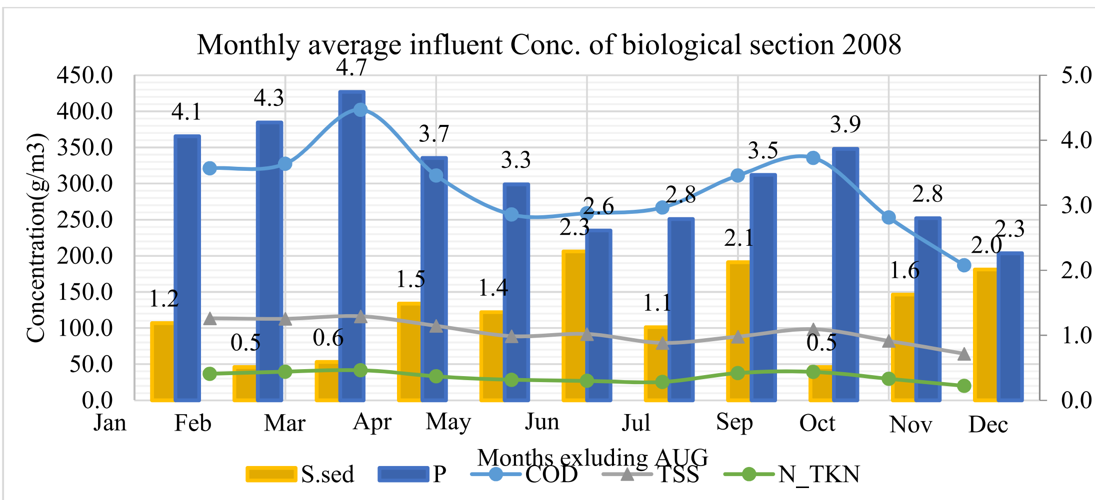
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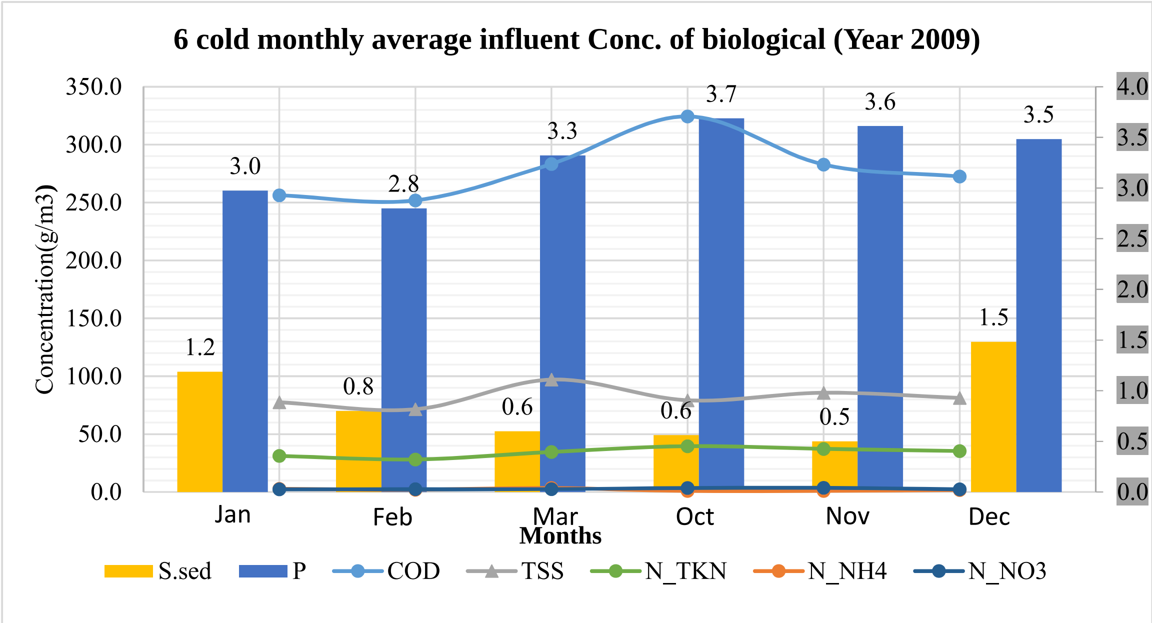
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Concentration(g/m3)
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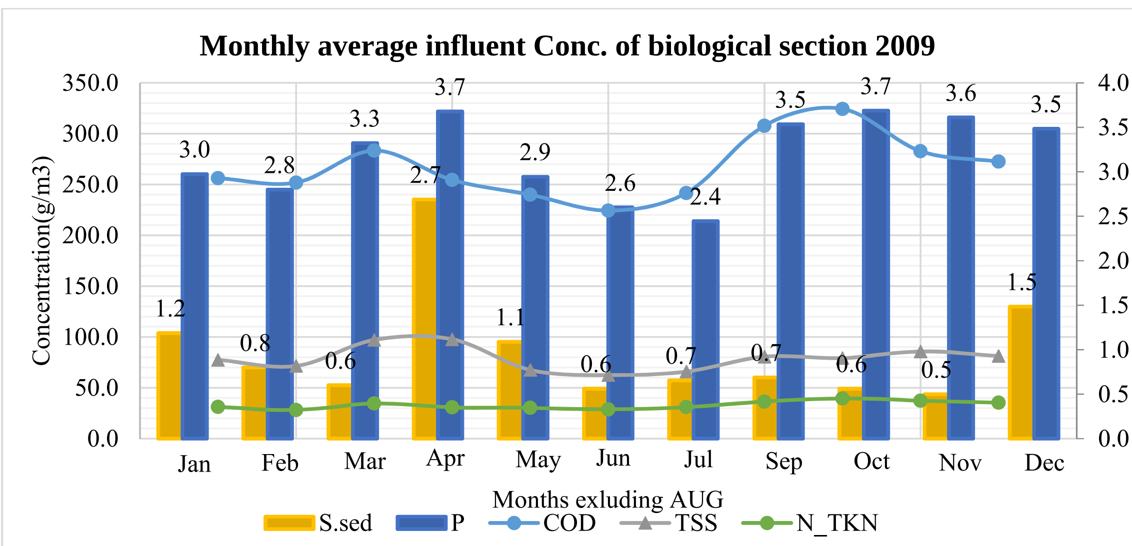
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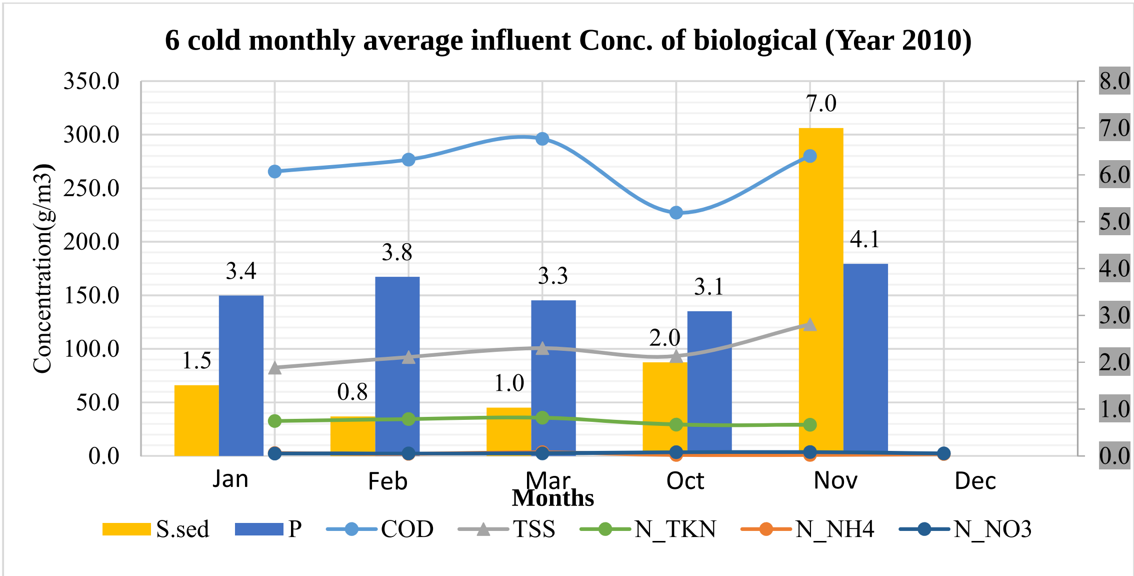
6 cold monthly average influent Conc. of biological (Year 2010)
Concentration(g/m3)
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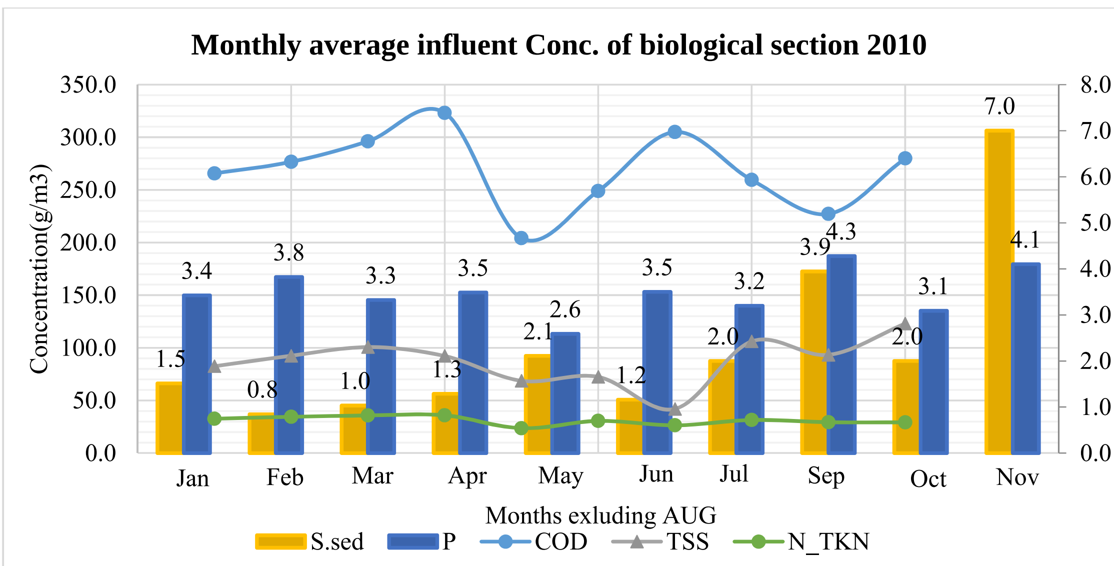
Monthly average influent Conc. of biological section 2010
Concentration(g/m3)
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APPENDIX 2

Name / Model	ZeeWeed (ZW500D)
Membrane material	PVDF, braided
Pore size, μm	0.04
Filament outside diameter, mm	1.9
Module dimensions,	(a) 1837 × 844 × 49
length × width × depth, mm	(b) 2198 × 844 × 49
Module effective membrane area, m^2	(a) 25.5 & 27.9
	(b) 31.6 & 34.4
Cassette dimensions,	(a) 1744 × 738 × 2208
length × width × depth, mm	(b) 1744 × 738 × 2568
	2122 × 1745 × 2590
Number of modules per cassette	(a) 16
	(b) 16, 48
Total membrane area per module, m^2	(a) 408, 446.4
	(b) 505.6, 550.4
	1516.8, 1651.2
Packing density, m^2 membrane area/m^3	
Internal module volume	
Clean water permeability, LMH/bar	(a) 0.42, 0.39
Recommended membrane aeration rate,	(b) 0.34, 0.31
Nm^3/h per m^2 membrane area	0.31, 0.29

Judd book, 2010

Membrane or Module Proprietary Name/model	**515 Panel RM/RW Module**
Membrane material	
Pore size, µm	
Panel dimensions, length x width x thickness, mm	1560 x 575 x 6, dual nozzle
Panel effective membrane area, m^2	1.45
Module dimensions, length x width x height, mm	2250-2930 x 575 x 2490, RM 2250-2930 x 575 x 4290, RW
Number of panels per unit	150-200, RM 300-400, RW
Total membrane area per unit, m^2	217-290, RM 435-580, RW
Clean water permeability, LMH/bar	
Maximum operating transmembrane pressure, bar	
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.42 (RM); 0.29 (RW)

Judd book 2010

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