DEVELOPMENT AND ASSESSMENT OF A MEMBRANE BIOREACTOR FOR WASTEWATER TREATMENT

PETER OGBEMUDIA OSIFO

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Dissertation submitted in compliance with the requirements of the Master's Degree in Technology in the Department of Biotechnology, Technikon Natal, Durban

Approved for final submission

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14-5-2001
Date

2001

Declaration

I hereby declare that the dissertation is my own work, unless stated to the contrary in the text, and that it has not been submitted in part, or in whole to any other Technikon/University.

[redacted]

P. O. Osifo

May, 2001

Dedication

THIS DISSERTATION IS DEDICATED TO MY PARENTS

ACKNOWLEDGEMENTS

I wish to express my sincere gratitude to the following people for their assistance in the completion of this dissertation.

I. Professor Hamanth Kasan and Mr Faizal Bux for understanding my situations and supporting me in making this work a reality.

II. Dr V L Pillay for his valuable input, guidance and encouragement.

III. My colleague in the research centre, for their various assistance and discussions.

IV. Ravi Govender, Rajin Kasavelu, Blaise Atkinson and Loshnee Nair for your support and assistance.

V. Professor C . Visvanathan for his advise and help especially during the WISA MTD workshop in Drakensberg.

VI. My good friends; Prosper Momubanghan,Sampson Osadorlor, Gavin Eweka, Solomon Obasogie, Sunny Ogunrunbi, and Rev. Frank and Philomena Ogagba for your moral and financial support.

VII. My wife for the patience and tolerance she showed through the years I took to complete this project.

VIII. My parents for their interest and encouragement.

IX. CWWR , Natal Technikon for financing this project.

ABSTRACT

A woven fibre membrane unit was developed locally and used for this study. The membrane unit was submerged in the aerobic tank of an anoxic-aerobic biological treatment process to treat dairy effluent. Another unit similar to the process above was also set up but with a gravity settling tank replacing the membrane unit. Kinetic studies were done on both processes and their results compared. Effluent characteristics of both processes were also compared. For the membrane bioreactor, the effectiveness of two backflush methods ie air and water, were tested. In this study it was discovered that air backflush could remove the clogging formed at external and internal pores of the membrane better than water backflush measured in terms of rate of pressure recovery and energy consumption. The effluent quality of the membrane bioreactor (MBR) in terms of COD, nitrogen, suspended solids and turbidity was excellent under all conditions tested and was better than conventional activated sludge system. The COD removed based on the average value in MBR was 96.02 % compared to 94 % conventional system. A mass balance of the processes showed that 79 % in the form of nitrogen fed into both the MBR and conventional systems was denitrified. The effluent suspended solids from the MBR was not determinable while that of conventional system was 2.09 mg/L. Maximum specific nitrification rate determined graphically was 3.0 d^{-1} and 2.10 d^{-1} in conventional and MBR respectively. The smaller value of nitrification rate in MBR could be attributed to large number of nitrifier volatile suspended solids in the system. The estimated true yield and decay rate coefficients of conventional system is 0.204 kgVSS/kgCOD and 0.013 d^{-1} respectively. The flux obtained from this study is between 22 and 92 $L/m^2.h$. However, the average flux value is 57 $L/m^2.h$ for the whole period of operation. Regular flux decline observed during operation was improved through regular backwash with air and chemical cleaning. The transmembrane pressure did not as whole increase more than 55 kPa due to frequent cleaning of the membrane surface.

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LIST OF ABBREVIATIONS

A	ammonia concentration
A_1, A_2	viral coefficients
A_{nx}. Eff.	effluent from anoxic reactor
ADUF	anaerobic digestion ultrafiltration
AS	activated sludge
ASP	activated sludge process
BEPR	biological excess phosphorus removal
BOD	biological oxygen demand
BOD_5	five days Biological Oxygen Demand
BOD_L	ultimate Biological Oxygen Demand
C_G	concentration of solute at the membrane surface
C_B	bulk solute concentration
COD	chemical oxygen demand
CWWR	Centre for Water and Wastewater Research
dc	concentration change
DO	dissolved oxygen
DWF	dry weather flow
DWAF	Department of Water Affairs and forestry
dx	finite change along x-direction
Eff	effluent
EDTA	ethylene diamine tetraacetic acid
F/M	food to microorganism ratio
f_d	residual fraction of the living biomass non-biodegradable
HRT	hydraulic retention time
Inf	influent
J	permeate flux
k_d	endogenous decay rate

K_N	half saturation constant for NO_3-N
K_S	half saturation constant for substrate as COD
K_{on}	half saturation constant for oxygen concentartion
MBR	membrane bioraector
MF	microfiltration
MLSS	mixed liquor suspended solids
MLE	Modified Ludsack Ettinger
MLVSS	mixed liquor volatile suspended solids (same as VSS)
MTD	Membrane Technical Division
MUCT	Modified University of Cape Town
MWCO	molecular weight cut off
N	nitrogen
N_1,N_2	nitrate concentration at location 1 and 2 of flow diagram
nm	not measurable
NF	nano filtration
NPV	net positive value
NVSS	nitrifiers volatile suspended solids
OA	oxygen absorption
OUR	oxygen uptake rate
P	phosphorus
P_{in}	pressure inside membrane module
P_{atm}.	atmospheric pressure (same as pressure outside membrane module)
Q	flow rate; Qo = influent flow rate, Qe = effluent flow rate, Qw = sludge wasting flow rate
r_d	endogenous decay rate
r_g	net rate of biomass growth
r_{su}	rate of substrate utilization
r_x	rate of biomass growth
RAS	return activated sludge
RBCOD	readily biodegradable COD
R_m	membrane permeateability constant
R_p	polarised layer resistance

R_{ds}	resistance offered by dissolved solids on membrane wall
R_{ss}	resistance offered by suspended solids on membrane wall
R_t	total resistance offered by dissolved and suspended on membrane wall
RWQO	receiving water quality objectives
S	substrate concentration measured as COD, S_o= influent substrate concentration, S_e = effluent concentration
S_b	biodegradable COD
S_{ti}	total COD, S_{te} = total effluent COD, S_{ti} = total influent COD
S_u	unbiodegradable COD, S_{ui} = unbiodegradable influent COD, S_{ue} = unbiodegradable effluent COD, S_{up} = unbiodegradable particulate COD
SBCOD	slowly biodegradable COD
SRT	solids retention time
SS	suspended solids
STD	standard deviation
t_{perm}	time for collecting permeate of volume Vperm
TKN	Total Kjeldahl Nitrogen
U	specific substrate utilization rate
U_N	specific nitrification rate
U_{dN}	specific denitrification rate
U_v	specific substrate utilisation rate based on the VSS
UCT	University of Cape Town
UF	ultrafiltration
V	volume; Vr = reactor volume
VSS	volatile suspended solids (same as MLVSS)
V_{perm}	volume of permeate collected
WAS	waste activated sludge
WISA	Water Institute of South Africa
WRC	Water Research Commission
WWTW	Waste Water Treatment Works
X, X_a	concentration of active biomass
X_d	concentration of inert endogenous residue
X_{io}	concentration of biological inert organic matter from the feed

X_{ii}	concentration of biological inert refractory inorganic matter from the feed
X_{io}^{o}	concentration of inert organic
X_{ii}^{o}	concentration of inert inorganic
X_v	biomass concentration in terms of VSS
Y	yield coefficient
θ	HRT
θ_c	SRT
θ_c^m	minimum solids retention time
∞	infinity
α_N	nitrate removal coefficient
σ	standard deviation
μ_m	maximum specific growth rate of biomass
ΔP	pressure drop across cake
$\Delta\Pi$	osmotic pressure difference

CHAPTER ONE

INTRODUCTION

The demand for higher pollution control and greater amount of useable water has created a need for more effective treatment processes for domestic and industrial water and waste water. For the past 20 to 30 years, researchers have embarked on a development program geared toward the utilization of membranes in complete sewage treatment. The conventional method of treatment has been modified with membrane units either submerged or coupled to the aeration tanks. The use of activated sludge systems followed by ultrafiltration (UF)/microfiltration (MF) membrane units have been investigated (Grethein,1978; Chaize and Huyard, 1991; Nell and Kafaar,1995). The insertion of membrane modules directly into bioreactor has also been investigated (Yamamoto *et al.*,1989; Chiemchaisri *et al.*, 1992; Visvanathan *et al.*, 1997).

The objectives of this project was to develop a membrane bioreactor with a locally made membrane unit for the treatment of dairy wastewater and compare the treatment efficiency of the membrane bioreactor to an activated sludge systems . To this effect a brief review is made on the environmental impact of dairy effluent and South African legislation on industrial effluent. The following section is concerned with the issues of legislation and regulation on industrial effluent.

1.1 EFFLUENT LEGISLATION

1.1.1 Current Legislation

Ever increasing numbers of industries over the past three decades has given rise to concern over different forms of chemical and physical pollution affecting water bodies. This led to more stringent measures through the amendment of Public Health Act (1919) and Water Act,(1956).
The act imposes certain control such as the:

a) the purification of effluent to a predetermined standard before discharge to the original water source;

b) the requirement of a permit to use certain quantities of water;

c) effluent purification becomes integral part of the industrial process;

d) the requirement of a permit for the erection or enlargement of water care works (a work used for the treatment of water intended for human consumption or food processing or treatment of sewage(DWAF,1986)).

The Act attempted to maintain a balance between the demand for water and deterioration in quality by imposing certain duties and exemptions upon the users of water for industrial purposes. Two of the more significant duties are, the duty to purify effluent and the duty to return the water and effluent to the point of origin. The duties play an important role in pollution control as the onus is impacted upon the user to provide evidence to the Department of Water and Forestry (DWAF) that requirements have been complied with, thereby promoting self regulation on the part of the industry. Table 1.1. highlights South African environmental legislation on waste water emissions and health and safety in the food and dairy industry.

Table 1.1. South Africa environmental laws affecting waste water emissions and health and safety in the industry (Gilfillan, 1997).

Act	Principal focus	Problems targeted
Water Act 1956	Protection of water environment and effluent standard	Water quality and water reuse
Environmental Conservation Act (1965)	Environmental protection	Protection of natural and social environment
Health Act (1977)	Control of pollution	Effect of pollutants on health
National Occupational Health and Safety Act	Ensuring adequate health and safety levels	Workers health and safety

1.1.2 Water Quality and Legislation Control

The majority of industrial effluents such as the dairy effluent, require treatment to make it suitable for reuse or to prevent pollution in the receiving water course. The DWAF manage the water resources of South Africa, and the following points highlight their water quality policy:

- o pollution must be reduced at source;
- o discharge effluent must meet the minimum standard required;
- o enforcement of stringent standards if the minimum effluent standard is insufficient to maintain the fitness of the water body; and
- o exemptions for effluent standards will be considered only as a last resort.

An industrialist requires a permit from the DWAF to use more than 150 m^3 of water per day. The application includes the nature and the volume of the effluent that will be generated and how this effluent will be purified and disposed. Before the DWAF allows waste to be disposed of into the water environment, it will require the industry involved to demonstrate that all possible efforts have been made to meet stipulated standards (Gilfillan, 1997).

Water quality is been managed on the basis of uniform effluent standards. All industrial effluent is obliged to comply with uniform standards such as the general and special standard. Table 1.2. gives examples of some of the standards. The standards have the potential to be updated to incorporate the newest pollution abatement technology which gives rise to this advantage for its simplicity and ease of easy to enforcement. The disadvantages of adhering to these standard are, it offers no incentive for industries to locate at the most environmentally advantageous site and it ignores the affected receiving water bodies.

Table 1.2 Some General and Special Standard for Effluent (DWAF, 1986)

Parameter	General standard	Special standard
Colour, odour, or taste	No detectable colour, odour or taste	No detectable colour, odour or taste
pH	5.5-9.5	5.5-7.5
Temperature (°C)	<35	<25
OA (mg/L)	<10	<5
COD (mg/L)	<75	<30
Conductivity	<75 mS/m above that of intake water	15% above that of intake water
Suspended solids (mg/L)	<25	<10
Soap oil and grease (mg/L)	<2.5	nil
Sodium (mg/L)	<90 above that of intake water	<50 above that of intake water
Nitrate as N (mg/L)	nil	1.5
Total Chromium, Cr (mg/L)	0.50	0.50
Phenolic compound as phenols (mg/L)	0.1	0.02
Lead (as Pb) mg/L	0.1	0.1

1.2 Treatment Method Adopted By Industry

Currently, most industrial effluents are treated at sewage treatment plants by a number of different processes. This treatment process results in an acceptable reduction in pollution load. Very few industries use on site treatment methods before discharging into the water course or sewage plant. In most cases, primary or primary-secondary treatment methods are used. Primary treatment entails the settling out of any solid particulate matter in settling tanks or basins. Periodically, the basins are drained and the sludge that has accumulated at the bottom is removed and buried/incinerated or treated in anaerobic digesters. Secondary treatment is accomplished through the oxidation of organic matter in the liquid effluent (effectively reducing the COD load) using resident micro-organisms as in the case of activated sludge system or a biofilm in the case of trickling filters. During trickling filter treatment, the effluent is trickled through a particulate bed containing an attached growth biofilm which oxidizes organic effluent as they filter through.

Most waste water treatment works employ activated sludge treatment processes. The activated sludge process is a much quicker as, it involves vigorous aeration of sewage, which results in floc formation (Droste, 1997). The flocs are allowed to settle and are then added to the fresh sewage through repeated recycles. At this stage the sludge is activated and is able to oxidize organic compounds very quickly.

The non-biodegrable organic pollutants and the nutrient; soluble nitrogen and orthophosphate, are removed using advanced treatment methods. Nutrient removal is also possible with the modified activated sludge process such as Phoredox, UCT, MUCT (WRC 1986).

1.3 Dairy Effluent Quality

Exit effluent properties promulgated by DWAF is shown in Table 1.2. A survey conducted by Sutton (1997), of industrial effluent produced in Kwa-Zulu Natal, shows that almost all the industries evaluated produced effluent that is of high COD concentration. Among the industries investigated are the dairy, textile, food and meat industries.

Water management in the dairy industry is well documented (Steffen and Kirsten, 1997) but effluent production and disposals remain a problematic issue for the dairy effluent. The dairy industry consumes approximately 4.5 million m^3 of water per annum in over 150 dairies in the Republic of South Africa (RSA). Between 75% and 95% of water intake is discharged as effluent (Steffen and Kirsten, 1989). A survey by Steffen and Kirsten (1989) showed that South African dairies apply either very basic or inefficient treatment methods. Dairy effluent disposal in South Africa usually results in one of two problems; firstly, high treatment levies are charged by local authorities for industrial effluents, and secondly, further pollution can be caused when untreated effluents are either discharged into the environment or used directly as irrigation water (Strydom *et al.*, 1995). The second problem of disposal is more prevalent at large dairies situated near dairy farms in rural areas, where access to adequate waste water treatment works is not available. The major constituent of dairy effluent is milk. Details of the effluent properties collected for this study is shown in Appendix 3 (Kasavelu, 1999). To enable the dairy industry to contribute to water conservation, an efficient and cost effective effluent treatment method has to be developed.

1.4 BIOLOGICAL WASTE TREATMENT

1.4.1 Aerobic Biological Treatment

Biological waste treatment is primarily used to remove dissolved and particulate organic matter in waste water. Some suspended organics will also be metabolized. Because of the natural flocculation and settling characteristics of the biomass, other suspended matter can be removed through sedimentation (WRC 1986).

The degradation of organic matter in water is a result of the presence of microorganism in the receiving bodies of water and is a natural occurring process. High organic loads in a waste water will upset the biocenosis of the receiving bodies of water and cause other undesirable effects. Biological treatment is engineered to accelerate natural decay processes and neutralise or stabilise the waste before it is finally discharged to receiving waters.

This aerobic biological treatment can be effected through different types of engineering design and configuration. The present biological treatment processes are conventional activated sludge, trickling filter, rotating biological contactor, sequential biological reactor, and MBR to mention a few. Some of these processes were modified or can be modified to suit a particular waste treatment in order to achieve a desire end product (Metcalf and Eddy, 1994).

1.4.2 Microorganisms In Aerobic Biological Treatment

Microorganisms in the waste water system promote all forms of reactions in the system, in fact they are the biological engine driving the process. Droste, (1997) defined biomass in a reactor as the sum total of all living organisms in the process. Other organic matter present, suspended or dissolved, may have associated with living biomass at some time but this matter is not part of the biological engine driving the process.

The microorganisms are used to convert the colloidal and the dissolved carbonaceous organic matter into various gases and cell tissue. The resulting cell tissue (biomass) can be removed from the treated waste water by gravity because the cell is slightly denser than that of water. The cell tissue produced from the organic matter must be removed from the solution. Their presence in the effluent will be measured as COD (Droste, 1997). If cell tissue is not removed, the only treatment done is that associated with the bacterial conversion of a portion of the organic matter originally present to various gaseous end products. The use of membrane technology to separate the cell tissue from the waste water is a recent development which have proved to be more efficient because the membrane pore size is smaller than the size of the cell particles (Visvanathan *et al.*, 1997).

1.4.3 Activated Sludge System

The activated sludge process was developed in the early 1900 by Ardern and Locket who used fill and draw reactors (a batch process) to successfully treat wastes in a short period of time (Urbain *et al.*, 1997). Continuous flow reactors were designed shortly thereafter and regularly used because of problems in controlling a number of batch reactors through fill-react-settle-draw cycles with variable influent flow rates.

For the treatment of municipal waste water an activated sludge process is normally used. Figure 1.1. shows schematic flow process of conventional activated sludge. In the aeration tanks conditions are created that stimulate the growth of various species of microorganisms, so that the pollutants are taken up and converted to cell material. The mixture of activated sludge and water is a suspension which subsequently is separated into cleaned waste water and thickened activated sludge.

The settling tank is determining or limiting factor for the operation of an activated sludge process (ASP) because the settling processes in the secondary treatment depends mainly on the biomass concentration in the aeration tank. To achieve a safe separation of activated sludge and clear waste water, the average biomass concentration in the activated sludge tank, measured as mixed liquor suspended solids (MLSS), should be 3 000-4 000 mg/L and should not exceed 5 000 mg/L (Engelhardt *et al.*, 1998). Over the years attempt have been to increase the working biomass. These development have brought improvement, but there is still plenty of scope for highly efficient systems to be developed.

1.5 MEMBRANE BIOREACTOR (MBR)

Membrane bioreactors are a combination of two processes; a biological degradation reactor and a membrane separation unit. Both must be combined as a single process not as a downstream incorporation as tertiary treatment unit.

Since the 1970's, pressure driving membrane separation processes microfiltration (MF) and ultrafiltration (UF) have been developed to replace clarifiers traditionally used in activated sludge treatment system (Fane *et al.*,1979). The modifications have generally been devised with a specific function in mind, but overall aims are to improve effluent quality and to also reduce construction and operating cost.

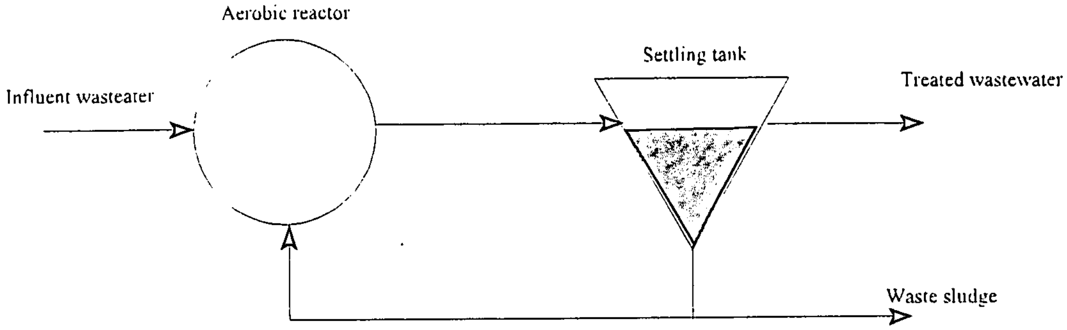


Figure 1.1 - Conventional activated sludge system

Figure 1.2. shows the two methods of operation most common for wastewater treatment. Separate stage system with cross flow membrane modules are separated from the aeration tank (Grethlein,1978; Nell and Kafaar,1995; Chaize and Huyard,1997). The feed flows parallel to the membrane surface at a high speed and pressures between 300 and 500 kPa. As a result, high specific permeate flows of approximately 100 to 300 $L/m^2.h$ are possible (Seyfried *et al.*, 1998). At the same time the specific energy consumption is fairly high, it lies between 3 and 7 kWh/m^3 permeate.

In the case of vacuum-driven membrane modules, pressure is not applied directly to the feed but to the permeate instead. The membrane modules are submerged in the aeration tank, usually above the aerators to remove the cake-layer with the help of rising air bubbles (Yamamoto *et al.*, 1989; Chiemchaisri *et al.*, 1992). Depending upon the construction of the modules, water or air backflush can reduce the cake layer at the surface of the membrane and to some degree the fouling. Both the specific energy consumption, approximately 0.2 to 0.5kWh/m^3 permeate, and the specific permeate flow, approximately 20 $L/m^2.h$ permeate, are one magnitude of order lower than those of cross flow modules. Therefore, a much larger membrane surface is needed (Seyfried *et al.*, 1998).

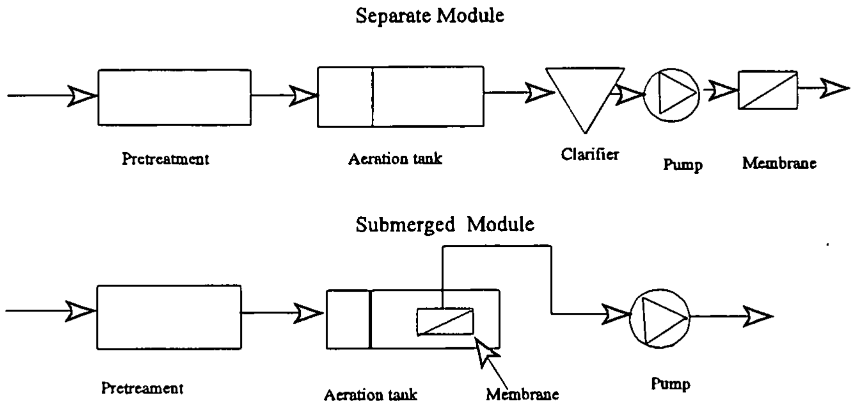


Figure 1.2 Flow chart for separate and submerged membrane modules (Seyfried *et at., 1998).*

Most researchers (Yamamoto, 2000; Fan *et al.*, 1996; De silva *et al.*, 1998; Visvanathan *et al.*, 1997; Chemchieriasri *et al.*, 1993) found that a combination of membrane modules and activated sludge system offer many benefits. The membrane bioreactor process allows total retention of biomass regardless of its properties (eg. size of flocs and age) and guarantees a high quality of treated water. This process enables a combination of high volumetric throughput and high mass loadings that ensure a better operation reliability and stability. Complete nitrification is possible as a result of long sludge age (Davies *et al.*, 1998). Sludge from the settling tank into effluent as a results of insufficient denitrification in the activated sludge are not possible with membrane bioreactor and therefore, sludge bulking are eliminated.

1.6 Process Economics

Research studies carried out by Davies and co-worker (1998) show that capital and operating cost of membrane bioreactors have not been competitive to that of conventional activated sludge systems. Table 1.3. shows the capital and operating cost estimates for membrane plants and conventional ASP at 1.0 and 16.58 M*l*/d dry weather flow (DWF) treatment capacities. Membrane modules accounted for 70% of the capital cost of the larger plant before land and civil cost were taken into consideration.

Whole life cost for each of the options were calculated. The unit treatment cost in each case was estimated from the net positive value (NPV) of the options and the average throughput rate. When land and civil cost were taken into consideration, unit treatment cost by membrane bioreactors started from 8.5 pence/m^3 for 1M*l*/d (DWF) plant and reduced to 6.8 pence/m^3 for a 16.58 M*l*/d (DWF) treatment plant based on full treatment up to 3 DWF maximum. The corresponding treatment cost by conventional technology was 6.1 p/m^3 for a 1M*l*/d DWF and 2.3 p/m^3 for 16.58 M*l*/d DWF. From the analysis of Davies *et al.*, 1998, it seems that unit treatment costs when using membranes is much higher than when using conventional system. There is optimum working capacity between 1.0 M*l*/d and 16.58 M*l*/d that will result in economic compromises with membrane systems, that is membrane is competitive for treatment capacities up to 3.5 M*l*/d DWF but only if the full treatment is reduced to 2 DWF. And also membrane could be made competitive for a treatment capacities up to 8 M*l*/d with full treatment to 3 DWF or 15 M*l*/d treatment capacity for a full treatment to 2 DWF maximum if the costs of membrane is reduced by 50%.

Table1.3 Comparison between the cost of membrane bioreactor and convention ASP (Davies *et al.*, 1998)

PARAMETERS	MEMBRANE BIOREACTOR		CONVENTIONAL ASP	
			A*	B*
DWF, Ml/d	1	16.5	1	16.5
Max. Throughput,Ml/d	3	22.5	3	22.5
Average Throughput, Ml/d	1.4	49.7	1.4	49.7
Band Screen	59,304	156,085	59,304	156,085
ASP	-	-	519,000	1,669,562
Membrane System	997,858	15,740,131	-	-
Land	7,000	136,429	21,900	325,573
Civil	126,000	956,095	380,000	1,787,783
TOTAL CAPEX	**1,190,162**	**16,990,103**	**980,204**	**3,942,259**
Manpower	20,000	50,000	20,000	50,000
Chemicals	4,200	5,000	4,200	5,000
Power	16,176	275,000	13,000	148,000
Membrane Replacement	61,420	1,018,350	-	-
Others	11,173	160,994	19,000	62,460
ANNUAL RUNNING COST	**112,969**	**1,509,344**	**56,200**	**264,730**
Unit treatment cost p/m^3	8.5	6.8	6.1	2.3

A* = Cost of ASP without primary settling tank

B* = Cost of ASP with primary and final settling tank

Note: the estimated costs in Table 1.3 are in GBP: £ : P. exchange rate GBP : South Africa Rand = 11.47 (GBP = Great Britain Pounds)

1.7 Project Objectives and Approach

The objective of this project was to develop and assess the performance of a MBR using locally made membrane unit and compare the performance of the membrane bioreactor unit to a conventional activated sludge systems on their basis of treatment efficiency. In order to achieve this objectives the project was approached in the following manner by:

- investigating the optimal backflushing method with water and air for the membrane system

- investigating the performance of MBR and conventional activated sludge systems for COD and nitrogen removal using dairy effluent as a feed. This also involves determining the kinetic coefficients using the data collected from both the MBR and AS systems.

1.8 DISSERTATION STRUCTURE

Chapter 1: Introduction, industrial effluent control in SA through legislation, and project objective.

Chapter 2: Literature review; Overseas and local application of membrane bioreactor for waste water treatment. Review of operating parameters governing activated sludge system and principle and application of woven fibre microfiltration membrane.

Chapter 3: Model development for steady state and non-steady state system.

Chapter 4: Experimental

Chapter 5: Results, discussion and recommendations

CHAPTER TWO

LITERATURE REVIEW

2.1 MBR SYSTEM DEVELOPMENT

2.1.1 Brief history

Most of the MBR plants existing presently are mainly overseas. The majority of these plants are situated in Japan, the United Kingdom (UK) and the United States (U S). The development of MBR started in the US. The early pioneers was Dorr-Oliver then located in Stanford, Connecticut, who began work on membrane bioreactor technology in the mid-1960s, but did not develop system commercially in North America. Rather, the first commercial application of a membrane bioreactor system was in the early 1970s by Thetford Systems of Ann Arbor, Michigan (Hadi and Pierre, 1999).

Between 1974-1980, the company installed about 27 plants treating and recycling flush water from small commercial facilities using cross flow techniques. Since the early 1980's Thetford's process has started to be used in larger facilities such as buildings, shopping centers, industrial parks, sport facilities and other places where flush water recycling is required to reduce waste water discharge to sewer.

In 1989, Kubota corporation installed the first submerged membrane pilot system, with the first commercial plant following in August,1991 (Steve, 1998). Since then, over 150 plants have been installed treating a wide range of effluents including sewage, sludge liquors, industrial and food processing waste water and grey water for re-use.

2.1.2 Overseas Experience

The use of submerged membrane modules was anticipated by Yamamoto *et al.*, (1989) who applied hollow fibers for direct membrane separation to wastewater treatment in activated sludge. A microfiltration membrane made of polyethylene, with 0.1 μm pore size and 0.3 m^3 effective surface area was used. The water was filtered from the supplied wastewater through suction. Various operational modes were applied to investigate stable flux conditions as well as organic and nitrogen removal. The system was run for 120 days at volumetric organic loading of 1.5 $kg/m^3.d$ through intermittent suction (10 minute "on" and 10 minute "off") with low pressure at 13 kPa. The flux obtained from this work was 5.04 $L/m^2.h$ and more than 95 % COD removal was achieved. Through intermittent aeration mode 60 % of nitrogen was removed though denitrification.

Chaize and Huyard, (1991), investigated solid retention time (SRT) and hydraulic retention time (HRT) on membrane bioreactor potentialities. Two modelling approaches were tested (maintenance phenomena, single sludge simulation) with unsuccessful prediction. Organic stabilization and nitrogen removal was successfully accomplished by Chiemchairi *et al.*, (1992) using two hollow fibre membrane modules (Mitshibishi Rayon co., Ltd) of 0.3m^2 surface area each. The membrane module were submerged in a bid reactor tank of about 60.5 litres. Highly turbulent conditions was created within the separation zone in conjunction with jet aeration inside the membrane modules, this is done to reduce sludge accumulation on the membrane surface during filtration. The average effluent COD obtained from this study was 20.8mg/l and the average flux was of 0.2m^3/m^2d under continuous suction.

Visvanathan *et al.*, (1997) applied air back flushing to dislodging the particles formed at the surface of the membrane during their studies using 0.2μm hollow fibre UF membranes for low strength domestic wastewater. The air back flushing was used in controlling the concentration polarization that causes flux decline during the separation process. From the studies it was found that 15min filtration and 15min air back flushing gave the best results for both flux stability and net cumulative permeate volume. This cyclic operation could not completely remove the clogging however, 90%

of both COD and TKN reduction was achieved.

Recent developments in membrane bioreactor technology enable Urbain *et al.*, (1997), and De Silva *et al.*, (1998), to successfully carry out population dynamic of various organisms in the system. Through the use of an advance mathematical model, a followed up of the work of Furumai and Rittmann, (1992), Urbain *et al.*, (1997), and De Silva *el al.*,(1998) were able to quantify the relationships among heterotrophic bacteria and autotrophic bacteria and their chemical constituent in MBR. The model was based on the following assumptions; no biomass in the effluent, partial removal of biomass associated products by the membrane and the dissolved oxygen is approximately 4 mg/L in the aerobic and 0.5 mg/L in the anoxic periods. The model explains the key trends in the cyclic operation i.e. ammonia consumption during aerobic periods and the steady rise during anoxic periods; nitrate is produced only during aerobic periods, but declines in the anoxic periods; and the soluble COD mainly consists of biomass associated product and is relatively constant.

Pressure driven crossflow filtration was used by Fane *et al.*, (1979); Grethlein, (1978); Chaize and Huyard, (1991) coupled with a completely mixed reactor to separate treated wastewater from activated sludge. The work of Fane and others (1979) was based on modeling the MBR through the concepts of particle rejection at the surface of the membrane. Grethlein, (1978) studied the treatment of domestic wastewater using septic tank-membrane system. The BOD and nitrate reduction was 90 and 75 % respectively. Re-circulated flow was used to mix the content of the septic tank and provide control of the concentration polarization at the surface of the membrane.

Over the years various pilot plant studies have been carried out with membrane bioreactors. At Kingston Treatment Works (UK), three membrane treatment units were retrofitted into an old activated sludge package plant. The pilot plant was installed by Wessex Water in conjunction with Welsh and Southwest water. The capacity of the plant is 100 m^3/day treating screen crude sewage effluent at Kingston Treatment Works. The plant according to Steve, (1998) is now in its four years of operation with very high effluent quality. The sludge production is less than 0.4 kg ds/kg BOD.

Porlock is a picturesque village located on the edge of Exmoor National Park in UK, with a population of approximately 4 000 in the catchment area. Before the installation of the membrane unit, effluent from the treatment works was discharge in to a nearby sea. The requirement was thus for a high quality compact treatment process that could blend in with the surroundings. The capacity of the pilot plant installed is 1 900 m^3/d. The plant contains 24 Kubota membrane units in four aeration compartments, the treated effluent is removed under gravity. The efficiency of removal for the following parameters; biological oxygen demand (BOD),suspended solids (SS) and faecal coliforms from the pilot study are respectively 98.1, 99.5 and 99.9 percent (Steve, 1998).

2.1.3 South African Experience

In South Africa (S.A) little have been developed in the use of membrane bioreactor processes for wastewater treatment as compares to the development of conventional methods. The increasing use of membrane in recent time for wastewater treatment in SA was as a results of innovation into polymer science material at the University of Stellenbosch (Hurndal *et al.*, 1998) and the development of membrane modules by Bintech (Pty) Ltd and Mabraatech (Pty) Ltd.

Pilot-plant Studies

In 1996 an UF unit was coupled to a anaerobic digestion unit by Nell and Kafaar to treat industrial wastewater. The UF membrane modules used for the laboratory scale study were locally made by Bintech, Ltd, Paar. Two digestion units were used; anaerobic digestion ultrafiltration (ADUF) unit 1 and 2. The membrane area used for this studies was 0.22m^2 with molecular weight cut off (MWCO) of 40,000. The membrane material was made of polyether sulphone and the module outside diameter was 9mm. The average space loading into ADUF units was 2.5 kgCOD/m^3.d. Overall system performance of the two units were poor despite 98 % COD removal. The loading rate into the ADUF units was between 11 and 15 kgCOD/m^3.d over a day HRT. These values shows the poor performance of the units as compared to other units operated in the past. Both ADUF unit 1 and 2 suffered- high flux decline. Cleaning with warm water was able to improve the life of the membrane by 10% temporarily. An average flux decline rates of 3.4 and 2.9 L/m^2.d was recorded during the entire test period.

Strohwarld and Ross, (1992) conducted an investigation on brewery waste using a membrane bioreactor (UF module coupled with activated sludge). Urea was added to the industrial waste to correct nitrogen deficiency. COD reductions of between 96 and 99 % were achieved at a space loading rate of 15 kgCOD/m^3.d. The COD of the final effluent (UF permeate) was general below 100mg/L. Fouling of membrane was not experienced and membrane flux remained stable. In this study the membrane flux was significantly influenced by linear velocity, sludge concentration and metabolic condition in the digester. Flux was optimized by maintaining the linear velocity at a point where the pressure drop equaled the maximum allowable operating pressure. Sludge concentration was simultaneously reduced to a point where the maximum sludge loading rate was not exceeded.

Up flow anaerobic sludge bed digesters were used in the Western Cape to treat fruit wastewater. Due to over loading and sludge loss in the effluent during peak periods, an alternate solution was necessary.

Strohwald, (1993) carried out a laboratory investigation into the possible application of ADUF process to solve this problem. A 100 liters polyethylene reactor (active sludge volume 50 liters) was combined with a low cost ultrafiltration unit. The membrane unit was tubular (9mm o.d), with total filtration area of 0.4 m^2. The molecular weight cut off (MWCO) of the membrane is 40000 Da (2 × 20 tube configuration) and the material is made of polyethersulphone. The UF permeate was recycled back to the digester, the excess being wasted. The concentrated biomass was returned to the digester after passing through a heat exchanger.

The results obtained after 121 days of continuous operation showed that a mean space loading rate of 1.46 kgCOD/m^3.d could be obtained at COD reduction rate in excess of 96% and mean HRT of 2.3 days. The loading rate in excess of 1.5 kg COD/m^3.d resulted in deterioration of the COD reduction potential causing a sharp increase in the COD content of the UF permeate. An average flux from this work was 14.8 L/m^2-h without the need for chemical cleaning. The membrane flux was reported to decrease rapidly at MLSS values greater than 20 g/L. Maintaining the MLSS below this value seems to be one of the prerequisites for limiting flux loss. Linear flow velocity was shown to affect the flux directly. A flux increase of 20 to 30 L/m^2.h can be expected for every meter per second velocity increase.

Brand *et al.*, 1999 investigated the use of membrane bioreactor for industrial effluent treatment. Effluent from acid extraction plant in Secunda contains mainly organic acid i.e acetic and propionic acid. This was treated in activated sludge system coupled with UF membrane units (2 units). The UF has a MWCO of 4 0000 and a surface area of 3 m^2. The results of this studies shows that COD and SS were reduced to an average value of 93% and 2mg/L.

Full-scale Application

One of the full scale application of MBR experienced in the recent time in SA is the ADUF application used to process maize effluent in Meyerton Mill of Messrs African Products (Ross *et al.*, 1992). This process was used to treat effluent produced during the manufacture of conflour, natural and modified starch, glucose syrups, maize and dextrins.

The results after 15 months of operation, have illustrated the merit of the process for the production of a colloid-free effluent at a mean COD removal efficiency of 97%. A mean space loading of 3 kgCOD/m^3.d was maintained during very large variations in the feed load. Permeate flux was given as between 8 and 37 L/m^2.h with a linear velocity of 1.6 m/s and inlet operation pressure of 450 kPa. Periodic cleaning of the polyethersulphone membrane with EDTA was carried out after 13 months of operation.

2.2 Process Configurations

Most of studies carried out since 1989 with membrane reactor mainly involved organic removal. Even though nitrogen has to be removed it is through intermittent aeration (Yamamoto *et al.*, 1989;Chiemchaisri *et al.*, 1992; Fan *et al.*, 1996; De Silva *et al.*, 1998). Nitrogen removal generally involves nitrification and denitrification processes. Two basic configuration have been developed, the Ludsack-Ettinger configuration and Wuhrmann configuration (WRC 1984., Orhon and Artan, 1994) as shown in figure 2.1.a. In the Ludsack-Ettinger configuration shown in figure 2.1.a, two reactors are connected in series partially separated without intermediate settling. Sewage is fed to the first anoxic reactor to supply organic matter used as the energy and carbon source in denitrification. The denitrified effluent is aerated in the second reactor which secures nitrification. The partial separation between the reactors induces an interchange between the anoxic and aerated zones, and the nitrified mixed liquor of the aerated zone entering the anoxic volume become subjected to denitrification. This process was later modified by Barnard (Orhon and Artan., 1994), who completely separated the anoxic and aerobic reactors, recycling the settled activated sludge to the anoxic reactor and providing an additional internal recycle from the aerobic to the anoxic zone. This modification is known as Modified Ludsack-Ettinger (MLE) process. The advantages of this process is that, process control is simple, smaller anoxic volume is required, and denitrification rate is higher due to direct introduction of sewage containing high degradable organic waste into the anoxic. The only disadvantage is that complete denitrification is not possible.

In Wuhrmann configuration, two reactors are connected in series, the first aerobic and the second anoxic. Its denitrification potential relies upon the electron acceptor demand of the endogenous biomass in the anoxic zone. Nitrification takes place in the first reactor and the nitrified effluent is transferred from the first reactor to the second (anoxic) reactor through normal flow for denitrification at a relatively slow rate dictated by endogenous decay (Furumai and Rittmann,1992). The influent waste water may be completely nitrified due to sufficient long sludge age. Here internal recycle is not required but sludge recycle is possible (see figure 2.1. b).

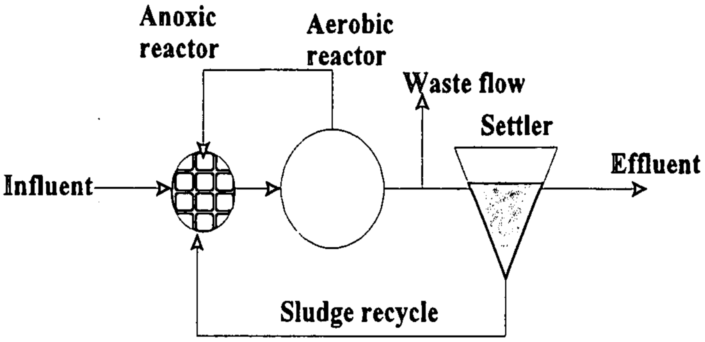


Figure 2.1a: Modified Ludzack-Ettinger nitrification-denitrification activated sludge process.

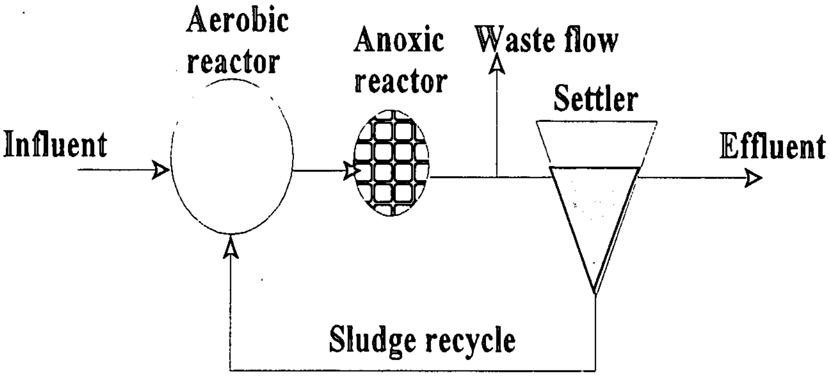


Figure 2.1b: Wuhrmann nitrification-denitrification activated sludge process.

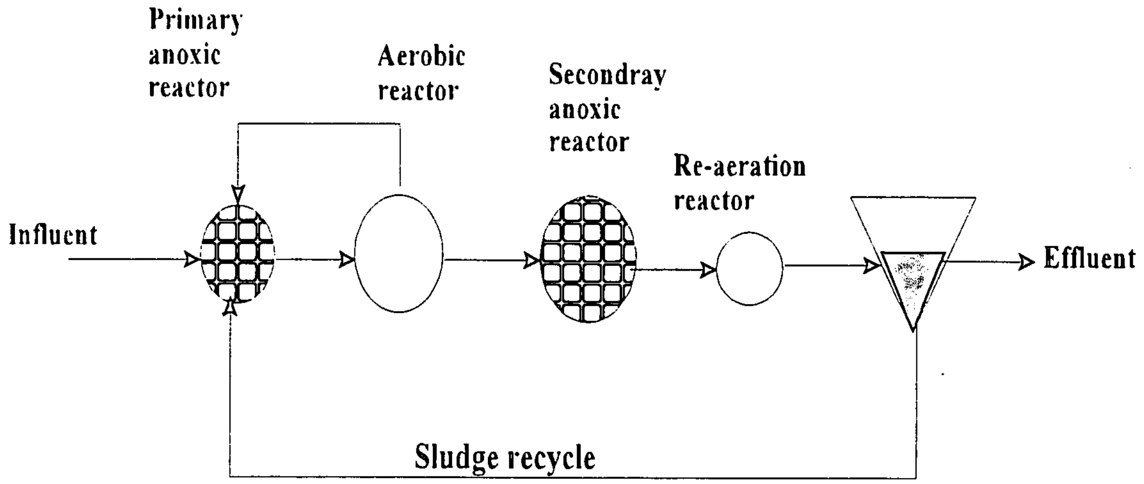


Figure 2.1c The Bardenpho process for nitrogen removal

2.2.1 Bardenpho Process

The four stage Bardenpho process was developed in order to achieve a good denitrification potential (WRC 1984; Orhon and Artan,1994). This process is a combination of MLE and Wuhrmann processes. As shown in figure 2.1.c, the process consists of primary anoxic reactor, aerobic reactor, secondary anoxic reactor and re-aeration reactor. The nitrified effluent from the first aerobic reactor is recycle internally to the primary anoxic reactor for denitrification. The nitrate that escape denitrification in the anoxic reactor is discharge to the secondary anoxic reactor for denitrification to produce a relatively nitrate free effluent. The re-aeration reactor was incorporated in order to strip nitrogen bubbles generated in the secondary anoxic reactor attached to the sludge floc. Ammonia release from the sludge in the second anoxic reactor is also nitrified in the last aerobic zone. A modification of Bardenpho process (addition of fifth stage) is used for combined nitrogen and phosphorus removal (Orhon and Artan,1994).

2.3 Reactor analysis

2.3.1 Aerobic reactor

In aerobic reactor carbon oxidation and nitrification takes place. Two groups of microorganism are said to be responsible for these reactions; they are heterotrophs and autotrophs (Ekama and Wentzel, 1999., Atkinson, 1999). The heterotrophs grow and oxidize carbonaceous organic matter in both the aerobic and the anoxic reactors; in the aerobic reactor it uses the molecular oxygen as the electron acceptor while in the anoxic reactor nitrate is used as electron acceptor. The growth mechanism of heterotrophs in an activated sludge system is characterized by the following stoichiometric equation

Oxidation and synthesis :

$$COHN + O_2 + Nutrients \rightarrow CO_2 + NH_3 + C_5H_7NO_2 + \text{other end products} \quad \ldots\ldots\ldots\ldots\ldots 2.1$$

Where COHN denotes organic matter and $C_5H_7NO_2$ biomass.

Endogenous respiration

$$C_5H_7NO_2 + 5O_2 \rightarrow 5CO_2 + H_2O + NH_3 + \text{Energy} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots 2.2$$

Equation 2.2. represent endogenous mass loss. This is referred to as endogenous metabolism, strictly speaking endogenous metabolism describes a strategy which is adopted by microorganisms when deprived of an external source of nutrient and a preferred term is organism decay.

The autotrophs carries out its reactions through nitrification. Nitrification may be described as the oxidization of reduced nitrogen compounds such as proteins (Connel, 1998). The essential feature of nitrification which is of engineering importance is that it is carried out by strictly autotrophic bacteria. Two microbia species are responsible in order to achieve nitrification: Nitrosomonas and Nitrobacter.

Nitrosomonas catalyze oxidation of ammonia to nitrite using molecular oxygen, see equation 2.3. whereas Nitrobacter further oxidize nitrite to nitrate using oxygen derived from the water molecule as shown in equation 2.4.

Nitrosomonas : $NH_3 + CO_2 + O_2 = \text{Cell} + NO_2^-$ 2.3

Nitrobacter : $NO_2^- + CO_2 + O_2 = \text{Cell} + NO_3^-$ 2.4

In contrast to heterotrophs, nitrifiers use carbon dioxide (inorganic carbon) rather than organic carbon for new cells synthesis. The conversion of carbon dioxide to organic cell tissue is a reductive process that requires a net input of energy. Autotrophic organisms must therefore spend more of their energy for synthesis than do heterotrophs, resulting in general lower growth rates among the autotrophs that is while sludge age of 10 or higher is required to achieve good nitrification (Dold *et al.*, 1980).

2.3.2 Anoxic reactor

In the anoxic reactor, organic substrate is oxidized using nitrate as terminal electron acceptor as shown in equation 2.5. below. The nitrified mixed liquor in the aerobic reactor is recycled internally to the anoxic reactor where they are denitrified. Microorganism responsible for this type of reactions are the heterotrophic bacteria (Argaman and Miller, 1979) hence a facultative bacteria since it can also survive aerobic and anaerobic conditions.

The facultative bacteria that use nitrate require a readily biodegradable source of carbon, this can either be supplied by the carbon in the wastewater or the carbon remaining in the bacteria cell tissue the use of methanol.

$$2\,NO_3^- + 10e^- + 12H^+ \rightarrow N_2 + 6H_2O \quad \text{..2.5}$$

2.3.3 Phosphorus Removal

Recent development in the use of biological treatment processes for the removal of phosphorus can be found in the work of Ekama and Wenzel,1999; Orhon and Artan 1994; Wentzel *et al.,* 1992. Phosphorus in wastewater is mainly the orthophosphate (PO_4^{3-}), polyphosphate ($P_2O_7^{4-}$) and the organically bonded phosphorus. The PO_4^{3-} and the organically bonded phosphorus are responsible for about 70% of municipal influent phosphorus (WRC 1986). Microorganisms need phosphorus for normal cell growth and energy transport, and as a result 10-30% of the influent phosphorus is removed during secondary biological treatment process. Biological treatment method has been used to stress or stimulate the growth of microorganisms so that they will take up more phosphorus than that required for normal cell synthesis (Atkinson,1999). However, many full scale plants have been built to achieve biological excess phosphorus removal (BEPR). The process plant configuration was such that an anaerobic reactor was included in the four stage Bardenpho process having observed that biological phosphorus removal was possible when an anaerobic reactor was followed by aerobic reactor (Ekama *et al* 1986). In most cases, the phosphorus removal attainable through biological excess phosphorus removal (BEPR) is limited, the concept of chemical phosphorus removal still persists as a supplement to BEPR or, in many cases, the sole method of phosphorus removal.

2.4 PROCESS CONTROL PARAMETERS

The parameter of fundamental importance in the design and control of activated sludge process is the sludge age. The other control parameter is the loading rate which is represented in of volumetric loading, organic loading, F/M ratio and floc loading. Noran (1991) defined loading rate as a rate at which settled sewage is applied to an activated sludge reactor. The magnitude of the loading rate will influence the biological behavior of sludge in many ways including the COD removal rate, setteability, and in the overall the efficiency of the plant.

Other principal factors reported by most worker (Dold *et al.*, 1980) used in process control are (1) maintaining the dissolved oxygen level in the aeration tank, (2) regulating the amount of recycled activated sludge (RAS) and (3) controlling the waste activated sludge (WAS). The mixed liquor suspended solids concentration (MLSS) can also be use as a controlling parameter. The recycled activated sludge is important in maintaining MLSS concentration (the "M" in the F/M ratio) and WAS is important in controlling θ_c. Oxygen uptake rate (OUR) is receiving serious attention as a means of monitoring and controlling activated sludge system (Boyle *et al.*, 1989; Wentzel *et al.*, 1995).

2.4.1 Volumetric loading

This parameter determines the amount of time which the sewage will undergo aeration in the reactor and is define as:

$$Retention\text{-}time(d)=\frac{Reactor\text{-}volume(m^3)}{Total\text{-}daily\text{-}flow(m^3/d)}$$

The above expression is known as the hydraulic retention time (HRT). The HRT of activated sludge lies within the ranges of 4-8 hrs (Metcalf and Eddy, 1994) and in the absence of sludge recycle it has the same value as the sludge age.

2.4.2 Organic loading

The organic loading is a measure of the amount of COD which is applied per unit volume of aeration tank capacity and it is express as:

$$Organic\text{-}Loading\text{-}(kgCOD/m^3\text{-}d) = \frac{Influent\text{-}flow\text{-}rate\text{-}(m^3/d) \times Influent\text{-}COD\text{-}(kg/m^3)}{Reactor\text{-}volume(m^3)}$$

The diurnal mass loading or the change in COD loading with the time of the day has a significant effect on the biological system. This variation has effect on the effluent quality, sludge settling properties and the efficiency of the plant performance.

In order to have an accurate values of mass loading in terms of COD it is advisable that a composite sample comprising 24 hourly samples of equal volume should be obtain (Metcalf and Eddy, 1994). The daily mass loading rates (COD or BOD) using hourly data can be found from the following expression

Daily mass loading, kg/d = $\sum^{24}_{i=1}$(concentration, mg/L)(flow rate, Mgal/h)(96)

2.4.3 Process loading Factor (F/M Ratio)

The food to microorganism ratio is one of the most important parameters used in controlling activated sludge systems. It is linearly related to sludge age and it is only this form of loading which the operator has in controlling the process. The F:M ratio also called the process loading factor,U, is represented as $(So - Se) / X\theta$, where θ is the hydraulic retention time (HRT). The magnitude of the loading rate (F:M ratio) will influence the biological behavior of the sludge which in many ways includes COD removal rate, sludge setteability and the degree of nitrification (Noran, 1991). The usefulness of this parameter on the effect of net specific growth of microorganism will be dealt with later. Substrate utilization rate as a function of the F:M ratio and process efficiency is related according to the equation:

$$U=(\frac{F/M}{100})E$$

where E is the process efficiency.

2.4.4 Sludge age (θ_c)

As mentioned above sludge age and F/M ratio are the two most important parameter in design and control of activated sludge systems. The sludge age, often called solid retention time (SRT) is an approximation to the mean solids residence time in the reactor system. It is expressed as:

$$Sludge-age(\theta_c)=\frac{Total-solids-in-the-reactor(kg)}{Total-solids-wasted-daily(kg/d)}(d)$$

Through experience in operating of conventional activated sludge reactors, it is found that θ_c should usually be between 3-14 days (Droste,1997) in order to produce a biological floc which can be handled easily. It has be reported that for $\theta_c < 3$ days, the biomass is not dense enough to settle easily, producing bulking sludge (Brenner, 1990). For $\theta_c > 14$ days, the floc particles are too small to settle rapidly and the fraction of the living cells in the biomass is low. Good sludge settling properties are essential for an efficient gravity settling operation and stable activated sludge process (Noran, 1991). Since the sludge age largely governs how well floc will settle, a value is chosen based upon experience and the type of sludge generated. Sludge age can be controlled either by the sludge wasting rate from the bottom of clarifier or by direct abstraction from the reactor.

The relationship between the sludge age (θ_c) and the applied loading rate (Specific utilization rate, U) is given in equation 3.5. section 3.2.1. The substrate utilization rate can be evaluated by determining the difference between the influent and the effluent COD. Judging from the expression of the specific utilization above, the evaluation of the active biomass cannot be measure directly this makes the use of U an impractical control parameter in activated sludge process. In WRC (1986) model, the importance of sludge age over other control parameters was stressed. With the use of θ_c it is not necessary to determine the amount of active sludge mass in the system nor the need to evaluate the amount of food utilized. The use of θ_c is simply based on the fact that, to control the growth rate of microorganism and hence their degree of waste stabilization, a specific percentage of the cell mass in the system must be waste each day. In nitrification environment the sludge age is usually high, normally exceeding 10 days dueto the slow growth rate of nitrifiers.

2.4.5 Oxygen uptake rate (OUR)

The use of oxygen uptake rate is receiving recognition as a means of monitoring and controlling activated sludge system. OUR is based on the rate at which microorganisms in activated sludge systems uses oxygen as they consume substrate. It is therefore taken as a measure of biological activity in the system. High OUR's indicate high biological activity and vice versa. A series of batch tests have been conducted (Wentzel *et al.*, 1995a; Ubisi *et al.*, 1997) on the use of OUR in estimating heterotrophs active biomass in the influent wastewater. Referring to the work of Wentzel *et al.*,(1995), OUR is measured by taking a sample of mixed liquor, saturated with dissolved oxygen (DO), and using DO probe to measure the rate of decrease of DO with time, the value obtain is normally reported as mg O_2/L.min or mgO_2/L.h.

2.4.6 MLSS Concentration

During operation of activated sludge systems,there is constant wash out of sludge from the reactor to the clarifier. In order to maintained a constant concentration of MLSS in the reactor, settled sludge in the clarifier is returned to the entrance of the reactor. RAS rate is determined by material balance around the reactor or settling tank depending on the desired MLSS in the system. In membrane bioreactors no settling tank is required and in most cases there is no sludge wasting from the reactor. High MLSS concentration of up to 45 gMLSS/L has been reported (Yamamoto *et al.,* 1989). The significance of this is the low F/M ratio experienced and plant size can also be reduced due to high MLSS in the system.

According to the recent findings by Yamamoto (2000), when sludge concentration exceeds a certain limit, permeate flux rapidly declines due to dramatic rise in viscosity of the sludge mixture. Limits for the filtration of activated sludge are 30,000-40,000 mg/L. Most practical applications are operated at less than 20,000 mg/L. A guideline of 10,000-20,000 mg/L is recommended for submerge membrane separation activated sludge systems for on -site domestic wastewater treatment (Yamamoto, 2000). Up till now, no proper define correlation between permeate flux and sludge concentration has been reported.

2.4.7 Recycle rates

When considering nutrient or nitrogen removal in system incorporating anoxic-aerobic zones or anaerobic-anoxic-aerobic zones both sludge and mixed liquor are recycled. The purpose of sludge recycling is to maintain a desire level of MLSS concentration while internal mixed liquor recycling is required for denitrification or for nutrient removal depending on the desire treatment method. For this type of study which involves nitrogen removal, the nitrified mixed liquor is internally recycled to the anoxic zone for denitrification. The use of the internal recycle rate as a control parameter based on concentration of nitrate in the system can be found in the work of Lilley *et al.,(*1997).

2.4.8 Temperature effects on the process

The effects of temperature on biological reaction rate constants is very important in assessing the overall efficiency of a biological treatment process. There have been a series of investigation in the past on how this parameter affects various processes. From the WRC (1986) report it was deduced that for every 6 °C drop in temperature the maximum specific growth rate value will halve which implies that the minimum sludge age will double. It was also reported that temperature sensitivity of half saturation constant has no effect on the minimum sludge age but it does affect the efficiency of nitrification. This was also noted by Argaman and Brenner (1986).

Apart from the temperature influence on metabolic activities of the organism in the process, it also has a profound effect on gas transfer rates. Activated sludge process is a mixture of suspended solids required certain minimum level of dissolves oxygen, and the solubility of gases like oxygen is function of temperature (Metcalf and Eddy, 1994).

2.4.8 Effect of pH variation

In nitrification and denitrification processes the optimum pH for bacteria growth has been found to lies between 6.5-7.5. (Dold *et al.*, 1980). During nitrification, the autotrophs use CO_2 as their source of energy for cell growth. This is derived in form of carbonate from the wastewater. The utilization of this bicarbonate is associated with a reduction in the alkalinity. If this alkalinity is not controlled or supply by the addition of hydroxide then there will be a drop in pH. Fortunately enough, in the anoxic zone, the organism utilizes protons in the reduction of nitrate then. Thus mixed liquor tends to become alkaline compared with the acidity produced during nitrification. The balance of pH is almost the same in the system with transfer of low alkaline sludge from aerobic to anoxic and high alkaline sludge from anoxic to aerobic.

2.4.9 Nutrient requirements

The major nutrient requirements are nitrogen and phosphorus. The amount of nutrients required for biological processes or activated sludge depend on the amount of biomass formed and removed from the process. The composition of biomass represented as molecular formula by Orhon and Artan (1994) is approximately $C_5H_7O_2N$. Phosphorus is not included in this formula but there is approximately 0.1 mole percent of phosphorus per mole of biomass. Table 2.1. represents a reported data describing the composition of aerobic biomass in terms of major constituents.

Table2.1 Composition of aerobic biomass (Fraser, 1975)

Element	% Weight
Carbon	49.0
Hydrogen	6.0
Oxygen	27.0
Nitrogen	11.0
phosphorus	2.5
Other	4.5
	100

In domestic wastewater, an excess of N_2 and P is required to satisfy the metabolic requirements of the bacteria in the secondary treatment. However, in industrial wastes, it may be necessary to add nutrients to the wastewater to sustain the biological reactions. The rule of thumb that applies to activated sludge process as given by Droste, (1997) is that the ratio of influent degradable matter expressed on an ultimate BOD or COD basis to N_2 and P should be COD:N:P = 100:5:1 on a mass basis.

2.5 Membrane Filtration

Swart *et al.*, (1996) defined membrane filtration as the retention of suspended and dissolved solids, of certain size, by a semipermeable membrane. The membrane separation process may be driven by pressure or an electrical potential. Membrane separation is characterized by the types of separation carried out which may depend on the pores size of the particular membrane. Different separation processes include microfiltration (MF), ultrafiltration (UF) and reverse osmosis (RO). Electrodialysis (ED) processes are based on the transport of ions through membranes as a result of an electrical driving force.

2.5.1 Membrane Materials and Properties

The membrane fabric to be used in this study is similar to that used by Pillay (1991) and Swart (1993). The membrane is within the microfiltration range with pore size between 1-5 μm as shown figure 2.2. The fabric is woven to form a longitudinal array of tubes which is referred to as a tube curtain. The curtain construction is such that continuous and isolated tubes are formed from a single homogeneous cloth, as an array of side-by-side tubes in the warp direction. When pressurized the collapsible array swells to form rigid cylindrical tubes.

A single tube of required length can be cut from the tube curtain. Sealant is normally use to seal the joint along the seam of the fabric. The sealant should be free of air bubbles that would otherwise form a pinhole when set. The formation of pinhole leads to poor permeate quality. For more details about properties and construction of woven fibre membrane refer to Swart (1993). For the purpose of this work the construction is detailed in section 4.1.1.

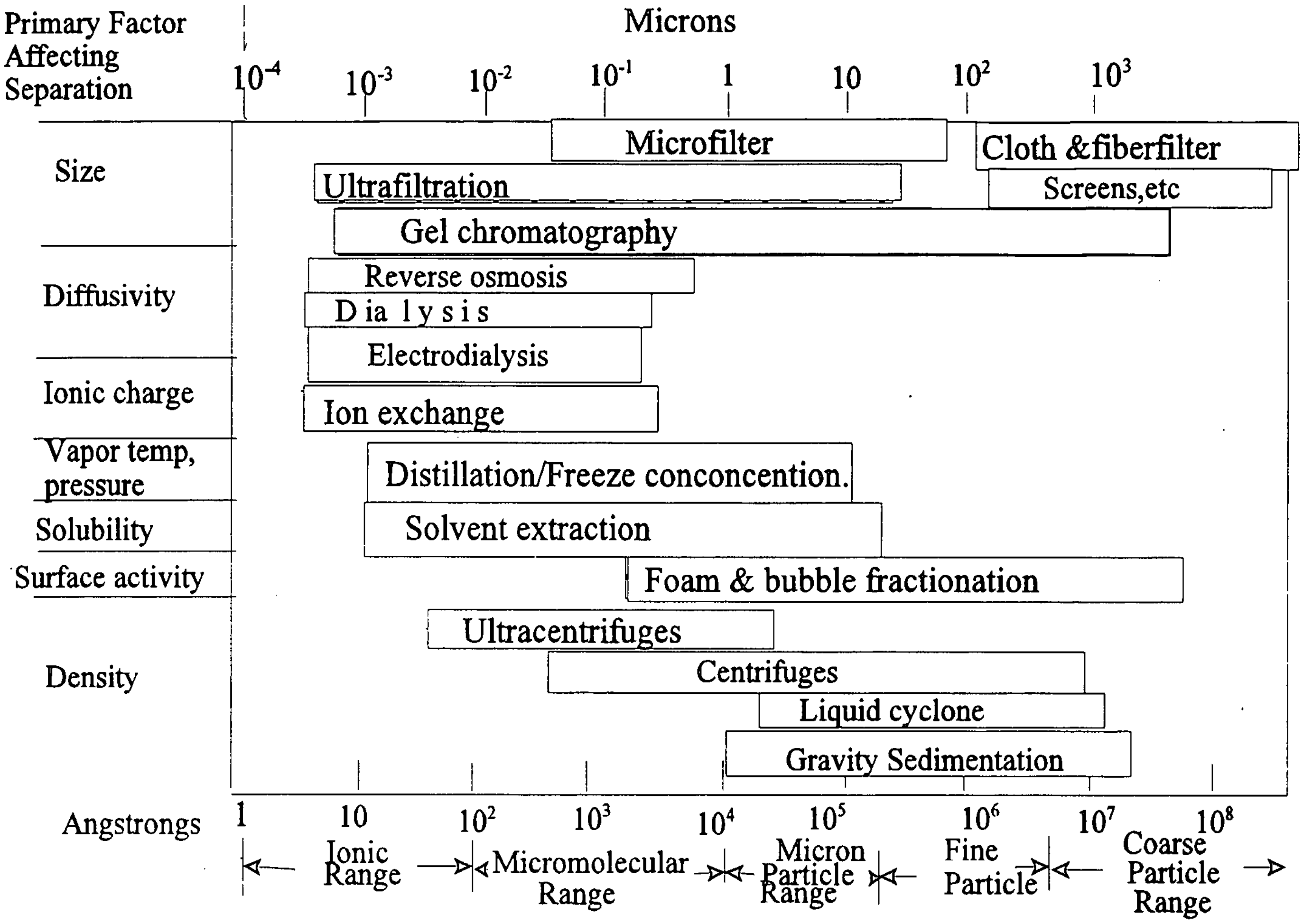


Figure 2.2 **Useful range of separation processes, showing the range of particle or molecular size covered by each process and the primary factor governing each separation process (Cheryan., 1986)**

2.5.2 Principles of Operation

In microfiltration technique one distinguishes between two operating procedures:

- dynamic operation of "cross-flow filtration" and
- static operation of "dead-end filtration".

2.5.2.1 Cross flow filtration

In cross flow filtration a flow (the so called feed) is pump across the membrane surface at pressures between 300 and 500 kPa. The feed has to control and limit the development of the coating on the membrane surface, and thus represents an important design feature. The feed behavior can be describe as that similar to the developed boundary layers whenever a fluid flow past a solid surface (whether the surface is porous or not).

In UF and MF the feed solution is brought to the membrane surface by convective transport and because the water is forced through the membrane at a much higher rate than the solute molecules, a build-up of solute species is observed at the solution-membrane interface. This is shown schematically in Figure 2.3. This solute build-up is known as concentration polarization and is chiefly responsible for the marked deviation in flux compared to pure water flux.

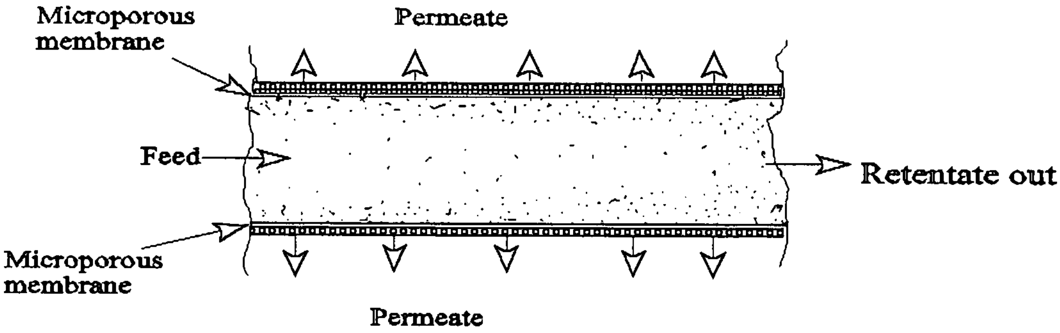


Figure 2.3 Schematic presentation of cross-flow filtration.

The solids deposited at the surface of membrane form a concentration boundary layer and is a function of the water removed and the gross fluid mechanics of the system. The importance of the hydrodynamic condition of the solute (macromolecules) streams expresses itself through the shear force that is exerted at the membrane-solution interface which have been modelled by Pillay (1991). The higher this shear force, the easier it is for the solute molecules concentrated near the wall to diffuse back into the bulk solution hence, in activated sludge tanks a high cross flow velocity is needed to promote depolarization, the wall shear play a role in scouring the suspended matter near the membrane surface and thus reducing fouling (Blatt,1971). These principle of operations are applicable to all the ADUF systems describe above (section 2.1.1. & 2.1.2.).

2.5.2.2 Dead-end filtration

Static operation or dead-end operation of membrane systems has the reject valve outlet closed. In the case of submerged membrane the permeate is sucked through the membrane and a negative pressure of 0.2-1.0 bar is built up. The retained particles settle on the membrane surface. In order to prevent break down of the permeate flow, the coating is removed from the membrane surface. Cleaning is done with water or air backflush.

Steve, 1998, described the phenomenon of upward flow of mixed liquor across the submerged membrane surface as similar to cross flow filtration. The energy required for aeration of the mixed liquor generate an upward cross-flow over the membrane as shown in Figure 2.4.(a), thereby keeping the fouling at the membrane surface to minimum. Cross-flow filtration mode can be contrasted to standard dead-end filtration mode as given in Figure2.4.(b), where the cake thickness increases continuously with time. This results in permeate flux that decreases continuously and rapidly. Usually the fluxes obtained with cross-flow mode are significantly greater than that obtained from dead-end mode. However, it has been found that for low turbidity feed streams, fluxes obtained in cross-flow and dead-end may not differ substantially (Pillay 1991., Swart 1993). In these instance, it is preferable to operate in the dead-end mode, since the pumping requirements, energy consumption etc. are considerably reduced.

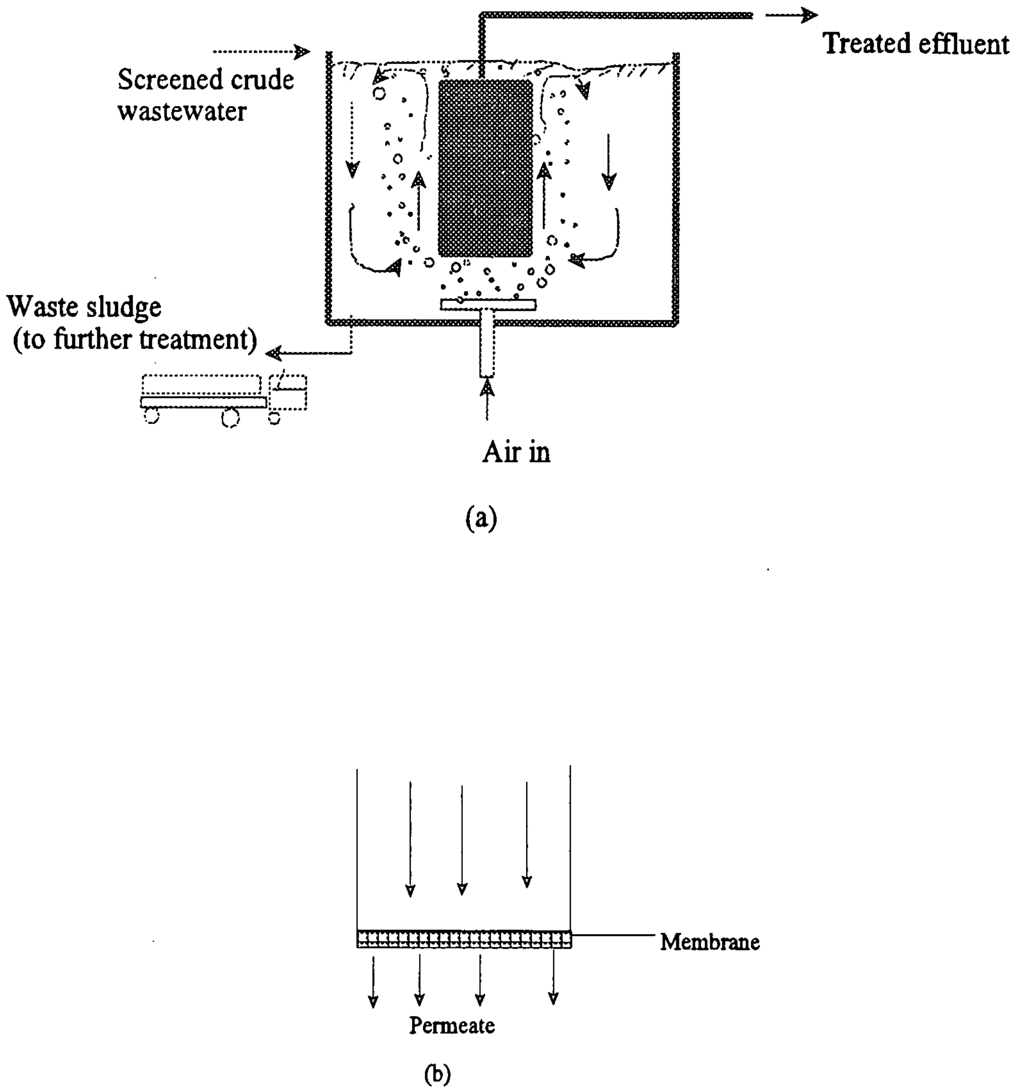


Figure 2.4 Schematic shows (a) principle of operation of submerged MBR in wastewater treatment process, (b) dead-end filtration

2.5.3 CONCENTRATION POLARIZATION AND FOULING

During microfiltration processes, flux decline is generally observed with time. When the transmembrane pressure is increased, the flux will also increase linearly, but there comes a point when the flux becomes independent of the transmembrane pressure. This deviation from linearity is observed at high pressure regardless of other operating conditions due to consolidation of the gel-polarized layer (Meien and Ronaldo, 1994; Wijmans *et al.*, 1984). With pure solvent e.g. water, the flux produced is much higher than for the suspended solution as shown in Figure 2.5. Recent findings by Yamamoto (2000) have shown that there is no guarantee that there will be linear relationship between operating pressure and permeate flux. An initial increase in pressure sometimes causes a decline in the permeate flux.

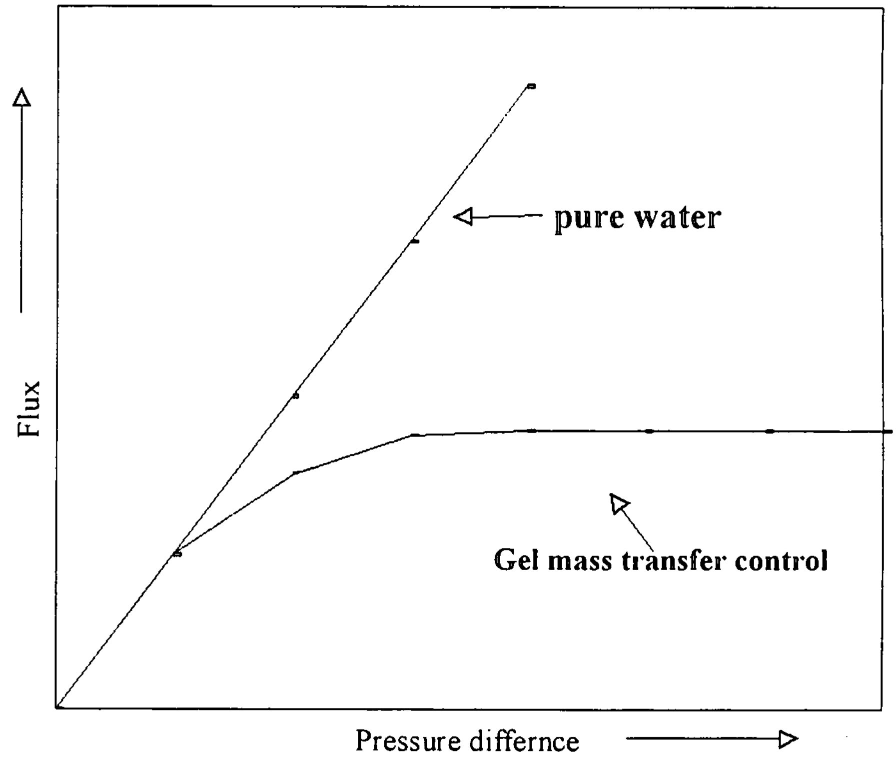


Figure 2.5: Dependence of flux on operating pressure (Swart et at., 1996)

Several reasons are thought to be responsible for flux decline by most researchers (Fane *et al.*,1979; Cheryan.,1986). These are: (1) Membrane compaction and deterioration; this was considered unlikely by Fane et al., (1979). The reasons given was because of low transmembrane pressure used (5 to 200 kPa) and the mechanical stability of the membrane hardly permit compaction at that range of pressure. (2) Membrane fouling or clogging by macro molecules; this phenomenon is known to be significant in a solution containing a spectrum of macromolecules as in the case of activated sludge mixed liquor.(3) Changes in gel layer which increases its resistance to flow; this is due to concentration polarization resulting in a localized increase in solute concentration or higher local osmotic pressure at the membrane surface.

From Figure 2.6. the initial rapid flux decline with time is caused by concentration polarization due to the formation of gel layer and the process can be reversed by changing the operating conditions. The decline in flux with time reaches a steady state at a point. At this point the gel layer remains constant. The long term flux decline that preceded the steady-state which remained almost constant throughout the time period is due to cake deposit at the surface of the membrane (Pillay ,1991; Swart *et al.*,1996).

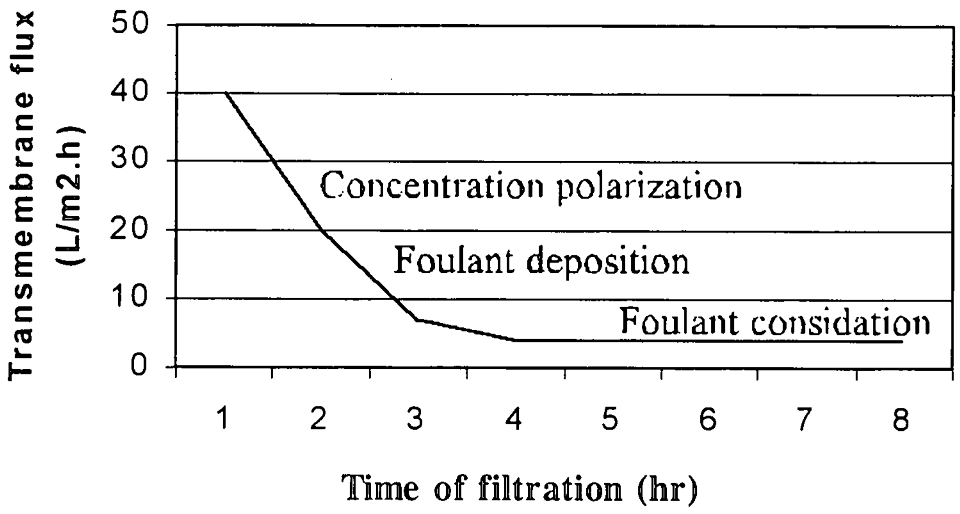


Figure 2.6: Long term Flux decline

The phenomenon occurring at the membrane surface during UF/MF processes is also depicted schematically in Figure 2.7. It explains the factors that could contribute to concentration polarization and fouling. During MF of a solution of macromolecules, there is a radial migration of the solute by convective transport toward the membrane surface. Solute build up at the wall will cause a concentration gradient within the boundary layer with the formation of gel layer. This causes a back transport of solute into the bulk solution due to the molecular diffusion effect (Cho *et al.*, 1999; Pillay, 1991). At some point, steady-state is reach, as explained earlier, whereby the rate of convective transport of the solute into the gel layer at the transmembrane surface is equal to the rate of diffusion of the solute back into the bulk solution. The solution concentration in the gel layer reaches a maximum, that is no more solute molecules can be accommodated due to the close packed nature of the solute molecules. This concentration, if high enough, could cause solutes to precipitate (crystallize) and foul the membrane.

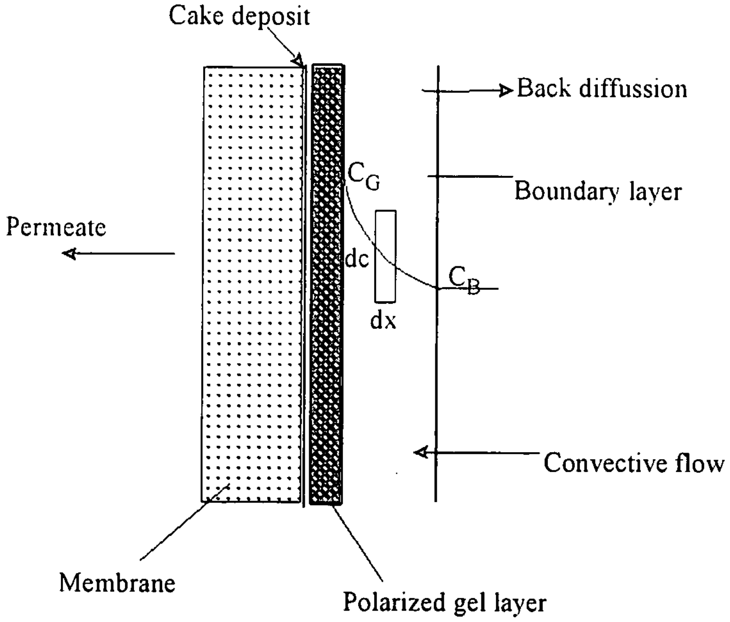


Figure 2.7 Schematic of concentration polarization during UF of macromolecules solutes, showing the building up of gel polarise layer and associate boundary layer.

One well known approach of avoiding solute precipitation is to improve the scouring effect by increasing the tangential flow velocity, but this method has a drawback of increased energy consumption. As stated by Yamamoto (2000), critical flux concept is another approach to control fouling. The critical flux is ideally defined as the flux below which fouling takes place. It is also important to reduce compaction of the cake layer. Low pressure filtration and intermittent filtration, has been found to be the effective means of achieving this (Visvanathan *et al.*, 1997).

2.5.4 Mathematical Models of Fouling

The operating parameters that explained the behaviour of long term flux decline with time in MF and UF are not easily assessable. Several models with varying levels of complexity have been proposed in the recent years (Pillay, 1991; Wijmans *et al.*,1984). For design purposes, a highly sophisticated model, which accounts for every detail of long term flux decline with time of the process, is hardly justified as a results of too many assumptions. The model should, however, be based on mechanistic principles so that it can be extrapolated, and be capable of predicting performance with reasonable accuracy. In fact the problem has been an inability to precisely model the fact occurring near the membrane wall. Several attempts have been made by some researchers (Fane *et al.*, 1979; Swart *et al.*, 1995) to model concentration polarization and fouling at the membrane surface in activated sludge. Activated sludge is a complex mixture of dissolved and suspended solids. The dissolved solids are composed of high molecular weight fatty acids, proteins polysaccharides and other biodegradable products as well as unreacted substrate and inorganic substances. The suspended solids are composed of bacterial cells and insoluble matter, so under these conditions, the membrane surface is vulnerable to polarization and fouling effects which make the modeling of the system a difficult task.

However, a number of MF/UF processes have been mathematically modelled to predict the effects of concentration polarization and fouling. In these studies mixed liquor suspended solids (MLSS) concentration were used to correlate flux effects. The reasons are, in activated sludge MLSS is the dominant solid is readily measured and predictable from activated sludge kinetic models (Fane *et al.*,1979)

2.5.4.1 GEL POLARIZATION MODEL

This model explained that transmembrane flux is independent of applied pressure. The model shows how the flux varies with bulk concentration and the concentration of gel at the membrane surface.

$$J = K_g \ln C_G / C_B \qquad \text{.......(2.6)}$$

Where J is the permeate flux in units of $L/m^2 h$, K_g is the mass transfer coefficient, having the same units as the flux, C_G is the concentration at the membrane surface in mg/L, C_B is the bulk solution concentration in mg/L.

In this model, one of the most significant parameter for controlling flux is the mass transfer coefficient, K_g; the ratio of diffussivity to the boundary layer thickness over which the concentration gradient exist. If the rate at which solute is transferred back into bulk solution is increased, then the thickness of the gel layer will decrease there by increasing the permeate flux. This can be achieved by high cross flow velocity or high scouring velocity (Yamamoto, 2000; Cheryan,1986).

The expression relating K_g to the physical properties of the solid solution , flow dimensions, and operating parameters exist in a lot of literatures(George and George,1980; Cheryan,1986). The coefficient K_g can be evaluated from the relationship between the dimensionless groups of Reynolds number (Re), Schmidt number(Sc) and Sherwood number (Sh) (McCabe and Smith,1994).

2.5.4.2 THE RESISTANCE MODEL

This model explained the effect of pressure on transmembrane flux throughout the period of flux decline during the UF/MF process. It showed that flux was inversely related to the total resistance at the surface of the membrane and directly as the transmembrane pressure drop. As can be seen from the previous expression (equation 2.6.) there is no pressure terms. This model predicts that flux is related to transmembrane pressure drop as given by equation 2.7. below.

$$J = \Delta P/(R_m + R_{ss} + R_{ds}) = \Delta P / R_t \quad \text{---}(2.7)$$

Where, R_m is intrinsic membrane resistance determined using pure water as the feed. R_{ss} and R_{ds} are resistances offered by the suspended solids and dissolves solids respectively and R_t is the total resistance. The use of this model was tested by Fane *et al.*, (1979) in their model for combined UF/AS system. A similar model of this nature cited in literature for macromolecules or colloidal feed stream (Meien and Ronaldo, 1994). In this case, flux is define as

$$J = \Delta P/(R_{m'} + R_p) \quad \text{---}(2.8)$$

Rm' is the addition of the intrinsic membrane resistance and the fouling layer resistance which are assume to be unaffected by the operating parameters. R_p is the polarized layer resistance. This term is a function of applied pressure and it comprises of two resistances; the gel polarized layer and the associated boundary layer. At low pressure, assuming fouling layer resistance is not too large ($R_m << R_p$), the flux will become independent of pressure and approach a limiting value which will be a function of the variable affecting the mass transfer properties of system.

2.5.4.3 OSMOTIC PRESSURE MODEL

Within the transmembrane pressure range in UF/MF processes the osmotic pressure value is ignored with little error. However, at times the concentration of the macromolecules might be too high that the effects can no longer be considered negligible (Wijmans *et al.*, 1986). According to (Swart *et al* 1995) the osmotic pressure model is applied when the gel layer does not form. High concentrated macromolecular solutes are still however, concentrated on the surface of the membrane. This explains the experimental results obtained when a solution is filtered, no gel layer is form and the retention is one hundred percent (Wijmans *et al.*, 1984). The equations relating membrane flux, transmembrane pressure drop and osmotic pressure difference are:

$$J = (\Delta P - \Delta \Pi)/R_m \quad \text{---(2.9.)}$$

Where ΔP(kPa) and $\Delta \Pi$ (kPa) are the transmembrane pressure drop and osmotic pressure difference across the membrane surface, respectively. This model assumes that the deviation from pure water occur mainly due to the osmotic pressure at the surface of the membrane. In order to relate the osmotic pressure to the concentration of macromolecules at the surface of the membrane, the following functional relationship was proposed by van't Hoff (Cheryan 1986., Wijmans *et al.*, 1986)

$$\Delta \Pi = A_1C + A_2C + A_3C + \text{--------} \quad \text{---------------(2.10.)}$$

where, $C = C_m$ is the concentration at the membrane surface, and A_1, A_2, A_3,----------, are known as the "viral coefficients", and are constant.

Equation (2.10.) can be relate to concentration in general form as :

$$\Delta \Pi = f(C_m) \quad \text{--(2.11.)}$$

To evaluate C_m the gel polarised model based on film theory from equation 2.6, was used

$$C_m = C_B \exp (J / K) \quad \text{--(2.12.)}$$

and substituting equation (2.12.) into equation (2.9.) gives:

$$J = (\Delta P - f(C_m \exp(J / K))/R_m \quad \text{--(2.13.)}$$

Thus increasing transmembrane pressure will at first increase the flux momentarily. As a result of more convective transport of solutes to membrane surface the value of C_m increases, thereby increasing $\Delta\Pi$ and resulting in a decrease in the driving force ($\Delta P - \Delta\Pi$) and flux.

CHAPTER THREE

3 MODEL DEVELOPMENT IN MEMBRANE BIOREACTOR SYSTEM

3.1 Theory

To fully understand the performance of MBR and to have a design tool, a steady state mathematical model was developed. The components of any mathematical model for single sludge systems consists of a set of equations that describe the process kinetics, material balance and stoichiometry. More sophisticated models which describe the changes that take place within the reactor with various conditions of feed rate and organic loading rate have been developed (De Silva *et al.*, 1998; Chaize and Huyard, 1991).

Here it is demonstrated how the model, coupled with chemical analysis systematical allowed in-depth interpretation of MBR performance. Chemical analysis and model interpretation identify advantages of aerobic-anoxic cycling (intermittent aeration and non-aeration systems) as compared to continuous aerobic-anoxic operation. The model discussed here is similar to those existing in literature (WRC 1986; Yamamoto et al., 1989; MetCalf and Eddy, 1991; De Silva *et al.*, 1998; Urbain *et al.*, 1997; Furumai and Rittmann, 1992).

3.2 Steady-state material balance

3.2.1 Material balance over solids

The assumption made here is that the species involved are in constant ratio (heterotrophic and autotrophic) and are homogeneously mixed throughout the entire system under a given set of steady-state conditions. From Figure 3.1., the material balance over solids is given as

Accumulation = Inflow - Outflow + Net growth

$$V\frac{dX}{dt}=Q_oX_o-(Q-Q_w)X_e-Q_wX+r_gV \ldots\ldots\ldots\ldots 3.1$$

Assuming that the biomass concentration in the input and output are zero and steady state condition prevail, equation 3.1. reduces to:

$$\frac{Q_wX}{V}=r_g=r_X-r_d \ldots\ldots\ldots\ldots 3.2$$

Where,

r_g = net rate of biomass growth, mass/unit volume .time

r_d = endogenous decay rate, mass/unit volume .time

r_x = rate of biomass growth, mass/unit volume .time

The relationship between substrate utilization rate, r_u, and biomass growth rate is given as

$r_x = -Yr_{su}$, Where, r_{su} = substrate utilization rate , mass/unit volume . time

and between endogenous decay rate and biomass concentration is

$r_d = -k_dX$

Where,

Y = Yield coefficient, (biomass produced)/(substrate consumed . time)

k_d = decay coefficient, d^{-1}

Substituting in equation 3.2

$$\frac{Q_w X}{VX} = -Y\frac{r_{su}}{X} - k_d$$

The left hand side represent the reciprocal of sludge age (as discussed in section 2.4.4.). Therefore

$$\frac{1}{\theta_c} = -Y\frac{r_{su}}{X} - k_d \quad \text{.......... 3.3}$$

the term $-r_{su}/x = U$ is the specific substrate utilization rate

$$U = \frac{S_o - S_e}{\theta X} = \frac{Q}{V_r}\frac{(S_o - S_e)}{X} \quad \text{.......... 3.4}$$

Where, V_r is the reactor volume and V_r/Q = HRT, hydraulic retention time. Therefore

$$\frac{1}{\theta_c} = YU - k_d \quad \text{.......... 3.5}$$

The mass concentration of the biomass can be estimated for by substituting for U in equation 3.4. to give

$$X=\frac{\theta_c Y(S_o-S_e)}{\theta(1+k_d\theta_c)} \quad \text{..........} 3.6$$

Equation 3.6. is used to determine the concentration of active biomass in the system without taking into consideration the inert biomass present in the system.

3.2.2 Mass balance over substrate

Accumulation = input - output - consumption

$$V\frac{dS}{dt}=Q_oS_o-Q_eS_e-(r_{su})V$$

At steady-state, the left hand side of the equation is equals to zero. Applying the Monod rate law, to give the relationship

$$-r_{su}=\frac{r_X}{Y}=\frac{\mu_m SX}{Y(k_s+S)}$$

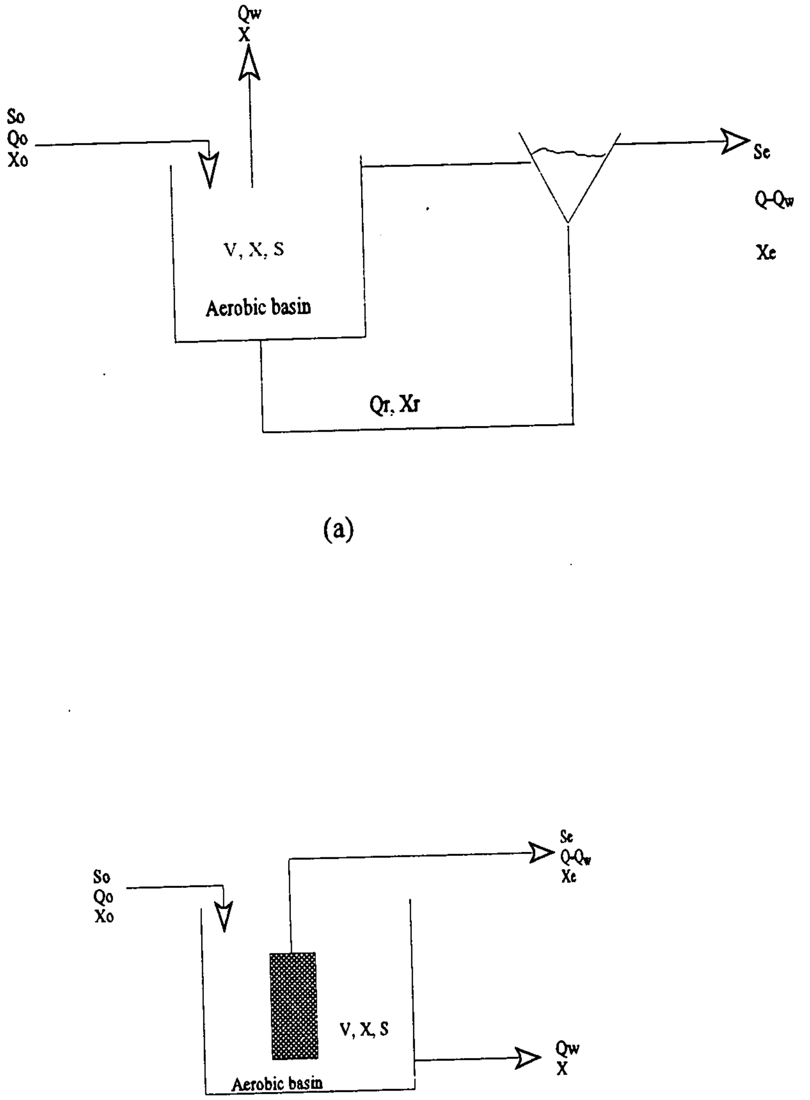
$$U=-\frac{r_{su}}{X}=\frac{kS}{k_s+S} \quad \text{..........} 3.7$$

Where,

k = maximum substrate utilization rate constant (mg substrate/mg biomass . day)

K_s = substrate saturation constant or half-velocity constant , substrate concentration at one-half the maximum growth rate (mg/L)

S = substrate concentration surrounding the biomass (mg/L)



(a)

(b)

Figure 3.1 Aerobic reactors with (a) Settling system (b) membrane system

Taking the reciprocal of both sides of equation 3.7., results in

$$\frac{1}{U}=\frac{K_s}{kS}+\frac{1}{k} \quad \text{..................................} 3.8$$

In a completely mixed reactor, the concentration of the substrate in the reactor is equal to the concentration of the substrate of the effluent therefore, $S = S_e$, and from figure 3.1.a and 3.2.b, if no wastage occurred, the concentration of biomass within the system increased to a large proportion. From equation 3.5., as $\theta_c \rightarrow \infty$, the equation becomes $0 = YU - k_d$, therefore $U = k_d / Y = F/M$. The experimental determined values of k_d and Y can be used to estimate F/M ratio and check if the estimated F/M ratio is in agreement with experimentally measured values.

Equation 3.5. and 3.8. are commonly used in estimating the kinetic constants Y, k_d, k, and K_s. From equation 3.5. it is clear that a plot of $1/\theta_c$ versus U should results a straight line of gradient Y and intercept $-k_d$. Similarly from equation 3.8., the plot of 1/U versus 1/S should result in a straight line of gradient K_s/k and intercept 1/k.

3.2.3 Process sludge

The value of the biomass; X in equation 3.5. above is assumed to be the active microbial mass of sludge and may be approximated by either the MLSS or MLVSS, both of which can be measured directly. However, both the MLSS and MLVSS measure materials apart from active biomass and may be considered to be composed of :

$$X(MLSS) = X_a + X_d + X_{io}$$

$$X(MLSS) = X\,(MLVSS) + X_{ii}$$

Where,

X_a = the concentration of active biomass

X_d = the concentration of inert endogenous residue

X_{io} = the concentration of particulate inert organic matter from the feed

X_{ii} = the concentration of particulate inert refractory inorganic from feed.

The inert endogenous residues can come from the decomposition of the live cells and will accumulate within the reactor at high sludge ages. Assuming that the residual fraction of the live biomass non-biodegradable fraction is defined as f_d therefore

$$X_d = f_d\, k_d\, X_a\, \theta_c \qquad \text{.......... 3.9}$$

The inert organic or inorganic solids entering with the feed will also accumulate within the reactor at high sludge ages, given as

$$X_{io} = X_{io}^{o}(\frac{\theta_c}{\theta})$$

$$X_{ii}=X_{ii}^{o}(\frac{\theta_c}{\theta})$$

Where X^{o}_{io} and X^{o}_{ii} are the inert organic and inorganic respectively. Therefore, the MLSS in the reactor or system becomes:

$$X(MLSS)=X_a(1+f_d k_d \theta_c)+(X_{io}^{o}+X_{ii}^{o})\frac{\theta_c}{\theta}$$

Assuming there are no inert solids in the feed therefore, MLSS = MLVSS = X_v and are approximated by

$$X_v=X_a(1+f_d k_d \theta_c) \qquad \text{3.10}$$

If equation 3.6. is corrected for inert solids from endogenous residence, the measured solids concentration is

$$X_v=\frac{Y(S_o-S_e)}{(1+k_d\theta_c)}(1+f_d k_d \theta_c)\frac{\theta_c}{\theta} \qquad \text{3.11}$$

The relationship between substrate utilization rate and sludge age as given by equation 3.4. is

$$U_v = \frac{1 + k_d\theta_c}{Y(1 + f_d k_d \theta_c)\theta_c} \quad \text{.......................................} 3.12$$

The error in estimating the values of Y and k_d is significant when a linear relation is assumed (see equation 3.5.). At low sludge ages equation 3.5. can be used directly with the measured value of MLSS or MLVSS approximating X_a (Sundstrom and Klei,1979). At high sludge ages ($\theta_c > 10$ days), the values of Y and k_d are determine using equation 3.12. An improved method can be seen in the work of Fraser (1975) by taking logarithms of both side of equation 3.12.

$$\log U_v = \log \frac{1 + k_d\theta_c}{(1 + f_d k_d \theta_c)\theta_c} - \log Y$$

Plot $1/\theta_c$ versus $(1 + k_d\theta_c)/(1 + f_d k_d X_a \theta_c)\,\theta_c$ on log-log plot by assuming $Y = 1$ for a selected range of k_d, say $k_d = 0$ to $k_d = 0.20$ .The curvature for each plot is dependent on the value of k_d. It has been found that the value f_d for domestic wastewater lies between 10-20% (WRC 1986). In this work the value f_d has been assumed to be 20%. Using the experimental data to plot $U_v = (S_o - S_e) / X_v\theta_c = 1/\theta_c$ a comparison to theoretical curves is made. The values of k_d attached to that theoretical curve which most closely fits the experimental curve is taken as k_d, this represents decay rate of biomass used in this study. The Y value is estimated from the displacement of the theoretical curves (when Y = 1) away from the experimental curve.

Again, strictly for active biomass, the observed yield coefficient (Y_{obs}) can be calculated by the following equation

$$Y_{obs} = \frac{Y}{1 + k_d\theta_c} \quad \text{...} 12a$$

where Y is often referred to as the true yield, as it represents the active amount of organism which have been synthesis from a given amount of substrate. The observed yield (Y_{obs}) is the yield value determine experimentally.

3.2.4 Estimation of Unbiodegradable Substrate as COD

The total COD of wastewater is made up of a biodegradable fraction, an unbiodegradable fraction and an active biomass fraction. The biodegradable fraction is measured as the ultimate biological oxygen demand (BOD_L). The biodegradable fraction of different wastewater differs. For example, in dairy effluent, the biodegradable fraction is between 80-96%.

The use of S in equation 3.8., involves only the biodegradable fraction of the COD of the liquor surrounding the bacteria. In this case, the BOD_L is a more useful measure than COD. BOD measurements are tedious and of variable accuracy depending on seeding dilution water, careful temperature control,and the presence of toxic components (WRC, 1986). The unbiodegradable fraction can be determined without BOD measurements (Fraser, 1975). Assuming no active biomass in effluent used.

Let S_t = total COD (mg/l)

S_b = biodegradable COD (mg/l)

S_u = unbiodegrable COD (mg/l)

$$S_t = S_b + S_u$$

Therefore, the total influent COD will be, (see figure 3.1.)

$$S_i = S_{ti} = S_{bi} + S_{ui}$$

Also for the effluent

$$S_e = S_{te} = S_{be} + S_{ue}$$

At steady-state, the influent and effluent unbiodegradable are the same, i.e.

$$S_{ui} = S_{ue} = S_u$$

$$\therefore S_i - S_e = (S_{bi} + S_{ui}) - (S_{be} + S_{ue})$$

$$= S_{bi} + S_{be}$$

Therefore the definition (or unit) of substrate utilization rate,U is the same, and it does not depend on whether the substrate concentration is defined in terms of COD or BOD_L.

The COD concentration, S of equation 3.7. is the effluent biodegradable COD. S_{be} can be express as $S = S_{be} = S_e - S_u$. S_e is the measured effluent COD, while S_u is obtained graphically as:

$$U = \frac{kS}{k_s + S} \qquad \text{3.7}$$

however, when $S = S_{be} = 0$, $U = 0$.

Plotting S_e ($= S_{be} + S_u$) versus U from experimental results, and extrapolating the straight line through the data to intercept the S_e axis, the value of S_e at the intercept is the concentration of unbiodegradable COD in the influent.

It should be noted that : (1) For the above method to be successful, a constant value of influent COD is necessary. (2)The value of K_s in equation 3.7. is expressed as biodegradable COD. (3)The straight line fit of equation 3.7. is acceptable if $K_s >> S$. (4) The substrate utilization rate correction, $U = U_v(1 + f_d k_d X_a \theta_c)$, is applicable in plotting U against S_e if greater degree of accuracy is sought.

The value of S_u is estimated from $S_{be} = S_e - S_u$, after calculating S_u from the plot above since S_e is the measured effluent value. The calculated S_{be} values is used as data for the plot of equation 3.7.

Alternative methods used in estimating unbiodegradable COD, S_u, is cited in WRC(1986). A laboratory unit was operated at two different sludge age >3 days. The COD of the filtrate through the 45μm filter paper is estimated, (S_u). The unbiodegradable particulate COD (S_{up}), is calculated by subtracting soluble COD (S_{be}) and soluble unbiodegradable COD (S_u) from the total COD (S_t) to give $S_{up} = S_t - b_{be} - S_u$.

3.2.5 Effluent Substrate Concentration

From equation 3.5.

$$U=\frac{1+\theta_c k_d}{\theta_c Y}$$

Substituting the above equation into equation 3.7. results in

$$\frac{kS}{K_s+S}=\frac{1+\theta_c k_d}{\theta_c Y}$$

$$S=\frac{K_s(1+\theta_c k_d)}{\theta_c(Yk-k_d)-1} \quad \text{.................................} 3.13$$

Thus, if the kinetic coefficients are known, equation 3.13. can be used to predict the biodegradable effluent from the reactor . It is clear from equation 3.13. that the concentration of the biodegradable effluent depends on the sludge age and is independent of the influent concentration, the feed rate, and hydraulic retention time. A plot of S versus θ_c can be used to specify the minimum sludge age needed for a particular waste stabilization to occur.

This minimum sludge age is the critical value of sludge age for which the biomass are washed out or wasted from the system faster than they can be produce. As substrate utilization is increased, the sludge age needed to affect a particular treatment decreases. At sludge age minimum, θ_c^m , the substrate utilization rate is maximum which can be represented by the following equation (re-writing equation 3.5.)

$$\frac{1}{\theta_c^m} = kY - k_d \quad \text{.......................................} 3.14$$

Where

k = maximum substrate utilization rate

θ_c^m = minimum sludge age

For design purposes, the value of the minimum sludge age is usually multiply by a safety factor to ensure adequate time for waste treatment process (WRC, 1986; Metcalf and Eddy ,1994)

3.3 MBR With Aerobic-Anoxic Systems

3.3.1 Nitrification

It is generally accepted that under given environmental conditions, nitrification follows Monod's rate law with ammonia and oxygen as substrates (Argaman and Brenner,1986).
The rate of substrate utilization is given as applied to the equation below

$$U_n = k_{nm}[\frac{N}{K_n+N}][\frac{O}{K_{on}+O}] \qquad 3.15$$

U_n the specific substrate utilization rate by nitrifiers, depends on N (NH_4^+-N concentration) and O(oxygen concentration) both with Monod or hyperbolic expressions. k_{nm} is the maximum specific rate of electron-donor utilization for nitrification. K_n and K_{no} are the half-maximum rate concentrations for NH_4^+-N and dissolved oxygen respectively. The multiplicative relation was shown to be the best representation of dual limitation by electron donor and acceptor (De Silva *et al.*, 1998).

If dissolved oxygen concentration in the aerobic basin (see figure 3.2.) is maintained well above the half saturation value for oxygen, the rate term for oxygen approaches unity and ammonia is the sole limiting substrate. At higher ammonia concentration the reaction becomes zero order (De Silva *et al.*, 1998; WRC, 1984; Argaman *et al.*, 1986). Argaman and co-worker (1986) showed that nearly complete nitrification take place at higher aerobic solid retention time (SRT) while the degree of nitrification decreases at lower aerobic SRT values. Researchers (Argaman and Brenner, 1986; WRC 1986) have argued that because of the extreme non linearity of the Monod equation in the vicinity of the half-saturation concentration, it is not practical to design for a desired ammonia effluent near this value using the Monod equation. If some what higher effluent concentrations are acceptable one may use a zero order rate equation and a maximum specific rate which is easier to assess.

An alternative approach to this, given by (Ekama *at al.*, 1986; Dold *et al.*, 1980) is based on the critical aerobic SRT for nitrification. If lower effluent ammonia is desired the critical SRT approach may be used (similar to equation 3.14.) and safety factor can be applied or specific nitrification corresponding to the desired effluent ammonia can be determined. If nitrification occurred in a single sludge age involving aerobic-anoxic system the aerobic SRT for nitrification is define as

$$\theta_{cA} = \theta_c \frac{V_A}{V_T}$$

Where,

θ_{cA} = aerobic STR, day

θ_c = overall SRT, day

V_A = volume of aerobic reactor, L

V_T = volume of entire system, L

and the equation for critical aerobic SRT relating to maximum specific nitrification rate is

$$\theta_{cA}^{m} = \frac{1}{k_n Y_n - k_{dn}}$$

Where,

$\theta_{cA}{}^{m}$ = critical aerobic SRT, days

k_n = maximum specific nitrification rate, mg NH_4^+-N/mg VSS. d

Y_A = nitrification yield coefficient, mg VSS/ mg NH_4^+-N

k_{dn} = endogenous decay coefficient for nitrifiers, d^{-1}

3.3.2 Biological Denitrification

This biological reaction is known as disseminative reduction of nitrogen or denitrification and involves the reduction of nitrate or nitrite present in wastewater to gaseous nitrogen which escapes to the atmosphere. This reaction normally takes place in the anoxic basin (see figure 3.2.) of wastewater treatment process.

Many investigators reported zero order denitrification with respect to nitrate down to very low nitrate concentrations (Dold *et al.*, 1980; Noran, 1991). The denitrification rate using multiplicative Monod kinetics as given by De Silva *et al.*,(1998) is :

$$U_{dN}=k_{dm}[\frac{S}{K_{sd}+S}][\frac{N_3}{K_{nd}+N_3}][\frac{K_{od}}{K_{od}+O}] \quad \text{......................} 3.16$$

From equation 3.15. it is clear that denitrification rate U_{dN} is inhibited by a high concentration of oxygen. Oxygen concentration well above K_{od} makes the right-hand term inside the last parenthesis go to zero. U_{dN} is also controlled by the substrate COD concentration and nitrate concentration.

Argaman and Brenner (1986) found that the rate of nitrate removal in denitrification processes is carbon limited, of which the rate controlling step is the rate removal of COD. For the range of COD removal studied, the removal rate was best described by first-order kinetics with a non-biodegradable residual. The integrated form of the rate expression is:

$$U_{dN}=\frac{S_1-S_2}{X_h\theta_d}=K_{SH}(S_2-S_u) \qquad 3.17$$

Where,

S_1, S_2 = COD concentration in influent and effluent of anoxic respectively.

K_{SH} = COD removal rate coefficient, L/mg.day

θ_d = once through residence time in anoxic basin, days

Investigator have also shown that nitrate reduction in the primary anoxic basin is proportional to two

$$N_1-N_2=\alpha S_{bi}+k_2X_a\theta_d \qquad 3.18$$

rates occurring simultaneously (WRC, 1986). Nitrate reduction as shown by equation 3.17. has two phases. The first phase of denitrification is proportional to the influent biodegradable COD while the second phase is independent of the nitrate concentration but a function of the active biomass (X_a) in the reactor. It was hypothesised that the rapid first phase of denitrification is associated with the utilization of readily biodegradable COD (RBCOD) in the influent and that the second phase of denitrification is governed by the slowly biodegradable COD (SBCOD) and the organism death (WRC 1986; Dold *et al.*, 1980).

N_1, N_2 = the influent and effluent nitrate concentration in anoxic reactor

α = denitrification constant attributable to the readily biodegradable COD,

(mg NO_3-N/mg RBCOD)

k_2 = second denitrification rate constant (mg NO_3-N/mg VSS.d)

X_a = active mass concentration (mg VSS/L)

b_i = biodegradable COD concentration (mg/L)

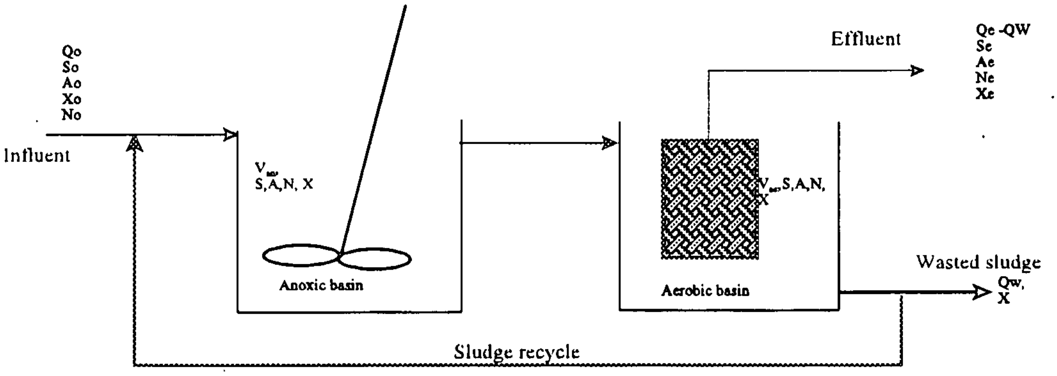


Figure 3.2 Bench scale single sludge treatment unit

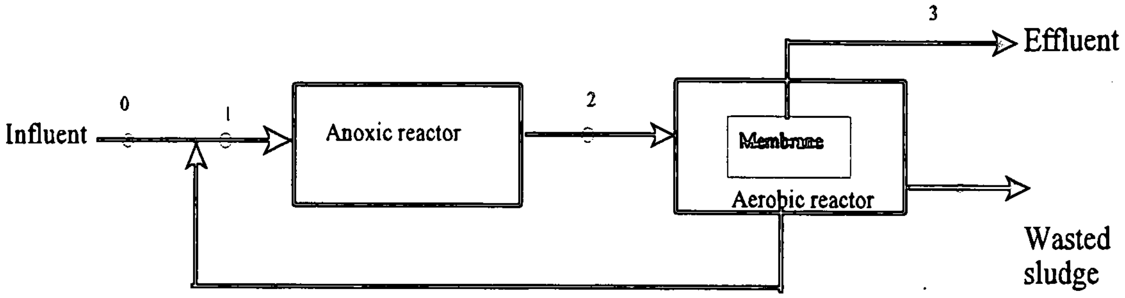


Figure 3.3 Basic definition for MBR systems

3.4 Summary

Equation 3.5. and 3.8. will be used in estimating the kinetic constants Y, k_d, k, and K_s using the data collected from MBR and activated sludge system. From equation 3.5. it is clear that a plot of $1/\theta_c$ versus U should results a straight line of gradient Y and intercept $-k_d$. Similarly from equation 3.8., the plot of 1/U versus 1/S should result in a straight line of gradient K_s/k and intercept 1/k.

The nitrifiers utilization rate as expressed by equation 3.15 will be used in estimating the maximum substrate utilization rate of the nitrifiers. The nitrifiers utilization rate and sludge age are related to the concentration of nitrifiers in the reactor. Since activated sludge has a limited sludge age within which to achieve its treatment efficiency and MBR can be operated at long sludge age therefore, a plot of nitrifiers utilization rate versus effluent ammonia concentration from MBR and activated sludge system will be used in estimating maximum nitrification rate and may be used in comparing the treatment efficiency of both systems.

CHAPTER FOUR

4 EXPERIMENTAL

4.1 Materials

4.1.1 Woven fibre membrane material and module design

The membrane units were fabricated at the Department of Chemical Engineering, M L Sultan Technikon, Durban. The module consists of a single length of woven polyester hose i.d. 25 mm wrapped around the outside of a PVC porous tube and of length 175 mm. Both tube ends were properly glued to a straight connector and the other ends connected to silicone tubing that runs through the pump. The treated water that permeate through the membrane from the outside of the membrane-porous tube was collected from the lumen of the tube by suction.

The first 35 mm and the last 15 mm of the tube was rendered impermeable by the application of a commercial resin. The effective length of the porous tube was thus 125 mm yielding an effective filtration area of 98 cm^2. The pore size of the woven fibre membrane was 1-5 μm.

Reactor material

Two rectangular reactors , 26 cm × 26 cm × 32 cm was made of transparent acrylic plastic sheet with total working volume of 50 litres (25 litres each).

Pump

Two peristaltic pumps were used for the study; A Watson-Marlow pump (model 313S/D) and Masterflex pump (model 752010). The Watson Marlow pump has a front panel mounted potentiometer for manual speed control, as well as instant reverse and power switches.

It was connected with two pump-heads both having the same speed rate. The flow out of the pump-head was determined by the diameter of tubing across the pump head. One of the two pump head was used to feed the reactor from the influent tank while the other was used to recycle mixed liquor from the aerobic to anoxic reactor or recycle sludge from the settling tank to anoxic reactor.

The Masterflex pump was connected with a single roller head that was used to extract permeate from the membrane module. Marprene tubing with i.d. 3.2, 4.8. and 6.1 mm were used across each pump head. The size used depends on the desired flow rate.

Settling tank:
An inclined cylindrical transparent PVC settling tank was used to thicken the effluent sludge from the activated sludge. A stirrer was installed driven by 7 rpm motor. The specification of the settling tank are

Total volume = 2.5 L

Operating Volume = 2.2 L

Inclination = 25° from the vertical.

Instrumentation :
Flow rate: Feed and recirculation flow rate was measured by the pump setting with respect to the particular diameter of the tubing being used. The pumps were calibrated to check for the accuracy of the flow rate, it was discovered that the range of flow rates given by the pump manual almost equalled the calibrated values.

Pressure: A pressure gauge with a range of -100 to 150 kPa was placed along the tubing line between the outlet of the membrane module and suction pump to measure the pressure drop across the membrane.

pH: The pH and the temperature of the sludge was measured using a multi-meter.

Concentration measurement: Appendix 2 gives detailed references to all analyses conducted for the determination of chemical composition.

4.2 METHODS

4.2.1 Short-term experiment on membrane filtration

The effect of suction pressure and mixed liquor suspended solid concentration on short performance of the locally made membrane module were studied. The schematic diagram of the experimental set-up is shown in figure 4.1. The sludge used was seeded from Darville Wastewater Treatment Plant and was diluted or concentrated to the desired concentration. The reactor volume was 22 litres and was well mixed by aeration from the bottom of the reactor. The membrane module was immersed and suspended vertically in the reactor. All experimental runs was done for several hours.

The following operational modes were studied (i) 60:15 (ii) 120 : 15 (iii) continuous operation. Where 60:15 implies 60 minutes of filtration and 15 minutes of air backwash. Similar operation was carried out using water as backwash.

4.2.2 Laboratory pilot scale activated sludge unit

Experimental work was centered on pilot scale test of continuous single sludge unit. The pilot plant shown in figure 4.1. was composed of completely mixed biological reactor. The total working volume of the two reactors is 40 litres with aerobic volume fraction of 57.5 %.

Feed was pumped into the anoxic zone of the unit. Mixed liquor was recycled from the bottom of clarifier to the anoxic reactor. Air diffuser was used to aerate the aerobic reactor and the anoxic reactor was mixed with electric stirrer.

The reactor was originally seeded with mixture of sewage sludge from Darville Wastewater Treatment Works, Pietermariztburg and then acclimatized for a period of 7-10 days. Feed consist of dairy effluent collected from Clover, Queensburg. A supply tank was filled with the raw waste almost every day which was continuously stirred with a variable speed mixer. For detailed effluent characteristics see appendix 3. Temperature and pH of the mixed liquor were controlled in the range of 22-27 °C and 7.0- 8.4 respectively. DO concentration was monitored constantly at a level of < 0.5 in the anoxic reactor and between 2-6 in the aerobic reactor using oxygen probe meter.

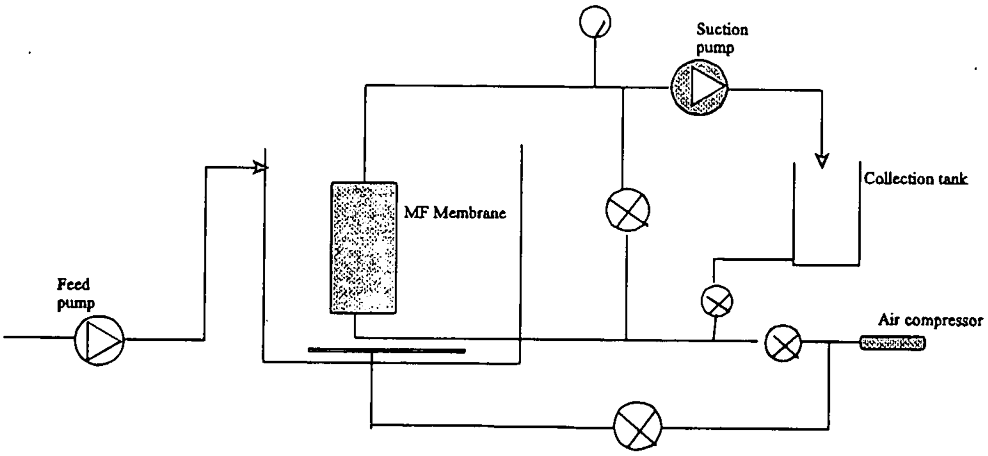


Figure 4.1 hybrid membrane bioreactor

4.2.3 Laboratory Pilot-Scale MBR

The modified system is depicted in Figure 4.2. It consists of two reactor zones, the aerobic zone for carbon oxidation and nitrification and the anoxic zone for denitrification. The raw wastewater was introduced into the anoxic tank after screening through a 750μm circular disc screen. The total working volume of the reactors is 40 litres separated into an aerobic volume fraction of 55 percent. One membrane unit (1-5μm) with total surface area 9 8 cm^2 was submerged in the aerobic tank with the top and bottom ends of the membrane connected to the suction and back-flushing pumps respectively.

The volumetric loading into the system was between 0.486 -2.283 kg COD/m^3/d. An air diffuser was used to create aeration in the aerobic tank, while an electric stirrer which rotates at about 40 rpm was installed in the anoxic tank to keep the biomass in suspension. The pH was maintained between 7.2 and 8.2 while dissolved oxygen was maintained between 2.0 - 6.0 in the aerobic tank and below 0.5 in the anoxic tank. The effect of different HRT on the performance of the system was investigated. The HRT of the membrane bioreactor process was varied between 4.0 to 12.7 hours by controlling the effluent permeate fluxes at 43.2, 69.1, 103.4, and 138.0 L/d. The change in trans-membrane pressure was monitor daily to check for the clogging of membrane during operation. Solids Retention Time (SRT) in the reactor was appriximately 100 days since there was no sludge wasting (except sample taken for analysis) during the first phase of experiment. The MLSS concentration was determined every two to three days.

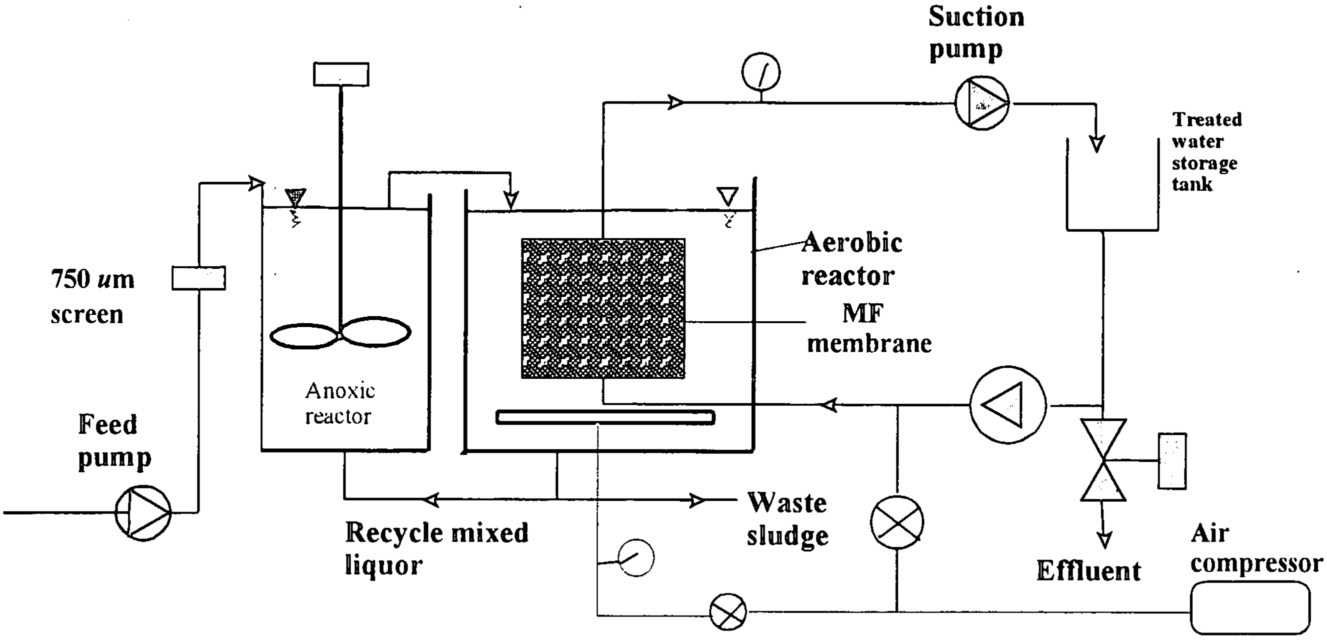


Figure 4.2: Schematic diagram of hybrid membrane bioreactor

CHAPTER FIVE

RESULTS

5.1 General

5.2 Activated Sludge Units

The values of all the results obtained from this experimental study are presented in full in Appendices 4 and 5. The values presented in the tables are average values obtained every 2 to 3 days. The tables of calculated values for removal rates are presented in relevant sections of this chapter. The calculated values of substrate removal rate of Appendix 5 are presented in Table 5.1.

Sigma Plot and Microsoft Power Point are the computer soft ware programs used to fit the data. Linear regression was used to fit various straight lines .

The laboratory studies were performed in two runs, run 1 and run 2 . The first run was performed using aerobic-anoxic basin for carbon oxidation, nitrification and denitrification. The results are in Appendix 4 (Table A4.1-A4.4). The results obtained from run 2 which was carried out under continuous aeration for carbon oxidation alone are displayed in Appendix 5 (Table A5.1-A5.4).

Generally, the influent COD for both runs was between 400-950 mg/L. The raw dairy effluent was diluted to give the desired COD concentration. The effluent collected have little or no ammonia in it. It was spike with a solution of ammonia sulphate and potassium hydrogen phosphate to give a solution equal to or higher than the ratio 100:5:1 (COD:NH_4^+-N:P). Analysis of the solids was done on very small volumes of sludge collected from the reactor so as not to affect the sludge age.

5.2.1 Determination of Yield (Y) and endogenous decay coefficients (k_d)

It was shown in section 3.2. that in order to determine Y and k_d a linear fit to a set of data points of $1/\theta_c$ against U had to be found (see equation 3.4.). This can also be done using equation 3.11 with the correction factors but the ranges of sludge age used for this study does not permit the use of equation 3.11. Figure 5.1. shows the data points of Table 5.1. plotted as $1/\theta_c$ versus U, the straight was fitted according to least squares method.

The value of Y obtained from the slope of Figure 5.1. is 0.204 mg VSS/mg COD and that of k_d is 0.013 d^{-1}. The correction factor of equation 3.10 is $X_v = X_a(1 + f_d k_d X_a \theta_c)$ and if by putting the value of k_d obtained from the plot into the correction factor and if f_d is taken to be 20 % (Sundstrom and Klei, 1979), it becomes $(1 + 0.0026 \theta_c)$. The effect is only 7.8 % when the sludge age is 30 days therefore, the factor can be neglected. The use of correction factor becomes significant at a very long sludge age, in this case the reciprocal of the sludge is plotted against U_v as indicated by equation 3.12 on a log-log plot. When using equation 3.12. different values of k_d are selected. The best fit with Y = 1, that corresponds to the straight line of equation 3.10 is the acceptable fit.

5.2.2 Determination of K_S, k and S_u

The determination of K_S, k, and S_u using graphical method was stressed in section 3.2.2. It was also shown in section 3.2.4., that unbiodegradable soluble COD in the influent is equal to the unbiodegradable COD exiting the system. The use of COD utilization rate in the anoxic reactor versus effluent COD in run 2 enabled the estimation the unbiodegradable COD in the system.

Before determining K_S and k the value of unbiodegradable must be found. This can also be done assuming K_S is much greater than $S = (S_e - S_u)$ and from equation 3.8., a linear relationship exists between U_V and $(S_e - S_u)$, and the intercept of the straight line fit to the data and $S_e - S_u$ axis would give the value of S_u. Inspection of Figure 5.2. revealed that a plot of U_v versus (Se - Su) resulted in a straight line which meant that the assumption that Ks >> (Se - Su) is correct. The point of interception of the straight line and x-axis is the value of unbiodegradable and is equal to 17.5 mg/L. Figure 5.3 shows the experimental points plotted as $1/U_v$ versus $1/(S_e - S_u)$. From the slope and intercept, the value of $k = 3.70\ d^{-1}$, $K_S = 17.5$ mg/L and $S_u = 16.75$ mg/L.

Table 5.1 Laboratory pilot scale aerobic digestion results.

This table is made from Table A5.1 to A5.4 (Appendix 5)

Table	Feed L/d	S_i mgCOD/L	$S_e \pm \sigma S_e$ mgCOD/L	$U_v \pm \sigma U_v$ mgCOD/mgVSS.d	MLSS mg/L	VSS ±σVSS mg/L	θ_c day
TableA5.1	69.1	457.38	18.98 ±1.09	0.405 ±0.03	4713.25	3222.25 ±50.98	14.29
TableA5.2	69.1	461.63	19.41 ±1.16	0.460 ±0.02	3817.38	285 6.13±55.23	11.76
TableA5.3	69.1	463.57	19.13 ±0.99	0.560 ±0.02	3299.14	2377.00 ±35.91	10.52
TableA5.4	69.1	457.29	21.50 ±1.14	0.770 ±0.03	2787.29	1704.14 ±48.23	7.14

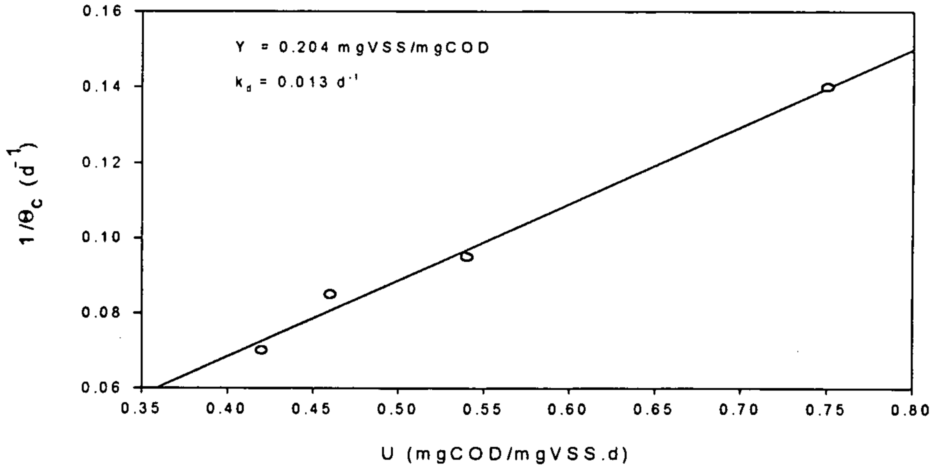


Figure 5.1 Determination of Y and k_d using linear relation

Table 5.2 Laboratory pilot scale aerobic reactor results.

This table is made from Table A5.1 to A5.4 (Appendix 5)

Table	1/(S_e - S_u) ± σ 1/(S_e - S_u) L/COD	1/U_v ±σ(1/Uv) mgVSS/mgCOD.d	VSS ±σVSS mg/L
TableA5.1	0.457 ±0.103	2.47 ±0.20	3222.25 ±50.98
Table A5.2	0.376 ±0.130	2.17 ±0.07	285 6.13±55.23
TableA5.3	0.360 ±0.125	1.79 ±0.08	2377.00 ±35.91
TableA5.3	0.273 ±0.467	1.30 ±0.0.05	1704.14 ±48.23

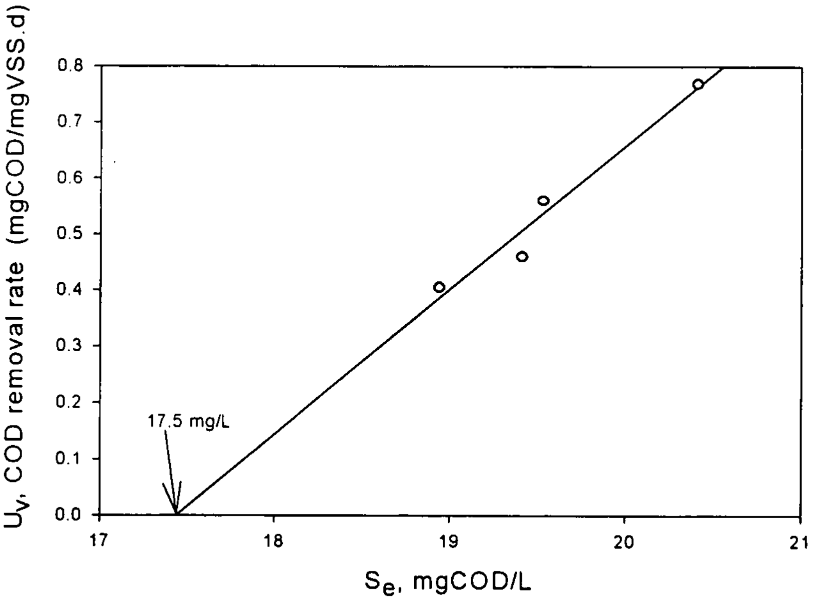


Figure 5.2 Determination of S_u

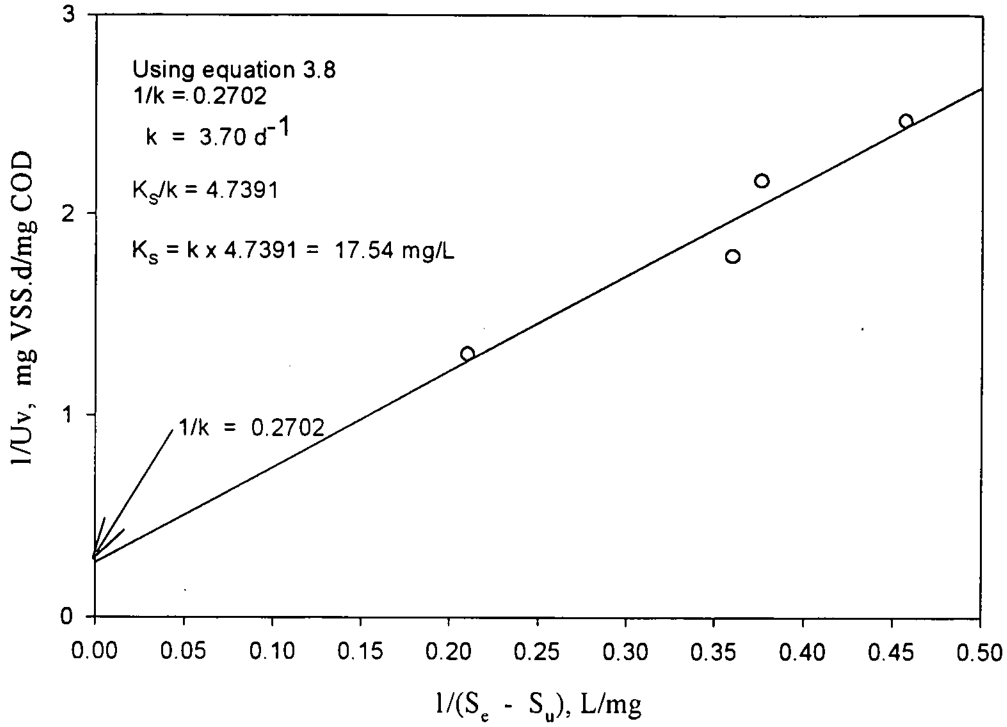


Figure 5.3 Determination of k and K_S

5.2.3 Kinetic Coefficients

The kinetic coefficients determined from experimental data is used to predict the performance of the aerobic reactor using equation 3.13. and substituting various values of coefficients i.e. K_S, k, k_d, Y and S_u (determined by experimental data). The effluent COD concentration can be determined for a selected value of sludge age . By inspection of equation 3.13:

$$S_e = S_u + \frac{K_s(1+\theta_c k_d)}{\theta_c(Yk-k_d)-1} \text{..............................} 3.13.$$

the expression shows that effluent concentration is dependent upon the kinetic coefficients which means that the efficiency of the system depends on the accuracy of the coefficients.

The solid concentration in the aerobic system can be predicted by equation 3.10. given as

$$X_v = \frac{Y(S_o - S_e)}{(1+k_d\theta_c)} \frac{\theta_c}{\theta}(1+f_d k_d \theta_c)$$

When SRT and HRT are equivalent, the concentration of solids in the reactor is linearly related to the effluent concentration since all coefficients are known including S_o. If $\theta_c \neq \theta$, the concentration of the solids (biomass) in the reactor will increase as the sludge age increases.

5.3 Laboratory Pilot-Scale Aerobic-Anoxic Reactor

Results obtained from this experimental study are shown in Appendix 4. Table 5.3. is a summary of the data of Appendix 4 (Table A4.1 to A4.4). Method of operation of this run is similar to the aerobic reaction process but the estimation of kinetic coefficients is more complex because of the reactors in series involved. Table 5.3. shows nitrate removal rate, COD removal rate from the anoxic reactor and nitrification rate in the system.

5.3.1 Determination of COD Removal Rate Coefficient in Anoxic Reactor;K_{SH}, S_u , b_N and α_N

DO concentration in anoxic reactor was always kept below 0.5 mg/L. This reduces the complexity of the Monod multiplicative kinetics given by equation 3.16. The nitrate removal rate has been reported to follow zero order kinetics with respect to the nitrate concentration (WRC 1986; Noran 1991; De Silva *et al.*, 1998). Steady-state material balance around the system described by Figure 3.2. gives estimated values of denitrification rate given as (Argaman and Brenner,1986).

$$U_{dN}=\frac{(A_o-A_e)+(N_o-N_e)-\lambda_N(S_o-S_e)}{\theta_D X_h} \quad \text{.....................................} 5.3.1a$$

Where A_o and A_e = influent and effluent ammonia concentration respectively (mg/L)
N_o and N_e= influent and effluent nitrate concentration respectively (mg/L)
S_o and S_e = influent and effluent COD concentration respectively (mg/L)
Θ_c = HRT through anoxic reactor
Xh = heterotrophs biomass concentration (mg/L)
λ = Ammonia assimilation factor (mg NH_4^+-N/mg COD)
U_{dN} = Specific denitrification rate (mgNO_3-N/mgVSS.d)

Inspection of data of Figure 5.4. shows that removal rate of both nitrate and COD in anoxic reactor are linearly related defined by equation 5.3.1b. The data used for this plot are based on average values given in Table 5.3. and $\lambda = 0.031$ based on the findings by Argaman and Brenner,(1986). The rates as estimated are approximate values because no attempt was made to do population studies on various species involved. Heterotrophic biomass used was based on the approximated value of VSS and the nitrifier concentration was accounted for in terms of the ratio of COD to ammonia in the influent (Metcalf and Eddy, 1994).

The value of nitrate utilization coefficient ,α_N, obtained from the slope as plotted in Figure 5.4., is is 7.105 × 10^{-3} mg NO_3-N/mg COD and b_N,the denitrifier endogenous rate from the intercept is 0.206 d^{-1}.

Figure 5.5. shows the average values of COD removal rate as plotted against the effluent COD concentration from the anoxic reactor. Inspection of data of Table 5.3. and effluent COD from anoxic reactor shows that the COD removal rate ($U_{COD,AN}$) is linearly related to the effluent COD concentration in the anoxic basin. Extention of the straight line to the x-axis gives an approximate value of the unbiodegadable COD(Argaman *et al.*, 1986). The slope of the straight line is the COD removal rate coefficient represented as K_{SH} = 5.10 L/mg.

$$U_{COD,AN} = \alpha_N U_{dN,AN} + b_N \qquad \text{.................................... 5.3.1b}$$

Table 5.3. Estimated nitrate removal rate ,COD removal rate and nitrification rate.

θ_c= 18.5 day

Date	$U_{dN,AN}$ (mg NO3-N/mgVSS.d)	$U_{COD,AN}$ (mgCOD/mg VSS.day)	U_N (mg NH4+-N/mg NVSS)
26/02/99	0.160	0.59	2.08
01/03/99	0.160	0.57	1.92
04/03/99	0.065	0.57	1.93
07/03/99	0.160	0.60	2.04
10/03/99	0.156	0.60	1.89
13/03/99	0.191	0.57	1.90
16/03/99	1.139	0.64	2.12
19/03/99	0.266	0.59	3.15
22/03/99	0.290	0.59	3.38
25/03/99	0.210.	0.60	2.40
Average	0.280	0.59	2.281
σ	0.0308	0.021	0.543

Table 5.3 (cont.)

θ_c= 15 day

Date	$U_{dN,AN}$ (mg NO3-N/mg VSS.day)	$U_{COD,AN}$ (mgCOD/mg VSS.day)	U_N (mg NH4+-N/mg NVSS)
30/03/99	0.200	0.67	2.60
02/04/99	0.230	0.67	2.19
05/04/99	0.230	0.66	2.54
08/04/99	0.193	0.67	2.23
11/04/99	0.205	0.73	2.14
13/04/99	0.171	0.59	2.39
16/04/99	0.206	0.71	2.56
18/05/99	0.211	0.71	2.49
20/05/99	0.278	0.69	2.45
22/05/99	0.188	0.68	2.32
Average	0.211	0.68	2.42
σ	0.0295	0.038	0.164

Table 5.3 (cont.)

θ_c= 10.2 day

Date	$U_{dN,AN}$ (mg NO3-N/mgVSS.d)	$U_{COD,AN}$ (mgCOD/mg VSS.day)	U_N (mg NH4+-N/mg NVSS)
26/04/99	0.220	0.75	3.02
28/04/99	0.150	0.76	2.38
30/04/99	0.241	0.72	2.83
02/05/99	0.248	0.74	2.96
04/05/99	0.198	0.75	2.62
06/05/99	0.236	0.73	2.81
08/05/99	0.198	0.69	2.24
10/05/99	0.311	0.75	3.01
12/05/99	0.264	0.76	2.49
14/05/99	0.198.	0.78	2.75
Average	0.226	0.743	2.712
σ	0.044	0.025	0.270

Table 5.3 (cont.)

θ_c= 7.2 day

Date	U_{dN} (mg NO3-N/mgVSS.d)	$U_{COD,AN}$ (mgCOD/mg VSS.day)	U_N (mg NH4+-N/mg NVSS)
18/05/99	0.178	0.87	2.60
20/05/99	0.072	0.73	0.64
22/05/99	0.141	0.83	2.17
24/05/99	0.254	0.86	3.03
28/05/99	0.168	0.82	2.94
30/05/99	0.277	0.81	2.89
02/06/99	0.211	0.75	2.64
04/06/99	0.169	0.70	1.66
06/06/99	0.222	0.75	2.56
08/06/99	0.216	0.75	2.65
Average	0.191	0.788	2.358
σ	0.059	0.059	0.164

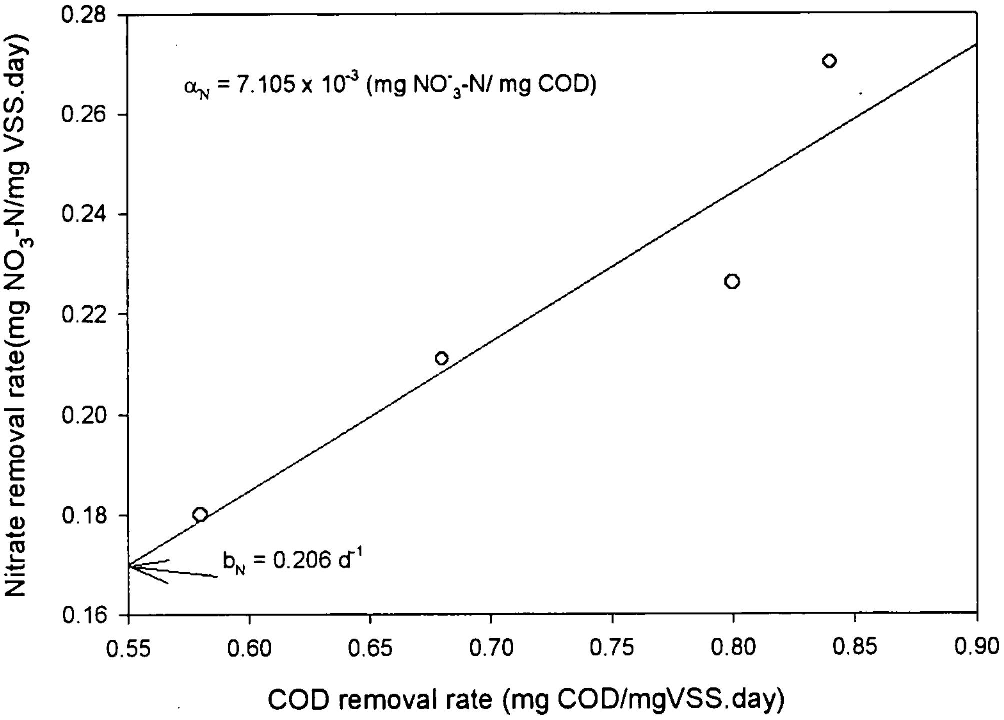


Figure 5.4 Determination of nitrate consumption coefficients.

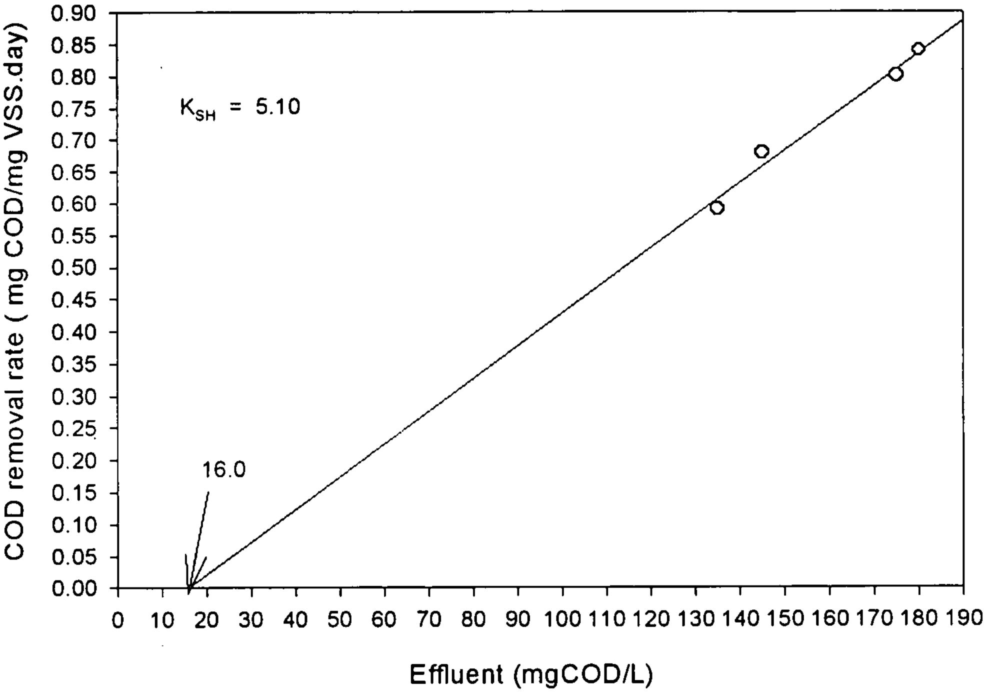


Figure 5.5 Determination of COD removal coefficients.

5.3.2 Determination of Maximum Specific Nitrification Rate

See Table 5.3 for values of specific nitrification rate, U_N, and the corresponding effluent ammonia concentration in Appendix 4. Estimated values of nitrification rate and the corresponded effluent ammonia concentration are plotted in Figure 5.6. For effluent ammonia concentration between 2.0 and 40 mg/L ,the maximum specific nitrification is 3.0 mg NH_4^+-N/mg NVSS.day. For lower effluent concentrations, lower rates are obtained. The rate in Figure 5.6. is expressed in terms of nitrifiers biomass, thus the use of NVSS instead of volatile suspended solids (VSS).

To determine nitrifier fraction, the total biomass concentration must be known which can be obtained experimentally. The estimated nitrifier fraction was based on the influent COD to ammonia ratio as given by Metcalf and Eddy, (1994). In this study, for a ratio of 3, 4, and 5 nitrifiers fraction used is 0.083, 0.064 and 0.053 respectively.

By inspection of data of Figure 5.6. it is difficult to determine the maximum specific nitrification rate value at lower effluent concentrations which will render erroneous results regarding half saturation coefficients. For this reason, it is suggested that a safety factor should be used by using the maximum rate to determine the minimum sludge age then multiply by the safety factor to give the correct minimum sludge age under which nitrification will occurred (WRC, 1986; Argaman and Brenner, 1986).

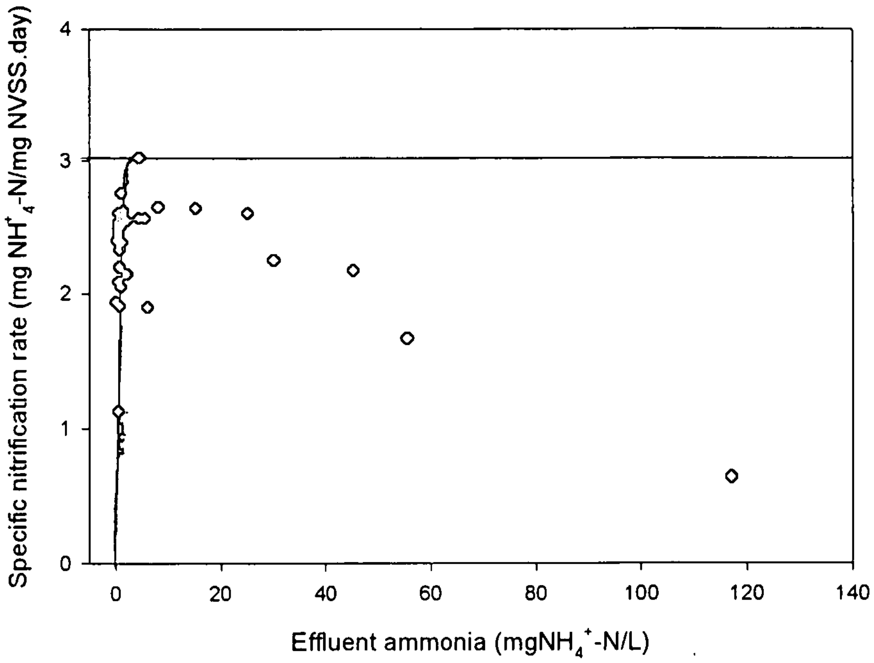


Figure 5.6 Correlation of specific nitrification with effluent ammonia.

5.4 Short Term Performance of Woven Fibre Membrane

5.4.1 Air Backflushing Method

The effect of pressure drop with time was examined for the different runs considered in this study. Figure 5.7 shows the time rate of change of pressure drop and flux for different hours of operation. The flux is calculated from the amount of water filtered per unit time per unit surface area of the membrane. Under constant filtration, the permeate flux was expected to constant. The gradual drop in permeate flux with time as shown by curve AC is due to changes in tubing diameter within the pump rollers-head. A slightly decrease is experienced each time the rollers passes over the tubing. The decrease in tubing diameter cause the volume flow rate across the pump head to decrease which

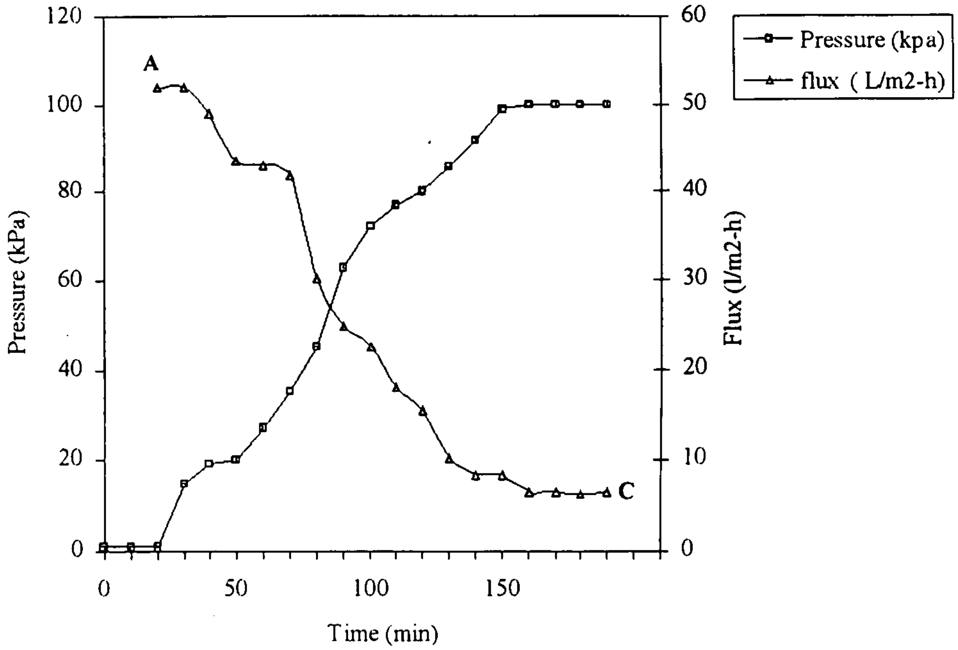


Figure 5.7 Permeate flux obtained in continuous operation.

ultimately affect the membrane flux (Pillay, 2000). The pressure force applied by the pump which was kept constant in the study could not keep the pressure difference across the membrane constant because the pressure difference was dependent on a dynamic status of the membrane that is, the extent of clogging inside the membrane as well as the extent of formation of sludge cake on the outer surface of the membrane. Cake formation at the membrane surface is responsible for the increase in transmembrane pressure drop. This can be explained with the use of equation 2.8. (section 2.5.4.2.).

ΔP is the trans-membrane pressure drop as shown in Figure 5.8. It increases with increasing in cake thickness. P_{in} and P_{atm} is the pressure on the inside and outside the membrane respectively. P_{atm} is atmospheric pressure. During suction, a vacuum is created on the inside of the module creating a lower pressure on the inside than outside. The imbalance of pressure across the membrane

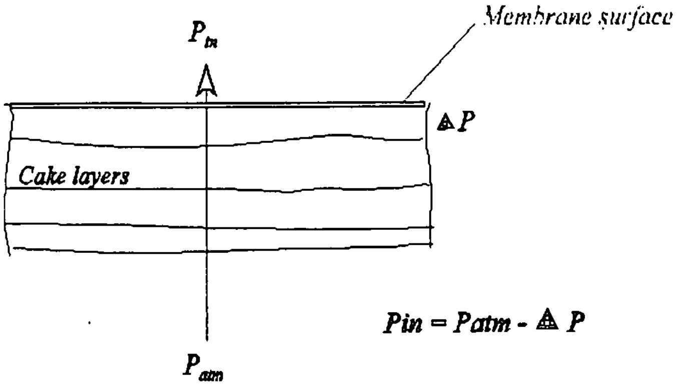


Figure 5.8 -Cake formation at membrane surface

resulted in the flow of mixed liquor toward the membrane surface. The particles of lower molecular weight cut off (MWCO) permeate through the membrane while those greater are retained at the membrane surface or within the pore of the membrane. Proper aeration around the membrane module helps to shear off some of the cake formed at the membrane surface otherwise there will be cake build up which will increase the pressure drop across the membrane. Since the pressure outside the membrane is constant (atmospheric) and an increase in trans-membrane pressure resulted in negative pressure increase inside the membrane as can be seen in figure 5.8.

Figures 5.9., 5.10. and 5.11. shows the performances at different mode of operation. The time rate of pressure drop is used to assess the performance of the membrane system. Continuous operation shows rapid decrease in flux with time, while the cyclic operation (discontinuous mode) can partially recover pressure after air diffusion period. The cyclic operation could not completely removed the clogging which was noticed by gradual increase of pressure drop with time but it could improved the pressure recovering and the flux up to about 200 % compared to continuous operation.

Figure 5.9. shows cycle of 8 hours filtration and 15 minutes air backflush. This cycle was attempted to see if long term filtration will show a significant improvement as compared to other cycle experimented. As can be seen , the rate of pressure increased after the first 8 hours of filtration indicates that this method cannot be applied to long term operation.

On the other hand, in the case of 120:15 operation (see Figure 5.10.), the filtration time was reduced but the air backflush time was kept constant. The 15 minutes backflush time that was kept constant has been found to be the optimum time for backflush (Visvanathan, 1997). Normally, when the suction pump is switched on filtration begin immediately and it take the filtrate some time to fill the cavity of the membrane module before appearing at the pump delivery outlet for collection. For this study it took almost about 5-15 minutes for permeate to appear from delivery end. This time lag could not permit the study of 15:15 (15 minutes filtration and 15 minutes air backflush) and 30:15 cycles because of the low volume of filtrate that was obtained. However, 120:15 operation cycle has more cumulative permeate volume as compared to 8hr:15min operation cycle which is attributed to low rate of pressure increase after air diffusion. Figure 5.11. shows 60:15 operation mode. This operation has more cycle per day and the rate of pressure recovery is similar to 120:15 cycle but more cumulative permeate volume is obtained from the latter method because of long filtration period.

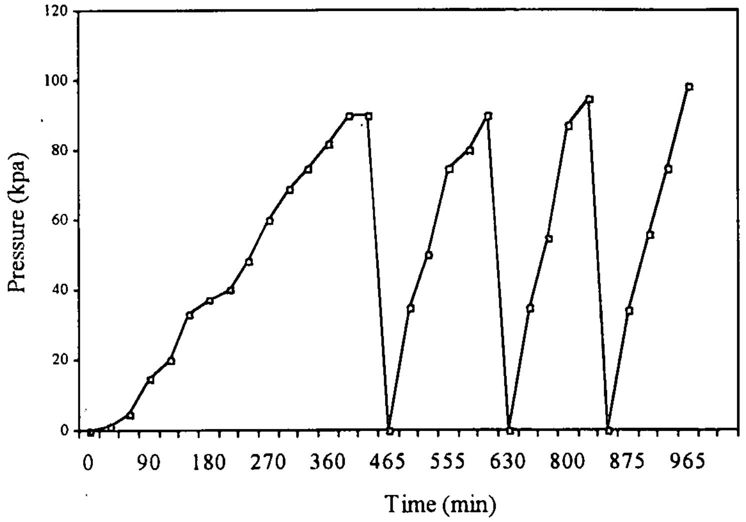


Figure 5.9 Variation of transmembrane pressure with time at 8 hours and 15 minutes cycle operation

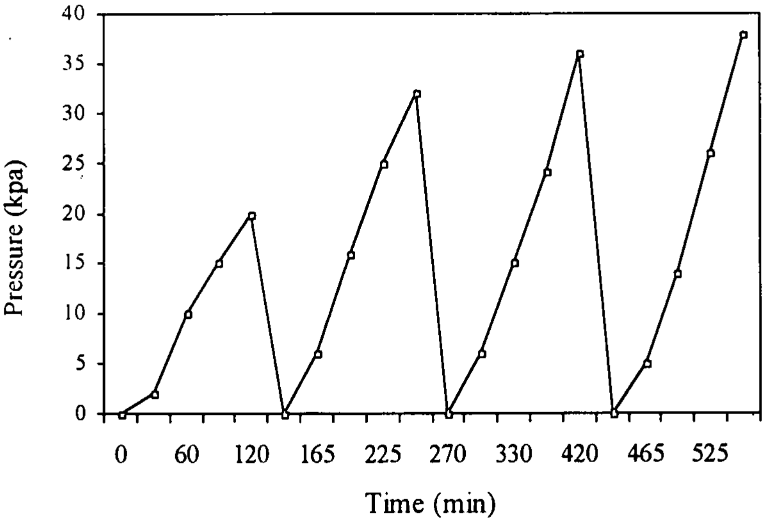


Figure 5.10 Variation of transmembrane pressure with time at 120:15

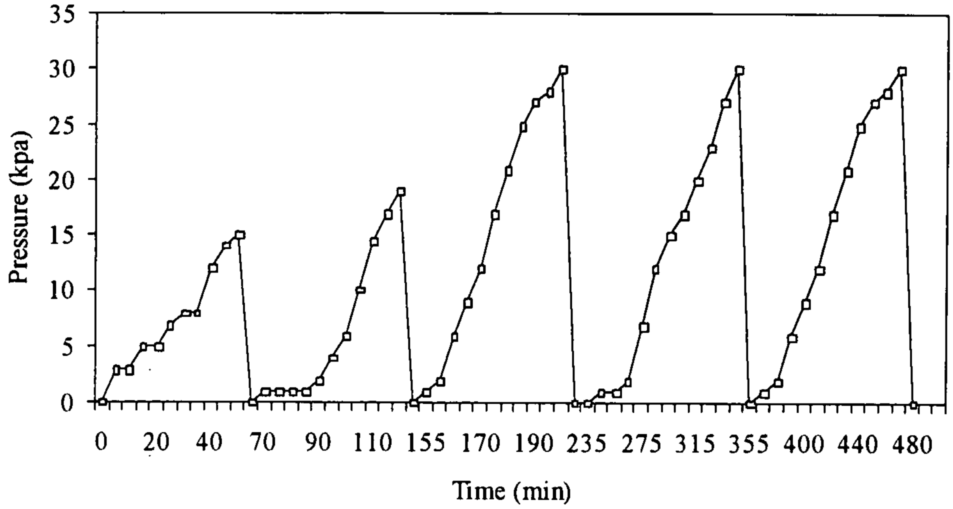


Figure 5.11 Variation of transmembrane pressure with time at 60:15

5.4.2 Water Backflush Method

Figures 5.12., 5.13. and 5.14. shows the profiles of pressure drop using water as backflush. A drastic pressure drop was experienced for 8hours: 15min cycle indicating that this cycle cannot be used for long term study. In the case of 120mins: 15mins operation, the rate of pressure increase is lower than for 8hours: 15mins cycle operation but higher when compares to the same air-filtration cycle. Also similar trends was observed for 60mins: 15mins operation. From Figures 5.12., 5.13. and 5.14.it is clear that water cycle operation could not effectively remove the clogging from the membrane because of the rapid pressure increase a after backflush. For example during the 120mins:15mins cycle, the pressure increase in the first step of filtration-backflush operation was three times the initial increase rate and almost 6 times the initial rate in second cycle of operation. This phenomenon can be explain as follows: after filtration the membrane module was removed from the reactor and immersed in a bucket filled with clean water. The backflush was then carried out with either water or air. With water, it was observed that the backflushing water could not wash away the sludge or cake layer formed at the membrane surface. From this observation it appeared as though the sludge layer was cemented to the membrane surface. The water forced the sludge layer to break , creating an opening for the water to flow through without necessarily washing away the sludge layer from the surface.

As with air, the sparging nature of air through the membrane pores into the liquid help to create turbulence or eddies near the membrane wall that washes away the sludge particles formed around the surface. This is why the filtration-air cycle is more efficient since the cake formed at the membrane surface is washed away easily.

In conclusion, due to the inefficiency of the water backflush, the air backwflushing method was used in the long-term study. The increased pressure with time during filtration shows that continuous suction cannot be applied in the long term study. On the other hand 120mins:15mins operation gave more cumulative permeate volume than 60mins:15mins cycle because of the filtration time, although the rate of pressure increase in the former cycle was slightly greater than that of the latter. A cycle of 120mins:15mins was used for long term experimental study on the basis of permeate volume with the average rate of transmembrane pressure increase for each successive cycle operation is about 30 kPa.

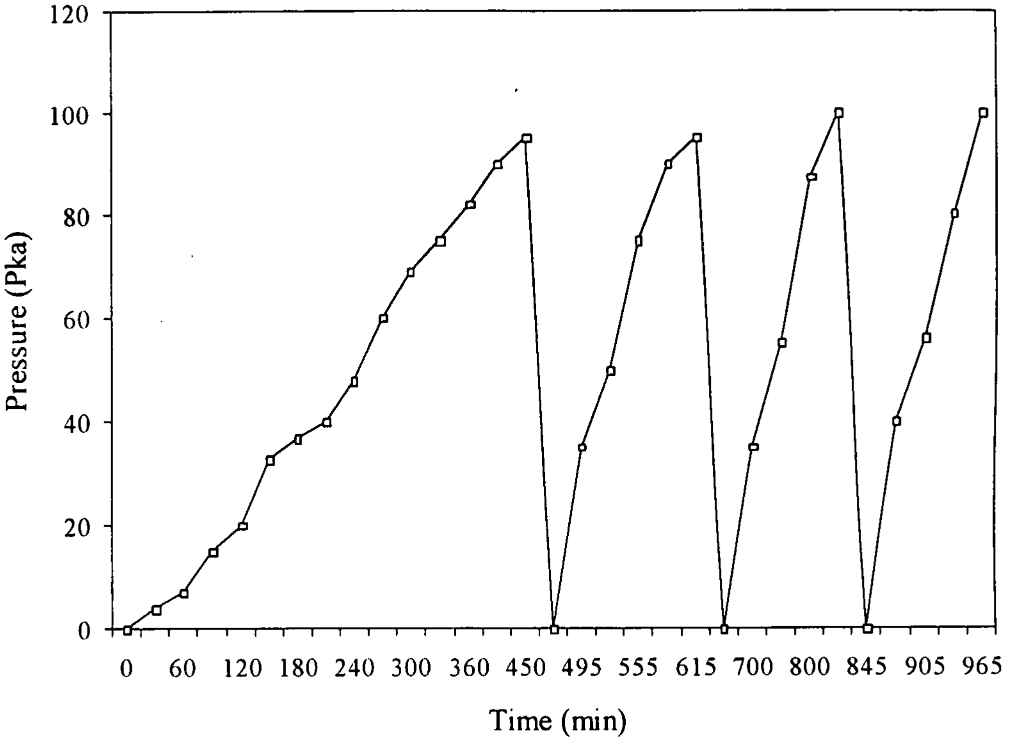


Figure 5.12 Variation of transmembrane pressure with time at 8 hours and 15 minutes cycle operation (water-backflush)

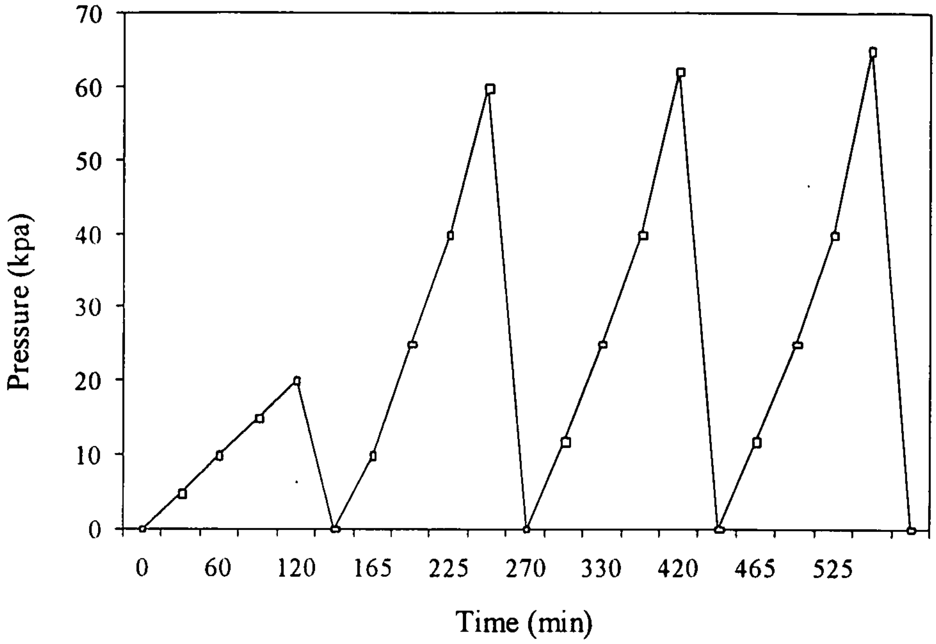


Figure 5.13 Variation of transmembrane pressure with time at 120:15 (water-backflush)

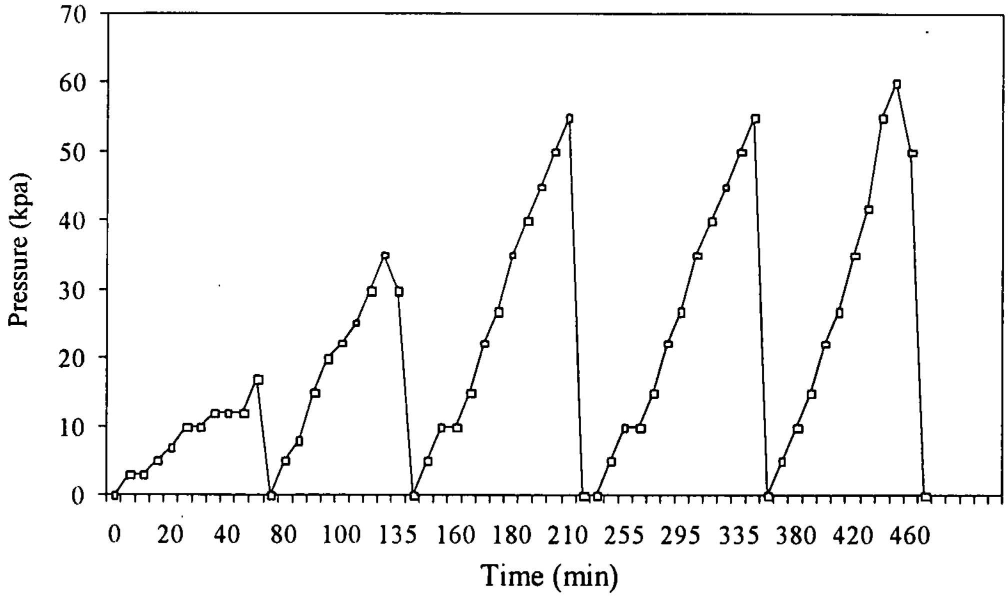


Figure 5.14 Variation of transmembrane pressure with tme at 60:15 (water-backflush)

5.5 Long Term Performance of MBR

5.5.1 Laboratory-scale MBR with Aerobic-Anoxic systems

The results obtained from the laboratory scale studies are depicted in Table A6 (Appendix 6). Table A6 shows COD removal rate in the anoxic, nitrate removal rate in anoxic and nitrification rate in the aerobic reactor. The method of operation used here is similar to that discussed in section 5.3. The the effluent analysis was based on the filtrate from the membrane while the previous analysis was based on effluent from the settling tank. COD removal rate ($U_{COD,AN}$) and nitrate utilization rate ($U_{dN,AN}$) were plotted based on the concept of Argaman and Brenner,(1986) and the data was fitted according to the least square method. The WRC (1986) model could not be used directly because the major components in the influent were not fractionated into various components before use. The prediction of the model (see equation 3.18.) is that the denitrification rate is governed by two rate steps occurring simultaneously. The first step postulates that denitrification is proportional to the concentration of biodegadable COD in the influent while the second step postulates that denitrification is independent of nitrate concentration but are in first order with biomass concentration in the reactor (WRC, 1986).

5.5.2 Determination of COD removal rate coefficient ($U_{COD,AN}$), nitrate removal rate coefficient ($U_{dN,AN}$), and nitrate endogenous rate coefficient (b_N)

Table A6 (Appendix 6) shows the estimated figures of COD removal rate, nitrate removal rate and nitrification rate. The denitrification rates varied over a rather wide range with effluent nitrate concentration. It became evidence that the zero order denitrification rate given by equation 5.3.1a, was inappropriate for this study. Further examination of the data revealed a definite correlation between denitrification and COD concentration in the anoxic reactor. It was concluded that in this system, the denitrification rate was carbon limited and its rate was controlled by the rate of COD removal (Ekama *et al.*, 1999). Figure 5.15. shows the graph of COD removal rate versus effluent

COD concentration. The removal was best described by a first order reaction with non-biodegradable residues as given by equation 3.16. From the slope and intercept of figure 5.15, it was determined that for this wastewater, K_{SH} = 0.064 L/mg.d and S_u = 16.5 mg/L.

The data from this study are in good agreement with that obtained from the activated sludge system (see figure 5.2.). The value of S_u obtained from the activated sludge system was 17.5 mg/L, the difference may be due to experimental error. The average value of these two figures was used for subsequent calculation in this work. The COD utilization rate was found to be proportional to nitrate utilization rate as can be seen in figure 5.16. The Nitrate utilization rate coefficient, as determined by the slope of the best fit was 0.338 mg NO_3-N/mg COD and the endogenous rate determined by the intercept is -0.038 $d^{-1.}$

5.5.3 Determination of maximum specific nitrification rate in MBR

The values of specific nitrification rate and the corresponding effluent ammonia concentration from the membrane is in Appendix 6. The maximum value of the nitrification rate as can be seen in Figure 5.17. is 2.10 d^{-1}. Refer to section 5.3.2 for similar explanation.

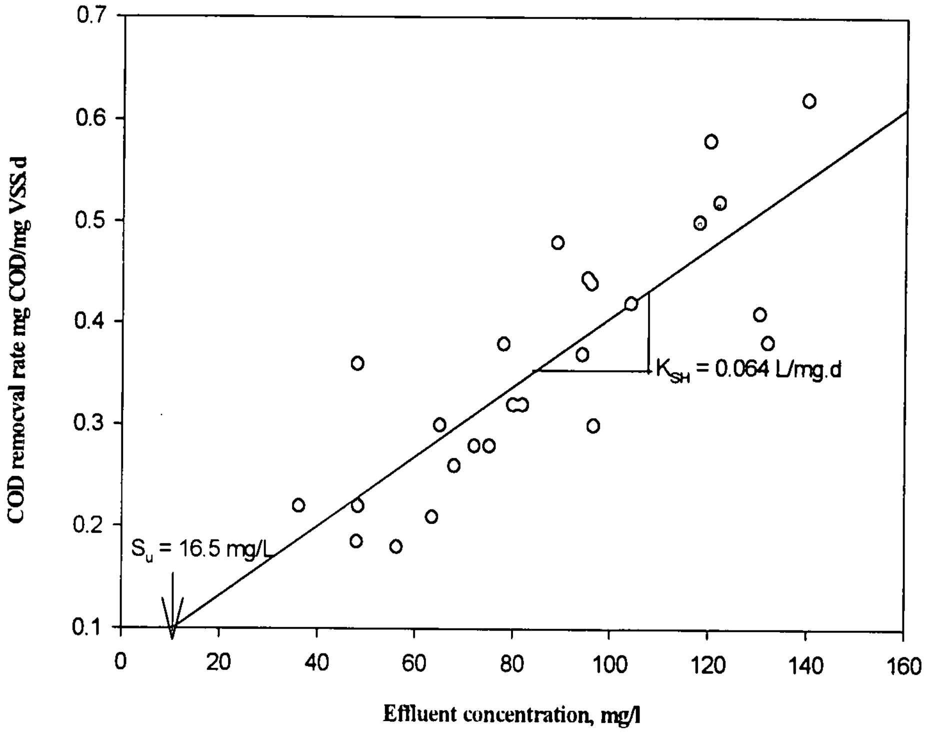


Figure 5.15 Determination of COD removal coefficients

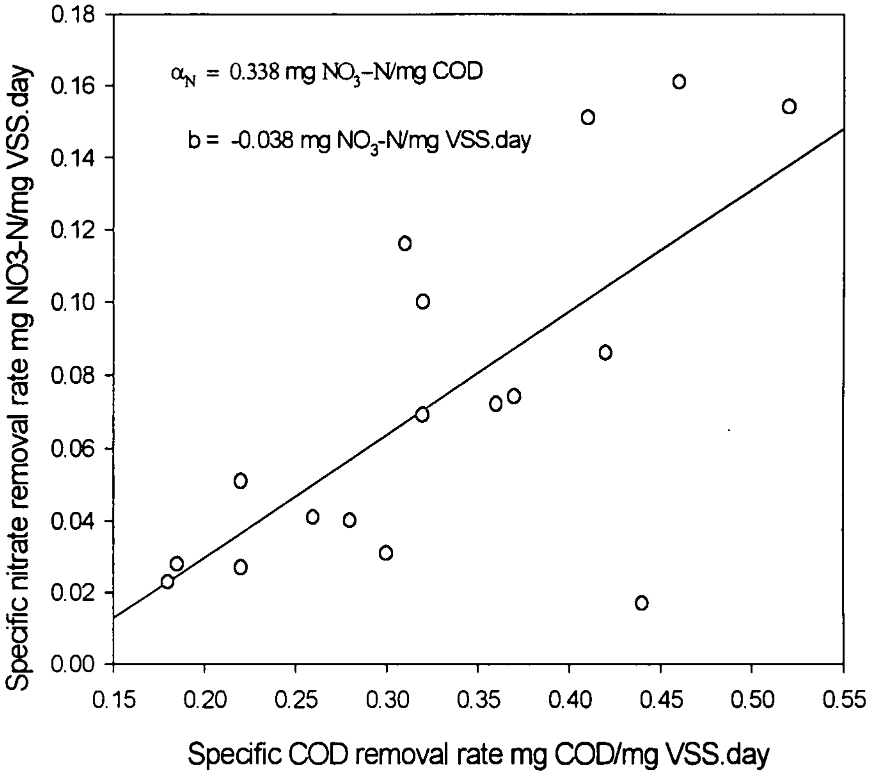


Figure 5.16 Determination of nitrate consumption coefficients

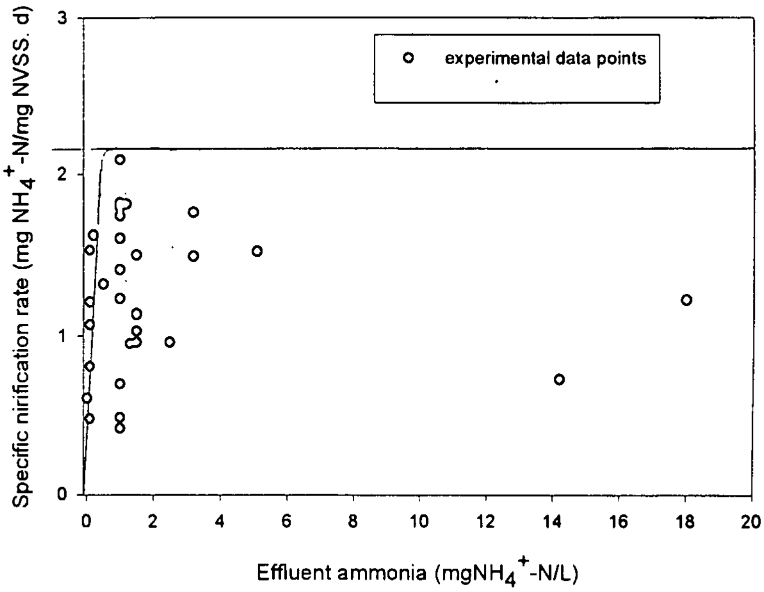


Figure 5.17 Correlation of specific nitrification rate with effluent ammonia

5.5.4 MBR conditions during intermittent aeration

The aerobic side of Figure 4.2. was used for the laboratory scale study with the anoxic reactor removed. The system was operated for 120 minutes of aerating and non-aerating cycle. Table A7 (Appendix 7) shows the results obtained from the cycles of operation. The results presented are daily average of the measured values taken during each cycle. The HRT of the system is 8 hours. The SRT was controlled at 20 days by continuous sludge removal from the bioreactor with all other conditions constant until steady state.

Figure 5.18. shows the cycle time of oxygen concentration during each period of aeration and non-aeration . The dissolved oxygen concentrate was high in the 120 minutes aeration period but dropped to very low values during the 120 minutes of anoxic period.

Figures 5.19. and 5.20. shows COD and ammonia effluent concentrations during each cycle time as determine experimentally. The trends in the cycle of operation for the effluent COD concentration shows that overall COD removal of about 95 % of COD are removed during the whole cycle. The NH_4^+-N was nitrified only during the aerobic period and rose during the anoxic period. The daily average effluent NH_4^+-N concentration obtained from intermittent operation mode is higher than during the 71 days of operation without sludge wasting. A mass balance on N shows that an average of 98 % of NH_4^+-N put into the system was removed as N and 80 % of the influent N was denitrified to N_2. Figure 5.21. shows effluent nitrate during each cycle time.

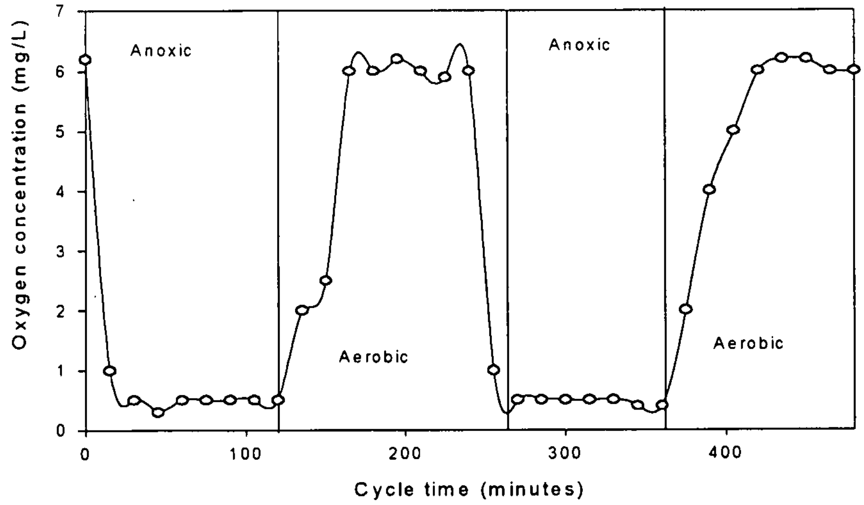


Figure 5.18 - Variation of oxygen conc. within aerobic-anoxic cycle

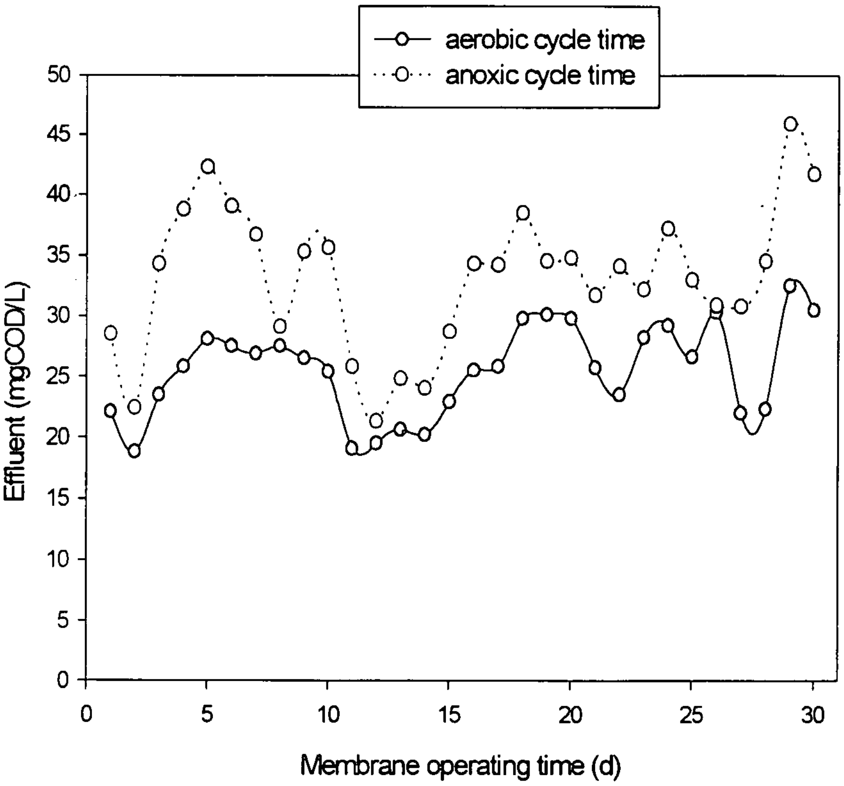


Figure 5.19 - Effluent COD concentration during intermittent aeration

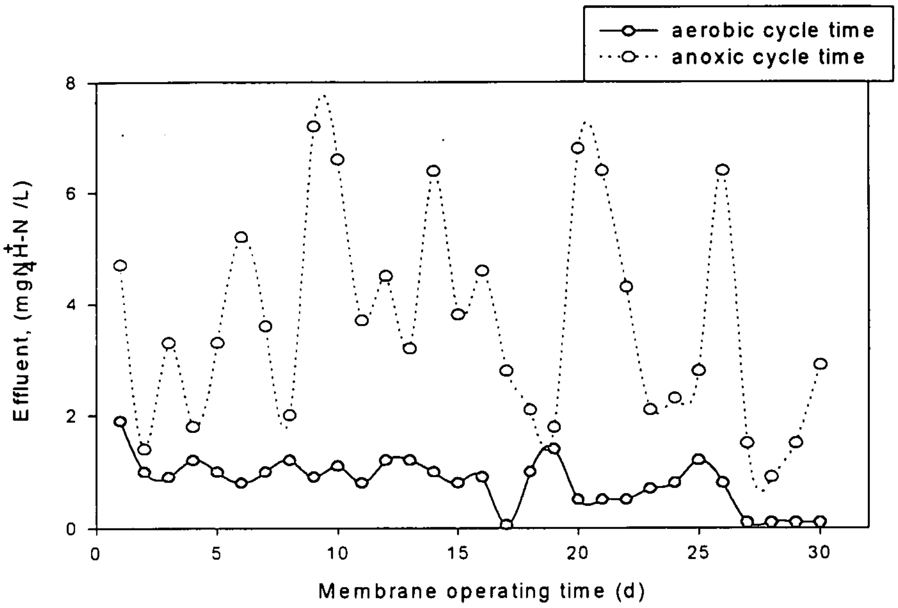


Figure 5.20 - Effluent ammonia concentration during intermittent aeration

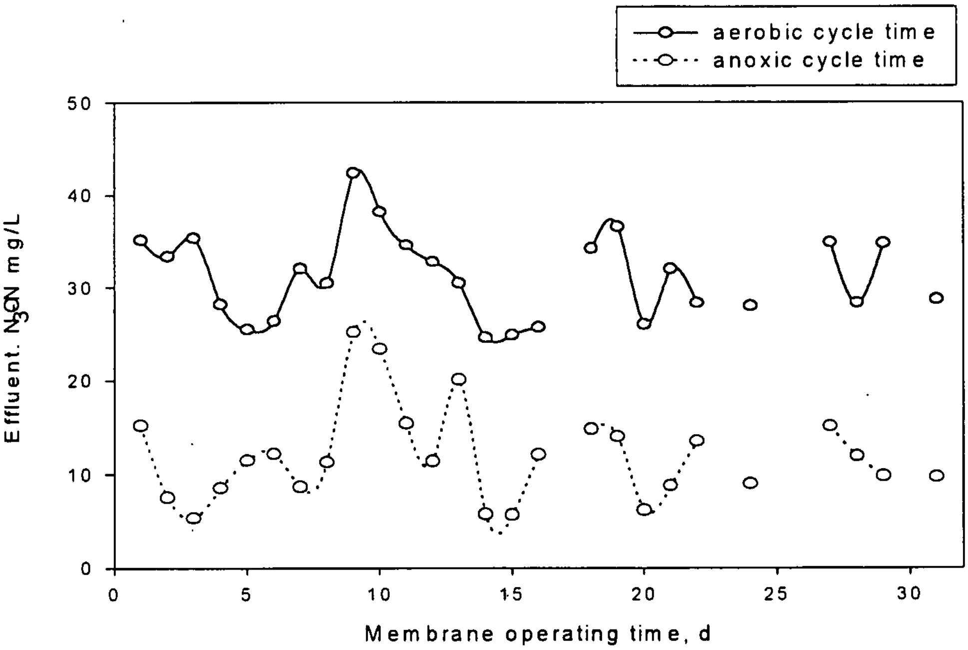


Figure 5.21 - Effluent nitrate concentration during intermittent aeration

5.5.5 Process performance

During the first seventy one days of operation, the influent COD was varied between 400-950 mg/L whereas the effluent values from the system at this period remains at an average of 24 mg/L. The COD removed based on average values was 96.02 %.

After 71 days of operation, ie, during the period of intermittent aeration operation, when the sludge age was maintained at 20 days, the effluent COD observed was almost the same as the effluent from the previous (71 days) operation. The general trends of effluent COD and % COD removed is shown in Figure 5.22. The only significant difference between the two operation is the way in which COD was induced in the different modes of operation. As shown in Figure 5.23., within the 71 days of operation (with aerobic-anoxic system), the organic load was kept constant for 22 days and thereafter was forced to experience a diurnal effect for the rest of the operational mode. The second phase of the operation mode (intermittent aeration), experienced a constant loading effect. The organic load into the system as a whole lies between 0.486-2.283 kgCOD/m^3.d.

Figure 5.24. shows the biological load in the entire period of operation (from 0.078-0.602 kgCOD/kgVSS.d). It is higher during intermittent aeration time. The fluctuating effect experienced from the first seventy one days of operation is due mainly to the changes in MLSS and the applied COD. The MLSS concentration tends to increase with operation time. This means that microorganisms degrade the residual organic matter and synthesis new cells. After about 71 days of operation, MLSS concentration increased by 4.3 times the initial concentration maintaining more than 14.0 gMLSS/L as shown in Figure 5.25.

The shift in operation mode (from aerobic-anoxic systems to intermittent mode) as mentioned in the above discussion was investigated whether any improvement in the amount of nitrogen removal was possible. With the latter method, average effluent ammonia increased slightly and the percent total nitrogen removed decreased from 85.5 % to 79.0 %. The trends in the influent , effluent, % nitrification and % nitrogen denitrified is shown in Figures 5.26. and 5.27. A closed evaluation of figure 5.27. shows that % nitrification is higher with the initial operation mode used in the first 71 days of operation. On the overall more than 97 % nitrification is possible.

Figure 5.28. shows the variation of average daily permeate flux and transmembrane pressure during the membrane operating period. The flux obtained from this study lies between 22-92 L/m^2-h, with the average flux value being 57 L/m^2-h for the entire period of operation. Regular flux decline observed during operation was improved through regular backflush with air and chemical cleaning. The transmembrane pressure did not as whole exceed 55 kPa due to the frequent cleaning of the membrane surface.

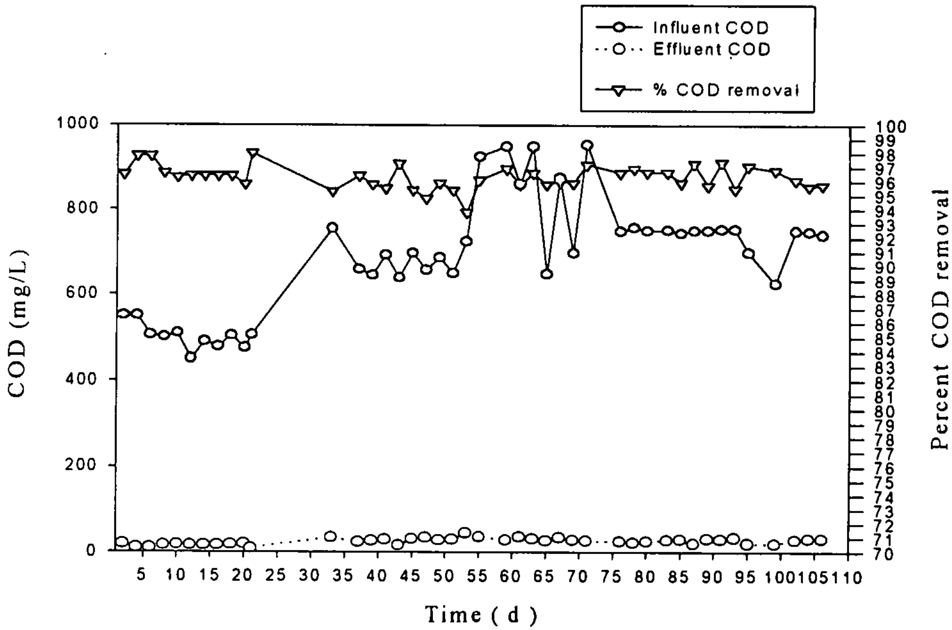


Figure 5.22 Daily observation COD removal by hybrid MBR system.

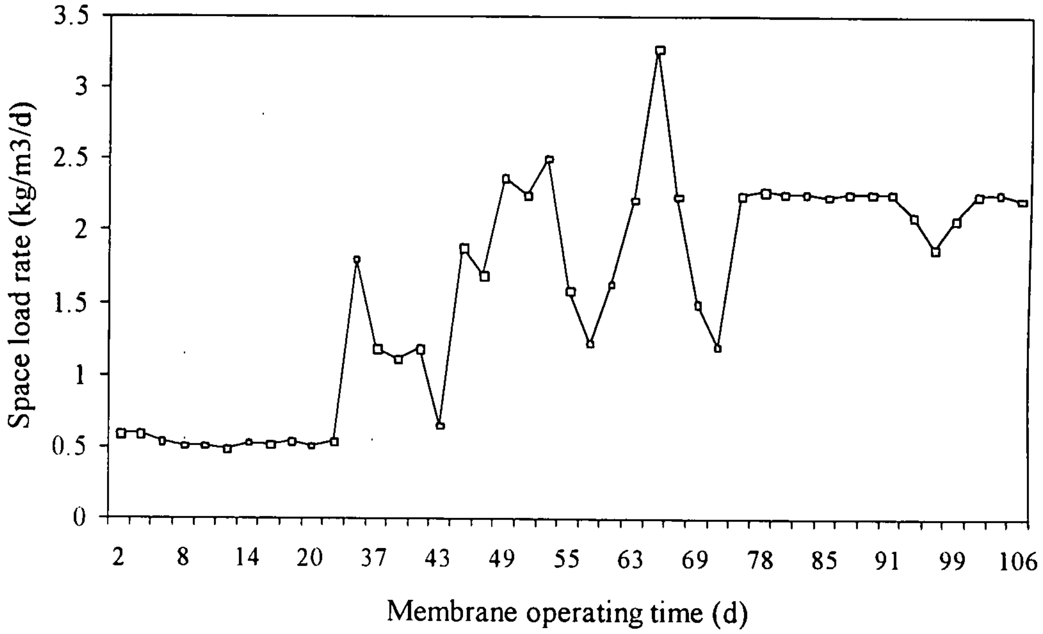


Figure 5.23 - Variation of space load rate Vs Time

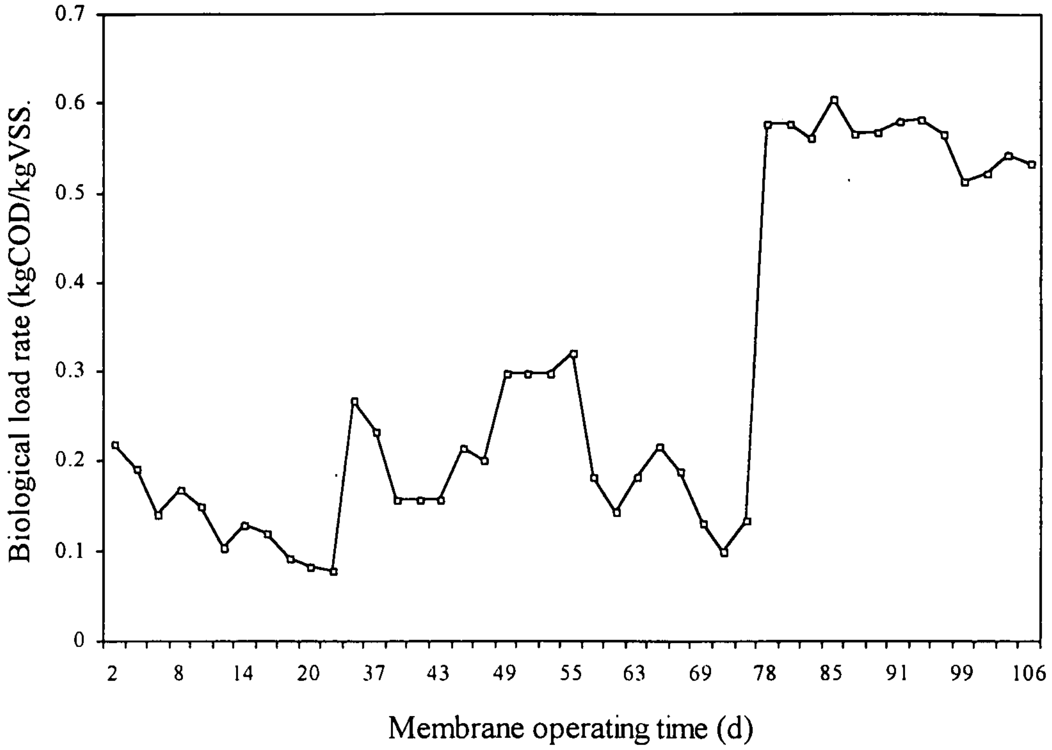


Figure 5.24 - Variation of Biological Load Rate Vs Time

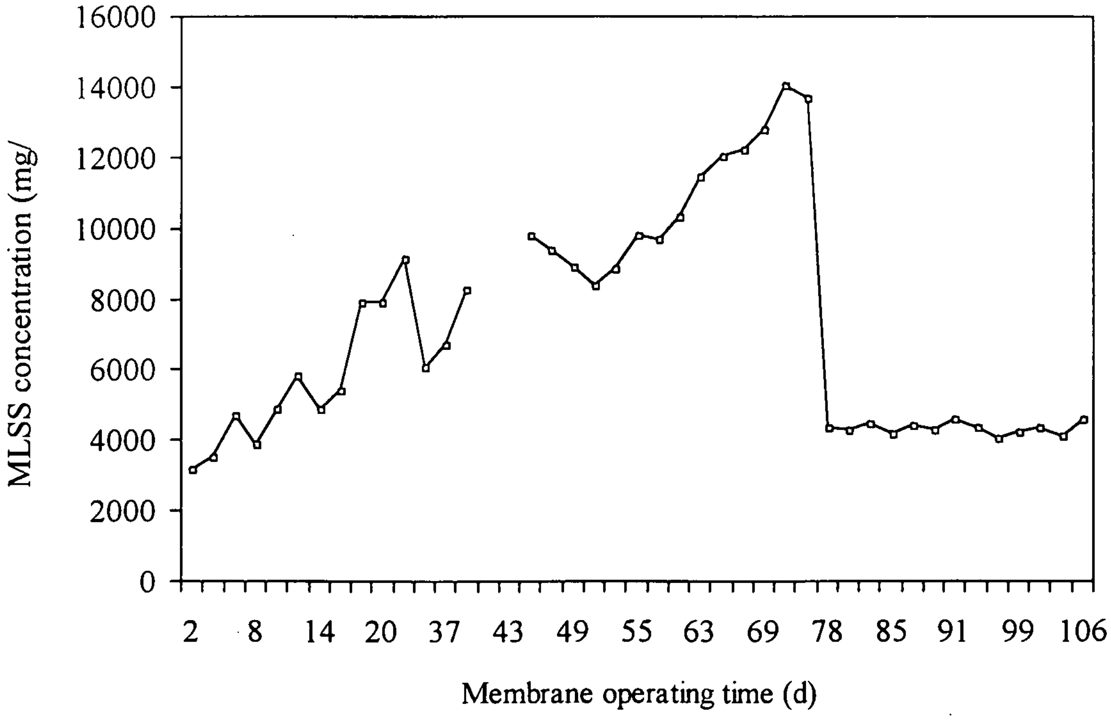


Figure 5.25 Variation of MLSS Concentration Vs Time

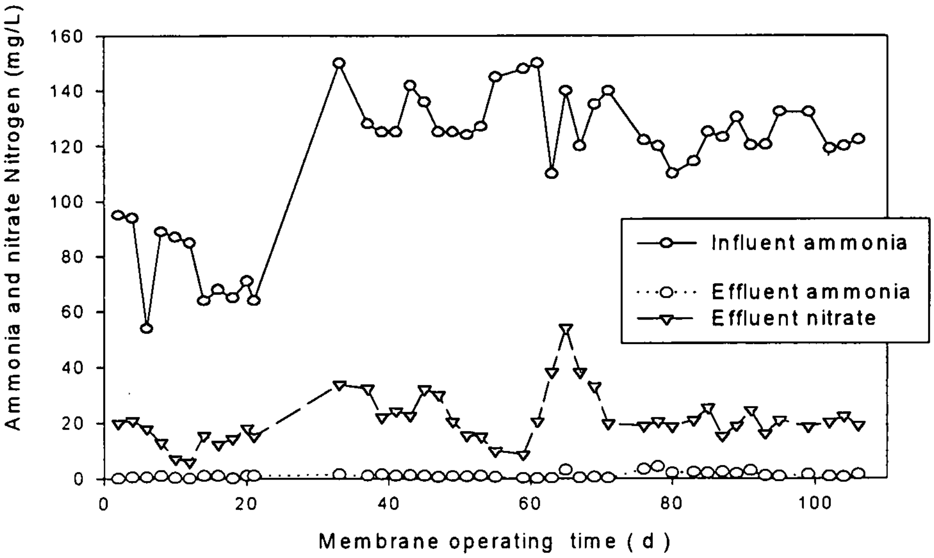


Figure 5.26 Variation of influent and effluent ammonia and nitrate Vs time

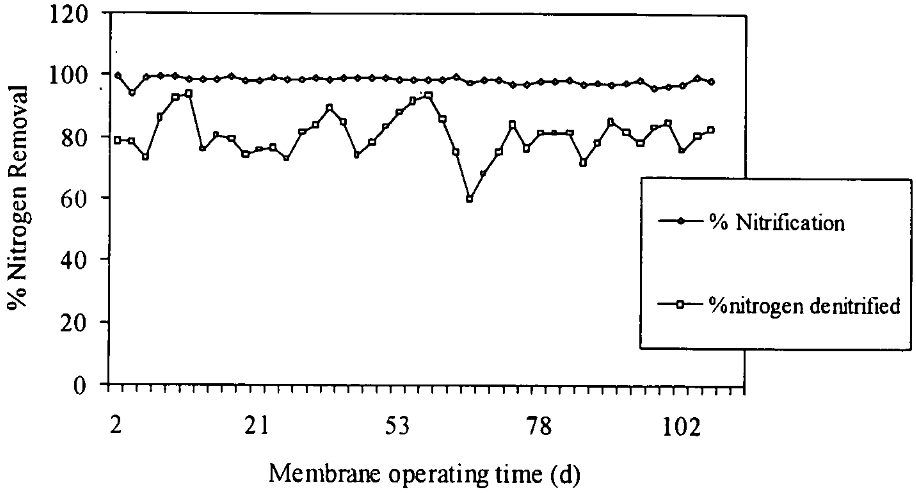


Figure 5.27 % Nitrogen removed Vs Time

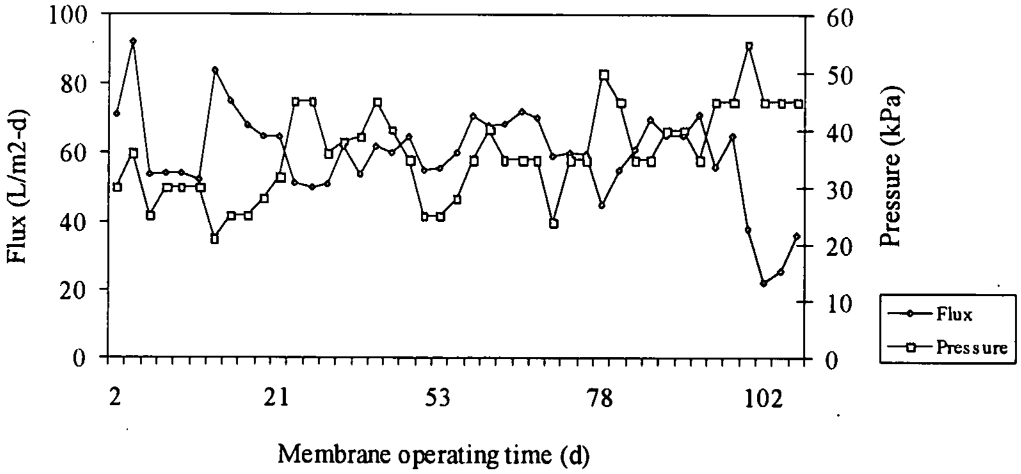


Figure 5.28 Variation of permeate flux and transmembrane pressure along membrane operating time

5.6 DISCUSSION

The kinetic coefficients obtained from this study are similar to those reported by Sundstrom and Klei (1979) although the effluent substrate used was described as milk waste which is similar in characteristics to dairy effluent. Table 5.4. shows the kinetic coefficients given by Sundstrom and Klei (1979) and those obtained from this study.

The plot of the data obtained from this study as shown in Figures 5.1., 5.2., and 5.3., are in fact based on the simplified model of equation 3.5 and 3.8 instead of equation 3.12. Therefore, it would have been insufficient to assume that the constant coefficients had been accurately determined on the strength of the best fit of the theoretical plot. The data processing procedures up till the use of equation 3.12. was only a simplified method of finding the constants that will yield good fit.

The only significant difference between the results obtained from this study and that reported by Sundstrom and Klei (1979) is the value of maximum substrate utilization rate. The conditions under which this study was carried out was not reported because the constants k, K_S and k_d are temperature dependent. It is unlikely that the wide difference in the value of k is attributed to temperature alone and the slight variance in effluent was presumed to play an important role.

The process procedures of section 3.2. model application up till equation 3.12 was only successfully applied to the study performed under conventional methods. The effect of different sludge age on MBR performance was not investigated. It was however not possible to determine the kinetic coefficients ie Y, K_S, k, and k_d from the data obtained from the MBR. The COD and the nitrate removal rate from conventional and MBR were evaluated as reported in Appendices 4 and 6 . Plots are shown in Figures 5.4. and 5.13. The nitrate removal rate coefficient obtained from the MBR is higher than that obtained from conventional methods. This may be attributed to higher concentration of VSS in the MBR as a result of long sludge ages.

Figure 5.6. and 5.17. show the plotted data of maximum specific nitrification rate obtained from

conventional methods and the MBR , respectively. Within the ranges of sludge age used for the conventional method, the maximum specific nitrification rate determined was approximately 3.0 d^{-1}. Close examination of Figure 5.6., shows that the saturation coefficient was difficult to determine due to the nature of the curve. This is evident in the the non-linearity of the data in the vicinity of 0.09 mg/L effluent ammonia concentration. For design purposes the maximum nitrification value obtained is used to determine the minimum sludge age required. A similar approach was adapted to results obtained from MBR. The maximum nitrification rate in the system was 2.10 d^{-1}. This implies that the nitrifying bacteria had a much higher growth rate in the MBR than during conventional treatment. Note that the maximum nitrification rate determined from both systems may not be too accurate as the correct autotrophic and heterotrophic fraction was not known. The autotroph fraction used is based on the influent COD/NH_4^+-N ratio.

Table 5.4 Comparison of kinetic coefficients

	Y (mg VSS)/(mg COD)	k_d (d^{-1})	k (d^{-1})	K_S (mg/L)	Basis
Sundstrom and Klei (1979)	0.157	0.012	0.051	20.85	COD
Laboratory AS system	0.204	0.013	3.70	17.4	COD

Comparison of effluent characteristics on the basis of turbidity and suspended solids shows that the effluent turbidity and suspended solids from the conventional activated sludge system was higher than those obtained from the MBR. As shown in Figure 5.29. the average effluent turbidity from the microfilter used in this study is 0.92 NTU which is well above the 0.5 NTU recommended standard. The effluent suspended solids from the MBR in most cases were undeterminable. Average suspended solids from conventional system was 2.7 mg/l (Table 5.5.). The high value of suspended solids from the convention system may be attributed to pin point floc formation during system operation. The pin point floc was as a results of sludge bulking. The cause of sludge bulking was not immediately known, but in most cases it was eliminated by reducing sludge wastage and internal recycle rates.

The long SRT with which the MBR was operated minimised excess sludge production. This is not unique as any process that achieves long SRT should yield similar results including conventional, AS systems. It is noted that, however, a membrane bioreactor easily satisfies a long solids retention time condition even at high organic loading without deteriorating the effluent quality nor losing the sludge simply because of membrane separation, by which a high sludge concentration is always maintained. This seem very difficult to be obtained by conventional AS systems utilizing gravity settling for solid-liquid separation.

Table 5.5 Analysis of effluent from MBR and AS systems

	AS system	MBR
Turbidity (NTU)	6.01	0.90
Suspended solids (mg/L)	2.09	nm

From theoretical point of view, the effect of long solids retention time is expressed using the following equation (Yamamoto *et al.*, 1989; Metcalf and Eddy, 1994; Fan *et al.*, 1996):

$$Y_{obs} = \frac{Y}{(1 + k_d \theta_c)}$$

In this experimental work,the value of Y was not determined using the MBR systems because two sludge ages ie 100 and 20 days was considered. If for instance the value of Y and k_d are 0.6 mgVSS/mgCOD and 0.05 d^{-1} respectively which is typical of MBR systems (Yamamoto *et al.*, 1989), the longer the SRT, the lower Y_{obs} will be. For example at SRT = 100 d, $Y_{obs} = 0.10$, which implies that only 10 % of influent COD is converted to activated sludge, or in other words, 90 % of influent COD is removed through oxidation. If STR = ∞, $Y_{obs} = 0$ indicating that influent COD is completely oxidised. Ignoring the accumulation of inorganic suspended solids, very long SRT are

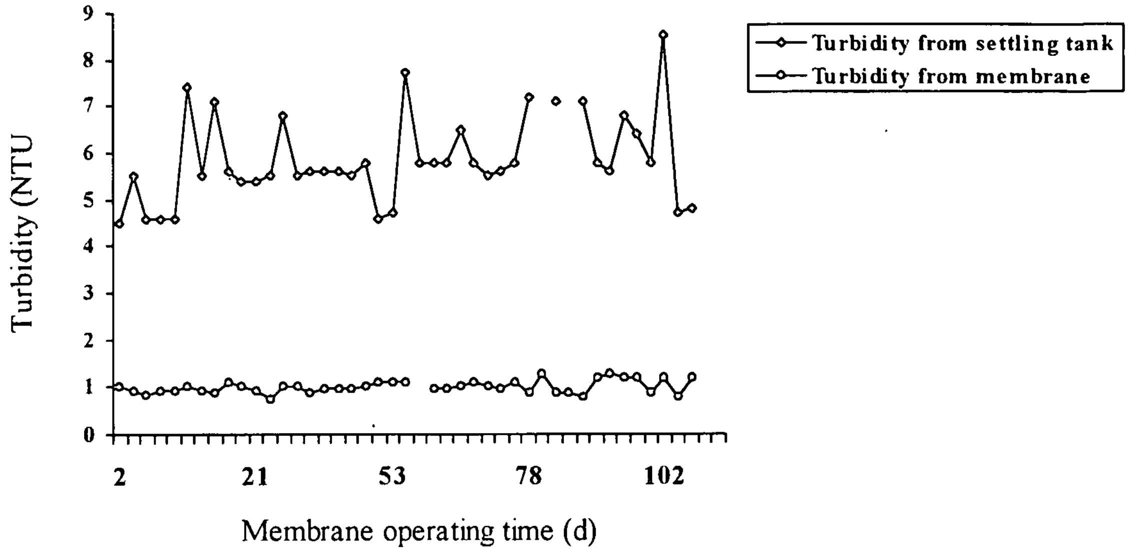


Figure 5.29 Effluent turbidity from membrane and settling tank

possible with MBR under reasonable volumetric loading rates, but impossible with conventional suspended growth systems, because of the withdrawal of suspended solids with effluent and bad settling properties at high concentration. A study done on MBR's with synthetic waste suggested that sludge removal is not necessary for biological reasons (Chiechaisri *et al.*, 1992). Sludge wasting is required in raw wastewater due to accumulation of inorganic suspended solids.

During the long performance of MBR, problems were experienced with foam generation. Causes were not investigated but attempts were made to control foaming throughout the experimental work. The foaming always occurred periodically, frequently when organic loading was increased, but often for no obvious reason.

The most reasonable explanation without positive proof will be similar to that given by Nell and Kafaar (1995). They suggested that foam is generated by vigorous mixing action through aeration in the aerobic reactor which affects the biomass causing it to release polymeric substances which promote foam formation. Foaming always lead to loss of biomass through entrainment especially at biomass concentrations exceeding 10000 mgMLSS/L. The effect was corrected by covering the aerobic reactor with a lid.

Membrane fouling appears to be a problem that has to be managed by regular cleaning and air backflush. The average flux experienced during the first 71 days was higher than during the last 36 days of the investigation (Figure 5.25.). Regular cleaning was done during the last 36 days of experimentation by soaking the membrane in a container filled with a sodium hypochlorite solution.

5.7 CONCLUSION

This study proved that the membrane air backflush/aeration process plays an integral part in the improvement of flux by removing external deposits on the membrane. Defouling of the membrane prevents the compaction of the cake layer and reduces the internal pore clogging of the membrane. Although the cyclic operation with air backflush could not completely remove the clogging, the flux was improved up to 200 % as compared to continuous operation.

The membrane cleaning process adopted in this study was found to be adequate in removing mainly external membrane resistance. Nevertheless, it is necessary to introduce a chemical cleaning procedure for complete removal of external and internal deposits, caused by macromolecular adsorption.

COD removal with MBR treatment was greater than 96.02 % with effluent concentrations below an average value of 28 mg/L. Because of the high SRT in the bioreactor, significant COD removal variance was recorded for the various organic loads investigated. Effluent quality in terms of suspended solids and nitrogen was extremely good.

From the process efficiency point of view, the MBR produced better quality effluent than conventional biological treatment processes and could totally eliminate the common operational problem of sludge bulking. However, further purification is needed since the effluent produced from the MBR is well above the recommended 0.5 NTU turbidity standard.

5.8 RECOMMENDATIONS

Attention should be given to process stability. In this study, 106 stable operation could be achieved through intermittent suction with low pressure. However, it is still uncertain whether the process remains stable for a longer periods of operation as the membrane has the inherent drawback of severe clogging at the surface. It requires improvement and continuous suction should be implemented.

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APPENDICES

APPENDIX 1

Calculation of permeate flux and statistical formulae

Permeate flux :

$$J(\frac{L}{m^2h})=\frac{Permeate\text{-}flowrate}{Area}=\frac{3600}{1}(\frac{L}{s.m^2}.\frac{s}{h})$$

$$=\frac{V_{perm}}{t_{perm}}.\frac{1}{2\Pi RL}.\frac{3600}{1}$$

$$=\frac{10\times10^{-3}}{t_{perm}}.\frac{1}{2\Pi(0.025)(0.125)}.\frac{3600}{1}$$

$$=\frac{1833.46}{t_{perm}}$$

Statistical Formulae :

$$Average=\bar{x}=\sum_{i=1}^{n}\frac{x_i}{n}$$

$$Standard\text{-}deviation=\sigma=[\frac{\sum_{i=1}^{n}(x_i-\bar{x})}{n-1}]^{\frac{1}{2}}$$

n = number of data points

x = data point.

APPENDIX 2

Details of analytical procedures:

The literature references of all analyses are given, and any modifications of the methods are described in full.

- Chemical oxygen demand (COD)

  Spectroquant analysis method using SQ 118 Merck photometer for all ranges of COD tested.

- Saline and free ammonia

  Spectroquant analysis method using SQ 118 Merck photometer for all ranges of ammonia tested.

- Nitrate

  Spectroquant analysis method using SQ 118 Merck photometer for all ranges of nitrate tested.

- MLSS concentration (standard method

Laboratory 20ml samples were taken from aerobic reactor and filtered through a 45um glass fibre. The residue retained on the filter is dried at 103 °C for least one hour and then weighed.

- VSS concentration

The dried solids resulting from the MLSS analysis detailed above were ignited at 550 °C for about 30 minutes and weighed. The loss in weight was recorded as VSS.

APPENDIX 3

Effluent characteristics from dairy industry (Average value from July 1998 - June 1999)

Date	COD mg/L	pH	conductivity (ms/m)	soap oil and grease(mg/L)	Settleable solids (ml/L)
July 1998	3800	10.2	200.0	120	0.8
August 1998	4000	10.0	140.0	110	0.5
September 1998	4000	10.8	210.0	180	0.4
October 1998	4000	11.0	-	-	-
November 1998	3800	11.0	280.0	180	-
December 1998	3500	10.6	210.0	100	0.3
January 1999	3900	10.2	180.0	100	0.4
February 1999	3800	9.0	140.0	100	0.5
March 1999	3800	9.8	160.0	100	0.1
April 1999	4200	10.4	205.0	105	0.2
May 1999	4100	10.4	190.0	98	0.2
June 1999	4200	10.0	300.0	100	8.0

APPENDIX 4

Results obtained from laboratory pilot scale aerobic-anoxic systems. The complete set of data recorded are presented in Table A4.1-A4.4

Table A4.1

Date	Feed	COD, mg/L			NH4+-N, mg/l			NO3-N,mg/l			MLVSS		Temp	Sludge	Turbidity	S/Solid mg/L	
	L/d	Inf.	Eff.	Anx.Eff	Inf.	Eff.	Anx.Eff	Inf.	Eff.	Anx.Eff	mg/L	pH	oC	age	(NTU)	influent	effluent
		S0	S3	S2	A0	A3	A2	N0	N3	N2				(d)			
26/02/99	69.1	645	23.5	124.0	156	0.5	62.2	0.1	31.2	29.3	3550	7.9	22.5	18.5	4.5	110	2.24
01/03/99	69.1	650	39.2	128.1	148	0.2	20.1	0.0	18.2	15.5	3640	7.5	22.1	18.5	5.5	-	-
04/03/99	69.1	650	41.0	135.0	148	0.1	102.5	0.0	108.1	65.5	3621	8.1	22.1	18.5	4.6	98	2.08
07/03/99	69.3	650	31.5	131.0	150	1.0	55.2	1.0	35.2	15.1	3452	7.9	22.5	18.5	-	-	-
10/03/99	69.1	680	35.2	148.1	147	6.0	35.8	1.0	29.8	10.8	3528	7.9	22.2	18.5	-	45	1.98
13/03/99	69.1	645	29.8	125.0	148	0.7	35.5	0.1	8.5	7.2	3650	7.5	22.5	18.5	7.4	48	-
16/03/99	69.2	680	35.5	145.0	148	0.5	22.5	0.1	46.6	23.1	3298	8.1	22.5	18.5	5.5	68	2.06
19/03/99	69.1	654	34.8	142.3	235	0.6	30.0	0.1	20.1	12.0	3500	7.6	22.4	18.5	7.4	68	2.6
22/03/99	69.1	651	29.8	138.1	250	1.0	19.5	0.0	14.5	9.2	3486	7.2	25.1	18.5	-	98	1.98
25/03/99	69.1	687	33.2	145.0	185	0.2	24.4	0.1	17.0	10.2	3582	7.8	24.2	18.5	7.1	-	-
Average											3530.7				6.0	76.43	2.16

Table A4.2 Laboratory pilot scale results for aerobic-anioxic systems

Date	Feed L/d	COD, mg/L Inf. So	COD, mg/L Eff. S3	COD, mg/L Anx.Eff S2	NH4+-N, mg/l Inf. Ao	NH4+-N, mg/l Eff. A3	NH4+-N, mg/l Anx.Eff A2	NO3-N,mg/l Inf. No	NO3-N,mg/l Eff. N3	NO3-N,mg/l Anx.Eff N2	MLVSS mg/l	pH	Temp oC	Sludge age (d)	Turbidity (NTU)	S/Solid mg/L influent	S/Solid mg/L effluent
30/03/99	69.1	650	30.2	150.3	165	0.5	80.0	0.0	35.0	25.0	2975	7.4	22.2	15	5.6	98	2.58
02/04/99	69.1	650	39.1	148.1	180	0.7	45.2	1.0	25.5	19.2	2985	8.2	22.5	15	5.4	101	2.57
05/04/99	69.1	652	42.0	156.0	165	2.1	57.5	1.1	11.0	9.9	3025	7.2	23.2	15	-	100	3.01
08/04/99	69.1	645	18.2	150.2	164	0.5	85.5	1.0	40.8	24.4	2972	7.5	22.5	15	5.5	89	2.58
11/04/99	69.1	750	28.5	165.0	167	2.1	65.2	0.1	23.2	18.0	3200	8.2	23.1	15	6.8	-	-
13/04/99	69.1	575	32.6	148.2	145	0.2	65.0	1.0	39.6	23.3	2874	7.5	22.2	15	5.5	-	-
16/04/99	69.1	700	35.0	168.0	166	4.4	90.0	1.2	30.2	33.5	2980	8.1	23.5	15	-	-	-
18/04/99	69.1	698	36.0	162.2	158	1.0	65.4	1.2	20.8	11.4	3000	7.8	22.5	15	5.6	96	1.98
20/04/99	69.1	655	35.8	95.0	168	0.4	55.5	0.1	19.4	11.4	3234	8.2	22.5	15	5.6	98	2.35
22/04/99	69.1	650	29.5	100.0	160	0.7	95.0	0.1	25.2	19.5	3256	7.5	23.5	15	-	96	-
Average											3050.1				5.7	96.86	2.51

Table A4.3 Laboratory pilot scale results for aerobic-anioxic systems

Date	Feed	COD, mg/L			NH4+-N, mg/l			NO3-N,mg/l			MLVSS	pH	Temp	Sludge	Turbidity	S/Solid mg/L	
	L/d	Inf.	Eff.	Anx.Eff	Inf.	Eff.	Anx.Eff	Inf.	Eff.	Anx.Eff	mg/l		oC	age			
		So	S3	S2	Ao	A3	A2	No	N3	N2				(d)	(NTU)	influent	effluent
26/04/99	69.1	650	49.5	180.0	165	4.5	95.1	2.0	40.2	25.8	2519	7.8	22.5	10.2	5.5	95	1.54
28/04/99	69.1	750	40.2	192.0	148	1.1	65.4	1.0	60.0	45.7	2923	8.4	23.4	10.2	5.8	90	2.01
30/04/99	69.1	650	35.0	200.2	150	0.8	101.0	0.1	18.2	22.5	2492	7.5	23.5	10.2	-	-	-
02/05/99	69.1	650	35.4	184.0	169	1.0	100.2	0.1	20.4	11.7	2530	7.8	22.5	10.2	-	-	-
04/05/99	69.1	667	35.0	182.0	145	1.5	98.2	1.0	35.6	26.9	2598	7.5	22.5	10.2	7.1	56	1.98
06/05/99	69.1	625	37.5	175.3	146	1.0	65.8	0.1	18.5	10.8	2455	7.8	22.6	10.2	5.8	56	1.98
08/05/99	69.1	598	35.2	170.9	250	30.1	65.5	1.0	115.5	94.3	2478	8.1	22.1	10.2	5.8	45	1.96
10/05/99	69.1	650	29.7	200.7	200	2.5	98.4	0.1	30.5	29.0	2395	7.4	22.5	10.2	-	-	-
12/05/99	69.1	655	35.0	168.0	175	0.4	65.5	1.0	25.2	22.2	2563	8.1	23.4	10.2	6.5	56	0.98
14/05/99	69.1	680	37.5	120.1	168	1.0	95.2	0.1	45.0	38.2	2871	7.9	22.4	10.2	5.8	-	-
Average											2582.4				6.04	66.33	1.74

Table A4.4 Laboratory pilot scale results for aerobic-anioxic systems

Date	Feed	COD, mg/L			NH4+-N, mg/l			NO3-N,mg/l			MLVSS		Temp	Sludge	Turbidity	S/Solid mg/L	
	L/d	Inf. So	Eff. S3	Anx.Eff S2	Inf. Ao	Eff. A3	Anx.Eff A2	Inf. No	Eff. N3	Anx.Eff N2	mg/l	pH	oC	age (d)	(NTU)	influent	effluent
18/05/99	69.1	685	30.5	200.0	148	25.1	101.0	1.0	45.2	40.1	2225	8.1	22.5	7.2	-	-	-
20/05/99	69.1	650	45.1	192	151	117.0	130.0	2.0	10.0	68.2	2502	7.6	22.3	7.2	5.6	56	2.02
22/05/99	69.1	680	25.5	204.1	150	45.2	95.0	1.0	55.5	45.5	2290	7.5	22.3	7.2	5.8	58	2.35
24/05/99	69.1	650	38.2	147	145	0.8	35.0	0.1	20.5	19.6	2250	7.9	22.5	7.2	7.2	45	1.98
28/05/99	69.1	655	35.1	183	145	1.5	45.4	0.1	65.8	54.2	2314	8.1	21.5	7.2	-	-	-
30/05/99	69.1	680	30.5	183	155	5.5	68.8	1.0	30.6	19.5	2450	7.5	22.4	7.2	7.1	46	0.95
02/06/99	69.1	599	30.3	155.9	145	15.2	101.0	1.2	25.8	20.2	2326	7.6	22.5	7.2	-	-	-
04/06/99	69.1	650	35.1	197.8	145	55.5	99.5	1.0	45.2	30.6	2568	7.1	21.6	7.2	7.1	65	2.5
06/06/99	69.1	655	39.6	200.0	148	5.5	46.1	0.1	15.8	11.5	2634	7.5	21.5	7.2	5.8	65	2.05
08/06/99	69.1	680	25.5	145.0	149	8.1	55.6	0.1	25.8	20.5	2513	7.5	22.5	7.2	5.6	56	1.98
Average											2407.2				6.3	55.86	1.98

APPENDIX 5

Results obtained from laboratory pilot plant aerobic digestion.
The complete set of data recorded during the steady state period at various sludge ages is presented in Table A5.1-A5.4

Table A5.1

Date	Feed L/d	COD, mg/L Inf.	COD, mg/L Eff.	NH4+-N, mg/L Inf.	NH4+-N, mg/L Eff.	MLSS mg/L	M LVSS mg/L	pH	Temp oC	Sludge age day
12/06/1999	69.1	470	20.8	180	135	4692	3225	7.8	24.5	14.29
15/06/99	69.1	454	17.0	98	30.2	4600	3129	8.2	25.1	14.29
18/06/99	69.1	450	19.8	54.5	0.29	4874	3312	7.2	22.3	14.29
21/06/99	69.1	498	19.2	54.0	0.29	4576	3251	7.1	24.3	14.29
24/06/99	69.1	424	20.1	54.0	0.41	4860	3222	8.02	24.1	14.29
27/06/99	69.1	458	19.8	51.2	0.17	4682	3198	7.5	22.4	14.29
30/06/99	69.1	481	18.5	49.8	1.2	4652	3218	7.5	24.8	14.29
02/07/1999	69.1	400	19.3	54.0	4.4	4770	3223	7.4	23.1	14.29
Average						4713.25	3222.25			
STD							50.9783			

Table A5.2

Date	Feed	COD, mg/L		NH4+-N, mg/l		MLSS	MVSS		Temp	Sludge age
	L/d	Inf.	Eff.	Inf.	Eff.	mg//	mg/l	pH	oC	day
06/07/1999	69.1	459	18.5	150	45.0	3830	2850	6.9	24.2	11.76
08/07/1999	69.1	450	19.5	145	29.0	4100	2861	7.1	23.2	11.76
10/07/1999	69.1	465	22.0	65	1.5	3863	2848	6.9	23.1	11.76
13/07/99	69.1	485	22.2	56	0.5	3710	2789	7.4	24.5	11.76
17/07/99	69.1	460	21.0	54	0.6	3668	2910	7.2	23.8	11.76
19/07/99	69.1	455	19.2	56	1.0	3664	2845	7.5	25	11.76
21/07/99	69.1	450	20.5	54.3	0.8	4012	2792	7.2	23.1	11.76
23/07/99	69.1	469	19.8	59	0.5	3692	2954	6.9	23.2	11.76
						3817.38	2856.13			
							55.2279			

Table A5.3

Date	Feed L/d	COD, mg/L Inf.	Eff.	NH4+-N, mg/l Inf.	Eff.	MLSS mg//	MVSS mg/l	pH	Temp oC	Sludge age day
27/07/99	69.1	452	21.4	65	1.1	3256	2388	6.9	23.3	10.52
29/07/99	69.1	457	19.2	63	0.2	2955	2375	7.2	22.8	10.52
01/08/1999	69.1	490	18.5	59	0.5	3422	2389	7.2	22.9	10.52
04/08/1999	69.1	485	18.6	63	0.4	3503	2299	7.6	23.1	10.52
07/08/1999	69.1	450	19.8	61	0.8	3347	2410	7.5	23.3	10.52
09/08/1999	69.1	455	20.0	98	10.1	3162	2391	7.01	23.2	10.52
11/08/1999	69.1	456	19.2	158	60.8	3449	2387	7.9	22.7	10.52
						3299.14	2377			
							35.9119			

Table A5.4

Date	Feed L/d	COD, mg/L Inf.	Eff.	NH4+-N, mg/l Inf.	Eff.	MLSS mg/l	MVSS mg/l	pH	Temp oC	Sludge age day
15/08/99	69.1	450	17.5	65.0	5.5	2821	1725	7.5	22.9	7.14
17/08/99	69.1	462	19.2	75.6	15.1	2614	1719	7.3	23.2	7.14
19/08/99	69.1	460	18.9	68.1	1.0	2696	1698	7.01	23.2	7.14
21/08/99	69.1	450	17.5	65.0	0.8	3163	1730	7.5	23.3	7.14
23/08/99	69.1	460	18.0	65.2	10.2	2819	1598	7.5	24.1	7.14
25/08/99	69.1	465	20.5	59.8	8.4	2780	1734	7.8	22.8	7.14
26/08/99	69.1	454	11	64.2	2.1	2618	1725	7.4	22.6	7.14
						2787.29	1704.14			
							48.2266			

APPENDIX 6

Results obtaied from MBR with anoxic-aerobic reactor systems

Table A6-Experimental data collected from MBR with aerobic-anoxic systems

nm* = not measurable

	Feed	COD, mg/L			NH4+-N, mg/l			NO3-N,mg/l				g/L		mg/L	mg/L	oC	
Date	L/d	Inf.	Eff.	Anx.Eff	Inf.	Eff.	Anx.Eff	Inf.	Eff.	Anx.Eff	MLSS	MVSS	pH	Influent	Effluent	Temperature	Tubidity
		So	S3	S2	Ao	A3	A2	No	N3	N2				S/Solid	S/Solid*		(NTU)
05/09/'99	43.2	550	19.50	104	95	0.1	33.6	0.1	20.0	17.5	3.17	2.70	7.50	98	nm*	24.5	1.01
07/09/'99	"	550	11.10	94.2	94	0.5	30.0	1.0	20.8	20.1	3.56	3.10	7.40	102	nm	25.3	0.91
09/09/'99	"	505	10.30	65.0	54	0.5	10.5	0.1	18.0	13.6	4.70	3.85	7.4	110	nm	25.1	0.85
11/09/'99	"	500	17.20	48.0	89	0.1	25.2	1.0	13.0	9.5	3.88	3.22	7.4	52	nm	25.3	0.92
13/9/99	"	510	18.7	80.0	87	0.4	30.0	1.0	7.2	10.1	4.90	3.64	7.1	47	nm	24.4	0.92
15/9/99	"	450	16.2	48.2	85	0.1	33.2	1.0	6.0	12.6	5.87	4.64	7.2	45	nm	25.1	0.91
17/9/99	"	490	17.8	72.1	64	1.0	25.0	1.0	15.6	11.0	4.92	4.08	7.2	98	nm	24.8	1.02
19/9/99	"	478	17.0	68.0	68	1.0	25.4	0.1	12.3	10.9	5.42	4.25	7.4	120	nm	24.4	0.92
21/9/99	"	504	18.0	48.0	65	0.1	20.1	1.0	14.4	13.3	7.90	5.93	7.5	120	nm	24.2	0.87
23/9/99	"	475	20.1	36.1	71	1.0	24.0	1.0	18.2	12.5	7.94	6.19	7.2	115	nm	24.6	1.1
24/9/99	"	505	9.8	56.2	64	1.0	19.8	1.1	15.2	12.5	9.18	6.98	7.4	104	nm	24.5	1.02
06/10/'99	69.1	756	35.0	122	150	1.5	88.4	1.0	33.7	26.5	6.10	4.86	6.5	130	nm	23.5	0.91
10/10/'99	"	659	24.0	96.0	124	1.0	96.0	0.1	32.3	28.2	6.75	5.12	7.4	125	nm	25.1	0.75
Average														97.3846154		24.67692308	0.93153846

Table A6 (cont.) -Experimental data collected from MBR with aerobic-anoxic system

Date	UDN mgNO3-N/ mgVSS.d	UCOD (anoxic) mgCOD/ mgVSS.d	UN mgNH4-N/ mgNVSS.d	Flux L/m2-h	Press. kPa	Space Load kgCOD/m3-d	% COD Removal	% NH4-N Removal	% Nitrogen Denitrified	Biological Load Rate kgCOD/kgVSS.d
05/09/'99	0.086	0.42	1.53	71.3	30	0.594	96.5	99.90	78.95	0.220
07/09/'99	0.074	0.37	1.32	92.3	36	0.595	97.8	94.00	78.4	0.192
09/09/'99	0.031	0.3	0.61	53.86	25	0.547	97.8	99.07	73.33	0.142
11/09/'99	0.072	0.36	1.21	54.1	30	0.509	96.6	99.88	86.4	0.168
13/9/99	0.065	0.32	1.07	54.1	55	0.551	96.3	99.54	92.76	0.151
15/9/99	0.051	0.22	0.81	52.1	30	0.486	96.4	98.44	94	0.104
17/9/99	0.04	0.28	0.7	84.0	21	0.529	96.4	98.44	75.63	0.130
19/9/99	0.041	0.26	0.7	75.00	25	0.516	96.4	98.52	80.59	0.121
21/9/99	0.028	0.19	0.48	75.0	25	0.544	95.8	99.69	79.08	0.092
23/9/99	0.027	0.22	0.49	68.0	28	0.513	98	98.31	74.08	0.0829
24/9/99	0.023	0.18	0.42	65.0	32	0.545	95.3	98.13	75.53	0.078
06/10/'99	0.154	0.52	1.5	51.20	45	1.806	96.4	99.00	76.53	0.269
10/10/'99	0.164	0.44	1.6	50.20	45	1.2	95.8	98.55	72.58	0.234
Average	0.065846154	0.313846154	0.956923077	65.0892308	32.8461538	0.687307692	96.5769231	98.57	79.8353846	0.153

Table A6 (cont.)

Date	Feed		COD, mg/L		NH4+-N, mg/l			NO3-N,mg/l			g/l			mg/l	mg/ l	oC	
	L/d	Inf. So	Eff. S3	Anx.Eff S2	Inf. Ao	Eff. A3	Anx.Eff A2	Inf. No	Eff. N3	Anx.Eff N2	MLSS	MVSS	pH	S/Solid S/Solid	S/Solid S/Solid*	Temperature	Tubidity (NTU)
12/10/'99	69.1	645	27.0	82.0	120	1.5	60.4	1.0	22.0	24.5	8.28	7.04	7.6	118	nm	25.5	1.01
14/10/99	"	694	30.0	120.2	150	1.0	94.1	1.0	24.2	20.9	--	--	7.2	99	nm	25.2	1
16/10/99	103.7	639	17.0	74.0	225	2.6	127.2	1.2	22.6	30.2	--	--	7.3	101	nm	25.3	0.89
18/10/99	"	698	32.0	130.0	220	2.0	98.0	1.2	32.1	15.1	9.82	8.45	7.3	98	nm	25.3	0.95
20/10/99	'	657	35.0	132.0	125	1.5	25.5	1.0	32.0	24.2	9.36	8.42	7.6	104	nm	24.1	0.95
22/10/99	138.2	687	28.5	120.0	125	1.0	30.5	0.1	30.0	24.4	8.93	7.95	7.6	114	nm	24.2	0.96
24/10/99	"	650	30.2	130.3	124	1.0	42.1	1.0	20.5	15.2	8.40	7.55	7.4	103	nm	25.2	1.02
26/10/99	"	725	45.0	91.2	127	1.0	35.2	1.0	15.5	11.0	8.86	7.83	7.2	115	nm	25.7	1.11
28/10/99	69.1	925	36.4	96.5	145	1.5	48.9	1.0	15.0	11.2	9.80	8.73	7.4	112	nm	24.6	1.11
30/10/99	"	714	27.3	78.0	148	2.5	60.3	1.0	10.0	9.8	9.70	8.54	7.2	132	nm	24.3	1.1
02/11/1999	"	950	29.0	95.2	150	2.2	40.0	1.1	8.7	9.2	10.34	9.00	7.2	144	nm	24.8	0.95
04/11/1999	103.7	860	36.4	90.2	150	1.8	38.0	0.9	20.5	16.2	11.43	10.29	7.5	119	nm	25.2	0.95
06/11/1999	138.2	950	32.4	140	154	0.2	56	1.0	38.2	24.5	12.06	10.97	7.5	122	nm	24.3	0.96
08/11/1999	"	648	27.3	98.2	140	3.2	25.0	1.0	54.0	26.0	12.21	11.24	7.1	98	nm	26.1	1.01
Average													7.3462	112.384615		24.94615385	0.99692308

Table A6 (cont.)

Date	UDN mgNO3-N/ mgVSS.d	UCOD (anoxic) mgCOD/ mgVSS.d	UN mgNH4-N/ mgNVSS.d	Flux L/m2-h	Press. kPa	Space Load kgCOD/m3-d	% COD Removal	% NH4-N Removal	% Nitrogen Denitrified	Biological Load Rate kgCOD/kgVSS.d
12/10/99	0.1	0.32	1.03	50.9	39	1.114	95.5	98.75	81.25	0.158
14/10/99	-	-	-	62.0	38	1.198	97.3	99.40	83.93	-
16/10/99	-	-	-	53.8	39	1.657	97.3	98.80	89.33	-
18/10/99	0.151	0.41	1.41	62.0	45	1.89	95.4	99.09	85.05	0.214
20/10/99	0.081	0.38	1.14	60.0	40	1.703	94.8	98.80	74.00	0.202
22/10/99	0.118	0.58	1.74	65.0	35	2.373	95.9	99.20	78.42	0.299
24/10/99	0.132	0.56	1.82	55.0	25	2.245	95.4	99.19	83.47	0.297
26/10/99	0.138	0.41	1.79	55.5	25	2.504	93.8	99.21	87.78	0.320
28/10/99	0.072	0.38	1.13	60.4	28	1.598	96.1	98.96	89.31	0.183
30/10/99	0.074	0.3	0.96	71.0	35	1.233	96.2	98.83	92.22	0.144
02/11/1999	0.07	0.38	0.73	68.2	40	1.641	96.9	98.53	93.47	0.182
04/11/1999	0.141	0.45	1.22	68.5	35	2.22	95.8	98.80	85.73	0.217
06/11/1999	0.195	0.62	1.81	72.1	35	3.282	96.6	99.87	75.71	0.30
08/11/1999	0.101	0.4	1.76	70.5	35	2.239	96.0	97.77	59.86	0.190
Average	0.115727273	0.442727273	1.41	63.3846154	35	1.983307692	95.9615385	98.9576923	82.9446154	0.231636364

Table A6 (cont.)

Date	Feed	COD, mg/L			NH4+-N, mg/l			NO3-N,mg/l			g/l			mg/l	mg/ l	oC	
	L/d	Inf.	Eff.	Anx.Eff	Inf.	Eff.	Anx.Eff	Inf.	Eff.	Anx.Eff	MLSS	MVSS	pH	S/Solid	S/Solid	Temperature	Tubidity
		So	S3	S2	Ao	A3	A2	No	N3	N2				S/Solid	S/Solid*		(NTU)
10/11/2011	69.1	875	35.0	75.2	120	1.3	34.2	1.1	38.0	23.3	12.84	11.43	7.5	108	nm	24.5	1.11
12/11/1999	"	698	28.5	63.4	135	1.5	28.2	1.0	33.0	19.0	14.06	12.10	7.4	95	nm	25.1	0.98
14/11/99	69.1	954	27.0	105.0	140	3.2	28.0	1.0	19.7	20.0	13.67	12.30	7.5	121	nm	24.6	0.99
Average																24.84461538	1.01569231

Table A6 (cont.)

Date	UDN mgNO3-N/ mgVSS.d	UCOD (anoxic) mgCOD/ mgVSS.d	UN mgNH4-N/ mgNVSS.d	Flux L/m2-h	Press. kPa	Space Load kgCOD/m3-d	% COD Removal	% NH4-N Removal	% Nitrogen Denitrified	Biological Load Rate kgCOD/kgVSS.d
10/11/2011	0.102	0.28	0.95	59.4	24	1.511	95.9	98.89	68.17	0.132
12/11/1999	0.17	0.21	0.96	60.1	35	1.205	96.1	98.89	75.19	0.100
14/11/99	0.169	0.42	1.49	59.5	35	1.648	95.3	97.71	84.36	0.134
Average	0.120972727	0.388272727	1.242	65.3084615	33.7	1.856030769	96.0861538	98.7207692	80.6964615	0.181363636

APPENDIX 7

Results obtained from MBR with intermittent aeration and sludge wasting.
(Sludge age used is 20 days)

Table A7

The value in the parentheses indicate :average, aerobic, and anoxic values.

Date	Feed	COD, mg/L		NH4-N, mg/L		NO3-N, mg/L		S/Solid, mg/L		Tubidity	% COD
	L/d	influent	effluent	influent	effluent	influent	effluent	influent	effluent	NTU	Removal
19/11/99	69.1	748.5	(25.3)(22.1)(28.5)	122.2	(3.3)(1.9)(4.7)	-	(25.2)(35.2)(15.2)	56.2	nm	1.1	96.61
20/11/99	69.1	744.2	(20.6)(18.8)(22.4)	120.0	(1.2)(1.0)(1.4)	-	(20.5)(37.4)(36.0)	48.6	nm	1.2	97.23
21/11/99	69.1	758.4	(28.9)(23.5)(34.3)	120.0	(2.1)(0.9)(3.3)	-	(20.4)(35.4)(5.4)	98.7	nm	0.9	96.19
22/11/99	69.1	750	(32.3)(25.8)(38.8)	120.6	(1.4)(1.0)(1.8)	-	(18.4)(28.2)(8.6)	100.2	nm	1.1	95.69
23/11/99	69.1	750.3	(35.2)(28.1)(42.3)	110.1	(2.1)(0.9)(3.3)	-	(18.5)(25.5)(11.5)	96.5	nm	1.3	95.31
24/11/99	69.1	755.5	(33.3)(27.5)(39.1)	120.5	(3.2)(1.2)(5.2)	-	(19.3)(26.4)(12.2)	96.2	nm	1.1	95.59
25/11/99	69.1	758.1	(31.8)(26.9)(36.7)	114.3	(2.3)(1.0)(3.6)	-	(20.4)(32.1)(8.7)	85.7	nm	1.1	95.81
26/11/99	69.1	752.3	(28.3)(27.5)(29.1)	120.8	(1.4)(0.8)(2.0)	-	(20.9)(30.5)(11.3)	96.5	nm	0.9	96.24
27/11/99	69.1	752.3	(30.9)(26.5)(35.3)	120.0	(4.1)(1.0)(7.2)	-	(34.0)(42.4)(25.2)	100.2	nm	0.8	95.89
28/11/99	69.1	745.5	(30.5)(25.4)(35.6)	125.0	(3.9)(1.2)(6.6)	-	(30.8)(38.2)(23.4)	110.5	nm	0.9	95.91
29/11/99	69.1	745.2	(19.3)(19.1)(15.5)	128.0	(2.3)(0.9)(3.7)	-	(25.0)(34.6)(15.4)	100	nm	0.5	97.41
30/11/99	69.1	750.3	(20.4)(19.5)(21.3)	120.1	(2.8)(1.1)(4.5)	-	(22.1)(32.8)(11.4)	102.2	nm	0.8	97.28
01/12/99	69.1	750	(22.7)(20.6)(24.8)	125.3	(2.0)(0.8)(3.2)	-	(25.3)(30.5)(20.1)	98.8	nm	1.1	97.05
Average	69.1	750.8153846		120.5307692				91.56153846		0.984615385	96.32384615

Table A7 (cont.)

Date	% NH4-N Removal	%Nitrogen Denitrified	MLSS mg/L	MVSS mg/L	pH	oC Temp.	Flux L/m2-h	MF Press(kPa)	Space Load kgCOD/m3-d	Biological Load kgCOD/kgVSS.d
19/11/99	97.30	76.67	4354	3890.21	7.81	24.10	45.1	50.00	2.249	0.578
20/11/99	99.00	82.03	4337	3875.2	7.52	25.20	55.2	45.00	2.236	0.577
21/11/99	98.25	81.25	-	-	7.54	25.20	61.2	35.00	2.278	-
22/11/99	98.84	83.64	4479	4002	7.91	25.20	70	35.00	2.253	0.563
23/11/99	98.09	81.22	-	-	8.1	24.60	65.3	40.00	2.254	-
24/11/99	97.34	81.66	4195	3748	7.54	25.30	65.3	40.00	2.270	0.606
25/11/99	97.99	80.14	4428	3956	7.41	23.60	71.4	35.00	2.278	0.576
26/11/99	98.84	81.54	-	-	7.42	25.50	55.8	45.00	2.260	-
27/11/99	96.58	68.25	4538	4115	7.41	24.50	65.5	45.00	2.260	0.549
28/11/99	96.88	72.24	4296	3896	7.4	24.50	38.1	55.00	2.240	0.575
29/11/99	98.20	78.67	4363	3956	7.51	24.60	22.2	45.00	2.239	0.566
30/11/99	97.67	78.27	-	-	7.21	24.90	25.2	45.00	2.254	-
01/12/99	98.40	78.21	4577	4150	7.34	24.80	36	45.00	2.253	0.543
	97.95333586	78.75307692		3954.267778	7.547692308	24.76923077	52.02307692	43.07692308	2.255710569	0.570267025

Table A7 (cont.) Results obtained from MBR with intermittent aeration and sludge wasting
(Sludge age used is 20 days)

Date	Feed	COD, mg/L		NH4-N, mg/L		NO3-N, mg/L		S/Solid, mg/L		Tubidity	% COD
	L/d	influent	effluent	influent	effluent	influent	effluent	influent	effluent	NTU	Removal
02/12/99	69.1	750.5	(22.1)(20.2)(24.0)	130.4	(3.8)(1.2)(6.4)	-	(15.2)(28.9)(1.5)	96.6	nm	1.2	97.06
03/12/99	69.1	750	(25.8)(22.9)(28.7)	120.1	(2.5)(1.2)(3.8)	-	(15.3)(26.7)(3.9)	96.5	nm	1.1	96.56
04/12/99	69.1	753.2	(29.9)(25.5)(34.3)	120.3	(2.8)(1.0)(4.6)	-	(18.9)(25.7)(12.1)	98.6	nm	1.3	93.03
05/12/99	69.1	759.8	(30.0)(25.8)(34.2)	120.3	(1.8)(0.8)(2.8)	-	- - -	100.5	nm	1.2	96.05
06/12/99	69.1	753.5	(34.2)(29.8)(38.5)	120.4	(1.5)(0.9)(2.1)	-	(24.5)(34.2)(14.8)	112	nm	1.2	95.46
07/12/99	69.1	750.4	(32.5)(30.1)(34.5)	121.5	(0.9)(0.01)(1.79)	-	(25.3)(36.6)14.0)	97.2	nm	1.2	95.67
08/12/99	69.1	698.9	(32.3)(29.8)(34.8)	120.3	(3.9)(1.0)96.8)	-	(16.1)(30.2)(2.2)	110.4	nm	1.2	95.38
09/12/99	69.1	661.3	(28.7)(25.7)(31.7)	120.0	(3.9)(1.4)(6.4)	-	(20.4)(32.0)(8.8)	110.2	nm	1.2	95.66
10/12/99	69.1	625.1	(28.8)(23.5)(34.1)	130.6	(2.4)(0.5)(4.3)	-	(20.9)(28.3)(13.5)	100	nm	0.9	95.39
11/12/99	69.1	658.8	(30.2)(28.2)(32.2)	130.8	(1.3)(0.5)(2.1)	-	- - -	95.2	nm	0.8	95.42
12/12/99	69.1	691.7	(33.2)(29.2)(37.2)	132.2	(1.5)(0.7)(2.3)	-	(18.5)(28.0)(9.0)	75.5	nm	1.2	95.2
13/12/99	69.1	731	(29.8)(26.6)(33.0)	130.2	(1.8)(0.8)(2.8)	-	- - -	-	nm	1.1	95.92
14/12/99	69.1	750	(30.6)(30.3)(30.9)	120.5	(3.8)(1.2)(6.4)	-	- - -	48.7	nm	1.2	95.92
15/12/99	69.1	749	(26.4)(22.0)(30.8)	120.9	(4.1)(0.8)(7.4)	-	(25.0)(34.9)(15.1)	56.2	nm	0.8	96.48
Average	69.1	717.9		123.7				91.75		1.107692308	95.54923077

Table A7 (cont.)

Date	% NH4-N Removal	%Nitrogen Denitrified	MLSS mg/L	MVSS mg/L	pH	oC Temp.	Flux L/m2-h	MF Press(kPa)	Space Load kgCOD/m3-d	Biological Load kgCOD/kgVSS.d
02/12/'99	97.09	85.43	4369	3962	7.38	24.80	34.8	45.00	2.255	0.569
03/12/'99	97.92	85.18	4273	3875	6.91	24.80	55.2	35.00	2.253	0.581
04/12/'99	97.67	81.96	4301	3900	7.12	25.20	62.5	35.00	2.263	0.580
05/12/'99	98.50	-	-	-	7.32	25.10	65.6	40.00	2.283	-
06/12/'99	98.75	78.41	4284	3885	7.32	24.80	60.4	45.00	2.264	0.583
07/12/'99	99.26	78.44	4187	3797	7.15	23.90	54.8	35.00	2.254	0.594
08/12/'99	96.76	83.37	4080	3700	7.32	25.20	58.5	35.00	2.100	0.567
09/12/'99	96.76	79.75	4323	3920	7.54	26.10	60.1	37.00	1.987	0.507
10/12/'99	98.16	82.16	-	-	6.98	24.10	58.4	50.00	1.878	-
11/12/'99	99.01	-	4250	3854	7.5	24.60	62.3	45.00	1.979	0.514
12/12/'99	98.87	84.87	4397	3987	7.45	25.50	38.7	55.00	2.078	0.521
13/12/'99	98.62	-	4417	4005	7.62	24.80	40.2	55.00	2.196	0.548
14/12/'99	96.85	-	-	-	7.52	23.90	38.6	47.00	2.253	-
15/12/'99	96.61	75.93	4570	4144	7.54	23.60	30.1	45.00	2.250	0.543
	97.979286	81.11888889		3906.7	7.33	24.73846154	52.72307692	43	2.156821304	0.553863983

Table A7 (cont.)

Date	Feed	COD, mg/L		NH4-N, mg/L		NO3-N, mg/L		S/Solid, mg/L		Tubidity	% COD
	L/d	influent	effluent	influent	effluent	influent	effluent	influent	effluent	NTU	Removal
16/12/'99	69.1	748.1	(28.4)(22.3)(34.5)	119.2	(0.8)(0.1)(1.5)	-	(20.1)(28.3)(11.9)	58.5	nm	0.9	96.2
17/12/'99	69.1	753.2	(32.5)(25.8)(39.2)	120.4	(0.5)(0.1)(0.9)	-	(22.3)(34.8)(9.8)	56.2	nm	1.1	95.69
18/12/'99	69.1	740.1	(39.2)(32.5)(45.9)	120	(0.8)(0.1)(1.5)	-	- - -	-	nm	1.2	94.7
19/12/'99	69.1	738.6	(36.1)(30.5)41.7)	122.2	(1.5)(0.1)(2.9)	-	(19.2)(28.7)(9.7)	58.2	nm	1.2	95.11
Average	69.1	727.84		124.01				67.53125		1.060769231	95.61892308

Table A7 (cont.)

Date	% NH4-N Removal	%Nitrogen Denitrified	MLSS mg/L	MVSS mg/L	pH	oC Temp.	Flux L/m2-h	MF Press(kPa)	Space Load kgCOD/m3-d	Biological Load kgCOD/kgVSS.d
16/12/99	99.33	82.47	4643	4210	7.39	25.00	29.3	55.00	2.248	0.534
17/12/99	99.58	81.06	-	-	7.35	24.80	62.4	45.00	2.263	-
18/12/99	99.33	-	4659	4225	7.39	25.00	58.4	40.00	2.224	0.526
19/12/99	98.77	83.06	4585	4158	7.45	25.10	54.6	45.00	2.219	0.534
	98.49428509	81.41814815		4061.2275	7.454	24.70384615	46.73230769	47.5	2.186684522	0.534226848