DEPARTMENT OF CHEMICAL ENGINEERING

UNIVERSITY OF CAPE TOWN

FINAL REPORT
TO THE
WATER RESEARCH COMMISSION OF SOUTH AFRICA

on

AN EXPLORATORY INVESTIGATION OF CROSSFLOW MICROFILTRATION FOR SOLID/LIQUID SEPARATION IN BIOLOGICAL WASTEWATER TREATMENT

by

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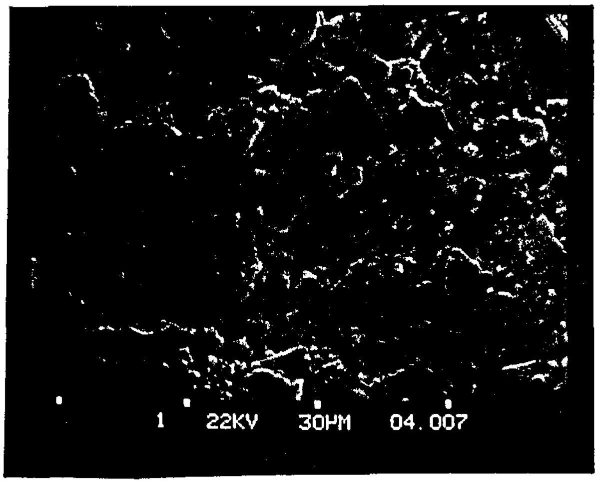

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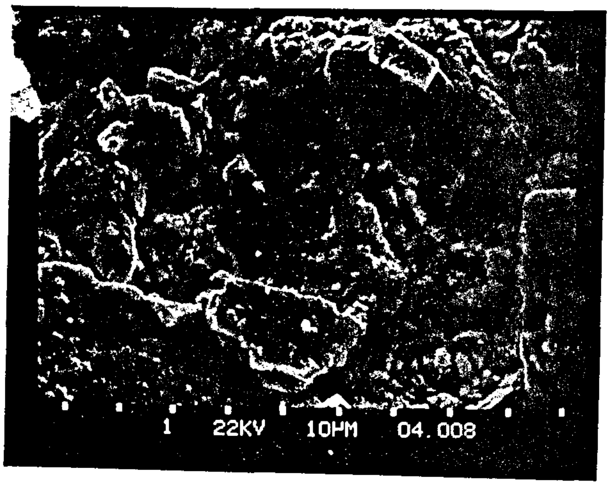

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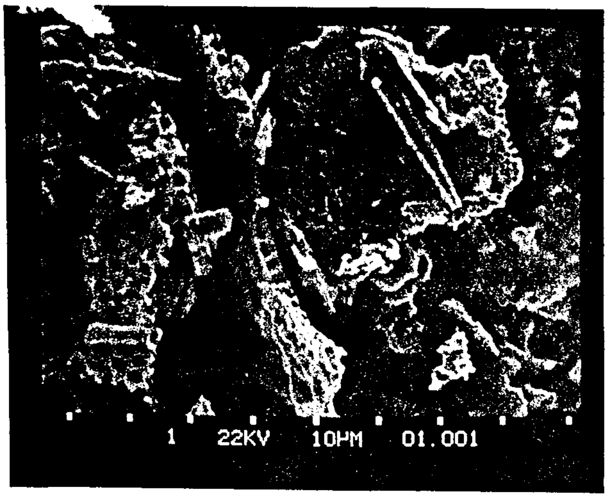

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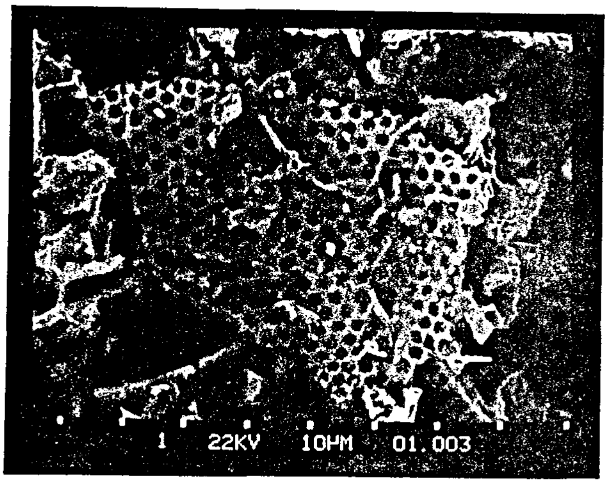

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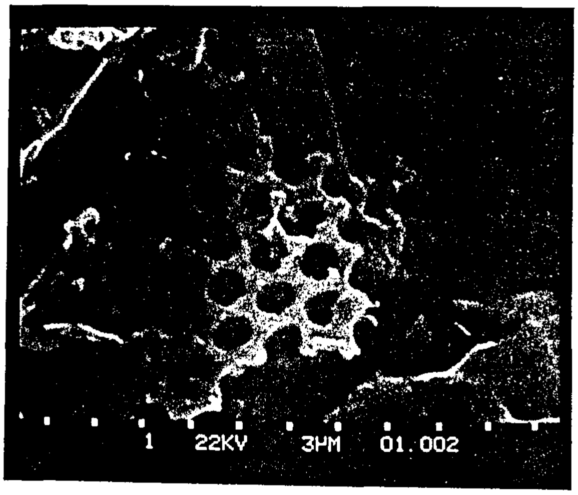

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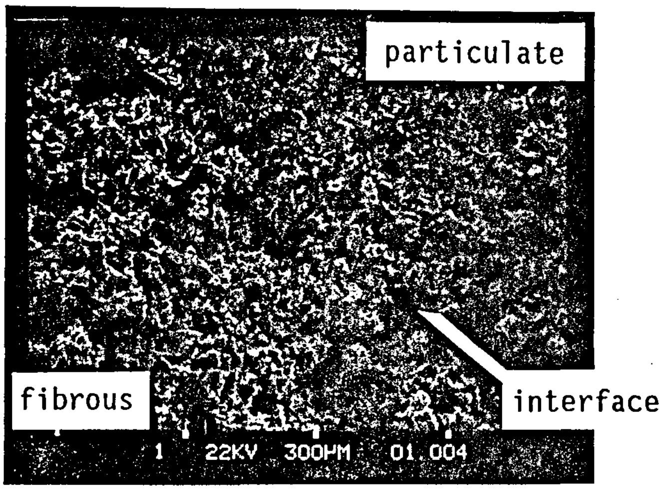

6. Swimline DE Interface
(magnification 100)

FIGURE 3.3 : Electron Micrographs of Precoat Layers

EXECUTIVE SUMMARY

This report contains the results of an exploratory investigation into the application of Crossflow Microfiltration (CFMF) for solid/liquid separation in two biological wastewater treatment systems. The systems chosen were the Upflow Anaerobic Sludge Bed (UASB) reactor and the aerobic Activated Sludge systems. The main objective of the investigation was to determine whether or not CFMF adversely affected the performance of the two systems.

Solid/liquid separation is required to achieve acceptable process performance and/or suitable effluent quality in terms of suspended solids content. At higher suspended solids concentrations in the reactor, its capacity is increased, reducing the reactor size requirements.

During crossflow filtration, the fluid flows tangentially across the filter membrane surface. The shear forces remove the filter cake that would build up during conventional dead-end filtration, maintaining a high filtrate flux. A literature survey highlighted the significant progress that has been made in the development of suitable membranes.

A laboratory scale CFMF unit with a one meter 12 mm diameter woven polyester tube (Schweiz Seidengazefabriek AG, Thal) membrane was constructed. The CFMF unit could be connected to a precoat tank or to a bioreactor. Filter aid (Kulu 15A limestone, (Lewis and Everett)) was applied as a precoat layer to prevent irreversible fouling of the membrane by biological growth in the membrane pores. The CFMF unit was operated at pressures of between 2,0 and 2,5 kPa and at a crossflow velocity of 2,2 to 3,6 $m.s^{-1}$.

Waste activated sludge was used in initial tests to determine the conditions required for the formation of a stable precoat layer. Test data obtained were compared with those obtained by the Natal University group. The tests indicated that the CFMF unit should be operated at between 2,0 and 2,5 kPa with a crossflow velocity of 2,2 to 3,6 $m.s^{-1}$.

The precoat slurry concentration was also an important factor for stable precoat layer formation. A slurry concentration of 1 $g.l^{-1}$ was found to be adequate. Optimisation of the filtration process was beyond the scope of this investigation. A minimum steady state flux of 15 $l.m^{-2}h^{-1}$ was required to balance the influent flow to the reactor.

A UASB reactor is essentially a vertical plug-flow reactor to which wastewater is fed at the bottom. The biological sludge is generated in the form of pellets which have good settleability characteristics. They form a dense bed at the bottom of the reactor largely solving the solid/liquid separation problem. Pellets break up into fine particles at the top of the bed, however, raising the effluent Chemical Oxygen Demand (COD) and the suspended solids concentration, lowering its quality.

Pelletised sludge from an operating UASB reactor was used for the start-up of the UASB reactor in this study. The amount of synthetic wastewater fed to the reactor was increased until 15 l of feed with a COD of 5000 $mg.l^{-1}$ was fed to the reactor daily. The reactor was operated at 35°C. Once the COD removal had reached 90 per cent, the reactor was taken as having stabilised. The suspended solids concentration was measured as 1 $g.l^{-1}$.

Filtration of the liquid above the pelletised bed commenced, effluent having previously been drawn off from the surface of the liquid in the reactor. The sludge bed did not appear to be disturbed during the filtration process.

The CFMF unit flux dropped to below the 15 $l.m^{-2}h^{-1}$ minimum due to the low filterability of the UASB sludge. This prompted an investigation into different types of precoat material. Commercially available diatomaceous earth (Swimline Diatom Powder, Swimline (Pty) Ltd.) was found to give the highest flux (around 20 $l.m^{-2}h^{-1}$), without the loss of the high effluent quality attained with limestone, and was used in the remainder of the experimental work.

The suspended solids concentration at the top of the reactor was increased from 1,0 to 5,9 $g.l^{-1}$ in 17 days using CFMF. The maximum solids concentration was not reached during the course of this study. The COD removal was increased from 90 to 98 per cent. This was attributed to the possible growth of aerobic organisms on the residual soluble COD material at the top of the reactor that had been aerated by the turbulent return of the CFMF reject. This did not affect the operation of the anaerobic sludge bed.

Recirculation of the slurry through the CFMF unit changed its nature. The average particle size was reduced from 36 µm to 16 µm, lowering the filterability of the sludge.

The CFMF effluent was of a high quality. No suspended solids were detected on 19 of the 31 days and the concentration was 50 $mg.l^{-1}$ on average, this was considerably below the 500 $mg.l^{-1}$ normally found in effluent from the gravity settling tanks conventionally used for solid/liquid separation. The membrane did not have to be cleaned for up to four days.

A ten litre activated sludge system was coupled to the CFMF unit in place of the UASB reactor. The CFMF unit operating conditions were unchanged.

The size of the settling tanks, conventionally used for solid/liquid separation, is dependent on the reactor solids concentration and the settleability of the sludge (which may vary due to process upsets). The cost of the tanks limit the concentration at which the bioreactor may be operated. CFMF opens up the possibility of operation at higher solids concentrations.

Activated sludge wasted from reactors fed a similar synthetic wastewater feed was used for reactor start-up. The reactor was maintained at 25°C. There were six experimental stages at which steady state was reached. The hydraulic load on the system was 15 $l.day^{-1}$, whilst the COD was increased from 7,5 $g.day^{-1}$ to 45,0 $g.day^{-1}$ and the sludge age increased from 5 to 20 days.

The suspended solids concentration in the reactor was increased from 0,97 $g.l^{-1}$ to 19,47 $g.l^{-1}$, representing a nearly four-fold increase in the maximum concentration at which conventional systems are operated. The maximum concentration at which the system could be operated was not reached during the experimental work.

The suspended solids concentration was only 71 per cent (on average) of that predicted by steady-state activated sludge theory. Growth of microorganisms in the feed tank and feed piping was one potential cause of the discrepancy. The kinetic constants used for the prediction were based on a slurry that included larger organisms than those present in this study. The COD removal was between 98 and 99 per cent once steady state conditions had been reached at the end of each stage.

The energy input due to recirculation changed the nature of the sludge. The average particle size was reduced from 103 μm to 2 μm, making the sludge unsettleable. Microscopic examination confirmed that organisms larger than 2 μm were not present in the sludge. The reactor efficiency had not, however, been impaired. The energy input in a full-scale process should be lower so the size reduction may not be as great.

The sludge viscosity increased with increasing solids concentration, reducing the CFMF flux. The dependence of flux on slurry viscosity and the dependence of viscosity on temperature, highlights operating temperature as an important design consideration in CFMF-bioreactor coupled systems. As activated sludge processes follow seasonal temperature changes the flux may vary as much as 20 per cent between winter and summer.

No suspended solids were detected in the CFMF effluent during 46,8 per cent of the operation. On average the concentration was 0,013 $g.l^{-1}$. Theoretically no solids should pass through the CFMF membrane, but growth of microbes on the residual COD material present in the effluent is most likely to have led to the detection of solids. At a suspended solids concentration of 11 $g.l^{-1}$ the crossflow velocity had to be increased from 2,2 to 3,6 $m.s^{-1}$ to maintain the minimum flux. Cleaning was usually necessary only every four days.

It was concluded that the coupling of both biosystems with CFMF enhanced their performance. The COD removal remained high at the elevated solids concentrations whilst the CFMF effluent quality was better than that possible with gravity settling. In both cases the maximum solids concentration at which the systems could be operated was not reached.

The increase in sludge viscosity with solids concentration and the effect of temperature were concluded to be important considerations in the design of bioreactor-CFMF coupled systems.

It is recommended that, after the successful operation of both of the laboratory reactors, the operation of pilot-scale reactors coupled with CFMF units should be investigated. Constraints such as economic factors and the aeration of the activated sludge systems should be studied. An optimum reactor suspended solids concentration should be determined, as well as the maximum suspended solids concentrations at which the reactors can be operated. This is likely to arise from a comparison of the cost saving of the reduced reactor size against the cost and physical constraints of aeration, sludge handling, pumping and the cost of the CFMF system. As the viability of a bioreactor-CFMF coupled system is dependent on the CFMF flux, attention must be directed at the selection of precoat materials and the technique of precoating. The implementation of membranes that do not require precoating should reduce both operating costs and the system complexity.

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ABBREVIATIONS

AS : Activated Sludge

CFMF : Crossflow Microfiltration

COD : Chemical Oxygen Demand

DE : Diatomaceous Earth

MLSS : Mixed Liquor Suspended Solids

MLVSS : Mixed Liquor Volatile Suspended Solids

PVC : Polyvinyl Chloride

RO : Reverse Osmosis

TDS : Total Dissolved Solids

TSS : Total Suspended Solids

UASB : Upflow Anaerobic Sludge Bed

UF : Ultrafiltration

VSS : Volatile Suspended Solids

SYMBOLS

b_h : Heterotrophic organism endogenous respiration rate, d^{-1}

b_n : Nitrification organism endogenous respiration rate, d^{-1}

f : Unbiodegradable fraction of active mass, mg VSS(mg VSS)$^{-1}$

f_{cv} : COD to VSS ration of active sludge mass, mg VSS(mg COD)$^{-1}$

Kr : Conversion rate of biodegradable organic nitrogen to free and saline ammonia by active mass, l.(mg VSS)$^{-1}$d^{-1}

Kn : Half saturation coefficient in Monod equation for nitrification, mg N.l^{-1}

Y_h : Heterotrophic organism yield coefficient, mg VSS(mg COD)$^{-1}$

Y_n : Nitrifier organism yield coefficient, mg VSS(mg VSS)$^{-1}$

μ_{nm} : Specific growth rate of nitrifiers, d^{-1}

CHAPTER ONE

INTRODUCTION

1.1 OBJECTIVE

The aim of this project was to conduct an exploratory experimental investigation into the application of crossflow microfiltration (CFMF) for solid/liquid separation in biological wastewater treatment systems. The membrane selected for the experimental investigation was of the self-supporting, woven cloth type with a filter precoating.

Two types of treatment system were selected for the preliminary study :

(i) The upflow anaerobic sludge bed (UASB) reactor
(ii) The single reactor aerobic activated sludge system.

The principle objective of the study was to determine whether or not Crossflow Microfiltration for solids/liquid separation and operation of the systems at high solids concentrations would have a detrimental effect on the performance of the two biological systems. Optimisation of the filter performance was not an objective.

The following objectives were initially proposed for this project :

(i) To obtain guidlines as to the solids concentration at which units may be operated;
(ii) Assess the performance with respect to any rejection of chemical species (soluble organics, NH_3, NO_3, and PO_4) in the wastewater;
(iii) Identification of operating problems encountered with filtration as the mode of separation;
(iv) Identification of problems encountered with operating an activated sludge process at high suspended solids concentrations: for example, difficulties in achieving sufficient oxygen transfer.

All the above objectives, excepting objective (ii) were achieved in this experimental work.

1.2 BACKGROUND

In recent years, significant progress in solids/liquid separation by membrane systems has been made worldwide. In South Africa one of the specific areas of interest has been the crossflow mode of filtration. This interest was initiated by research studies, principally at the Universities of Natal and Stellenbosch, and has now advanced to pilot and full-scale applications of crossflow filtration.

To date there has been only limited evaluation of crossflow filtration in the context of biological wastewater treatment. Due to the benefits derived in other areas, and because solid/liquid separation is an important unit operation in wastewater treatment, it seemed appropriate to initiate an investigation into crossflow filtration in wastewater treatment applications. For this initial study two types of biological system were selected : namely, anaerobic UASB and aerobic activated sludge.

Both the UASB and the activated sludge systems require solid/liquid separation to achieve acceptable process performance and/or suitable effluent quality in terms of suspended solids content. In the following sections the individual requirements of these two systems are discussed.

1.3 UPFLOW ANAEROBIC SLUDGE BED (UASB) REACTOR

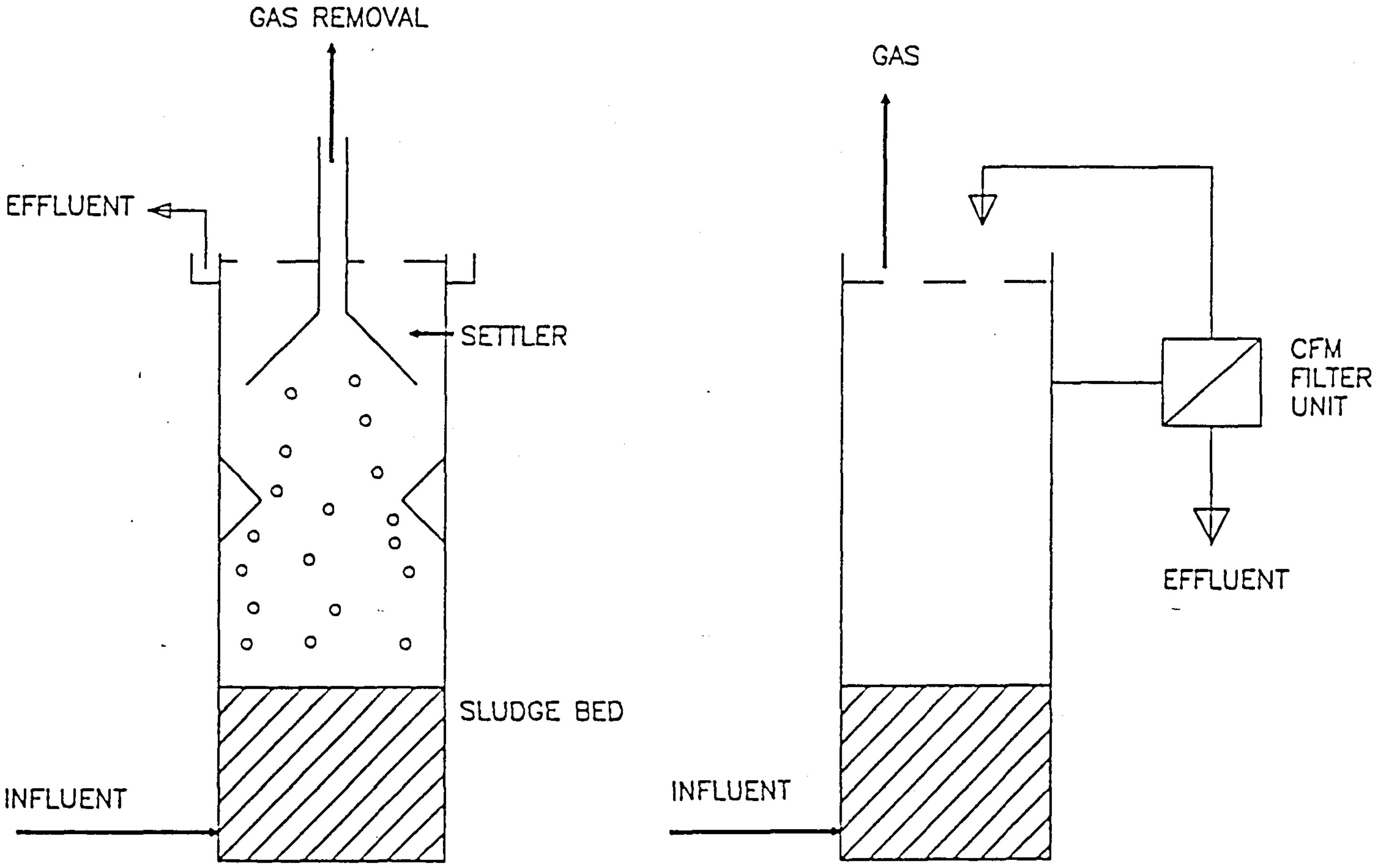
In the anaerobic digestion process, organic compounds are transformed into methane and carbon dioxide in the absence of free oxygen. There are essentially two stages: acidogenisis and methanogenisis. During the first stage, complex organic compounds are hydrolysed and partially degraded to give simple organics, such as fatty acids, together with production of carbon dioxide and hydrogen. During the second stage, the products from the first stage are converted to methane and carbon dioxide (Anderson et al, 1986). The microorganisms in an anaerobic system reproduce more slowly than those in aerobic systems, so a longer minimum solids retention time is required for process stability and minimum sludge production.

Recent developments in anaerobic wastewater treatment have hinged on techniques for improving solids retention within the bioreactor. One such system is the upflow anaerobic sludge bed (UASB) reactor. This is basically a vertical plug flow reactor with a solids/liquid/gas separation device at the top of the reactor (Fig. 1.1(a)). Wastewater is fed to the bottom of the reactor by a distributor comprising evenly-spaced nozzles. As the effluent rises through the reactor, anaerobic bacteria digest its organic content and produce the carbon dioxide and methane mixture known as biogas. The key feature of this mode of operation is that with most wastewaters the sludge generated is in the form of dense pellets which form a sludge bed in the reactor.

Sam-Soon et al (1987) proposed a mechanism for the pellet formation. It is thought that Methanobacterium Strain AZ, which uses hydrogen as its only energy source, is stimulated by the high hydrogen partial pressure at the base of the reactor This organism has the characteristic that it is able to produce all of its amino acids, with the exception of cysteine, at a high rate. The organism's growth therefore is limited by the availability of cysteine from external sources. With this limitation of cysteine, extracellular long chain polypeptides are formed as a result of the excess amino acid production and these bind the species with other organisms to form sludge pellets.

The pellets have good settleability characteristics and in this way, the UASB reactor has overcome the problems of solid-liquid separation to a certain extent - in some cases a settling tank was no longer required, the final product simply being the treated wastewater that is drawn off the top of the reactor.

It has been found, however, (Sam-Soon et al, 1987) that there is an upper inactive zone in the reactor where pellet break-up is observed, resulting in the production of fine organic particles. Bubbles of biogas rising through this zone keep the fine material in suspension above the sludge bed. The fine suspended solids are often not removed by a settling tank and consequently reduce the quality of the effluent. Generally, the concentration of suspended solids in the effluent from the standard UASB system is in the region of 300 to 600 mg suspended solids per litre. This solids content will contribute approximately 500 to 1000 mgl^{-1} to the effluent COD concentration. Therefore, even though the soluble COD removal



(a) STANDARD WITH SETTLER UNIT

(b) EFFLUENT REMOVAL BY FILTRATION

FIGURE 1.1 : UASB REACTOR CONFIGURATIONS

FIGURE 1.1 : UASB Reactor Configurations

in a UASB system may be very effective, the overall removal in terms of the total effluent COD is not as efficient. For example, say the influent COD concentration was 5000 $mg.l^{-1}$ and the effluent filtered and unfiltered COD concentrations 500 and 1000 $mg.l^{-1}$, respectively. The soluble COD removal would be 90 per cent; however, the overall COD removal would only be 80 per cent.

An alternative means for solids/liquid separation in the standard UASB reactor configuration would be to use the method shown schematically in Fig. 1.1(b). In this system a stream taken off near the top of the reactor circulates through a CFMF unit; reject passes back to the top of the reactor and permeate constitutes the effluent. This arrangement holds several advantages over the standard UASB system:

(i) There is no longer the requirement for the solid/liquid/gas separator unit within the reactor; this would simplify the reactor construction.

(ii) Fine material would be excluded from the effluent and concentrated in the reactor, giving a solids-free effluent.

(iii) The concentration of fine suspended solids above the sludge bed can be controlled through intermittent wastage from the top of the reactor. Because the sludge production is small and the concentration high, solids disposal would not constitute a major problem.

(iv) By removing organic fine suspended solids from the effluent, the organic content of the effluent in terms of COD concentration would be reduced.

The effectiveness of filtration as a means for removing the suspended fine organic material needs to be evaluated as an alternative to the conventional UASB settling tank unit.

1.4 ACTIVATED SLUDGE SYSTEM

In aerobic biological processes, the wastewater is contacted with bacteria and protozoa in the presence of free oxygen. The most commonly used system is the "activated sludge process", where the microbial population is a flocculant suspension in an aerated tank; atmospheric oxygen or compressed air is used for aeration. Mixed liquor from the aeration tank passes to a solid/liquid separation device; thickened solids are returned to the tank, and clear liquid is removed as effluent (Fig. 1.2).

Solids/liquid separation is an important unit operation in the activated sludge systems for two reasons :

(i) For retaining biomass in the system thereby increasing the solids retention time in the system above the hydraulic retention time;

(ii) For clarifying the effluent from the system.

Traditionally, gravity settling in secondary settling tanks has been used for solid-liquid separation in activated sludge systems. The settling tank combines the function of a clarifier (producing a clarified final effluent) and a thickener (producing continuous underflow of thickened sludge to return to the biological reactor). Should the settling tank fail in either of these two functions, sludge will be carried over the effluent weirs and escape with the effluent. Besides delivering an effluent of poor quality, loss of sludge could affect the behaviour of the biological process by uncontrolled reduction of the sludge age to below that required for proper plant performance.

The conditions in the biological reactor affect the settling and clarification characteristics of the sludge. For example, under-aeration reduces the settleability of the sludge; over-aeration may lead to pin-point floc formation and poor clarification even though the sludge maintains good settling characteristics. Thus, the functions of the biological process and that of the secondary settling tank interact upon

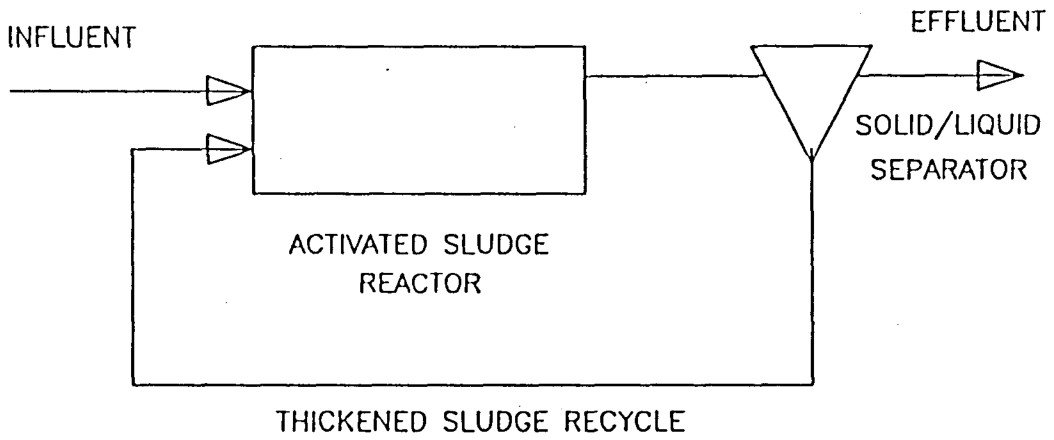


FIGURE 1.2 : ACTIVATED SLUDGE SYSTEM

each other and the design of the one cannot be undertaken independently of the other - failure in either of these two unit processes causes failure of design objectives.

The size (area) of the settling tank must be such that for the given solids and hydraulic loading from the biological reactor, the rate of settling of solids exceeds the upflow velocity of liquid in the tank. For a sludge with certain settleability characteristics the settling rate is strongly dependent on the concentration of solids in the feed to the settler from the reactor. As the reactor solids concentration increases, so the settling velocity decreases demanding a larger settling tank size (area) i.e.

Settler Size = f(Reactor solids concentration)

The mass of solids in the reactor is determined by the Chemical Oxygen Demand (COD) of the influent. Therefore the size of the reactor is dependent on the selected operating solids concentration i.e.

Reactor Volume = Mass of solids/
Selected reactor solids concentration

From this discussion it is apparent that the selected reactor solids concentration directly influences both the size of the settling tank as well as the reactor itself. High reactor solids concentrations will yield smaller reactor volumes, but at the expense of larger settling tanks. In practice, experience has shown that reactor solids concentrations of between 3000 and 5000 mgl^{-1} are appropriate when gravity settling tanks are used for solids/liquid separation.

If a solid/liquid separation were used where the separation did not constrain operation to low reactor biomass concentrations, then the volume of the process could be reduced. In addition, if the means of separation was no longer dependent on gravity, the problem of poor settleability affecting the effluent quality would be resolved. These requirements point towards considering filtration as an alternative solid/liquid separation process for the activated sludge systems.

1.5 OUTLINE OF INVESTIGATION

The report is divided into four principle sub-sections.

(i) The reasons for investigating crossflow filtration for solid/liquid separation in biological wastewater treatment systems have been established above. In crossflow filtration different filter media, all of which provide for solid/liquid separation can be utilised. For example, Reverse Osmosis membranes are used as the filter medium if the objective is to reject molecular and colloidal species from the filter permeate in addition to solids. Alternatively, if the only objective is solids/liquid separation, then a more porous membrane such as a Micro Filter may be used as qthe filter medium. To aid in the selection of the filter medium appropriate for the current investigation, a review of past experience with Crossflow Filtration in biological systems was undertaken.

(ii) Chapter Three describes the design and commissioning of the lab-scale crossflow filtration system utilised in the investigation.

(iii) Chapter Four reports on the study of UASB performance.

(iv) Chapter Five reports on the Activated Sludge system investigation.

Conclusions specific to the UASB and the Activated Sludge study are listed in Chapters Four and Five respectively. General conclusions and recommendations are listed in Chapter Six.

CHAPTER TWO

CROSSFLOW FILTRATION FOR SOLID/LIQUID SEPARATION IN BIOLOGICAL SYSTEMS

2.1 INTRODUCTION

Filtration for solids/liquid separation and clarification of effluents appears to have an important application in biological wastewater treatment systems, such as the activated sludge system and the upflow anaerobic sludge bed system.

In conventional "dead-end" filtration, the fluid flows perpendicularly to the membrane surface as shown in Fig. 2.1. The filtration flux declines rapidly, as the solids present in the fluid are deposited on the membrane during the filtration process, forming a thick filter cake of continually decreasing porosity. This is especially true with biological solids as these are gelatinous and compressible. Crossflow filtration methods were developed to overcome this problem. The filter cake formed by solids deposition is swept away as the fluid flows across the membrane (Fig. 2.1). The greater the crossflow velocity, the larger the shear forces on the membrane surface and the less cake buildup.

In this chapter experience with crossflow filtration in fermentation applications and the limited wastewater treatment applications is reviewed. Although this study is directed at application to wastewater treatment only, the review includes fermentation applications as this is the area where the technology was developed. It is likely that in many instances the experience with fermentation systems will extend to the wastewater treatment applications.

2.2 MEMBRANE TYPES

There are essentially three membrane processes that use the crossflow technique: Reverse Osmosis; Ultra Filtration; and Crossflow

Microfiltration. The processes are categorised by their membrane pore size. Figure 2.2 shows the range of pore size classifications.

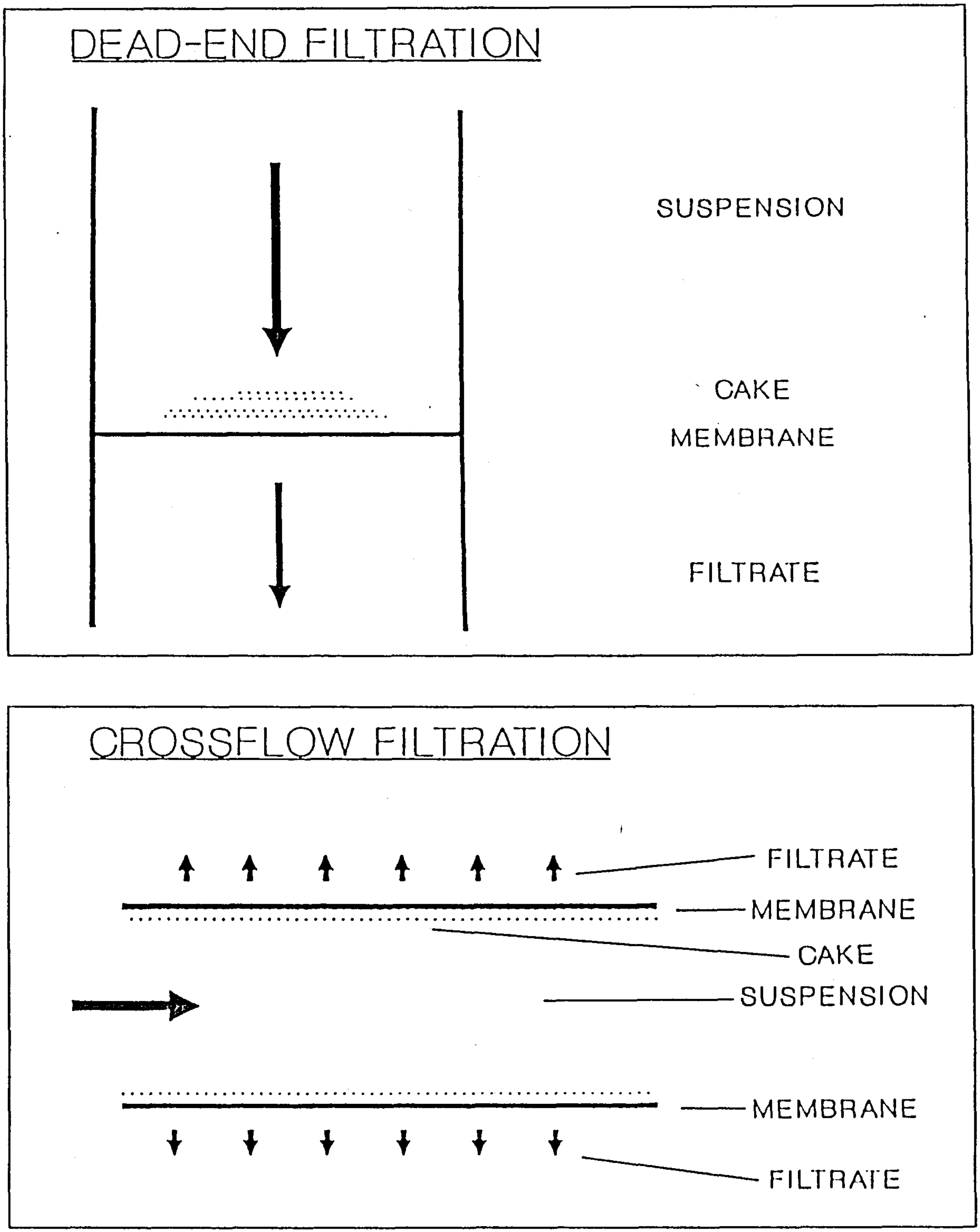


FIGURE 2.1 : Dead-end and Crossflow Filtration

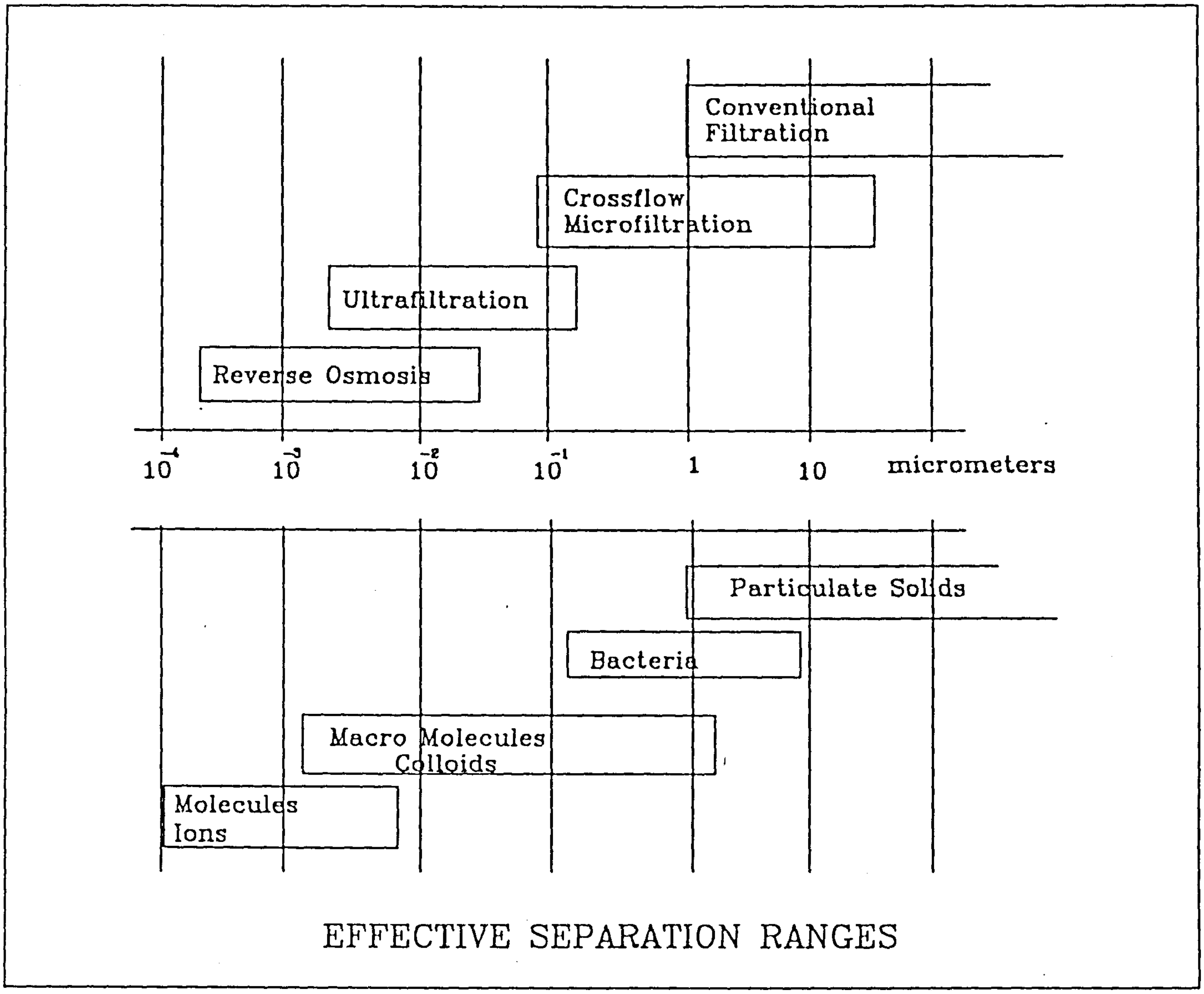


FIGURE 2.2 : Effective Separation Ranges

Reverse Osmosis (RO) membranes have the smallest pores, and are capable of preventing the passage of molecules of greater than 150 molecular weight. RO membranes reject both inorganic and ionic organic solutes up to 99 per cent. The small pore size requires the application of a large pressure (2500 - 6000 kPa) to the liquid to overcome the osmotic pressure hindering the separation of salts from water. The upper limit in RO concentration is approximately 70 gl^{-1}.

RO first became prominent in the 1950's (Hart, 1985), having been developed for the desalination of sea and brackish water, a process that has remained its major application.

Ultrafiltration (UF) membranes have larger pores than those found in RO membranes, and are generally unable to desalt fluids as the ions pass straight through the membrane. The principle function of these membranes is to reject macromolecules and colloids. The membranes are generally spiral-wound modules comprising layers of membrane, spacer and impervious sheet. The fluid flows across the membrane, whilst the permeate flows in the spiral towards the center of the module and the perforated permeate collection pipe. UF membranes are operated at relatively low pressures (100 - 800 kPa) because the rejected macromolecules and colloids do not exert a significant osmotic pressure.

Crossflow Microfiltration (CFMF) membranes are used for the retention of particulates, microorganisms and certain colloids. There is some overlap in the size range in which UF and CFMF are applied. Crossflow microfiltration systems are relatively new and require the least pressure (100 - 300 kPa) to drive the separation process, due to the relatively large pore size. The membrane modules are usually self-supported tubes through which the fluid is pumped, or plate and frame configurations.

2.3 CROSSFLOW FILTRATION IN FERMENTATION SYSTEMS

The size of material in fermentation broths ranges from tenths of nanometers to millimeters at the other extreme, posing considerable separation problems (Michaels and Matson, 1985). Additionally, the broths are dilute and a complex spectrum of components comprising feed material, product as well as the microorganisms themselves.

Solid/liquid separation in biological systems is generally used for two reasons. Initially, solid/liquid separation was solely applied for the clarification of bioreactor effluent to recover the desired product. The product can either be the liquid filtrate, or the solid biological material itself. The desired purity of the product largely determines the type of separation used in "downstream processing".

The second use for solids/liquid separation arose from the quest for higher productivity from the usually dilute biological systems. The biological cells were separated and recycled back to the bioreactor, establishing far higher cell concentrations than normal. High rate product formation is achieved and greater loadings per unit volume of reactor become possible.

The uncoupling of the solid and the liquid retention times in the reactor also enables operation above the critical "wash-out" hydraulic dilution rate (i.e. the maximum growth rate of a particular organism). The time required for start-up of a biological system is also reduced, as cells are retained even though dilution rates may exceed the initial growth rate.

The continuous removal of the bioproduct from the broth is particularly beneficial in fermentations where it may inhibit the growth of the organisms present, such as the ethanol inhibition of Saccharomyces cerevisiae. The inhibitory effect may be particularly prevalent in high concentration cultivation.

Traditionally the methods for solid/liquid separation were through gravity settling, centrifugation, rotary vacuum filtration or thermal separation (e.g. evaporation). Problems have been encountered with these methods for a variety of reasons, amongst these :

(i) As fermentation "solids" contain a large amount of water, their density is little greater than that of water. This presents a problem in separation processes, such as settling or centrifugation, that are reliant on a density difference for separation.

(ii) A large number of the products are labile compounds, ruling out thermal separation processes.

(iii) With processes such as batch centrifugation or vacuum filtration, the solids must be re-suspended before return to the bioreactor.

(iv) Processes such as centrifugation and vacuum filtration were found in many instances to cause cell damage.

Cell immobilisation is a technique that was developed for increasing bioreactor cell concentration as an alternative method to physical solids/liquid separation. With this method cells are bound in beads of jelly-like material and retained in a reactor. The technique has proved successful in a number of fermentation applications; however, its application is not without problems, and its scope is limited. For example,

(i) The method only applies to situations where liquid product is the objective.

(ii) Pre-processing costs to immobilise the cells are high and the processes often complex.

Membrane separation techniques have largely replaced the traditional methods in recent years. It has been found that many of the problems previously encountered could be circumvented with membrane separation. Examples of the use of membranes for clarification include : the clarification of beer tank bottoms to increase the yield of beer, (Le, 1987); and in the recovery of labile antibiotics from fermentation broths. Examples of the successful application of systems using solids recycle include : high rate ethanol production, (Nishizawa *et al*, 1983; Damiano *et al*, 1985; Cheryan and Mehaia, 1984); high rate lactic acid production, (Ohleyer *et al*,1985); anaerobic sludge digestion process, (Bindoff *et al*, 1987).

Some of the advantages of membrane separation are as follows:

(i) The process is not dependent on density differences to effect separation.

(ii) No phase change is required and the separation is performed at ambient temperature.

(iii) Crossflow filtration operation allows ready application in continuous flow systems or in batch or semi-batch operations.

(iv) The process additionally prevents excessive cell damage - an important consideration for cell recycle operations (Le and Atkinson, 1985).

(v) No complex cell immobilisation procedures are required.

(vi) The gas evolved during the fermentation leaves through the conventional gas port in the reactor, eliminating a significant problem found in immobilised cell reactors (Cheryan and Mehaia, 1984).

Membrane separation processes have been successfully developed for a wide range of applications, although the fouling tendencies of colloidal material and antifoams have often plagued their viability, particularly for RO and UF applications where the membrane pore sizes are small. Heavy, frequent fouling increases operating costs by requiring membrane replacement or cleaning.

2.3.1 EXPERIENCE WITH REVERSE OSMOSIS

RO membranes have the smallest pore sizes of the three membrane types considered here. Because of the problems of fouling by particulate and colloidal material and because of the low filtrate fluxes, RO membranes have not proved viable when used in conjunction with biological systems. Rather, application of RO generally is limited to downstream processing of "clean" liquids from which the foulants have been removed; for example, in the desalination and final polishing of treated wastewater.

The high capital and running costs of RO systems further restrict its application. It is not surprising, therefore, that RO is most commonly encountered in the extraction of high value products in the pharmaceutical industry. For example, RO has replaced vacuum evaporation for the concentration of antibiotics which are often extremely labile.

2.3.2 EXPERIENCE WITH ULTRAFILTRATION

The first ultrafiltration membranes with a reasonable water permeability were developed in the 1960's by Sourirajan and Loeb (1962) with the aim of desalinating water. They were found to be more applicable for the retention of macrosolutes, especially in sewage treatment. The modules manufactured today range from flat plate and frame types, to sandwich modules, tubular systems, capillary devices and hollow fibre cartridges.

Ultrafiltration finds an important application in biotechnology for the separation and removal of low molecular mass compounds and harvesting of biomass - especially of bacteria and cellular debris. The first experience with UF in this context was in 1968, when Michaels (Michaels and Matson, 1985) proposed the use of the membranes for the retention of biocatalysts in a continuously operated membrane reactor. Blatt *et al* (1968) also reported work on such a system in the same year. Since that time, there has been extensive research and development of UF applications in biotechnology. Certain of the important experiences are discussed below.

Bautista *et al* (1986) used ultrafiltration as the concentration step in hyaluronate lyase production. Concentration by conventional methods, such as solvent precipitation, takes a long time, and there is some loss in activity. Solvent precipitation is an expensive operation, with high explosion and health risks, including the cost of the disposal of the waste solvent. The low energy requirements, absence of need for the addition of chemical contaminants and the room temperature operation of UF has lead to its successful implementation. The 7,5 to 10 fold concentration of filtrate was found to be possible with little loss in activity. Excessive recirculation to obtain greater concentrations transferred too large an amount of heat to the solution, causing deactivation of the extremely labile enzymes.

Deeslie and Cheryan (1982) found that decay in reactor activity was a major problem hindering the commercial application of UF. They highlighted various factors causing the loss in activity in the reactors :

(i) Although enzyme leakage (the active biocatalyst) was found to occur predominantly during the first two hours of operation, it did not decrease the product output significantly due to the activity of the enzymes that were retained. Enzyme leakage may be more significant with less active organisms.

(ii) Thermal deactivation, caused by heat transfer by pumping in the filtration circuit, was found to cause a dramatic loss in activity.

(iii) Although high shear rates are often attributed as being the cause of excessive deactivation by the excessive rupture of cells, they did not find any evidence of it in their hollow fibre module.

(iv) Extensive fouling of the ultrafiltration membranes is another cause of decreased productivity. The remaining enzyme was sufficiently active to make up the loss within a fairly short space of time, so no decrease in the product output occurred.

Nishizawa *et al* (1983) investigated ethanol production with cell recycle, using a hollow fiber UF module for cell retention. To examine the performance of the cell-recycle system, continuous cultures were grown at various concentrations and dilution rates. The cell viability was kept above 90 per cent in all of the runs. The following observations were made :

(i) Above a dilution rate of 0,35 h^{-1} the productivity of ethanol from glucose declined at any cell concentration. This indicated that at high dilution rates the productivity decreased due to the accumulation of by-products other than ethanol.

(ii) Non-growing cells were significantly inhibited by ethanol compared with growing cells. This is a disadvantage at high cell concentrations, because non-growing cells are often used as they enable feed glucose to be used mainly for ethanol production and not the growth and production of cells. Approximately 10 per cent of the substrate can be saved for ethanol production in such a system. Another practical aspect of the non-growing cell system is that infection by other organisms is prevented due to the lack of nitrogen and other nutritionally rich compounds.

(iii) The maximum productivity of the system was found to be 27 $g.l^{-1}h^{-1}$ with 85 $g.l^{-1}$ cell concentration - a value 20-fold higher than batch culture and 12-fold more than a continuous single-stage fermentation without the recycle of cells.

Ohleyer *et al* (1985) studied the continuous production of lactic acid in a cell recycle reactor. Operation with a filtration rate of up to 18,5 $l.m^{-2}.h^{-1}$ was achieved with a cell concentration of 140 $g.l^{-1}$, which was close to the theoretical maximum packing volume of cells. No cell purge was required, because it was found that at steady-state, no cell growth occurred. Lactic acid inhibition was severe above 60 $g.l^{-1}$.

Damiano *et al* (1985) used ultrafiltration membranes for the recycling of yeast cells. Greater inhibition constants were necessary in the growth models at high recycle ratios due to the toxins that were retained in the system, by the UF molecular weight cut-off.

The metabolism of the yeast was found to vary depending on the recycle ratio. Fermentation was mostly aerobic at low recycle ratios, whilst it became anaerobic at high recycle ratios. Oxygen transfer to the system was held constant and it appeared to be the limiting factor at high recycle ratios.

Hoffman et al (1987) investigated the use of CFMF and UF membranes for the separation process :

(i) They concluded that UF membrane flux decline was not as severe as that of CFMF membranes after five hours of operation. The observation was made using their particular Saccharomyces cerevisiae fermentation broth.

(ii) At high recycle ratios, the amount of substrate that was wasted was reduced. The substrate utilisation was found to be 0,16 g cells.g^{-1} glucose when there was no recycle, whilst it decreased to 0,01 g cells.g^{-1} glucose at 97 per cent recycle. Generally, the optimum recycle ratio can be determined by a balance of fermentor productivity and the amount of wasted substrate.

2.3.3 EXPERIENCE WITH CROSSFLOW MICROFILTRATION

From the literature, it appears that CFMF has received attention in biotechnological applications only subsequent to RO and UF. This appears surprising as CFMF is in fact the simplest membrane technology, with the lowest capital and operating cost requirements. Perhaps the reason is that initially RO and UF were regarded as a means for attaining both solid/liquid separation and preferential removal of low molecular mass liquid products, i.e. simultaneous rejection of solids and macromolecular solutes. Only once the severity of the problems of RO and UF membrane fouling were recognised, did CFMF acquire prominence as a more cost effective means of solid/liquid separation and as a pre-treatment stage for RO and UF.

Self-supported hoses, similar to the inexpensive woven polyester membranes used by Bindoff et al (1985), have been developed for CFMF. A wide range of properties are available due to the different types of weave and filter aid combinations. The hose is available in essentially unlimited lengths. The permeabilities are high, and self-supported membranes do not impede the flow of the suspension over the membrane.

A great deal of research has gone into the high-rate fermentation of *Saccharomyces cerevisiae* to produce ethanol.

Cheryan and Mehaia (1984) reported that the productivity of *Saccharomyces cerevisiae* fermentation of glucose was greatly enhanced by using CFMF for cell recycle.

Cell concentrations varied from 14 to 100 $g.l^{-1}$, allowing the use of dilution rates as high as 0,5 to 2,0 h^{-1} for complete utilisation of a 10 per cent glucose feed. Productivity of 100 g ethanol.$l^{-1}h^{-1}$ could be maintained for long periods of time.

Cheryan and Mehaia (1984) compared the ethanol productivity of batch, continuous culture and immobilised cell systems. The results of the survey are shown in table 2.1 below. One can see the dramatic improvement in the productivity when crossflow microfiltration is used.

TABLE 2.1 : COMPARISON OF PRODUCTIVITY OF DIFFERENT FERMENTOR SYSTEMS FOR THE PRODUCTION OF ETHANOL

System	Organism	Feed	Concentrations		Prod.	REF
			Sugar $g\ l^{-1}$	Ethanol $g.l^{-1}$	$g.l^{-1}h^{-1}$	
Batch						
	S. cerevisiae	Glucose	100	49	2,4	a
	S. cerevisiae	Glucose	200	98	4,0	a
	Z. mobilis	Glucose	300	127	5,1	b
	K. fragilis	Lactose	150	75	3,5	c
Continuous Culture						
	S. cerevisiae	Glucose	160	31	4,1	d
	Z. mobilis	Glucose	150	55	11,0	b
Immobilised Cells						
	Z. mobilis	Glucose	100	44	29,0	e
	Z. mobilis	Glucose	150	74	57,0	f
	S. cerevisiae	Glucose	197	71	25,0	g
Membrane Recycle						
	Z. mobilis	Glucose	100	45	120,0	b
	Z. mobilis	Glucose	140	70	120,0	b
	K. fragilis	Lactose	150	75	70,0	c
	S. cerevisiae	Glucose	100	49	100,0	a
	S. cerevisiae	Glucose	200	65	130,0	a

Z. : *Zymomonas*
K. : *Kluyveromyces*
S. : *Saccharomyces*

TABLE OF REFERENCES

a : Cheryan and Mehaia (1984)
b : Rogers *et al* (1982)
c : Ghose and Tyagki (1979)
d : Margaritis *et al* (1981)
e : Klein and Kressdorf (1983)
f : Ghose and Bandyopadhyay (1980)

For long-term stability of the system, the total fermentation volume was held constant by matching the feed rate with the permeate rate. Cells were bled from the system intermittently to maintain the cell concentration at the required value.

When the system was operated at 50 and 100 $g.l^{-1}$ the glucose conversion was essentially 100 per cent, with corresponding productivities of 49 and 96 $g.l^{-1}h^{-1}$. On increasing the glucose concentration to 200 $g.l^{-1}$ the productivity was increased to 130 $g.l^{-1}h^{-1}$, but the substrate utilisation dropped to 70 per cent.

Cheryan and Mehaia (1984) suggest that it could be possible to increase the substrate utilisation by lowering the dilution rate or increasing the cell concentration. The outlet ethanol concentration would be correspondingly higher.

Cheryan and Mehaia (1983), found that the productivity of the fermentation of lactose to ethanol by Kluyveromyces fragilis was significantly improved by the coupling of a cross-flow membrane module to the primary fermentor, in a semi-closed loop configuration. An ethanol productivity of 240 $g.l^{-1}hr^{-1}$, 80 times better than batch fermentation productivity, was obtained at a dilution rate of 6 hr^{-1} with an inlet lactose concentration of 150 $g.l^{-1}$ and cell concentration of 150 $g.l^{-1}$.

2.3.3.1 Constraints on Concentration and Productivity

Limitations as to the cell, or slurry, concentration that can be used with CFMF have been reported. Although reported principally for CFMF systems, these limitations would also apply to other membrane systems.

The constraints may be physical, such as pumping circuit difficulty due to increased slurry viscosity at high concentrations. Other constraints may be inherent in a particular system. For example, the rate of oxygen transfer can limit the cell growth rate in aerobic rate of oxygen transfer can limit the cell growth rate in aerobic fermentations at very high cell concentrations (Enzminger and Asenjo, 1986).

Sreenath and Jeffries (1987) found that cell recycle did not sufficiently improve the rate of production for the process to become economically viable during their xylose fermentation using *Candida shehatae* ATCC 22984 to produce ethanol. The xylose fermenting organisms need to be modified to give them the properties found in *Saccharomyces* to improve the output.

The optimum level of reactor biomass concentration is also determined by the balance between reactor cost and membrane cost (Anderson *et al*, 1986).

Taniguchi *et al* (1987) used a ceramic microfiltration membrane for high concentration cultivation of lactic acid bacteria *Streptococcus cremoris* and *Lactobacillus casei*. With the removal of lactate, the period of logarithmic growth rate was extended and productivities of 19-fold and 9-fold, respectively, greater than those of corresponding batch fermentations were obtained.

Lee *et al* (1980) found that different strains of yeast actually fouled the membrane differently. This seems to indicate that the observation made by Hoffmann *et al* (1987), that UF membranes were better than CFMF membranes for the separation of a *Saccharomyces cerevisiae* broth was restricted to their particular strain and not generally the case. An average pore size of 0.02 μm is necessary to avoid "growing through" and to allow flux recovery after cleaning.

There is a fairly high energy saving when separation is performed by CFMF as opposed to high shear rate UF. Lee *et al* (1980) found CFMF specific energy consumption to be 4 $Wm^{-2}sA^{-1}$ compared to 1000 $Wm^{-2}sA^{-1}$ for UF. The low cost pressure housing and modular design of CFMF minimise problems with industrial scale-up.

Murkes (1986) mentioned the phenomenon of secondary membrane formation on the surface of the CFMF membrane due to the deposition of a layer of particles. He found that secondary membrane actually prevented particles smaller than the CFMF membrane pore size from passing through, allowing a membrane of larger pore size to be used. This is advantageous as the pressure requirements for driving the separation process are reduced due to the larger membrane pore size.

Hoffmann *et al* (1987) used membranes to improve bioreactor performance in *Zymomonas mobilis* fermentation. They developed relationships for calculating the specific substrate utilisation rate and cell mass concentration with cell retention, using data obtained from fermentation without cell retention. They also found that with high gas hold-up, the filtration flux decreased due to the membrane being hydrophobilised by the gas, having been hydrophilised by the ethanol.

Le (1987) investigated the CFMF recovery of beer from beer tank bottoms. During the investigation an interesting relationship between the flux and pressure was found. Two membranes with different pore sizes switched roles as being the one with the higher flux, on increasing the pressure. Le suggested that microfiltration occurs with bridging of particles across the filtration matrix. At high pressures such bridges may break up, resulting in a decrease in the efficiency of the filtration. The bridges appear to break at different pressures with different pore sizes. It is interesting that one can increase the flux at a particular pressure, purely by using the optimum pore size for that pressure. The upper bound is obviously when the pore size is too large to effect adequate separation.

2.4 CROSSFLOW FILTRATION IN BIOLOGICAL WASTEWATER TREATMENT

Initially interest in membrane separation techniques for wastewater treatment centered on tertiary treatment aspects. With the developments in biotechnological applications, however, there has been increasing interest in applying membrane technology for solids/liquid separation directly in conjunction with the biological wastewater treatment stage.

2.4.1 REVERSE OSMOSIS APPLICATIONS

In waste water treatment, RO is generally used for the final polishing and purification of the effluent water from the process.

The study of Wojcik *et al* (1980) showed that RO was a very effective process for the removal of bacteriological and chemical contaminants when combined with proper pretreatment of the feed water. The pretreatment section in the plant used by Wojcik *et al* comprised two clarifier and two biological digester stages.

Arora and Trompeter (1983) conducted work on the fouling of RO membranes in wastewater applications. The following were discovered :

(i) Processes that include biological treatment as well as treatment such as coagulation, flocculation and sedimentation result in far superior filtration than similar ones with no biological treatment.

(ii) The removal of food sources for bacterial growth by the biological system reduced the severity of organic fouling of the RO membrane. Chlorination was used to discourage bacterial growth, but some of the RO membranes were susceptible to chlorine attack, requiring dechlorination prior to filtration.

(iii) Inorganic fouling of the membrane was identical to that found treating normal water and not only wastewater. The inorganic fouling could be reduced by softening the water, removing iron and manganese and reducing the water recovery.

(iv) The researchers concluded that more research was required before the process could be universally recommended. The success that they had had was at fairly high expense due to the pretreatment required before the filtration step.

Slater *et al* (1983) reviewed research into RO for municipal wastewater treatment. The main uses were in the treatment of secondary and tertiary effluent as well as some work on the treatment of primary effluents and digester supernatants.

Their review included work conducted by Fischer and Lowell (1970) who used both flat plate and tubular RO membranes for the treatment of raw sewage, primary and secondary effluent, as well as digester supernatant. The filtration flux was maintained for 20 days whilst operating on primary and secondary effluent, although results were very dependent on the composition of the wastewater. The composition dependence is significant as Feige and Smith (1974) found that there was a rapid decline in product water flow when their RO system was fed secondary sewage.

The results obtained by Fischer and Lowell from digester supernatant were similar to those from raw sewage as the flux dropped to zero after two days. The high level of fine suspended and colloidal matter, was thought to be the main cause of the flux decline.

Membrane fouling was found (Lim and Johnston, 1976) to be more pronounced for membranes with a low rejection capacity. This means that RO with low filtration fluxes, due to high rejection, are the most applicable in wastewater treatment. To achieve the necessary high capacity a large membrane surface area would be required, at a correspondingly high expense.

Van den Huvel *et al* (1981) found that RO was an excellent method for concentrating primary settled, municipal wastewater prior to anaerobic digestion. The Biological Oxygen Demand, (BOD) was increased approximately seven times by RO concentration, before it was fed to an upflow anaerobic digester. This reduced the hydraulic load on the reactor.

In contrast to both Fischer and Lowell (1970) and Feige and Smith (1974), Van den Huvel *et al* found no rapid flux decline due to membrane fouling. This is probably due to the nature of the sewage used for the experimental work and although Van den Huvel *et al* were successful in their investigation, it seems that RO could not generally be used with the same success for filtration of primary sludge.

Yanagi and Mori (1980) used an advanced RO membrane process with automatic sponge ball cleaning for treating secondary sewage effluent on which Feige and Smith had had no success. The cost of membrane replacement was, however, triple the cost of the electrical power required by the process and the major operating expense.

One of the drawbacks of reverse osmosis as a separation process, other than high operating costs, is that trace metals are accumulated in the system. This leaves the filtrate deficient in these trace metals which often have to be replaced downstream.

2.4.2 ULTRAFILTRATION APPLICATIONS

The use of ultrafiltration membranes for the reclamation of wastewater has been investigated with varying degrees of success. The main problem, as found by Olsen and Haagensen (1983) and Sachs and Zisner (1977) amongst others, is the maintenance of a reasonable flux over an extended period with acceptable cleaning frequency. Recent developments in UF modules, have alleviated this problem to a great extent.

Olsen and Haagensen (1983) used UF to filter secondary effluent. BOD in this effluent was found to be reduced by over 50 per cent by the filtration process. They also investigated using the ultrafiltration filtrate as the feed for a RO plant. The capacity of the RO plant was improved above that possible when it was fed secondary effluent, but the cost of operating both systems was found to be economically unviable.

Inoue *et al* (1981) compared several different types of UF modules available for treating secondary effluent in place of the conventional coagulation, sedimentation and sand filtration. Increasing operation pressures and sponge ball cleaning of the modules was necessary to prevent clogging and maintain a reasonable flux. The researchers concluded that it would be necessary to evaluate lower pressure membrane processes for operating costs to become acceptable.

Arika *et al* (1977) operated an activated sludge ultrafiltration combined process for the reclamation of domestic wastewater. The only modification of the conventional treatment process was the substitution of the ultrafiltration system for the final clarifier. The process operation was considerably different due to the capabilities of the UF membranes.

The 15 $m^3.day^{-1}$ plant was operated successfully for ten months, during which time the effluent was of a good quality, viz. BOD < $2mg.l^{-1}$, COD (Mn) < 10 $mg.l^{-1}$ and no suspended solids and bacilli detected.

The objective of the research was to develop a compact sewage treatment system for processing waste, generated in a single building, to produce water for non-potable purposes.

Several factors influenced the design of the plant. Although the quality of the influent waste water from a single domestic source can vary considerably, the system had to maintain a high-quality effluent. The system had to be compact enough to be installed in the individual building or factory. The sludge formation had to be minimised, to minimise solids accumulation requiring periodic wasting. The process had to run without operator supervision to remain economically viable. Furthermore, the overall reclamation cost had to be low for the plant to be acceptable.

Activated sludge was continuously drawn from the aeration tank to feed the UF modules. An 80 mesh rotating screen was used to prevent gross objects and fiber entering the system. The product water was removed through the membranes as filtrate, whilst the filter reject was returned to the tank.

The system was designed for continuous operation, requiring the storage of factory effluent produced during the eight to nine hour working day. The design capacity was equivalent to the factory's minimum domestic waste output. The air flowrate to the 4 m^3 aeration tank was optimised during the first three months of operation and fixed at 36 m^3h^{-1}. During the same period, the velocity of turbulent flow over the membrane surface was fixed at 1,8 ms^{-1} and the operating pressure optimised. The process temperature was not controlled in any way, that of the influent varying between 6 and 26°C.

The influent BOD ranged from 20,7 to 421,3 $mg.l^{-1}$ and the COD between 24,0 and 333,0 $mg.l^{-1}$. During the ten month period, the effluent quality remained fairly stable, corresponding to a BOD removal of 99,1 per cent and COD removal of 93,6 per cent.

The membrane had a molecular weight rating of 24 000 , ensuring that all detectable suspended solids (0,45 μm and above) were retained. This meant that biologically non-degradable substances were accumulated gradually. During the test period, Mixed Liquor Suspended Solids (MLSS) increased by 92 kg and Mixed Liquor Volatile Suspended Solids (MLVSS) increased by 44 kg in the system, with a total incoming BOD of 535 kg.

There were several benefits derived from the system. The UF system completely separated the sludge from the water, eliminating the sludge bulking phenomenon often associated with nutrient removing activated sludge systems. The formation of excess sludge was minimised by the high digestion rate, reducing the operating costs. The system consists of only two units, requiring less frequent maintenance than a conventional system.

Because the plant was compact, an aeration tank retention time of six hours could be used because the final clarifier had been omitted. There was also greater freedom of layout.

The system was shown to be able to handle wide fluctuations in the influent quality as a result of the UF system maintaining a high level of MLSS in the aeration tank.

This research has indicated that a similar system, using a CFMF membrane, which would be operated below the 13,33 kPa driving pressure required by this system, could prove to be very successful. It would be interesting to compare the accumulation of non-biologically degradable material if a CFMF system were used in place of a UF system.

Ross *et al* (1989) used a UF membrane for solids/liquid separation in anaerobic digestion systems. The digester output was withdrawn from the top the digester and fed to the UF module. Concentrated sludge was then returned to the digester. The sludge recirculation provided sufficient mixing in the digester to enhance its performance. The South African produced UF unit (Bintech (Pty) Ltd. Paarl) was operated at a pressure of 500 kPa. The membrane had an asymmetrical structure with a thin skin of pore size 0,001 to 0,1 micron supported by a more porous layer. The linear velocities used were in the region 2,0 to 3,5 $m s^{-1}$.

The Anaerobic Digester Ultrafiltration (ADUF) process was completely enclosed reducing odours in the vicinity of the plant. No strict control of the Solids Retention Time (SRT) was necessary, the solids concentration merely being controlled by its influence on the permeate flux from the UF unit. The size of the external UF unit was able to be increased to allow a greater hydraulic throughput.

The researchers also emphasised that the choice between a mesophilic (35°C) and a thermophilic (55°C) digestion system was important in the design. At the higher temperature, the sludge viscosity is reduced and the flux from the UF unit is higher as a result.

2.4.3 CROSSFLOW MICROFILTRATION APPLICATIONS

Membrane filtration has been an available alternative for some time, though there have been problems with its implementation. The cost of a membrane system is directly proportional to the rate of clean water production. With a low-cost product such as water, both ultrafiltration and reverse osmosis systems have proved to be too expensive. The flux rate was low and the heavy membrane fouling required frequent replacement/regeneration of the modules. The microfiltration membrane has become an economic option due to the increased pore size significantly raising the flux. The system is also operated at a much lower pressure, reducing energy requirements (Tran, 1985).

Bindoff *et al* (1987) have used a CFMF process developed in South Africa to concentrate waste activated sludge and anaerobic sludge from a conventional sewage works. In this system the membrane comprised a tubular woven cloth with a precoat of 15 μm limestone.

The researchers found that activated sludge could be concentrated from 0,5 to 5 per cent total solids (m/v). The permeate quality did, however, deteriorate at the higher feed concentrations and with the duration of the run.

The solids content of a fill-and-draw pilot-scale anaerobic digester was increased from 2,6 to 5,5 per cent (m/v) total solids, using CFMF. The organic loading to the digester was doubled (1,8 to 3,1 kg volatile solids per m^3 per day) the solid retention time being held constant at 26 days and the liquid retention time decreased to 14 days. The permeate from the CFMF unit was actually better than the supernatant from a comparable digester (122 versus 570 $mg.l^{-1}$ suspended solids).

In a later report, Bindoff *et al* (1987) state that the operating load on a 2300 l pilot digester, operating on various sludge streams, was increased from 3,6 kg volatile solids per day to 7,7 kg volatile solids per day. There were no ill effects on the digester performance over the test period of 213 days, the permeate quality remaining good. The volatile acid/alkalinity ratio was maintained at 0,15, whilst the destruction of volatile solids remained at 63,4 %, highlighting the successful operation of the pilot digester coupled to the crossflow microfilter.

Krauth and Staab (1988) investigated the substitution of the final clarifier by membrane filtration within an activated sludge system that was operating at increased pressure. The system is operated at increased pressure to improve the oxygen supply.

Tests were carried out on four membrane types in the study; on the basis of pore size it appears that three could be regarded as UF and one as CFMF membrane. Effective solid/liquid separation was attained with four membranes. It is of interest, however, to note that problems of membrane fouling were most predominant with the CFMF and not the UF membranes.

2.4.4 OTHER SYSTEMS OPERATING WITH RETAINED BIOMASS

Anderson *et al* (1986) investigated the use of porous membranes for retaining the biomass in a biological wastewater treatment process. Two reactors, with porous membrane liners, enabled the acidogenic and methanogenic steps in anaerobic digestion to be carried out separately at their own optimal operating conditions.

Optimum, or maximum, treatment capacity of the system was not reached, although the maximum feed COD concentration was 100 000 $mg.l^{-1}$, which is generally higher than most industrial wastewaters. To increase the organic loading rate, the feed rate was increased, but the reactor effluent was filtered to retain the solids in the reactor for the required Solids Retention Time. In this way, the SRT is greater than the Hydraulic Retention Time (HRT). The maximum attainable filtration rate of 0,77 $l.m^{-2}h^{-1}$ proved to be the limiting factor in their experimental work.

Anderson *et al* suggest that better solid-liquid separation, as in crossflow microfiltration, would enable the loading rate to be increased.

2.4.5 SUMMARY OF CROSSFLOW SYSTEMS USED IN BIOLOGICAL WASTEWATER TREATMENT

The information summarised in the tables below is categorised into the three membrane types RO, UF and CFMF. As much operating detail as possible is given. The data give some indication of the wide variation in operating conditions and the fluxes that were obtained from the different membrane systems.

TABLE 2.2 : REVERSE OSMOSIS MEMBRANES

AUTHOR	FILTER DEVICE	APPLICATION	OPERATING PARAMETERS
Fischer and Lowell (1970)	Flat Plate and Tubular Cellulose Acetate membranes	Raw Sewage; Primary and Secondary effluent; and Activated Sludge	Flux only maintained during filtration of Primary and secondary effluent. Otherwise membranes heavily fouled. No operating Conditions given.
Thomas and Mixon (1972)	Cellulose Acetate Membranes	Primary Effluent	Initial flux 66,7 to 75.0 $l.m^{-2}h^{-1}$ Pressure = 5520 to 6900 Pa Optimum Velocity between 0,25 and 0,50 ms^{-1}. Below this fouling severe.
Hart (1985)	ZF99 tubular membranes	Abattoir Effluent	Pressure =2500-5000 kPa Flux = 0,2 $lm^{-2}h^{-1}bar^{-1}$ TSS = 65 $g.l^{-1}$

TABLE 2.2 : REVERSE OSMOSIS MEMBRANES (Continued)

AUTHOR	FILTER DEVICE	APPLICATION	OPERATING PARAMETERS
Tsunge and Mori (1977)	Nitto Electrical Cellulose Acetate Membranes	Secondary Treated Effluent	Pressure = 4119 kPa Feed Flow = 0,43 $l.s^{-1}$
Wojcik *et al* (1980)	Cellulose Acetate	Secondary Treated Effluent	Feed TDS = 1 $g.l^{-1}$ Flowrate = 0,76 $l.s^{-1}$
Yanagi and Mori (1980)	NTR-50	Secondary Municipal Effluent	Pressure 1961 kPa Serious Fouling
Yanagi and Mori (1980)	Tight NTR-90 With spongeball cleaning	Secondary Municipal Effluent	Pressure = 3,92 kPa I.D. = 12 mm Length = 2,5 m Flux = 27,08 to 31,25 $l.m^{-2}h^{-1}$

TABLE 2.3 : ULTRAFILTRATION MEMBRANES

AUTHOR	FILTER DEVICE	APPLICATION	OPERATING PARAMETERS
Arika *et al* (1977)	IOPOR Type XP-24	Activated Sludge	Velocity = 1,8 $m.s^{-1}$ Pressure = 133,3 kPa Flux = 28,54 $l.m^{-2}h^{-1}$ Effluent SS = 0,0 $g.l^{-1}$ No membrane cleaning for four months
Inoue *et al* (1981)	Hollow Fiber (SF 301) Polyvinyl Alcohol Pore size 0,04μm	Secondary Sewage Effluent	Pressure = 100 to 500 kPa Velocity = 0,43 - 0,93 $m.s^{-1}$ Flux = 0,10 - 0,18 $m^3.h^{-1}$ Backflushing used
Inoue *et al* (1981)	Tubular (MRT-104) Polyacrlonitrile Pore size gives 90 % rejection of 13000 MW	Secondary Sewage Effluent	Pressure = 100 to 500 kPa Velocity = 1,0 - 2,4 $m.s^{-1}$ Flux = 0,16 $m^3.h^{-1}$
Krauth and Staab (1988)	Abcor HFM251 Abcor HFM276 Enka Microdyn Hoechst E11 (25 000 MW cut)	Activated Sludge	The Hoechst E11 membrane found to be superior than others. E11 operation data : Pressure = 200 kPa Velocity = 8,5 $m.s^{-1}$ Flux = 290 $l.m^{-2}h^{-1}$ Pressure = 300 kPa Velocity = 8,5 $m.s^{-1}$ Flux = 445 $l.m^{-2}h^{-1}$ MLSS = 3,27 $g.l^{-1}$

TABLE 2.3 : ULTRAFILTRATION MEMBRANES (Continued)

AUTHOR	FILTER DEVICE	APPLICATION	OPERATING PARAMETERS
Olsen and Haagensen (1983)	Plate and Frame GR61PP	Municipal Secondary Effluent	Pressure = 300 kPa Influent = 0,06 $l.s^{-1}$ Flux = 70 $l.m^{-2}h^{-1}$
Ross *et al* (1989)	Tubular Asymmetric Polyethersulphone (Bintech, Paarl)	Anaerobic Digester Sludge	Pressure = 500 kPa Velocity = 2,0 - 3,5 $m.s^{-1}$ Flux = 300 $l.m^{-2}h^{-1}$ At 25°C TSS = 43 $g.l^{-1}$
Sachs and Zisner (1977)	Tubular HUF33040 With sponge ball cleaning	Oxidation Pond Effluent	Pressure = 80 kPa Velocity = 2,99 $m.s^{-1}$ Flux = 75 $l.m^{-2}h^{-1}$

TABLE 2.4 : CROSSFLOW MICROFILTRATION MEMBRANES

AUTHOR	FILTER DEVICE	APPLICATION	OPERATING PARAMETERS
Bindoff (1985)	Tubular Woven Polyester with precoat : Diatomaceous Earth Limestone Limestone and Cellulose	Primary Digested Sludge	Pressure = 250 kPa Velocity = 2,0 to 2,3 $m.s^{-1}$ Best fluxes were obtained with a limestone precoat Flux = 69 $l.m^{-2}h^{-1}$ SS = 3 $g.l^{-1}$
Bindoff *et al* (1988)	Woven Fabric Tubes with precoat : Limestone 25 μm Tube Diameter = 25 mm	Waste Activated Sludge Anaerobic Digested Sludge	Pressure = 100 - 600 kPa Velocity = 1,0 - 3,0 $m.s^{-1}$ Flux = 91,8 $l.m^{-2}h^{-1}$ TSS = 5 $g.l^{-1}$ Flux = 18,3 $l.m^{-2}h^{-1}$ TSS = 24 $g.l^{-1}$
Treffery-Goatley *et al* (1987)	Polyester Fabric Woven Tubes (Schweiz Seidengazefabrik AG Thal-Switzerland) Diameter = 12 mm with precoat : Limestone 15 μm Limestone 15 μm + alum Alum	Clarification of River Water Tertiary Sewage Works Effluent	Pressure = 180 kPa Velocity = 1 $m.s^{-1}$ Best flux was achieved with limestone-alum mix Flux = 380 $l.m^{-2}h^{-1}$

2.5 CONCLUSIONS

The following aspects are apparent from a review of membrane filtration applications in filtration and biological wastewater treatment :

(i) RO has been applied successfully in desalination applications where the feed water is clear of solids and colloidal matter; with these present fouling problems are encountered. RO is not suited for solid/liquid separation.

(ii) The application of both UF and CFMF membranes for solid/liquid separation has been demonstrated by the successful operation of biological systems.

(iii) In certain fermentation applications with UF rejection of toxins, and accumulation in the bioreactor, has lead to inhibition at high cell concentrations. This effect does not occur with CFMF.

(iv) At high cell concentrations in both UF and CFMF fermentation systems, product inhibition has been observed. This is unlikely to be a consideration in wastewater treatment applications where the objective is to minimise soluble product formation.

(v) In aerobic biological systems operated at high cell concentrations difficulties have been encountered with supplying sufficient oxygen.

(vi) In general CFMF has both lower capital and operating costs than UF, and for this reason would appear more suited to wastewater treatment applications. In both fermentation and wastewater treatment applications cases have been reported where UF membranes have not been fouled as readily as CFMF membranes. It would appear that the nature of the biological matter is crucial in determining which system is preferable.

(vii) Detailed information on the influence of CFMF membrane pore size and/or precoat on filtration performance is generally unavailable in the literature. Also, the reported values of attainable filtrate flux and frequency of membrane cleaning vary widely. Because these factors are critical in determining the viability of using membrane filtration, attention should be directed at determining realistic performance data.

CHAPTER THREE

EXPERIMENTAL CROSSFLOW MICROFILTRATION SYSTEM

3.1. EXPERIMENTAL EQUIPMENT

3.1.1 CROSSFLOW MICROFILTRATION UNIT

3.1.1.1 Membrane

The CFMF membrane used in this investigation was a woven polyester tube (Schweiz Seidengazefabrik AG, Thal) with a 12 mm diameter and active length of 1,00 m. The tubing became rigid and fairly self-supporting when liquid is pumped through it under pressure. The membrane has been used in the filtration of various wastewaters by Bindoff *et al* (1987) at Natal University.

A precoat layer is deposited on the inner surface of the membrane prior to the filtration of a biological slurry. The layer prevents the biological material from coming into contact with the filter cloth, minimising biological fouling. The precoat can largely determine the filtration flux and quality of the filtrate due to the nature of the filter cake that is formed.

The membrane was clamped onto two PVC nipples, teflon tape being used to prevent leakage from either end. The inlet-side nipple was fitted with a gland that allowed it to be moved horizontally, enabling the membrane to be stretched taut.

Approximately 15 cm of both ends of the membrane were coated with a rubber sealant (Shoe Patch, distributed by Game, Set and Match, South Africa) to eliminate end effects. The turbulence at the entrance and exit of the module tends to disturb the precoat layer, lowering filtrate quality due to resultant pinholing.

1,33 m of PVC guttering beneath the membrane collected the filtrate along its length. A removable section of similar guttering could be inverted and fitted onto the lower PVC gutter to form a housing which lowered the rate of evaporation from the membrane surface. As a large amount of light was blocked out, the growth of photosynthetic aerobic biological material on the outside of the membrane was discouraged.

3.1.1.2 Pressure Gauge

A pressure gauge (Capegauge Instrumentation, Range 0 - 600 kPa) was fitted with pressure tappings on the influent and effluent sides of the membrane. This enabled the pressure drop over the membrane length to be determined and an average transmembrane pressure to be estimated.

3.1.2 CFMF UNIT PUMP

A C32M Monopump (Wm. Spilhaus & Co. (WP) (Pty) Ltd) was used to pump liquid through the CFMF unit. The pump was fitted with a variable speed drive to enable a wide variety of pumping pressures. and velocities to be selected. The maximum operating pressure was 600 kPa, the pump being able to deliver 1,30 $m^{-3}h$. The pump was generally operated at a pressure between 200 and 300 kPa and flowrate of 0,79 $m^{-3}h$.

3.1.3 PUMPING CIRCUITS

There are essentially two different circuits that pass liquid through the CFMF module - the precoat and filtration circuits. Selection of pumping circuit was achieved via a two-way PVC valve at the inlet and outlet of the filter tube, to adjust concentrate flow, and a two-way valve on the line from the PVC membrane housing to vary the membrane filtrate return between the precoat tank and reactor.

3.1.3.1 Precoat Circuit

The C32M Monopump was used to pump the precoat slurry, contained in the precoat tank, through the CFMF module to deposit a precoat layer on the surface of the membrane. This was performed prior to filtration of the biological slurry.

The precoat circuit operated on a complete recycle basis, the concentrate and filtrate both being returned directly to the precoat tank. The filtrate flowed under gravity.

The precoat tank was manually agitated during the precoating process to ensure that the precoat did not settle out of the suspension.

3.1.3.2 Filtration Circuit

Once the precoat layer had stabilised, the flow from the precoat tank was stopped and the biological slurry, from the bioreactor, was fed to the CFMF unit via the monopump.

The concentrate was returned directly to the top of the reactor, whilst the filtrate flowed, under gravity, into a surge tank.

The surge tank had an overflow pipe approximately 10 cm from the base to return excess filtrate back to the top of the reactor. The remainder of the filtrate was pumped from the surge tank by the Watson Marlow pump, as system effluent, at the same rate as the feed.

3.2. EXPERIMENTAL METHOD

3.2.1 MEMBRANE PRECOATING

Selection of the precoat that was used in this experimental work is discussed in Section 3.3. In this section only the general method used to deposit the precoat layer on the membrane is discussed.

15 g of the precoat was mixed with approximately 16 litres of standard potable water in the precoat tank. The suspension was manually agitated during precoating to prevent the precoat from settling out.

The C32M monopump was started, valves having been selected so that complete recycle through the CFMF unit back to the precoat tank was possible. The operating pressure and velocity was the same as that chosen for the filtration of the biological slurry.

The slurry was pumped through the CFMF unit for 15 minutes, by which time the precoat layer had stabilised. The pump influent was then changed from the precoat slurry to the effluent from the bioreactor. The valve had to be switched over as quickly as possible to prevent a pressure bump that would strip the precoat off the membrane surface.

Once the dark biological slurry had passed through the CFMF membrane (visible through the clear hose following the CFMF membrane a few seconds after changeover), the concentrate return was switched from the precoat tank to the reactor.

At this stage the filtrate return was directed to the surge tank.

It was important to wait until the biological slurry had been pumped through the CFMF unit to prevent precoat slurry from entering the reactor. Obviously there was still a small volume loss/gain in the reactor, but this was corrected for during make-up to account for evaporative losses.

The precoat tank was drained and filled with clean water in preparation for membrane cleaning.

3.2.2 MEMBRANE CLEANING

Depending on the quality of the precoat layer, the filtration flux declined below the required 15 $l.day^{-1}$ after a period of between one and four days.

For the cleaning operation the pump influent was changed from the reactor effluent to clean water contained in the precoat tank. Once the water had passed through the CFMF unit, the valves were switched to return it as well as the filtrate, to the precoat tank.

The weir valve after the membrane unit was fully opened to reduce the pumping pressure. This allowed the tube to be pinched manually. Pinching the tube increased the surface shear, stripping the precoat from the membrane.

The water in the precoat tank was replaced periodically. Once all of the precoat layer had been removed and the filtration flux returned to its initial clean water flux rate, the precoating process discussed in Section 3.2.1 was repeated.

3.3 PRECOATING TRIALS

Several filtration tests were conducted using waste activated sludge (WAS) prior to the coupling of the CFMF unit to the UASB reactor. WAS was chosen so that the operation of the CFMF unit constructed for this investigation could be compared with operating data collected by Bindoff *et al* at Natal University using the same type of membrane to filter WAS.

A further investigation into the actual type of precoat required was necessitated by the low filterability of the UASB biological slurry. This investigation was performed after the CFMF unit had been coupled with the UASB reactor.

3.3.1 PRESSURE REQUIREMENTS

Initially, a centrifugal pump was used as the pressure pump for the CFMF unit. It was able to generate a crossflow velocity of 2 $m.s^{-1}$ at a maximum inlet pressure of 150 kPa.

Bindoff (1988) mentions that one of the requirements for a thick, stable precoat is a minimum outlet pressure of 150 kPa, so the crossflow velocity was decreased to 1 $m.s^{-1}$, to enable operation at pressure of 170 kPa. The pressure drop over the membrane was still between 30 and 40 kPa, resulting in the membrane outlet pressure being below the required 150 kPa. In addition, the flow generated by the pump, under these operating conditions was unstable.

Filtration experiments were performed at both sets of operating conditions, using waste activated sludge and it was found that the fluxes after 24 hours were typically in the region 11 $l.m^{-2}h^{-1}$, less than the required 15 $l.m^{-2}h^{-1}$.

When the results were compared with those of Bindoff (1985), they appeared to be similar to the data that was obtained when no precoat was used, (13,5 to 14,2 $l.m^{-2}h^{-1}$). This indicated that a stable precoat layer had not been established prior to filtration.

The centrifugal pump was replaced with a C32M monopump with a variable speed drive that allowed flexibility in the setting of operating conditions.

During the remainder of the experimental work, a pressure of 250 to 300 kPa, with a minimum crossflow velocity of 2 $m.s^{-1}$, was used. The flow from the monopump was a great deal more stable than that from the centrifugal pump and the outlet pressure was well above the minimum, facilitating effective precoating.

3.3.2 INITIAL PRECOAT CONCENTRATION

Although the same precoat operating pressure and crossflow velocity conditions as used by Bindoff (1985) were applied, the results obtained from waste activated sludge filtration still indicated that a stable precoat layer had not been formed.

Various operating pressure, crossflow velocity combinations were investigated in an attempt to find conditions where a stable precoat could be formed. The variation in flux was negligible.

At this stage the volume of the precoat tank was decreased from 30 l to 16 l to determine whether the initial concentration of the precoat slurry had any effect. Up until this point 1,9 g of precoat was used to make the precoat slurry in the precoat tank. This corresponded to a filter loading of 0,05 $kg.m^{-2}$ as specified in the operating conditions given by Bindoff (1987). In Bindoff (1988) the loading for a Kulu 15A precoat was, however, given as 0,3 to 0,4 $kg.m^{-2}$. For this loading approximately 15,2 g of precoat was required. The fluxes obtained, when a precoat slurry of this concentration was used, were dramatically higher as shown in Fig. 3.1.

This indicates the significant dependence of the formation of a thick, stable filter cake on the initial concentration of the precoat slurry.

3.3.3 TYPE OF PRECOAT

The different types of precoat that were investigated in this experimental work are discussed in the following sections. Due to the

low filterability of the UASB biological effluent, it was necessary to find a precoat that was able to maintain the filtrate flux above the required 15 l.day^{-1} (15 l.m^{-2}h^{-1})for a minimum of 24 hours.

The flux-time relationships for the various types of precoat are shown in Fig. 3.2.

3.3.3.1 Kulu 15A Limestone (Lewis and Everett)

It was found that the "steady state" flux, obtained after several hours' filtration of the UASB effluent, was unacceptably low when Kulu 15A limestone, was used as precoat.

After 24 hours the flux had generally deteriorated to below the required 15 l.m^{-2}h^{-1}, as shown in Fig. 3.2. The filtrate quality was, however, still very good - essentially no suspended solids being detected.

A particle size analysis was performed using a Malvern Particle Size Analyser (Malvern Instruments). The D(50%) was found to be 17 μm. Detailed results of the size analysis are given in Appendix 4.

3.3.3.2 Diatomaceous Earth (Holpro Chemical Corp.)

Bindoff (1985) investigated the precoat properties of various types of filter aids. In the experimental work, diatomaceous earth (DE)

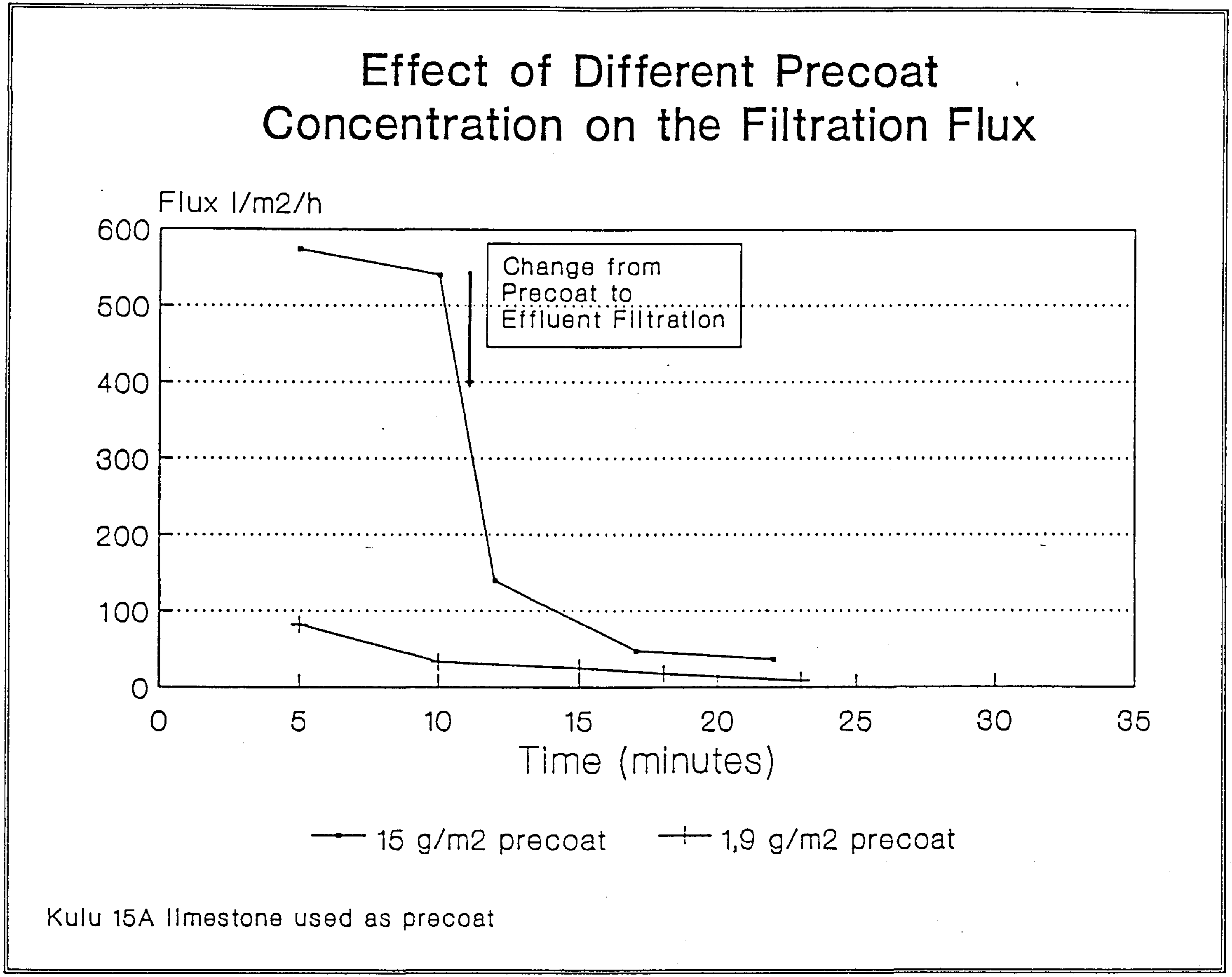


FIGURE 3.1 : Effect of Different Precoat Concentration on the Filtration Flux

gave fluxes that were comparable to those when limestone was used as the precoat material.

Due to the difference in the nature of the particles of the two materials, it was hoped that the different type of precoat formed by the diatomaceous earth precoat would better aid the filtration.

Diatomaceous earth (Holpro Chemical Corporation), at size specification of 14 to 19 per cent retention on a 25 μm screen, was selected for a trial filtration experiment. 15 g in 15 l of tap water, as previously, was used to make the precoat slurry. The same pump setting, as in the Kulu 15A limestone trials, was used (velocity 2,5 $m.s^{-1}$; pressure 250 kPa).

It was found that the flux during precoating was extremely high, with the result that the crossflow velocity was very low. All of the diatomaceous earth flowing into the membrane tube was deposited at the outlet, where the crossflow had become negligible. This eventually lead to the complete blocking of the tube.

The membrane was cleaned and a new slurry made. The pump rate was increased from 50 to 70 per cent, to provide the necessary crossflow velocity and operating pressure, to ensure that the precoat formed uniformly. The flux-time relationship is shown in Fig. 3.2. The filtrate quality was extremely poor, hardly any separation appearing to have been affected.

The filtrate flux actually increased immediately after the change from precoating to effluent filtration, thereafter the flux declined rapidly and the run was terminated. The increase in flux on changeover, followed by the sharp decline, indicates that a substantial amount of the protective precoat layer had been removed and the membrane cloth rapidly fouled. The particles were larger than the Kulu 15A particles.

In a separate experiment slurries,of the same concentration (by mass), were made with the various precoats. The slurries were then filtered through Whatman Number 1 filter paper in Buchner funnels

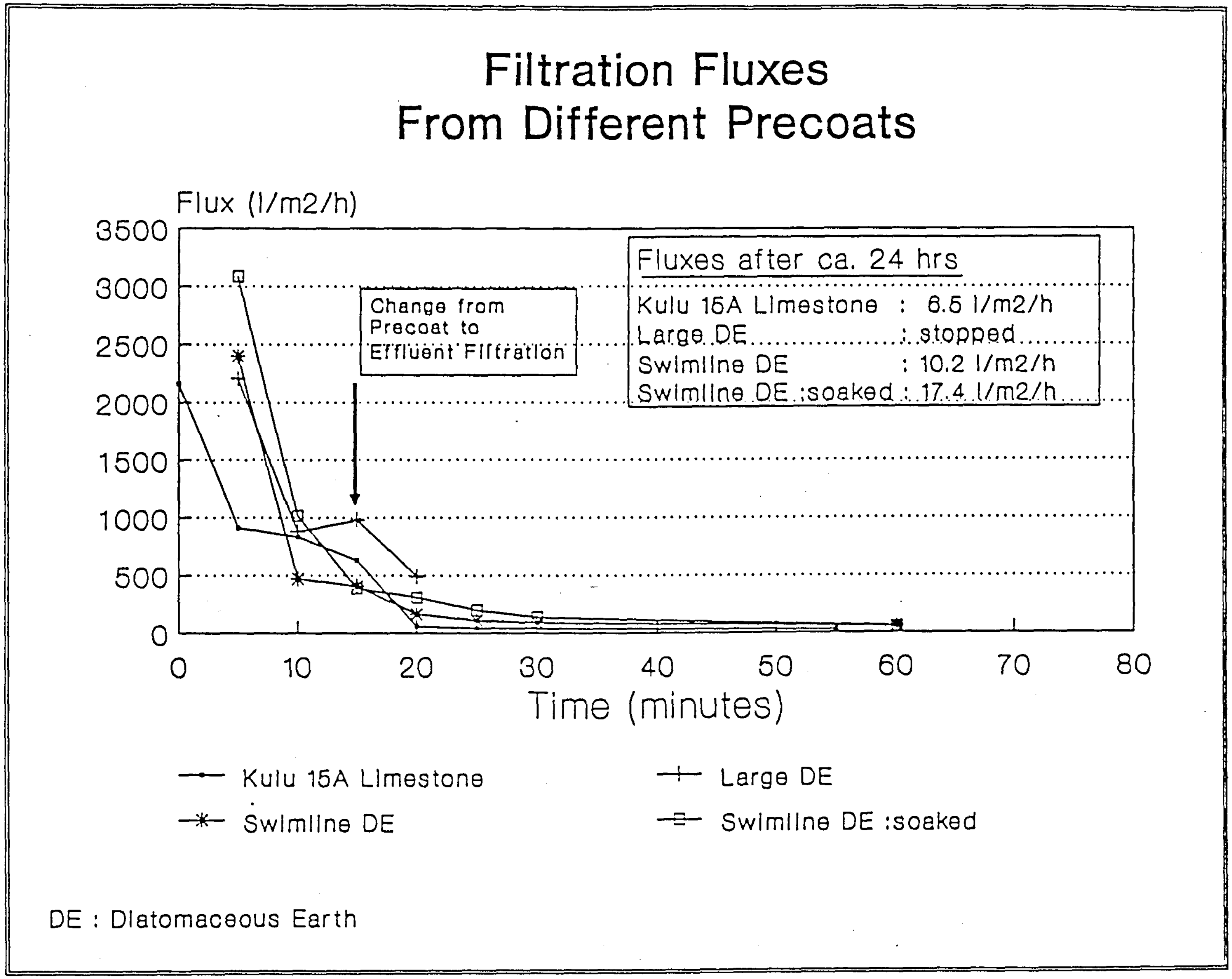


FIGURE 3.2 : Filtration Fluxes from Different Precoats

using the same pressure difference. The resulting filter cakes were dried at room temperature.

The Kulu 15A limestone filter cake was much more compact than that obtained from the diatomaceous earth, due to the small particle size. It was also much thinner due to the difference in densities.

The diatomaceous earth cake was more friable for its comparative thickness, indicating that the precoat layer formed on a CFMF membrane was likely to be less stable. This had been found to be valid in the CFMF filtration trial discussed above.

3.3.3.3 Swimline Diatomaceous Earth (Swimline,(Pty) Ltd.)

Swimline Diatomaceous Earth, with a smaller average particle size than the Holpro Diatomaceous Earth, was investigated to ascertain whether smaller particles would improve the precoat properties.

A flux-time relationship is shown in Fig. 3.2. The filtrate flux was slightly higher than that obtained with the Kulu 15A.

The initial filtrate quality was fairly poor, though substantially better than that obtained with the Holpro diatomaceous earth (DE). After approximately one hour, the filtrate quality had improved significantly, becoming comparable to that obtained when using Kulu 15A.

The phenomenon of increased flux quality with time when diatomaceous earth was used as a precoat was also observed by Bindoff (1985).

The Holpro DE manufacturer's specifications indicate that the diatomaceous earth water absorption is 280 per cent. The absorption of water was thought to be the underlying reason why the filtrate quality improved significantly within an hour. As the DE absorbs water it would swell, decreasing the poresize in the precoat layer, enabling a better separation to be effected.

Trials were conducted with diatomaceous earth that had been presoaked in water for an hour, before being added to the precoat tank.

A higher "steady state" flux was obtained with the soaked Swimline DE precoat than when it had been added directly to the precoat tank at the start of precoating.

The pre-swollen DE is able to form a precoat layer with the smaller pore size during the precoating process. This prevents excessive blinding of the pores by biological particles once filtration commences. If swelling occurs once the filtration has commenced, biological particles enter the precoat matrix and become trapped as the pore size decreases, blinding the pore and lowering the "steady state" flux that can be obtained.

Presoaked Swimline DE was chosen for use in the remainder of the experimental work on filtration of UASB effluent. It was found that the system could be operated for up to four days, before the filtrate flux became unacceptably low and reprecoating became necessary.

The Swimline DE did not appear to be pure DE, as it also contained pink fibrous material. This fibrous material probably enhanced the stability of the precoat layer. A sample of the layer collected from the inner surface of the membrane showed two distinct layers - the coarser, fibrous material in contact with the membrane and the finer, white DE material in contact with the fluid.

A Malvern Particle Size Analysis conducted on the Swimline DE gave a D(50%) of 50 μm. Detailed results are given in appendix 4. This was significantly larger than the average size of the Kulu 15A particles. The different nature of the filter cake that was formed due to the fibrous material present in the Swimline DE probably lead to the comparable separation efficiency. It was found that the size distribution was unimodal - i.e. a distinct size difference between the fibrous and not fibrous particles was not indicated.

Although the operating pressure of 200 kPa was similar to that used in Kulu 15A filtration, the crossflow velocity was 1 to 2 m.s^{-1} higher towards the end of the runs, when filtrate flux was low. As filtration flux is dependent on crossflow velocity, the increase probably also contributed to the higher fluxes. As the required separation was being achieved, the operation of the system with a Kulu 15A precoat with the higher velocity was not investigated.

3.3.4 ELECTRON MICROGRAPHS OF PRECOAT LAYER

Electron micrographs were taken of the precoat layers formed by Kulu 15A limestone and Swimline Diatomaceous Earth.

Micrograph 1 in Fig. 3.3 and Micrograph 1 in Fig. 3.4 show the different surfaces formed by Kulu 15A limestone and Swimline DE respectively, both magnified 1000 times. The former is more densely packed, due to the increased number of smaller particles that are evident.

The other major difference between the cakes is the honeycomb structure of the DE. The pores are approximately 2 μm in diameter. This structure probably helps in the formation of a filter cake with a high porosity. This structure is clearly shown in micrograph 2 of Fig. 3.4 which was taken at 1500 times magnification.

As discussed in Section 3.3.3.3, the Swimline DE contains two types of material. The pink, fibrous material formed a layer directly on the surface of the membrane, whilst the DE was deposited in a layer above this. The cross-section of precoat layer in micrograph 2 of Fig. 3.5 shows the interface between these layers at a magnification of 100 times. The difference in porosity of the layers is clear, the DE layer being far more compact, as the particles are smaller.

1. Kulu 15A Precoat
   (magnification 1000)

2. Kulu 15A Precoat
   (magnification 1000)

FIGURE 3.3 : Electron Micrographs of Precoat Layers

1. Swimline DE
   (magnification 1000)

2. Swimline DE
   (magnification 1500)

FIGURE 3.4 : Electron Micrographs of Precoat Layers

1. Swimline DE
(magnification 3500)

2. Swimline DE Interface
(magnification 100)

FIGURE 3.5 : Electron Micrographs of Precoat Layers

3.3.5 CONCLUSIONS FROM PRECOAT STUDY

1. The best fluxes in this study were obtained when using a commercially available swimming pool diatomaceous earth (Diatom Powder, Swimline (Pty.) Ltd.). This material appears to contain two fractions : a fibrous fraction which forms a layer against the cloth membrane and a fine particular fraction which builds up over the fibrous layer; that is, in contact with the slurry mix.

2. When the diatomaceous earth is used as a precoat, the filtrate quality is initially poor, but increases appreciably within the first 30 to 60 minutes of filtration until a solids-free filtration is observed. By presoaking and swelling the diatomaceous earth prior to precoating a better filtrate flux is obtained due to improved precoat formation before blinding by biological material can occur.

3. The initial concentration of precoat material in the precoat tank appears to exert a major influence on the nature of the precoat layer formed and the resulting flux. A slurry concentration of approximately 1 g of precoat per litre was found to be adequate.

CHAPTER FOUR

UASB EXPERIMENTAL INVESTIGATION

4.1. EXPERIMENTAL EQUIPMENT

A schematic diagram of the Upflow Anaerobic Sludge Bed reactor and crossflow microfiltration, (CFMF) unit is given in Fig. 4.1.

4.1.1 UPFLOW ANAEROBIC SLUDGE BED (UASB) REACTOR

The UASB reactor was constructed from PVC and comprised a 10 cm diameter, vertical cylinder, 95 cm high. The cylinder was sealed at the bottom. An open frustum 22 cm high and reaching a maximum diameter of 50 cm was joined to the top of the reactor column. The operating volume of the reactor was approximately 9 litres.

4.1.1.1 Feed to the UASB Reactor

The synthetic wastewater feed was fed to the reactor from the feed tank by a Watson Marlow 502 S peristaltic pump (Aeromix (Pty) Ltd.) fitted with a double 303 pumphead. The double head allowed the system influent and efluent to be balanced. The pump rate could be varied between 0 and 100 per cent of full capacity - the capacity being dependent on the diameter of the tubing used. The normal setting was at a delivery rate of 15 l.day^{-1}.

After the pump, the feed line was split into four separate streams connected to four symmetrically positioned inlets at the bottom of the UASB reactor. This ensured that the feed was well distributed at the base of the sludge bed.

4.1.1.2 Slurry Removal and Return

The slurry was removed from 20 cm below the top of the cylinder and pumped via the C32M monopump through the CFMF unit.

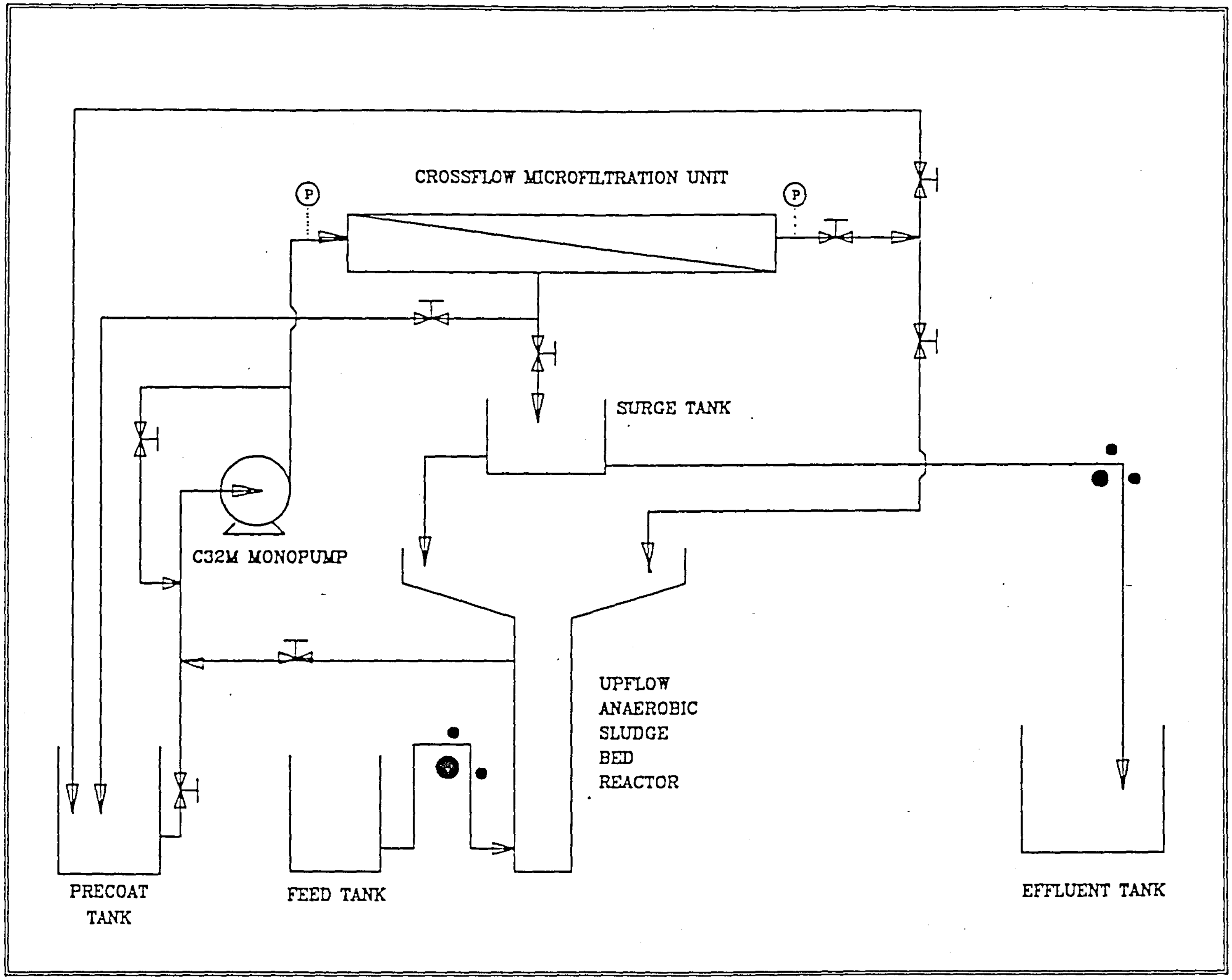


FIGURE 4.1 : Schematic Diagram of UASB Reactor and Crossflow Microfiltration Unit

The section of pipe between the reactor and the pump inlet was steel reinforced to prevent collapse of the hose due to the strong suction of the positive displacement pump. The hose to and from the Monopump was flexible to minimise the vibrations transmitted to the rig that tend to disturb the membrane precoat.

Reject from the CFMF unit was returned to the top of the reactor, tangentially at the rim of the frustum. The induced vortex-like flow of the returned stream was designed to dissipate the energy of the stream returned from the CFMF unit, so as not to disrupt the pelletised sludge bed at the bottom of the reactor.

4.1.1.3 Temperature Control

The temperature was controlled at 30°C by a temperature controller built by the Chemical Engineering Department, University of Cape Town.

Two heating tapes, (HCS 30W, 220 V, Wireohms) were fixed to the outer surface of the reactor using glass insulation tape (3M), one around the lower half and one around the upper half of the reactor. The heating tapes were operated independently by the controller using on-off control. The controller took readings from two thermistors placed in the reactor. One of the thermistors was positioned 38 cm from the bottom of the cylindrical section, coinciding with the top of the sludge bed. The other thermistor was held in a polystyrene block floating on the surface of the liquid in the frustum. The dual control enabled the temperature of the sludge bed and the frustum section to be controlled separately.

When a single control system and probe was used, it was found that the temperature of the cylindrical part of the reactor, close to the sludge bed, would reach 30°C a short while after the heating tape was switched on. The large volume of liquid in the frustum was hardly affected. This presented a problem as the temperature gradient increased at night during winter, (Room temperature below 10°C) forcing the bed temperature below the set point. The heating coil would then provide insufficient heat even when on continuously. This problem was eliminated when the liquid in the frustum maintained at the set point at all times. Once filtration started, heat was added to the liquid in this section by the pump.

Copper tubing was coiled in the frustum and in the reactor, down to the level of the slurry outlet pipe. Water was circulated through the pipe to remove the heat generated by pumping once filtration started.

4.2 EXPERIMENTAL METHOD

4.2.1 SYNTHETIC WASTEWATER FEED

The synthetic wastewater fed to the UASB reactor comprised glucose, essential trace metals, nutrients and a buffer. This growth medium has been used extensively for research on UASB systems by the Department of Civil Engineering, University of Cape Town.

Once the system had stabilised, 15 litres of the wastewater, containing material with a COD concentration of 5000 $mg.l^{-1}$, was fed to the reactor each day.

The buffer ($NaHCO_3$), was used to maintain the pH at the top of the reactor at approximately 7,2. This prevented inhibition of the microorganisms by the acids produced by the acidogens.

The composition of the feed is detailed in Appendix 1.

4.2.2 UASB REACTOR OPERATION

4.2.2.1 Chemical Oxygen Demand

The stability of the reactor was determined by monitoring the percentage removal of the COD material. The detailed method is given in Appendix 2. The COD removal was an important indicator of the activity of the pelletised sludge. A drop in the COD removal during constant COD loading conditions would indicate that the stabilised system had been adversely affected by the build-up of solids in the reactor.

The COD was monitored daily during reactor start-up, but once the system had stabilised at the maximum COD loading, it was only determined two to three times per week.

4.2.2.2 Solids Concentration

The solids concentration at the top of the reactor, as well as in the filtrate from the CFMF membrane, was determined using the method outlined in Appendix 2. This analysis was performed daily, once filtration had started.

4.3 START-UP OF THE UPFLOW ANAEROBIC SLUDGE BED REACTOR

Figure 4.2 shows the percentage COD loading and removal during the start-up of the UASB reactor.

The reactor was partially filled with tap water before any sludge was added. 10 litres of sludge, that had been fed on the same synthetic wastewater as used in this experimental work, was then added to the UASB reactor. The sludge contained a high percentage of pelletised material.

The sludge was allowed to settle overnight and form a bed at the bottom of the reactor. During this period the controller was operative, but no feed medium was supplied.

The next day, feeding was started at full hydraulic load, but at half the COD loading, i.e. 2500 $mg.l^{-1}$. As filtration would only start once the system had stabilised, the influent was balanced by effluent drawn off from the UASB reactor liquid surface by the Watson Marlow peristaltic pump.

After two weeks' operation at the above conditions, the percentage COD removal was found to be only 47 per cent. A thermocouple was used to probe the sludge bed and check the temperature. Due to the plug flow nature of

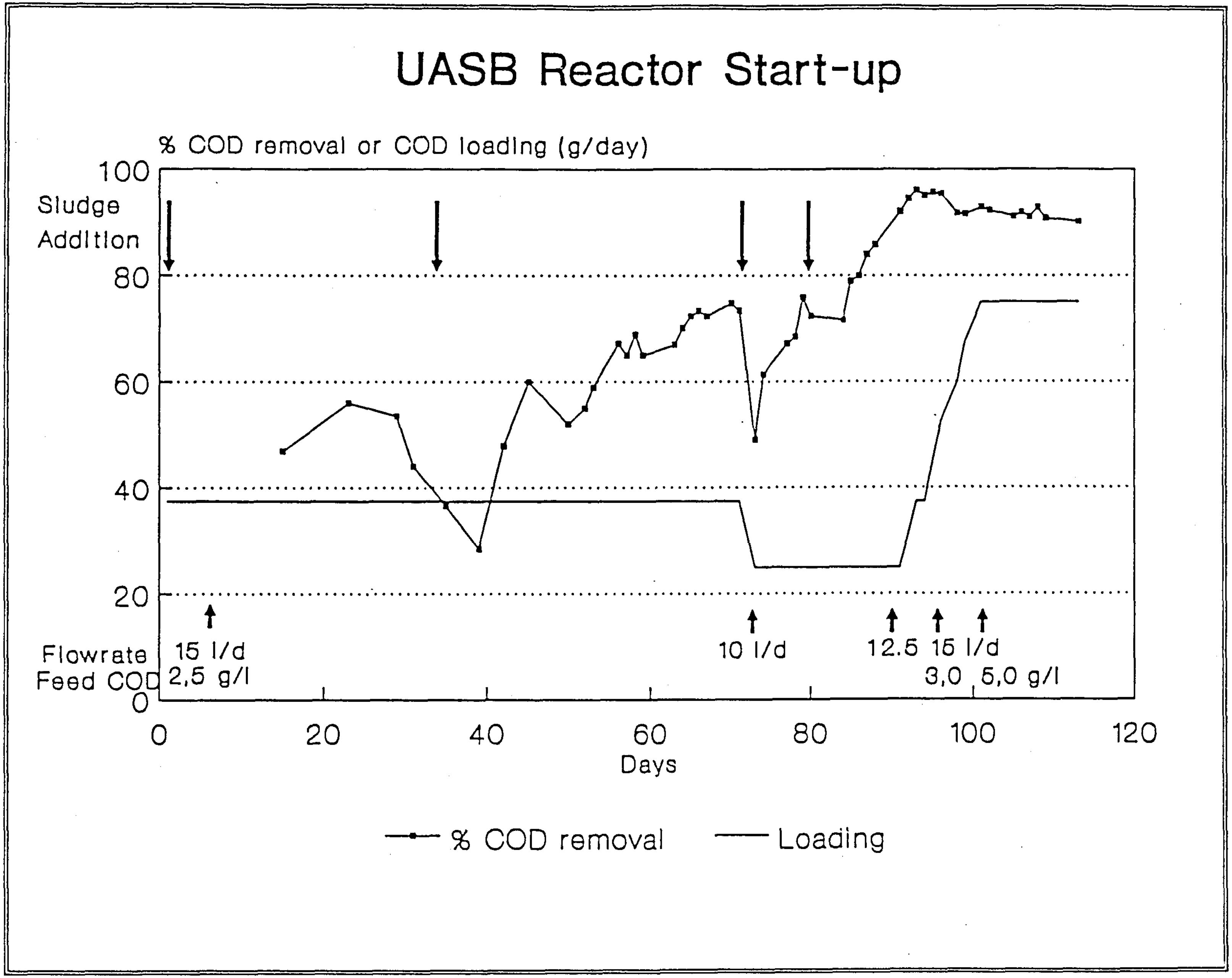


FIGURE 4.2 : UASB Reactor Start-up

the system, there were pockets of sludge that were several degrees below the 30°C set point. The ability of the sludge to degrade COD material is extremely dependent on the temperature of the environment in which it grows. During the following few days the efficiency of the temperature controller was improved and the pockets were essentially eliminated. Details of the improvement are discussed in Section 4.1.1.3.

As the COD removal did not increase substantially over the next two weeks, more sludge was added to the reactor. Only a few litres of sludge was added during this and subsequent additions. Approximately one litre was pelletised. The COD removal appeared to decrease initially when sludge was added to the reactor as additional, undegraded biological material was introduced along with the pellets.

By day 66, the COD removal had increased to approximately 73 per cent, but seemed to have stabilised at this point. The anaerobic sludge has a very low specific growth rate, necessitating the correct balancing of the sludge and the COD load.

1,3 g of yeast extract (Difco) was added to the daily 15 l feed to ensure that the growth was not nutrient limited. A 1:1 recycle was also implemented, recycling 15 l.day^{-1} of liquid from the top of the reactor through the sludge bed again to ensure that channeling was not the cause of the low COD removal.

During the next 5 days the COD showed negligible improvement. The recycle was stopped and more sludge was added on days 72 and 81 as it became available. The hydraulic load was also decreased to 10 l.day^{-1}.

From day 84, the COD removal increased steadily, reaching 94 per cent by day 92. At this point the hydraulic load was increased to 12,5 l.day^{-1} and to 15 l.day^{-1} the following day.

Between days 95 and 100 the COD load was incremented by 500 mg.l^{-1} daily, until the desired level of 5000 mg.l^{-1} was reached. As the load on the reactor was increased, a slight downward trend in the COD removal by the system can be seen in Fig. 4.2. This trend is probably due to the fact that the COD loading was increased from 31,05 to 75,0 g.COD.day^{-1} over approximately 20 days. Insufficient biomass had been generated within this relatively short time (growth of anaerobic organisms is slow) to maintain the former COD removal. The reactor had previously been slightly underloaded as the COD removal was a few percent above the normal 90 per cent removal obtained from UASB reactors.

The reactor was stabilsed on the 15 $l.day^{-1}$ of 5000 $mg.COD.l^{-1}$ feed for a further twelve days during which the COD removal was 91,7 per cent on average. UASB reactors operated by the Civil Engineering Department, University of Cape Town, have been found to have similar removals.

At this point, the reactor had reached stable operation at the desired hydraulic and COD loading and filtration was able to commence.

4.4 EFFECT OF PUMPING ON THE NATURE OF THE SLUDGE

Once filtration commenced it was noticed that the filterability of the efluent from the UASB reactor decreased considerably. The suspension could no longer be filtered through Whatman Number 1 filter paper during the COD analysis.

A sample of the slurry was taken for analysis on a Malvern 2600/3600 Particle Size Analyser (Malvern Instruments). Another sample was taken from a similar UASB reactor that was operating without filtration in the Department of Civil Engineering, University of Cape Town. Detailed results of the laser size analysis are included in Appendix 4.

The D(50%) of the sample taken from the UASB reactor operating without filtration was 36 μm, compared to the D(50%) of 16 μm obtained from the system operating with filtration. The results indicated that the particle size was being reduced substantially (approximately halved) during pumping and filtration.

The observed decrease in the filterability of the slurry was attributable to the smaller particle size. The smaller particles are likely to penetrate the precoat layer more easily as well as form a denser, far less permeable filtercake. Particle size was listed by Novak (1986) as one of the important factors affecting the filterability of activated sludge in his experiments.

The particles found in a UASB system are generally composed of fine biological material, generated by the breakdown of the pelletised sludge at the top of the sludge bed. These fines are probably still bound together by the polypeptide chains that bind the pellets - bonds readily broken by the shear during the pumping and filtration operations.

In addition to the decrease in the particle size, the COD removal increased to between 98 and 99 per cent. This high COD removal is not normally found in UASB systems and was attributed to the growth of aerobic organisms in the liquid in the frustum of the UASB reactor. The liquid in this zone tends to be oxygenated by the turbulence caused by the return of the concentrate from the CFMF unit. The presence of the aerobes must also have an affect on the nature of the slurry. In a larger system, the turbulence would be less significant and aerobes may not be found to be present to the same extent.

4.5 PERFORMANCE OF THE UASB SYSTEM

4.5.1 INCREASE IN SOLIDS CONCENTRATION

Figure 4.3 shows the solids build-up in the UASB reactor as well as the solids present in the filtrate from the CFMF unit. Solids data were only collected after the last sludge addition to the reactor had been made.

During the first 17 days after the last sludge addition, the solids concentration was increased from 1,0 gl^{-1} to 5,9 gl^{-1}. The solids retained were those generated by the UASB system as well as those resulting from the aerobic growth in the frustum section of the reactor.

Fluctuations in the concentration of solids in the reactor were due to slight imbalances between the amount of filtrate removed and wastewater fed. Losses due to splashing and evaporation (approximately 0,6 $l.day^{-1}$) were made-up using filtrate before samples were taken.

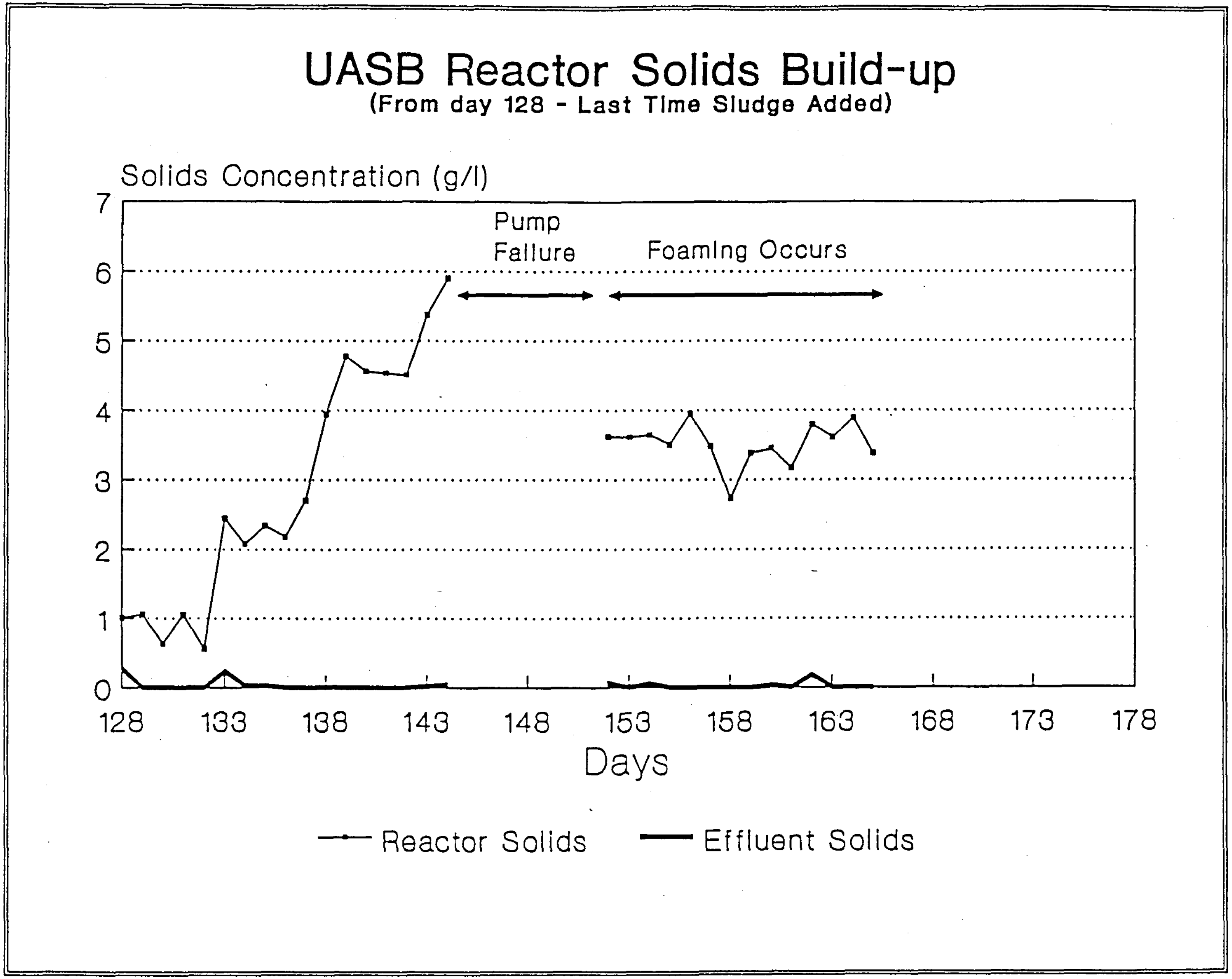


FIGURE 4.3 : UASB Reactor Solids Build-up
(From day 128 - last time sludge added).

Tabulated values of the daily reactor and effluent suspended solids concentrations are listed in Appendix 3, Table 3B. Suspended solids were not detected in the filtrate from the CFMF unit in nineteen of the thirt-one daily measurements. On one occasion the solids concentration in the filtrate was 0,276 $g.l^{-1}$. Generally, however, the average solids concentration was less than 0,050 $g.l^{-1}$. This range is similar to that obtained by Bindoff (1985) with a diatomaceous earth precoat (0,196 to 0,019 $g.l^{-1}$). The amount of solids detected depended on the condition of the precoat layer, especially as the filtrate quality improves substantially during the first hours after precoating.

In UASB systems receiving the same influent, but with gravity settling units, the solids concentration is usually in the region of 0,500 $g.l^{-1}$.

On day 146, the C32M Monopump motor overheated as it was apparently undersized and unable to operate continuously. During the following week the reactor feed was stopped whilst the pump motor was repaired.

The filtration process and reactor feed were restarted on day 151. A considerable amount of foaming occurred, a problem that had not been encountered up until this stage. In a full-scale process, the turbulence created by the return of concentrate from the CFMF unit would not be as great, reducing the significance of foaming. The foam was not particularly stable as it dispersed when filtration was stopped whilst precoating.

In the experimental process, however, the foam entrained a large portion of the solids that were produced, removing these from the reactor. Figure 4.3 shows that during the period of foaming there was essentially no solids build-up. The percentage COD removal was approximately 98 per cent, indicating that the system was active and that solids were being generated. The filtrate solids concentration also remained generally undetectable, implying that solids should have been accumulating.

Unfortunately, on day 165 the motor overheated once more, excessive wear having been noticed on the carbon brushes as previously. The experiment was terminated at this stage.

If operation had been continued, silicone antifoam could have been used to disperse the foam, if it had persisted. The draw-back when using antifoams is that they tend to affect the membrane, often lowering the flux considerably by changing the hydrophilicity.

4.5.2 COD PERFORMANCE

Figure 4.4 shows the percentage COD removal from just prior to the start of filtration.

Before filtration began, the COD removal had settled to between 91 and 92 percent. Immediately after filtration started, the percent removal dropped sharply. The feed pump rate was found to have become unstable after the replacement of the peristaltic tubing.

Approximately one litre of pelletised sludge was added to the reactor to dampen the effect of an influent COD fluctuation in the future. The peristaltic feed tubing was adjusted to ensure that the flow to the reactor was constant.

At this stage, the percentage COD removal increased rapidly, reaching a percentage removal of between 98 and 99 per cent. The growth of aerobic organisms in the frustum section was thought to have been a possible cause of the increased COD removal capacity.

4.5.3 FILTRATE FLUX AND FREQUENCY OF PRECOATING

Using a precoat of Swimline diatomaceous earth, enabled a "steady-state" flux of approximately 20 $l.m^{-2}h^{-1}$ to be attained. The flux declined gradually with time, falling below the desired 15 $l.m^{-2}h^{-1}$, necessitating membrane cleaning and re-precoating.

Continuous operation for up to four days, was found to be possible without membrane cleaning. During this period the filtrate quality remained high and there was insignificant build-up of biological growth on the outer surface of the membrane.

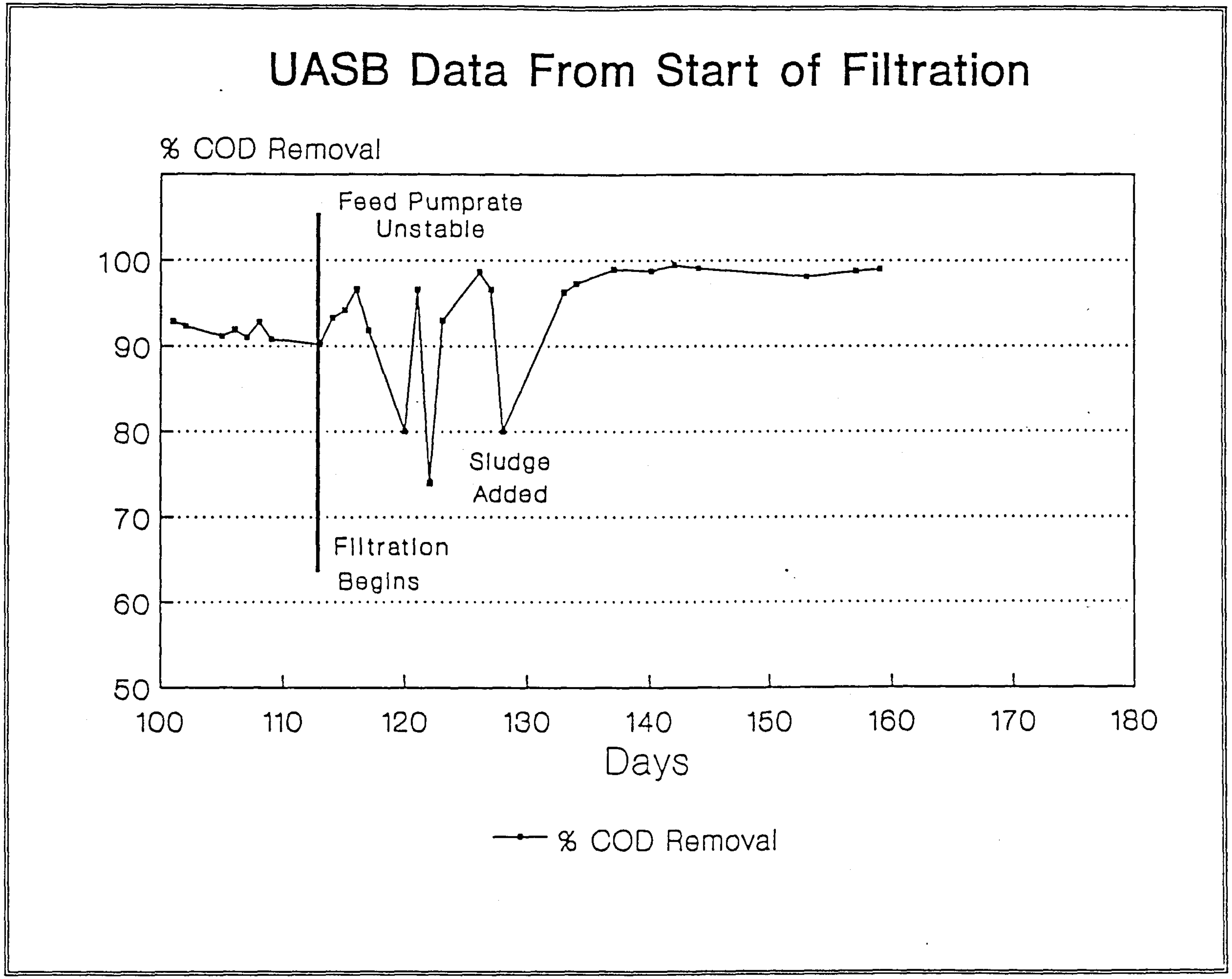


FIGURE 4.4 : UASB COD Data from Start of Filtration

4.6 CONCLUSIONS FROM THE UASB STUDY

1. When the CFMF unit was used to withdraw effluent from the UASB system the suspended solids concentration above the sludge bed was increased from less than 1,0 g.l^{-1} to 5,9 g.l^{-1} during the 17 days of trouble-free operation with the CFMF unit. The rate of increase was near linear. The maximum concentration at which the reactor could be operated was not reached as the experimental work had to be terminated due to pump failure.

2. The effluent from the system (CFMF permeate) was of very high quality with respect to solids content. On 19 of the 31 test days no solids were detectable in the permeate. On the remaining 12 test days the suspended solids concentration generally was less than 0,050 g.l^{-1}. The UASB-CFMF coupled system effluent contained a far lower suspended solids concentration than UASB systems receiving the same influent, but with gravity settling units, where the solids concentration is usually in the region of 0,500 g.l^{-1}.

3. UASB operation in terms of soluble COD removal was not adversely affected by CFMF operation. In fact, the COD removal increased from the normal level of approximately 90 percent to between 98 and 99 percent.

4. The very high COD removal can apparently be ascribed to growth of aerobic organisms in the upper section of the reactor. Oxygen supplied via the turbulence at the top of the reactor (caused by the CFMF reject stream) and substrate undegraded in the UASB sludge bed provide the requirements for growth of these organisms. This growth serves to polish the UASB effluent, giving a significant increase in COD removal. The introduction of oxygen does not upset anaerobic UASB operation.

5. The highest CFMF flux values obtained during the study were only approximately 25 to 50 per cent of the values observed in research by the Natal University group, using the same type of membrane and waste activated sludge.

6. Poor fluxes were obtained when using Kulu 15A limestone as precoat. This material has been used successfully by the Natal University group.

7. Pumping during filtration decreased the average particle size normally found in the UASB effluent (from 36 to 16 μm), substantially lowering its filterability. The effect may, however, be less significant in a larger UASB reactor system where the larger volume of slurry available will require a smaller turnover of sludge via the pump in order to extract the required effluent volume.

8. The concept of the aerobic polishing in the top zone of the UASB reactor should be patented.

CHAPTER FIVE

ACTIVATED SLUDGE EXPERIMENTAL INVESTIGATION

5.1 EQUIPMENT

A Schematic diagram of the activated sludge reactor and the crossflow microfiltration (CFMF) unit is given in Fig. 5.1.

The CFMF used in this investigation is the same as that used in the UASB reactor system. A full description of the unit is given in Chapter 3, Section 3.1.1.

5.1.1 ACTIVATED SLUDGE REACTOR

The activated sludge (AS) reactor was manufactured from PVC. The cylindrical reactor had a diameter of 0,22 m and height of 0,5 m, giving the reactor a capacity of 15 litres. The reactor was, however, only operated at a volume of 8 litres, which in addition to the 2 litre volume of the piping connecting the reactor with the CFMF unit gave a process volume of 10 litres.

The hold-up volume remaining in the reactor gave the additional process capacity required for small imbalances between the influent and effluent flowrates that occur due to filtration flux decline.

5.1.1.1 Agitation

The reactor was fitted with an agitator to thoroughly mix the reactor feed and the biomass, maintaining constant stirred tank reactor (CSTR) conditions.

There were also four PVC baffles symmetrically placed, running vertically down the reactor wall to facilitate good mixing.

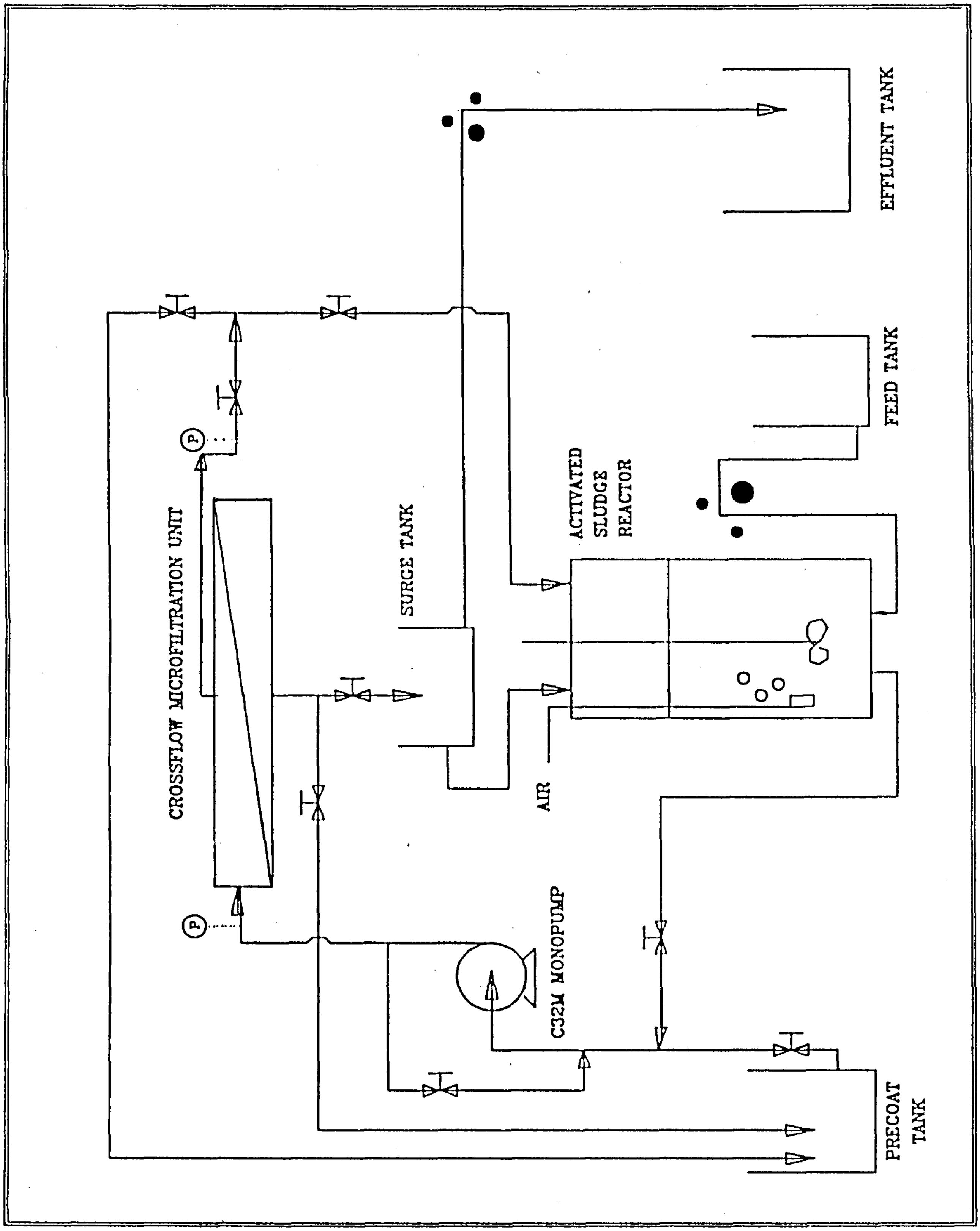


FIGURE 5.1 : Schematic Diagram of Activated Sludge Reactor and Crossflow Microfiltration Unit

5.1.1.2 Feed to the Activated Sludge Reactor

The synthetic wastewater feed was fed directly into the bottom of the reactor. No distributor was necessary as the reactor contents were thoroughly mixed. The feed was pumped from the feed tank into the reactor by the Watson Marlow 502S peristaltic pump, that was used in the UASB reactor investigation, at rates of up to 15 litres per day.

5.1.1.3 Effluent Removal from Activated Sludge Reactor

Once filtration had commenced, permeate from the CFMF unit was removed from the surge tank, using the Watson Marlow peristaltic pump, to balance the influent flow. Prior to the start of filtration, the reactor was operated on a fill and draw basis explained in section 5.3.

5.1.1.4 Slurry Removal and Return

The same C32M Monopump as used in the UASB system, was used to remove sludge from the reactor and pass it through the CFMF unit. The sludge was withdrawn from the bottom of the reactor and returned at the top. The turbulence caused by the sludge return from the CFMF unit aided both reactor aeration and mixing.

5.1.1.5 Temperature Control

The reactor temperature was maintained at 25°C by a controller built at the Chemical Engineering Department, University of Cape Town. A thermistor probe was floated on the surface of the liquid in the reactor to measure the temperature.

A heating tape (HCS 30W, 220 V, Wireohms) was fitted to the outside of the reactor with glass insulation tape. The controller was able to switch the heating tape on to maintain the reactor temperature when necessary.

A copper cooling coil was fitted into the reactor and water passed through it continuously. During the summer it was necessary to remove the excess heat entering the system due to the C32M pump.

5.1.1.6 Aeration

A standard sintered glass frit, as used in aquaria, was used to provide a fine source of bubbles to aerate the system. A Shiruba SP107 aquarium pump was used to pump air through a glass rod onto which the frit was attached.

5.2 EXPERIMENTAL METHOD

5.2.1 SYNTHETIC WASTEWATER FEED TO THE ACTIVATED SLUDGE REACTOR

A synthetic waste water feed has been developed and used by the Civil Engineering Department, University of Cape Town as feed for their research Activated Sludge Reactors.

The feed comprised : Vitamins; Micro nutrients; Macro nutrients; Complex carbohydrates; Soluble, and readily biodegradable soluble material. The composition of the feed is given in Appendix 1.

The fats and oils included in the synthetic wastewater used by the Civil Engineering Department were omitted and ammonium chloride was substituted for the organic nitrogen component as the latter lead to a great deal of foaming in the reactor due to the high degree of turbulence once filtration commenced. Glucose was used to make up the COD shortfall caused by the above changes.

5.2.1.1 COD Loading

The influent flow per day was kept constant at 15 $l.day^{-1}$ throughout the investigation, once the activated sludge reactor had passed through the start-up phase. Table 5.1 below, shows the stages of the experimental work and the respective COD loading. Some of the stages relied on an increase in the sludge age in the system to increase the amount of biomass rather than an increase in influent COD loading.

TABLE 5.1 : CONDITION AT EACH EXPERIMENTAL STAGE

Stage	Hydraulic Load [l.day^{-1}]	Sludge Age [days]	COD Conc. [mg.l^{-1}]	COD Loading [g.day^{-1}]
1	15	5	500	7,5
2	15	5	1000	15,0
3	15	5	2000	30,0
4	15	15	2000	30,0
5	15	20	2000	30,0
6	15	20	3000	45,0

5.2.2 ACTIVATED SLUDGE REACTOR OPERATION

5.2.2.1 Chemical Oxygen Demand

As in the UASB investigation, the stability of the reactor was determined by monitoring the percentage removal of the influent COD material. Permeate samples (i.e. reactor effluent) were taken several times a week and analysed according to the method detailed in Appendix 2.

5.2.2.2 Solids Concentration

A sample of sludge was removed from the reactor using a siphon tube with a wide internal diameter so as not to exclude large or filamentous particles. A sample of the permeate was also taken.

These samples were analysed for their total suspended solids content as discussed in Appendix 2, section 4. It should be noted, however, that the reactor sludge sample was centrifuged for 15 minutes at 6000 rpm and the supernatant poured off. The supernatant was not filtered (contrary to the methodology in Appendix 2) as the small amount of solids lost was assumed negligible in comparison to those remaining in the centrifuge tube.

Total suspended solids present in both the reactor sludge and the effluent was determined on a daily basis throughout the experimental work; permeate/effluent samples only becoming available once had filtration commenced.

5.2.2.3 Dissolved Oxygen

The dissolved oxygen concentration was measured periodically using a dissolved oxygen meter to ensure that the reactor was adequately aerated. It was vital that aerobic conditions were maintained in the reactor when operating at the high solids concentrations reached in this experimental work.

Although the aeration of activated sludge reactor forms a vital part of the efficient reactor operation, no attempt was made in this experimental study to investigate the type of oxygen transfer equipment (and its associated cost) that would be required on a full-scale plant. It was assumed that the reactor aeration achieved during this experimental work was possible on a full-scale plant.

5.2.2.4 Maintenance of Sludge Age

To maintain the sludge at the desired age a volume, Qw, was wasted on a daily basis by siphoning sludge directly out of the reactor into a measuring cylinder.

Qw is defined as follows :

$$Qw = V/Rs$$

where : V is defined as the reactor volume
Rs is the desired sludge age in days

The volume of the piping connecting the CFMF unit with the reactor was included in the total reactor volume used in this calculation.

5.3 START-UP OF THE ACTIVATED SLUDGE REACTOR

Activated sludge that was used for reactor start-up was obtained from reactors that had a five-day sludge age and were treating a synthetic wastewater similar to the one used in this experimental work. (The reactors were being operated in the Department of Civil Engineering, University of Cape Town).

The reactor was filled with five litres of tap water to which approximately 1,5 liters of concentrated sludge was added. The synthetic wastewater feed rate was set at 5 litres per day.

During start-up (the first nine days of operation) the reactor was operated on a fill and draw basis i.e. the agitator was stopped and the sludge allowed to settle so that a volume of supernatant, equivalent to the total daily influent, could be drawn off.

Additional sludge was added to the reactor periodically during the start-up so that after nine days the reactor suspended solids concentration had reached 0,95 $g.l^{-1}$. The COD removal was 95,9 per cent. At this stage filtration using the CFMF unit for solids/liquid separation commenced.

After one day of operation using the filtration circuit, an excessive amount foam was found to have been generated. To reduce the foaming, the organic nitrogen component of the feed (found to produce the most foam on agitation) was replaced with an equivalent amount (with respect to nitrogen) of ammonium chloride. This substantially reduced the foaming to an acceptable level. The organic nitrogen composition is given in Appendix 1 along with the other components of the synthetic wastewater.

5.4 THE EFFECT OF PUMPING THROUGH THE CFMF CIRCUIT ON THE NATURE OF THE SLUDGE

The sludge in the reactor was circulated approximately 1900 per day. The energy input to the system was found to change the nature of the sludge significantly.

5.4.1 SLUDGE PARTICLE SIZE

A sample was taken from an activated sludge reactor operating at the Civil Engineering Department as well as from the reactor used in this investigation. A size analysis was performed using a Malvern Particle Size Analyser. Detailed results are included in Appendix 4.

It was found that the D(50%) particle diameter for the Civil Engineering sample was 103 μm, whilst the reactor used in this study was found to have a D(50%) particle diameter of only 2 μm. The C32M Monopump has a comparatively low shear and would not normally be attributed as the cause for this observed size reduction. There is a reduction of pressure over the back pressure valve on the outlet from the CFMF membrane of two bars. As the pressure reduction occurs over a very short distance, this is most probably the cause of cell disruption.

In a full scale plant the pressure would have to be released more gradually, probably by means of a long section of vertical piping.

It has been suggested (Hart, 1989) that the decrease D(50%) particle size could have lead to improved oxygen transfer rates. Currently, the oxygen gradient thought to exist in larger-sized flocs needs to be investigated.

5.4.2 SLUDGE SETTLEABILITY

Due to the decrease in the average particle size, mentioned in section 5.4.1, the sludge was found to be unsettleable due the formation of a stable colloid. This is significant as it could pose a problem in the downstream processing (dewatering) of the sludge wasted from such a system.

The CFMF process requires a certain minimum crossflow velocity and operating pressure to effect separation no matter what length of membrane is used. There is very little pressure drop per unit length of membrane (approximately 2 to 4 $kPa.m^{-1}$), allowing for a considerable increase in the membrane length (and increased total permeate production) with virtually the same pumping requirements as the lab-scale system used in this investigation. This means that in a full-scale plant, the circulation rate per day through the CFMF unit would be much lower. The latter as well as a more gradual pressure release after the CFMF membrane would probably improve the sludge settleability and dewatering characteristics by not breaking the flocs to as great an extent and also allow greater time for reflocculation to occur.

5.4.3 MICROORGANISM POPULATION

Sludge samples from the conventional reactor at Civil Engineering and the reactor coupled to the CFMF unit were compared under the microscope.

The sludges were found to be very different in nature. Predators such as amoebae and psuedomonas, generally with a much larger size than the other bacteria, were only found to be present in the conventional sludge. Considering the relative particle sizes of the two sludges one would expect these larger microorganisms not to be found in the CFMF-activated sludge system.

The COD removal, discussed in section 5.5.2, was not adversely affected by the absence of these microorganisms as the degrading of the biological material is primarily performed by the smaller bacteria that were still present. The only parameters likely to be affected are the death rate of the microorganisms (due to the lack of predators) and the actual mass of sludge maintained by a particular influent COD (The mass is likely to be lower than expected as the larger organisms are not present).

Filamentous organisms were also not found in the CFMF-Activated sludge system. This has a direct influence on the settleability of the sludge. A certain proportion of filamentous organisms present in the microbial population tends to aid the settling process (Ekama, 1988). In 'pin-point' floc the lack of sufficient filamentous organisms results in the formation of small flocs with no real strength or structure. The flocs are easily sheared and broken up by relatively low turbulence. The formation of stable flocs greatly improves sludge settleability.

5.4.4 SLUDGE VISCOSITY

The sludge viscosity was determined at various total suspended solids concentrations for both the conventional activated sludge and that from the CFMF-activated sludge system.

The viscosity was measured using an Ostwald Tube Viscometer. The experimental method is detailed in Appendix 2, Section 5. An Ostwald viscometer, as opposed to a circulating bob-type viscometer, was chosen so that the shear on the sludge samples was kept to a minimum. The results of the viscosity analysis are shown in table 5.2 below.

TABLE 5.2 : RESULTS OF VISCOSITY MEASUREMENTS

Sample Origin *	TSS Conc. $[g.l^{-1}]$	Viscosity $[Pa.s] \times 10^4$
AS-CFMF	1,62	9,70
AS-CFMF	1,93	9,75
AS-CFMF	4,83	11,03
AS-CFMF	10,72	15,77
AS	2,22	10,41
AS	3,66	13,47
Water	0,00	8,95

* AS-CFMF : Activated Sludge from system with CFMF
AS : Activated Sludge from conventional reactor

The data shown in the table above indicates that the viscosity of the sludge from the AS-CFMF system is slightly lower than that from a conventional reactor at a similar suspended solids concentration.

On increasing the solids concentration from 1,93 to 4,83 $g.l^{-1}$ (an increase of 150 per cent) the viscosity increased by 13 per cent. There was, however, a approximately 43 per cent increase in the viscosity when the solids concentration increased from 4,83 to 10,72 $g.l^{-1}$. This shows that the viscosity tends to increase exponentially with increasing solids concentration.

The change in viscosity brought on by increases in the solids concentration has a direct effect on the Reynold's Number which distinguishes between turbulent and laminar flow. CFMF is dependent on operation in the turbulent flow regime to achieve effective shear at the membrane surface. As viscosity increases, the Reynold's number decreases, signifying decreased turbulence. This is observed as a decrease in the filtrate flux from the CFMF unit with increasing solids concentration.

At the highest solids concentration at which viscosity was measured, 10,72 $g.l^{-1}$, the viscosity was found to be nearly double that of water. The reactor was operated at suspended solids concentrations above 10,72 $g.l^{-1}$ and the consequent lowering of the permeate flux from the CFMF unit due to the increasing sludge viscosity is discussed further in section 5.5.3.

Obviously there will be a critical solids concentration at which the flow changes from turbulent to laminar and in the design an operation of a coupled system it is important that this is accounted for. Ross *et al* (1989) using an anaerobic sludge found that the changeover between laminar and turbulent flow occurred at a digester solids concentration of approximately 40 $g.l^{-1}$. This crossover value will depend on the type of sludge as well as the amount of energy input in the filtration circuit.

Operating temperature also affects the viscosity of the sludge. A higher operating temperature increases the flux from the filtration system. Ross *et al* (1989) found that the viscosity closely followed the Andrade equation (Findlay, 1953)

$$\text{Viscosity} = Ae^{B/RT}$$

where A and B are constants for a given liquid.

The dependence of the flux on the sludge temperature is likely to be significant where there is a large variation between summer and winter operating temperatures. Ross *et al* (1989) found that the flux from their UF system increased 2 per cent per °C change in sludge temperature. A winter temperature commonly used in the design of activated sludge plants is 14°C whilst summer temperatures can reach 25°C. This means that the change in flux could be in the region of 20 per cent between summer and winter.

As Ross *et al* (1989) suggest the choice between a mesophilic (35°C) and a thermophilc (55°C) system is an important design consideration (if the choice is available) when using a membrane process for solids/liquid separation, due to the potentially higher fluxes in the thermophilic region.

Due to the decrease in the flux from the CFMF unit as solids concentration increases, additional membrane area will be required to produce the same amount of permeate, increasing the cost of the filtration unit. In the cost optimisation for the process a solids concentration will be found where the overall cost of the reactor and CFMF unit is a minimum. If the reactor were operated at a very low solids concentration it would be extremely large in size. The flux from the CFMF unit would be high, so less membrane would be required and the filtration cost would be low. At the other extreme, if the biological reactor were operated at a high solids concentration, the reactor would be small, yet the permeate flux would be very low increasing the cost and size of the CFMF unit. The optimum reactor solids concentration would lie between these two scenarios.

The optimum reactor solids concentration is not only limited by the cost of the solids/liquid separation device, but is also restricted by the cost and capabilities of aerators used for oxygen transfer in the biological reactor.

5.5 PERFORMANCE OF THE ACTIVATED SLUDGE REACTOR

5.5.1 INCREASE IN THE REACTOR SOLIDS CONCENTRATION

The changes in the solids concentration during the various stages of the investigation are shown in Fig. 5.2. The conditions at each particular stage (i.e hydraulic and COD loading and the reactor sludge age) are given in Table 5.1, section 5.2.1.1 .

Steady-state activated sludge theory (Marais and Ekama, 1976) was used to estimate the expected total suspended solids (TSS) concentration in the reactor at the various conditions. Kinetic constants used in the theoretical prediction are given in Table 5.3. The predictions are also shown in Fig. 5.2.

On changing from one stage to the next, no sludge was wasted from the reactor so that the solids concentration would build up more rapidly. Sludge wasting was recommenced after several days to obtain steady state conditions at the particular sludge age.

It can be seen (Fig. 5.2) that the actual experimental solids concentration (once steady-state conditions had been established) were on average only 71 per cent of the expected value. Table 5.4, below gives the numerical values of the predicted and experimental reactor suspended solids concentrations.

TABLE 5.3 : ACTIVATED SLUDGE KINETIC PARAMETERS

PARAMATER		(at 25°C)
Heterotrophic Organisms		
Y_h	mg VSS(mg COD)$^{-1}$	0,45
b_h	d^{-1}	0,28
Kr	l.(mg VSS)$^{-1}$d^{-1}	0,017
f_{cv}	mg VSS(mg COD)$^{-1}$	1,48
f	mg VSS(mg VSS)$^{-1}$	0,2
Nitrification Organisms		
Y_n	mg VSS(mg VSS)$^{-1}$	0,1
b_n	d^{-1}	0,05
K_n	mg N.l^{-1}	1,79
ϕ_m	d^{-1}	0,80

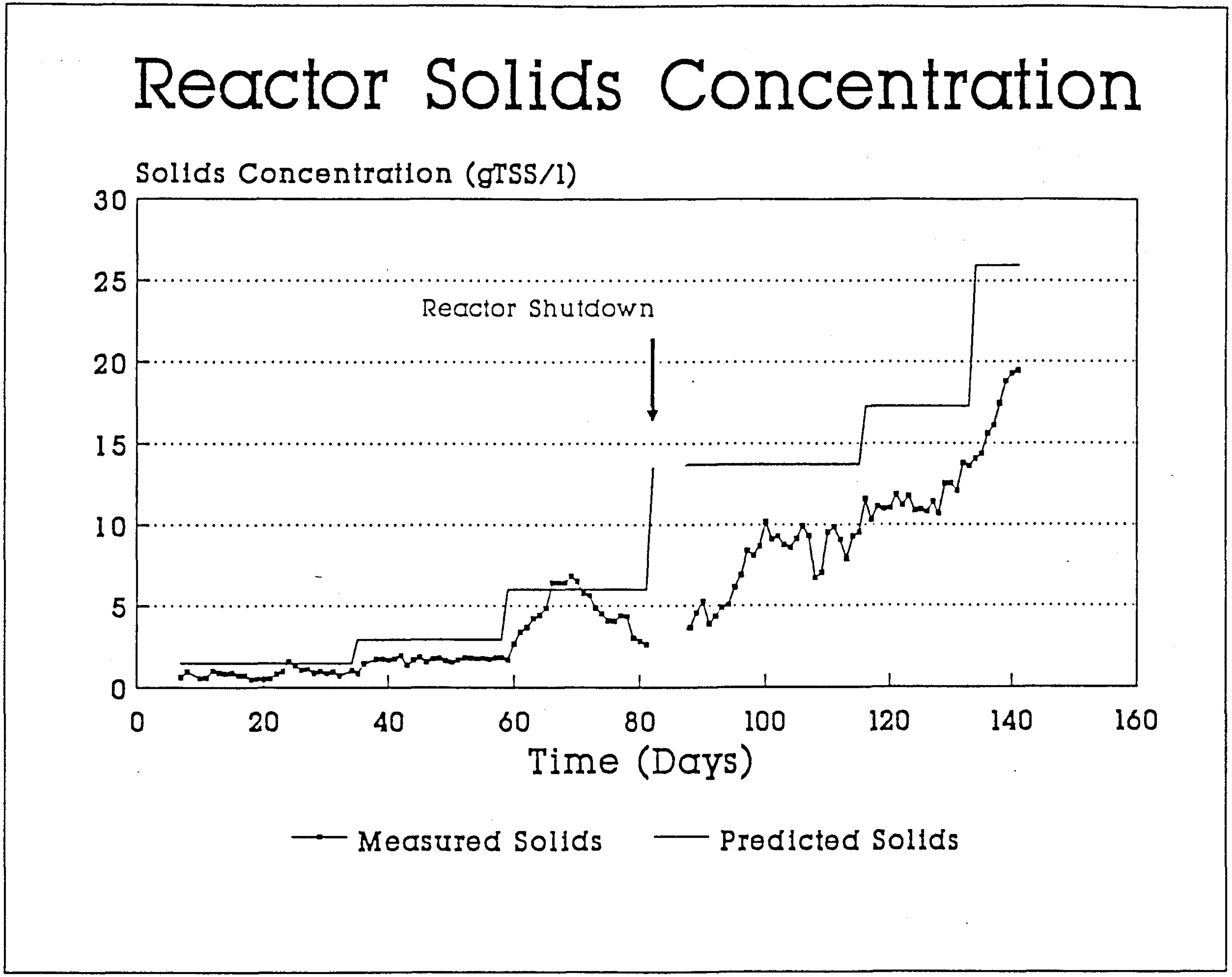


FIGURE 5.2 : Activated Sludge Reactor Solids Concentration

TABLE 5.4 : PREDICTED AND EXPERIMENTAL REACTOR SOLIDS CONCENTRATIONS

Stage	Predicted Solids conc. [$g.l^{-1}$]	Experimental Solids conc. [$g.l^{-1}$]	Experimental divided by Predicted Value
1	1,478	1,040	0,707
2	2,950	1,840	0,623
3	6,000	4,384	0,731
4	13,700	9,528	0,654
5	17,300	13,602	0,787
6	25,950	19,468 *	0,750
			(ave : 0,710)

* In this case steady state was not reached due to a temperature fluctuation.

Sludge wasting during the third stage was only restarted after the measured solids concentration in the reactor had exceeded the predicted value. Once the sludge wasting was restarted, the solids concentration again dropped to a concentration only 73 per cent of the predicted value. This showed that the biological system approached this off-set value from above as well as below, ruling out the possibility that insufficient time had been allowed for the solids concentration to reach the predicted value.

It is possible that the computer predicted value of the reactor solids concentration tended to be an over-estimate as the steady-state theory kinetic data was based on a conventional activated sludge system where predators were present in the microorganism population (i.e. their absence in this investigation being responsible for the loss in biomass. The fact that predators were not observed in the system is discussed in section 5.4.3).

Another reason for the difference between predicted and experimental solids concentrations in the reactor was the significant growth of microorganisms in the feed tank and the piping connecting it with the reactor. The microorganisms reduced the COD load actually entering the reactor. The biomass produced in the feed tank and piping was, however, not accounted for in the measurements of the reactor biomass supported by a particular COD loading. This was probably the most significant cause of the reduction in mass.

The microorganisms were introduced into the unsterilised reactor feed from the surroundings. They were able to establish themselves in substantial quantities within 24 hours despite the feed tank and the piping to the reactor being cleaning daily with boiling water.

One possible way in which the growth could have been reduced was to have refrigerated the feed tank to maintain it below 4°C, where microorganism growth would have been inhibited. The feed flowrate into the reactor was sufficiently small for the reactor operating temperature not to be affected. The reactor was also provided with a means of heating should it have been necessary.

On day 82 the feed pipe at the bottom of the reactor broke, causing the reactor to drain completely. Sludge was collected from the catch-tray and put back into the reactor along with additional sludge wasted from reactors at the Department of Civil Engineering as during start-up. This incident caused a drop in the reactor solids concentration that is indicated in Fig. 5.2.

The maximum concentration reached was 19,47 $g.l^{-1}$. This represents a nearly four-fold increase in the maximum concentration at which conventional activated sludge systems operate.

Unfortunately during the attempt to achieve steady state at this concentration the reactor temperature controller malfunctioned leading to an increase in the operating temperature, destroying the microorganism population. The excessive foaming associated with the organism death effectively emptied the reactor contents. At this stage the experimental work was terminated, although prior to the temperature run-away the biological system had still not been adversely affected by operation at the high concentration.

5.5.2 COD REMOVAL AND REACTOR PERFORMANCE

The COD loading and the percentage removal by the activated sludge system are shown in Fig. 5.3. The COD removal was determined to monitor the effects of operation at high suspended solids concentrations on the reactor performance.

Immediately after an increase in the COD loading on the reactor a drop in the percentage removal was obtained. This was to be expected until the biomass had increased sufficiently to degrade the additional COD material entering. This should be taken into account when looking at Fig. 5.3. The high percentage removals that would normally be obtained by the continuous operation of the reactor at any of the particular stages were only indicated at the end of each stage.

Once steady state had been reached at the various stages, the COD removal was between 97 and 99 per cent. The COD removal had generally exceeded 97 percent within 7 days of the increase in loading. This fast recovery rate was attributable to the more rapid increase in the biomass concentration as wasting had been suspended for a short period after the increase in loading.

The maintenance of this high removal at each stage indicated that the system was operating efficiently at the high suspended solids concentrations that were reached.

5.5.3 FILTRATE FLUX AND FREQUENCY OF PRECOATING

The CFMF unit was initially operated at a pressure of 2,0 to 2,5 bar and a crossflow velocity of 2,2 ms^{-1}. The crossflow velocity was raised to 3,6 ms^{-1} when the reactor suspended solids concentration exceeded 11 $g.l^{-1}$.

The days on which the membrane was reprecoated are indicated in the table of experimental data in Appendix 5. The steady state flux lay between 20 and 25 $l.m^{-2}h^{-1}$. This was higher than the flux that had been

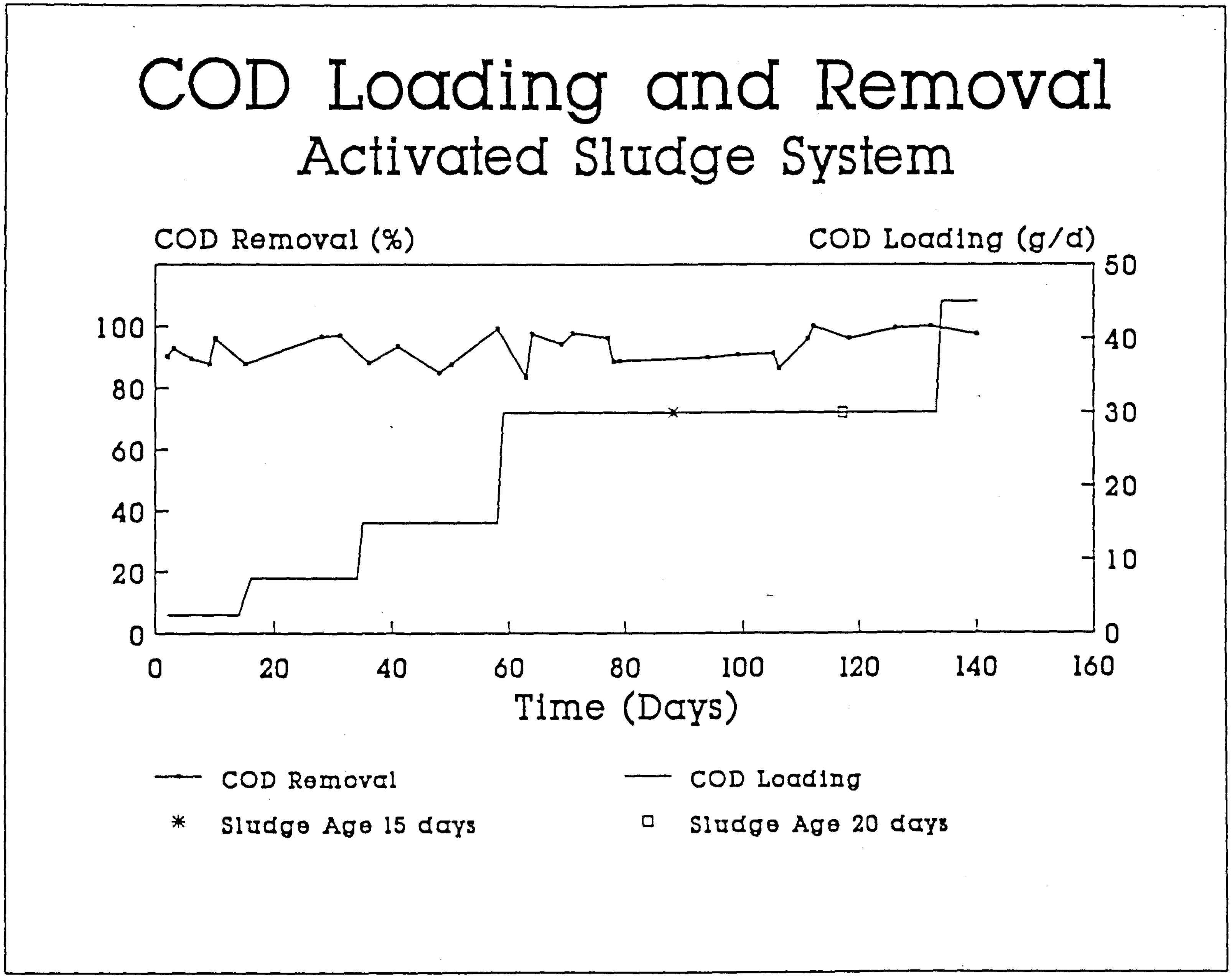


FIGURE 5.3 : COD Loading and Removal in the Activated Sludge System

obtained during filtration of the UASB effluent in the first part of this project.

At low concentrations of solids (around 1 $g.l^{-1}$) the flux remained above the 15 $l.m^{-2}h^{-1}$ required to balance the influent flow, for periods of up to 12 days before the membrane required cleaning and reprecoating.

When the reactor solids concentration reached 11 $g.l^{-1}$ it was found that even daily precoating was insufficient to maintain the flux at the level required to balance the influent flowrate. At this stage the crossflow velocity was increased to 3,6 ms^{-1} where the frequency of precoating was again reduced to between one and two times per week.

The effluent/permeate quality remained good throughout the experimental work, though the quality tended to deteriorate just prior to when reprecoating became necessary indicating the break-down or loss of the precoat layer. The suspended solids concentration detected in the effluent is shown in Fig. 5.4, along with the occassions when precoating and membrane cleaning were performed. On average the suspended solids concentration detected in the effluent was 0,013 $g.l^{-1}$ and on 46,8 per cent of the days during operation no suspended solids were detected in the effluent.

It was found that at the high COD loadings that the effluent suspended solids concentration was higher than at the lower COD loadings. This could have been brought about by the growth of microorganisms on the residual COD material in the permeate after filtration. Obviously there would be a higher residual COD in the effluent at high reactor COD loadings, although the percentage COD removal at low and high COD loadings was essentially the same.

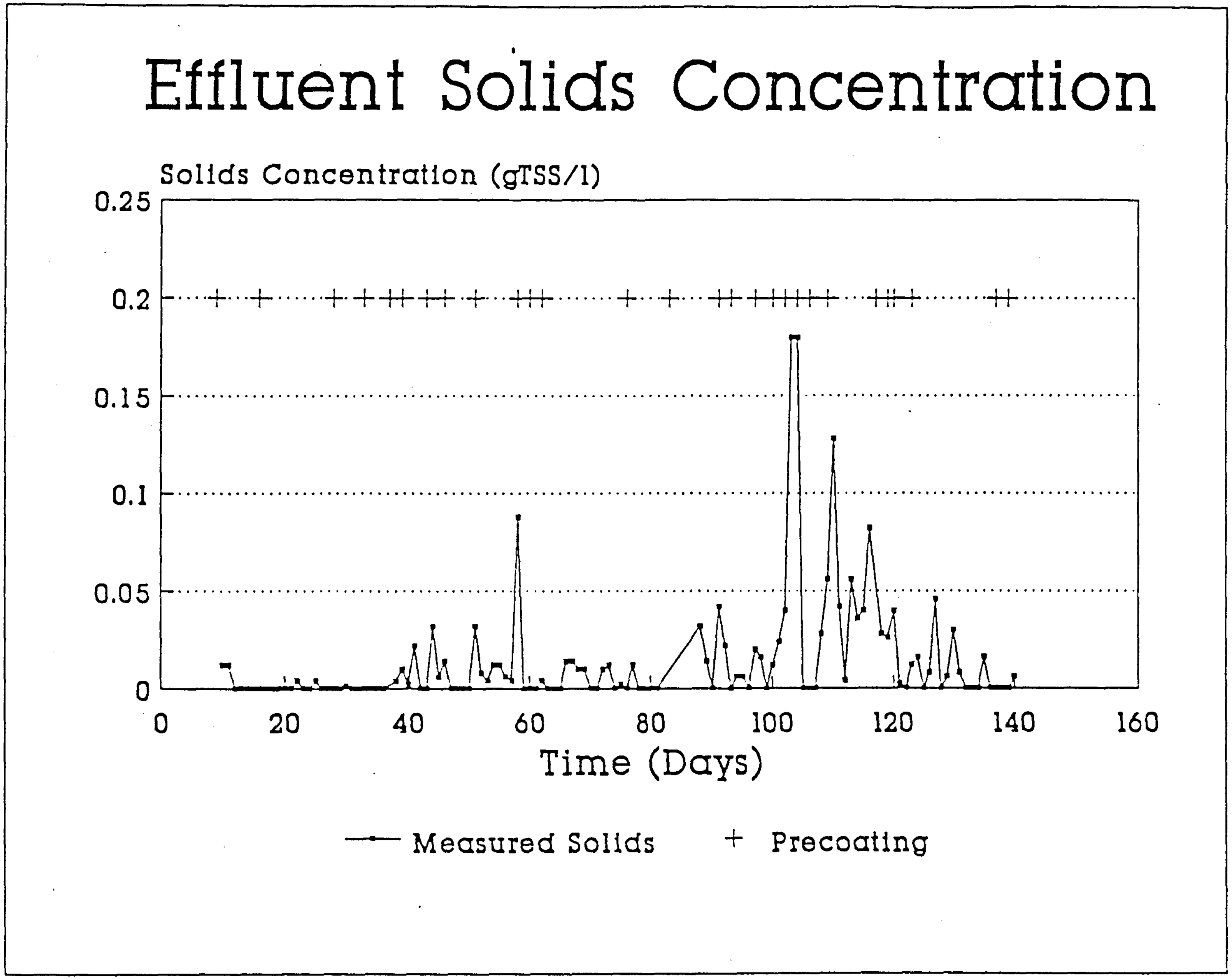


FIGURE 5.4 : Effluent Solids Concentration and Precoating in the CFMF-Activated Sludge System

5.6 CONCLUSIONS FROM THE ACTIVATED SLUDGE STUDY

1. The suspended solids concentration in the activated sludge reactor was successfully increased from 0,95 $g.l^{-1}$ to 19,47 $g.l^{-1}$ by the use of CFMF for solids/liquid separation. The maximum concentration at which the system could be operated had not been reached when the experimental work was terminated due to temperature controller malfunction.

2. The operation of the activated sludge reactor at suspended solids concentrations of up to nearly four times the normal level did not adversely affect the efficient operation of the reactor. At the end of each stage (i.e. at a range of intermediate values between 0,95 $g.l^{-1}$ to 19,47 $g.l^{-1}$) the COD removal was above 97 per cent. After a shock increase in COD loading, the reactor recovered to a COD removal of above 97 per cent within a few days. By suspending solids wasting after the step COD increase, the solids build-up was more rapid, enhancing the reactor recovery time. After three days sludge wasting was generally recommenced.

3. The permeate from the CFMF unit was of a high quality throughout the experimental investigation. On 46,8 per cent of the days during which the unit was operated, no suspended solids were detected in the effluent. The average suspended solids concentration detected was 0,013 $g.l^{-1}$. The higher concentration of suspended solids detected at higher COD loadings indicated that there was a degree of biomass growth on the residual COD material after filtration.

4. It is possible that the decrease in particle size from the 103μm found in conventional activated sludges to 2μm enhanced oxygen transfer in the reactor. The mechanical transfer of oxygen was not investigated in this study and it was assumed that the oxygen transfer achieved in this experimental work could be achieved on a full-scale plant.

5. The decrease in the particle size and the elimination of the filamentous organisms makes the sludge unsettleable. This poses a potential waste sludge dewatering problem depending on the method used for the dewatering process. By releasing the sludge pressure more gradually (such as in a long, vertical section of piping) after it is passed through the CFMF filter the particle size may not be decreased as significantly.

5.7 OTHER OBSERVATIONS FROM THE ACTIVATED SLUDGE STUDY

1. The growth of organisms in the feed tank and feed line as well as the use of kinetic constants derived from systems where predatory microorganisms were present, were atributed as causes of the measured reactor solids concentration being 71 per cent (on average) of the predicted value.

2. The CFMF flux of between 20 and 25 $l.m^{-2}h^{-1}$ obtained during the filtration of the activated sludge was higher than that obtained during the UASB experimental work. This emphasised the different filterability of different types of sludges. At solids concentrations above 11 $g.l^{-1}$ it was necessary to increase the crossflow velocity from 2,2 $m.s^{-1}$ to 3,6 $m.s^{-1}$ to achieve the same flux as obtained at concentrations below 11 $g.l^{-1}$. The membrane only required reprecoating approximately every four days.

3. Although the organic nitrogen component of the synthetic wastewater feed was found to be the main cause of excessive foaming this is unlikely to be encountered in a full-scale process due to the lower degree of turbulence.

4. The pumping of activated sludge through the CFMF filtration circuit affects the nature of the sludge, lowering the average particle size, settleability, and viscosity of the sludge. The larger microorganisms and filaments normally found in an activated sludge population were eliminated.

5. The removal of the larger microorganisms from the activated sludge population did not affect the COD removal efficiency of the reactor.

6. The sludge viscosity was found to increase exponentially with increasing suspended solids concentration. The viscosity of the sludge is an important factor in the design of an activated sludge reactor-CFMF coupled process as the flux obtained from the CFMF unit (and consequently the filtration area and cost of the solids/liquid separation) is directly affected by it.

CHAPTER SIX

CONCLUSIONS AND RECOMMENDATIONS

6.1 CONCLUSIONS

This section contains the major conclusions drawn from Chapters Three, Four and Five. Detailed conclusions are given at the end of each of these chapters.

1. When the CFMF unit was used to withdraw effluent from the UASB system the suspended solids concentration above the sludge bed was increased from less than 1 $g.l^{-1}$ to 5,9 $g.l^{-1}$ during the 17 days of trouble-free operation with the CFMF unit. The rate of increase was near linear. The maximum solids concentration at which the system could operate was not reached during the course of this experimental work due to pump failure leading to the termination of the study.

2. UASB operation in terms of soluble COD removal was not adversely affected by the high suspended solids concentration or CFMF operation. The COD removal was increased from the normal level of approximately 90 per cent to between 98 and 99 per cent. This probably resulted from the additional polishing effect of aerobic organisms growing on the residual COD material in the top zone of the reactor. The upper zone was probably aerated by the turbulent return of the CFMF reject stream. Unfortunately, no dissolved oxygen measurements were made in the top section of the reactor during operation.

3. The suspended solids concentration in the activated sludge reactor was successfully increased from 0,95 $g.l^{-1}$ to 19,47 $g.l^{-1}$ by the use of CFMF for solids/liquid separation. Again, the maximum suspended solids concentration at which the reactor could be operated was not reached during this experimental work.

4. The operation of the activated sludge reactor at suspended solids concentrations between 0,95 $g.l^{-1}$ to 19,47 $g.l^{-1}$ did not adversely affect the efficient operation of the reactor. In each case, the COD removal was above 97 per cent once steady state had been reached.

5. The nature of both biological sludges was changed by the recirculation through the CFMF unit. The average particle size in the UASB effluent was approximately halved (from 36 to 16 μm), whilst the activated sludge average particle size was reduced to approximately 50 times smaller (from 103 to 2 μm) than that found in conventional reactors. The activated sludge became unsettleable and the larger microorganisms and filaments normally found in such a sludge were eliminated.

   The change in the nature of the sludges did not adversely affect the operation of either bioreactor, though the CFMF unit flux was impaired by the poor filterability of the sludges. The change in nature should be less significant in full-scale systems where the larger volume of slurry will reduce the number of times that it has to be passed through the CFMF unit.

6. It is possible that the decrease in particle size from the 103μm found in conventional activated sludges to 2μm enhanced oxygen transfer in the reactor. The mechanical transfer of oxygen to the system was beyond the scope of this investigation, although aerobic conditions were maintained in the laboratory reactor throughout the study.

7. Sludge viscosity is an important factor in the design of a bioreactor-CFMF coupled process as the flux obtained from the CFMF unit (and consequently the filtration area and cost of the solids/liquid separation) is directly affected by it. Pumping costs in a full-scale plant are also directly related to the viscosity. The activated sludge viscosity was found to increase exponentially with increasing suspended solids concentration. Operating temperature (as influenced by the biological system or seasonal weather conditions) also affects the viscosity of the sludge and the flux that can be obtained from a CFMF unit.

8. Commercial swimming pool diatomaceous earth (Diatom Powder, Swimline (Pty.) Ltd.) when used as a membrane precoat gave the highest steady-state fluxes (between 15 and 25 $l.m^{-2}h^{-1}$) during the filtration of the two biological slurries in this investigation. This material was different to other precoats as it acted as an asymmetric membrane: a

porous, fibrous fraction formed a layer against the cloth membrane onto which a fine, less porous, particulate fraction built up.

9. The initial concentration of precoat material in the precoat tank appears to exert a major influence on the nature of the precoat layer formed and the resulting flux. A slurry concentration of approximately 1 g of precoat per litre was found to be adequate.

10. The CFMF steady-state flux values of between 20 and 25 $l.m^{-2}h^{-1}$ obtained during the study were only approximately 50 per cent of the values observed at similar operating conditions in research by the Natal University group. During filtration of the activated sludge, however, reprecoating was only required every four days on average compared to the daily reprecoating performed by the Natal University group.

    The difference between the operation of the CFMF unit in this investigation and the unit used by the Natal University group, whilst operating at the same crossflow velocity, pressure and temperature conditions, emphasised the dependence of the permeate flux on the nature of the sludge filtered.

11. The permeate from the CFMF unit was of a high quality throughout the experimental investigation.

    On nineteen of the thirty-one test days, during UASB effluent filtration, no solids were detectable in the permeate. On the remaining 12 test days the suspended solids concentration generally was less than 0,050 $g.l^{-1}$. This suspended solids concentration was considerably lower than the 0,500 $g.l^{-1}$ obtained from a comparable reactor when gravity settling units are used.

    On 46,8 per cent of the days during which the unit filtered activated sludge, no suspended solids were detected in the effluent. The average suspended solids concentration detected was 0,013 $g.l^{-1}$. The higher concentration of suspended solids detected at higher COD loadings indicated that there was a degree of biomass growth on the residual COD material after filtration.

6.2 RECOMMENDATIONS

1. After the successful operation of both the laboratory UASB and activated sludge reactors at suspended solids concentrations considerably above those found in conventional systems, research into the operation of pilot-scale reactors coupled to CFMF units should be performed. During these experimental investigations, the maximum solids concentration at which these reactors can be operated should be determined.

2. Constraints (such as economic factors) that were not encountered whilst using the laboratory-scale process should be investigated. The reactor solids concentration must be optimised by the economic evaluation of savings on the reactor size by operation at a high suspended solids concentration and the elimination of the gravity settling tank, versus the cost of the CFMF unit and associated higher pumping, waste sludge handling and other costs.

3. The effective aeration of the activated sludge must be considered. It is possible that depending on the method used for aeration, the cost or physical mass transfer of oxygen into the sludge may act as a further constraint on the reactor solids concentration.

4. CFMF flux and operating conditions are key factors in determining the viability of the bioreactor-CFMF coupled systems. A base of information should be established on CFMF performance with different types of biological sludges, types of membrane, operating conditions, precoat material and precoating technique.

5. Particular attention should be directed at the selection of precoat materials and the technique of precoating. A relationship between the average particle size of the precoat material required to efficiently filter a particular biological slurry of a certain particle size needs to be investigated. The precoat material used in this experimental work indicates that the combination of two materials, leading to the formation of an asymmetric precoat layer of fine pore size material supported by more porous material may increase fluxes substantially.

6. The implementation of membranes that do not require precoating should be investigated. Eliminating the precoating step would reduce operating costs. As indicated in Chapter Three, there are UF membranes that do not need precoating, yet operate at conditions similar to those of CFMF. Membranes falling into the CFMF category could be developed locally by companies such as Bintech and Explochem.

7. In the design of a process using CFMF for solids/liquid separation, the operating temperature is an important consideration. Due to the decrease in the viscosity of the sludge and associated increase in the CFMF flux, as high an operating temperature as possible should be selected.

8. The effect of the particle size on oxygen transfer in the activated sludge reactors as well as its effect on sludge settleability and viscosity needs to be investigated.

9. The concept of aerobic polishing in the top zone of the UASB reactor should be patented.

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APPENDIX ONE

COMPOSITION OF SYNTHETIC WASTEWATER FEED

1. SYNTHETIC WASTEWATER FEED FOR UASB REACTOR

The synthetic wastewater feed comprised glucose, trace metal solution, nutrients as well as a buffer (Gabb, Civil Engineering Department, University of Cape Town). Table 1A shows the quantities required for 15 l of fed solution, with a COD content of 5000 mg.l^{-1}. The buffer is used to maintain the pH at the top of the reactor at approximately 7,2. A typical analysis of the municipal water used in this experiemental owrk is given in Tables 1L and 1M.

Tables 1B and 1C give the compositions of the nutrients and trace metal solutions respectively.

TABLE 1A : COMPOSITION OF 15 l FEED SOLUTION

Glucose monohydrate	77,25 g
Trace metal solution	50 ml
Nutrients solution	750 ml
Buffer ($NaHCO_3$)	150 g
Tap water	to make 15 l

TABLE 1B : COMPOSITION OF 5 l NUTRIENT SOLUTION

COMPOUND	FORMULA	MASS (g)
Ammonium chloride	NH_4Cl	50,0
Dipotassium hydrogen orthophosphate	K_2HPO_4	20,0
Tap water		to make 5 l

TABLE 1C : COMPOSITION OF 1 l TRACE METAL SOLUTION

COMPOUND	FORMULA	MASS (g)
Boric acid	H_3BO_3	0,05
Iron(II) chloride	$FeCl_2.2H_2O$	2,00
Zinc chloride	ZnCl	0,05
Manganese sulphate	$MgSO_4$	0,50
Copper(II) chloride	$CuCl_2.2H_2O$	0,03
Ammonium molybdate	$(NH_4)_6Mo_7O_2.4H_2O$	0,05
Aluminium chloride	$AlCl_3.6H_2O$	0.05
Cobalt(II) chloride	$CoCl_2.6H_2O$	2,00
Hydrochloric acid	HCl	1 ml
Magnesium chloride	$MgCl_2.6H_2O$	2,13
Manganous chloride	$MnCl_2.4H_2O$	0,25
EDTA		0,05
Nickel chloride	$NiCl_2.6H_2O$	0,25
Potassium iodide	KI	0,05
Tap water		to make 1 l

2. SYNTHETIC WASTEWATER FED TO THE ACTIVATED SLUDGE REACTOR

Table 1D gives the composition of 15 l of the synthetic wastewater feed with a COD concentration of 1000 mg.l^{-1} (Gabb, Civil Engineering Department, University of Cape Town). A typical analysis of the Cape Town municipal water is given in Tables IL and 1M. During the course of the experimental investigation the influent COD was raised to 3000 mg.l^{-1}. The quantities of the various components of the synthetic wastewater shown in the table below were simply increased by the same factor as the desired increase in COD concentration (i.e. to obtain a feed with 3000 mgCOD.l^{-1} all of the quantities in the table were multiplied by three). The make-up water was still, however, only used to make the solution up to 15 l.

The compositions of the solutions mentioned in Table 1D are given in the tables that follow. All of the individual solutions were stored below 4°C to prevent as much biological growth as possible.

TABLE 1D : COMPOSITION OF 15 l FEED SOLUTION
(with a COD of 1000 mgl^{-1})

Vitamins	45 ml
Biodegradable solutes	30 ml
Micro inorganic nutrients	45 ml
Macro inorganic nutrients	255 ml
Complex carbohydrates	255 ml
Organic nitrogen/replacement	210 ml
Additional micronutrients	75 ml
Glucose monohydrate	10,7 g
Tap water	to make 15 l

TABLE 1E : COMPOSITION OF 1 l VITAMIN SOLUTION

COMPOUND	MASS (g)
Pantothenic acid	0,28
Nicotinic acid	0,28
D-Biotin	0,014
Folic acid	0,014
Pyridoxine HCL	0,28
4-Aminobenzoic acid	0,28
Inositol (meso) inact.	0,28
Thiaminium dichloride	0,28
Riboflavin	0,28
Choline chloride	0,28

TABLE 1F : COMPOSITION OF 1 l BIODEGRADABLE SOLUTES

COMPOUND	MASS (g)
Lactose	0,66
Sodium acetate	2,76
Sodium succinate	1,74
Lactic acid	4 ml
Sodium citrate	5,28
Ethanol	1,8 ml
Butanol	0,9 ml
Glucose monohydrate	1,32
Glycerol	1,08

TABLE 1G : COMPOSITION OF 1 l MICRO INORGANIC NUTRIENTS

COMPOUND	MASS (g)
$FeSO_4.7H_2O$	0,66
$ZnSO_4.7H_2O$	2,76
$MnSO_4.4H_2O$	1,74
$CuSO_4.5H_2O$	4 ml
$CoCl_2.6H_2O$	5,28
$Na_2MoO_4.2H_2O$	1,8 ml
H_3BO_3	0,9 ml
KI	1,32

TABLE 1H : COMPOSITION OF 5 l MACRO INORGANIC NUTRIENTS

COMPOUND	MASS (g)
NH_4Cl	54,0
K_2HPO_4	19,0
KH_2PO_4	1,0
$MgCl_2.6H_2O$	66,0
$CaCl_2.2H_2O$	16,5

TABLE 1I : COMPOSITION OF 5 l COMPLEX CARBOHYDRATES

COMPOUND	MASS (g)
Starch	16,85
Cellulose	13,20
Agar agar	2,65
Dextrin	23,05

TABLE 1J : COMPOSITION OF 2 l ORGANIC NITROGEN

COMPOUND	MASS (g)
Casein	2,70
Peptone	5,05
Yeast	5,05
Gelatin	3,95

Note that after excessive foaming was encountered on commencement of filtration, the organic solution above was replaced by 28 $g.l^{-1}$ of ammonium chloride. The loss in COD was corrected for by the addition of glucose. This correction has already been indicated in Table 1D.

TABLE 1K : COMPOSITION OF 2 l ADDITIONAL MICRO NUTRIENTS

COMPOUND	MASS (g)
$NiCl_2.6H_2O$	0,100
$AlCl_3.6H_2O$	0,040
NH_4VO_3	0,010
Na_2SeO_3	0,004
TiO_2	0,016
$Na_2WO_4.2H_2O$	0,006

TABLE 1L : TYPICAL ANALYSIS OF MUNICIPAL WATER USED
(Morrison, 1989)

SPECIES	CONCENTRATION $[mg.l^{-1}]$
Cl^-	22
SO_4^{2-}	16
K^+	0,4
Na^+	12
Ca^{2+}	16
Mg^{2+}	1,3
Al^{3+}	0,28
Fe^{2+}	0,044
Mn^{2+}	0,028
F^-	0,04
Silicon	1
N (as ammmonia)	< 0,01
PO_4^{2-}	0,025

TABLE 1M : FURTHER ANALYSIS OF MUNICIPAL WATER USED
(Morrison, 1989)

PARAMETER	(at 20°C)
Conductivity	15 $mS.m^{-1}$
pH	9,1
Turbidity	0,6 NTU
Colour	< 5
Total Dissolved Solids	92 $mg.l^{-1}$
Hardness	48 $mg.l^{-1}$
Alkalinity	29 $mg.l^{-1}$

APPENDIX TWO

ANALYTICAL TECHNIQUES

1. CHEMICAL OXYGEN DEMAND (COD) ANALYSIS

The COD analysis was performed according to the potassium dichromate procedure outline in "Standard Methods". Suspended solid material was removed prior to analysis by centrifuging the sample at 4000 rpm for 20 minutes in a Beckman 45 centrifuge. The supernatant was then passed through 0,45 μm Millipore Filter paper.

2. TOTAL SOLIDS

Total solids analysis was only performed during the UASB reactor investigation.

Samples taken from the suspension at the top of the UASB reactor as well as the filtrate, were analysed for total solids content.

50 ml of each sample was poured into a preweighed evaporating basin. The sample was dried at 105°C for 24 hours, then placed in a dessicator to cool to room temperature. Once cooled, the basin was reweighed. The difference between the weights, when multiplied by 20, gave the total solids content in $g.l^{-1}$.

Before a sample was removed from the top of the UASB reactor, solids that had settled on the sloping sides of the cone at the top of the reactor, were resuspended.

3. TOTAL VOLATILE SOLIDS

The samples that had been processed for total solids concentration were placed in a furnace at 600°C for 20 minutes to oxidise the carbon content. The samples were removed from the furnace and placed in a dessicator to cool to room temperature. Once cool, the basins were reweighed - the difference between this weight and that found after drying at 105°C for 24 hours, multiplied by 20, gave the mass of volatile solids present in the sample in $g.l^{-1}$.

4. TOTAL SUSPENDED SOLIDS

Total suspended solids analysis was conducted on samples of both the reactor contents and the filtrate. The UASB reactor was sampled as outlined in section 2 above. The biomass that had attached to the activated sludge reactor walls was brushed off prior to sampling.

50 mls of each sample was filtered through preweighed 0,45 μm Millipore filter paper. The filter paper was dried in an oven at 105°C for 24 hours, in preweighed evaporating basins. After the drying was complete, the basins were placed in a dessicator to cool to room temperature. The gain in mass, multiplied by 20, gave the total suspended solids in g/l.

Due to the high biological solids content of the sample that was removed from the UASB and Activated Sludge reactors, it was necessary to centrifuge the sample in a Beckman 45 Centrifuge, at 4000 rpm for 20 minutes, prior to filtration. Centrifugation of the sample removed a sufficient proportion of the solids to prevent the build-up of an impermeable cake on the filter paper.

Only the supernatant liquid was then filtered through the Millipore filter paper and the paper and filter cake placed in the evaporating basin. The solid residue at the bottom of the centrifuge tube was added directly to the evaporating basin, distilled water being used to wash all the solids out of the tube.

5. DILUTED SLUDGE VOLUME INDEX (DSVI)

The Diluted Sludge Volume Index (DSVI) is defined as the volume (in ml) occupied by 1g of sludge after 30 minutes of settling in a 1 litre measuring cylinder with the requirement that the settled volume after 30 minutes is less than 200 ml.

Approximately 250 ml of reactor sample is poured into a clear 1 litre measuring cylinder. The volume is made up to 1 litre by adding the liquid effluent from the reactor (in this experimental work, the filtrate from the CFMF unit). The cylinder is inverted several times to resuspend the mixed liquor. The sludge is allowed to settle for 30 minutes before the level at which the settled volume of sludge is recorded.

To calculate the DSVI, the total suspended solids concentration (discussed above) must be known. The settled volume is multiplied by 4 (to make the 250 ml sample used up to 1000 ml) and then this value is divided by the TSS concentration in $g.l^{-1}$ to give the DSVI in $ml.g^{-1}$.

6. ACTIVATED SLUDGE VISCOSITY

The viscosity of the activated sludge samples was measured using an Ostwald Viscometer. An Ostwald Viscometer essentially comprises a narrow bore glass pipe connecting an upper and lower reservoir on opposite arms of a U-tube.

As much of the viscometer as possible was immersed in a constant temperature bath. 30 ml of the sample was pipetted into the viscometer and the system allowed to equilibrate for 10 minutes with the surrounding water. The water bath temperature was set at 28°C (Close to the temperature at which the activated sludge reactor was operated allowing for a slight temperature increase during flow through the filtration circuit due to energy input).

The sample was then pumped to the upper reservoir using a pipette-filler. The vacuum was released and the time taken for the sample to flow through the narrow bore pipe between two markers was recorded. This was repeated until the measurements corresponded closely. Repeatability is generally good.

The suspended solids concentration of each sample was determined using the method described in section 4. This was used to calculate the density of the sample by assuming that the solids were suspended in pure water.

The viscometer was calibrated by measuring the time taken for distilled water (introduced and equilibrated as above) to flow between the two markers. The water viscosity and density at the particular temperature used can be found in literature such as Perry (The Chemical Engineer's Handbook). Once this calibration has been performed, the unknown viscosities may be calculated using the following equation.

$$\eta_2 = \frac{\rho_2 * t_2 *}{\rho_1 * t_1} \eta_1$$

Where η_1 : Water viscosity from tables (cp)
η_2 : Unknown viscosity (cp)
ρ_1 : Water density (g.l^{-1})
ρ_2 : Sample density (g.l^{-1})
t_1 : Time for water to flow between markers (s)
t_2 : Time for sample to flow between markers (s)

APPENDIX THREE

UASB OPERATING DATA

TABLE 3A : CHEMICAL OXYGEN DEMAND (COD)
LOADING AND REMOVAL DATA

DAY	INFLUENT COD [mg.l^{-1}]	FEED RATE [l.day^{-1}]	COD LOADING [g.day^{-1}]	COD REM [%]
1	0	0	0	
2	2500	15	37,5	
3	2500	15	37,5	
4	2500	15	37,5	
5	2500	15	37,5	
6	2500	15	37,5	
7	2500	15	37,5	
8	2500	15	37,5	
9	2500	15	37,5	
10	2500	15	37,5	
11	2500	15	37,5	
12	2500	15	37,5	
13	2500	15	37,5	
14	2500	15	37,5	
15	2500	15	37,5	46,9
16	2500	15	37,5	
17	2500	15	37,5	
18	2500	15	37,5	
19	2500	15	37,5	
20	2500	15	37,5	
21	2500	15	37,5	
22	2500	15	37,5	
23	2500	15	37,5	56,0
24	2500	15	37,5	
25	2500	15	37,5	
26	2500	15	37,5	
27	2500	15	37,5	
28	2500	15	37,5	
29	2500	15	37,5	53,5
30	2500	15	37,5	
31	2500	15	37,5	44,0
32	2500	15	37,5	
33	2500	15	37,5	
34	2500	15	37,5	
35	2500	15	37,5	36,6
36	2500	15	37,5	
37	2500	15	37,5	
38	2500	15	37,5	
39	2500	15	37,5	28,5
40	2500	15	37,5	
41	2500	15	37,5	
42	2500	15	37,5	47,9

TABLE 3A CONTINUED

DAY	INFLUENT COD $[\mathrm{mg.l^{-1}}]$	FEED RATE $[\mathrm{l.day^{-1}}]$	COD LOADING $[\mathrm{g.day^{-1}}]$	COD REM [%]
43	2500	15	37,5	
44	2500	15	37,5	
45	2500	15	37,5	60,0
46	2500	15	37,5	
47	2500	15	37,5	
48	2500	15	37,5	
49	2500	15	37,5	
50	2500	15	37,5	52,0
51	2500	15	37,5	
52	2500	15	37,5	55,0
53	2500	15	37,5	59,0
54	2500	15	37,5	
55	2500	15	37,5	
56	2500	15	37,5	67,2
57	2500	15	37,5	65,0
58	2500	15	37,5	69,0
59	2500	15	37,5	65,0
60	2500	15	37,5	
61	2500	15	37,5	
62	2500	15	37,5	
63	2500	15	37,5	67,0
64	2500	15	37,5	70,2
65	2500	15	37,5	72,4
66	2500	15	37,5	73,4
67	2500	15	37,5	72,4
68	2500	15	37,5	
69	2500	15	37,5	
70	2500	15	37,5	74,8
71	2500	15	37,5	73,4
72	2500	10	25,0	
73	2500	10	25,0	49,0
74	2500	10	25,0	61,3
75	2500	10	25,0	
76	2500	10	25,0	
77	2500	10	25,0	67,2
78	2500	10	25,0	68,5
79	2500	10	25,0	75,9
80	2500	10	25,0	72,4
81	2500	10	25,0	
82	2500	10	25,0	
83	2500	10	25,0	
84	2500	10	25,0	71,6
85	2500	10	25,0	79,0
86	2500	10	25,0	80,0
87	2500	10	25,0	84,0

TABLE 3A CONTINUED

DAY	INFLUENT COD [mg.l^{-1}]	FEED RATE [l.day^{-1}]	COD LOADING [g.day^{-1}]	COD REM [%]
88	2500	10	25,0	85,8
89	2500	10	25,0	
90	2500	10	25,0	
91	2500	10	25,0	92,0
92	2500	12,5	31,3	94,5
93	2500	15	37,5	96,1
94	2500	15	37,5	95,1
95	3000	15	45,0	95,7
96	3500	15	52,5	95,4
97	4000	15	60,0	
98	4000	15	60,0	91,7
99	4500	15	67,5	91,6
100	5000	15	75,0	
101	5000	15	75,0	92,9
102	5000	15	75,0	92,3
103	5000	15	75,0	
104	5000	15	75,0	
105	5000	15	75,0	91,1
106	5000	15	75,0	91,9
107	5000	15	75,0	91,0
108	5000	15	75,0	92,8
109	5000	15	75,0	90,8
110	5000	15	75,0	
111	5000	15	75,0	
112	5000	15	75,0	
113	5000	15	75,0	90,1
114	5000	15	75,0	93,3
115	5000	15	75,0	94,2
116	5000	15	75,0	96,6
117	5000	15	75,0	91,8
118	5000	15	75,0	
119	5000	15	75,0	
120	5000	15	75,0	80,0
121	0	15	0,0	96,6
122	2500	15	37,5	74,0
123	2500	15	37,5	93,0
124	5000	15	75,0	
125	5000	15	75,0	
126	5000	15	75,0	98,7
127	5000	15	75,0	96,6
128	5000	15	75,0	80,0
129	2500	15	37,5	<80,0
130	2500	15	37,5	<80,0
131	2500	15	37,5	
132	2500	15	37,5	

TABLE 3A CONTINUED

DAY	INFLUENT COD $[mg.l^{-1}]$	FEED RATE $[l.day^{-1}]$	COD LOADING $[g.day^{-1}]$	COD REM [%]
133	2500	15	37,5	96,3
134	5000	15	75,0	97,3
135	5000	15	75,0	
136	5000	15	75,0	
137	5000	15	75,0	99,0
138	5000	15	75,0	
139	5000	15	75,0	
140	5000	15	75,0	98,8
141	5000	15	75,0	
142	5000	15	75,0	99,5
143	5000	15	75,0	
144	5000	15	75,0	99,1
145	5000	15	75,0	
146	5000	15	75,0	
147	5000	15	75,0	
148	5000	15	75,0	
149	5000	15	75,0	
150	5000	15	75,0	
151	2500	15	37,5	
152	2500	15	37,5	
153	2500	15	37,5	98,1
154	4000	15	60,0	
155	4000	15	60,0	
156	5000	15	75,0	
157	5000	15	75,0	98,8
158	5000	15	75,0	
159	5000	15	75,0	99,0
160	5000	15	75,0	
161	5000	15	75,0	
162	5000	15	75,0	
163	5000	15	75,0	
164	5000	15	75,0	
165	5000	15	75,0	

TABLE 3B : SOLIDS CONCENTRATION DATA

DAY	REACTOR SOLIDS [g.l^{-1}]	FILTRATE SOLIDS [g.l^{-1}]
128	1,0080	0,2760
129	1,0600	ND
130	0,6320	ND
131	1,0520	ND
132	0,5600	ND
133	2,4440	0,2360
134	2,0740	0,0320
135	2,3340	0,0280
136	2,1780	ND
137	2,7060	ND
138	3,9460	0,0080
139	4,7760	0,0020
140	4,5660	ND
141	4,5400	ND
142	4,5120	ND
143	5,3780	0,0220
144	5,9040	0,0480
145	Pump failure	
146		
147		
148		
149		
150		
151		
152	3,6180	0,0660
153	3,6140	ND
154	3,6460	0,0620
155	3,5060	ND
156	3,9540	ND
157	3,4840	ND
158	2,7300	ND
159	3,3860	ND
160	3,4520	0,0320
161	3,1680	ND
162	3,7920	0,1820
163	3,6140	ND
164	3,8880	ND
165	3,3800	ND

* ND : no solids detected

TABLE 3C : EFFECT OF DIFFERENT PRECOAT CONCENTRATION ON FILTRATE FLUX

TIME (mins)	15 $g.m^{-2}$ PRECOAT FLUX ($l.m^{-2}.h^{-1}$)	1,9 $g.m^{-2}$ PRECOAT FLUX ($l.m^{-2}.h^{-1}$)
5	574,0	81,5
10	540,0	32,7
12	139,4 (sludge)	
15		24,8
17	47,0	
18		17,4 (sludge)
22	36,0	
23		8,4

(sludge) = changeover from precoating to sludge filtration

Crossflow Veocity : 2,0 $m.s^{-1}$
Pressure : 250 kPa

TABLE 3D : FILTRATION FLUXES FROM DIFFERENT PRECOATS

A : Kulu 15A Limestone
B : Large Diatomaceous Earth (Holpro)
C : Swimline Diatomaceous Earth
D : Swimline Diatomaceous Earth : Soaked
* : Changeover from precoating to sludge filtration

TIME (min)	A FLUX $(l.m^{-2}.h^{-1})$	B FLUX $(l.m^{-2}.h^{-1})$	C FLUX $(l.m^{-2}.h^{-1})$	D FLUX $(l.m^{-2}.h^{-1})$
0	2160,0			
5	911,4	2207,5	2400,0	3085,7
10	830,8	881,6	469,6	1018,4
15	* 626,1	* 981,8	* 407,5	* 388,1
20	59,0	490,9	163,5	308,6
5	41,6	stopped	110,2	196,4
30			88,2	139,6
55	28,4			
60			55,1	55,6
24 hrs	6,5		10,2	17,4

Crossflow Velocity = 2,0 $m.s^{-1}$
Pressure = 300 kPa

APPENDIX FOUR

MALVERN PARTICLE SIZE ANALYSIS

Sample from UASB Reactor without CFMF Unit

MALVERN 2600/3600 PARTICLE SIZE ANALYSIS

SIZE	WEIGHT	WEIGHT IN BAND		LIGHT ENERGY	
[μm]	% UNDER	[μm]	[%]	CALCULATED	MEASURED
188,0	100,0	188,0 - 87,2	0,3	903	911
87,2	99,7	87,2 - 53,5	2,3	1179	1186
53,5	97,4	53,5 - 37,6	6,1	1400	1413
37,6	91,3	37,6 - 28,1	10,9	1567	1577
28,1	80,5	28,1 - 21,5	14,9	1755	1770
21,5	65,6	21,5 - 16,7	15,0	1896	1910
16,7	50,6	16,7 - 13,0	12,2	2014	2037
13,0	38,4	13,0 - 10,1	10,4	2047	2047
10,1	27,9	10,1 - 7,9	9,3	1987	2017
7,9	18,7	7,9 - 6,2	6,9	1884	1898
6,2	11,8	6,2 - 4,8	4,3	1710	1715
4,8	7,5	4,8 - 3,8	2,7	1491	1500
3,8	4,8	3,8 - 3,0	1,6	1320	1358
3,0	3,2	3,0 - 2,4	0,9	1294	1256
2,4	2,3	2,4 - 1,9	0,8	1376	1265
1,9	1,5				
D(50%) [μm] = 36,07		V M D [μm] = 47,02		S M D [μm] = 25,47	
D(10%) [μm] = 12,01		D(90%) [μm] = 90,44		SPAN = 2,17	

Sample from UASB Reactor with CFMF Unit

SIZE	WEIGHT	WEIGHT IN BAND		LIGHT ENERGY	
[µm]	% UNDER	[µm]	[%]	CALCULATED	MEASURED
188,0	100,0	188,0 - 87,2	3,4	155	144
87,2	96,6	87,2 - 53,5	9,0	264	260
53,5	87,6	53,5 - 37,6	10,6	396	397
37,6	77,0	37,6 - 28,1	9,7	559	560
28,1	67,3	28,1 - 21,5	10,7	772	777
21,5	57,3	21,5 - 16,7	8,4	1011	1016
16,7	48,9	16,7 - 13,0	7,8	1267	1270
13,0	41,1	13,0 - 10,1	8,0	1470	1469
10,1	38,1	10,1 - 7,9	7,5	1623	1641
7,9	25,6	7,9 - 6,2	6,1	1745	1749
6,2	19,5	6,2 - 4,8	4,2	1793	1796
4,8	15,3	4,8 - 3,8	3,8	1795	1796
3,8	12,0	3,8 - 3,0	2,7	1805	1843
3,0	9,3	3,0 - 2,4	1,9	1895	1880
2,4	7,4	2,4 - 1,9	2,2	2047	2047
1,9	5,2				
D(50%) [µm] = 16,55		V M D [µm] = 19,74		S M D [µm] = 10,40	
D(10%) [µm] = 5,60		D(90%) [µm] = 36,45		SPAN = 1,86	

Sample of Kulu 15A Limestone

SIZE [µm]	WEIGHT % UNDER	WEIGHT IN BAND [µm]	WEIGHT IN BAND [%]	LIGHT ENERGY CALCULATED	LIGHT ENERGY MEASURED
188,0	100,0	188,0 - 87,2	3,4	99	99
87,2	96,6	87,2 - 53,5	9,0	146	146
53,5	87,6	53,5 - 37,6	10,6	195	197
37,6	77,0	37,6 - 28,1	9,7	248	247
28,1	67,3	28,1 - 21,5	10,0	320	323
21,5	57,3	21,5 - 16,7	8,4	409	409
16,7	48,9	16,7 - 13,0	7,8	524	525
13,0	41,1	13,0 - 10,1	8,0	650	651
10,1	38,1	10,1 - 7,9	7,5	789	794
7,9	25,6	7,9 - 6,2	6,1	945	944
6,2	19,5	6,2 - 4,8	4,2	1092	1097
4,8	15,3	4,8 - 3,8	3,9	1260	1258
3,8	12,0	3,8 - 3,0	2,7	1467	1488
3,0	9,3	3,0 - 2,4	1,9	1737	1728
2,4	7,4	2,4 - 1,9	2,2	2047	2047
1,9	5,2				
D(50%) [µm] = 17,33		V M D [µm] = 26,90		S M D [µm] = 7,50	
D(10%) [µm] = 3,21		D(90%) [µm] = 62,31		SPAN = 3,41	

Sample of Swimline Diatom Powder

SIZE	WEIGHT	WEIGHT IN BAND		LIGHT ENERGY	
[µm]	% UNDER	[µm]	[%]	CALCULATED	MEASURED
188,0	100,0	188,0 - 87,2	19,9	1300	1300
87,2	80,1	87,2 - 53,5	26,6	1474	1479
53,5	53,5	53,5 - 37,6	17,0	1546	1546
37,6	36,5	37,6 - 28,1	11,6	1544	1543
28,1	24,9	28,1 - 21,5	8,2	1600	1611
21,5	16,7	21,5 - 16,7	5,6	1651	1642
16,7	11,1	16,7 - 13,0	3,5	1676	1685
13,0	7,5	13,0 - 10,1	2,3	1659	1657
10,1	5,3	10,1 - 7,9	1,7	1624	1631
7,9	3,5	7,9 - 6,2	1,2	1607	1607
6,2	2,4	6,2 - 4,8	0,7	1597	1597
4,8	1,6	4,8 - 3,8	0,6	1616	1607
3,8	1,1	3,8 - 3,0	0,4	1679	1709
3,0	0,6	3,0 - 2,4	0,2	1832	1829
2,4	0,4	2,4 - 1,9	0,2	2047	2047
1,9	0,3				
D(50%) [µm] = 50,19		V M D [µm] = 61,85		S M D [µm] = 28,68	
D(10%) [µm] = 15,60		D(90%) [µm] = 137,23		SPAN = 2,42	

Sample from Activated Sludge Reactor Without CFMF

SIZE [µm]	WEIGHT % UNDER	WEIGHT IN BAND [µm]	[%]	LIGHT ENERGY CALCULATED	MEASURED
188,0	100,0	188,0 - 87,2	59,3	2047	2047
87,2	40,7	87,2 - 53,5	16,2	1448	1398
53,5	24,5	53,5 - 37,6	9,8	1129	1126
37,6	14,7	37,6 - 28,1	4,4	1024	1001
28,1	10,4	28,1 - 21,5	3,5	1005	982
21,5	6,9	21,5 - 16,7	2,6	976	973
16,7	4,3	16,7 - 13,0	1,4	995	977
13,0	2,8	13,0 - 10,1	1,0	974	940
10,1	1,8	10,1 - 7,9	0,8	898	900
7,9	1,0	7,9 - 6,2	0,5	859	845
6,2	0,5	6,2 - 4,8	0,3	807	785
4,8	0,2	4,8 - 3,8	0,1	736	725
3,8	0,0	3,8 - 3,0	0,0	702	602
3,0	0,0	3,0 - 2,4	0,0	738	673
2,4	0,0	2,4 - 1,9	0,0	832	670
1,9	0,0				
D(50%) [µm] =103,06		V M D [µm] = 100,75		S M D [µm] = 56,90	
D(10%) [µm] = 27,42		D(90%) [µm] = 171,01		SPAN = 1,39	

Sample from Activated Sludge Reactor with CFMF

SIZE [µm]	WEIGHT % UNDER	WEIGHT IN BAND [µm]	[%]	LIGHT ENERGY CALCULATED	MEASURED
118,4	100,0	118,4 - 54,9	0,0	5	0
54,9	100,0	54,9 - 33,7	0,0	10	1
33,7	100,0	33,7 - 23,7	0,0	16	9
23,7	100,0	23,7 - 17,7	0,0	26	20
17,7	100,0	17,7 - 13,6	0,0	48	39
13,6	100,0	13,6 - 10,5	0,4	71	70
10,5	99,6	10,5 - 8,2	3,3	118	119
8,2	96,3	8,2 - 6,4	5,1	190	189
6,4	91,2	6,4 - 5,0	8,2	292	293
5,0	83,0	5,0 - 3,9	8,1	435	435
3,9	74,9	3,9 - 3,0	3,4	619	620
3,0	71,4	3,0 - 2,4	4,5	861	852
2,4	66,9	2,4 - 1,9	9,7	1171	1181
1,9	57,2	1,9 - 1,5	7,4	1566	1561
1,5	49,9	1,5 - 1,2	12,5	2047	2047
1,2	37,4				
D(50%) [µm] =	1,53	V M D [µm] =	2,59	S M D [µm] =	1,25
D(10%) [µm] =	0,56	D(90%) [µm] =	6,16	SPAN =	3,67

APPENDIX FIVE

ACTIVATED SLUDGE SYSTEM OPERATING DATA

The following table contains the experimental data that was collected during the operation of the activated sludge system.

The following abbreviations and terms appear in the table :

+Buffer : Addition of $NaHCO_3$ buffer
DSVI : Diluted sludge volume index. (See appendix 2)
DO : Dissolved oxygen concentration
N : Nitrogen
ND : Not Detected (refers to solids in the effluent)
P : Indicates membrane cleaning and precoating
PRE : The column in which membrane cleaning is indicated
TKN : Total Kjeldahl Nitrogen
+sludge : Activated sludge addition to the system
Wasting : The activated sludge wasted from the system
XF Vel. : Crossflow Velocity in CFMF unit

The reactor start-up and the results that were obtained from the activated sludge system is discussed in chapter 5. The comments that appear in the following table are more fully explained there.

TABLE 5A : ACTIVATED SLUDGE SYSTEM OPERATING DATA

Day	Influent Flow [l/day]	COD Influent [mg/l]	COD Effluent [mg/l]	COD Loading [mg/day]	COD Removal [%]	SUSPENDED SOLIDS Reactor [g/l]	SUSPENDED SOLIDS Effluent [g/l]	Sludge Age [days]	P R E	Comments
1	5	500		2500				5		Start-up
2	5	500	49.0	2500	90.2			5		DSVI = 400
3	5	500	36.7	2500	92.7			5		Fill and Draw
4	5	500		2500				5		
5	5	500		2500				5		
6	5	500	53.0	2500	89.4			5		Sludge Added
7	5	500		2500		0.618		5		
8	5	500		2500		0.950		5		Sludge Added
9	5	500	61.2	2500	87.8			5	P	Filtration Starts
10	5	500	20.4	2500	95.9	0.548	0.012	5		Foaming
11	5	500		2500		0.594	0.012	5		
12	5	500		2500		1.004	ND	5		
13	5	500		2500		0.882	ND	5		Organic N replaced
14	5	500		2500		0.820	ND	5		
15	10	500	61.2	5000	87.8	0.868	ND	5		
16	15	500		7500		0.710	ND	5	P	
17	15	500		7500		0.702	ND	5		DO = 10 mg/l
18	15	500		7500		0.476	ND	5		
19	15	500		7500		0.516	ND	5		
20	15	500		7500		0.464	ND	5		wasted 2l
21	15	500		7500		0.550	ND	5		no wasting, glucose
22	15	500		7500		0.834	0.004	5		wasted 2l
23	15	500		7500		0.986	ND	5		
24	15	500		7500		1.562	ND	5		
25	15	500		7500		1.336	0.004	5		

TABLE 5A CONTINUED

Day	Influent Flow [l/day]	COD Influent [mg/l]	COD Effluent [mg/l]	COD Loading [mg/day]	COD Removal [%]	SUSPENDED SOLIDS Reactor [g/l]	Effluent [g/l]	Sludge Age [days]	P R E	Comments
26	15	500		7500		1.066	ND	5		
27	15	500		7500		1.120	ND	5		
28	15	500	16.3	7500	96.7	0.880	ND	5	P	DSVI = 50,5 ml/g
29	15	500		7500		0.990	ND	5		
30	15	500		7500		0.857	0.001	5		
31	15	500	15.0	7500	97.0	0.956	ND	5		
32	15	500		7500		0.704	ND	5		
33	15	500		7500				5	P	No wasting
34	15	500		7500		1.040	ND	5		Wasted 21
35	15	1000		15000		0.846	ND	5		No wasting
36	15	1000	118.3	15000	88.2	1.440	ND	5		
37	15	1000		15000				5	P	DO = 10,4 mg/l
38	15	1000		15000		1.747	0.004	5		
39	15	1000		15000		1.744	0.010	5	P	
40	15	1000		15000		1.661	0.002	5		
41	15	1000	65.3	15000	93.5	1.759	0.022	5		
42	15	1000		15000		1.942	ND	5		Wasting 21
43	15	1000		15000		1.375	ND	5	P	DSVI = no settling
44	15	1000		15000		1.672	0.032	5		
45	15	1000		15000		1.849	0.006	5		
46	15	1000		15000		1.558	0.014	5	P	
47	15	1000		15000		1.748	ND	5		
48	15	1000	151.0	15000	84.9	1.789	ND	5		
49	15	1000		15000		1.622	ND	5		
50	15	1000	122.4	15000	87.8	1.520	ND	5		DSVI = no settling

TABLE 5A CONTINUED

Day	Influent Flow [l/day]	COD Influent [mg/l]	COD Effluent [mg/l]	COD Loading [mg/day]	COD Removal [%]	SUSPENDED SOLIDS Reactor [g/l]	Effluent [g/l]	Sludge Age [days]	P R E	Comments
51	15	1000		15000		1.678	0.032	5	P	
52	15	1000		15000		1.829	0.008	5		
53	15	1000		15000		1.784	0.004	5		
54	15	1000		15000		1.755	0.012	5		
55	15	1000		15000		1.777	0.012	5		
56	15	1000	0.0	15000		1.707	0.006	5		
57	15	1000		15000		1.815	0.004	5	P	TKN = 650 mg N/l
58	15	1000	8.2	15000	99.2	1.840	0.088	5		No Wasting
59	15	2000		30000		1.679	ND	5	P	TKN :460 mg N/l
60	15	2000		30000		2.682	ND	5		[New membrane]
61	15	2000		30000		3.406	ND	5		
62	15	2000		30000		3.692	0.004	5	P	
63	15	2000	330.5	30000	83.5	4.208	ND	5		
64	15	2000	53.0	30000	97.4	4.436	ND	5		
65	15	2000		30000		4.840	ND	5		
66	15	2000		30000		6.398	0.014	5		
67	15	2000		30000		6.394	0.014	5		
68	15	2000		30000		6.400	0.010	5		
69	15	2000	118.2	30000	94.1	6.810	0.010	5		Wasting 21
70	15	2000		30000		6.490	ND	5		
71	15	2000	44.9	30000	97.7	5.760	ND	5		DO = 7,90 mg/l
72	15	2000		30000		5.644	0.010	5		
73	15	2000		30000		4.854	0.012	5		
74	15	2000		30000		4.494	ND	5		
75	15	2000		30000		4.070	0.002	5		

TABLE 5A CONTINUED

Day	Influent Flow	COD Influent	COD Effluent	COD Loading	COD Removal	SUSPENDED SOLIDS Reactor	Effluent	Sludge Age	PRE	Comments
	[l/day]	[mg/l]	[mg/l]	[mg/day]	[%]	[g/l]	[g/l]	[days]		
76	15	2000		30000		4.058	ND	5	P	
77	15	2000	77.5	30000	96.1	4.384	0.012	5		DSVI = no settling
78	15	2000	232.5	30000	88.4	4.324	ND	5		pH = 5,9 | +buffer
79	15	2000	22.4	30000	88.8	3.026	ND	5		pH = 8,2
80	15	2000		30000		2.822	ND	5		excessive foaming
81	15	2000		30000		2.622	ND	5		excessive foaming
82	15	2000		30000						reactor drained
83	15	2000		30000						No Wasting
84	15	2000		30000						
85	15	2000		30000					P	sludge added
86	15	2000		30000						
87	15	2000		30000						sludge added
88	15	2000		30000		3.676	0.032	15		
89	15	2000		30000		4.550	0.014	15		
90	15	2000		30000		5.260	ND	15		DO = 5,5 mg/l
91	15	2000		30000		3.910	0.042	15	P	Temp. high - foaming
92	15	2000		30000		4.351	0.022	15		+sludge after sampl.
93	15	2000		30000		4.883	ND	15	P	
94	15	2000	204.0	30000	89.8	5.064	0.006	15		
95	15	2000		30000		6.615	0.006	15		
96	15	2000		30000		6.894	ND	15		
97	15	2000		30000		8.383	0.020	15	P	
98	15	2000		30000		8.096	0.016	15		
99	15	2000	183.6	30000	90.8	8.670	ND	15		
100	15	2000		30000		10.161	0.012	15	P	Wasting 0.667 l

TABLE 5A CONTINUED

Day	Influent Flow [l/day]	COD Influent [mg/l]	COD Effluent [mg/l]	COD Loading [mg/day]	COD Removal [%]	SUSPENDED SOLIDS Reactor [g/l]	SUSPENDED SOLIDS Effluent [g/l]	Sludge Age [days]	PRE	Comments
101	15	2000		30000		9.092	0.024	15		pH = 6.51
102	15	2000		30000		9.283	0.040	15	P	DO = 6,5 mg/l
103	15	2000		30000		8.746	0.180	15		DSVI no settling
104	15	2000		30000		8.582	0.180	15	P	
105	15	2000	175.4	30000	91.2	9.122	ND	15		
106	15	2000	273.4	30000	86.3	9.890	ND	15		
107	15	2000		30000		9.282	ND	15	P	Membrane Replaced
108	15	2000		30000		6.713	0.028	15		
109	15	2000		30000		7.038	0.056	15	P	
110	15	2000		30000		9.516	0.128	15		
111	15	2000	81.6	30000	95.9	9.832	0.042	15		DO = 7 mg/l
112	15	2000	0.0	30000	0.0	9.056	0.004	15		
113	15	2000		30000		7.868	0.056	15		
114	15	2000		30000		9.265	0.036	15		
115	15	2000		30000		9.528	0.040	15		
116	15	2000		30000		11.558	0.082	15		
117	15	2000		30000		10.282		20	P	No Wasting
118	15	2000	81.6	30000	95.9	11.094	0.028	20		DO = 7 mg/l
119	15	2000		30000		10.958	0.026	20		DSVI no sett.
120	15	2000		30000		11.029	ND	20	P	
121	15	2000		30000		11.827	0.040	20	P	XF Vel. = 3,3 m/s
122	15	2000		30000		11.219	0.002	20		pH = 6.6
123	15	2000		30000		11.741	ND	20		Wasting 500 ml
124	15	2000		30000		10.873	0.012	20	P	
125	15	2000		30000		10.932	0.016	20		

TABLE 5A CONTINUED

Day	Influent Flow [l/day]	COD Influent [mg/l]	COD Effluent [mg/l]	COD Loading [mg/day]	COD Removal [%]	SUSPENDED SOLIDS Reactor [g/l]	SOLIDS Effluent [g/l]	Sludge Age [days]	P R E	Comments
126	15	2000	12.2	30000	99.4	10.772	ND	20		
127	15	2000		30000		11.382	0.008	20		
128	15	2000		30000		10.658	0.046	20		
129	15	2000		30000		12.510	ND	20		
130	15	2000		30000		12.540	0.006	20		
131	15	2000		30000		12.064	0.030	20		DO = 7 mg/l
132	15	2000	0.0	30000	100.0	13.762	0.008	20		
133	15	2000		30000		13.602	ND	20		
134	15	2000		30000		14.050	ND	20		No Wasting
135	15	2000		30000		14.354	ND	20		
136	15	3000		45000		15.590	0.016	20		
137	15	3000		45000		16.102	ND	20		
138	15	3000		45000		17.455	ND	20	P	
139	15	3000		45000		18.822	ND	20		
140	15	3000	81.6	45000	97.3	19.285	ND	20	P	
141	15	3000		45000		19.468	0.006	20		
142								20		Overheated Reactor TERMINATE