CRANFIELD UNIVERSITY

SCHOOL OF WATER SCIENCES

PhD THESIS

Academic Years 1992-1995

Allan Mann

A Comparison of Floating and Sunken Media Biological Aerated Filters (BAF)

Supervisor Prof. T. Stephenson

March 1997

This thesis is submitted in a partial fulfilment of the requirements for the degree of Doctor of Philosophy

This thesis is dedicated to Oscar and Shorty with who's help I continue to persevere.

Abstract

The aim of this experimental work was to directly compare the performance of two types of media support in pilot scale biological aerated filter (BAF) reactors. The two media types were identical in shape and size except one was sunken with a relative density of 1.05 and the other floating with a relative density of 0.92 and made of polypropylene.

Empty bed tracer studies were initially undertaken to ascertain the hydraulic characteristics of the media types under different process conditions. Almost ideal plug-flow was seen without aeration but with aeration some mixing and by-passing was seen which increased with higher aeration rates. Aerator design and positioning had little effect on the flow and that the sunken media would perform best in downflow and the floating media in upflow during biological treatment.

Two methods of start-up were employed during unsteady state analysis, activated sludge seeding and the use of the process liquid (settled domestic sewage) at the operational flowrate. Both methods showed a similar total start-up time of 28 d. At steady state the floating media removed 78% suspended solids (SS) and 75% soluble chemical oxygen demand (sCOD) compared to 66% and 68% respectively in the sunken media. At high flowrates and during shock loadings of up to 1.5 times the nominal flowrate the floating media again showed a better performance than the sunken media. With increasing shock loadings the recovery time increased with a corresponding decrease in solids and soluble COD removal rates. At the maximum SS loading of 1.397 kg m^{-3} d^{-1} (1.403 kg m^{-3} d^{-1} sCOD) only 35% (30% sCOD) removal was seen in the sunken media compared to 60% (40% sCOD) in the floating media.

The steady state results were used to produce an empirical model relating effluent soluble COD to influent COD and reactor height. Assuming plug flow the profiles showed a first order rate of reaction. A first order empirical model was then established based on that used for trickling filters, which produced constants dependent on the media type.

During tertiary treatment low ammonia loadings of 0.2 kg m^{-3} d^{-1} restricted autotrophic growth and led to preferential growth of heterotrophic bacteria. At ammonia loadings of up to 1.16 kg m^{-3} d^{-1} nitrification was rapid. Below 5 °C nitrification was minimal but almost full nitrification was achieved at approximately 10 °C. Though backwashing was carried out only on a weekly basis, overwashing appeared to have caused a gradual reduction in treatment efficiency. During nitrification the sunken media was more efficient at ammonia removal than the floating media, indicating that the autotrophic bacteria prefer the more open structure of the sunken media. Reactor profiles indicated a zero to half order reaction.

Acknowledgements

I am grateful to the UK Engineering and Physical Sciences Research Council (EPSRC) for there financial support during my studentship and continuing research. Also many thanks to Severn Trent Water and Tetra (Europe) Ltd for their support in funding my latest research on Biological Aerated Filters.

My sincere thanks go to my supervisor, Professor Tom Stephenson, the staff and students in the School of Water Sciences at Cranfield University for their help, observations and criticisms.

Contents

Notation

A	Cross-sectional area of reactor	m^2
B_V	Volumetric loading	$kg\ m^{-3}\ d^{-1}$
C	sCOD concentration	$kg\ m^{-3}$
Ci	Influent sCOD concentration	$kg\ m^{-3}$
C_t	Tracer Concentration	$mol.\ l^{-1}$
D	Dispersion coefficient	$m^2\ s^{-1}$
H	Height up reactor	m
k	Reaction rate constant	d^{-1}
k'	Biomass constant	$kg\ m^{-3}$
k^*	Overall process constant	$kg\ m^{-3}\ d^{-1}$
L	Length	m
n	Media constant	-
N	Number of tanks	-
Q	Volumetric Flowrate	$m^3\ d^{-1}$
R^2	R-Square Value	-
t	Time	min.
$\bar{t}$	Tracer mean residence time	min.
t_m	Mean residence time (volume / flowrate)	min.
t_p	Peak residence time	min.
u	Liquid velocity	$m\ s^{-1}$
v	Volumetric flowrate	$l\ min^{-1}$
V	Volume	l
σ_θ^2	Normalised variance	-
τ	Normalised residence time	-

Figures

Tables

Chapter 1.

INTRODUCTION AND SYNOPSIS.

Biological aerated filters have been shown over the last 15 years to be very effective in both secondary and tertiary treatment of domestic, municipal and industrial wastewaters. Though the process has been introduced and developed in such a short time, such plants have been used in many cases in preference to other more established treatment types such as activated sludge plants and trickling filters (Stensel and Reiber, 1983). The main benefits of biological aerated filters over other treatment types are their small footprint size, ability to treat high organic loadings and resistance to shock loadings (Smith *et al.*, 1990).

Though the process has been well developed since its first introduction in the late 1970s and early 1980s, the number of variable process parameters involved in designing such plants has meant that the ultimate aim to design the ideal plant has been difficult to attain despite manufacturers claims (Smith *et al.*, 1990). One reason for the difficulty in designing the ideal reactor is the variability in the type of wastewaters to be treated, whether the waste is industrial such as landfill leachate or dairy effluent, or general domestic wastewater (Rusten, 1984; Rundle, 1986). Thus to design the appropriate reactor the concentration as well as variation in the nutrient types to be removed, such as suspended solids COD / BOD and ammonia, must be considered .

Many factors need to be accounted for when designing biological aerated filters. Firstly there is the choice between using the reactors in upflow (co-current) or downflow (counter-current) (Grasmick *et al.*, 1984). This decision depends on a number of factors as both method have benefits. Downflow reactors are limited to a maximum flowrate which is dependant on the height and pressure drop across the reactor bed. Upflow reactors do not have this restriction and flowrate is dependant on the pump capacity. Downflow reactors show better nitrification ability when used in combination with carbonaceous removal in single reactors. Upflow and downflow reactors show different hydraulic characteristics which are also influenced by the aeration rate and the type of media support used (Vedry *et al.*, 1994).

The type of media used is probably the most important factor to account for when designing BAFs. A number of media parameters need to be considered such as density, size, shape and surface texture. The density of the media affects the flowrate, backwashing rate and filtration capacity (Toettrup *et al.*, 1994). The size, shape and surface texture control the available surface area and attachment of the biofilm as well influencing the liquid flow, solids filtration, backwashing and oxygen transfer rate (Quickenden *et al.*, 1992; Smith and Marsh, 1995; Smith and Brignal, 19956). Thus each parameter involved in designing BAFs may affect other parameters as well as accounting for the performance of the reactors under different process conditions.

Other factors that need to be considered for optimum performance of BAF reactors are the backwashing rate and regularity, bed height and aeration rate (Quickenden *et al.*, 1992). Though capital costs in building BAF plants are comparable to those for alternative treatment methods, it is the running costs that have, in the past, restricted BAF use especially the aeration and backwashing costs (Carrand *et al.*, 1990). Subsequently the most recent developments have been to optimise aeration backwashing rates with the possibility of eliminating backwashing altogether by continuously recycling the media.

In this work the primary factor investigated was the effect the density of the media has on reactor performance under different process conditions. Initial investigation was on the flow characteristics within the reactors using the floating and sunken medias. This was undertaken using sodium chloride as a tracer in the empty beds both upflow and downflow with different aeration and liquid rates. This method of analysis was quick and simple but yielded valuable information (Levenspiel, 1972). The results indicated that for the best performance, utilising the maximum bed volume, sunken media reactors should be run downflow and floating media reactors should be run upflow. The tracer analysis also showed the effect aeration had on the systems. Without air the reactors were shown to be almost ideal plug-flow (Sater and Levenspiel, 1966). With aeration some mixing and by-passing within the reactors was seen which increased with increased aeration which led to a reduction in the volume of bed utilised (Canziani, 1988). Thus optimisation of the aeration rate in reactors is required not only to reduce running costs and obtain maximum oxygen utilisation but also to maintain the maximum bed volume used for biological growth and nutrient removal (Lee and Stensel, 1986). Thus the tracer studies indicated that in upflow reactors floating media would perform better than sunken media in biological treatment.

Following the tracer studies, start-up of the reactors was undertaken. The two methods employed, activated sludge seeding and the use of the process liquid (settled domestic sewage) (Smith *et al.*, 1990; Bacquet *et al.*, 1991; Park and Ganzczarczyk, 1994) at the operational flowrate, showed that the total start-up time for each was similar. Though the time was similar, the use of seeding produced a more stable biofilm, indicated by a more controlled effluent quality. The use of seeding showed a more rapid response but the initial time involved in recycling the activated sludge resulted in a longer start-up time. The use of the process liquid without seeding showed an initial rapid response though to obtain high removal rates of over 40% required a long period of time. The floating media showed a more rapid start-up with both suspended solids removal and soluble COD removal. This appears to be caused by the greater ability of the floating media to filter solids. In the case of activated sludge seeding it allowed biomass flocs tobe held within the bed giving the biomass a longer contact time with the media and subsequently improved biofilm attachment. From results showed that activated sludge is recommended for improved start-up of BAF reactors, though a number of factors must be taken into account. The biomass should be taken from an appropriate source, depending on the type of treatment (secondary or tertiary). Improved start-up may be seen if the seed was taken from an alternative fixed-film process such as trickling filter or rotating biological contactor humas as such sludge will contain a higher proportion of fixed biomass than suspended biomass compared to activated sludge.

During periods of increased flowrates and shock loadings the floating media, again, showed a better performance than the sunken media. With increased flowrates there was a rapid drop in the quality of the effluent leaving the reactors. This appeared to be due to sloughing of the biomass through increased shear which caused an increase in the suspended solids concentration and a reduction in soluble COD removal. At flowrate increases at the lower range recovery was within approximately 8 h but at the higher range recovery was much slower with a large reduction in the total removal rates. The reason for this poor response was most likely due to the media used. To obtain media identical in shape and size but with different densities, the use of plastic was required. This resulted in the media having a smooth surface when extruded to form the grains. It can be concluded from this that for a reactor to withstand shock loadings, a rough texture media is required for better biomass attachment. Though the consequence of this is a higher backwash requirement (Hamoda *et al.*, 1987; Gray, 1993).

The earlier tracer studies indicated that the flow through the reactors was not ideal plug-flow but some mixing and by-passing occurred due to the air flow (Mushu, 1990). Plug-flow was assumed though to attempt to produce a basic empirical model for BAFs (Grady, 1983; Bishop and Rittman, 1995). Soluble COD measurements taken across the height of the reactor produced good first order profiles which suggested an overall first order reaction with soluble COD removal. A basic first order empirical model was thus established to describe reactor performance relating effluent soluble COD to influent soluble COD and reactor height. The constants produced in this model, which was based on that used for trickling filters, were dependant on the media type. Other factors though affect these constants. These include liquid and air flowrates, as these affect the hydraulic characteristics of the reactors, and the rate of backwashing. This model was produced as a base for future work since these factors need to be accounted for.

Nitrification within the reactors was limited during secondary treatment. When the reactors were run for tertiary treatment performance was restricted by a number of factors. Low ammonia loadings during start-up restricted growth, possibly through arestriction in nutrient availability and through preferential growth of carbonaceous removing bacteria. Once ammonia loading increased start-up was rapid. Thus high ammonia loadings are required during start-up of nitrifying reactors to allow optimum treatment. Temperature also showed some effect on treatment efficiency. At temperatures below 5 °C nitrification was minimal but almost full nitrification was achieved at approximately 10 °C indicating good performance at low temperatures (Iida and Teranishi, 1984). Backwashing affected performance over the experimental period. Unlike backwashing in secondary processes which is required approximately every 12 to 48 h due to the rapid growth of carbonaceous bacteria, the slower growing nitrifying bacteria require less frequent backwashing (Rogalla and Payraudeau, 1988; Dillon and

Thomas, 1990; Smith and Hardy, 1992). Though backwashing was carried out only on a weekly basis during this experimental work, there appeared to be a gradual reduction in treatment efficiency, possibly through overwashing. This indicates the need to carry out backwashing based on pressure drop across reactor beds rather than on a time basis, both to reduce process costs and reduce the possibility of overwashing (Robinson *et al.*, 1994).

Backwashing is an important parameter in the use of biological aerated filters Smith and Brignal, 1996). The rate and frequency of backwashing ultimately controls the amount of biomass within reactors and the subsequent treatment efficiency (Dillon and Thomas, 1990). The direction of flow, air scour rate and the frequency of backwashing depends on the type of media and the process flowrate (Dillon and Thomas, 1990). The use of downflow drainage of liquid during the experimental analysis in these upflow reactors resulted in no blockage of the reactors thus no additional pumping of backwash water was required. The floating media though showed a better response to this type of backwashing than the sunken media which would more likely benefit from upflow liquid backwashing.

Overall the results have indicated that the ideal system using floating media should have process liquid upflow in the direction of bed compression, and downflow backwashing against the line of bed compression. In sunken media reactor the opposite parameters should be used with process liquid downflow and backwashing upflow. The ideal media is required to be optimised. It should be small in size to allow good filtration but not so small to result in rapid blockage of the reactor and the surface texture should be rough but this roughness should be limited to restrict media attrition and backwash requirements.

REFERENCES

Bacquet, G., Joret, J.C., Rogalla, F., Bourbigot, M.M. (1991). Biofilm Start-Up and Control in Aerated Biofilter. *Environ. Tech.,* 12, 747-756.

Bishop, P.L., Rittman, B.E. (1995). Modelling Heterogeneity in Biofilms. Report or the Discussion Session. *Wat. Sci. Tech.*, 32, 8, 263-265.

Carrand, G., Capon, B., Rasconi, A., Brenner, R. (1990). Elimination of Carbonaceous and Nitrogenous Pollutants by a Twin Stage Fixed Growth Process. *Wat. Sci. Tech.*, 22, 1/2, 261-272.

Canziani, R. (1988). Submerged Aerated Biofilters. IV - Aeration Characteristics. *Ingegneria Ambientale*, 17, 627-636.

Dillon, G.R., Thomas, V.K. (1990). A Pilot-Scale Evaluation of the 'Biocarbone Process' for the Treatment of Settled Sewage and for Tertiary Nitrification of Secondary Effluent. *Wat. Sci. Tech.*, 22, 1/2, 305-316.

Grady, L. (1983). Modelling of Biological Fixed Films - A State-of-the-Art Review. Fixed-Film Biological Processes for Wastewater Treatment. Wu, Y.C., Smith, E.D. (Editors), New Jersey, USA, 75-134.

Grasmick , A., Elmaleh, S., Yahi, H. (1984). Nitrification by Attached Cell Reactors Aerated at Co- or Counter-Current. Experimental Data and Modelling. *Wat. Res.*, 18, 885-891.

Iida, Y., Teranishi, A. (1984). Nitrogen Removal from Municipal Wastewater by a Single Submerged Filter. *J.WPCF.*, 56, 251-258.

Lee, K.M., Stensel, H.D. (1986). Aeration and Substrate Utilisation in a Sparged Packed Bed Biofilm Reactor. *J. WPCF.*, 58, 1066-1072.

Levenspiel, O. (1972) Chemical Reaction Engineering. *2nd Ed., Wiley, New York.*

Mushu, Y. (1990). The Use of Dispersed Flow Models in Design of Biofilm Reactors. *Wat. Air Soil Pollut.*, 53, 297-314.

Park, J.W., Ganczarczyk, J.J. (1994). Gravity Separation of Biomass Wahed Out from an Aerated Submerged Filter. *Environ. Tech.*, 22, 181-189

Quickenden, J., Mittal, R., Gros, H. (1992). Effluent Nutrient Removal with Sulzer Biopur and Filtration Systems. *Europ. Conf. Nut. Rem. from Wastewater,* Sept., Wakefield.

Robinson, A.B., Brignal, W.J., Smith, A.J. (1994). Construction and Operation of a Submerged Aerated Filter Sewage Treatment Works. *J.IWEM.*, 8, 215-227.

Rogalla, F., Payaudeau, M. (1988). Tertiary Nitrification with Fixed Biomass Reactors. *Wat. Supply,* 6, 347-354.

Rundle, H.(1996). Experiences with Biological Aerated Filters for Treatment of Settled Sewage and Dairy Effluent. *Proc. 2nd BAF Symp.,* Cranfield, UK.

Rusten, B. (1984). Wastewater Treatment with Aerated Submerged Biological Filters. *J. WPCF,* 56, 424-431.

Sater, V.E., Levenspiel, O. (1966). Two Phase Flow in Packed Beds. *I & E C Fundamentals,* 5, 86-92.

Smith, A.J., Brignal, W.B. (1996). Trouble Shooting and Optimisation of BAF Systems. *Proc. 2nd BAF Symp.,* Cranfield, UK.

Smith, A.J., Hardy, P.J. (1992). High Rate Sewage Treatment Using Biological Aerated Filters. *J. IWEM.*, 6 ,179-193.

Smith, A.J., Marsh, P.(1995). Enhanced Wastewater Treatment with Lamella Tube Settlers, Sunbmerged Aerated Filters and Ultra-Violet Radiation. *Proc. Water Environ. Fed. 68th Annual Conf.*, Miami, 627-637.

Smith, A.J., Quinn, J.J., Hardy, P.J. (1990). The Development of an Aerated Filter Package Plant. Proc. *1st Int. Conf. Advances in Wat. Treat. Environ. Mang.*, 27-29 June, Lyon, France.

Stensel, H.D., Reiber, S. (1983). Industrial Wastewater Treatment with a New Biological Fixed-Film System. *Environ. Prog.*, 2, 110-115.

Toettrup, H., Rogalla, F., Vidal, A., Harremoes, P. (1994). The Treatment Trilogy of Floating Filters : From Pilot to Prototype to Plant. *Wat. Sci. Tech.*, 29, 10-11, 23-32.

Vedry, B., Paffoni, C., Gousailles, M., Bernard, C. (1994). First Months Operation of Two Biofilter Prototypes in the Waste Water Plant of Acheres. *Wat. Sci. Tech.*, 29, 10-11, 23-32.

Chapter 2.

SUBMERGED GRANULAR AND STRUCTURED MEDIA BIOLOGICAL AERATED REACTORS - A REVIEW.

Submitted for Publication,

Water Research.

SUBMERGED GRANULAR AND STRUCTURED MEDIA BIOLOGICAL AERATED REACTORS - A REVIEW.

Allan Mann and Tom Stephenson.

School of Water Sciences, Cranfield University, Cranfield, Bedford. MK43 0AL, UK.

INTRODUCTION

Introduction

Since the late 1970s interest in 'novel' fixed-film wastewater treatments has greatly increased, with research carried out on process such as fluidised beds, rotating biological contactors, contact aerators and biological aerated filters (BAFs) (Stensel and Reiber, 1983; Stephenson *et al.*, 1993). Contact aerators have been used since the beginning of the century, especially in the United States but it was only in the early 1980s that processes using structured plastic media instead of natural media were developed (Quickenden *et al.*, 1992). Parallel with the work on submerged structured media, development was carried out on granular media biological aerated filters (Smith *et al.*, 1990; Stephenson *et al.*, 1993). The primary functions of these two treatment types is to remove influent organic matter and suspended solids. Organic substrates are removed by the attached biomass which also removes a proportion of the suspended solids (Bishop and Rittman, 1995). Most of the suspended solids though are removed physically through settlement in structured media systems or filtration in granular media reactors (Pujol *et al.*, 1992b).

The use of blocks or modules of structured media allow easy design and construction of reactors but also allows easy access within the reactors if necessary (Clarke, 1996). The high voidage of such media causes the flow within reactors to be completely mixed, resulting in good oxygen transfer and thus high nutrient removal. The use of structured media though limits solids removal and thus secondary clarification is required (Boller, 1987). Limited work on randomly packed media such as Flocor and pumice has also been carried out (Young *et al.*, 1975). The design of granular media reactors is much more complex. Due to the plug flow characteristics of such reactors different design parameters affect the amount of mixing and biological growth that occurs and subsequently affects performance (Mann *et al.*, 1995). The media type affects also BAF reactor performance through variations in size, shape, density and roughness. Reactors may also be upflow or downflow, i.e. co-current or counter-current with the air flow (Tschui *et al.*, 1994). The plug flow nature of such reactors also limits the oxygen transfer efficiency, though hold-up of the air within the media improves the transfer rate. Increasing the aeration rate also leads to improves oxygen transfer though this also increases mixing. The mixed flow that occurs when structured media is used limits such reactors to one process type (carbonaceous removal or nitrification) (Faup *et al.*, 1982). With granular media BAFs the flow is plug flow and thus it is possible to carry out solids removal, carbonaceous removal, nitrification and by having part of the bed unaerated, denitrification, in one unit. Backwashing is required to remove excess biological growth and captured solids. Structured media reactors require only infrequent or no backwashing whereas granular media reactors require washing on a regular basis, the frequency of which depends on the process type (secondary or tertiary) to avoid clogging (Park and Ganczarczyk, 1994).

There are claimed to be a great number of benefits with the use of structured and granular media submerged aerated bioreactors over other treatment methods (Dumbleton, 1992a; Dumbleton, 1992b). The modular design and compactness of such reactors, which may be covered or underground gives them the benefit of a small 'footprint' process (Gilles, 1990; Pujol *et al.*, 1992b; Churchley, 1995). Start-up is rapid and high effluent quality is obtainable, even at high organic or hydraulic loadings. Though plant costs are comparable to those for similar processes, running costs tend to be high primarily due to the cost of backwashing and aeration. Part of the research carried out at present is to optimise the process and improve the control of parameters such as backwashing frequency and aeration rate (Pearce , 1996; Clarke, 1996).

History and Development

The first mention of submerged aerated reactors using fixed media occurred in 1913, when layers of slate were used as support media for biological growth. These 'submerged contact aerators' or 'submerged aerators' are still in use today, predominantly in the United States and called Contact aerators (Clark, 1930; Rittman and McCarty, 1980). By the late 1920s only limited work had been carried out on such processes intermediate between activated sludge and trickling filters (Stephenson *et al.*, 1993). These processes used natural materials such as brushwood or straw and avoided using fine grain materials such as sand due to clogging problems (Buswell and Pearson, 1929). In the 1930s work was carried out using 'Tank Filters', or 'Emscherfilters'. These reactors used coarse slag as the support media. Initially the slag was employed to increase the air residence time and thus the oxygen utilisation in activated sludge plants. It was found though that the slag retained and encouraged biomass growth. Analysis of this biological matter indicated that it more closely resembled humus biomass found in trickling filters than that found in activated sludge (Bach, 1937).

In the United States work continued throughout the 1930s, 40s and 50s on the development of contact aerators and the Hays process which were structured media rather than granular media processes (Griffith, 1943; Lackey and Dixon, 1943; Fair *et al.*, 1948; Fair *et al.*, 1950; Wilford and Conlon, 1957). During the 1960s two areas of work were carried out in Britain. Firstly in 1964 a patent was brought out describing a submerged aerated process using coke as the biological support media (Albright and Wilson, 1964). In 1967 work was carried out using a support medium of short PVC tubes (Ministry of Technology, 1967).

By the late 1960s and early 1970s work continued on submerged reactors using large coarse media. Initially these upflow reactors were used for anaerobic treatment but were investigated for aerobic treatment through both direct and indirect oxygenation and used primarily for nitrification (McCarty and Haug, 1971; Haug and McCarty, 1972). In 1975 packed-bed reactors (PBRs) were investigated. Again aerobic PBRs were developed from anaerobic filters but differed from the previous work by using air and not oxygen.

They also used silica sand, anthracite coal and plastic media (Young *et al.*, 1975; McHarness *et al.*, 1975). As well as in the United States, work was carried out in Japan on submerged fixed media processes (Iwai *et al.*, 1977). It was only in the late 1970s and early 1980s that the BAF was developed in Europe as a more compact but as efficient process as trickling filters and activated sludge plants (Pujol *et al*, 1994). Throughout the 1980s BAFs were known by a variety of names including packed bed reactors, submerged aerated filters and immersed aero-filters (Young and Stewart, 1979; Faup *et al.*, 1982; Rusten, 1984). Work continued during the 1980s in the development of BAFs specifically for secondary and tertiary, taking place primarily in North America, Europe and Japan (Condren, 1990; Canler and Perret, 1994; Carrio *et al.*, 1995). It was only in the late 1980s and early 1990s development work began in the UK (Dillon and Thomas, 1990; Smith and Hardy, 1992). Development has now expanded to the Far East and China (Peng *et al.*, 1995).

Between the late 1980s and the present day work on developing submerged aerated biofilters has continued with a large number of proprietary system now available (Table 1). At present such reactors are used in denitrification and phosphorus removal as well as solids and carbonaceous removal (Rogalla *et al.*, 1990a; Sammut *et al.*, 1992; Henze, 1995).

TABLE 1. Proprietary Submerged, Aerated, Fixed Film Systems.

Process Name	Media	Liquid Flow	Manufacturer
ABF Biofilter	Structured	Downflow	Waterwise Services Ltd.
BAF	Structured	Up/Downflow	Copa Ltd.
Biobead	Floating	Upflow	Brightwater Engineering Ltd.
Biocarbone	Sunken	Downflow	Biwater / OTV
Biofor	Sunken	Upflow	Degremont UK Ltd.
Biopur	Structured	Downflow	John Brown Eng. / Sulzer
Biostyr	Floating	Upflow	Biwater / OTV
ColOX	Sunken	Up/Downflow	Tetra (Europe) Ltd.
CTX Bioreactor	Structured	Up/Downflow	Hodge Stetfield Ltd.
FAST	Structured	Up/Downflow	Promech
Hi-Paf	Structured	Up/Downflow	WPL Ltd.
REBAF	Floating	Up/Downflow	Comenco
SAFe	Sunken	Downflow	PWT Projects Ltd.
SAM	Structured	Downflow	EWS Clearwater

PLANT DESIGN

Liquid Flow and Hydraulics

The hydraulic residence time and hydrodynamics of reactors affect biofilm growth and subsequently nutrient removal. Flow characteristics are influenced by the liquid flow, air flow and media used, especially with the use of granular media (Le Cloirec and Martin, 1984; Lee and Welander, 1994).

Structured media reactors have been found to be completely mixed due to the high void volume. This high degree of mixing improves oxygen and substrate transfer though complete mixing reduces the efficiency of single reactors to remove high levels of more than one substrate (BOD or ammonia). This is primarily due to competition between bacterial types (Boller *et al.*, 1994; Zhang and Bishop, 1994a). On the other hand plug flow found in granular medias produces zones of different nutrient strengths and thus allows the growth of different bacterial species. One advantage of completely mixed reactors over plug flow reactors is the effect of toxic or inhibitory substances. Unlike plug flow systems, in completely mixed systems inhibitory substances are rapidly dispersed and do not remain concentrated thus the effect on reactor performance is limited (Chen and Cheng, 1994).

Structured and granular media reactors may be run co-current or counter-current to the air flow i.e. upflow or downflow (Fig 1 and 2) (Grasmick *et al.*, 1984). In upflow systems the raw wastewater is introduced at the base of the reactor, with the treated wastewater leaving at the top, thus avoiding odour problems through air stripping (Iida and Teranishi, 1984; Newbigging *et al.*, 1995). It has also been claimed that upflow systems use the volume of the reactor more efficiently. Subsequently suspended solids and biomass within reactors is held more uniformly, which allows high influent flowrates to be used, longer run times and lower backwash volumes. Better oxygenation is obtainable in upflow filters through non-coalasence of air bubbles. In downflow systems liquid enters at the top, thus avoiding blockage of influent nozzles that may occur in upflow systems (Desbos *et al.*, 1990; Newbigging *et al.*, 1995). Downflow systems also make it possible to introduce air part way up the reactor bed, thus a denitrification zone may be produced at the base of the bed, following carbonaceous removal and nitrification (Bacquet *et al.*, 1991). The high oxygen level required for nitrification occur at the base of beds, thus high ammonia removal is possible in downflow systems where carbonaceous levels at the base of beds are low (Gonzalez-Martinez and Wilderer, 1991). Downflow systems, unlike upflow systems, claim to prevent intermixing and integrating of the support media. This could be particularly advantageous for BOD and nitrification single-step filters or effluent polishing reactors. This is because intermixing of heterotrophic and autotrophic bacteria is subsequently reduced, reducing competition (Faup *et al.*, 1982; Newbigging *et al.*, 1995).

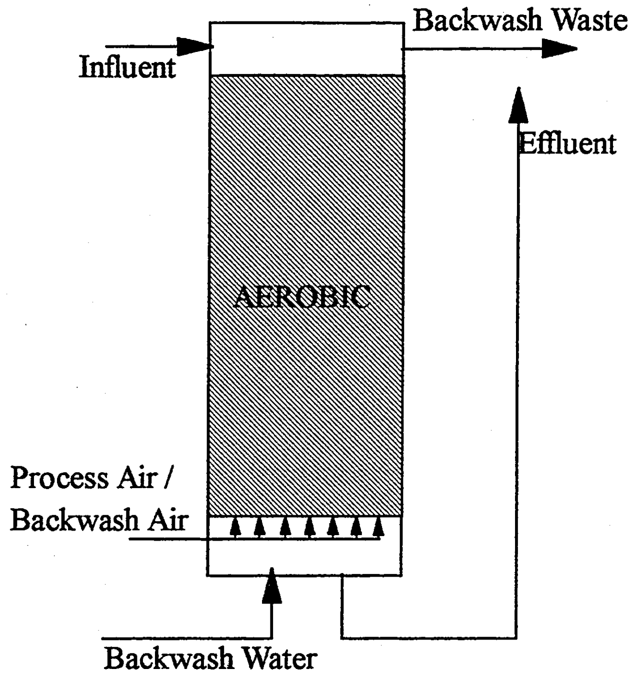

Figure 1a. Downflow Structured Media Reactor

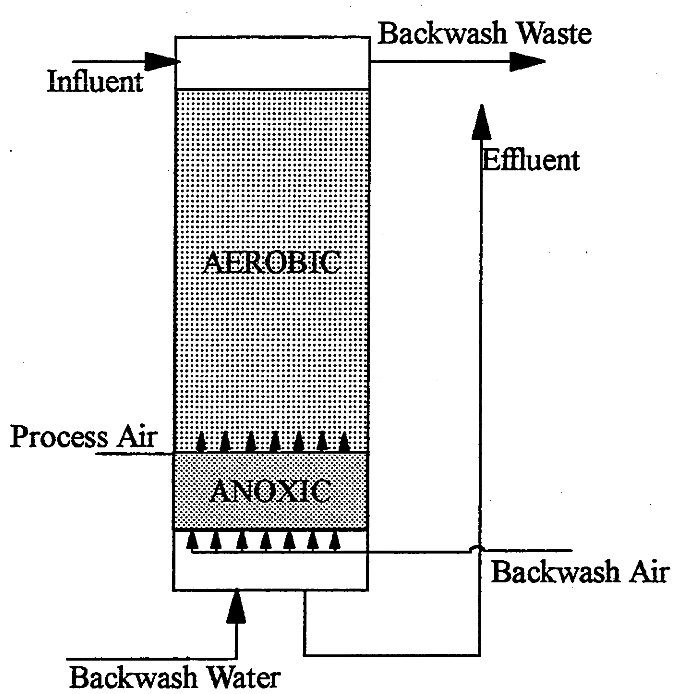

Figure 1b. Downflow Granular Media Reactor (With Anoxic Zone).

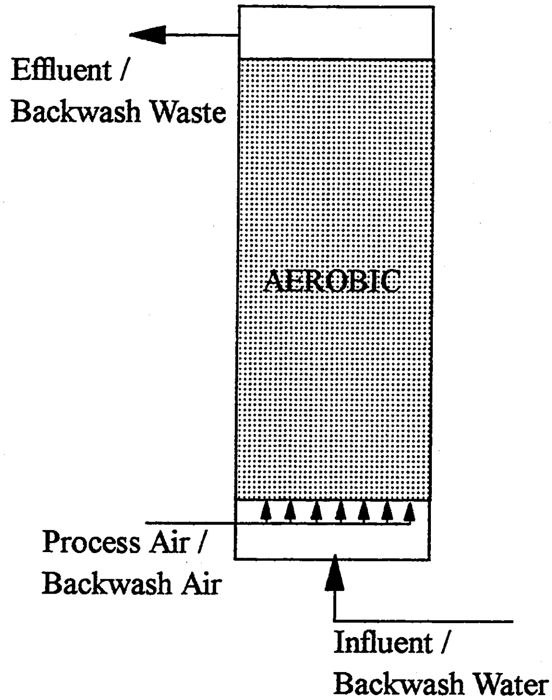

Figure 2a. Upflow Granular Media Reactor

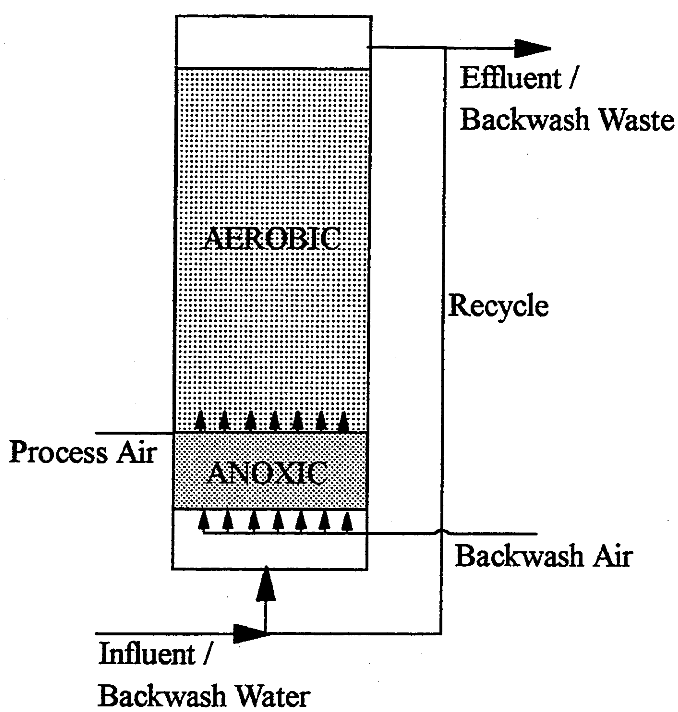

Figure 2b. Upflow Granular Media Reactor (With Recycle and Anoxic Zone).

During backwashing, submerged aerated filter media is subject to a degree of regrading after the wash cycle (Akunna *et al.*, 1994). In downflow systems the filter media regrades with a tendency for the media holding the heterotrophic bacteria , which is slightly less dense to move to the top of the reactor and the media holding nitrifying autotrophic bacteria to move to the base of the reactor. In upflow systems mixing of biofilm populations causes reduced treatment efficiency (Wilderer *et al.*, 1995). BAFs require backwashing to remove excess biofilm and solids entrapped within the media. Upflow systems have the advantage over downflow systems since backwash water can be stored above the reactor whereas in downflow systems additional storage tanks are required to hold the backwash water. Although upflow systems are claimed to require less process air, this volume of backwash water above the media can result in higher requirements for backwash blowers and the water supply is limited should prolonged washing be required (Jepsen and Jansen, 1993; Visvanathan and Nhien, 1995).

Media Types

During the early development of submerged, aerated, fixed film reactors, the media used tended to be slate plates or corrugated asbestos plates. Later developments used random media such as Flocor or pumice previously used in trickling filters (Ruffer and Rosenwinkel, 1984; Mesdaghinia, 1986). Present developments tend use structured modular plastic media, granular media denser than water or granular media lighter than water (Table 2).

Modular media made of materials such as polypropylene has been used in place of man-made random media such as 'Flocor'. This is because, though both types have similar specific surface areas (200-500 m^2m^{-3}), modular media allows ease of construction, easy access to the base of reactors if required and may be replaced easily or built on and in many cases modular media has replaced media used in trickling filters (Carrand *et al.*, 1990; Gros, 1991; Quickenden *et al.*, 1992). The high void voidage of structured media restricts the filtration capability but reduces the possibility of clogging (Rusten and Odegaard, 1986; Boller, 1987; Anderson, 1989). Thus generally secondary clarifiers are required except when wastewaters with only low concentrations of suspended solids are to be treated, such as in tertiary treatment, or the liquid and air velocities are low enough to allow settlement within reactors (Boller, 1987; Churchley *et al.*, 1995). Though post clarifiers are often required backwashing is generally not required as excess biomass tends to be removed from the base of reactors or in the secondary clarifier, following sloughing. Structured media also has the benefit of no media loss or attrition and low head loss (Gray, 1993).

The use of granular media and its related filtration ability removes the need for secondary clarifiers but subsequent backwashing is required to remove excess biomass and captured

TABLE 2. Types of Media Used in Lab, Pilot and Full Scale Analysis

Media	Media Type	Media Size (mm)	Liquid Flow	Reference
Structured	Retriform Plastic	-	Upflow	Iida and Teranishi, 1984
	Corrugated Plastic	-	Downflow	Ryhiner *et al.*, 1992 Tschui *et al.*, 1994
	Plastic (PLASdek)	-	Upflow	Rusten, 1984 Rusten & Thorvaldsen, 1983
Sunken	Expanded Clay	~ 1.3	Upflow	Sagberg *et al.*, 1992
		6 - 8		Fruhen *et al.*, 1994
		~ 1.9		Smith and Hardy, 1992
		-		Carrand *et al.*, 1990
		~ 3.5		Vedry *et al.*, 1994
	Vitrified Clay	3 - 4	Downflow	Clapp *et al.*, 1994
	Expanded Shale	3 - 6	Downflow	Bacquet *et al.*, 1991
		2 - 6		Robinson *et al.*, 1994
		-		Smith and Hardy, 1992
		3 - 6 (Nit.)		Dillon and Thomas, 1990
		2 - 5 (Carb.)		Dillon and Thomas, 1990
		-		Paffoni *et al.*, 1990
		-		Rogalla and Payraideau, 1988
		2.5 - 3.5		Tschui *et al.*, 1994
	Porous Stone	20 - 35	Upflow	Costa Reis and Sant'Anna, 1985
	Slag	~ 40	Downflow	Dee *et al.*, 1994
	Activated Carbon	4 - 6	Upflow	Ros and Mejac, 1991
	Granular	~ 2.8	Upflow	Jimenez *et al*, 1987
		3 - 6	Downflow	Pujol *et al.*, 1994
		2.5 - 3.5	Upflow	Pujol *et al.*, 1994
	Polypropylene	2.3 - 2.7	Upflow	Mann and Stephenson, 1996
Floating	Expanded Polystyrene	2 - 3.5 3 - 3.5	Upflow	Visvanathan and Nhien, 1995 Tschui *et al.*, 1994
	Polypropylene	2.3 - 2.7	Upflow	Mann and Stephenson, 1996

solids (Dillon and Thomas, 1990; Smith and Hardy, 1992; Wheale and Cooper-Smith, 1992; Gray, 1993). Though the hydraulic nature of structured media reactors allows better oxygen and substrate transfer than granular media reactors through mixing, the higher specific surface area (1000-1500 m^2m^{-3}) of the granular media allows a higher biomass concentration. The type of granular media used varies. Sunken medias tend to be made of natural products such as shale, activated carbon or phyllosilicates such as baked clay or pozzolana (Ros and Mejac, 1991; Dauthuille, 1990; Pujol *et al.*, 1992b; Vedry *et al.*, 1994). Floating media though tends to be produced from man-made products such as polyurethane or polystyrene (Gray, 1993; Vedry *et al.*, 1994; Altemeier *et al.*, 1995; Andersen *et al.*, 1995).

The size of granular media used affects the efficiency of treatment (Smith and Marsh, 1995; Cantwell, 1996). The use of large media causes a reduction in nutrient and solids removal through high void spaces and a reduced area available for biofilm growth. Large media though reduces backwashing requirements and thus reduces process costs (Costa Reis, 1985; Pujol *et al.*, 1994). Small media allows good filtration and a large surface area for biofilm growth though more frequent and higher rate backwashing is required (Smith and Brignal, 1996). Fine media is thus ideal for effluent polishing where there is only a small fraction of suspended solids and larger media should be used for roughing with intermediately sized material used for general treatment (Quickenden *et al.*, 1992).

Media size consequently affects biofilm growth through the area available for biofilm growth and hydraulic characteristics of reactors, aeration and oxygen utilisation rates and backwashing frequency. The shape of the media also affects reactor performance. To obtain the greatest surface area to volume ratio, i.e. the maximum growth area per volume of media, the media should be spherical in shape. On the other hand irregular grains have been found to improve performance, possibly through break-up of air bubbles passing through the reactor or through variation in the size of the void spaces (Sagberg and Berg, 1996). A media characteristic that affect reactor performance is the density. Most BAF reactors use sunken media, whether the process is upflow or downflow. Backwashing of such medias is required to be upflow to allow partial or total bed fluidisation, requiring high pumping rates. A more recent advance is the use of floating media (Toettrup *et al.*, 1994). This media works in reverse to standard gravity filtration. Process liquid flow tends to be upflow requiring only simple backwashing, usually through gravity drainage. The final media parameter that affects reactor performance is the roughness. Smooth media surfaces restrict biofilm growth as the biofilm is unable to adequately attach itself to such medias. Reactors that use such media are unstable with variable linear air and liquid velocities causing biofilm sloughing. Smooth medias though reduce the need for high rate backwashing and thus reduce process costs. Rougher medias though allow bacteria to firmly attach themselves, producing more stable biofilms but require higher backwashing rates. Rough medias have also been found to improve solids retention (Ruffer and Rosenwinkel, 1984).

Aeration

Aeration constitutes a large proportion of the running costs when operating biological aerated filters (Smith and Brignal, 1996). In structured media reactors, variation in the aeration rate largely unaffects the parameters of such reactors due to their completely mixed nature (Rusten, 1984). In granular media reactors the rate of aeration is critical for optimum performance. In such reactors the maximum amount of oxygen utilised (2-22%) and oxygen transfer rate is low. Thus only relatively low aeration rates are required. If the aeration rate is too low though, there is insufficient substrate removal and anaerobic zones may be created within the reactor. In addition there may be early blockage of the bed close to the influent inlet, resulting in insufficient bed utilisation. If the aeration rate is too high scouring of the biofilm and a reduction of the solids removal efficiency may occur. It has also been found that low aeration rates lead to better oxygen transfer but also cause a reduction in the ammonia removal efficiency (Pearce, 1996).

The rate of aeration is primarily dependant on the media type used. Oxygen transfer to the bulk liquid is controlled by the size of bubbles produced and the hold-up time within the reactors. The size of the air bubbles within the media is not controlled by the aeration system and thus is largely unaffected by the type of aerator used (Caziani, 1988; Mann *et al.*, 1995). These are further controlled by the shape, size and density of the media (Sagberg and Berg, 1996). As well as the media type, the rate and direction of liquid flow affects the optimum aeration rate. In upflow systems, liquid and air flows in the same direction and variation in one of these parameters largely unaffects the other. In downflow systems liquid and air flows are in opposition. Consequently low liquid flowrates with high aeration rates leads to increased hold-up of the liquid. With high liquid flowrates and low aeration rates there is subsequent hold-up of the air (Pearce, 1996).

To monitor and control the required aeration rate for optimum reactor performance, measurement of the dissolved oxygen concentration is not sufficient. Required oxygen not only comes from the bulk liquid but also from direct contact between the biofilm and air bubbles. Thus the best method for controlling the aeration rate is through off-gas measurements (Lee and Stensel, 1986; Pearce, 1996).

Backwashing

Backwashing has two functions, to remove captured solids and remove excess biomass in granular media reactors and predominantly to remove excess biomass when used for structured media (Park and Ganczarczyk, 1994; Smith and Brignal, 1996). Backwashing rates and intervals are thus determined by the solid captured and the biofilm growth rate. Solids removal depends on factors such as the properties of the captured solids (size and inertness), the media support used (shape and size), the biofilm structure and the hydraulic characteristics of the reactors used (nominal flowrate and aeration rate) (Arvin and Harremoes, 1989). The amount and size of the particulate material in the influent may also affect the biofilm growth which is also affected by the reactor and media design, nature of the wastewater to be treated and growth and substrate kinetics (Jimenez *et al.*, 1987). There are two parameters in backwashing, dislodging of filtered solids and excess biomass through high shear from high rate aeration (air scour) and the removal of these dislodged solids through flushing with high rate liquid flows (Quickenden *et al.*, 1992).

In structured packing backwashing tends to be carried out infrequently, when used. In reactors that use such media solids removal from the process liquid is carried out either through settlement or through bioadsorption (Ryhiner *et al.*, 1994). Captured solids and excess biomass are removed either periodically from the base of the reactor or in the secondary clarifier (Chudoba, *et al.*, 1991). In granular medias clogging occurs, thus frequent backwashing is required (Pujol *et al.*, 1992a). The frequency depends primarily on the size of the media and the characteristics of the wastewater to be treated (Smith and Brignal, 1996). Reactors with stable, rapidly growing biofilms involved in carbonaceous removal (secondary treatment) require more frequent and high rate backwashing than reactors with less stable and slower growing biofilms used in nitrification (tertiary treatment) (Dillon and Thomas, 1990). The shape, density size and therefore voidage of the media affects the pumping requirements needed to wash the bed. Other factors that influence the frequency of backwashing through the influence on biofilm growth include organic loading rates and biomass growth rate (Robinson *et al.*, 1994). BAFs are normally used for secondary or tertiary treatment. When used for secondary treatment, backwashing is carried out every 12 to 48 h. In tertiary treatment backwashing is less frequent due to lower biofilm growth through reduced nutrient levels and subsequently carried out only on a weekly basis (Dillon and Thomas, 1990; Bacquet *et al.*, 1991, Smith and Hardy, 1992).

There are three methods used to determine the frequency of backwashing. The simplest method used is to backwash on a regular basis, usually every 24h. This method does not require additional monitoring equipment but does not allow the optimum time period between washes to be used (Dillon and Thomas, 1990; Bacquet *et al.*, 1991; Wheale and Cooper-Smith, 1995). The second method used is to carry out backwashing when a pre-set head loss is reached. This method allows for the treatment of variable strength wastewaters though optimisation of the time between backwashings (Faup *et al.*, 1982; Jimenez *et al.*, 1986; Amar *et al.*, 1986; Paffoni *et al.*, 1990; Jepson and Jansen, 1993). The third method monitors effluent quality and subsequently controls backwash frequency through reactor performance (Gray, 1993).

The type of backwashing carried out is determined largely by the process configuration and on each individual process. Each process usually goes through stages of 1) air scour, 2) air scour with liquid backwash and 3) liquid backwash only, though simple use of air scour and liquid drainage has been used successfully in upflow reactors (Quickenden *et* *al*., 1992). Generally backwashing is carried out upflow in reactors that use sunken media, in opposition to bed compression whether the process liquid is either upflow or downflow (Jimenez *et al*., 1987; Kurbiel, 1987; Rogalla and Bourbigot, 1990; Pujol *et al*., 1992a). Reactors using floating media though backwash downflow, again in the opposite direction to media compression (Rogalla and Bourbigot, 1990; Jepsen and Jansen, 1993; Vedry *et al*., 1994; Visvanathan and Nhien, 1995).

Normal backwashing removes approximately 30% of the biomass without reducing reactor performance to any great degree (Bacquet *et al*., 1991). This results in a recovery time of up to two hours (Pujol *et al*., 1992; Robinson *et al*., 1994; Visvanathan and Nhien, 1995). Underwashing, which causes premature headloss and solids breakthrough, results in more frequent backwashing making the process uneconomical. Overwashing on the other hand causes loss of effluent quality through reduced filtration capability and low biomass concentration (Robinson *et al*., 1994). Thus a larger proportion of the biomass is removed resulting in longer recovery time for sufficient biomass growth to take place (Amar *et al*., 1986).

Due to the high operational costs of backwashing, optimisation is necessary to minimise energy consumption and the quantities of air and wash water used (Carrand *et al*., 1990). Generally the amount of liquid required for backwashing is 2 to 15% of the total treated water in BAFs which compares to 0.9 to 8.3% used in sand and anthracite filters (Ruffer and Roenwinkel, 1984: Paffoni *et al*., 1990; Dillon and Thomas, 1990; Park and Ganczarczyk, 1994; Wheale and Cooper-Smith, 1995). Other considerations such as the use of low backwash water flows allows the process liquid piping to be used for backwashing and thus reduces capital costs (Kirichenko and Drushlyak, 1988). At the same time though it is necessary to maximise the time between washes, avoid media loss and maintain a good population of active biomass within reactors (Carrand *et al*., 1990).

FIXED FILMS

Biofilm Structure and Microbiology

The structure of fixed-film biofilms is very complex and depends on environmental conditions and biofilm age (Lazarova and Manem, 1995). Though much work has been carried out on the kinetics of substrate removal, work on biofilm structure has been limited (Christensen *et al*., 1989). There are two types of bacteria in biofilms, active and inactive. The active bacteria is responsible for substrate removal from wastewaters and the inactive biofilm for the observed biomass build-up. Thus one requirement in maintaining an efficient process is keep the inactive biomass to a minimum and maintaining a thin active biofilm (Moreau, *et al*., 1994). Biofilm thickness is controlled by factors such as the growth rate and decay of active biomass and the attachment and accumulation of particulates and inert substances to the biofilm (Hoehn and Ray, 1973; Arvin and Harremoes, 1989). Biofilm thickness and solids retention is also influenced by the hydraulic characteristics of reactors. High gas and liquid flowrates cause high shear within reactors, which limits the thickness of biofims through erosion and sloughing (Lewandowski and Walser, 1991; Wilderer *et al.*, 1995). The properties of the biofilm vary with the depth and the biofilm thickness, which is not uniform over the biofilm surface. Thus substrate and biofilm kinetics are not uniform (Bishop and Rittman, 1995). Low biofilm densities have been found to inhibit substrate transport. Thus in biofilms of higher density inhibition decreases (Christensen *et al.*, 1989). Backwashing influences species distribution within reactors through bed movement (Wilderer *et al.*, 1995). The distribution of micro-organisms within a biofilm though is also dependant on the growth competition in the biofilm. The biofilm growth and decay rates are also dependant on the substrate concentration within the biofilm. Subsequently a variation in the type and amount of different species populations occurs at different depths within a biofilm (Arvin and Harremoes, 1989).

In structured media reactors that show well mixed flow, the low bulk liquid substrate concentration favours the growth of filamentous bacteria. On the other hand plug flow as seen in granular reactors, where there tends to be high inlet substrate concentrations, favours the growth of the more dense floc forming bacteria. Similarly the growth of filamentous bacteria is also favoured by high soluble substrate concentrations while high particulate concentrations favours the growth of floc formers. (Cao and Alaerts, 1995). In structured and granular media reactors the physical structure and the physiological structure of biofilms is thus affected by the hydrodynamics of reactors (Wilderer *et al.*, 1995).

BAF reactors treating settled domestic wastewaters have been found to contain a variety of bacterial genera (Zinebi *et al.*, 1994). A large proportion of bacteria is comprised of mucous colony forming bacteria (Smith and Brignal, 1996). Other genera include *Acinetobacter* (23%), *Enterobacteriaceae* (11%), *Aeromonas* (9%), *Flavobacterium* (6%), *Alcaligenes* (7%), *Pseudomonas* (4%) and *Moraxella* (3%). Zoning of particular species has not been found but the amount of each microbial species from different reactor heights has been found to vary (Giuliano and Joret, 1989). During accumulation the predominant bacteria type develops from *Proteobacteria* to *Mycobacterium* and finally *Methylcbacterium* (Fujie *et al.*, 1994). As well as bacteria fixed-film biofilms contain other organisms such as protozoa and worms which influence the population dynamics of the bacteria in the biofilm and subsequently reactor performance through predation (De Beer and Muyzer, 1995).

Mechanistic and Empirical Models

Modelling of biological systems can be categorised into two types mechanistic and empirical (Grady, 1983; Tchobanoglous and Schroeder, 1985). Mechanistic models deal with individual phenomena, applying kinetics and other reactor engineering principles and look at processes on a micro scale using factors such as biofilm growth rate and nutrient transport (Williamson and McCarty, 1976; Rittman and McCarty, 1980; Avaev, 1986; Capdeville and Nguyen, 1989; Arvin and Harremoes, 1989). Many of these models work on a one dimensional basis and often assume that certain parameters are constant which leads to inaccurate results (Furumai and Rittman, 1994; Quek *et al.*, 1995). Because some constants are considered as universal these simplified mechanistic models are easy to interpret (Harremoes, 1982; Arvin and Harremoes, 1989, Zeghal *et al.*, 1995). One assumption is that flow through granular beds is ideal plug flow though it has been found that some mixing does occur. The degree of mixing depends on the liquid and air flowrates and velocities within the reactor which subsequently affects biological performance (Jenning *et al.*, 1976; Grasmick *et al.*, 1980; Mushu, 1990). Due to the difficulties in measuring biofilm thickness models have assumed them to be uniform (Riemer *et al.*, 1980; Lee and Stensel, 1986, Hamoda, 1989; Lewandowski *et al.*, 1994; Lewandowski *et al.*, 1995). Recently though research has shown the biofilms are not uniform and properties such as density and porosity vary with depth and diversity of species occurs in different places. Thus to obtain an accurate picture of the performance of biofilms, many factors need to be accounted for with the use of two or three dimensional modelling (Bishop and Rittman, 1995). In contrast empirical models directly relate input variables such as organic loading or substrate concentration to output variables such substrate removal rate. Subsequently parameters such as mass transfer coefficients and oxygen utilisation rates do not need to be measured or directly accounted for (Meunier and Williamson, 1981, Hamoda, 1989).

Models specifically for the performance of structured and granular media reactors have been very limited and tended to be mechanistic rather that empirical (Jennings *et al.*, 1976; Hamoda, 1989). These models vary from first order to zero order depending on whether biofilm efficiency is limited by diffusion, i.e. whether external mass transfer or internal mass transfer is limiting (Harremoes, 1977; Hamoda, 1989; Janning *et al.*, 1995). The accuracy of these models varies. Though they tend to be valid, the required conditions with which these models work are limited and required to be stable over a small range (Hamoda, 1989).

<u>Biological Growth Kinetics</u>

When modelling fixed film systems work tends to be based on substrate utilisation and kinetics rather than growth kinetics. Thus minimal work has been carried out specifically on submerged fixed-film aerated bioreactors. A major problem involved in analysing biological kinetics is knowledge of the active bacterial concentration within biofilms and calculating the maximum specific growth rate and growth yield, which is often approximated as universally constant (Arvin and Harremoes, 1989).

In nutrient removal, one of the rate limiting steps is the biofilm thickness. With thin biofilms the substrate fully penetrates the biofilm and thus zero order substrate kinetics are involved. With thicker biofilms substrate penetration is limited (Arvin and

Harremoes, 1989). Bacterial growth within biofilms has been shown to follow an exponential pattern when there is no substrate limitation. It has also been found that any increase in biofilm thickness does not directly follow an increase in substrate removal rate. Thus a doubling of the biofilm thickness may result in a ten fold increase in the substrate removal. (Christensen *et al.*, 1989). Substrate concentration subsequently inhibits biological growth kinetics only when it is limited. Another factor that influences the growth kinetics of biofilms is the presence of other organisms. These organisms such as protozoa predate on the biofilm and thus influence factors such as the mass transport rate growth rates but especially the death rate and thus the overall rate of growth of the biofilm (De Beer and Muyzer, 1995).

<u>Substrate Removal Kinetics</u>

The kinetics of substrate utilisation are based on the Monod and Michaelis-Menton formulae. Here there is a gradual transition from first order to zero order kinetics which depends on the thickness and density of the biofilm. This is effected by the concentration of bacteria and the dependence between rate of growth and substrate concentration (Arvin and Harremoes, 1989).

The primary characteristic of biofilm reactor kinetics is the diffusion resistance to substrate movement and the products within the biofilm. Thus reactor performance depends on whether substrate can partially or fully penetrate the biofilm (Arvin and Harremoes, 1989). The most simple expression of diffusion resistance assumes a zero order rate of reaction (Arvin and Harremoes, 1989). More accurate results are obtained by taking into account the transition from zero to first order kinetics but the difference is only slight (Harremoes, 1977). There are three possible rate limiting factors in substrate transport : the biofilm thickness, the electron donors and the electron acceptors. With thin biofilms substrates can fully penetrate the biomass, making the biofilm thickness the rate limiting factor and the reaction zero order. Where there are limited electron donor or acceptors the substrate can only partially penetrate the biofilm, thus substrate removal is half-order (Arvin and Harremoes, 1989; Christensen *et al.*, 1989; Christiansen *et al.*, 1994).

Substrate transport was thought to be due to mainly molecular diffusion though recent work has shown that the mechanism of mass transport is more complex (Lewandowski *et al.*, 1995). Substrate removal kinetics have also tended to rely on soluble substrate removal and do not take into account non-diffusable particulate matter removal. The main processes involved in solids breakdown are the attachment of the particulate matter to the surface of the biofilm and subsequent extracellular degradation of these particulates. Very little is known about this form of solids removal with only collision factors used in some cases to account for the effect of solids (Arvin and Harremoes, 1989; Bishop and Rittman, 1995).

Oxygen Utilisation

Previous studies have found that substrate removal rates can be controlled not only by substrate concentration but also by oxygen concentration (Meunier and Williamson, 1981; Grady, 1983; Lee and Stensel, 1986). It has also been found that oxygen utilisation affects and is affected by the substrate utilisation (Reiber and Stensel, 1985; Lee and Stensel, 1986). Any increase in the substrate loading results in an increase in oxygen utilisation but bulk liquid oxygen concentrations show only slight changes (Reiber and Stensel, 1985). In reactors that use pre-aeration or pre-oxygenation, any available oxygen is dissolved within the bulk liquid. Subsequently substrate removal is limited by the DO concentration. In sparged reactors such as BAFs oxygen is supplied by interfacial contact between air bubbles and the biofilm as well as oxygen dissolved in the bulk liquid (Lee and Stensel, 1986). With low substrate loadings the primary source of oxygen is from the bulk liquid. As loadings increase oxygen is increasingly supplied by interfacial contact between air bubbles and the biofilm. With increased flow the boundary layer between the biofilm and bulk liquid is reduced which reduces the external mass transfer resistance and thus increases the oxygen mass transfer rate directly from air bubbles as well as from the bulk liquid (Zhang and Bishop, 1994b). BAFs are thus able to remove higher substrate loadings as they are not limited by bulk oxygen concentrations. This high availability of oxygen also promotes nitrification due to the high oxygen requirements which is inclined to be the main rate limiting step (Boller *et al.*, 1994; Cecen and Gonenc, 1994; Cecen and Gonenc, 1995). Thus any accurate model relating oxygen utilisation and substrate removal needs to take in account both oxygen transfer through the bulk liquid and direct contact between the air and biofilm. Models though should not rely on K_La values calculated from clean water experiments as these tend to cause underestimation of the oxygen transfer efficiency (Lee and Stensel, 1986).

Temperature Effects

In biological wastewater treatment systems the rate of substrate removal and thus performance is directly related to the temperature of the system (Fdez-Polanco *et al.*, 1994). It has been found though that variable and extreme temperatures has only a limited effect on the performance of granular and structured media submerged aerated biofilters (Iida and Teranishi, 1984; Paffoni et al., 1990; Sakuma et al., 1993). At temperatures down to 10°C there is almost no effect on the performance of such reactors. At temperatures below 5°C there is some decrease in suspended solids removal, BOD removal and nitrification with evidence that prolonged very low temperatures cause loss of biofilm (Koutsakos *et al.*, 1992). Though low temperatures cause a reduction in reactor performance, this can be improved by increasing the retention time (McCarty and Haug, 1971; Harremoes, 1982; Fdz-Polanco *et al.*, 1994). The temperature at which maximum performance occurs depends on the treatment type. In carbonaceous removal the temperature at which the maximum growth rate and substrate removal occurs is approximately 38°C (Visvanathan and Nhien, 1995). This compares with a temperature of approximately 29°C for nitrification, with a decrease of 2% in nitrification rate for every drop of 1°C (Fdz-Polanco, 1994). At higher temperatures the effect on the performance is related to the biofilm population. At high temperatures above 41°C the type of micro-organisms become thermophilic rather than mesophilic and thus alters the biological growth and substrate removal rates (Visvanathan and Nhien, 1995).

Although performance increases with temperature it has been found that the amount of biological matter decreases with temperature, with a drop of two thirds in the microbial mass between 5°C and 35°C with a subsequent decrease in the yield coefficient (Fujie *et al.*, 1994).

SUBSTRATE REMOVAL

Solids Filtration and Carbonaceous Removal

The most widespread use of submerged fixed-film biological reactors is in the removal of carbonaceous material (Table 3). Work in the early 1980s established the potential of such reactors with the advent of granular BAFs which also allowed solids filtration (Legise *et al.*, 1980; Stensel and Reiber, 1983; Hamoda and Abd-El-Bary, 1987).

Solids are removed by two methods. Particles tend to be captured on the on surface of biofilms or through filtration in granular medias. These particles are subsequently either removed through hydrolysis by exoenzymes or in the case of inert substances, with excess biomass removed during backwashing (Nielson and Harremoes, 1995). In structured medias removal is primarily through hydrolysis, bioadsorbsion or through settlement while in granular medias it is mainly through filtration (Ryhiner *et al.*, 1994). The amount of solids removed depends on the media size (Jimenez *et al.*, 1987). In fine media reactors solids are easily removed but low void spaces reduces the total capacity of such reactors, though performance is comparable to that for sand filters (Kuribayashi, 1992). Large media reactors have a higher capacity to hold solids but the minimum size of filtered solid is limited (Ryhiner *et al.*, 1994). Higher solids concentrations though may inhibit biological performance by reducing the oxygen transfer efficiency (Sundararajan and Ju, 1995).

The main benefits of BAF reactors over other secondary processes are the high potential removal rates per volume of media, high resistance to shock loadings and the production of high quality effluent (Lilly *et al.*, 1991; Rogalla and Sibony, 1995). Levels of up to 4.1 kg BOD removed m^{-3} d^{-1} have been reported with around 90% removal (Dillon and Thomas, 1990; Smith and Hardy, 1992). These values compare to 0.06 kg BOD removed m^{-3} d^{-1} h for trickling filters, 0.35 kg BOD removed m^{-3} d^{-1} h for oxidation ditches and 0.42 kg BOD removed m^{-3} d^{-1} h for activated sludge plants (Smith and

TABLE 3. Loadings Obtained During Carbonaceous Removal and Nitrification.

Treatment Type	Loading ($kg\ m^{-3}\ d^{-1}$)	Plant Size (m^3)	Reference
Carbonaceous	1.29 - 1.6	0.217	Faup *et al.*, 1982
Removal	< 15.4	0.048	Rusten and Thorvaldsen, 1983
(kg COD $m^{-3}\ d^{-1}$)	0.18 - 1.25	2.88	Iida and Teranishi, 1984
	0.034 -1.325	0.048	Rusten, 1984
	3.3 - 15.4	0.0085	Costa Reis and Sant'Anna, 1985
	< 3.25	Full Scale	Carrand *et al.*, 1990
	2.5 - 4.6 (BOD)	0.141	Dillon and Thomas, 1990
	0.7 - 5	0.14	Rogalla *et al.*, 1990
	< 10	Full Scale	Rogalla *et al.*, 1990
	< 15	0.2 - 0.3	Rogalla and Bourbigot, 1990
	8 - 10	0.019	Bacquet *et al.*, 1991
	3.7 - 4.1	0.005 - 0.01	Ros and Mejac, 1991
	2 - 9	0.002	Ryhiner *et al.*, 1992
	0.83 - 2.65 (BOD)	Full Scale	Smith and Hardy, 1992
	1.17 - 2.34 (BOD)	0.007	Clapp *et al.*, 1994
	4.8 - 8	Full Scale	Pujol *et al.*, 1994
	0.5 - 6.3	Full Scale	Pujol *et al.*, 1994
	< 14.6	0.091	Visvanathan and Nhien, 1995
Nitrification	1.08 - 1.27	0.217	Faup *et al.*, 1982
(kg NH_3-N $m^{-3}\ d^{-1}$)	0.03 - 0.14	2.88	Iida and Teranishi, 1984
	0.38 - 0.76	Full Scale	Rogalla and Payraudeau, 1988
	< 0.46	Full Scale	Carrand *et al.*, 1990
	0.46 - 1.05 (KjN)	0.141	Dillon and Thomas, 1990
	< 0.8	Full Scale	Paffoni *et al.*, 1990
	< 0.6	0.14	Rogalla *et al.*, 1990
	< 1	0.2 - 0.3	Rogalla and Bourbigot, 1990
	0.054 - 0.173	0.0374	Jimenez *et al.*, 1991
	0.1 - 0.8	8	Sagberg *et al.*, 1992
	0.19 - 0.35	Full Scale	Smith and Hardy, 1992
	< 0.1	0.01	Dee *et al.*, 1994
	0.12 - 0.535	0.739	Fruhen *et al.*, 1994
	< 0.85	Full Scale	Vedry *et al.*, 1994

Hardy, 1992; Pujol *et al.*, 1994). Much higher loadings of up to 8 kg BOD removed m^{-3} d^{-1} are possible though there is also a subsequent reduction in effluent quality and a drop in removal efficiency by 10 to 20% (Amar et al., 1986, Rogalla *et al.*, 1994). The reason for such high removal rates is the high biomass concentration available to treat the wastewater within BAFs (Rogalla *et al*, 1991; Smith and Hardy, 1992). The high achievable removal rates also allow short detention times of less than 1 h (Le Cloirec and Martin, 1984; Rogalla *et al.*, 1991: Rogalla and Sibony, 1992). Though shorter detention times and higher loadings can be achieved the resulting effluent quality produced may still remain high with less than 10 mg l^{-1} suspended solids possible (Rogalla and Sibony, 1992).

Fixed-film reactors have been found to perform better than suspended growth reactors during shock loadings of primary effluent (Young and Stewart, 1979). The plug flow in granular media reactors shock organic loadings are treated through gradual reduction in concentration. In structured media reactors, which are completely mixed, dilution of shock organic loads occurs but there is also an immediate increase in effluent substrate concentration (Rusten, 1984). High peak loadings though may also cause breakthrough due to limited treatment efficiency and acclimatisation at lower nutrient concentrations (Ruffer and Rosenwinkel, 1984). An additional benefit of structured and granular media plants is the production of high quality sludges. In granular media reactors, sludge produced showed good settlability with sludge yields of 0.63-1.06 kg SS per kg BOD (Amar *et al.*, 1986; Stensel *et al.*, 1988; Dillon and Thomas, 1990; Rogalla and Sibony, 1992). In structured media reactors the sludge produced was much lower at 0.15 to 0.25 kg SS per kg BOD (Rusten, 1984). Though structured and granular media reactors show many benefits over other secondary treatment methods there is one basic drawback. The amount of carbonaceous material removed is restricted due to the low oxygen transfer efficiency with a drop of up to 22%, limiting oxygen availability. Consequently media type and reactor height affect substrate removal through variation in air contact time (Rogalla and Sibony, 1992).

Nitrification

Generally submerged aerated biofilters perform well in nitrification with removal rates up to 1.27 kg NH_3-N m^{-3} d^{-1} achieved when used for tertiary treatment (Table 3) (Faup *et al.*, 1928; Saintpierre, 1988; Bourbigot and Lacamp, 1989; Pujol *et al.*, 1994; Fruhen *et al.*, 1994; Meaney and Strickland, 1994). Secondary treatment reactors used primarily to remove carbonaceous material have been found to treat ammonia loadings of up to 1 kg NH_3-N m^{-3} d^{-1} in conjunction with high COD loadings (Rogalla and Bourbigot, 1990). There are two methods of ammonia removal, assimilation and oxidation. In tertiary nitrification reactors the majority of ammonia is removed through cell synthesis (assimilation) and oxidation by autotrophic bacteria but in secondary reactors some ammonial nitrogen is required for the growth of heterotrophic bacteria (Akunna *et al.*, 1994). Oxidation is preferred as nitrogen may be removed in subsequent denitrification.

Assimilated nitrogen is bound within the biomass and may only be removed through backwashing (Jimenez *et al.*, 1987).

Other tertiary treatment processes are inclined to be highly influenced by variations in temperature during carbonaceous removal. Temperature though has only a slight influence on the nitrification rate in submerged bioreactors even at very low temperatures, though maximum activity occurs at 28-29 °C (Harremoes, 1982; Iida and Teranishi, 1984; Paffoni *et al.*, 1990; Fdz-Polanco *et al.*, 1994).

Inhibition of nitrification in submerged aerated bioreactors can take place in a number of ways; through competition by non nitrifying biological species, predation by higher organisms and through the influence of solids, substrate concentration and toxic substances (Oslislo and Lewandowski, 1985; Beg and Hassan, 1987; Rogalla and Payraudeau, 1988; Anderson, 1989; Lee and Welander, 1994). Biological competition is predominantly depend on the nutrient concentration of the wastewater to be treated (Boller *et al.*, 1994). Though some carbonaceous material, primarily carbon dioxide, is required for growth of autotrophic nitrifying bacteria, high concentrations of carbonaceous matter promote the growth of faster growing heterotrophic bacteria resulting in limited autotrophic growth and subsequent nitrification reduction (Faup *et al.*, 1982; Rogalla and Payaudreau, 1988; Al-Haddad *et al.*, 1991; Akunna *et al.*, 1994). High solids concentrations inhibit nitrification through adsorption onto the biofilm surface. This adsorption causes the growth of thick biofilms and promotes the growth of heterotrophic bacteria (Boller *et al.*, 1994). Thus in reactors treating both carbonaceous material and ammonia, nitrification efficiency is dependent on the amount of solids and carbonaceous material present. Nitrification efficiency may be high at low loadings, but as loadings are increased efficiency drops even when the C : N ratio is constant (Al-Haddad *et al.*, 1991). As well as high levels of carbonaceous material, low levels also inhibit nitrification by restricting the carbon source for biological growth of the nitrifying bacteria thus requiring the addition of an external carbon source (Sammut *et al.*, 1992; Furumai and Rittman, 1994). High level of ammonia promote the growth of nitrifiers but may result in a build up of nitrites without subsequent denitrification and very high levels inhibit *Nitrobacter* activity (Fdz-Polanco, 1994). This rapid nitrifier growth though also leads to a short start up time (Haug and McCarty, 1972; Harremoes, 1982; Hao and Chen, 1994). Oxygen concentration is a further factor that influences nitrification. The removal of carbonaceous material leads to a reduction in oxygen levels. High levels of oxygen are required for effective nitrification with levels greater than 4 mg l^{-1} of dissolved oxygen required, thus ammonia removal tends to be limited by oxygen availability but ammonia limitation may occur at a bulk oxygen : to bulk ammonia ratio of 2.5-4 to 1. This results in accumulation of nitrite (Boller *et al.*, 1994; Cecen and Gonenc, 1994; Chen and Chang, 1995).

Sludge production in nitrification is much lower than that produced in carbonaceous removal. Values of 0.13-0.32 kg SS per kg BOD have been measured (Iida and

Teranishi, 1984). Thus backwashing is required only infrequently and controls nitrification but also allows the redistribution of biomass within a bed (Fruhen *et al.*, 1994). This improves the efficiency of tertiary reactors especially during start-up when biomass predominantly grows at the inlet where ammonia concentration is at its highest (Abd-El-Bary, 1977; Hao and Chen, 1994).

Denitrification

The versatility of BAF reactors has allowed them to be used purely for denitrification simply by operating reactors without aeration (Jepsen and Jansen, 1993; Eichinger *et al.*, 1995; Janning *et al.*, 1995). The amount of denitrification in aerobic fixed film reactors depends on a number of conditions. The first of these conditions involves the positioning of the aerators within the reactors. If the aerators are placed part way up the reactor height, an anoxic zone is left below the aerators (Rogalla and Bourbigot, 1990). Generally anoxic denitrification zones are used in high performance reactors where there is high BOD and ammonia removal. The second condition where denitrification takes place is where aeration is limited or there is poor air distribution within reactors allowing anoxic pockets within the reactors to develop. This form of denitrification tends to occur predominantly in granular media reactors (Jimenez *et al.*, 1986).

Denitrification takes place following nitrification where the majority of carbonaceous material has already been removed, additional carbon is thus required to allow denitrification to take place often through the addition of secondary sewage or other sources such as methanol (Sagberg *et al.*, 1992; Carrand *et al.*, 1992; Hwang *et al.*, 1994; Zeghal *et al.*, 1995). Thus to obtain high denitrification rates of up to 5 kg NO_3-N m^{-3} are only available when an external carbon source is required in excess (Jepsen and Jansen, 1993; Pujol *et al.*, 1994). One method to solve the problem of carbon availability is to use a pre-denitrification reactor prior to the main carbonaceous / nitrifying reactor and employing recycle (Henze, 1995). Another alternative is by using recycle in granular media upflow reactors with the aerator placed part way up the media bed or without recycle in structured media beds (Dee *et al.*, 1994; Cecen and Gonenc, 1995; Smith and Marsh, 1995). In nitrifying reactors limiting oxygen availability encourages denitrification (Akunna *et al.*, 1994; Cecen and Gonenc, 1994). When recycle is used in granular media nitrification-denitrification reactors, the recycle ratio determines the nitrification efficiency (Jimenez *et al.*, 1986).

Phosphorus Removal

The most recent advances in designing submerged media aerated biofilters involve there use for phosphorus removal (Sagberg *et al.*, 1992; Goncalves *et al.*, 1994b). There are two methods of phosphate removal carried out in granular biological aerated filters. Firstly the use of flocculation with filtration and settlement and secondly biological uptake. Until the late 1980s phosphorus removal tended to be carried out using chemical precipitation, the use of suspended biomass or a combination of the two (Rogalla *et al.*, 1990a; Sammut *et al.*, 1992). The three dominating continuous or sidestream processes were the Bardenpho, A/O and PhoStrip processes (Gonzalez-Martinez and Wilderer, 1991). By the late 1980s work began to develop BAF type reactors for phosphorus removal.

Phosphorus flocculation and precipitation may be carried out at any point in the wastewater treatment process making this method more flexible than biological removal (Henze, 1995). The types of chemicals used for precipitation tend to be iron salts, aluminium salts or polyelectrolytes (Tanaka *et al.*, 1991; Henze, 1995). With biological aerated filters additional equipment is not required when flocculators are added prior either to the primary influent or primary effluent but large amounts are required making this method uneconomical. Less flocculator is required if used latter in treatment, though this may require additional filtration or settling equipment (Rogalla *et al.*, 1990). The use of coagulants and filtration in BAFs allows effluent phosphorus values to be reduced to 0.2 mg l^{-1} (Rogalla *et al.*, 1990a; Sammut *et al.*, 1992).

Fixed film reactors require less reagent addition than chemical precipitation processes and produce less sludge (Sammut *et al.*, 1992). In addition to this, BAFs have other benefits over suspended growth systems. In biological phosphorus removal the phosphorus is removed when it is taken in by the biomass. With biomass concentrations up to 5 times greater in fixed-film systems than suspended growth systems there is a greater potential in fixed media systems for phosphorus removal. Due to the parameters involved in phosphate removal, the use of such treatment is limited. The organisms used in phosphate removal are required to undergo both aerobic and anaerobic stages (Andersen et al, 1995). Thus fixed-film reactors are required to run under batch, semi-batch conditions or through the use of recycle (Gonzalez-Martinez and Wilderer, 1991; Goncalves and Rogalla, 1992; Goncalves *et al.*, 1994b). This is required to build up a high concentration of phosphorus within the biofilm to allow removal through subsequent backwashing (Sammut *et al.*, 1992; Goncalves *et al.*, 1994a)

The main parameters involved in determining phosphate removal efficiency are the times of the aerobic and anaerobic phases and the carbonaceous loading. In addition two possible inhibitors to phosphorus uptake are the availability of a degradable carbon source and excess nitrate concentrations. A biodegradable carbon source is simple to supply within a system but to avoid high nitrate levels, nitrification must be coupled with efficient denitrification or avoided all together prior to phosphorus removal (Henze, 1995).

APPLICATIONS

The primary use of submerged media biological aerated reactors is in the treatment of domestic wastewaters. The adaptability of such plants allows them to be used in a number of situations. Not only are new plants constructed in preference to other secondary and tertiary processes such as trickling filters but the small footprint of such plants promotes their use in upgrading established works (Sagberg *et al.*, 1992; Budge and Gorrie, 1996). There two reasons for the upgrading of established sewage treatment works. Rapid population growth and redistribution have caused existing works to be overloaded. Many of these works are in built up areas where there is limited space for expansion (Pujol *et al.*, 1992a; Lessel, 1994; Rogalla *et al.*, 1994; Brewer, 1996). Secondly recent discharge constraints have required high standards of effluent quality to be obtained (Robinson *et al.*, 1994). As well as the small footprint space required, the compact design and the ability to conceal BAF plants underground makes them ideal for environmentally sensitive sites such as coastal areas (Gilles, 1990; Rogalla *et al.*, 1990b; Horan, 1992; Pujol *et al.*, 1992b; Morris, 1993; Monro, 1995; Sampa and Tanaka, 1995). Not only have submerged media biological aerated reactors been used to treat a wide variety of domestic wastewater strengths, but the high effluent qualities achievable has allowed recycling on the water for domestic as well as industrial use (Adachi and Fuchu, 1991).

Less use of submerged aerated biological reactors for high strength industrial wastewater treatment has been made than for the treatment of domestic wastewater (Kleiber *et al.*, 1994). One of the primary industrial applications of such reactors is in the treatment of wastewaters contaminated by petroleum products and related substances such as phenols which tend to be toxic (Hamoda *et al.*, 1987; Bouwer *et al.*, 1992; Hamoda, 1993). Plants treating domestic waste have been found to treat wastewaters containing petroleum products (hydrocarbons) with concentrations of up to 20 mg l^{-1}, at efficiencies of up to 95%. These plants used beds of crushed media, 2 m in height (Kirichenko and Drushlyak, 1988; Drushlyak and Kirichenko, 1990). When levels of up to 1000 mg l^{-1} of hydrocarbon products were present, granular BAFs were found to perform well when compared with suspended growth reactors and fluidised bed reactors. The granular BAF had a greater capacity at removing hydrocarbon products than the suspended growth reactor but not as much as the fluidised bed reactor. The BAF reactor also showed better resistance to shock hydraulic and organic loads but showed a poorer aeration capability and oxygen utilisation (Holladay *et al.*, 1978). When structured media reactors were used to treat such wastes they showed poorer removal rates and less resistance to shock loadings but improved oxygen utilisation (Hamoda *et al.*, 1987,; Hamoda and Al-Haddad, 1987). Another major application for such reactors is for the treatment of food industry effluents which tend to be highly variable. These wastes come from dairies and the production of dairy products, meat packing plants and slaughterhouses, and contain high levels of BOD (up to 3000 mg l^{-1}), oil and grease (up to 500 mg l^{-1}) and protein (up to 1000 mg l^{-1}) (Rusten, 1983; Rundle, 1996). When compared with activated sludge processes, structured media reactors were showed to be superior in treatment especially on a volumetric basis. Up to 15.4 kg COD removed m^{-3} was removed in the structured media reactor compared to 4.16 kg COD removed m^{-3} in the activated sludge plant (Rusten and Thorvaldsen, 1983). Other applications include the treatment of paper mill effluents, lagoon effluents, rolling mill wastewaters, distillery wastewaters, landfill leachates and in wastewaters with high ammonia levels (Costa Reis and Sant'Anna, 1985: Baldwin and Ellis, 1987; Schlegel, 1988; Adachi and Fuchu, 1991, Duryea *et al.*, 1993; Rovel *et al.*, 1994; Isohata and Kawabata, 1994).

CONCLUSIONS

Over the last 15 to 20 years there has been a great deal of development into granular and structured media biological aerated reactors. These reactors have been shown to be comparable in performance to other secondary and tertiary methods and in many cases has superseded established processes (Stensel and Reiber, 1983). The main benefits of such reactors over other treatment methods include their small footprint size, high biomass concentration, their ability to treat high loading rates and resistance to shock loads (Smith *et at.*, 1990).

Though biological aerated reactors using structured and granular medias have replaced many other processes it is unlikely, at present, that BAF plants will become the main treatment method in all sewage treatment works. In areas such as small communities where space is not limited, trickling filters, which require only limited maintenance and low energy requirements have remained the preferred method of treatment. There is still plenty of potential in the design of structured and granular media reactors especially for use in areas of limited ground space, variable strength wastewaters, updating established plants and high strength wastewaters (Stensel *et al.*, 1988).

Though reactors have been designed for the removal of BOD, suspended solids, ammonia, nitrates, nitrites and phosphorus, there are still some aspects of BAF treatment that require further investigation and optimisation. Limited work has been carried out on the biofilm structure and kinetics though the performance of reactors has been shown to be virtually unaffected by low temperatures (Iida, 1984; Paffoni, 1990). Thus the development of an accurate empirical or mechanistic model is required that takes into account factors such as aeration rate, temperature, media type and nutrient concentration (Hamoda, 1989). Such a model would be a great benefit in the design of reactors.

At present the primary objectives in the development of reactors are in reducing costs and process optimisation (Schlegel, 1995). The two areas are specifically being investigated. These are backwashing requirements and aeration rates. Subsequently work is at present being carried out to reduce backwashing times and even eliminating it through the use of continual media recycling (Stephenson, 1996). Aeration is at present being optimised by the use of direct off-gas measurements and thus controlling the aeration rate through the measurement of oxygen utilised (Pearce, 1996). Design of biological aerated reactors has thus reached a point where the basic concept is well understood but fine tuning is required to obtain, what may hope to be, an almost ideal treatment process.

REFERENCES

Abd-El-Bary, M.F., Eways, M.J. (1977). Biological Nitrification in Contact Aeration Systems. *Water Sew. Works*, 91-93.

Adachi, S., Fuchu, Y. (1991). Reclamation and Reuse of Wastewater by Biological Aerated Filter Process. *Water Sci. Tech*, 24, 9, 195-204.

Akunna, J., Bizeau, C., Moletta, R., Bernet, N., Heduit, A. (1994). Combined Organic Carbon and Complete Nitrogen Removal Using Anaerobic and Aerobic Upflow Filters. *Water Sci. Tech.*, 30, 12, 297-306.

Albright and Wilson. (1964). Patent No. 971,338.

Al-Haddad, A.A., Zeidan, M.O., Hamoda, M.F. (1991). Nitrification in the Aerated Submerged Fixed-Film (ASFF) Bioreactor. *J. Biotech.*, 18, 115-128.

Altemeier, G., Gassen, M., Pfeffer, S. (1995). Abwasserreininigung mit Biofiltern in Herford. *Abwasser Technik*, 1, 42-45.

Amar, D., Partos, J., Granet, C., Faup, G.M., Audic, J.M. (1986). The Use of an Upflow Mixed Bed Reactor for Treatment of a Primary Settled Domestic Sewage. *Water Res.*, 20, 9-14.

Andersen, K.L., Bundgaard, E., Andersen, V.R., Hong, S.N., Heist, J.F. (1995). Nutrient Removal in Fixed-Film Systems. *Proc. Water Environ. Fed. 68th Annual Conf., Miami*, 581-590.

Anderson, B. (1989). Tentative Nitrogen Removal with Fixed Bed Processes in Malmo Sewage Treatment Plant.. *Technical Advances in Biofilm Reactors*, Nice, April, 265-276.

Arvin, E., Harremoes, P. (1989). Concepts and Models for Biofilm Reactors Performance. *Tech. Advances in Biofilm Reactors*, 4-6 April, Nice, France, 191-212.

Avaev, A.A. (1985). Modelling Mass Exchange in the Treatment of Fluid in a Biological Filter. *Soviet J. Water. Chem. Tech.*, 7, 7-9.

Bach, H. (1937). The 'Tank Filter' for the Purification of Sewage and Trade Wastes. *Water Works Sew.*, 389-393.

Bacquet, G., Joret, J.C., Rogalla, F., Bourbigot, M.M. (1991). Biofilm Start-Up and Control in Aerated Biofilter. *Environ. Technol.*, 12, 747-756.

Baldwin, W.C., Ellis, N.H. (1987). Packaged System Treats Landfill Leachate. *Pollut. Eng.*, Mar., 44-46.

Beg, S.A., Hassan, M.M. (1987). Effects of Inhibitors on Nitrification in a Packed-Bed Biological Flow Reactor. *Water Res.*, 21, 191-198.

Bishop, P.L., Rittman, B.E. (1995). Modelling Heterogeneity in Biofilms. Report of the Discussion Session. *Water Sci. Tech.*, 32, 8, 263-265.

Boller, M. (1987). Nutrient Removal from Wastewater. *Proc. 7th EWPCA Syposium,* Munich, 253-278.

Boller, M., Gujer, W., Tsuchi, M. (1994). Parameters Affecting Nitrifying Biofilm Reactors. *Water Sci. Tech.*, 19, 10-11, 1-11.

Bouwer, E.J., Chen, C.T., Li, Y.-H. (1992). Transformation of a Petroleum Mixture in Biofilms. *Water Sci. Tech.*, 26, 3-4, 637-646.

Brewer, P. (1996). Poole Wastewater Treatment BAF Plant. *Proc. 2nd BAF Symp., Cranfield, UK.*

Budge, F., Gorrie, D. (1996). Operational Trials of Different Proprietary Lamella and BAF Systems. *Proc. 2nd BAF Symp., Cranfield, UK.*

Buswell, A.M., Pearson, E.L. (1929). The Nidus (Nest) Rack, a Modern Development of the Travis Colloider. *Sew. Works Jour.*, 1, 187-195.

Canler, J.P., Perret, J.M. (1994). Biological Aerated Filters: Assessment of the Process Based on 12 Sewage Treatment Plants. *Water Sci. Tech.*, 29, 10-11, 13-22.

Cao, Y.S. Alaerts, G.J. (1995). Influence of Reactor Type and Shear Stress on Aerobic Biofilm Morphology, Population and Kinetics. *Water Res.*, 29, 1, 107-118.

Cantwell, A.D.C. (1996). Operating Performance and Future Development of the Biobead System. *Proc. 2nd BAF Symp., Cranfield, UK.*

Canziani, R. (1988). Submerged Aerated Biofilters. IV - Aeration Characteristics. *Ingegneria Ambientale,* 17, 627-636.

Capdeville, B., Nguyen, K.M. (1990). Kinetics and Modelling of Aerobic and Anaerobic Film Growth. *Water Sci. Tech.*, 22, 1/2, 149-170.

Carrand, G., Capon, B., Rasconi, A., Brenner, R. (1990). Elimination of Carbonaceous and Nitrogenous Pollutants by a Twin Stage Fixed Growth Process. *Water Sci. Tech.*, 22, 1/2, 261-272.

Carrio, L.A., Zhang, J., Pujnn, J., Mitchell, A. (1995). Biological Aerated Filtration; New York City's Experience. *Proc. Water Environ. Fed. 68th Annual Conf., Miami, 805-811.*

Cecen, F., Gonenc, I.E. (1994). Nitrogen Removal Characteristics of Nitrification and Denitrification Filters. *Water Sci. Tech.*, 29, 10-11, 409-416.

Cecen, F., Gonenc, I.E. (1995). Criteria for Nitrification and Denitrification of High-Strength Wastes in Two Upflow Submerged Filters. *Water Environ. Res.*, 67, 132-142.

Chen, S., Cheng, S. (1994). The Enhancement of Nitrification by Indirect Aeration and Kinetic Control in a Submerged Biofilm Reactor. *Water Sci. Tech.*, 30, 11, 131-142.

Christensen, F.R., Holm Kristensen, G., La Cour Jansen, J. (1989). Biofilm Structure - An Important and Neglected Parameter in Waste Water Treatment. *Water Sci. Tech.*, 21, 805-814.

Christiansen, P., Hollesen, L., Harremoes, P. (1994). Liquid Film Diffusion on Reaction Rate in Submerged Biofilters. *Water Res.*, 29, 947-952.

Chudoba, J., Strakova, P., Kondo, M. (1991). Compartmentalised versus Completely-Mixed Biological Wastewater Treatment Systems. *Water Res.*, 25, 973-978.

Churchley, J., Rundle, H., Upton, J. (1995). The Use of BAF Technology to Update UK Sewage Treatment Works Performance. *Proc. Water Environ. Fed. 68th Annual Conf., Miami, 795-803*

Clapp, L.W., Talarczyk, M.R., Park, J.K., Boyle, W.C. (1994). Performance Comparison Between Activated Sludge and Fixed-Film Processes for Priority Pollutant Removals. *Water. Environ. Res.*, 66, 153-160.

Clark, H.W. (1930). Past and Present Developments in Sewage Disposal and Purification. *Sew. Works Jour.*, 2, 561-571.

Clarke, S. (1996). Recent Experience and Developments Using the Biopur BAF Process. *Proc. 2nd BAF Symp., Cranfield, UK.*

Condren, A.J. (1990). Technology Assessment of the Biological Aerated Filter. US EPA/600/52-90/015-8P.

Costa Reis, L.G., Sant'Anna, G.L. (1985). Aerobic Treatment of Concentrated Wastewater in a Submerged Bed Reactor. *Water Res.*, 19, 1341-1345.

Dauthuille, P. (1990). Le Biofor. *Tribune de l'Eau*, 546-547, 65-68.

De Beer, D., Muyzer, G. (1995). Multispecies Biofilms : Report from the Discussion Session. *Water Sci. Tech.*, 32, 8, 269-270.

Dee, A., James, N., Jones, I., Strickland, J., Upton., Cooper, P. (1994). Pre- or Post-Denitrification at Biological Filter Works ? A Case Study. *Water Sci. Tech.*, 29, 10-11, 145-155.

Desbos, g., Rogalla, F., Sibony, J., Bourbigot, M.M. (1990). Biofiltration as a Compact Technique for Small Wastewater Treatment Plants. *Water Sci. Tech.* 22, 3/4, 145-152.

Dillon, G.R., Thomas, V.K. (1990). A Pilot-Scale Evaluation of the 'Biocarbone Process' for the Treatment of Settled Sewage and for Tertiary Nitrification of Secondary Effluent. *Water Sci. Tech.*, 22, 1/2, 305-316.

Drushlyak, O.S., Kirichenko, A.G. (1990). Final Treatment of Effluents with Petroleum Products on Aerated Filters. *Khimiya i Tekhnologiya Vody*, 12, 1121-1122.

Dumbleton, B. (1992a). Flood of New Ideas in Fluid Market. *Const. Weekly*, Aug., 23-28.

Dumbleton, B. (1992b). UK Water Industry Invests 28,000M Pounds Sterling to Reach Environmental Goals. *Water Waste Int.*, 7, 21-25.

Duryea, D.J., Caroll, J.E., Sangrey, K.H. (1993). Nitrification of Lagoon Effluent Utilising the ColOX Biofiltration Process. Nitrification Utilising Biological Aerated Filters. *New England Water. Pol. Cont. Ass.* 1993 Annual Conf.

Eichinger, J., Kramer, P., Wilson, T.E. (1995). Denitrification in Deep Bed Filters - A Comparison of Filter Media. *Proc. Water Environ. Fed. 68th Annual Conf., Miami*, 777-782.

Fair, G.M., Fuhrman, R.E., Ruhhoff, C.C., Thomas, H.A., Mohlman, F.W. (1948). Sewage Treatment at Military Installations - Summary and Conclusions. *Sew. Works Jour.*, 20, 52-95.

Fair, G.M., Thomas, H.A. (1950). The Concept of Interface and Loading in Submerged Aerobic Biological Sewage Treatment Systems. *Sew. Pur.* 235-247.

Faup, G.-M., Leprince, A., Pannier, M. (1982). Biological Nitrification in an Upflow Fixed Bed Reactor (UFBR). *Water Sci. Tech.,* 14, 795-810.

Fdez-Polanco, F., Villaverde, S., Garcia, P.A. (1994). Temperature Effect on Nitrifying Bacteria Activity in Biofilters : Activation and Free Ammonia Inhibition. *Water Sci. Tech.*, 30, 11, 121-130.

Fruhen, M., Bocker, K., Eidens, S., Haaf, D., Liebeskind, M., Schmidt, F. (1994). Tertiary Nitrification in Pilot-Plant Plug-Flow Fixed Film Reactors with Long Term Ammonium Deficiency. *Water Sci. Tech.*, 29, 10-11, 61-67.

Fuije, K., Hu, H.Y., Tanaka, H., Urano, K. (1994). Ecological Studies of Aerobic Submerged Biofilters on the Basis of Respiratory Quinone Profiles. *Water Sci. Tech.*, 29, 7, 373-376.

Furumai, H., Rittman, B.E. (1994). Interpretation of Bacterial Activities in Nitrification Filters by a Biofilm Model Considering the Kinetics of Soluble Microbial Products. *Water Sci. Tech.*, 30, 11, 147-156.

Gilles, P. (1990). Industrial Scale Applications of Fixed Biomass on the Mediterranean Seaboard. Design, Operating Results. *Water Sci. Tech.*, 22, 1/2, 281-292.

Giuliano, C., Joret, J.C. (1988). Distribution, Characterisation and Activity of Microbial Biomass of an Aerobic Fixed-Bed Reactor. *Water Sci. Tech.*, 20, 11/12, 455-457.

Goncalves, R.F., Rogalla, F. (1992). Continuous Biological Phosphorus Removal in a Biofilm Reactor. *Water Sci. Tech.*, 26, 9-11, 2027-2030.

Goncalves, R.F., Le Grand, L., Rogalla, F. (1994a). Biological Phosphorus Uptake in Submerged Biofilters with Nitrogen Removal. *Water Sci. Tech.*, 29, 10-11, 135-143.

Goncalves, R.F., Nogueira, F.N., Le Grand, L. Rogalla, F. (1994b). Nitrogen and Biological Phosphorus Removal in Submerged Biofilters. *Water Sci. Tech.,* 30, 11, 1-12.

Gonzalez-Martinez, S., Wilderer, P.A. (1991). Phosphate Removal in a Biofilm Reactor. *Water. Sci. Tech.*, 23, 1405-1415.

Gonzalez-Martinez, S., Duque-Luciano, J. (1992). Aerobic Submerged Biofilm Reactors for Wastewater Treatment. *Water Res.* 26, 825-833.

Grady, L. (1983). Modelling of Biological Fixed Films - A State-of-the-Art Review. *Fixed-Film Biological Processes for Wastewater Treatment.* Wu, Y.C, Smith, E.D. (Editors), New Jersey, USA, 75-134.

Grasmick, A., Elmalen, S., Ben-Aim, R. (1980). Theory of Treatment by Submerged Biological Filtration. *Water Res.,* 13, 1137-1147.

Grasmick, A., Elmaleh, S., Yahi, H. (1984). Nitrification by Attached Cell Reactors Aerated at Co- or Counter-Current. Experimental Data and Modelling. *Water Res.*, 18, 885-891.

Gray, T.W. (1993). Biological Aerated Filters: The Solution to Small Footprint Plant. *Int. Water. Environ. Eng.*, Sum., 5-10

Griffith, L.B. (1943). Contact Aeration for Sewage Treatment. *Eng. News Rec.*, 28, 60-64

Gros, H. (1991). BIOPUR Systems and Bioactive Filters for Advanced Waste-Water Treatment. *Sulzer Technical Review*, 4/1991.

Hamoda, M.F. Al-Haddad, A.A. (1987). Investigation of Petroleum Refinery Effluent Treatment in an Aerobic Fixed-Film Biological System. *J. IWEM.*, 1, 239-246.

Hamoda, M.F., Al-Haddad, A.A., Abd-El-Bary, M.F. (1987). Treatment of Phenolic Wastes in an Aerated Submerged Fixed-Film (ASFF) Bioreactor. *J. Biotech.*, 5, 279-292.

Hamoda, M.F., Abd-El-Bary, M.F. (1987). Operating Characteristics of the Aerated Submerged Fixed-Film (ASFF) Bioreactor. *Water Res.*, 21, 939-947.

Hamoda, M.F. (1989). Kinetic Analysis of Aerated Submerged Fixed-Film (ASFF) Bioreactors. *Water Res.* 23, 1147-1154.

Hamoda, M.F. (1993). Behaviour of a Biological Fixed-Film System Treating Toxic and Non-Toxic Organic Wastes. *Water Sci. Tech.*, 28, 2, 155-163.

Hao, O.J., Chen, J.M. (1994). Factors Effecting Nitrite Build-Up in Submerged Filter System. *J. Environ. Eng.*, 120, 1298-1307.

Harremoes, P. (1977). Half-Order Reactions in Biofilm and Filter Kinetics. *Vatten,* 33, 122-143.

Harremoes, P. (1982). Criteria for Nitrification in Fixed Film Reactors. *Water Sci. Tech.*, 14, 167-187.

Haug, R.T., McCarty, P.L. (1972). Nitrification with Submerged Filters. *J.WPCF.*, 44, 2086-2102.

Henze, M. (1995) Nutrient Removal from Wastewater. *New World Water 1995*, 114-117.

Hoehn, R.C., Ray, A.D. (1973). Effects of Thickness on Bacterial Film. *J.WPCF.*, 45, 11, 2302-2320.

Holladay, D.W., Hancher, C.W., Scott, C.D., Chilcote, D.D. (1978). Biodegredation of Phenolic Waste Liquors in Stirred-Tank, Packed-Bed and Fluidised-Bed Reactors. *J.WPCF.*, 50, 2573-2589.

Horan, N. (1992). Reducing Pollution by Nutrient Removal. *Water Waste Treat.*, 35, 176.

Hwang, Y., Sakuma, h., Tanaka, T. (1994). Denitrification with Isopropanol as a Carbon Source in a Biofilm System. *Water Sci. Tech.*, 30, 11, 69-78.

Iida, Y., Teranishi, A. (1984). Nitrogen Removal from Municipal Wastewater by a Single Submerged Filter. *J.WPCF.*, 56, 251-258.

Isohata, Y., Kawabata, M. (1994). Development and Field Experience of Anticontact Packed-Bed Reactor for Wastewater Treatment. *Proc. 2nd Int. Conf. Fixed-Film Bio. Proc.*, Arlington, VA, 989-1013.

Iwai, S., Ohmori, H., Tanaka, T. (1977). An Advanced Sewage Treatment Process-Combination of Aerated Submerged Biological Filtration and Ultra-Filtration with Pulverised Activated Carbon. *Desalination,* 23, 29-36.

Janning, K.F., Harremoes, P., Nielsen, M. (1995). Evaluation and Modelling the Kinetics in a Full Scale Submerged Denitrification Filter. *Water Sci. Tech.*, 32, 8, 115-123.

Jennings, P.A., Snoeyink, V.L., Chian, E.V.K. (1976). Theoretical Model for a Submerged Biological Filter. *Biotech. Bioeng.*, 18, 1249-1273.

Jepsen, S.E., La Cour Jansen, J. (1993). Biological Filters for Post-Denitrification. *Water Sci. Tech.*, 27, 5-6, 369-379.

Jimenez, B., Capdeville, B., Roques, H., Faup, G.M. (1987). Design Considerations for a Nitrification-Denitrification Process Using Two Fixed-Bed Reactors in Series. *Water Sci. Tech.*, 19, 139-150.

Kirichenko, A.G., Drushlyak, O.G. (1988). Removal of Petroleum Products, SSAS and Biogenic Pollutants During Tertiary Treatment of Waste-Water in Aerated Filters. *Khimiya i Tekhnologiya Vody*, 10, 151-154.

Kleiber, B., Roudon, G., Bigot, B., Sibony, J. (1994). Assessment of Aerated Biofiltration at Industrial Scale. *Water Sci. Tech.*, 29, 10-11, 197-208.

Koutsakos, E., Smith, A.J., Brignal, W.J. (1992). Temperature Effects on the Performance of a Submerged Aerated Filter Process. *Proc. 15th Inter. Sym. Wastewater Treat.*, Montreal, 227-236.

Kurbiel, J. (1987). Upflow Contact Filtration of Biologically Treated Wastewater for Water Reclamation. *Tribune Cebedeau*, 520, 33-37.

Kuribayashi, S. (1992). Experimental Studies on Advanced Treatment of Wastewater for Amenity Use. *Water Sci. Tech.* 26, 9-11, 2401-2404.

Lacamp, B., Bourbigot, M.-M. (1989). Advanced Nitrogen Removal Processes for Drinking and Wastewater Treatment. *Proc. IAWPRC Conf.*, Guildford, UK, 437-443.

Lackey, J.B., Dixon, R.M. (1943). Some Aspects of the Hays Process of Sewage Treatment. *Sew. Works Jour.*, 15, 1139-1152.

Lazarova, V., Manem, J. (1995). Biofilm Characterisation and Activity Analysis in Water and Wastewater Treatment. *Water Res.*, 29, 2227-2245.

Le Cloirec, P., Martin, G. (1984). The Mean Residential Time Application in an Aerated Immersed Biological Filter. *Environ. Technol. Let.*, 5, 275-282.

Lee, K.M., Stensel, H.D. (1986). Aeration and Substrate Utilisation in a Sparged Packed-Bed Biofilm Reactor. *J.WPCF.*, 58, 1066-1072.

Lee, N.M., Welander, T. (1994). Influence of Predators on Nitrification in Aerobic Biofilm Processes. *Water Sci. Tech.*, 29, 7, 355-363.

Legise, J.P., Gilles, P., Moreaaud, H. (1980). A New Technology Development : The Biocarbone Aerated Filter. Proc. 53rd WPCF Conf., Las Vegas.

Lessel, T.H. (1994). Upgrading and Nitrification by Submerged Biofilm Reactors - Experiences from Large Scale Plants. *Water. Sci. Tech.*, 29, 10-11, 167-174.

Lewandowski, Z., Walser, G. (1991). Influence of Hydrodynamics on Biofilm Accumulation. *Proc. Technical Advances in Biofilm Reactors*, Nice, April, 619-624.

Lewandowski, Z., Stoodley, P., Altobelli, S., Fukushima, E. (1994). Hydrodynamics and Kinetics in Biofilm Systems - Recent Advances and New Problems. *Water Sci. Tech.*, 29, 10-11, 223-229.

Lewandowski, Z., Stoodley, P., Altobelli, S. (1995). Experimental and Conceptual Studies on Mass Transport in Biofilms. *Water Sci. Tech.*, 31, 1, 153-162.

Lilly, W., Bourn, G., Crabtree, H., Upton, J., Thomas, V. (1991). The Production of High Quality Effluents in Sewage Treatment Using the Biocarbone Process. *J. IWEM.*, 5, 123-133.

Mann, A.T., Fitzpatrick, C.S.B., Stephenson, T. (1995). A Comparison of Floating and Sunken Media Biological Aerated Filters Using Tracer Study Techniques. *Trans. IChemE. Pt. B*, 73, 137-143.

Mann A.T., Stephenson, T. (1996) Modelling Biological Aerated Filters for Wastewater Treatment. *Proc. 2nd BAF Symp., Cranfield, UK.*

McCarty, P.L., Haug, R.T. (1971). Nitrogen Removal from Wastewaters by Biological Nitrification and Denitrification. *Microbial Aspects of Pollution*, Sykes, G., Skinner, F.A. (Editors), Sym. No. 1, Society for Applied Bacteriology.

McHarness, D.D., Haug, R.T., McCarty, P.L. (1975). Field Studies of Nitrification with Submerged Filters. *J. WPCF.*, 47, 291-309.

Meaney, B.J., Strickland, J.E.T. (1994). Operating Experiences with Submerged Filters for Nitrification and Denitrification. *Water Sci. Tech.*, 29, 10-11, 119-125.

Mesdaghinia, A. (1986). Fixed Activated Sludge Makes Sewage Treatment Simple. *Water Sci. Tech.*, 18, 193-198.

Meunier, A.D., Williamson, K.L. (1981). Packed Bed Biofilm Reactors : Simplified Model. *J. Environ. Eng. Div. Proc. Am. Soc. Civ. Eng.*, 107, 1727-1735.

Ministry of Technology. (1968). Water Pollution Research. *HMSO*, London, UK, 100-103.

Monro, M. (1995), Cramped Style. *Water Bul.*, 671, 12-15.

Moreau, M. , Liu, Y., Capdeville, B., Audie, J.M., Calvez, L. (1994). Kinetic Behaviour of Heterotrophic and Autotrophic Biofilms in Wastewater Treatment Processes. *Water Sci. Tech.*, 29, 10-11, 385-391.

Morris, J. (1993). A Space Saving Innovation. *Water. Bul.*, 543, 8-9.

Mushu, Y. (1990). Use of Dispersed Flow Models in Design of Biofilm Reactors. *Water Air Soil Pollut.*, 53, 297-314.

Newbigging, M.L., Stephenson, J.P., Romano, L.S. (1995). Upflow or Downflow BAFs - Which Provides the Best Overall Performance. *Proc. Water Environ. Fed. 68th Annual Conf., Miami*, 783-794.

Nielsen, P.H., Harremoes, P. (1995). Solids : Report of the Discussion Session. *Water Sci. Tech.*, 32, 8, 273-275.

Oslislo, A., Lewandowski, Z. (1985). Inhibition of Nitrification in Packed Bed Reactors by Selected Organic Compounds. *Water Res.*, 19, 423-426.

Paffoni, C., Gousailles, M., Rogalla, F., Gilles, P. (1990). Aerated Biofilters for Nitrification and Effluent Polishing. *Water Sci. Tech.*, 22, 7/8, 181-189.

Park, J.W., Ganczarczyk, J.J., (1994). Gravity Separation of Biomass Washed Out from an Aerated Submerged Filter. *Environ. Tech.*, 22, 7/8, 181-189.

Pearce, P. (1996). Aeration Optimisation of Biological Aerated Filters. *Proc. 2nd BAF Symp., Cranfield, UK.*

Peng, J., Stevens, D.K., Yiang, X. (1995). A Pioneer Project of Wastewater Reuse in China. *Water Res.*, 29, 357-363.

Pujol, R., Canler, J.P., Iwema, A. (1992a). Biological Aerated Filters : An Attractive and Alternative Biological Process. *Water Sci. Tech.*, 26, 3-4, 693-702.

Pujol, R., Canler, J.P., Vachon, A., Vidou, P. (1992b). Biological Aerated Filters : An Adapted Biological Process for Wastewater from Coastal Areas. *Water Sci. Tech.*, 25, 12, 175-184.

Pujol, R., Hamon, M., Kandel, X., Lemmel, H. (1994). Biofilters : Flexible, Reliable Biological Reactors. *Water Sci. Tech.*, 29, 10-11, 33-38.

Quek, S.T., Ang, K.K., Ong, S.L., Tan, J.G. (1995). Reliability of Domestic-Waste Biofilm Reactors. *J. Environ. Eng.*, Nov., 785-790.

Quickenden, J., Mittal, R., Gros, H. (1992). Effluent Nutrient Removal with Sulzer Biopur and Filtration Systems. *Europ. Conf. Nut. Rem. from Wastewater,* Sept., Wakefield.

Reiber, S.H., Stensel, H.D. (1985). Oxygen Transfer in Fixed Film Systems. J. WPCF., 57, 135-140.

Riemer, M., Kristensen, G.H., Harremoes, P. (1980). Residence Time Distribution in Submerged Biofilters. *Water Res.*, 14, 949-958.

Rittman, B.E., McCarty, P.L. (1980). Design of Fixed-Film Processes with Steady-State Biofilm Model. *Prog. Water Tech.*, 12, 271-281.

Robinson, A.B., Brignal, W.J., Smith, A.J. (1994). Construction and Operation of a Submerged Aerated filter Sewage Treatment Works. *J. IWEM.*, 8, 215-227.

Rogalla, F, Payaudeau, M. (1988). Tertiary Nitrification with Fixed Biomass Reactors. *Water Supply*, 6, 347-354.

Rogalla, F, Payaudeau, M., Bacquet, G., Bourbigot, M.-M., Sibony, J., Gilles, P. (1990a). Nitrification and Phosphorus Precipitation with Biological Aerated Filters. *Res. J. WPCF.*, 62, 169-176.

Rogalla, F., Townshend, A., Sibony, J. (1990b). Innovative Wastewater Treatment Technology for Environmentally Sensitive Sites. IWEM Conf 1990, 45-1 - 45-11.

Rogalla, F., Bourbigot, M.-M. (1990). New Developments in Complete Nitrogen Removal with Biological Aerated Filters. *Water Sci. Tech.*, 22, 1/2, 273-280

Rogalla, F, Payaudeau, M., Sauvegrain, P., Sibony, J. (1991). Reduced Hydraulic Detention Time for Complete Nutrient Removal with Innovative Biological Reactors. *Water Sci. Tech.*, 24, 10, 217-229.

Rogalla, F., Sibony, J. (1992). Biocarbone Aerated Filters - Ten Years After : Past, Present and Plenty of Potential. *Water Sci. Tech.*, 26, 9-11, 2043-2048.

Rogalla, F., Lamouche, A., Specht, W., Kleiber, B. (1994). High Rate Aerated Biofilters for Plant Upgrading. *Water Sci. Tech.*, 29, 12, 207-216.

Rogalla, F., Sibony, J., Lacamp, B., Hansen, F. (1992). Aerated Biofilters - Recent European Examples. *IWEM Scientific Section Symposium in Advanced Sewage Treatment,* London, 33-53

Ros, M., Mejac, B. (1991). Treatment of Wastewater in an Upflow Packed-Bed Reactor. *Water Sci. Tech.* 24, 7, 81-88.

Rovel, J.M., Trudel, J.P., Lavallee, P., Schroeter, I. (1994). Paper Mill Effluent Treatment Using Biofiltration. *Water Sci. Tech.*, 29, 10-11, 217-222.

Ruffer, H., Rosenwinkel, K.H. (1984). The Use of Biofiltration for Further Wastewater Treatment. *Water Sci. Tech.*, 16, 241-260.

Rundle, H. (1996). Experiences with Biological Aerated Filters for Treatment of Settled Sewage and Dairy Effluent. *Proc. 2nd BAF Symp., Cranfield, UK.*

Rusten, B., Thorvaldsen, G. (1983). Treatment of Food Industrial Effluents - Activated Sludge Versus Aerated Submerged Biological Filters. E*nviron. Technol. Let.*, 4, 441-450.

Rusten, B. (1984). Wastewater Treatment with Aerated Submerged Biological Filters. *J.WPCF.*, 56, 424-431.

Rusten, B., Odegaard, H. (1986). Treatment of Food Industry Effluents in Aerated Submerged Biological Filters. *Vatten*, 42, 187-193.

Ryhiner, G., Birou, B., Gros, H. (1992). The Use of Submerged Structured Packings in Biofilm Reactors for Wastewater Treatment. *Water. Sci Tech.*, 26, 3-4, 723-731.

Ryhiner, G., Sorensen, K., Birou, B., Gros, H. (1994). Biofilm Reactors Configuration for Advanced Nutrient Removal. *Water Sci. Tech.*, 29, 10-11, 111-117.

Sagberg, P., Berg, K.G. (1996). Experiences with Biofor Reactors at VEAS, Norway. *Proc. 2nd BAF Symp., Cranfield, UK.*

Sagberg, P., Dauthuille, P., Hamon, M. (1992). Biofilm Reactors : A Compact Solution for the Upgrading of Waste Water Treatment Plants. *Water Sci. Tech.*, 26, 3-4, 33-742.

Saintpierre, O. (1988). Tertiary Nitrifying Immersed Biofilter with Plastic Media. *Environ. Technol. Let.*, 9, 1059-1072.

Sakuma, H., Tanaka, T., Maki, Y. (1993). Studies on Nitrification and COD removal Using Biological Aerated Filters. *Proc. 6th World Filt. Cong.*, Nagoya, Japan, 743-746.

Sammut, F., Rogalla, F., Goncalves, R.F., Penillard, P. (1992). Practical Experiences with Removing Nitrogen and Phosphorus on Aerated Biofilters. *Europ. Conf. Nut. Rem. from Wastewater,* Sept, Wakefield.

Sampa H., Tanaka, T. (1995). Pilot-Plant Study of a New Wastewater Treatment System. *J.CIWEM.*, 9, 564-572.

Schlegel, S. (1988). The Use of Submerged Biological Filters for Nitrification. *Water Sci. Tech.*, 20, 4/5, 177-187.

Schlegel, S. (1995). Saving Costs with Fixed-Bed Processes ?. *Korrespondenz Abwasser*, 42, 1343-1352.

Smith, A.J., Brignal, W.B. (1996). Trouble Shooting and Optimisation of BAF Systems. *Proc. 2nd BAF Symp., Cranfield, UK.*
Smith, A.J., Marsh, P. (1995). Enhanced Wastewater Treatment with Lamella Tube Settlers, Submerged Aerated Filters and Ultra-Violet Radiation. *Proc. Water Environ. Fed. 68th Annual Conf., Miami,* 627-637.
Smith, A.J., Quinn, J.J., Hardy, P.J. (1990). The Development of an Aerated Filter Package Plant. *Proc. 1st Int. Conf. Advances in Water Treat. Environ. Manag.*, 27-29 June, Lyon, France.
Smith, A.J., Hardy, P.J. (1992). High-Rate Sewage Treatment Using Biological Aerated Filters. *J. IWEM.*, 6, 179-193.
Stensel, H.D., Reiber, S. (1983). Industrial Wastewater Treatment with a New Biological Fixed-Film System. *Environ. Prog.*, 2, 110-115.
Stensel, H.D., Brenner, R.C., Lee, K.M., Melcer, H., Rakness, K. (1988). Biological Aerated Filter Evaluation. *J. Environ. Eng.*, 114, 655-671.
Stephenson, T. (1996). Development of the Recirculating Plastic Media Biological Aerated Filter (REBAF). *Proc. 2nd BAF Symp., Cranfield, UK.*
Stephenson, T., Mann, A. (1995). Biological Aerated Filters - Here to Stay. *Water Waste Treat.,* 38, 25-26.
Stephenson, T., Mann, A., Upton, J. (1993). The Small Footprint Wastewater Treatment Process. *Chem. Ind.*, 533-536.
Sundararajan, A., Lu, L.-K. (1995). Biological Oxygen Transfer Enhancement in Wastewater Treatment Systems. *Water Environ. Res.*, 67, 848-854.
Takamizawa, K., Fukunaga, I., Inoue, Z. (1993). Promotion of Nitrification and Denitrification by Recirculation of Effluent and Biofilm Process. *Environ. Tech.*, 14, 981-987.
Tanaka, K., Aoki, M., Yakahashi, S., Chida, S., Yasuda, T., Takagi, K., Yumoto, H., Kasakura, K. (1991). Study on Development of Phosphorus Removal Process by Contact Filtration. *Water Sci. Tech.*, 23, 739-745.
Toettrup, H., Rogalla, F., Vidal, A., Harremoes, P. (1994). The Treatment Trilogy of Floating Filters : From Pilot to Prototype to Plant. *Water Sci. Tech.*, 29, 10-11, 23-32.
Tchobanoglous, G., Schroeder, E.D. (1985) Water Quality : Characteristics, Modelling, Modification. Addison-Wesley, Reading MA.
Tschui, M., Boller, M., Gujer, W., Eugster, J., Mader, C., Stengel, C. (1994). Tertiary Nitrification in Aerated Pilot Biofilters. *Water Sci. Tech.*, 29, 10-11, 53-60.
Vedry, B., Paffoni, C., Gousailles, M., Bernard, C. (1994). First Months Operation of Two Biofilter Prototypes in the Waste Water Plant of Acheres. *Water Sci. Tech.*, 29, 10-11, 39-46.
Visvanathan, C., Nhien, T.T.H. (1995). Study on Aerated Biofilter Process Under High Temperature Conditions. *Environ. Technol.*, 16, 301-314.
Wheale, G., Cooper-Smith, G.D. (1992). The Use of Biocarbone for Advanced Sewage Treatment. *IWEM Scientific Section Symposium in Advanced Sewage Treatment*, London, Nov., 57-75.

Wheale, G., Cooper-Smith, G.D. (1995). Operational Experience with Biological Aerated Filters. *J. Chart. Inst. Water Environ. Manag.*, 9, 109-118.

Wilderer, P.A., Cunningham, A., Schlindler, U. (1995). Hydrodynamics and Shear Stress : Report from the Discussion Session. *Water. Sci. Tech.*, 32, 8, 271-272.

Wilford, J., Conlon, T.P. (1957). Contact Aeration Sewage Treatment Plants in New Jersey. *Sew. Ind. Wastes,* 29, 845-855.

Williamson, K., McCarty, P. (1976). Verification Studies of the Biofilm Model for Bacterial Substrate Utilisation. *J. WPCF.*, 48, 281-296.

Young, J.C., Baumann, E.R., Wall, D.J. (1975). Packed Bed Reactors for Secondary Effluent BOD and Ammonia Removal. *J.WPCF.*, 47, 46-56.

Young, J.C., Stewart, M.C. (1979). PBR - A New Addition to the AWT Family. *Water Wastes Eng.*, 20-25.

Zeghal, S., Nagem Nogueira, F., Salzer, C., Rogalla, F. (1995). Tertiary Denitrification Using an Upflow Floating Biofilter. *Proc. Water. Environ. Fed. 68th Annual Conf., Miami, 813-821.*

Zhang, T.C., Bishop, P.L. (1994a). Competition in Biofilms. *Water Sci. Tech.*, 29, 10-11, 263-270.

Zhang, T.C., Bishop, P.L. (1994b). Experimental Determination of the Dissolved Oxygen Boundary Layer and Mass Transfer Resistance Near the Fluid-Biofilm Interface. *Water Sci. Tech.*, 30, 47-59.

Zinebi, S., Henriette, C., Petitdemange, E., Joret, J.C. (1994). Identification and Characterisation of Bacterial Activities Involved in Wastewater Treatment by Aerobic Fixed-Bed Reactor. *Water Res.*, 28, 2575-2582.

Chapter 3.

A COMPARISON OF FLOATING AND SUNKEN MEDIA BIOLOGICAL AERATED FILTERS USING TRACER STUDY TECHNIQUES.

Published,

Transactions of the IChemE Pt B, May 1995, 137-143.

A COMPARISON OF FLOATING AND SUNKEN MEDIA BIOLOGICAL AERATED FILTERS USING TRACER STUDY TECHNIQUES.

Allan Mann, Caroline Fitzpatrick and Tom Stephenson.

School of Water Sciences, Cranfield University, Cranfield, Bedford. MK43 0AL, U.K.

ABSTRACT

Three phase fixed media wastewater treatment reactors, known as biological aerated filters (BAFs), combine aerobic treatment with biomass separation. Tracer studies have been undertaken using media both less dense and denser than water, with two aerator configurations. The reactors acted as near ideal plug flow systems without aeration. When aerated, channelling and mixing were increased. Variation in aerator design appeared to have almost no effect on flow through the bed. From the results, it is predicted that floating media will be best for high solids wastewater and sunken media for low solids wastewater.

KEYWORDS

Biological aerated filters; tracer studies; floating media; sunken media; mixing.

INTRODUCTION

Biological aerated filters (BAFs) are submerged media wastewater treatment reactors that combine both aerobic biological treatment and solids separation (Stephenson *et al.*, 1993). They use granular or fixed media as support for microbial biofilms. Biological aerated filters are a small footprint alternative to other aerobic processes and can operate at biochemical oxygen demand (BOD) loadings (0.7-2.8 kg BOD m^{-3} d^{-1}) that are much higher than those for trickling filters (0.1-0.4 kg BOD m^{-3} d^{-1}) and activated sludge plants (0.3-0.6 kg BOD $m^{-3}d^{-1}$) (Stensel and Reiber, 1983; Stephenson *et al.*, 1993). Carbonaceous BOD removal, solids filtration and nitrification can be carried out in one unit operation and alterations to the design allow denitrification and phosphate removal (Rogalla *et al.*, 1990).

The term BAF covers a broad range of designs and configurations. All BAF processes incorporate submerged media but the influent flow can be upwards or downwards and the media can be either structured or granular. Structured media reactors are better described as contact aerators (Huang and Fishburn, 1981; Yuanjin and Yangshan, 1981). Naturally based materials, e.g. pumice and expanded shale have been used in sunken media reactors and plastics, e.g. polystyrene and polypropylene, in floating media systems (Stephenson *et al.*, 1993). The use of plastics allows a great deal of control over the size, shape and density of the media particles and thus should enable an optimum reactor to be designed.

Biological aerated filters are three phase reactors with a stationary solid phase. There have been many investigations into three phase reactors and fixed beds but there has been little independent information on biofilm reactors of BAF design except where particles used are free moving in fluidized beds or in enhanced activated sludge processes (Zabrodsky, 1966; Riemer *et al.*, 1980; Rusten, 1984; Smith *et al.*, 1987; Canziani, 1988; Roustan *et al.*, 1990; Trinet *et al.*, 1991). Previous work undertaken on other two and three phase systems have shown near ideal mixing or plug flow. Three phase reactors, such as fluidized beds, and two phase reactors using liquid and gas, e.g. activated sludge plants, show near ideal mixing (Horan *et al.*, 1991; Fdez-Palanco *et al.*, 1994). Other two phase reactors with liquid only passing through packed beds, e.g. anaerobic fixed film reactors, have tended to show ideal plug flow (Sater and Levenspiel, 1966). In biofilm reactors the microbial populations depend on the hydraulic behaviour as well as other factors such as temperature and pH (Iida and Teranishi, 1984; Roustan *et al.*, 1990; Gonzales-Martinez and Duque-Luciana, 1991). The biofilm interacts with the bulk liquid in several ways, through biomass erosion and substrate transformation. Thus the liquid characteristics have a role in controlling the biofilm growth and structure (Arvin and Harremoes, 1990). Erosion is greatest when there is high shear stress on the biofilm thus reducing reactor efficiency. High shear stress comes from turbulent flow or high liquid velocities (Lewandowski and Walser, 1991). At low liquid velocities, i.e. laminar flow, there is likely to be little erosion of biofilm but resultant nutrient limitation causes low growth rate of the biofilm. Subsequently the maximum biofilm thickness occurs in the transition zone between laminar and turbulent flow (Lewandowski and Walser, 1991). Therefore mixing and flow dispersion within a BAF are critical to their operation (Coulson and Richardson, 1978). The current study compared a floating with a sunken particulate media in both upflow and downflow mode using tracer studies to study how the liquid flow is affected by the difference in media densities. The study also sought to find whether variation in aerator position and size affected flow through the reactor (Canziani, 1988).

THEORY

In ideal plug flow of liquid, a tracer pulse leaves a reactor after the theoretical hydraulic residence time (HRT) and is measured over a short space of time (Levenspiel, 1972; Bailey and Ollis, 1986). Most real life vessels undergo some deviation from this ideal due to channelling of liquid, stagnant areas within the vessel or some liquid re-circulation within the reactor.

Using tracer studies to produce residence time distribution (RTD) curves the non-ideality of flow through a vessel can be measured and the probable reasons deduced Levenspiel, 1972). Several numbers need to be found and these are the theoretical HRT,

$$t_m = \frac{V}{v} \tag{1}$$

the peak residence time, t_p,
and the mean residence time of the tracer,

$$t = \frac{\Sigma t C_t}{\Sigma C_t} \tag{2}$$

which can be normalised to

$$\tau = \frac{t}{\bar{t}} \tag{3}$$

Differences in the mean residence time of the tracer and theoretical HRT indicate channelling or stagnant zones. Differences in peak residence time and the mean residence time of the tracer indicate mixing. Another indication of the degree of mixing is the value of the dispersion number, D/uL. This number is calculated from the variance in the output tracer concentration values (Levenspiel, 1972).

For closed vessels or simple systems, where there is one inlet and one outlet (Nauman and Buffman, 1983), the relationship between variance and dispersion is;

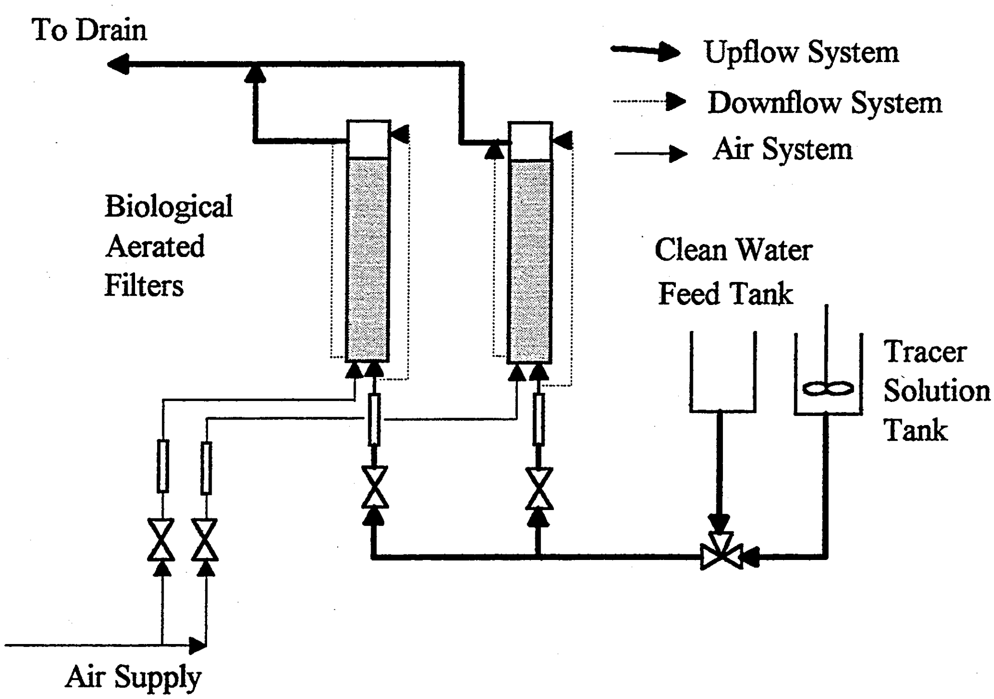

Figure 1. Flow Diagram of Reactor Set-Up.

$$\sigma_\theta^2 = 2\frac{D}{uL} - 2\left[\frac{D}{uL}\right]^2 (1 - e^{(1-uL/D)}) \tag{4}$$

Therefore through iteration D/uL can be found.

Using the tanks-in-series model the equivalent number of stirred tank reactors, N, can be found that equates to the performance of the experimental BAFs under each experimental condition. This can be found from the normalised variance;

$$\sigma_\theta^2 = \frac{1}{N} \tag{5}$$

To remove the effects of channelling / stagnant zones on the HRT profiles the variance was based on the tracer mean residence time and not on the volume / flowrate residence time.

MATERIALS AND METHODS

Two identical, cylindrical polyvinylchloride (PVC) reactors were built, one to hold a sunken media and the other to hold a floating media, with dimensions shown in Table 1. The vessels used were 0.2m in diameter which was approximately 50 mean particle diameters. This meant that wall effects were limited thus allowing data produced to be applied to larger scale reactors (Lang *et al.*, 1993). To keep the media in place rigid polypropylene mesh with 1.5mm diameter holes was placed at the top of the beds and held in place using clamps.

In the first liquid / air distribution design the air was introduced at the bottom of the reactor using a single 10 cm diameter filter nozzle positioned centrally and a 1.27 cm PVC L-bend was used to direct the incoming liquid towards the aerator (figure 2a). In the second design three 5 cm filter nozzles (Degremont) were placed around the base-plate with an identical liquid distributor placed centrally (figure 2b).

The reactors were designed to allow either liquid upflow or downflow. The media used was recycled polypropylene (Cookson plc, Cheshire, U.K.) which was re-extruded to form even particle sizes for both types of media. The pure polypropylene was used for the floating media with a relative density of 0.92 and a mixture of polypropylene (60%) and calcium carbonate (40%) (Britomya V, Croxton and Gary, Surrey, U.K.) was used to produce the sunken media of relative density 1.05.

Tracer studies were carried out at three different liquid flowrates of 0.35, 0.7 and 1.4 l min.$^{-1}$ with theoretical residence times of 72, 36 and 18 min. respectively. The studies were carried out for both reactors at each flowrate under both liquid downflow and upflow conditions with no air. These experiments were then repeated with compressed air injected at the bottom of the reactors at an air : liquid flowrate ratio of 10:1, typical of BAF systems.

TABLE 1. Media and Reactor Design.

Media Shape	Cylindrical
Media Size : Diameter (mm)	2.3 - 2.7
: Length (mm)	4 - 6
Reactor Height (m)	2.00
Reactor Diameter (m)	0.20
Height of Media In Reactor (m)	1.72
Media Volume (l)	54
Voidage (%)	42

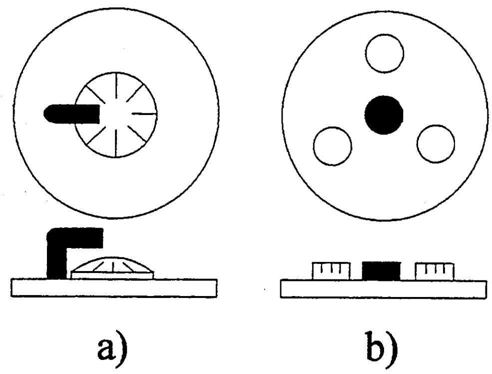

Figure 2. Air/Liquid Distribution: (a) Single Aerator; (b) Triple Aerator (White-Air Distributors, Black-Liquid Distributors).

The tracer used was a pulse of 0.1M solution of sodium chloride (GPR, Merck, Poole, U.K.) which was pumped into the reactor at the appropriate flowrate for 1 min. Since conductivity is directly related to molar concentration the resulting output tracer concentration was measured using a flow conductivity meter (LTH Electronics Ltd. Leicester, U.K.) every minute for up to 200 min. from the time of tracer input.

RESULTS AND CONCLUSION

The residence time distribution (RTD) curves were found for the three liquid flowrates 0.35, 0.7 and 1.4 l/min. Figures 3, 4, 5 and 6 show the results when using a liquid flowrate of 0.7 l/min for the single nozzle and triple nozzle aeration systems. The tracer fraction was calculated from the concentration reading taken at one minute intervals divided by the initial input concentration. From these graphs the points at which the mean residence time (t_m) and peak (t_p) residence times occurred were noted (Table 2). The variance in the measured points and thus the dispersion numbers were calculated from the conductivity readings (Table 3).

The change in aerator set-up did not result in a change in flow through the reactor. Mean and peak time showed almost no difference between the two designs (Table 2). Dispersion numbers for the non-aerated and downflow experiments were very similar for both distribution designs and those for the aerated, upflow experiments showed a slight reduction for the triple nozzle design. These results indicated that the aerator design had little effect on the flow and confirmed that the media itself dictated the flow of liquid and air (Canziani, 1988).

Overall the non-aerated experiments showed close to plug flow characteristics whereas the aerated experiments showed pronounced tailing of the curves similar to the results found in other studies, with and without biomass (Riemer *et al.*, 1980; Canziani, 1988).

TABLE 2. Tracer Study Results.

Liquid Flowrate (l/min)	Air Flowrate (l/min)	Theoretic HRT (min)	Sunken Media Peak Time (min)	Mean Time (min)	Floating Media Peak Time (min)	Mean Time (min)
<u>Single</u> <u>Aerator</u>						
Up						
0.35	0	72	65	71	60	65
0.7	0	36	33	36	32	36
1.4	0	18	16	17	16	17
0.35	3.5	72	36	47	33	47
0.7	7	36	16	27	21	29
1.4	14	18	11	14	13	14
Down						
0.35	0	72	57	61	69	72
0.7	0	36	29	30	34	35
1.4	0	18	16	17	16	17
0.35	3.5	72	25	46	44	55
0.7	7	36	15	28	22	31
1.4	14	18	11	16	12	14
<u>Triple</u> <u>Aerator</u>						
Up						
0.7	0	36	31	36	31	36
0.7	0.7	36	17	25	22	27
Down						
0.7	0	36	28	30	34	36
0.7	0.7	36	16	28	24	31

Reduction in the mean residence time, t, for the aerated experiments showed increased channelling or stagnant regions.

With no air passing through the reactors the sunken media in upflow and floating media in downflow showed almost no channelling but some degree of mixing at the lowest liquid flowrates. Thus under these conditions with a theoretical residence time of 72 min the mean times for upflow and downflow were 71 and 72 min respectively but the peak times dropped to 65 and 69 min. The sunken media in downflow and floating media in upflow showed some degree of mixing and a higher degree of channelling and stagnant zones. This small amount of mixing was confirmed by the low value of the dispersion numbers (Tables 3 and 5) and by the shape of the curves approximating that for ideal plug flow. As liquid flowrates increased, the difference in the mean and peak HRTs compared to the theoretical HRT was less, with a maximum difference of only 2 min at the highest flowrates. This demonstrated a reduction in mixing and channelling.

When air was injected there was a sharp decrease in the mean times for all flowrates, e.g., in upflow the sunken media reactor mean time dropped from 71 minutes to 47 minutes at a theoretical HRT of 72 min. This showed that aeration greatly increased the amount of channelling. At each flowrate the mean HRT was similar for both upflow and downflow with some difference between upflow and downflow for the peak HRT. With sunken media the peak HRTs were low for the downflow set-up (25 min compared to 36 min in upflow at a theoretical HRT of 72 min) whereas with floating media the peak HRT was low for the upflow set-up (47 min compared to 55 min in downflow at a theoretical HRT of 72 min).

Channelling within a packed bed means that localised high liquid velocities occur resulting in turbulent flow (Smith *et al.*, 1987). In these areas there will be reduced

TABLE 3. Dispersion Numbers for Tracer Experiments.

Liquid Flowrate (l/min)	Air Flowrate (l/min)	Upflow Sunken Media (D/uL)	Floating Media (D/uL)	Downflow Sunken Media (D/uL)	Floating Media (D/uL)
<u>Single Aerator</u>					
0.35	0	0.013	0.022	0.008	0.007
0.7	0	0.025	0.031	0.022	0.009
1.4	0	0.012	0.011	0.014	0.009
0.35	3.5	0.081	0.086	0.200	0.061
0.7	7	0.200	0.106	0.241	0.083
1.4	14	0.079	0.014	0.388	0.197
<u>Triple Aerator</u>					
0.7	0	0.022	0.023	0.022	0.010
0.7	7	0.135	0.095	0.241	0.081

TABLE 4. Tanks in Series Model - Equivalent Number of Tanks.

Liquid Flowrate (l/min)	Air Flowrate (l/min)	Upflow Sunken Media (D/uL)	Floating Media (D/uL)	Downflow Sunken Media (D/uL)	Floating Media (D/uL)
Design 1					
0.7	0	41	32	46	118
0.7	7	5	9	4	12
Design 2					
0.7	0	46	43	46	104
0.7	7	7	11	4	12

concentration of biomass and in some cases complete biomass absence is likely. Removal of some biomass in long term operations of most fixed film bioreactors is useful because of the build up of inactive biomass (Rittmann *et al.*, 1992). However, in BAFs this is achieved through backwashing (Stensel and Reiber, 1983). Therefore, in order to keep biomass concentration high channelling needs to be kept to a minimum. A degree of mixing is needed to allow time for the biomass to remove the maximum amount of nutrients from the liquid. Too much mixing can result in greater shear stress which is detrimental to biofilm growth (Riemer *et al.*, 1980). However, a certain amount of mixing is of benefit to the development of biofilm as it distributes nutrients. Thus a reactor may be very good at removing dissolved nutrients but will have poor solids filtration and vice versa.

When the tracer graphs (Figs 3-6) are compared the most obvious difference is that between aerated and non-aerated lines. Some of this difference could be due to an effective liquid void reduction due to entrapment of air; however the maximum actual void volume taken up by the air at any one time was less than 5% thus this had a minimal effect on altering the hydraulic residence time. The air enhanced the channelling effects and mixing within the reactors but may have also increased the proportion of stagnant regions.

If the dispersion numbers obtained in these experiments are compared, the non-aerated runs show a maximum dispersion number of 0.031, close to ideal plug flow. Whereas with the aerated runs the maximum dispersion number obtained was 0.241. These compare to values of 0.46 to 0.67 for activated sludge plants which are two phase and well mixed (Horan *et al.*, 1991). therefore aeration in BAFs causes a flow pattern somewhere between plug flow and well mixed.

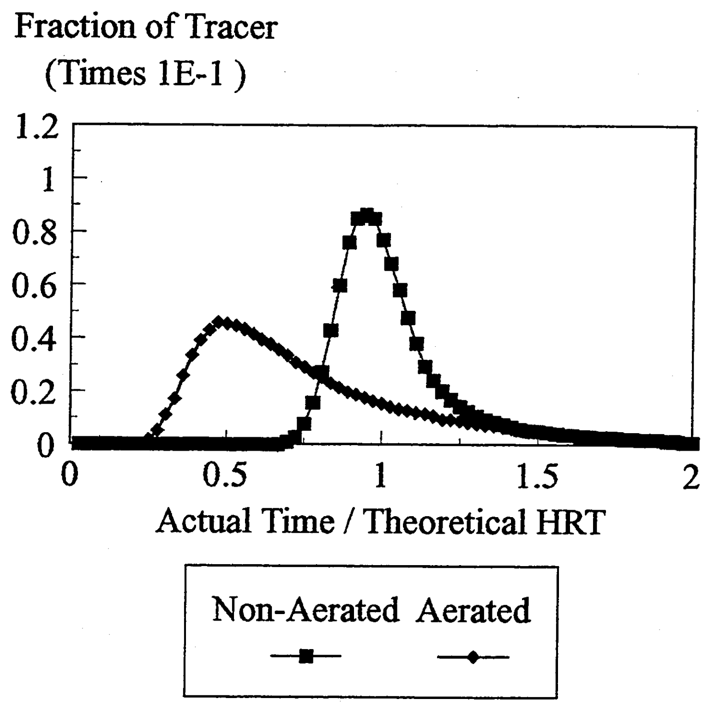

Figure 3a. Upflow, Sunken Media Tracer Curves with Single Aerator.

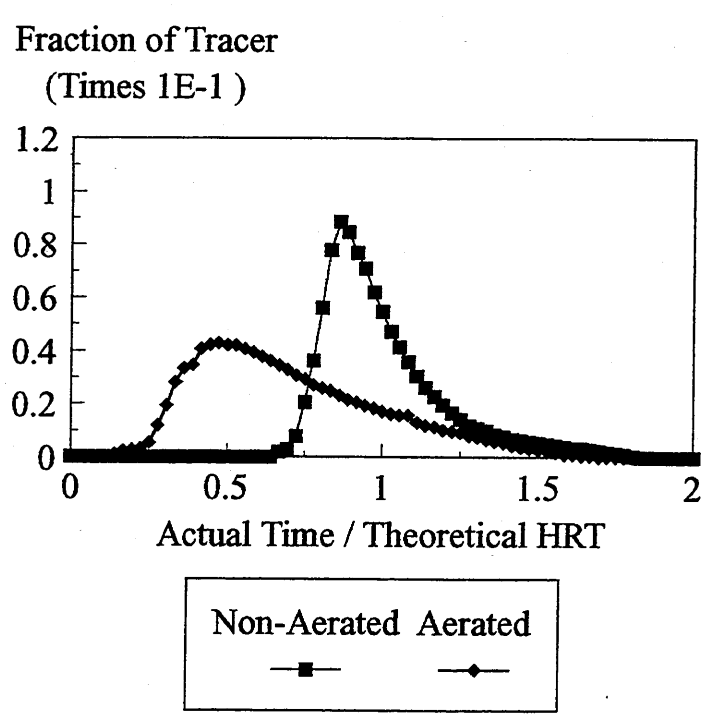

Figure 3b. Upflow, Sunken Media Tracer Curves with Triple Aerator.

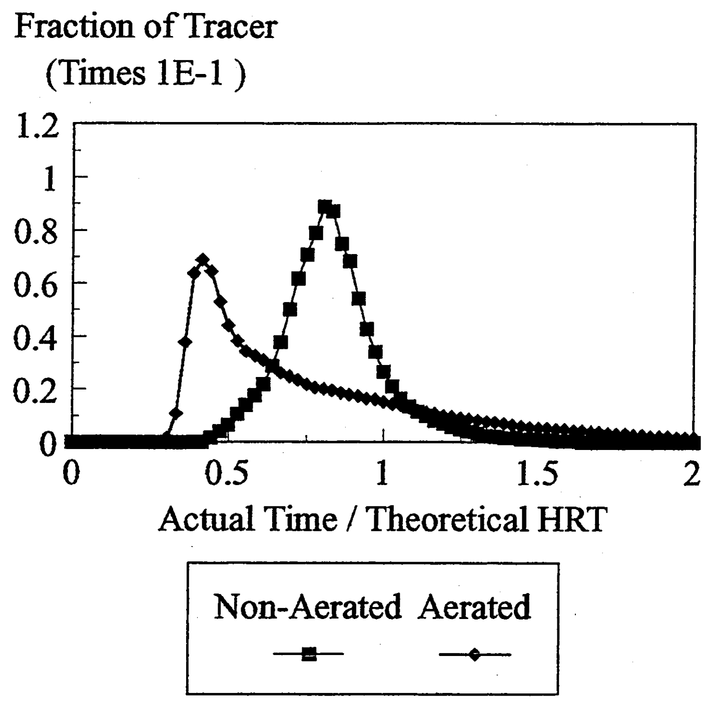

Figure 4a. Downflow, Sunken Media Tracer Curves with Single Aerator.

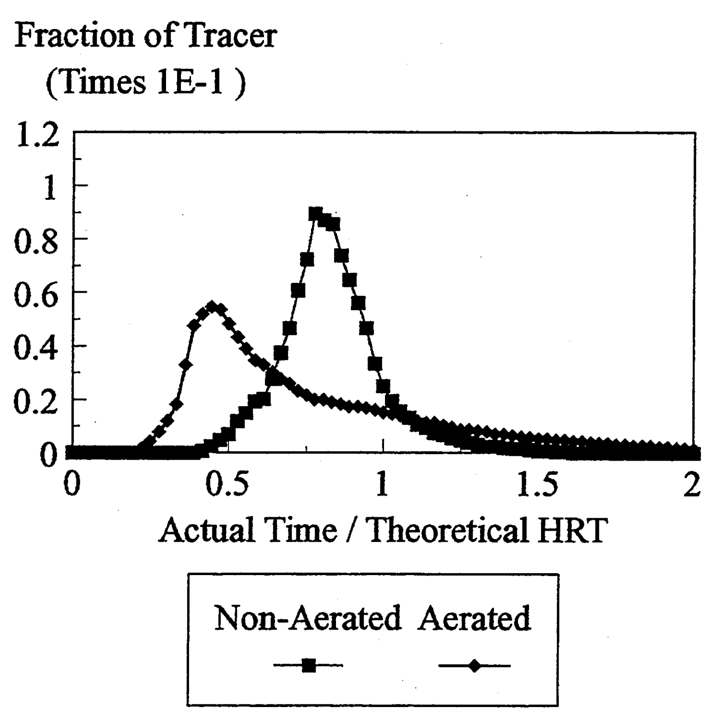

Figure 4b. Downflow, Sunken Media Tracer Curves with Triple Aerator.

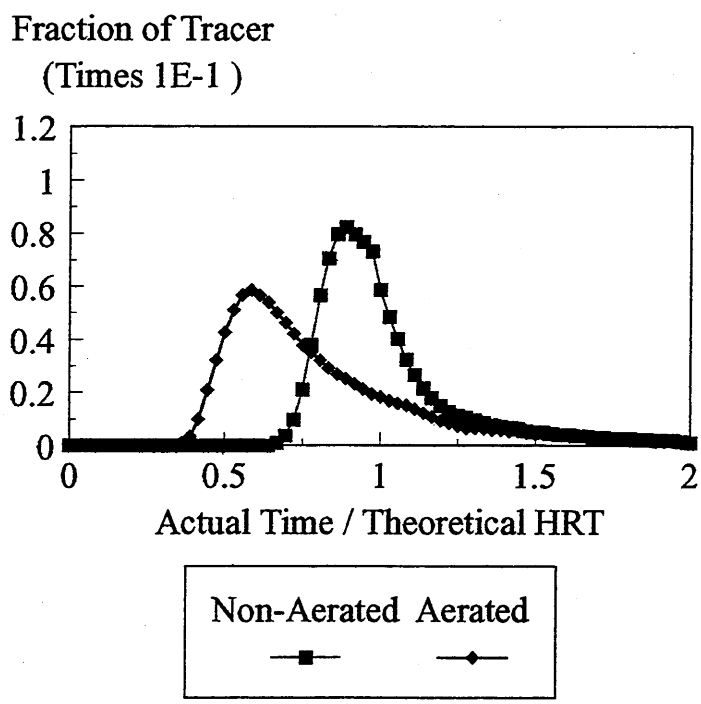

Figure 5a. Upflow, Floating Media Tracer Curves with Single Aerator.

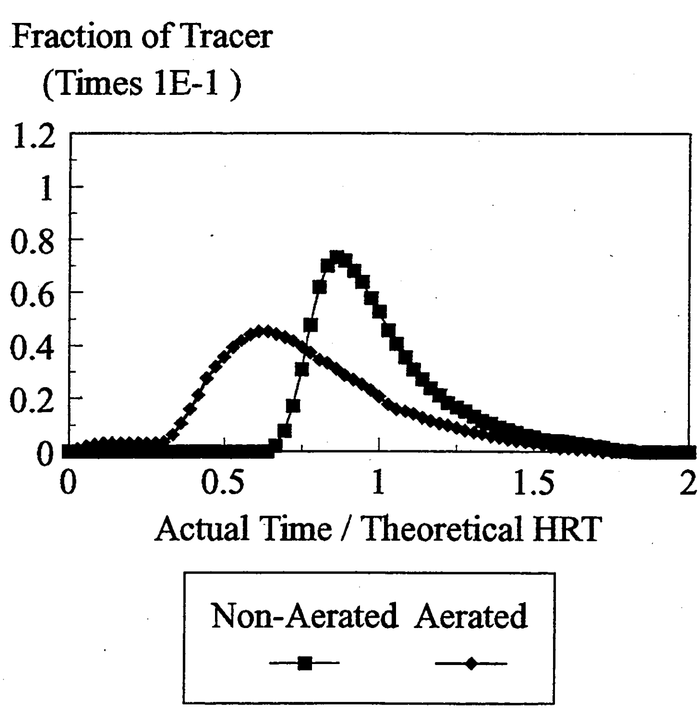

Figure 5b. Upflow, Floating Media Tracer Curves with Triple Aerator.

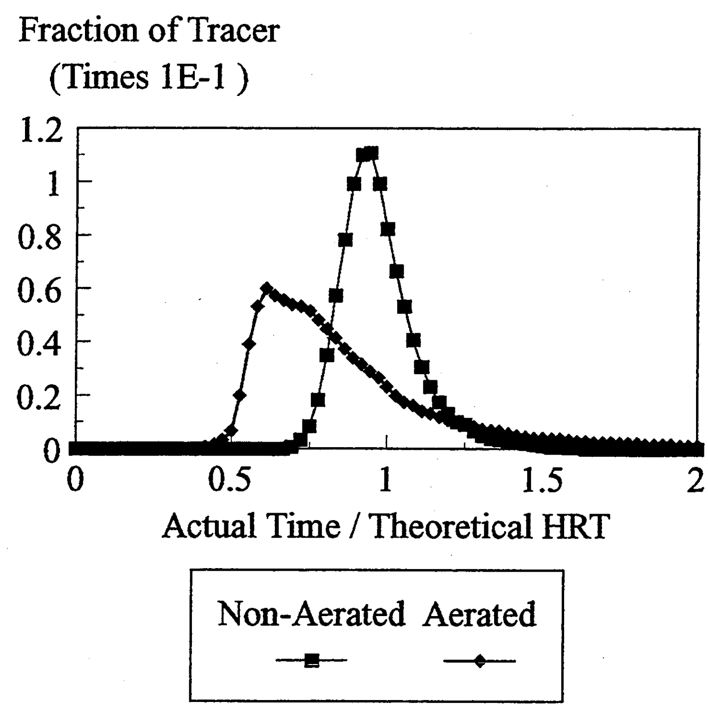

Figure 6a. Downflow, Floating Media Tracer Curves with Single Aerator.

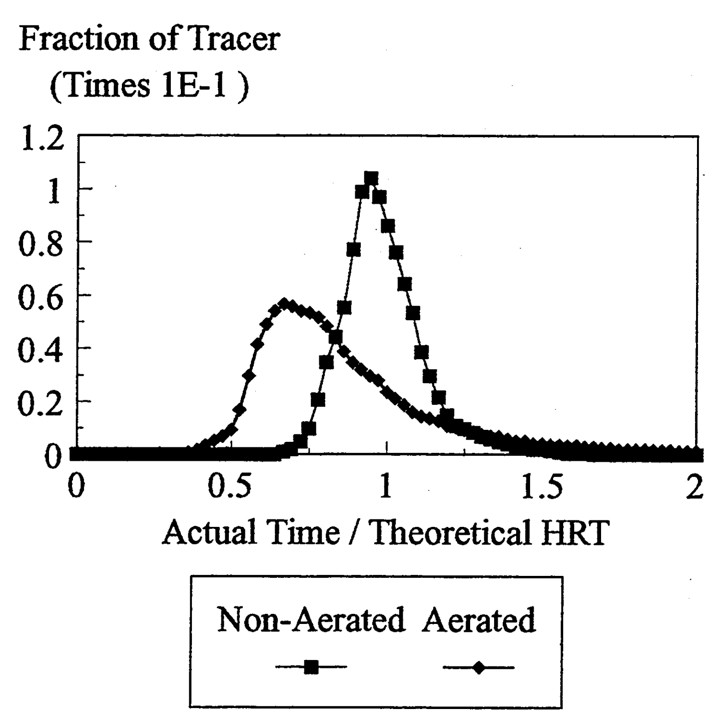

Figure 6b. Downflow, Floating Media Tracer Curves with Triple Aerator.

The results can be more easily compared by looking at the equivalent number of tanks in series (Table 4). The greater the number of tanks, the closer the approximation to plug flow. Channelling / mixing effects have been removed from the raw data to obtain these results, thus as with the dispersion numbers the values indicate the proportion of mixing to plug flow only.

Comparing the results for these experiments, there appeared to be little difference between the reactors at high liquid flowrates. As the liquid flowrates decrease the differences were more obvious. The results suggested that during start-up of biological fixed film reactors floating media especially with liquid upflow, would perform best for solids removal due to much lower mixing (shown by lower dispersion numbers). Higher mixing would tend to keep the solids in suspension within the macrostructure of the filter and increase the degree of sloughing of biomass through higher shear. Sunken media with liquid downflow would probably perform best for nutrient removal at start-up, especially at low flowrates through increased wastewater contact through mixing. From these results high suspended solids wastewater such as secondary domestic wastewater would best be processed using upflow floating media reactors. Wastewater containing low suspended solids such as landfill leachate or tertiary municipal wastewater would best be treated using sunken media downflow reactors. These results indicate how such reactor designs would perform only during start-up. They do not show how the reactors would perform at steady-state as biomass within the reactors is likely to alter the flow of liquid and air.

To confirm the effect that air and liquid flowrates have on different BAF configurations, biological studies are being undertaken to determine whether the tracer studies can be used to predict BAF performance.

NOMENCLATURE

C_t	Concentration	mol. l^{-1}
D	Dispersion coefficient	$m^2\ s^{-1}$
L	Length	m
t	Time	min.
$\bar{t}$	Tracer mean residence time	min.
t_m	Mean residence time (volume / flowrate)	min.
t_p	Peak residence time	min.
u	Liquid velocity	m s^{-1}
V	Volume	l
σ_θ^2	Normalised variance	
τ	Normalised residence time	
v	Volumetric flowrate	l $min.^{-1}$
N	Number of tanks	

ACKNOWLEDGEMENTS

One of us (A. Mann) is grateful to the U.K. Engineering and Physical Sciences Research Council (EPSRC) for a quota research studentship.

REFERENCES

Arvin E., Harremoes P. (1990). Concepts and Models for Biofilm Reactor Performance. *Wat. Sci. Tech.*, 22, 1/2, 171-192.

Bailey J.E., Ollis D.F. (1986). *Biochemical Engineering Fundamentals,* 2nd Edition, McGraw-Hill, New York, 533-586

Canziani R. (1988). Submerged Aerated Biofilters. IV - Aeration Characteristics. *Ingegneria Ambientale,* 17, 627-636.

Coulson J.M., Richardson J.F. (1978). *Chemical Engineering Vol. 2,* 3rd Edition, Pergamon, Oxford, 125-171.

Fdez-Palanco F., Real F.J., Garcia P.A. (1994). Behaviour of an Anaerobic / Aerobic Pilot Scale Fluidised Bed for the Simultaneous Removal of Carbon and Nitrogen. *Wat. Sci. Tech.*, 29, 10/11, 339-346.

Gonzale-Martinez S., Duque-Luciana J. (1992). Aerobic Submerged Biofilm Reactors for Wastewater Treatment. *Wat. Res.*, 26, 825-833.

Horan N.J., Parr J., Naylor P.J. (1991). Evaluation of Tracers for the Determination of the Mixing Characteristics of Activated Sludge Reactors. *Environ. Tech.*, 12, 603-608.

Huang C.S., Fishburn G.A. (1981). Contact Aeration Using Plastic Media - A Case Study. *Wat. Eng. Man.*, 128, 30-31.

Iida Y. and Teranishi A. (1984). Nitrogen Removal from Municipal Wastewater by a Single Submerged Filter. *J. WPCF.* 56, 251-258.

Lang J.S., Giron J.J., Hansen A.T., Tussel R.R., Hodges Jr. W.E. (1993). Investigating Filter Performance as a Function of the Ratio of Filter Size to Media Size. *Amer. Wat. Works Assoc.*, Oct., 122-130.

Levenspiel O. (1972). *Chemical Reaction Engineering,* 2nd Edition, Wiley, New York, 253-314

Lewandowski Z., Walser G. (1991). Influence of Hydrodynamics on Biofilm Accumulation. *Environ. Eng,* 11, 619-624.

Nauman E.B., Buffman B.A. (1983). *Mixing in Continuous Flow Systems,* 3-130, Wiley, New York

Riemer M., Kristensen G.H., Harremoes P. (1980). Residence Time Distribution in Submerged Biofilters. *Wat. Res.*, 14, 949-958.

Rittmann B. E., Trinet F., Amar D., Chang H.T. (1992). Measurement of the Activity of a Biofilm: Effects of Surface Loading and Detachment on a Three Phase, Liquid Fluidised Bed Reactor. *Wat. Sci. Tech.*, 26, 3/4, 585-594.

Rogalla F., Payraudeau M., Bacquet G., Bourbigot M.-M., Sibony J., Gilles P. (1990). Nitrification and Phosphorus Precipitation with Biological Aerated Filters. *J. WPCF.*, 62, 169-176.

Roustan M., Metral C., Capdeville B., Audic J. M., Marty A. (1990). Hydrodynamic Study of Three Phase Reactor with Bubble Supported Particles. *5th Med. Cong. Chem. Eng.*, Barcelona 5-7 Nov.

Rusten B. (1984). Wastewater Treatment with Aerated Biological Filters. *J. WPCF.*, 56, 424-431.

Sater V.E., Levenspiel O. (1966). Two Phase Flow in Packed Beds. *I & E C Fundamentals,* 5, 86-92.

Smith D. N., Stiegel G. J., Ruether J. A. (1987). *Encyclopaedia of Fluid Mechanics,* 6, 535-682.

Stensel H. D., Reiber S. (1983). Industrial Wastewater Treatment with a New Biological Fixed-Film System. *Environ Prog.*, 2, 110-115.

Stephenson T., Mann A., and Upton J. (1993). The Small Footprint Wastewater Treatment Process. *Chem. Ind.*, 533-536.

Trinet F., Heim R., Amar D., Chang H. T., Rittman B. E. (1991). Study of Bifilm and Fluidisation of Bioparticles in a Three Phase Liquid Fluidised Bed Reactor. *Wat. Sci. Tech.*, 23, 1347-1354.

Yuanjin Z., Yangshan W. (1981). A New Biological Treatment of Industrial Wastewater and Municipal Sewage - Biological Contact Aeration. *Wat. Ind. 81,* Brighton, June.

Zabrodsky S.S. (1966). *Hydrodynamics and Heat Transfer in Fluidized Beds,* 1-24, M.I.T. Press, Massachusetts.

Chapter 4.

PERFORMANCE OF FLOATING AND SUNKEN MEDIA BIOLOGICAL AERATED FILTERS UNDER UNSTEADY STATE CONDITIONS.

Submitted for Publication,

Water Research.

PERFORMANCE OF FLOATING AND SUNKEN MEDIA BIOLOGICAL AERATED FILTERS UNDER UNSTEADY STATE CONDITIONS.

Allan Mann and Tom Stephenson.

School of Water Sciences, Cranfield University, Cranfield, Bedford. MK43 0AL, UK.

ABSTRACT

The Biological Aerated Filter (BAF) combines both biological treatment and solids removal in one submerged, aerated three phase reactor. Over the last 20 years biological aerated filters have been developed not only as competitors for other secondary treatment systems such as trickling filters and activated sludge plants but also as tertiary treatment systems.

The aim of this work was to compare the performance of two such reactors, one containing a floating media and the other containing a sunken media under identical conditions during start-up and with increasing flowrates. Two methods of start-up were used. The first method used activated sludge as seed which was recycled through the reactor, the second method involved simply passing the process liquid (settled domestic sewage) through the reactor at the initial flowrate. Following a period of steady-state the flowrates were increased incrementally and the performance of the reactors analysed during and after each increase.

The results for the start-up indicated that the overall time to reach steady-state using both methods was approximately the same but due to the initial time required for the recycling of the activated sludge, this method took slightly longer. From an initial start-up loading of 0.486 kg m^{-3} d^{-1} (SS) and 0.568 kg m^{-3} d^{-1} (sCOD) suspended solids and soluble chemical oxygen demand removal rates dropped to below 50% when loadings of 1.397 kg m^{-3} d^{-1} (SS) and 1.403 kg m^{-3} d^{-1} (sCOD) were reached.

KEYWORDS

Biological aerated filter; increased organic loading; plastic media; start-up; tracer study.

INTRODUCTION

Aerobic submerged fixed film reactors use structured or granular media as support for biofilm growth. Contact Aerators use structured media which gives a large surface area for biofilm attachment but are limited in their filtration ability (Gros, 1992; Ryhiner *et al.*, 1992; Gray, 1993). Biological Aerated Filters though use granular media which has a larger surface area and improved filtration efficiency (Canler and Peret, 1994; Rogalla *et al.*, 1991; Stensel *et al.*, 1988; Dillon and Thomas, 1990). Thus carbonaceous removal, solids filtration and nitrification can be carried out in a single unit (Mann *et al.*, 1995). There are a large number design variations available which includes liquid flow direction (upflow or downflow) and media type (floating or sunken).

The most important factor that has led to the increased use of BAFs are their ability to treat high volumetric loading rates. Loading rates of 2.5 kg BOD_5 m^{-3} d^{-1} compared to 0.06 kg BOD_5 m^{-3} d^{-1} for percolating filters and 0.38 kg BOD_5 m^{-3} d^{-1} for activated sludge plants have been found (Smith *et al.*, 1990). This ability to treat high loadings is due to the high biomass density achieved within BAF reactors. Generally concentrations are 4 to 5 times higher than those found in suspended growth systems and specifically 8 to 9 times greater than in activated sludge plants (Smith *et al.*, 1990; Faup *et al.*, 1982).

Three important factors influence the growth of biofilm within BAF reactors. These are flowrate and nutrient concentrations of the wastewater to be treated, the backwashing regime used to remove excess biomass and the method of start-up (Bacquet *et al.*, 1991). There are three methods of start-up that have been used. Start-up of continuous reactors, initially as batch reactors followed by increasing flowrates, were found to take approximately 1 month to reach steady state at the process flowrate (Smith *et al.*, 1990). Secondly the use of the process liquid at either the nominal process flowrate, which was found to take 5 months (Bacquet *et al.*, 1991) or increasing the flowrate from an initial low value over a period of time (Smith *et al.*, 1990). Alternatively start-up may be carried out by seeding with activated sludge with steady state reached after 1.5 months (Hamoda and Abd-El-Bary, 1987; Park and Ganczarczyk, 1994; Faup *et al.*, 1982).

After initial start-up, the loading rate to a biological reactor is increased in a series of incremental steps, equivalent to shock loadings. Work carried out on anaerobic reactors has shown that fixed film reactors such as anaerobic filters and fluidised beds perform better during shock loadings than suspended biomass reactors such as upflow anaerobic sludge blanket reactors (Chung and Choi, 1993). Both fluidised beds and anaerobic filters have been reported to show only a slight decline in performance with shock loadings of 3 times the design loading (Tanaka and Matsuo, 1986; Barnes *et al.*, 1983). Similarly in aerobic systems fixed-film reactors were found to perform better than suspended biomass systems (Young *et al.*, 1975), but also perform better than anaerobic system (Kiestra and Eggers, 1986).

There are few published papers for fixed film reactors that include information on the start-up method used and only some of these give results collected during this period. The aims of this work was to compare two methods of initial start-up and subsequent shock loading effects. Firstly combining activated sludge seeding with recycle and secondly using an initial low flowrate which was increased in stages until SS and sCOD removal dropped to below 50%.

MATERIALS AND METHODS

Two identical polyvinylchloride (PVC) reactors were built 2m in height with a diameter of 0.2m. This diameter was designed to be approximately 50 mean particle diameters in size to limit wall effects. Limiting wall effects allowed the results produced in pilot scale reactors to be compared with results found in larger scale reactors (Lang *et al.*, 1993). The reactors were operated using settled domestic sewage introduced at the bottom of the reactors through a centrally placed 5 cm filter nozzle (Degremont). Aeration was carried out using three identical nozzles placed around the base plate. In addition sample ports were placed along the length of the reactors at 0.2m intervals. The medias were retained using a rigid polypropylene mesh with 1.5mm diameter holes, held at the top of the reactors using clamps (Mann *et al.*, 1995).

The two medias used were made of recycled polypropylene (Cookson plc, Cheshire, U.K.), identical in shape and size except that one was made of pure polypropylene, floating, with a relative density of 0.92 and the other contained a mixture of polypropylene (60%) and calcium carbonate (40%) (Britomya V, Croxton and Gary, Surrey, U.K.), sunken, with a relative density of 1.05.

Aeration was undertaken using a constant air : liquid ratio of 10 : 1. Throughout, the liquid temperature was maintained at 16.5 - 17.5 °C using a heat exchanger. Backwashing took place on a daily basis to remove excess biomass. During backwashing the process air and liquid were initially turned off and 2.5 l of liquid were drained from the bottom of the reactors. This was followed by an air scour of 25 l/min for 1 min. The reactors were left for 2 min and a further 7.5 l of liquid was drained. The process air and liquid flows were then resumed.

The work was carried out in two stages. The first stage investigated two methods of start-up. In one method settled sewage was passed upflow through the reactors at a process flowrate of 0.2 l/min. The second method involved recycling activated sludge through the reactor at this same flowrate and then changed to settled sewage after a week. The subsequent results were then compared. The second stage compared the performance of the two reactors over periods of increasing flowrate. Following a period of steady-state performance at the initial liquid flowrate of 0.2 l/min. (0.486 kg SS m^{-3} d^{-1}, 0.568 kg sCOD m^{-3} d^{-1}), the flowrate was increased to 0.3 l/min. (0.776 kg SS m^{-3} d^{-1}, 0.842 kg sCOD m^{-3} d^{-1}). Monitoring took place at the time of flow increase and over a period of time following the increase until steady-state was re-established. After a period of steady-state the flowrate was then increased again to 0.4 l/min. (1.035 kg SS m^{-3} d^{-1}, 1.097 kg sCOD m^{-3} d^{-1}) followed by 0.5 l/min. (1.397 kg SS m^{-3} d^{-1}, 1.403 kg sCOD m^{-3} d^{-1}) and the analysis repeated. Steady state was assumed to have occurred when a constant sCOD removal rate was maintained over a period of 5 d.

Samples were taken on a daily basis and were analysed for suspended solids, total chemical oxygen demand (tCOD), soluble chemical oxygen demand (sCOD), ammonia, nitrates, nitrites, temperature, pH and dissolved oxygen. Backwash solids and solids settlability were also measured. Biological oxygen demand (sBOD) was also measured initially to obtain BOD : COD ratios for influent and effluent (APHA, 1992).

RESULTS

Throughout the experimental period the effluent dissolved oxygen concentration remained above 6.5 mg/l. Effluent suspended solids concentrations ranged from 72 to 125 mg/l and sCOD concentrations ranged from 76 to 217 mg/l. Though the reactors performed well at sCOD and SS removal there was very little nitrification. Influent ammonia levels ranged from 14.6 to 33.2 mg/l. The maximum ammonia removal was 18% which was measured in the floating media reactor with the sunken media reactor showing a maximum of 11% removal. There was only a small maximum increase in nitrite and nitrate levels indicating of 1.6 mg/l (floating media) and 0.9 mg/l (sunken media).

The two methods of start-up required similar periods of time to reach steady state. Seeding with activated sludge required a start-up period of 30 d (Figure 1). This length of time included the 7 d during which activated sludge was recycled through the reactors. This compared to 28 d required to reach steady state using the process liquid only (Figure 2). Thus the use of activated sludge seeding reduced the time between the start of treatment of the process liquid and steady state but the time required for seeding and recycling extended this period. Subsequently one benefit seen when comparing the results was a more rapid establishment of biofilm using activated sludge and resulting more rapid increase in sCOD removal efficiency. Less scattering of points for the seeding results (Figure 1) compared to those for the process liquid start-up (Figure 2) also indicated the establishment of a more stable biofilm. For both start-up using activated sludge and using settled sewage, greater solids removal occurred once steady state had been reached with the floating media (80-90%) compared to that for the sunken media (50-60%). However sCOD removal at steady state for both start-up methods was similar (80%). After a steady state period of 12 weeks at a liquid flowrate of 0.2 l/min the flowrate was increased to 0.3 l/min (Figure 3). After the flowrate was increased at 09:00 (Figure 3), the solids removal dropped to almost zero for both reactors. Solids removal then gradually increased and after 7 h the removal efficiency reached 49% for the sunken

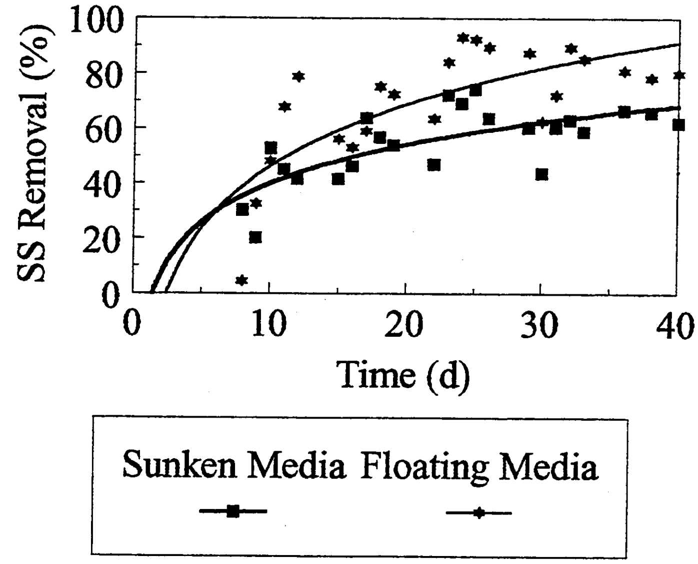

Figure 1a. Start-Up Using Activated Sludge (Suspended Solids Removal).

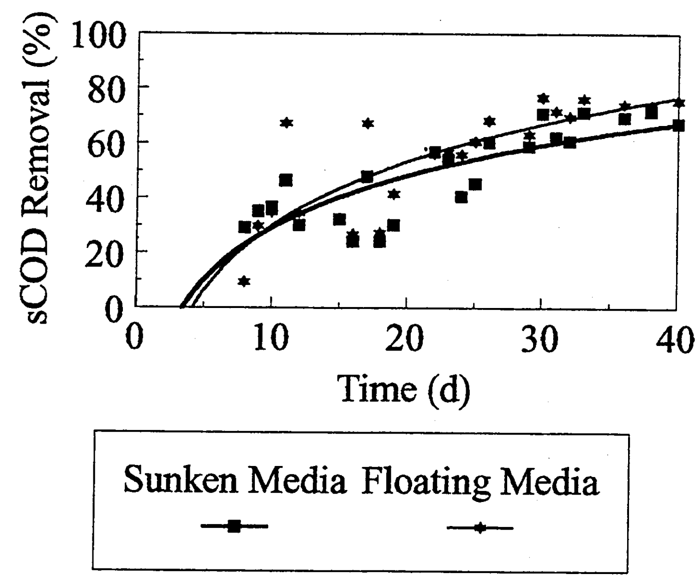

Figure 1b. Start-Up Using Activated Sludge (Soluble COD Removal).

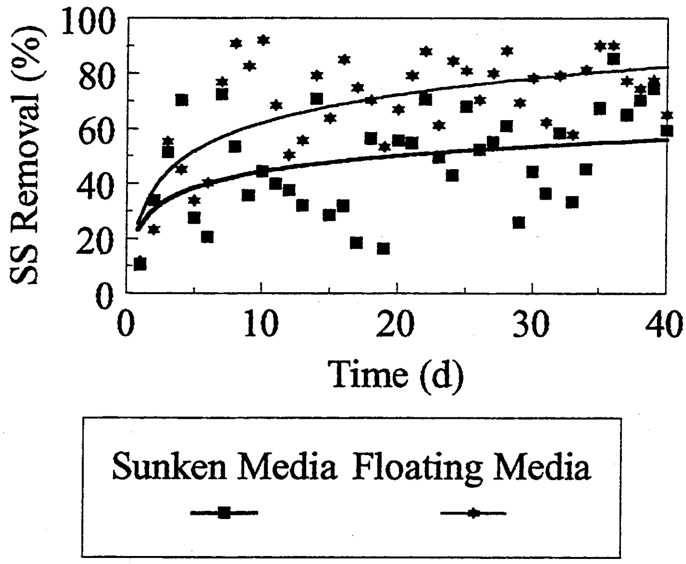

Figure 2a. Start-Up Using Settled Domestic Sewage (Suspended Solids Removal).

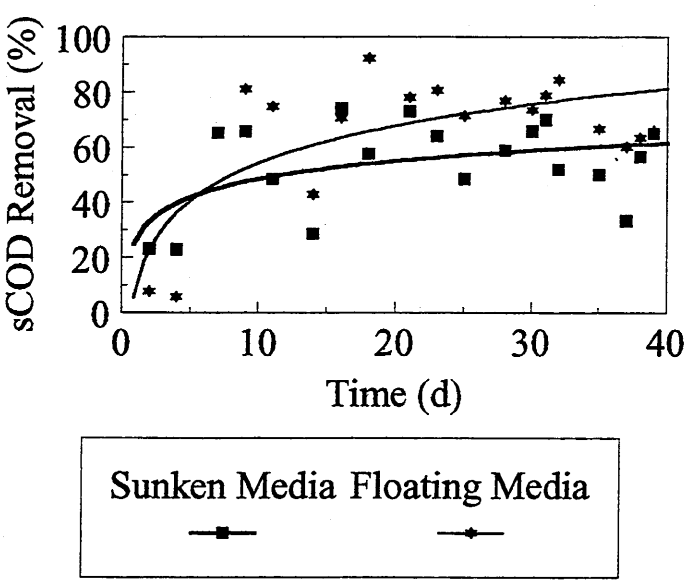

Figure 2b. Start-Up Using Settled Domestic Sewage (Soluble COD Removal).

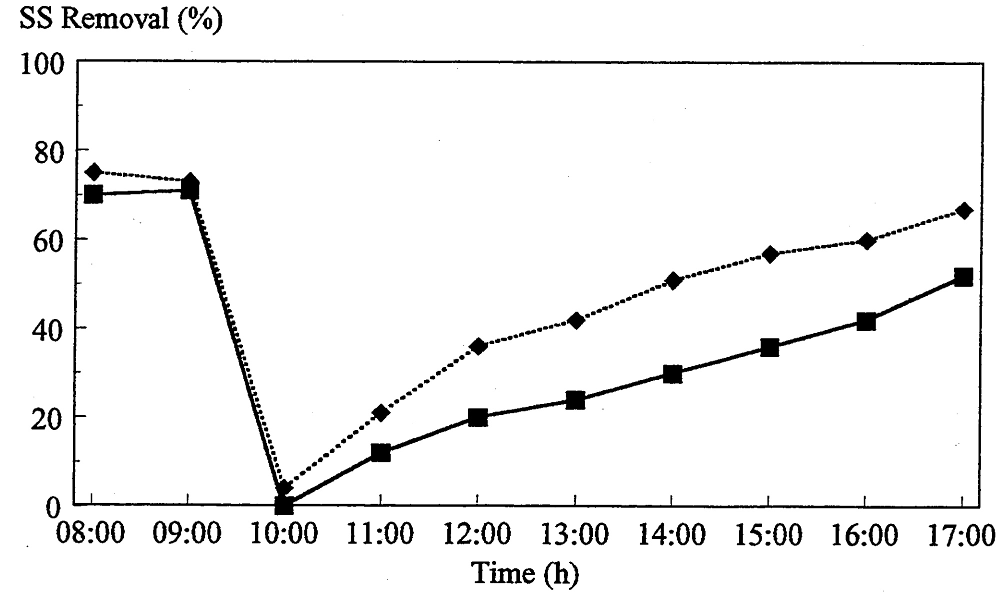

Figure 3a. Percentage Suspended Solids Removal During Flow Increase 0.2-0.3 l min^{-1}. (■-Sunken Media, ♦ -Floating Media)

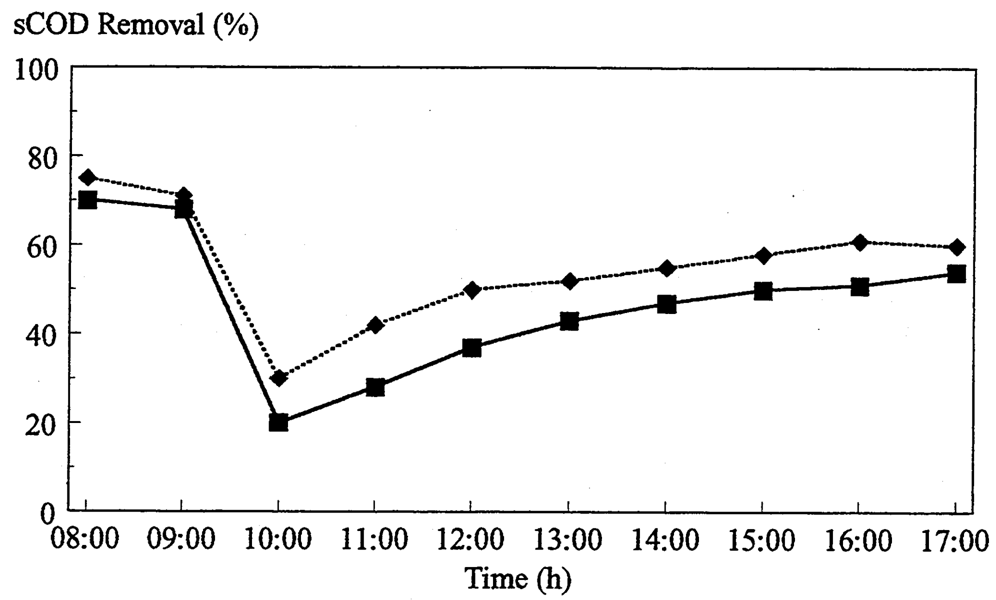

Figure 3b. Percentage Soluble COD Removal During Flow Increase 0.2-0.3 l min^{-1}. (■-Sunken Media, ♦ -Floating Media)

TABLE 1. Flows, Loadings, Mean Influent and Effluent SS and sCOD Concentrations and Standard Deviation ( ) at Steady State.
(r.t. = empty bed retention times)

		SS Conc. (mgl^{-1})			Soluble COD Conc. (mgl^{-1})*		
Liquid Flowrate ($l\ min.^{-1}$)	Loading ($kg\ m^{-3}\ d^{-1}$)	Influent	Effluent Sunken Media	Effluent Floating Media	Influent	Effluent Sunken Media	Effluent Floating Media
0.2	0.486 (SS)	91.2	30.7	19.7	106.5	33.7	26.3
(r.t. = 2.0)	0.568 (S_{COD})	(7.6)	(3.7)	(2.1)	(6.6)	(3.2)	(3.5)
0.3	0.776 (SS)	97.0	39.2	24.0	105.2	40.6	31.6
(r.t. = 1.3)	0.842 (S_{COD})	(8.4)	(5.1)	(3.6)	(6.5)	(5.2)	(3.4)
0.4	1.035 (SS)	97.0	48.0	28.6	102.8	51.4	39.8
(r.t. = 1.0)	1.097 (S_{COD})	(6.7)	(3.7)	(3.2)	(6.4)	(1.4)	(2.0)
0.5	1.397 (SS)	104.8	68.4	42.2	105.2	73.4	63.0
(r.t. = 0.8)	1.403 (S_{COD})	(6.0)	(3.8)	(2.5)	(6.7)	(2.6)	(3.7)

* BOD : COD ratio, influent = 0.45, effluent = 0.2

media and 64% for the floating media. Removal of sCOD dropped to 21% for the sunken media and 30% for the floating media and after 7 h removal efficiencies of 53% (sunken media) and 59% (floating media) were obtained. Steady-state was achieved after 4 d for both suspended solids and sCOD removal (Table 1), with an efficiency drop of 6% (SS) and 7% (sCOD) for the sunken media and 3% (SS) and 5% (sCOD) for the floating media.

After a further steady state period of 14 d the flowrate was increased to 0.4 l/min with a further efficiency drop of 9% (SS) and 11% (sCOD) for the sunken media and 5% (SS) and 6% (sCOD) for the floating media. Steady-state was reached after 7 d, and following a further 14 d at steady-state the liquid flow was increased to 0.5 l/min. At this increased flowrate the solids removal dropped to 0% for both reactors (Figure 5). Recovery was very slow with only 30% removal for the sunken media and 42% removal for the floating media after 22 hours. Similarly since the higher flowrate caused most of the biomass to be washed out at the point of flow increase, the sCOD removal efficiency dropped to almost 0%. Steady state was only achieved after 14 days, but at this point the solids removal efficiency had dropped to 35% (sunken media) and 60% (floating media) and the sCOD removal had dropped to 30% and 40% respectively. Results obtained at increasing flowrates indicated that the treatment efficiency of the reactors dropped

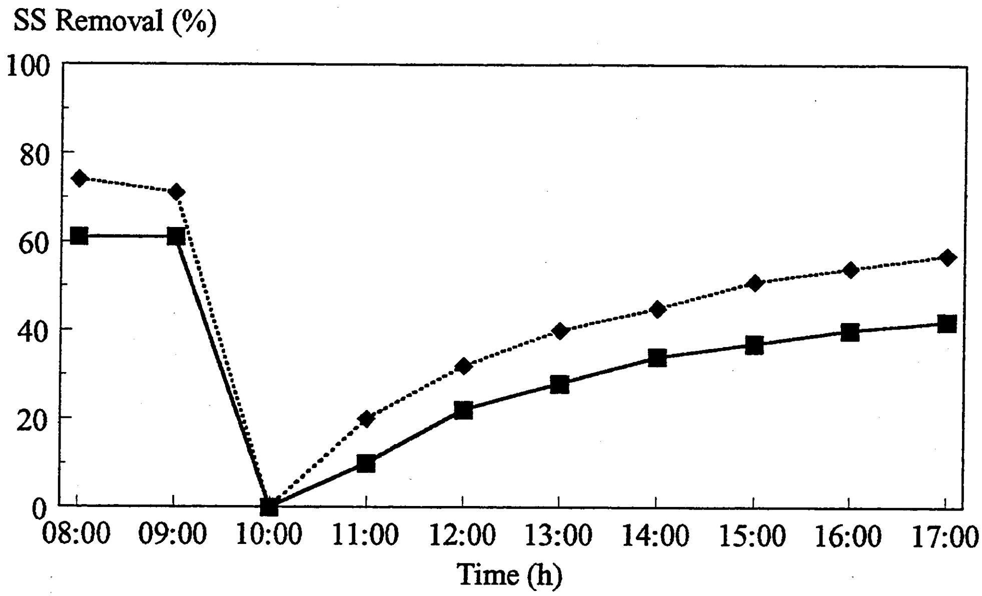

Figure 4a. Percentage Suspended Solids Removal During Flow Increase 0.3-0.4 $l\ min^{-1}$. (■-Sunken Media, ◆-Floating Media)

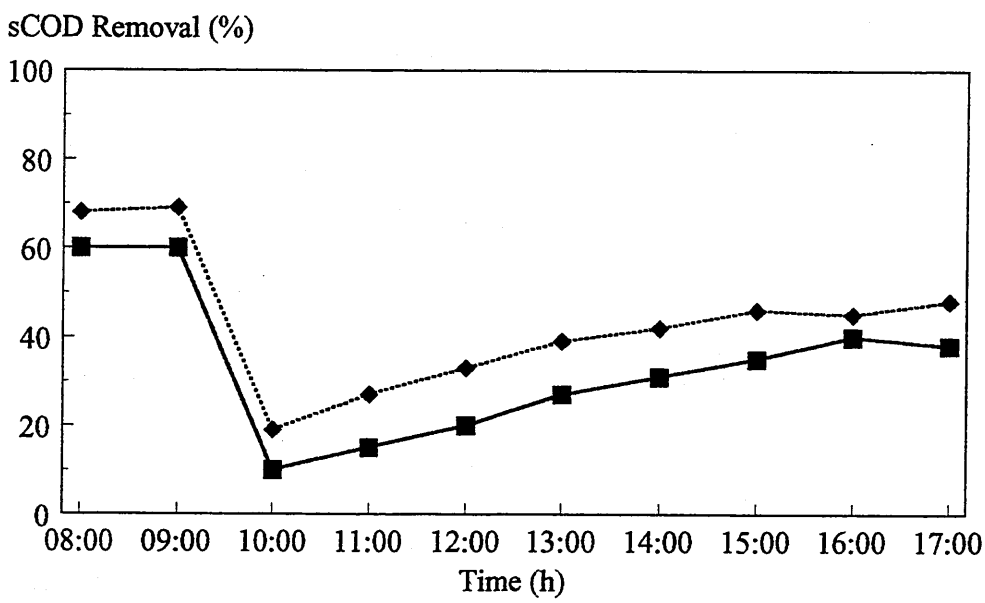

Figure 4b. Percentage Suspended Solids Removal During Flow Increase 0.3-0.4 $l\ min^{-1}$. (■-Sunken Media, ◆-Floating Media)

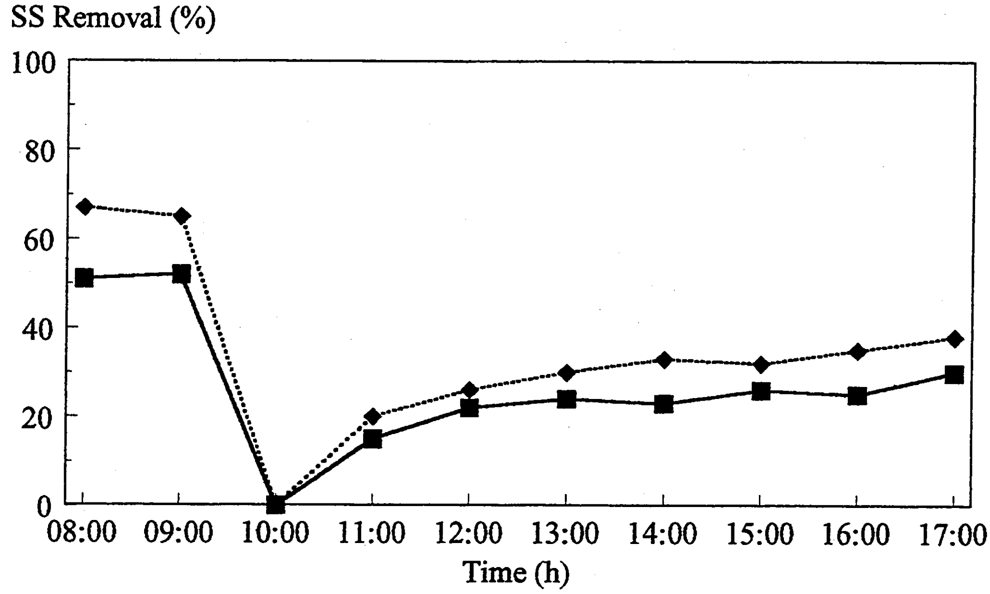

Figure 5a. Percentage Suspended Solids Removal During Flow Increase 0.4-0.5 l min^{-1}. (■ -Sunken Media, ◆ -Floating Media)

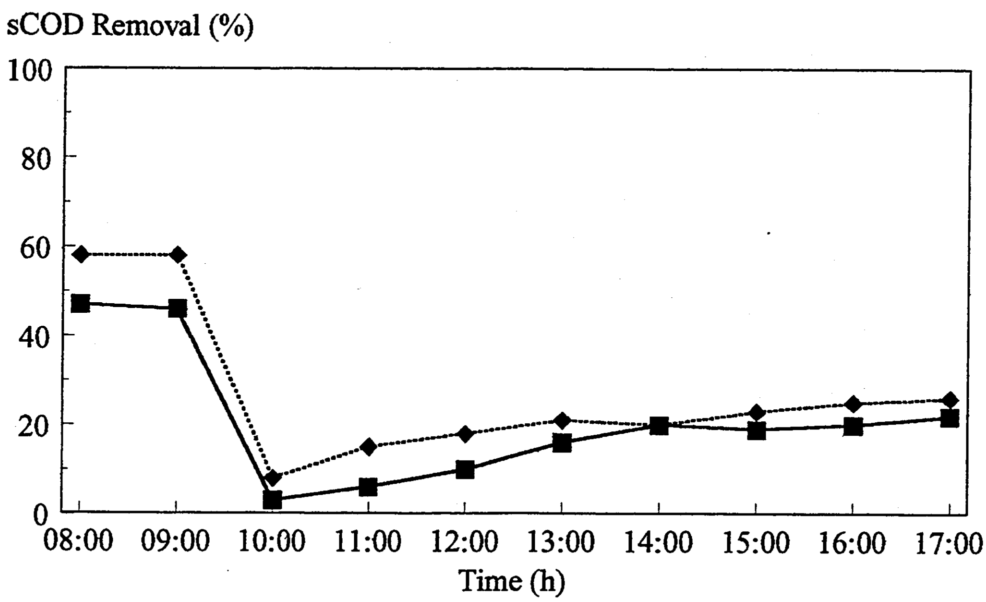

Figure 5b. Percentage Suspended Solids Removal During Flow Increase 0.4-0.5 l min^{-1}. (■ -Sunken Media, ◆ -Floating Media)

gradually rather than instantaneous at a single flowrate, though at a flowrate of 0.5 l/min, there was a marked decrease in all removal rates.

DISCUSSION

The method of start-up employed in fixed-film reactors affects their performance in different ways. Start-up of reactors using the process liquid at the nominal flowrate took longer than other methods (Bacquet *et al.*, 1991). But the higher shear produced caused fragmentary rather than mass detachment (Ohashi and Harada, 1994). Subsequently a more stable biomass was produced giving removal rates as high as 8 kg COD m^{-3} d^{-1} (80% COD removal) with backwashing having little effect on treatment efficiency (Bacquet *et al.*, 1991). This compared to maximum removal rates for activated sludge plants of 4.16 kg COD m^{-3} d^{-1} (Rusten and Thorvaldsen, 1983).

Using initial low flowrates during start-up caused thick biofilms to be produced reducing the stability of the biomass resulting in mass detachment (Ohashi and Harada, 1994). Ideally moderate shear is required to produce stable biofilm that is thin allowing improved mass transfer through reduced diffusion limitations (Lazarova and Manem, 1994). Seeding with activated sludge produced a high suspended biomass concentration which led to improved establishment of biomass within fixed-film reactors though an appropriate source of sludge was required (Faup *et al.*, 1982). From the results the use of activated sludge appeared to have produced a more rapid biofilm growth than that using the process liquid only. The biofilm also appeared to be more stable when seeding with activated sludge (Figure 1b), with less scattering of points which showed consistently improving sCOD removal rather than a large variance found when using process liquid only (Figure 2b).

The greater efficiency of the floating media at removing solids appeared to be due to compression of bed due to the buoyancy force of the media and flow of air and liquid acting upwards. On the other hand the negative buoyancy force of the sunken media acts downward thus opening up the bed. Greater mixing found in the clean sunken media reactor (Mann *et al*, 1995) indicated why initially better sCOD removal was seen in the reactor compared to the floating media reactor. However as the floating media reactor retained more solids which may have caused an increase in the mixing in the reactor, conditions improved for better biomass development. The low flowrates used during these experiments appeared to have encouraged the growth of unstable biomass through low shear (Ohashi and Harada, 1994). With increasing flow there was an increase in shear (Mann *et al*, 1995) which subsequently caused increased biofilm detachment (Zinebi *et al*, 1994). From this it can be concluded that there appeared to be little benefit in starting fixed film reactors using low flowrates and increasing the flow to the nominal value compared to starting at a high nominal flowrate. Though start-up using a high nominal flowrate takes longer, a stable biomass is produced (Bacquet *et al.*, 1991). Thus not only was biomass removed when liquid and air flowrates were increased but also these same conditions inhibited further growth of biomass using low flowrate start-up. This was indicated by a reduction in backwash solids produced as increased flowrates were used. Thus as flow was increased the was a subsequent decrease in sCOD removal.

Though reasonable organic shock loadings were attained, it appeared that it was the effect of the hydraulic shock loading that had the greater effect on the reactors. Fixed-film reactors have been found to be able to cope with organic shock loadings better than hydraulic shock loadings especially when using structured media due to the openness of the packing and the complete mixing of the liquid (Gray, 1993; Hamoda *et al.*, 1987). This apparent stability of the biomass has been attributed to the use of regular backwashing (Gray, 1993). The use of regular backwashing also improves biomass stability in granular media BAFs but due the added solids filtration capacity of BAF reactors any increase in the nominal flow causes an increase in the amount of mixing which dislodges these solids causing an increase in effluent solids concentration as in this study (Mann *et al.*, 1995; Tschui *et al.*, 1994).

Though the media used was not ideal due to its smooth surface the reactors performed well. At loadings of 0.57 kg sCOD m^{-3} d^{-1} (0.26 kg sBOD m^{-3} d^{-1} ,1.21 kg tCOD m^{-3} d^{-1} ,0.51 kg tBOD m^{-3} d^{-1}) the floating media attained 80% removal which compares well with that found for other processes (Gros, 1992; Rogalla *et al.*, 1991; Rusten, 1984; Smith *et al.*, 1990). The floating media performed better at the higher flowrates than the sunken media overall. This indicated that a floating media was more suited to upflow reactors.

Thus for fixed film reactors that are to be used over a long period of time start-up should be carried out using the nominal flowrate with activated sludge seeding as this ensures a stable biofilm that is resistant to high shear involved in shock loading, aids process recovery after backwashing (Bacquet *et al.*, 1991) and gives a thin biofilm for improved mass transfer (Lazarova and Manem, 1994).

ACKNOWLEDGEMENT

One of us (A. Mann) is grateful to the UK Engineering and Physical Research Sciences Council (EPSRC) for a quota research studentship.

REFERENCES

Abd-El-Bary, M.F., Ewayys, M.J. (1979). Biological Nitrification in Contact Aeration Systems. *Wat. Sew. Works*, 77, 91-93.

American Public Health Association. (1992). Standard Methods for the Examination of Water and Wastewater. 18th Edition, American Public Health Organisation, Washington, D.C.

Bacquet, G., Joret, J.C., Rogalla, F., Bourbigot, M.M. (1991). Biofilm Start-Up and Control in Aerated Biofilter. *Environ. Tech.*, 12, 747-756.

Canler, J.P., Perret, J.M. (1994). Biological Aerated Filters: Assessment of the Process Based on 12 Sewage Treatment Plants. *Wat. Sci. Tech.*, 29, 10/11, 13-22.

Chung, Y.C., Choi, Y.S. (1993). Microbial Activity and Performance of an Anaerobic Reactor Combining a Filter and a Sludge Bed. *Wat. Sci. Tech.*, 27, 1, 187-194.

Dillon, G.R., Thomas, V.K. (1990). A Pilot-Scale Evaluation of the 'Biocarbone Process' for the treatment of Settled Sewage and for Tertiary Nitrification of Secondary Effluent. *Wat. Sci. Tech.*, 22, 1/2, 305-316.

Gray, T.W. (1993). Biological Aerated Filters: The Solution to Small Foot Print Plant. *Int. Wat. Environ. Eng.*, Summer, 5-10.

Gros, H. (1992). BIOPUR Systems and Bioactive Filters for Advanced Waste-Water Treatment. *Sulzer Technical Review* 4/1991.

Faup, G.-M., Leprince, A., Pannier, M. (1982). Biological Nitrification in an Upflow Bed Reactor (UFBR). *Wat. Sci. Tech.* 14, Capetown, 795-810.

Gray, T.W. (1993). Biological Aerated Filters: The Solution to Small Footprint Plant. *Int. Wat. Environ. Eng.* Summer, 5-10.

Hamoda, M.F., Abd-El-Bary, M.F. (1987). Operating Characteristics of the Aerated Submerged Fixed-Film (ASFF) Bioreactor. *Wat. Res.*, 21, 939-947.

Hamoda, M.F., Al-Haddad, A.A., Abd-El-Bary, M.F. (1987). Treatment of Phenolic Wastes in an Aerated Submerged Fixed-Film (ASFF) Bioreactor. *J. Biotech.*, 5, 279-292.

Kiestra, H., Eggers, E. (1986). Treatment of Industrial Wastewaters. *Wat. Sci. Tech.*, 18, 3, 5-16.

Lang, J.S., Giron, J.J., Hansen, A.T., Tussel, R.R., Hodges Jr, W.E. (1993). Investigating Filter Performance as a Function of the Ratio of Filter Size to Media Size. *J. Amer. Wat. Works Assoc.*, Oct., 122-130.

Lazarova, V., Manem, J. (1994). Advances in Biofilm Aerobic Reactors Ensuring Effective Biofilm Activity Control. *Wat. Sci. Tech.*, 29, 10-11, 319-327.

Mann, A., Fitzpatrick, C.S.B., Stephenson, T. (1995). A Comparison of Floating and Sunken Media Biological Aerated Filters Using Tracer Study Techniques. *Trans. IChemE.*, 73, Pt. B, May, 137-143.

Ohashi, A., Harada, H. (1994). Characterisation o Detachment Mode of Biofilm Developed in an Attached-Growth Reactor. *Wat. Sci. Tech.*, 30, 11, 35-45.

Park, J-W., Ganczarczyk, J.J. (1994). Gravity Separation of Biomass Washed Out from Aerated Submerged Filter. *Env. Tech.*, 15, 945-955.

Rogalla, F., Payaudeau, M., Sauvegrain, P., Sibony, J. (1991). Reduced Hydraulic Detention Time for Complete Nutrient Removal with Innovative Biological Reactors. *Wat. Sci. Tech.*, 24, 10, 217-229.

Rusten, B. (1984). Wastewater Treatment with Aerated Submerged Biological Filters. *J.WPCF.*, 56, 422-425.

Ryhiner, G., Birou, B., Gros, H. (1992). The Use of Submerged Structured Packings in Biofilm Reactors for Wastewater Treatment. *Wat. Sci. Tech.*, 26, 3/4, 723-731.

Smith, A.J., Quinn, J.J., Hardy, P.J. (1990). The Development of an Aerated Filter Package Plant. *1st Int.Conf. Advances in Wat. Treat. and Environ. Man., Lyon, France 27-29 June.*

Stensel, H.D., Brenner, R.C., Lee, K.M., Melcer, H., Rakness, K. (1988). Biological Aerated Filter Evaluation. *J. Environ. Eng.*, 114, 655-671.

Tanaka, S., Matsuo, T. (1986). Treatment Characteristics of the Two Phase Anaerobic Digestion System Using an Upflow Filter. *Wat. Sci. Tech.*, 18, 7/8, 217-224.

Tschui, M., Boller, M., Gujer, W., Eugster, J., Mader C., Stengel, C. (1994). *Wat. Sci. Tech.*, 29, 10-11, 53-60.

Young, J.C., Baumann, E.R., Wall, D.J. (1975). Packed-Bed Reactors for Secondary Effluent BOD and Ammonia Removal. *J.WPCF.*, 47, 46-56.

Zinebi, S., Henriette, C., Petitdemange, E., Joret, J.C. (1994). Identification and Characterisation of Bacterial Activities Involved in Wastewater Treatment by Aerobic Fixed-Bed Reactor. *Wat. Res.*, 28, 2575-2582.

Chapter 5.

MODELLING BIOLOGICAL AERATED FILTERS FOR WASTEWATER TREATMENT.

Accepted for Publication,

Water Research.

MODELLING BIOLOGICAL AERATED FILTERS FOR WASTEWATER TREATMENT.

Allan Mann and Tom Stephenson.

School of Water Sciences, Cranfield University, Cranfield, Bedford. MK43 0AL, UK.

ABSTRACT

Biological aerated filters (BAFs) are submerged three-phase fixed media reactors for wastewater treatment. A major characteristic of BAF reactors is the use of granular media which allows solids separation as well as secondary or tertiary biological treatment in one unit. The aim of this work was to design a simple empirical model relating influent soluble chemical oxygen demand (sCOD) to effluent sCOD and reactor height, verify experimentally the suitability of this model and ascertain the relationship between model constants and reactor performance. The theoretical model was based on that designed for trickling filters as both fixed-film processes show a similar plug flow pattern.

Two reactors were set-up to run in parallel treating settled domestic sewage using media identical in size and shape except one was less dense than water (relative density 0.92) and the other denser than water (relative density 1.05). The reactors were run upflow with liquid flowrates of 0.29 to 0.58 m3 d-1 (0.2 to 0.5 l min^{-1}) and an air : liquid ratio of 10 : 1. After 4 weeks from start-up steady-state was reached. From this point, samples were taken at different heights along the reactors at timed intervals and profiles of sCOD removal against reactor height were produced. This analysis was repeated for a number of different flowrates and organic loadings (0.57 to 1.40 kg sCOD m^{-3} d^{-1}). The resulting data was then used with the empirical model, which was based on a first order reaction, to calculate the values of k* (overall process constant) and n (media factor). A higher value of k* was found for the floating media (55) compared with the value found for the sunken media (33). This indicated the greater efficiency of sCOD removal in the floating media. The values of the media constants were similar showing the similarity in the media shape and size.

KEYWORDS

Biological aerated filter; chemical oxygen demand; empirical model; wastewater treatment.

INTRODUCTION

Biological processes for wastewater treatment such as trickling filters and activated sludge plants have been used since the late 19th and early 20th centuries and are well established (Grady, 1983). During the late 1970's and 1980's a further process was developed, the biological aerated filter (BAF) (Legise *et al.*, 1980). Unlike other fixed film processes BAFs are able to carry out solids separation as well as aerobic biological treatment. Thus carbonaceous removal, solids filtration and nitrification can be carried out in a single unit (Stephenson *et al.*, 1993). Modification of reactors utilising an anaerobic zone also allows denitrification and phosphate removal (Rogalla *et al.*, 1991).

The origins of the submerged aerated biological reactor date back to the beginning of the century with the development of aerated reactors using submerged layers of slate followed by a secondary clarifier (Clark, 1930). These were known as submerged contact aerators and similar designs with randomly packed or structured plastic media are still used today. Several years later, in Germany, the Emscher or Tank Filter was introduced. The Emscher Filter used the higher surface area of coarse slag to promote a greater biomass concentration and thus give better treatment (Bach, 1937). The use of such reactors was limited due to the relative success of the activated sludge process. Using granular media BAFs, appeared in the late 1970's as a viable alternative to other biological wastewater treatment processes. The small footprint size, ease of construction and the ability to treat high organic loads have made the BAF an ideal alternative for upgrading existing wastewater treatment works (Stephenson *et al.*, 1993).

Modelling of biological fixed films can be considered in two categories; mechanistic and empirical models (Grady, 1983; Tchobanoglous and Schroeder, 1985). Mechanistic models deal with individual mechanistic phenomena with the application of kinetics and other reactor engineering principles (Williamson and McCarty, 1976; Rittmann and McCarty, 1980; Avaev, 1986; Logan *et al.*, 1987b; Capdeville and Nguyen, 1989). These models tend to look at processes on a micro scale, for example transport of nutrients across a biofilm and the use of the Monod equation (Arvin and Harremoes, 1989). Although mechanistic models can accurately describe bioreactor behaviour, being based on fundamental principles, they can require measurement of parameters not usually available to treatment plant designers and operators. Due to the many variables involved in the analysis of granular media fixed-film reactor performance, mechanistic modelling of such reactors is extremely complex. Many of these parameters are dependant on each other and subsequently must be included for a realistic model to be designed (Sarner, 1989).

To simplify mechanistic models some parameters are assumed to be constant. The most important and hardest parameter to account for in mechanistic modelling is the biofilm thickness and distribution. Unfortunately biofilm thickness is difficult to measure and in many models the distribution of the biofilm has been assumed to be uniform, though this has been found not to be true (Stensel and Lee, 1989; Hamoda, 1989; Lewandowski *et al.*, 1994). In many models ideal plug flow has also been assumed. Though plug flow is prevalent, it has been found to be non-ideal with increased mixing and dispersion at high flowrates (Mann *et al.*, 1995; Muslu, 1990). This variation in hydraulic dispersion and liquid velocity alters the hydrodynamic conditions within granular reactors, subsequently affecting the biological performance (Jennings *et al.*, 1976: Grasmick *et al.*, 1980; Muslu, 1990). Another important factor not always accounted for, that affects both mechanistic and empirical models, is the rate of aeration. Variation in aeration rate alters hydraulic flow as well as oxygen transfer efficiency through direct contact with the biofilm and dissolution into the bulk liquid. Thus varying the aeration rate affects oxygen transfer, flow conditions and shear (Lee and Stensel, 1986; Mann *et al.*, 1995).

In contrast to mechanistic models, empirical models merely relate input and output variables of a process to one another, i. e. relates hydraulic or organic loading rates to nutrient removal rate and does not require the measurement of parameters such as oxygen utilisation rate and mass transfer coefficients (Hamoda, 1989; Meunier and Williamson, 1981; Wu *et al.*, 1983). By measuring input and output variables the assumptions that uniformity of biomass distribution and ideal plug flow do not need to be made (Eckenfelder and Barnhart, 1963; Wu *et al.*, 1983).

Previous models for BAF type reactors have been mechanistic (Jennings *et al.*, 1976; Hamoda, 1989). The aim of this study was to produce a model of an empirical type relating influent soluble chemical oxygen demand (sCOD) to effluent sCOD and reactor height for specific medias in BAFs. This could act as a simple tool for design, selection and sizing of BAFs without having to develop a complex mechanistic model.

THEORY

Biological aerated filters and trickling filters show similar characteristics despite the fact that trickling filters are not submerged. Tracer studies carried out on trickling filters showed that the liquid undergoes plug flow movement through the beds (Sinkoff *et al.*, 1959). Similarly, tracer studies carried out to ascertain the hydrodynamic characteristics of BAF reactors showed a plug flow profile. These tracer studies were carried out on reactors identical to those used in these biological experiments. The results showed that more mixing and liquid bypassing occurred in the upflow reactor using sunken media than the upflow reactor using floating media. This indicated that biologically the floating media would show a better performance than the sunken media (Mann *et al.*, 1995). Thus the approach taken to design a model for BAF performance was similar to that used for trickling filters (Velz, 1948; Schulze, 1960; Eckenfelder *et al.*, 1961, 1963; Galler *et al.*, 1964; Logan *et al.*, 1987b). Following the design of a basic model, results obtained during biological treatment were used to examine the application of the model as in trickling filter studies (Harrison and Daigger, 1987; Logan *et al.*, 1987a; Matasci *et al.*, 1986).

The overall reaction rate for substrate utilisation by a biofilm can be described as first order. Thus

$$\frac{dC}{dt} = kC \tag{1}$$

where C is the sCOD concentration and k the rate constant at a specific time and sCOD concentration.

This integrates to

$$\frac{C}{C_i} = e^{-kt} \tag{2}$$

where C_i is the influent sCOD concentration.
t though can be related to the volumetric loading, thus

$$t = \frac{k'}{B_V} \tag{3}$$

where k' is the biomass constant which depends on the hydraulic characteristics of the reactor and B_V the volumetric loading
The volumetric loading can be re-written,

$$B_V = \frac{QC_i}{HA} \tag{4}$$

where Q is the volumetric flowrate, H the height and A the cross sectional area of the reactor.
Thus

$$t = \frac{k'HA}{QC_i} \tag{5}$$

and

$$\frac{C}{C_i} = e^{-mH} \text{ where } m = \frac{k^*A}{QC_i^n} \tag{6}$$

where n is a constant dependant on the media type, thus

$$\ln\left\{\frac{C}{C_I}\right\} = \frac{-k^*A}{QC_i^n}H \tag{7}$$

Plotting sCOD removal (C/C_i) against reactor height (H) gives a typical first order curve. By plotting $\ln(C/C_i)$ against H, for individual runs a series of values of the slope (-m) can be obtained for different values of C_i. From the plot of ln (m) against ln (C_i) the values of the slope (n) and y-axis intercept (k^*) can be found for each reactor.

MATERIALS AND METHODS

Two identical polyvinylchloride (PVC) reactors were built 2m in height with a diameter of 0.2m (Figure 1). This diameter was designed to be approximately 50 mean particle diameters in size to limit wall affects thus allowing the results produced to be applied to larger scale reactors (Lang *et al.*, 1993). The reactors were run with settled domestic sewage introduced at the bottom through a centrally placed 5 cm filter nozzle (Degremont). Aeration was carried out using three identical nozzles placed around the base plate. In addition sample ports were placed along the length of the reactors at 0.2m intervals. The medias were kept in place using a rigid polypropylene mesh with 1.5mm diameter holes, placed at the top of the reactors using clamps (Mann *et al.*, 1995).

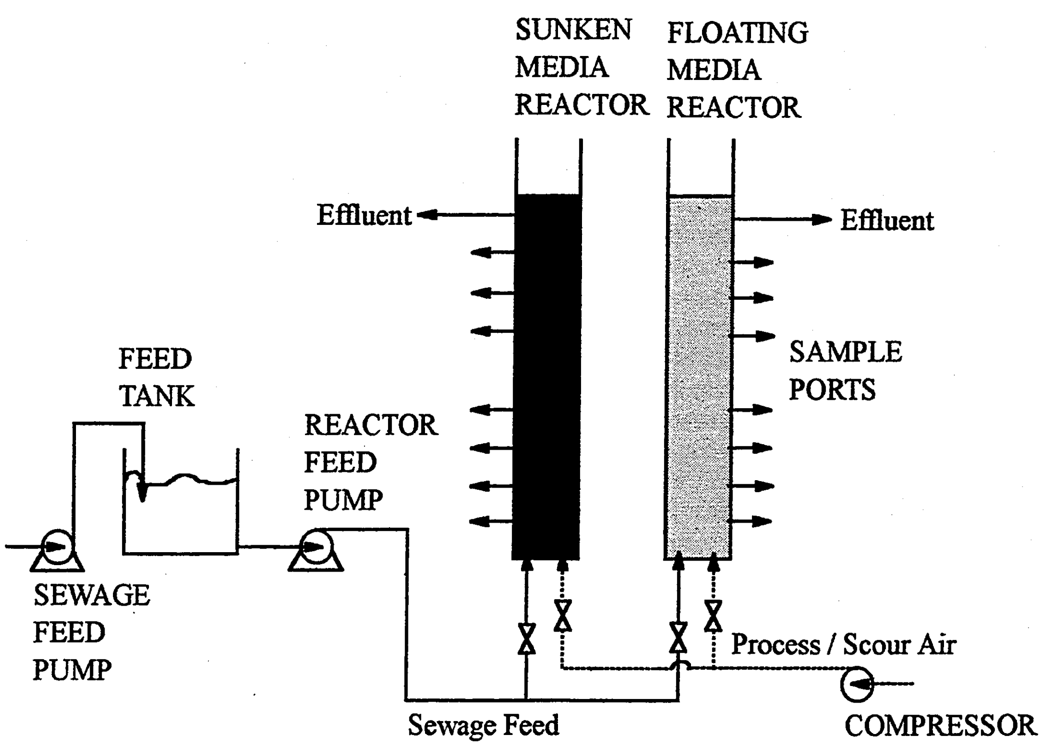

Figure 1. Experimental Set-Up.

TABLE 1. Media Specifications.

Media Shape	Cylindrical
Media Size : Diameter, mm	2.3 - 2.7
Length, mm	4 - 6
Media Relative Density : Sunken	1.05
Floating	0.92
Media Material : Sunken	Polypropylene / Calcium Carbonate
Floating	Polyprpylene
Voidage, %	42
Specific Surface Area, $m^2\ m^{-3}$	1,160

The two medias used were made of recycled polypropylene (Cookson plc, Cheshire, U.K.), identical in shape and size except that one was made of pure polypropylene, floating, with a relative density of 0.92 and the other, sunken media, contained a mixture of polypropylene (60%) and calcium carbonate (40%) (Britomya V, Croxton and Gary, Surrey, U.K.), with a relative density of 1.05 (Table 1).

Settled domestic sewage from Cranfield University's sewage treatment works was passed upflow through the reactors. Initially a liquid flowrate of 0.2 l/min was used and an air : liquid ratio of 10 : 1. After an initial start up period of 4 weeks the reactors were run at steady state for a period of time with backwashing taking place on a daily basis to remove excess biomass. Steady state was reached when the reactors achieved a constant sCOD rate of removal (+/- 5%) over a five day period. The flowrate was then increased to 0.3 l/min followed by a further steady state period and then increased again to 0.4 l/min with a subsequent steady state period, each time maintaining an air : liquid ratio of 10 : 1. Throughout, the liquid temperature was maintained at 16.5 - 17.5 °C using a heat exchanger and by lagging the reactors. Sampling took place each morning over this steady-state period, approximately 1 to 4 h prior to the daily backwashing.

To obtain sCOD profiles of the reactors, samples were taken at staggered time intervals from the influent pipe, sample ports and effluent pipe. These samples were filtered and the sCOD of the filtrate was then measured (APHA, 1992). In addition other analysis carried out throughout the work included, total and soluble biochemical oxygen demand (sBOD, tBOD), suspended solids, dissolved oxygen, pH, ammonia and temperature. Nitrate and nitrite values were measured at intervals but little removal was seen and subsequently analysis was made only infrequently.

TABLE 2. Influent and Effluent Process Conditions.

	Influent		Effluent (Sunken Media)		Effluent (Floating Media)
	Mean	Range	Mean	Range	Mean
Flowrate, Q ($m^3\ d^{-1}$)	-	0.29 - 0.58	-	0.29 - 0.58	-
Suspended Solids, ($mg\ l^{-1}$)	102	80 - 120	38	19 - 104	24
Soluble COD, ($mg\ l^{-1}$)	150	80 - 210	48	16 - 148	32
Total COD, ($mg\ l^{-1}$)	292	128 - 520	119	52 - 234	72
Ammonia, ($mg\ l^{-1}$)	21.2	15.2 - 30.6	19.1	14.4 - 30.4	17.9
Dissolved Oxygen, ($mg\ l^{-1}$)	1.1	0.8 - 1.6	7.6	5.5 - 8.6	7.5
pH	7.2	6.4 - 8.1	7	6.2 -7.9	7

RESULTS AND DISCUSSION

Start-up took four weeks of the total experimental period of 6 months. A maximum sCOD removal rate of approximately 90% was reached which dropped until steady state values of 68% sCOD removal for the sunken media and 75% sCOD removal for the floating media were attained. Over the steady state experimental period of 3 months these values varied from 60% to 75% for the sunken media and 71% to 85% for the floating media at the initial flowrate of 0.2 $l\ min^{-1}$. At increased flowrates the mean sCOD removal rate at steady state dropped, with only 36% sCOD removal for the sunken media and 40% sCOD removal for the floating media at a flowrate of 0.5 $l\ min^{-1}$. Over the experimental period influent conditions were maintained over a narrow range (Table 2). Effluent dissolved oxygen (DO) values remained at between 5.2 and 8.6 $mg\ l^{-1}$ ensuring oxygen availability did not become a limiting factor. Ammonia removal rates were low throughout indicating limited nitrification occurring. Though the model used was based on the sCOD values, the sBOD was also monitored over the experimental period. The sBOD : sCOD ratio was found to be 0.45 for the influent and 0.20 for the effluent thus allowing the model constants to be found, if required, in terms of sBOD removal.

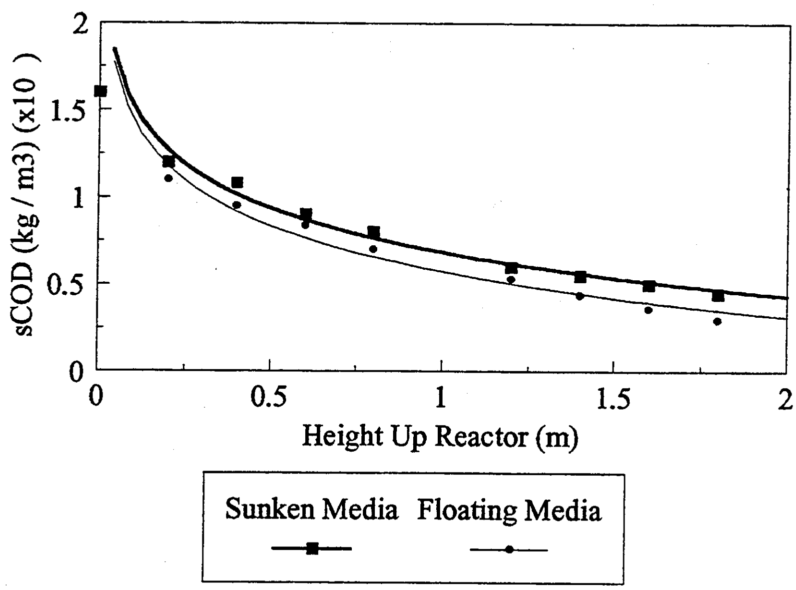

Figure 2. sCOD Removal Profile,
(Experimental Results, $C_i = 0.160$ kg m^{-3}).

Figure 2 shows a typical sCOD concentration profile obtained. To calculate values of n and k* nine sets of results were used with R^2 values between 0.978 and 0.993. R^2

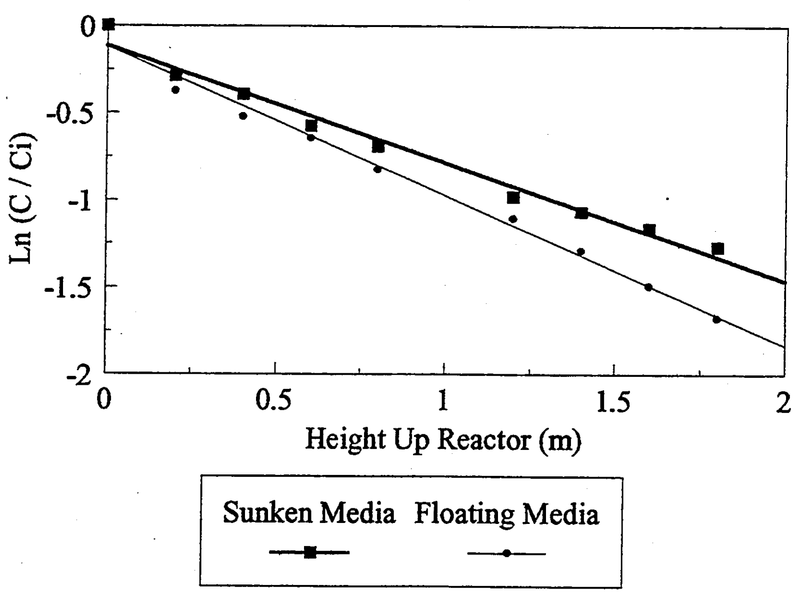

Figure 3. Logrithmic sCOD Removal Profile,
(Experimental Results, $C_i = 0.160$ kg m^{-3}).

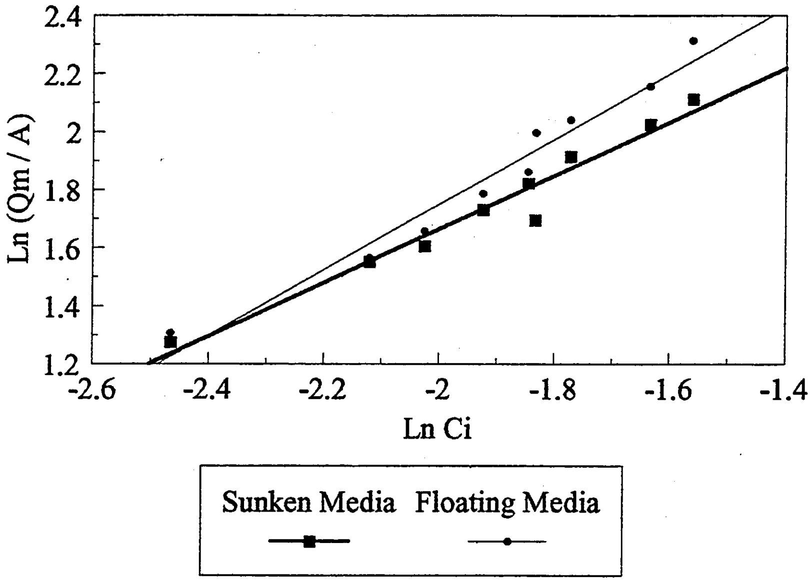

Figure 4. Calculation of n and k.*

indicates variation between the data points (dependant variables) and the regression or best fit line (independent variables). Values of R^2 range between 0 and 1 where values close to 1 indicate a close correlation between the points and the regression line. It was demonstrated that the greatest removal rate occurred at the base of the reactors then decreased exponentially as the liquid moved up the reactors and the sCOD became a limiting factor. The subsequent logarithmic graph of the same values (Figure 3) showed a close approximation to the 1st order equation (Equation 2). This corresponds to the current knowledge that in aerobic fixed film reactors the overall rate of reaction is 1st order (Grady, 1983; Capdeville and Nguyen, 1989; Muslu, 1990). R^2 values for the logarithmic graphs varied from 0.981 to 0.995. This confirmed that the initial part of the model could be applied to this process. Analysis made at different organic loadings were then used to find the constants n and k^* (Figure 4) with R^2 values of 0.958 for the sunken media line and 0.964 for the floating media line.

The results from the model (Table 3) showed the difference in the constants between the two reactors.

TABLE 3. Constant Results.

	k^*	n
Sunken Media	33	0.92
Floating Media	55	1.13

The k^* constant indicated the overall performance of the reactors in removing sCOD and controls the rate of removal over the length of the reactor at each influent sCOD concentration. The greater the value of k^* the lower the C/C_i values resulting in higher overall sCOD removal and a greater rate of removal. The difference in values of k^* found from the experimental results indicated a better performance for the floating media than for the sunken media, which was confirmed by sCOD removal values monitored daily over the experimental period. In the case of the media constant, n, similar values were obtained which indicated the similarity in the media designs. The value of n indicates the variation in the removal rate of sCOD over a range of influent sCOD concentrations and thus indicates the stability of the process. Thus reactors with low values of n would show a great variation in sCOD removal rates over a small range of influent sCOD concentrations. Subsequently these reactors would be unable to sufficiently treat highly variable wastewater concentrations or shock loadings. Removal of sCOD fitted the model well and over the period of 6 to 24 h from backwashing there was only a slight variation in sCOD removal rates. After backwashing the effluent solids were high, often greater than the influent concentration, due to dislodged solids leaving the reactors. After this period the effluent solids concentration dropped as biomass growth caused increased solids capture and the remaining dislodged solids left the reactors in the effluent. This continued until the concentration began to increase again, prior to backwashing, through sloughing of biomass (Lewandowski *et al.*, 1994; Cao and Alaerts, 1995). The mechanics of particulate matter removal differ from those of dissolved organic matter with indications that the two interact through particulate adsorption hindering the dissolved organic removal rate (Sarner, 1986). A great variation in the effluent solids over a 24 h period and only slight variation in the sCOD removal rate indicated that the suspended solids concentration did not affect sCOD removal efficiency. Thus no correction to the model was deemed necessary.

The value of k^* has a direct relationship to the removal of soluble COD from the process liquid and is thus an indicator of the active biofilm concentration within the reactor (Capdeville and Nguyen, 1989; Grady, 1983; Hoehn and Ray, 1973). There are a number of physical factors that influence biofilm growth which are in part dictated by media properties. The size, shape and density influence the flow through the packed bed and resulting shear on the biofilm. These factors along with surface roughness also dictate the surface area available for biofilm attachment. Flow direction and aeration rate are factors that also influence biofilm growth. As well as causing scouring of the biofilm, aeration, along with the flow direction, effects the compactness of the bed and thus the liquid flow passing through it (Cao and Alaerts, 1995; Hoehn and Ray, 1973; Logan *et al*, 1987a).

The difference in media types is shown by the value of n, the media constant. The values found for the sunken and floating media were similar (0.92 and 1.13) which would be expected for media similar in shape and surface texture. Since the only variation between the medias was the density, the difference in n values may be attributed to this and its

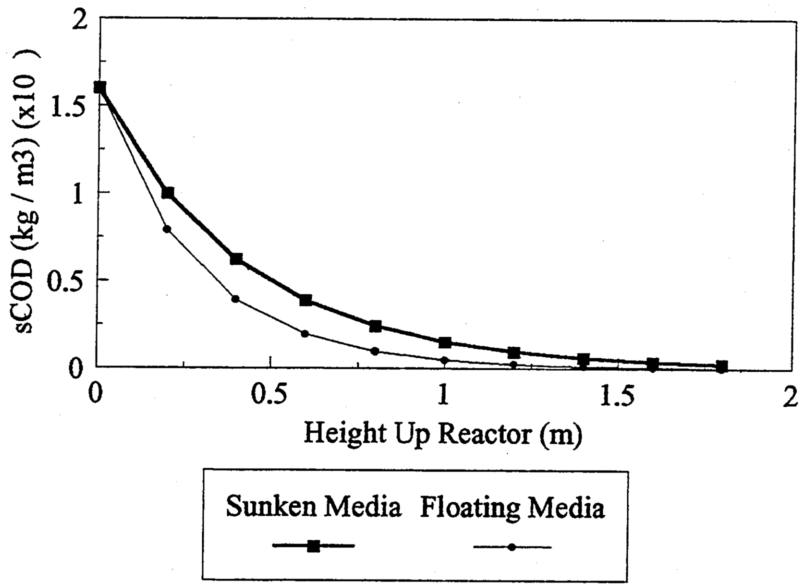

Figure 5. Model Prediction, C_i = 0.160 kg m^{-3}.

consequential effects on the flow through the reactors with the subsequent effect on biofilm attachment.

With the introduction of the process constants into the empirical equation (Equation 6) the subsequent graph (Figure 5) showed profiles similar to those found experimentally. The differences between predicted (Figure 5) and experimental (Figure 2) graphs indicated that the values of n and k* were too high possibly due to the narrow sCOD range , 80 to 200 mg l^{-1} (0.008 to 0.2 kg m^{-3}), over which the constant values were calculated.

Simple empirical models such as this do not require the kinetic knowledge required for mechanistic models and take into account the effects of large scale factors such as aeration rate and liquid velocity. Any attempt to accurately apply mechanistic models to such systems would involve a great number of variables requiring experimental confirmation. The use of this model thus allows a direct comparison between media performance and gives a simple method to ascertain the ideal media type to be used for a process and minimum height needed for a required discharge concent. The values of k* and n are specific to the media, under specific conditions of flow velocity and direction, temperature and aeration rate. Work on different granular media types give an improved understanding of the effect that shape, size and surface texture have on the values of the constants k^* and n and on the comparative performance (Eckenfelder and Barnhart, 1963; Harrison and Daigger, 1987). In addition constants could also be found for the media types under different process conditions such as downflow and with different aeration rates as this affects the amount of mixing within granular reactors and subsequently the performance in wastewater treatment (Mann *et al.*, 1995). Thus these experiments are required to obtain results that may be used to predict the ideal media required for individual plants. They would also allow the performance of media types to be predicted purely on their physical properties (Eckenfelder, 1961).

NOMENCLATURE

C	sCOD concentration	kg m^{-3}
C_i	Influent sCOD concentration	kg m^{-3}
k	Reaction rate constant	d^{-1}
t	Time	d
k'	Biomass constant	kg m^{-3}
B_V	Volumetric loading	kg m^{-3} d^{-1}
Q	Volumetric flowrate	m^{3} d^{-1}
H	Height up reactor	m
A	Cross-sectional area of reactor	m^{2}
k^*	Overall process constant	kg m^{-3} d
n	Media constant	-

ACKNOWLEDGEMENT

One of us (A.Mann) is grateful to the UK Engineering and Physical Research Sciences Council (EPSRC) for support.

REFERENCES

American Public Health Association. (1992). Standard Methods for the Examination of Water and Wastewater. 18th Edition, American Public health Organisation, Washington, D.C.

Arvin, E., Harremoes, P. (1989). Concepts and Models for Biofilm Reactor Performance. *Technical Advances in Biofilm Reactors.* Nice, 4-6 April, 191-212.

Avaev, A.A. (1985). Modelling Mass Exchange in the Treatment of Fluid in a Biological Filter. *Soviet J. Wat. Chem. Tech.*, 7, 7-9.

Bach , H. (1937). The Tank Filter, For the Purification of Sewage and Trade Wastes. *Wat. Works Sew.*, Oct., 389-393.

Cao, Y.S., Alaerts, G.J. (1995). Influence of Reactor Type and Shear Stress on Aerobic Biofilm Morphology, Population and Kinetics. *Wat. Res.*, 29, 107-118.

Capdeville, B., Nguyen, K.M. (1989). Kinetics and Modelling of Aerobic and Anaerobic Film Growth. *Wat. Sci. Tech.*, 22, 1/2, 149-170.

Clark, H.W. (1930). Past and Present Developments in Sewage Disposal and Purification. *Sew. Works Jour.* Oct., 561-571.

Eckenfelder, Jr., W.W. (1961). Trickling Filter Design and Performance. *J. San. Eng. Div.*, July, 33-45.

Eckenfelder, Jr., W.W., Barnhart, E.E. (1963). Performance of a High Rate Trickling Filter Using Selected Media. *J. Wat. Pollut. Control Fed.*, 35, 1535-1551.

Galler, W.S., Gotaas, H.B. (1964). Analysis of Biological Filter Variables. *J. San. Eng. Div.*, Dec., 59-79.

Grady, C.P.L. (1983). Modelling of Biological Fixed Films - A State-of-the-Art Review. In: Fixed-Film Biological Processes for Wastewater Treatment. Y. C. Wu and Smith E. D. (Ed.), Noyes Data Corporation, New Jersey, pp. 75-134.

Grasmick, A., Elmalen, S., Ben-Aim, R. (1980). Theory of Treatment by Submerged Biological Filtration. *Wat. Res.*, 13, 1137-1147.

Hamoda, M.F. (1989). Kinetic Analysis of Aerated Submerged Fixed-Film (ASFF) Bioreactors. *Wat. Res.*, 23, 1147-1154.

Harrison, J.R., Daigger, G.T. (1987) A Comparison of Trickling Filter Media. *J. Wat. Pollut. Control Fed.*, 59, 679-685.

Hoehn R.C., Ray, A.D. (1973). Effects of Thickness on Bacterial Film. *J. Wat. Pollut. Control Fed.*, 45, 11, 2302-2320.

Jennings, P.A., Snoeyink, V.L., Chian, E.V.K. (1976). Theoretical Model for a Submerged Biological Filter. *Biotech. Bioeng.*, 18, 1249-1273.

Lang, J.S., Giron, J.J., Hansen A.T., Tussel, R.R., Hodges Jr, W.E. (1993). Investigating Filter Performance as a Function of the Ratio of Filter Size to Media Size. *J. Amer. Wat. Works Assoc.*, Oct., 122-130.

Lee, K.M., Stensel, H.D. (1986). Aeration and Substrate Utilisation in a Sparged Packed-Bed Biofilm Reactor. *J. Wat. Pollut Control Fed.*, 58, 165-172.

Legise, J.P., Gilles, P., Mureaud, H. (1980). A New Development in the Biological Aerated Filter Bed Technology. *Presented at the 53rd Annual Wat. Pollut. Control Fed. Conference*, Las Vegas, Nevada.

Lewandowski, Z., Stoodley, P., Altobelli, S., Fukushima, E. (1994). Hydrodynamics and Kinetics in Biofilm Systems - Recent Advances and New Problems. *Wat. Sci. Tech.*, 29, 10-11, 223-229.

Logan, B.E., Hermanovicz, S.W., Parker, D.S. (1987a). Engineering Implications of a New Trickling Filter Model. J. *Wat. Pollut. Control Fed.*, 59, 1017-1028.

Logan, B.E., Hermanovicz, S.W., Parker, D.S. (1987b). A Fundamental Model for Trickling Filter Design. *J. Wat. Pollut. Control Fed.*, 59, 1029-1042.

Mann A., Fitzpatrick C.S.B., Stephenson T. (1995). A Comparison of Floating and Sunken Media Biological Aerated Filters Using Tracer Study Techniques. *Trans. I.Chem.E. Pt. B*, 73, 137-143.

Matasci, R.N., Kaempfer, C., Heidman, J.A. (1986). Full Scale Studies of the Trickling Filter / Solids Contact Process. *J. Wat. Pollut. Control Fed.*, 58, 1043-1049.

Meunier, A.D., Williamson, K.J. (1981). Packed Bed Biofilm Reactors: Simplified Model. *J. Environ. Eng. Div.*, 107, 307-317.
Muslu, Y. (1990). Use of Dispersed Flow Models in Design of Biofilm Reactors. *Wat. Air Soil Pollut.*, 53, 297-314.
Rittmann, B.E., McCarty P.L. (1980). Design of Fixed-Film Processes with Steady-State Biofilm Model. *Prog. Wat. Tech.*, 12, (Toronto), 271-281.
Rogalla, F.., Payaudeau, M., Sauvegrain, P., Sibony, J. (1991) Reduced Hydraulic Detention Time for Complete Nutrient Removal with Innovative Biological reactors. *Wat. Sci. Tech.*, 24, 10, 217-229.
Sarner, E. (1989) Removal of Particulate and Dissolved Organics in Aerobic Fixed-Film Biological Processes. *J. Wat. Pollut. Control Fed.*, 58, 165-172.
Schulze, K.L. (1960). Load and Efficiency of Trickling Filters. *J. Wat. Pollut. Control Fed.*, 32, 3, 245-261.
Sinkoff, M.D., Porges, R., McDermott, J.H. (1959). Mean Residence Time of a Liquid in a Trickling Filter. *Jour. San. Eng Div., Amer. Soc. Civil. Engr.*, 85, SA 6, 51-60.
Stephenson, T., Mann, A., Upton, J. (1993). The Small Footprint Wastewater Treatment Process. *Chem. Ind.*, 533-536.
Tchobanoglous, G., Schroeder, E.D. (1985). Water Quality: Charateristics, Modelling, Modification. Addison - Wesley, Reading, MA.
Velz, C. J. (1948). A Basic Law for the Performance of Biological Filters. Sew. Works Jour., 20, 607-617.
Williamson, K., McCarty, P. L. (1976). Verification Studies of the Biofilm Model for Bacterial Substrate Utilization. *J. Wat. Pollut. Control Fed.*, 48, 281-296.
Wu, Y.C., Wilson, R.C., Beckman, W.R. (1983). Modelling of Plastic Media Trickling Filters. In: Fixed-Film Biological Processes for Wastewater Treatment. Y.C. Wu and Smith E. D. (Ed.), Noyes

Chapter 6.

A COMPARISON OF FLOATING AND SUNKEN MEDIA BIOLOGICAL AERATED FILTERS FOR NITRIFICATION.

Submitted for Publication,

Journal of Chemical Technology and Biotechnology.

A COMPARISON OF FLOATING AND SUNKEN MEDIA BIOLOGICAL AERATED FILTERS FOR NITRIFICATION.

Allan Mann and Tom Stephenson.

School of Water Sciences, Cranfield University, Cranfield, Bedford. MK43 0AL, UK.

ABSTRACT

The versatility of the biological aerated filter (BAF) has made it an important process in wastewater treatment. These submerged three-phase fixed media reactors have been used in a wide variety of applications in wastewater treatment, such as primary treatment (solids removal) secondary treatment (COD and BOD removal) and tertiary treatment (nitrification).

The aim of this work was to investigate the biological start-up of two such reactors to remove ammonia (nitrification), one containing a sunken media (relative density 1.05) and the other containing a floating media (relative density 0.92) both identical in shape and size. These upflow reactors with a media volume of 0.054 m^3 were run in parallel using secondary effluent as the process liquid at a flowrate of 0.2 l min^{-1} and air : liquid ratio of 10 : 1.

The two factors that primarily affected performance were nutrient concentration in the process liquid and the backwashing frequency. Initial low ammonia levels (< 5 mgl^{-1}) resulted in poor start-up. With increased ammonia concentrations (20 mgl^{-1}) performance improved. The results also indicated that a less frequent backwashing regime would improve start-up of nitrification.

KEYWORDS

Biological aerated filter; nitrification; plastic media; backwashing.

INTRODUCTION

The biological aerated filter has been developed over the last twenty years and has become an economical alternative to other secondary systems for treating wastewater due to its small footprint size and high loading capability (Legise *et a*l., 1980). This treatment process was initially developed as a secondary process but later developed for tertiary treatment.

Many factors affect the start-up and performance of fixed-film nitrifying reactors. As well biological factors such as the establishment of predators (rotifers), the hydraulic residence time and hydrodynamics of reactors affect both biofilm growth and nutrient removal (Lee and Welander, 1994). The hydrodynamics of reactors vary with the media used, liquid flowrate and aeration rates (Mann *et al.*, 1995). The type of biofilm growth and surface attachment through slow growth rate varies with different turbulent conditions (Rogalla *et al.*, 1990b). Increased turbulence causes reduced biomass concentrations within reactors through shear but improves substrate flux when substrate concentrations are non-limiting (Kugaprasatham *et al.*, 1992). The hydrodynamics of reactors may vary over time through build up of filtered solids. This causes variable conditions within reactors which may inhibit nitrification efficiency (Murphy *et al.*, 1977). The residence time of liquids within reactors is primarily dependant on two factors, the temperature the process liquid and the concentration of the nutrients in the influent to be treated (McCarty and Haug, 1971).

Performance of many biological systems is dependant on temperature (Fdez-Polanco *et al.*, 1994). Variable temperatures though have been found to have little effect on the performance of BAF reactors except at extreme temperatures (Iida and Teranishi, 1984; Paffoni *et al.*, 1990; Sakuma *et al.*, 1993). Performance at low temperatures has been improved by simply increasing the retention time (McCarty and Haug, 1971; Harremoes, 1982; Fdz-Polanco *et al.*, 1994)

Substrate concentrations within the influent wastewater affect the nitrification rate and biomass growth within BAF reactors (Goncalves *et al.*, 1994). Chemical substances such as fluoride inhibit the nitrification process non-competitively (Beg and Hassan, 1987). Competitive inhibition occurs through the growth of carbonaceous removing bacteria which is promoted through high influent BOD concentrations though some BOD is required for nitrifying biofilm assimilation (Rogalla and Payraudeau, 1988). High BOD also utilises a large part of the oxygen available to the nitrfiers which require high oxygen concentrations to metabolise (Haug and McCarty, 1972 Boller *et al.*, 1994). Low solids concentrations are also required as these also promote carbonaceous biofilm growth though capture of filtered particles (Schelegel, 1988; Boller *et al.*, 1994). Ammonia concentration in the feed is important primarily when start-up is taking place. High ammonia concentrations increase nitrifying biofilm development leading to a short start-up time (Haug and McCarty, 1972; Harremoes, 1982). Ammonia removal through oxidation is required in preference to assimilation as only oxidised nitrogen may be removed through denitrification (Jimenez *et al.*, 1987). Assimilated nitrogen may only be removed through biofilm removal by backwashing. Thus long biomass residence times are required of around 5 d (compared to 12-24 h for secondary processes) to allow the majority of the ammonia to be oxidised (Dillon and Thomas, 1990).

The objectives of this work were to investigate the start-up performance of parallel sunken and floating media reactors for nitrification and the effect low temperatures have on treatment efficiency.

MATERIALS AND METHODS

Experiments were carried out using two identical polyvinylchloride (PVC) reactors run in parallel. Each reactor was cylindrical in shape with an overall height of 2 m and diameter of 0.2 m. The media height within the reactors was 1.7 m with a total volume of 0.054 m^3. Liquid influent was introduced at the base of each column using a 5 cm filter nozzle (Degremont) and aeration was carried out using three identical nozzles placed around the base plates. Sample ports were placed at 20 cm interval along the length of the reactor and a port was placed on the base of each reactor to allow removal of sludge and liquid from the bottom of the reactors. The medias were kept in place using rigid polypropylene mesh with 1.5 mm holes which was held in place at the top of the reactors using clamps.

The medias used were made of recycled polypropylene (Cookson plc, Cheshire, UK) which was re-extruded to form a media with a relative density 0.92 (floating) and mixed with a calcium carbonate filler (40%) to form a media identical in shape and size but with a relative density of 1.05 (sunken). Each media was cylindrical with a diameter of 2.4-2.7 mm and length 4-6 mm. The surface area of the media was 1160 m^2m^{-3} and voidage of 42%

The process liquid used in the study was partially nitrified secondary sewage from Cranfield University's sewage treatment works. The sewage was passed upflow through the reactors at a flowrate of 0.2 l min^{-1} with a residence time of 2 h. Aeration was carried out at a rate of 2 l min^{-1} (air:liquid ratio of 10:1). Backwashing was carried out on a weekly basis after an initial start-up period to remove excess biomass. To backwash the influent air and liquid flows were shut-down and 2.5 l of liquid were drained from the bottom of the reactors. Following this air scour was carried at a rate of 25 l min^{-1} for 1 min. The reactors were left for 2 min and a further 7.5 l of liquid were drained. The process air and liquid were then resumed. Analysis of suspended solids, total COD, ammonia nitrates and nitrites were made every alternate day with temperature, dissolved oxygen and pH checked daily (APHA, 1992).

RESULTS AND DISCUSSION

Three factors primarily influenced the performance of the reactors, influent nutrient quality, temperature and backwashing rate. Previous studies have also indicated that rate limiting step that influences nitrification is oxygen availability, as high levels of oxygen are required for nitrification to take place (Cecen and Gonenc, 1991; Boller *et al.*, 1994; Cecen and Gonenc, 1994). At the aeration rate used dissolved oxygen concentrations in the effluent remained high at 5 mg l^{-1} or greater. Off gas measurements of the air leaving the reactors showed no apparent reduction in oxygen concentrations (resolution > 0.1%) subsequently in was assumed that oxygen limitation did not inhibit the performance of the reactors. Throughout the experimental period pH remained between 7.6 and 8.2 which was in the optimum range for nitrification (Barnes and Bliss, 1983).

Over the first 15 d of start-up influent nutrient levels remained low at 25 to 75 mg l^{-1} SS (Figure 1a), 50 to 140 mg l^{-1} tCOD (Figure 1b) and 1 to 5 mg l^{-1} NH_3-N (Figure 1c). Over this period there was a increase in suspended solids removal reaching a maximum of 57 % for the sunken media and 78 % for the floating media (Figure 2a). Similarly tCOD removal increased over this period with greater removal in the floating media reaching 71 % compared to 58 % in the sunken media (Figure 2b). Ammonia removal was highest in the floating media attaining maximum of 60 % with only 40 % in the Sunken media.

On day 15 a temperature shock was carried out during which the mean reactor temperature was reduced from 8 to 4 °C with a corresponding increase in suspended

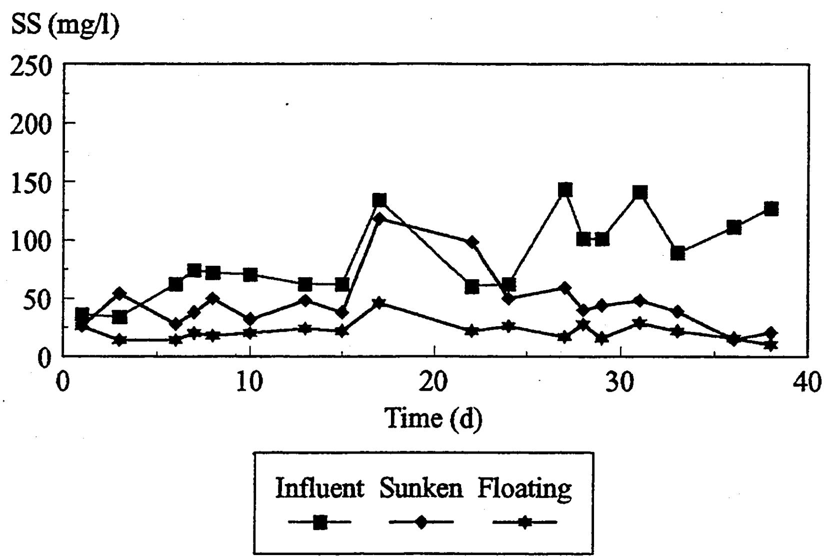

Figure 1a. Suspended Solids Concentrations During Start-Up and Shock Loadings.

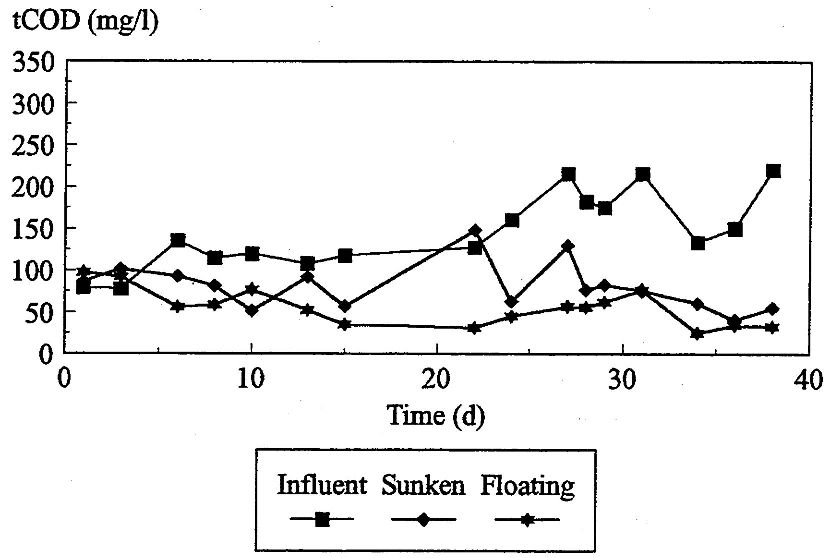

Figure 1b. tCOD Concentrations During Start-Up and Shock Loadings.

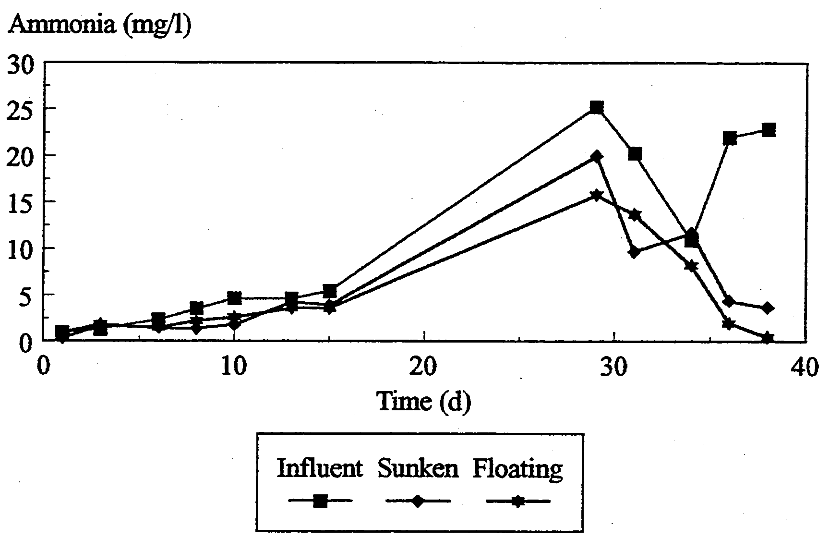

Figure 1c. Ammonia (NH_3-N) Concentrations During Start-Up and Shock Loadings.

solids, tCOD and ammonia loading. After a period of 3 d during which the reactor temperature was maintained at 3 to 5 °C, the temperature was returned to 8 +/- 3 °C and then maintained with nutrient loading remaining high.

Increased nutrient loading and reduced temperature showed the greatest effect on the sunken media performance. Suspended solids, tCOD and ammonia removal dropped rapidly to almost 0 %. In the floating media though almost no effect was seen in SS and tCOD removal and ammonia removal rate showed some reduction. After 7 d at the higher nutrient loadings and following the temperature shock the performance of the sunken media in suspended solids removal dropped to 0 % with only a reduction of 4 % in the floating media. There was then a rapid increase in performance with a maximum of 89 % removal in the sunken media and 83 % in the floating media (Figure 2a). The floating media performance was better than that for the sunken media in tCOD removal, with virtually no effect on the removal rate in the floating media but a rapid drop in the performance of the sunken media over the first 7 d following temperature shock. After 25 d at the higher loading (day 40) removal rates reached 88 % in the floating media from a value of 71 % 7 d after the temperature shock. This compares to a value of 79 % in the sunken media which rose from a value of 0 % following the temperature shock. These results showed that with suspended solids and total COD removal the floating media was able to withstand shock loadings, whereas the sunken media tended to initially show poor response and required several days to recover though the following recovery was rapid..

Unlike the increase in suspended solids and total COD loading where the floating media performance was largely unaffected and a decrease in the sunken media performance was seen, there appeared to be an improvement in ammonia removal performance in both media types with the higher ammonia loading though recovery took longer. Though the floating media performed better than the sunken media overall in ammonia removal, the

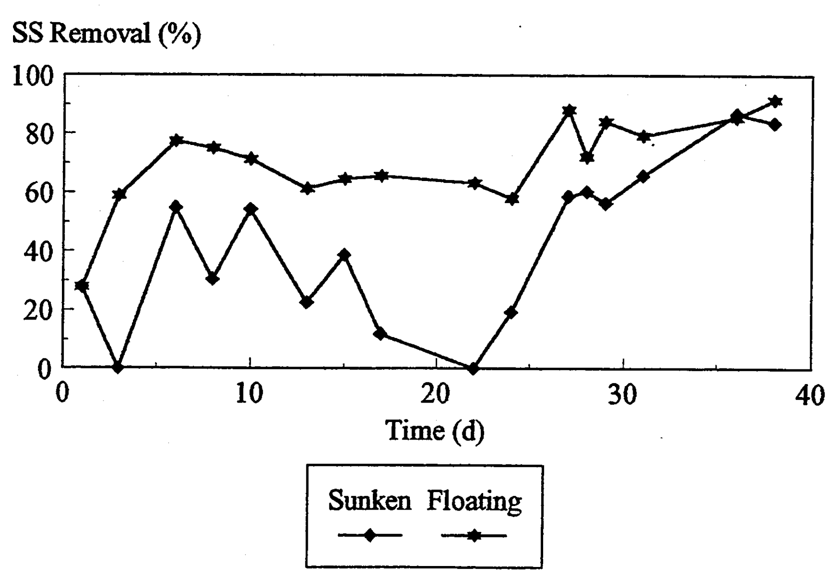

Figure 2a. Suspended Solids Removal Rates During Start-Up and Shock Loadings.

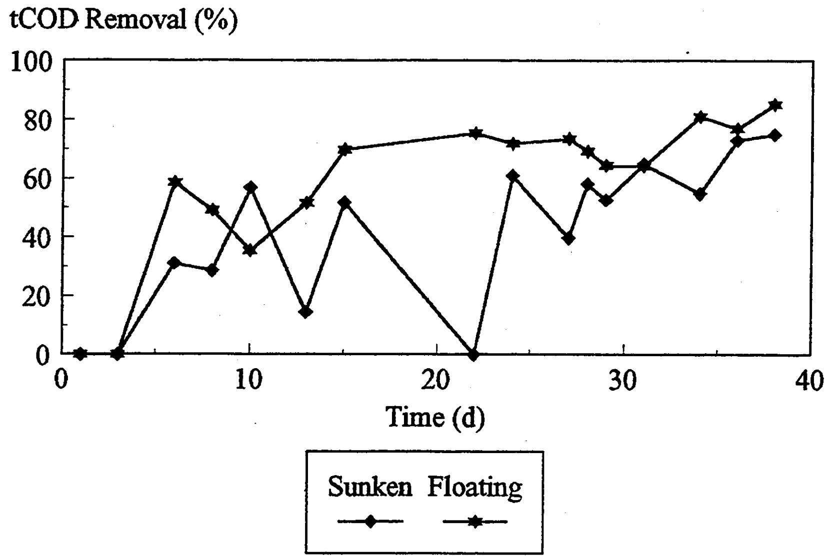

Figure 2b. tCOD Removal Rates During Start-Up and Shock Loadings.

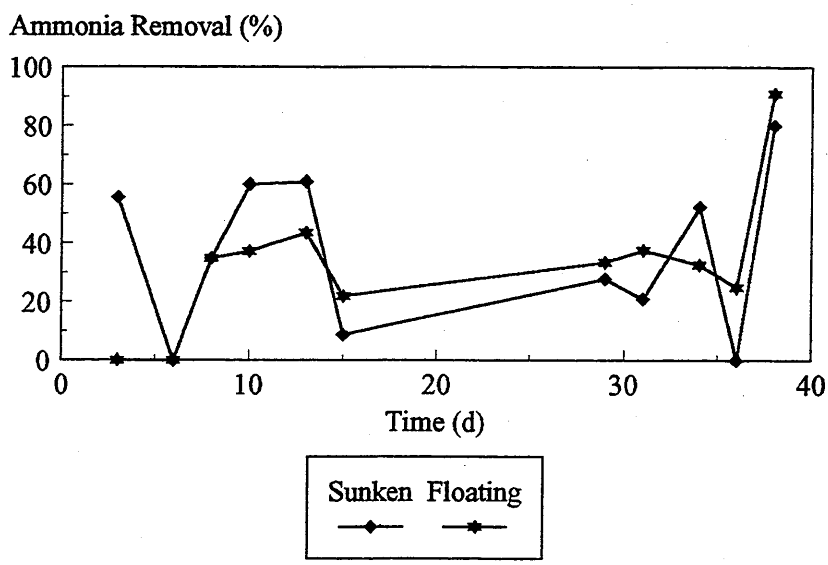

Figure 2c. Ammonia (NH_3-N) Removal Rates During Start-Up and Shock Loadings.

treatment profiles (Figure 2c) were similar in shape compared to those for SS and tCOD removal (Figures 2a and 2b).

After day 40 suspended solids, tCOD and ammonia removal rates decreased and some clogging of the reactors was observed. It was thus decided to backwash on day 44. Over the 5 d following backwashing SS and tCOD removal in the sunken media remained very low (Figure 3a and 3b). Following a rapid increase in performance over the next 2 d there was a subsequent decrease in SS and tCOD removal until backwashing was repeated 21 d after the first backwashing. Though a similar trend was seen in the floating media the effect on performance was not as great. Unlike the sunken media which

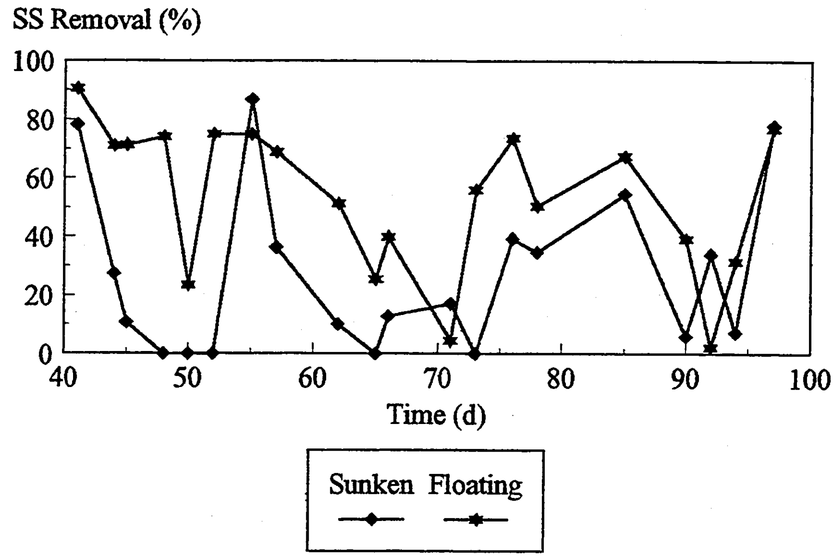

Figure 3a. Suspended Solids Removal Rates During Backwashing Period.

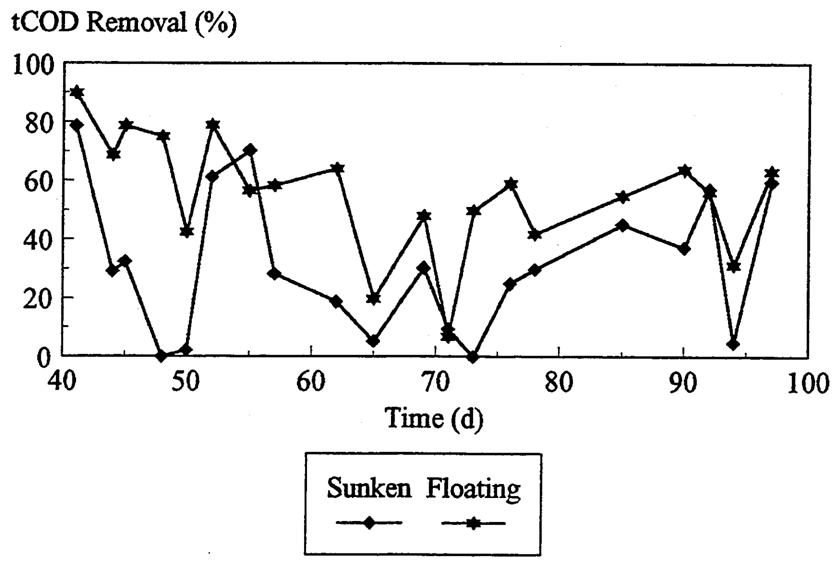

Figure 3b. tCOD Removal Rates During Backwashing Period.

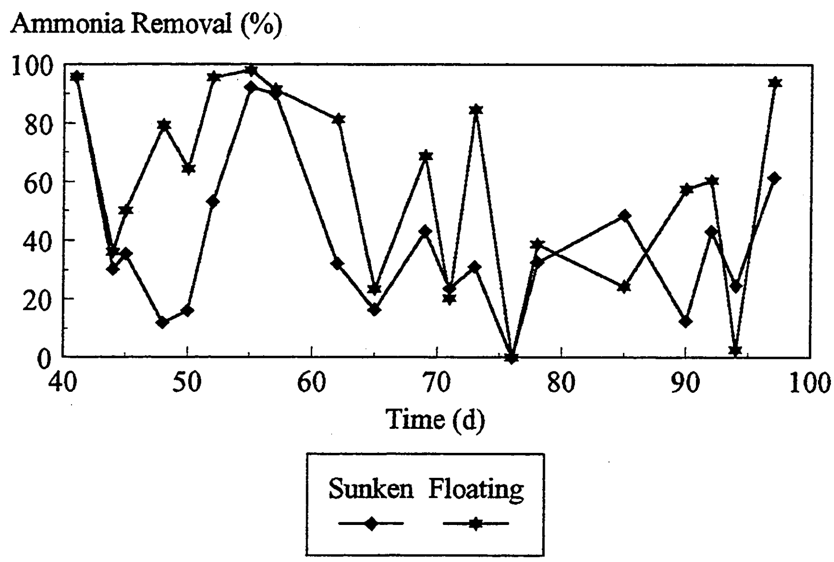

Figure 3c. Ammonia (NH_3-N) Removal Rates During Backwashing Period.

showed an initial decrease in ammonia removal the floating media showed an immediate increase in removal rate following backwashing. Ammonia removal reached 98 % in the floating media and 91 % in the sunken media though a performance similar to that for SS and tCOD removal was seen with a rapid drop in removal rates following an initial peak. Though backwashing was repeated only every 21 d (day 65 and 86) performance of the reactors remained variable and unstable over the 56 d following the first backwash.

Suspended solids removal rates of approximately 50% have been reported in nitrifying BAFs (Rogalla and Payraudeau, 1988). High suspended solids removal rates seen during these experiments suggested that high influent solids concentrations may have caused some inhibition to nitrifier growth. Concentrations of COD have been found to inhibit nitrification if limited or in excess (Al-Haddad et al., 1991; Sammut al., 1992). If limited, there will be an insufficient carbon source for biofilm growth, if in excess, the nutrients will promote rapid growth of carbonaceous removing bacteria in preference to slow growing nitrifiers (Faup et al., 1982; Rogalla and Payraudeau, 1988; Rogalla and Bourbigot, 1990a). During the experimental period sCOD remained below 64 mg l^{-1} with mean values of 46 mg l^{-1} during start-up and 51 mg l^{-1} during shock loading (Table 1).

Generally only partial sCOD removal rates of 20 to 30% have been found in tertiary BAFs with 90% ammonia removal (Sakuma et al., 1993; Rogalla and Payraudeau, 1988). This compares to mean values of 34% for the sunken media and 46% for the floating media. Thus only partial removal of the sCOD indicated that low sCOD concentrations

TABLE 1. Mean Values and Ranges () of Influent and Effluent Nutrient Concentration During Start-Up and Shock Loading.

	Influent		Effluent Sunken Media		Effluent Floating Media	
	Start-Up	Shock Load	Start-Up	Shock Load	Start-Up	Shock Load
Suspended Solids (mg/l)	59 (34-74)	103 (60-143)	39 (26-54)	52 (155-118)	20 (14-26)	23 (11-46)
Total COD (mg/l)	107 (77-135)	170 (127-220)	80 (52-101)	79 (41-148)	67 (36-98)	46 (25-77)
Soluble COD (mg/l)	46 (38-63)	51 (40-64)	35 (30-43)	38 (22-58)	38 (29-50)	27 (15-77)
Ammonia (mg/l)	3.2 (0.9-5.4)	20.3 (9.9-24.2)	2.1 (0.4-4.2)	9.9 (3.7-20)	2.3 (1.0-3.6)	8 (0.5-15.8)
Nitrites (mg/l)	0.41 (.24-.58)	0.52 (.22-.74)	0.34 (.27-.43)	0.44 (.22-.59)	0.64 (.32-.86)	0.23 (.02-.42)
Nitrates (mg/l)	3.7 (2.5-4.4)	3.4 (1.2-6.2)	4.4 (2.2-5.9)	4.9 (3.8-7.0)	4.2 (3.3-6.0)	4.6 (2.9-6.0)

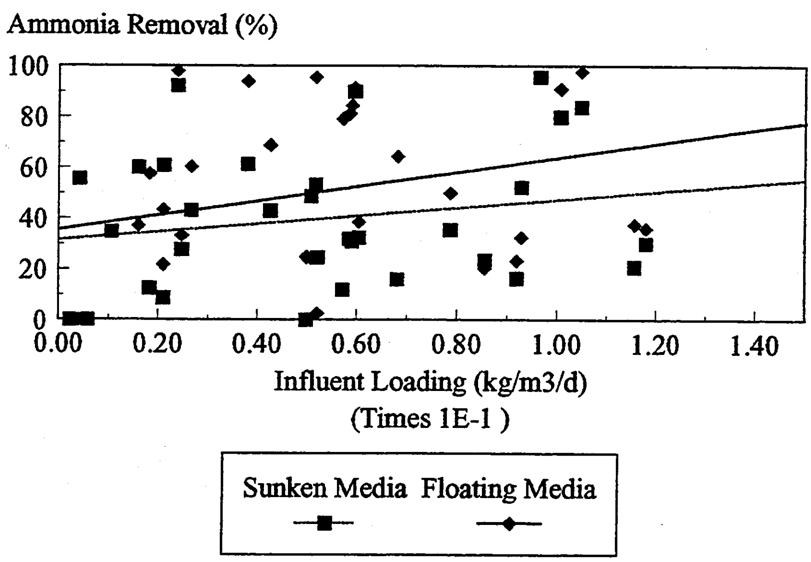

Figure 4. Ammonia Removal Rates versus Influent Ammonia Loading.

did not inhibit performance though high suspended solids removal rates indicated that there was preferential growth carbonaceous removing bacteria. Low concentrations of ammonia inhibit the start-up and performance of reactors through nutrient limitation. Indicated in the early stages of these experiments higher loadings of ammonia promoted the growth of nitrifiers and subsequently improved nitrification rates (Harremoes, 1982).

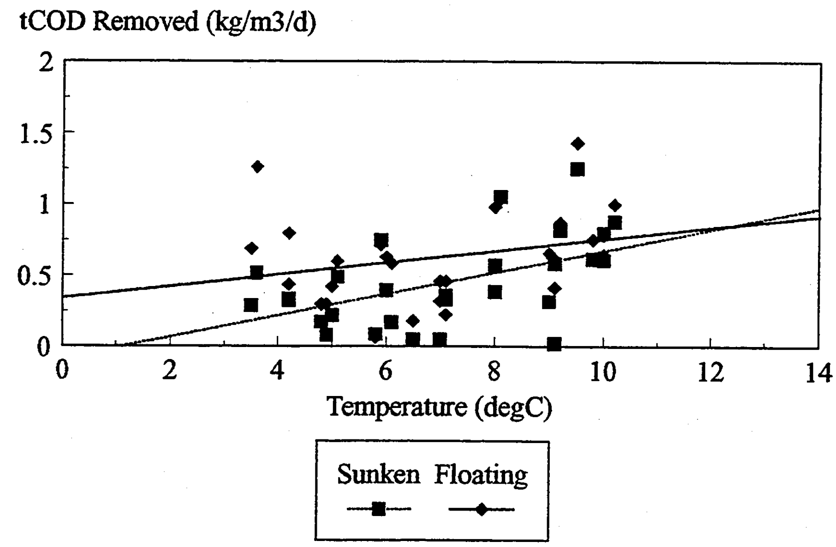

Figure 5a. Total COD Loading Removed with Increasing Temperature.

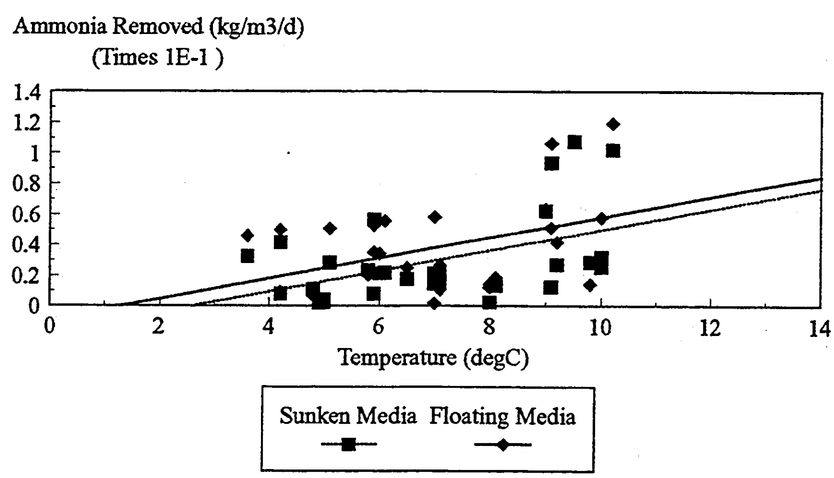

Figure 5b. Ammonia Loading Removed with Increasing Temperature.

Due to the variation in temperature and performance, no distinct correlation could be made between influent ammonia loading and removal rate (Figure 4).

The reactors were run between January and April during which time the influent temperature ranged from 3.4 °C to 10.2 °C. Over the experimental period there appeared to be little direct relationship between temperature and total COD removed (Figure 5a) with R^2 values of 0.261 for the sunken media and 0.077 for the floating media. Generally though the floating media appears to have performed better than the sunken media with removal rates as high as 80% at temperatures of 3.3 °C. Ammonia removal showed a better relationship with temperature (Figure 5b) with R^2 values of 0.281 for the sunken media and 0.181 for the floating media. Below 5 °C more than 60% removal was achieved, but at 10 °C almost full nitrification was obtained at removal loadings of 1.21 kg $m^{-3}d^{-1}$ for the floating media and 1.05 kg $m^{-3}d^{-1}$ for the sunken media. Again the floating media performed better than the sunken media. From the results it appeared that nitrification would cease at between 1.5 and 2 °C. Temperature has been found to make a great effect on biological reactor performance (Tschui *et al.*,, 1994) with up to 2% variation in nitrification performance occurring for every 1 °C change in temperature (Fdz-Polanco *et al.*, 1994). Though generally this is the case, BAF reactors have been found to perform well at variable temperatures. Though temperatures below 15 °C have been found to affect carbonaceous removal, low temperatures have been found to have only a minimal effect on nitrification (Harremoes, 1982; Iida, 1984; Paffoni, 1990). From these results the rate of increase in ammonia removal between 5 and 10 °C was found to be 3.06 in the sunken media and 2.34 in the floating media. Between temperatures of 10 and 20 °C the rate of change was calculated 2.35 in the sunken media and 2.15 in the

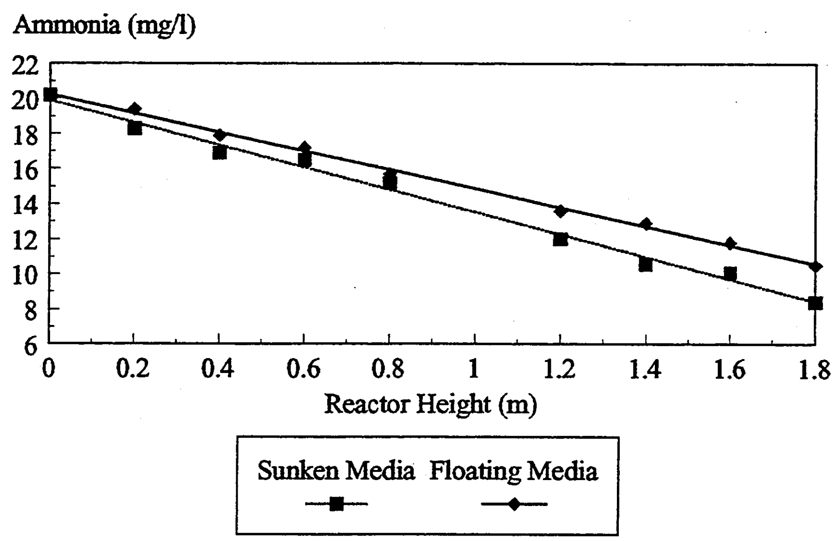

Figure 6. Reactor Profile During Ammonia Removal.

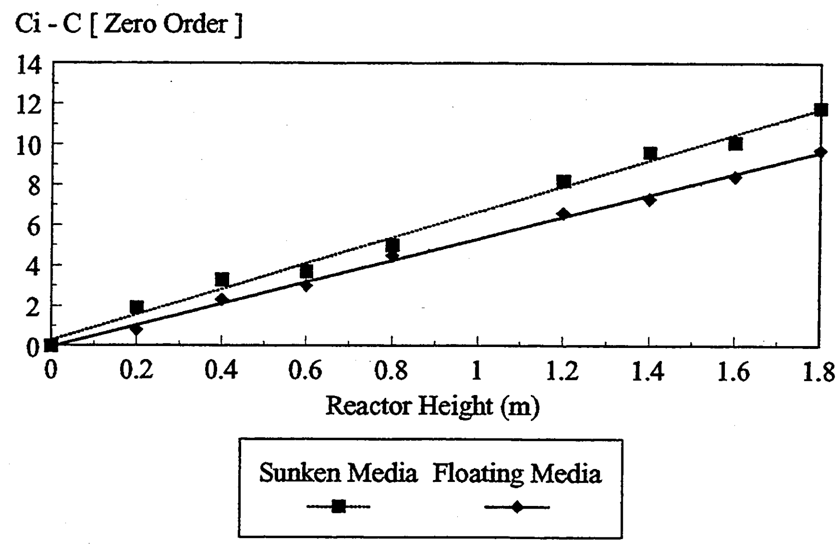

Figure 7a. Calibration of Reactor Profile as a Zero Order Reaction.

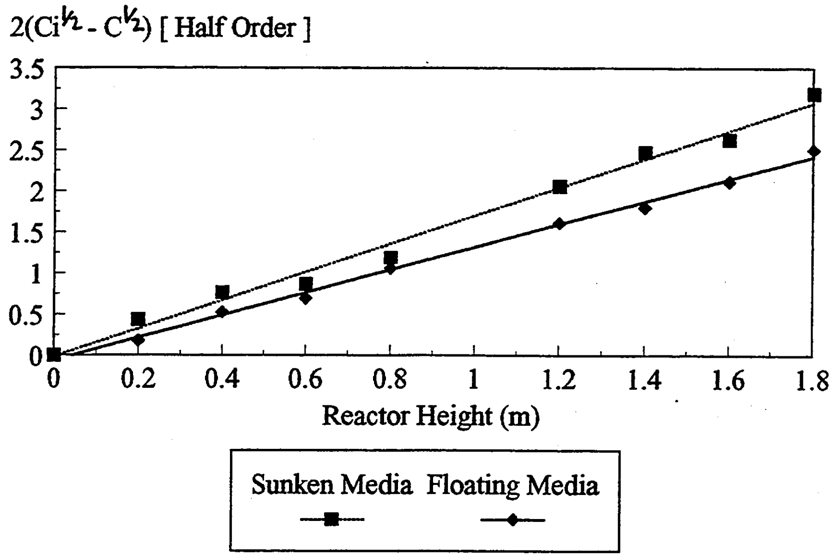

Figure 7b. Calibration of Reactor Profile as a Half Order Reaction.

floating media. This compared to a value of 2.33 found in rotating biological contactors (RBC). Similar Q_{10} values, which indicate the rate of growth change over a 10 °C range, of 2.59 in pure cultures and 3.32 in activated sludge have also been found (Barnes and Bliss, 1983).

Nutrient concentrations are leading parameters that affect nitrification within fixed-film reactors (Boller *et al.*, 1994). Solids concentrations in influent liquids have been found to inhibit performance in nitrifying reactors through absorption onto the biomass surface. This absorption causes thicker biofilms and inhibits the growth of nitrifiers. The solids also promote the growth of carbonaceous removing biofilms which further inhibit nitrifier growth (Boller *et al.*, 1994). From the results there appeared no direct correlation between suspended solids levels and ammonia removal rates.

Backwashing frequency appeared to be a major influence on reactor performance. In fixed-film reactors using structured media biofilm, profiles are produced with high concentrations at the inlet, decreasing towards the outlet as nutrient concentrations drop (Hao and Chen, 1994). In granular media reactors backwashing causes redistribution of biofilm producing a more even distribution throughout the reactor, though this more even biofilm distribution depends on backwash frequency (Fruhen *et al.*, 1994). Initial backwashing on day 44 caused a rapid increase in ammonia removal without a subsequent increase in effluent nitrate concentrations. From this it can be concluded the ammonia nitrogen was not oxidised but used for biomass assimilation (Jimenez et al., 1987; and Dillon Thomas, 1990). This continued until day 55. At this point there appeared to be limited ammonia assimilation and a drop in ammonia removal efficiency. It can be concluded from the results that during initial start-up regular backwashing promoted biofilm growth. Following the build up of the biofilm though regular backwashing inhibited the performance of the reactors and should be carried out only where a decline in nutrient or solids removal is seen. Subsequently the net withdrawal of biomass from the system was greater than the net growth rate of the nitrifying bacteria. Thus backwashing should be restricted to avoid washout. Though fixed-film processes tend to be stable under shock loads (Barnes and Bliss, 1983), during the experimental period the process appeared to be unstable through over backwashing.

Primary factors affecting nitrifying reactor performance are nutrient and gas diffusion. At low liquid velocities, found in treatment processes such as found in trickling filters external diffusion is a rate limiting factor. With higher velocities such as those found in

TABLE 2. Reaction Rate Constants for Ammonia Removal.

	Sunken Media	Floating Media
k (Zero Order) $kg\ m^{-3}\ s^{-1}$	1.54	1.38
k (Half Order) $(kg\ m^{-3})^{0.5}\ s^{-1}$	0.44	0.36

biological aerated filters, external diffusion is greater, thus reducing the potential of external diffusion becoming a rate limiting factor (Barnes and Bliss, 1983).

Unlike BOD removal which is a first order reaction, nitrification is a zero order reaction and thus independent of nutrient (ammonia) concentration. Other factors such as high BOD can influence the order of reaction, making the nitrification zero order reaction into a half or first order reaction (Arvin and Harremoes, 1989). Towards the end of the shock loading phase bed profiling was carried out, which measured ammonia concentration at different heights along the length of the reactors and thus at increasing treatment times (Figure 6). These profiles indicated a zero to half order reaction. To confirm the type of reaction the results were treated as both zero order (Figure 7a) and half order (Figure 7b). R^2 values of 0.992 for the sunken media and 0.997 for the floating media were found for a zero order reaction. These compare to 0.990 for the sunken media and 0.996 for the floating media for a half order reaction. Thus diffusion inhibition could not be ruled out. From these graph the reaction rate constant k was found for both orders of reaction (Table 2). These indicated a greater rate of ammonia removal in the sunken media than in the floating media.

Generally the floating media performed better than the sunken media in all aspects though nitrifying bacteria appear to be more efficient at ammonia removal in the sunken media than in the floating media. The high suspended solids removal efficiency in the floating media reactor may have led to some inhibition of nitrification, subsequently improved nitrification may be seen using wastewater with a low suspended solids concentration. Though backwashing was carried out only on a weekly basis improved performance may have been seen if backwashing is carried out only when reactor performance began to decline.

ACKNOWLEDGEMENTS

The authors are grateful to the UK Engineering and Physical Sciences Research Council (EPSRC), Severn Trent Water and Tetra (Europe) Ltd. for funding this research.

REFERENCES

Al-Haddad, A. A., Zeidan, M.O., Hamoda, M.F. (1991). Nitrification in the Aerated Submerged Fixed-Film (ASFF) Bioreactor. *Jour. Biotech.*, 18, 115-128.

American Public Health Association (1992). Standard Methods for the Examination of Water and Wastewater. 18th Edition, American Public Health Organisation, Washington, D.C.

Arvin, E., Harremoes, P. (1989). Concepts and Models for Biofilm Reactor Performance. *Tech. Advances in Biofilm Reactors,* 4-6 April, Nice, 191-212.

Barnes, D., Bliss, P.J. (1983). Biological Control of Nitrogen in Wastewater Treatment. *1st Ed. Spo, New York.*

Beg, S.A., Hassan, M.M. (1987). Effects of Inhibitors on Nitrification in a Packed-Bed Biological Flow Reactor. *Wat. Res.*, 21, 191-198.

Boller, M., Gujer, W., Tsuchi, M. (1994). Parameters Affecting Nitrifying Biofilm Reactors. *Wat. Sci. Tech.*, 29, 10-11, 1-11.

Cecen, F., Gonenc, I.E. (1991). Nitrification-Denitrification of High Strength Nitrogen Wastes in Two Up-Flow Submerged Filters. *Wat. Sci. Tech.*, 26, 9-11, 2225-2228.

Cecen, F. Gonenc, I.E. (1994). Nitrogen Removal Characteristics of Nitrification and Denitrification Filters. *Wat. Sci. Tech.*, 29, 10-11, 409-416.

Cecen, F., Gonenc, I.E. (1995). Criteria for Nitrification and Denitrification of High Strength Wastes in Two Upflow Submerged Filters. *Wat. Environ. Res.*, 67, 132-142.

Dillon, G.R., Thomas, V.K. (1990). A Pilot-Scale Evaluation of the 'Biocarbone Process' for the Treatment of Settled Sewage and for Tertiary Nitrification of Secondary Effluent. *Wat. Sci. Tech.*, 22, 1/2, 305-316

Faup, G.-M., Leprince, A., Pannier, M. (1982). Biological Nitrification in an Upflow Fixed Bed Reactor (UFBR). *Wat. Sci. Tech.*, 14, 795-810.

Fdez-Polanco, F., Villaverde, S., Garcia, P.A. (1994). Temperature Effect on Nitrifying Bacteria Activity in Biofilters: Activation and Free Ammonia Inhibition. *Wat. Sci. Tech.*, 30, 11, 121-130.

Fruhen, M., Bocker, K., Eidens, S., Haaf, D., Liebeskind., M., Schmidt, F. (1994). Tertiary Nitrification in Pilot-Plant Plug Flow Reactors with Long Term Ammonia Deficiency. *Wat. Sci. Tech.*, 29, 10-11, 61-67.

Goncalves, R.F. Le Grand, L., Rogalla, F. (1994). Biological Phosphorus Uptake in Submerged Biofilters with Nitrogen Removal. *Wat. Sci. Tech.* 29, 10-11, 135-143.

Hao, O.J., Chen, J.M. (1994). Factors Affecting Nitrate Build-Up in Submerged Filter Systems. *Jour. Environ. Eng.*, 120, 1298-1307.

Harremoes, P. (1982). Criteria for Nitrification in Fixed Film Reactors. *Wat. Sci. Tech.* 14, 167-187.

Haug, R.T., McCarty, P.L. (1972). Nitrification with Submerged Filters. *J. WPCF.*, 44, 2086-2102.

Iida, Y., Teranishi, A. (1984). Nitrogen Removal from Municipal Wastewater by a Single Submerged Filter. *J. WPCF.*, 56, 251-258.

Jimenez, B., Capdeville, B., Roques, H., Faup, G.M. (1987). Design Considerations for a Nitrification - Denitrification Process Using Two Fixed-Bed Reactors in Series. *Wat. Sci. Tech.*, 19, 139-150.

Kugaprasatham, S., Nagaoka, H., Ohgaki, S. (1992). Effect of Turbulence on Nitrifying Biofilms at Non-Limiting Substrate Conditions. *Wat. Res.*, 26, 1629-1638.

Lee, N.M., Welander, T. (1994). Influence of Predators on Nitrification in Aerobic Biofilm Processes. *Wat. Sci.Tech.*, 29, 7, 355-363.

Legise, J.P., Gilles, P., Mureaud, H. (1980). A New Development in the Biological Aerated Filter Bed Technology. Presented at the 53rd Annual WPCF Conference, Las Vegas, Nevada.

Mann, A., Fitzpatrick, C.S.B., Stephenson, T. (1995). A Comparison of Floating and Sunken Media Biological Aerated Filters Using Tracer Study Techniques. Trans. I.Chem.E. Pt. B, 73, 137-143.

McCarty, P.L., Haug, R.T. (1971). Nitrogen Removal from Wastewaters by Biological Nitrification and Denitrification. Microbial Aspects of Pollution, Sykes and Skinner (Edit.), Symposium Series No.1, Society for Applied Bacteriology.

Murphy, K.L., Sutton, P.M., Wilson, R.W., Jank, B.E. (1977). Nitrogen Control : Design Considerations for Supported Growth Systems. *J. WPCF*, 49, 549-557.

Paffoni, C., Gousailles, M., Rogalla, F., Gilles, P. (1990). Aerated Biofilters for Nitrification and Effluent Polishing. *Wat. Sci. Tech.*, 22, 7/8, 181-189.

Rogalla, F., Bourbigot, M-M. (1990a). New Developments in Complete Nitrogen Removal with Biological Aerated filters. *Wat. Sci. Tech.* 22, 1/2, 273-280.

Rogalla, F., Payraudeau, M. (1988). Tertiary Nitrification with Fixed Biomass Reactors. *Wat. Supply*, 6, 347-354.

Rogalla, F., Payraudeau, M., Baquet, G., Bourbigot, M., Sibony, J., Gilles, P. (1990b). Nitrification and Phosphorus Precipitation with Biological Aerated Filters. *J. WPCF.*, 62, 169-176.

Sakuma, H., Tanaka, T., Maki, Y. (1993). Studies on Nitrification and COD removal Using Biological Aerated Filters. *Proc. 6th World Filt. Cong.*, Nagoya, Japan, 743-746.

Sammut, F., Rogalla, F., Goncalves, R.F., Penillard, P. (1992). Practical Experiences with Removing Nitrogen and Phosphorus on Aerated Biofilters. *Europ. Conf. Nut. Rem. from Wastewater*, Wakefield, Sept., 92.

Schelegel, S. (1988). The Use of Submerged Biological Filters for Nitrification. *Wat. Sci. Tech.*, 20, 4/5, 177-187.

Tijhuis, L., van Loosdrecht, M.C.M., Hejnen, J.J. (1992). Nitrification with Biofilms. *Wat. Sci. Tech.*, 26, 9-11, 2207-2211.

Tschui, M., Boller, M., Gujer, W., Eugster, J., Mader, C., Stengel, C. (1994). Tertiary Nitrification in Aerated Biofilters. *Wat. Sci. Tech.*, 29, 10-11, 53-60.

Chapter 7.

SOLIDS REMOVAL FROM FLOATING AND SUNKEN MEDIA BIOLOGICAL AERATED FILTERS

Submitted for Publication,

Environmental Technology.

SOLIDS REMOVAL FROM FLOATING AND SUNKEN MEDIA BIOLOGICAL AERATED FILTERS.

Allan Mann and Tom Stephenson.

School of Water Sciences, Cranfield University, Cranfield, Bedford. MK43 0AL, UK.

ABSTRACT

Biological aerated filters (BAFs) are three-phased fixed media biological reactors, for wastewater treatment. Granular media BAFs combine solids removal with other biological treatments such as carbonaceous removal and nitrification. Subsequently periodic cleaning (backwashing) of such reactors is required to remove captured solids and excess biological growth. Backwashing is carried out using high rate liquid flows, high rate air flows (air scour) or a combination of the two.

The aim of this work was to compare the performance of two pilot-scale reactors run upflow, one containing a sunken media and the other containing a floating media. The amount of solids removed during each backwash was compared with to solids and nutrient removal performance of each reactor. The floating media showed better removal of both suspended solids and chemical oxygen demand (sCOD and tCOD). Since no clogging of either reactor was seen there was no build up of solids within the reactors over the experimental period. Thus larger amounts of solids removed from the floating media compared to the sunken media during backwashing showed greater solids capture and biological growth in the floating media.

KEYWORDS

Biological aerated filters, backwashing, floating media, sunken media.

INTRODUCTION

In the past research on backwashing has concentrated on filters used for solids removal from drinking water (Stevenson, 1994). In wastewater treatment, backwashing is also required for the removal of solids but in such reactors the solids removed are predominantly excess biomass (Park and Ganczarczyk, 1994). Thus backwashing has two functions, to remove captured solids and remove excess biomass. Backwashing is carried out in two ways. High rate aeration (air scour) which causes high shear which dislodges filtered solids, and excess biofilm and high rate liquid flows which are used to flush out this dislodged material (Quickenden *et al.*, 1992).

Solids removal depends on factors such as the properties of the captured solids (size and inertness), the media support used (shape and size), the biofilm structure and the hydraulic characteristics of the reactors used (nominal flowrate and aeration rate) (Arvin and Harremoes, 1989). The amount and size of the particulate material will primarily affect the biofilm growth but in granular media reactors it will also cause clogging of the media (Pujol *et al.*, 1992b). The media used in reactors will affect the filtration ability and biofilm thickness (Jimenez *et al.*, 1987). Structured media and large granular media have a limited filtration ability and also allow the growth of thick biofilm because of the large void spaces (Ryhiner *et al.*, 1994). Reactors using smaller granular media have a higher filtration ability and limited room for thick biofilms to form. Subsequently such reactors are prone to clogging. In addition rough surfaced media show improved solids retention (Ruffer and Rosenwinkel, 1984).

Biofilm structure affects the amount of backwashing required. In tern the backwashing influences species distribution within reactors through bed movement during backwashing (Wilderer *et al.*, 1995). Reactors with stable, rapidly growing biofilms involved in carbonaceous removal (secondary treatment) require more frequent and high rate backwashing than reactors with less stable and slower growing biofilms used in nitrification (tertiary treatment) (Dillon and Thomas, 1990).

Biofilm thickness and solids retention is influenced by the hydraulic characteristics of reactors. High gas and liquid flowrates cause high shear within reactors, which limits the thickness of biofims through erosion and sloughing (Wilderer *et al.*, 1995). Biofilm thickness is also controlled by factors such as the growth rate and decay of active biomass and the attachment and accumulation of particulates and inert substances to the biofilm (Hoehn and Ray, 1973; Arvin and Harremoes, 1989).

There are several designs of aerobic fixed film reactors used at present, which include non-submerged reactors such as trickling filters, partially submerged filters such as rotating biological contactors (RBC) and submerged reactors such as contact aerators and biological aerated filters (Arvin and Harremoes, 1989). Contact aerators and BAFs are similar in design but differ in the type of support media used. Contact aerators use structured packing or plates as the biofilm support. This type of media produces a high void volume within the reactors and a specific surface area of 200 to 500 m^2m^{-3}. Subsequently liquid is completely mixed in the reactors and the possibility of clogging is very small (Rusten and Odegaard, 1986; Boller, 1987; Anderson, 1989). In structured packing solids removal from wastewaters is carried out either through settlement due to low liquid velocities and a high void volume or through bioadsorption (Ryhiner *et al.*, 1994). Backwashing is only carried out in such reactors to remove excess biomass, if required. Generally though sloughed biomass falls to the bottom of the reactors and periodically removed (Chudoba, *et al.*, 1991). In BAFs granular material (specific surface area of 1000 to 1500 m^2m^{-3}) is used as the biofilm support and tend to show a plug-flow rather than a completely mixed liquid profile (Mann *et al.*, 1995). Thus in BAFs periodic backwashing is required to remove captured solids as well as excess biomass. As well as granular and structured media, some novel medias have been investigated. These include rope, fibre and brushwood used to support biofilms and geotextiles which also filter particulates (Buswell and Pearson, 1929; Valentis and Lesavre, 1989; Han-Chang and Thomas, 1993).

The frequency and rate of backwashing depends primarily on the type of media used. The shape, density size and therefore voidage of the media affects the pumping requirements needed to wash the bed. Other factors that influence the frequency of backwashing through the influence on biofilm growth include organic loading rates, biomass growth rate and wastewater characteristics (Robinson *et al.*., 1994). BAFs are normally used for secondary or tertiary treatment. When used for secondary treatment, backwashing is carried out every 12 to 48 h whereas in tertiary treatment backwashing is less frequent due to lower biofilm growth through reduced nutrient levels and subsequently carried out only on a weekly basis (Dillon and Thomas, 1990; Bacquet *et al.*, 1991, Smith and Hardy, 1992).

There are three methods used to determine the frequency of backwashing. The simplest method used is to backwash on a regular basis, usually every 24h. This method does not require additional monitoring equipment but does not allow the optimum time period between washes to be used (Dillon and Thomas, 1990; Bacquet *et al.*, 1991; Wheale and Cooper-Smith, 1995). The second method used is to carry out backwashing when a preset head loss is reached. This method allows for the treatment of variable strength wastewaters though optimisation of the time between backwashings (Faup *et al.*, 1982; Jimenez *et al.*, 1986; Amar *et al.*, 1986; Paffoni *et al.*, 1990; Jepson and Jansen, 1993). The third method monitors effluent quality and subsequently controls backwash frequency through reactor performance (Gray, 1993).

The type of backwashing carried out is determined largely by the process configuration and on each individual process. Each process usually goes through stages of 1) air scour, 2) air scour with liquid backwash and 3) liquid backwash only, though simple use of air scour and liquid draining has been used successfully (Quickenden *et al.*, 1992). Generally backwashing is carried out upflow in reactors that use sunken media, in opposition to bed compression whether the process liquid is either upflow or downflow (Table 1) (Jimenez *et al.*, 1987; Kurbiel, 1987; Rogalla and Bourbigot, 1990; Pujol *et al.*, 1992a). Reactors using floating media though backwash downflow, again in the opposite direction to media compression (Rogalla and Bourbigot, 1990; Jepsen and Jansen, 1993; Vedry *et al.*, 1994; Visvanathan and Nhien, 1995).

Due to the high operational costs of backwashing, optimisation is necessary to minimise energy consumption and the quantities of air and wash water used (Carrand et al., 1990). Other considerations such as the use of low backwash water flows allows the process liquid piping to be used for backwashing and thus reduces costs (Kirichenko and Drushlyak, 1988). At the same time though it is necessary to maximise the time between washs, avoid media loss and maintain a good population of active biomass within reactors (Carrand *et al.*, 1990).

MATERIALS AND METHODS

Two identical pilot scale reactors were set-up to run upflow in parallel.(Mann *et al.*, 1995). Each reactor was 2 m in height, with a diameter of 0.2 m and cylindrical in shape. The bed height was 1.72 m holding 54 l of media. One reactor contained sunken media (relative density, 1.05) and the other contained a floating media (relative density, 0.92). Each media type used was made of recycled polypropylene and identical in shape and size except for density. The media was cylindrical in shape with a length of 4 to 6 mm and diameter of 2.3 to 2.7 mm. The voidage of each bed was 42 % and specific surface area of 1160 m^2m^{-3}.

Initial work carried out on the two reactors involved tracer studies of the flow characteristics of the reactors (Mann *et al.*, 1995). Following this, analysis was undertaken on the biological performance of the reactors. This work was undertaken over the start-up period of the reactors for secondary treatment, during steady state and following shock loadings (increased flowrates). Process liquid and air flowrates were maintained at 0.2 l min^{-1} and 2 l min^{-1} respectively during the start-up and steady-state periods. The liquid flowrates were increased to 0.3, 0.4 and 0.5 l min^{-1} with air to liquid ratios of 10 : 1 during the shock loading periods.

Backwashing was carried out on a daily basis during each period of secondary treatment. The primary purpose of the work was to analyse the biological performance of the two types of media and backwashing was carried out only to avoid clogging of the reactors. The method used was found to be adequate and was thus maintained through out the experimental period. Backwashing involved the process liquid and air being turned off and 2.5 l of liquid drained from the base of the reactors which took approximately 30s (23 m h^{-1}). Air scour was then carried out for 1 min at 25 l min^{-1}. The reactors were then left for 2 min to allow the media to settle and a further 7.5 l of liquid was then drained taking 90 s (23 m h^{-1}). Process liquid and air flows were then resumed.

Throughout the experimental period, analysis of the influent and effluents were carried out on a daily basis. Measurements of suspended solids (SS), total chemical oxygen demand (tCOD), soluble chemical oxygen demand (sCOD), ammonia, temperature, pH and dissolved oxygen (DO) were made. Backwash liquids were measured for suspended solids concentration and settleability (APHA, 1992).

RESULTS AND DISCUSSION

Start-up of the reactors took approximately 30 d and defined when the value of three consecutive SS and tCOD removal rates remained constant. Over this period there was a large amount of variance in the suspended solids removal rate (Figure 1a). Through out the start-up period though, the floating media showed a better performance than the sunken media. After 35 d both reactors showed similar effluent suspended solids profiles, though the mean influent suspended solids increased and became more variable (Figures 1a and 1b). A similar performance was also seen in the total COD removal in the reactors (Figures 2a and 2b). During start-up, over the first 30 d, there was a large variance in the effluent quality especially in the sunken media, which also showed a poorer performance than the sunken media (Figure 2a). After day 30, during steady-state, both reactors showed a similar performance, though the floating media still performed better. Unlike the suspended solids profiles which showed increased effluent values with increased

TABLE 1. Mean and Suspended Solids and Total COD Removed at Different Flowrates.

	Sunken Media		Floating Media	
Flowrate (l/min)	SS (mg l^{-1}) / (kg $m^{-3}d^{-1}$)	tCOD (mg l^{-1}) / (kg $m^{-3}d^{-1}$)	SS (mg l^{-1}) / (kg $m^{-3}d^{-1}$)	tCOD (mg l^{-1}) / (kg $m^{-3}d^{-1}$)
0.2 (Start-Up)	57 / 0.304	153 / 0.816	82 / 0.437	206 / 1.099
0.2	90 / 0.480	201 / 1.072	106 / 0.565	240 / 1.280
0.3	58 / 0.464	122 / 0.976	73 / 0.584	147 / 1.176
0.4	49 / 0.523	100 / 1.067	68 / 0.725	131 / 1.387
0.5	36 / 0.480	68 / 0.907	63 / 0.840	105 / 1.400

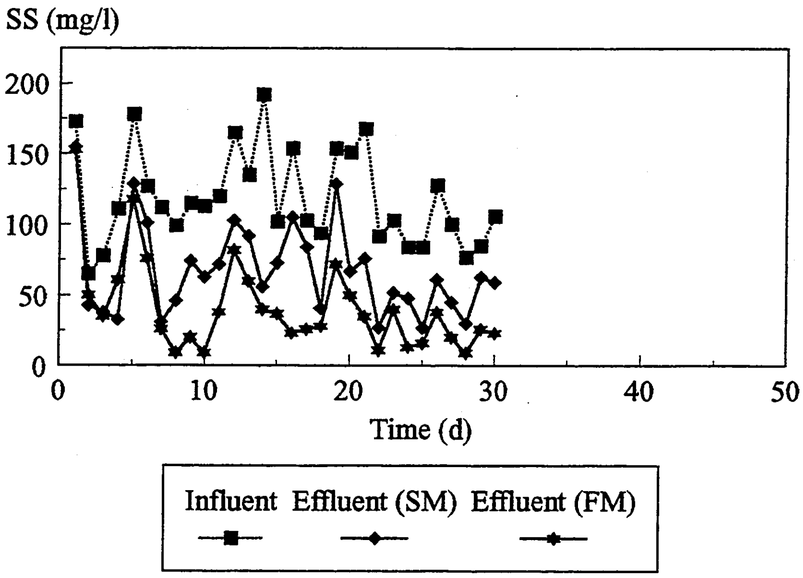

Figure 1a. Influent and Effluent Suspended Solids Concentrations During Start-Up.

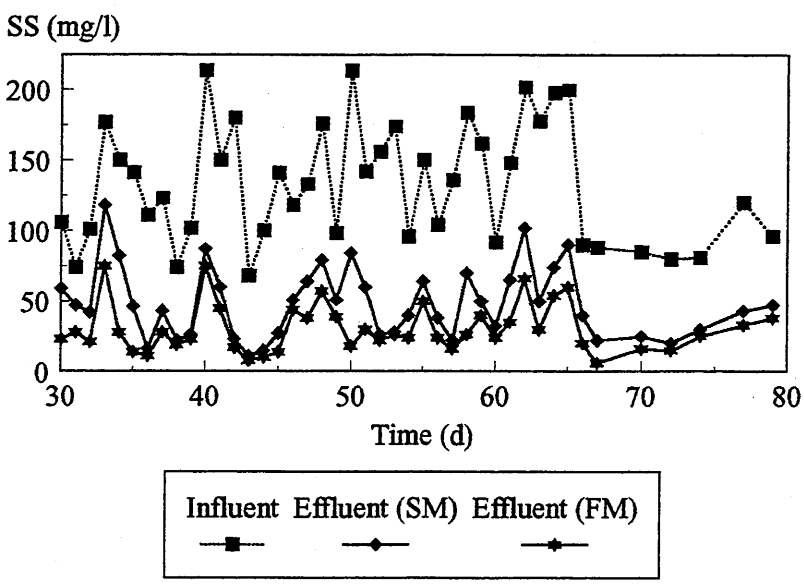

Figure 1b. Influent and Effluent Suspended Solids Concentrations During Steady-State.

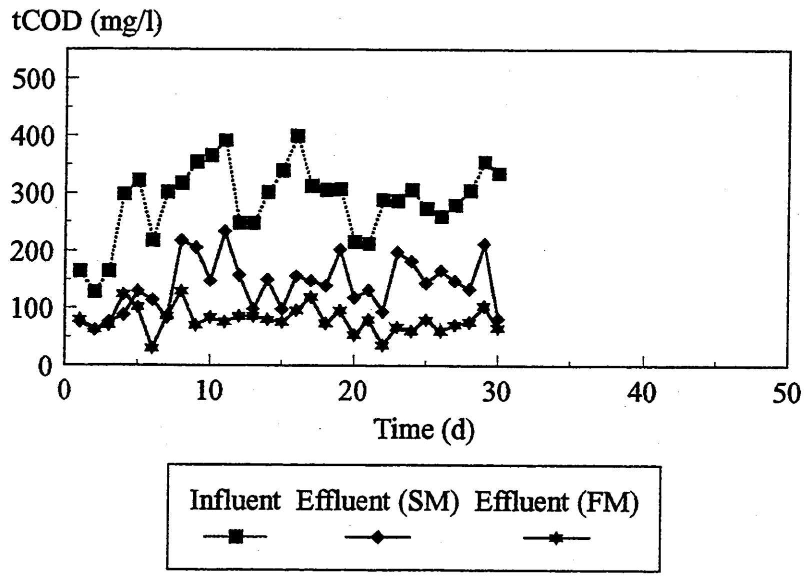

Figure 2a. Influent and Effluent Total COD Concentrations During Start-Up.

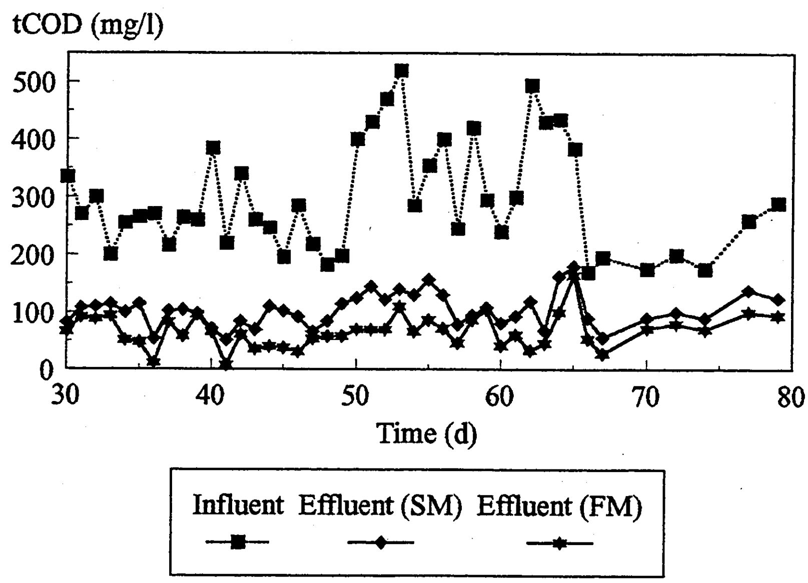

Figure 2b. Influent and Effluent Total COD Concentrations During Steady-State.

influent values, the total COD in the effluent remained stable, even though there was a large variance in the influent quality of 170 to 520 mg l^{-1} tCOD.

Following start-up there was a large increase in both mean suspended solids and mean total COD removed (Table 1). During start-up suspended solids effluent loading rapidly increased as influent loading increased, most noticably in the sunken media. During steady-state the effluent loading dropped, especially in the sunken media (Figure 3). A similar response was seen with the total COD. As influent total COD increased the effluent total COD increased more rapidly during start-up than at steady state (Figure 4).

When the reactors were run at higher flowrates there was an apparant decrease in effluent quality with a decrease in the amount of total COD removed (mg l^{-1}) (Table 1). With the sunken media the suspended solids loading removed appeared to remain largely unchanged with increased flowrates (Table 1). This showed that varriation in the hydraulic characteristics of the system had only a minor influence on the total solids capture efficiency of the sunken media. Whereas there was a very marked increase in the total amount of solids retained by the floating media, with an increase of 50% between flowrates of 0.2 and 0.5 l min^{-1}. This increase in efficiency in solids removal in the floating media may have been due to the increased upward force of the process liquid and air causing an increase in bed compaction which ultimately caused an improvement in filtration capability. The mean total COD loading removed in the sunken media showed only a slight varriation between the different flowrates, again indicating the limited effects flow varriation had on overall performance. In the floating media there was a slight improvement in the performance, though this may have been caused by the increasing suspended solids removal rate.

TABLE 2. Solids Removed During Backwashing (Flowrate = 0.2 l min^{-1}).

	Sunken Media		Floating Media	
	Mean (g)	Range (g)	Mean (g)	Range (g)
Start-Up				
1st Wash	1.06	0.30 - 3.50	1.08	0.20 - 3.54
2nd Wash	1.55	0.30 - 4.11	5.15	1.01 - 11.55
Overall	2.61	0.62 -6.92	6.22	1.76 - 12.06
Steady State				
1st Wash	0.77	0.28 - 1.35	0.84	0.30 - 2.51
2nd Wash	2.01	0.45 - 4.35	4.55	2.40 - 8.46
Overall	2.77	0.78 -5.20	5.39	2.75 - 8.99

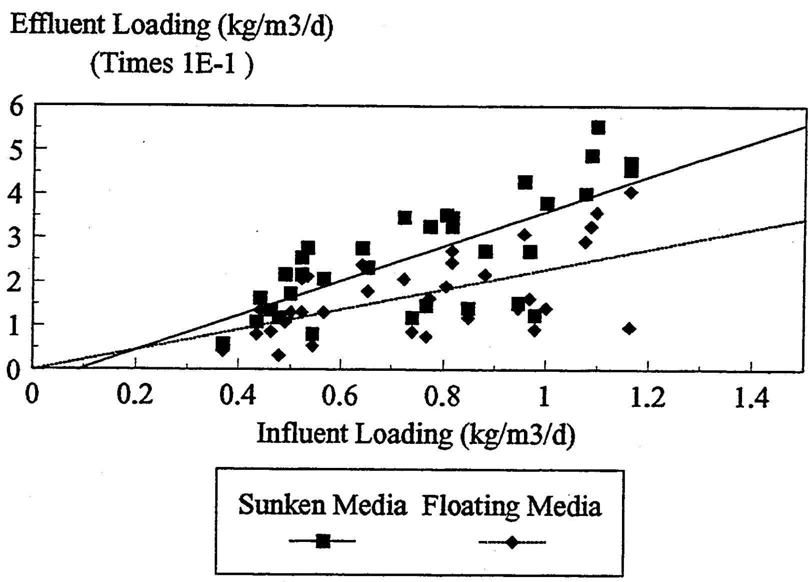

Figure 3. Effluent versus Influent Suspended Solids Loading During Steady-State.

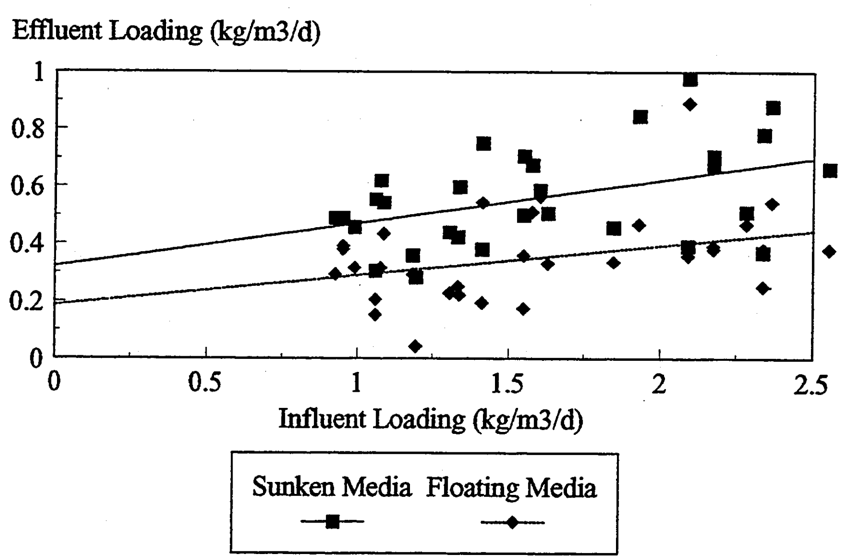

Figure 4. Effluent versus Influent Total COD Loading During Steady-State.

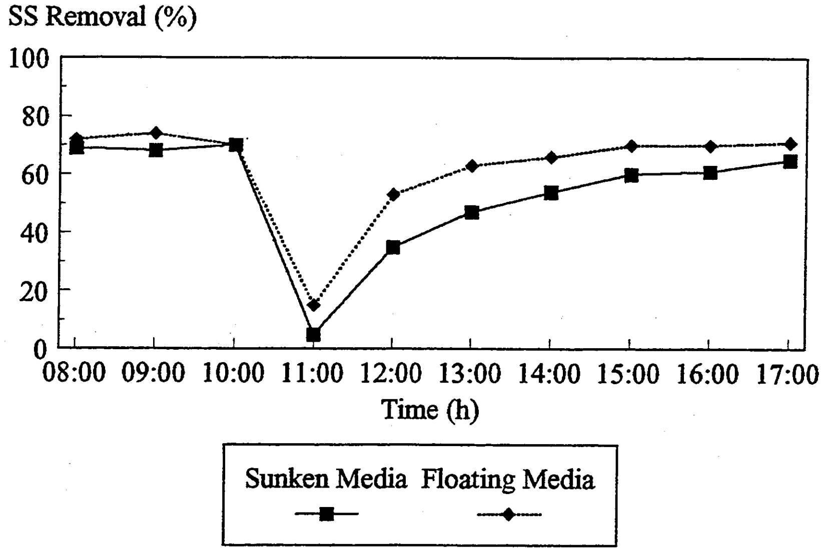

Figure 5a. Suspended Solids Removal Recovery Following Backwashing (10:00).

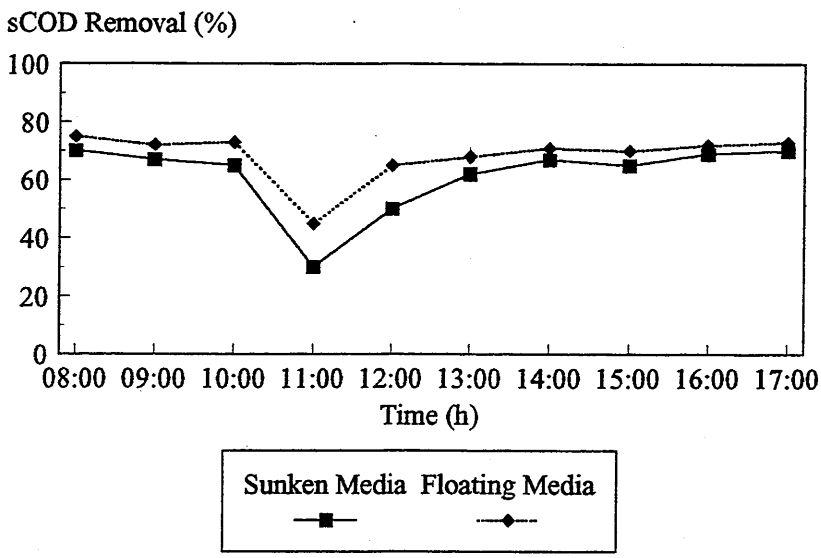

Figure 5b. Soluble COD Removal Recovery Following Backwashing (10:00).

With backwashing carried out every 24 h rapid recovery was necessary to obtain sufficient nutrient removal. At the lowest flowrate of 0.2 l min^{-1} there was an immediate drop in suspended solids removal following backwashing (Figure 5a). This appeared to be primarily caused by solids dislodged but not removed during backwashing. Within 3 h of backwashing, removal rates had improved to 34% from 5% after 1 h in the sunken media and 58% from 16% in the floating media. Full recovery only took place 7 h after backwashing in the sunken media and 5 h after backwashing in the floating media.

The effects were not as dramatic with the total COD removal (Figure 5b). In this case full recovery of both reactors took place 3 to 4 h following backwashing with a lower minimum removal rate and less rapid recovery occurring in the sunken media. At higher flowrates there was an increase in the recovery time. At the highest flowrate full recovery only occurred after 12 h in the floating media and 18 h in the sunken media.

There was a large difference in the amount of backwash solids removed in the two media types. Both media types produced similar amount of solids in the first wash prior to air scouring (Figures 6a and 6b) with mean vales of 0.77 g for the sunken media and 0.84 g for the floating media (Table 2). There was a large difference though in the amount of solids removed following air scour. In this case a mean value of 2.01 g of solids was removed from the sunken media and 4.55 g from the floating media, again indicating the higher solids removal and retention capability of the sunken media.

The sludge yield in the pilot reactors was low. During start-up the values were higher than at steady-state with 0.225 kg solids kg^{-1} of tCOD removed for the sunken media and 0.415 kg solids kg^{-1} of tCOD removed for the floating media (Table 3). The floating media carried out better SS and tCOD removal and produced average sludge yields of 0.310 kg solids kg^{-1} of tCOD removed (Table 3) at steady-state (Figure 8).

The sunken media did not perform as well and produced sludge yields of 0.200 kg solids kg^{-1} of tCOD removed (Figure 7). These values compare to 0.3 - 0.55 kg SST kg^{-1} COD

TABLE 3. Backwash Solids Produced per Kilogram Total COD Removed.

	Sunken Media (kg Solids / kg tCOD Removed)	Floating Media (kg Solids / kg tCOD Removed)
Start-Up	0.225	0.415
Steady-State	0.200	0.310

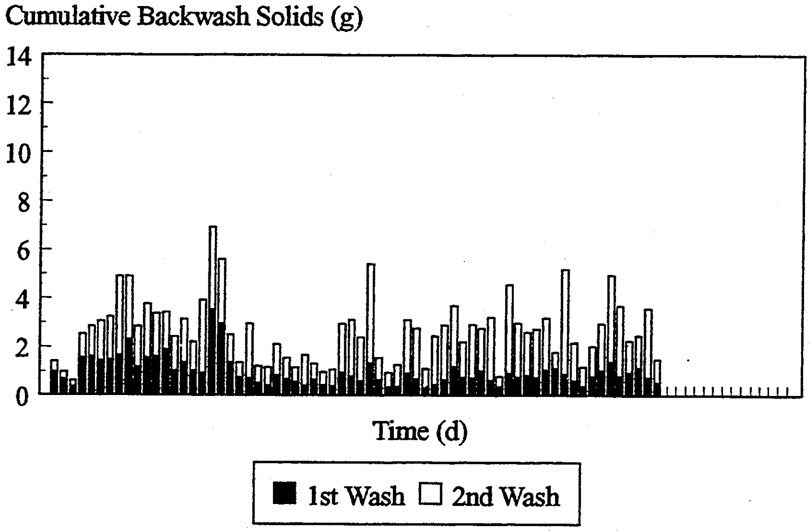

Figure 6a. Backwash Solids Removed During Backwashing (Sunken Media).

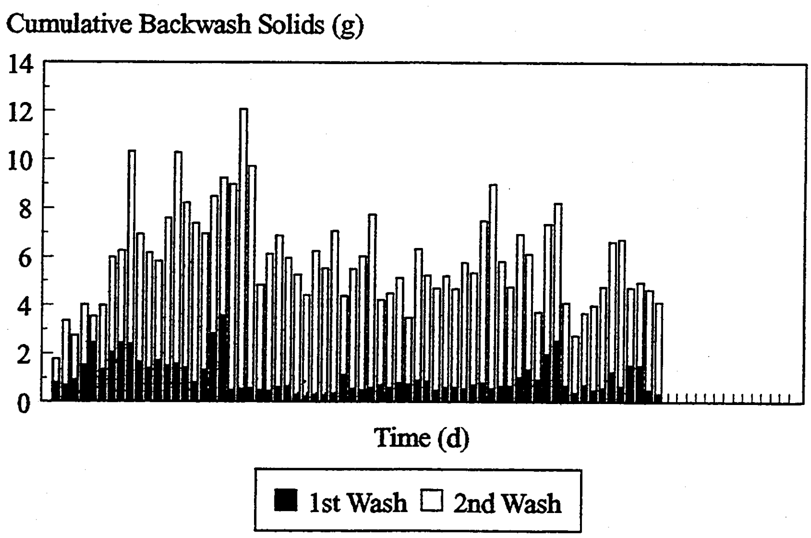

Figure 6b. Backwash Solids Removed During Backwashing (Floating Media).

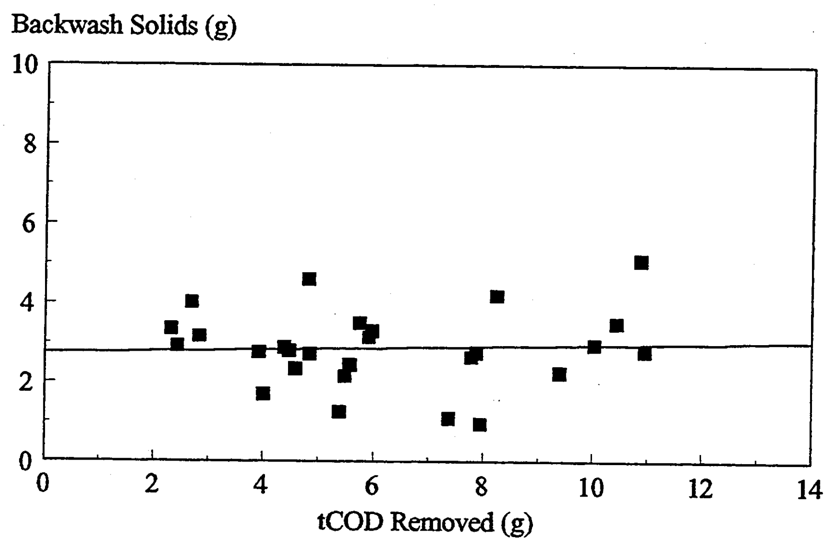

Figure 7. Solids Produced versus Total COD Removed During Steady-State (Sunken Media).

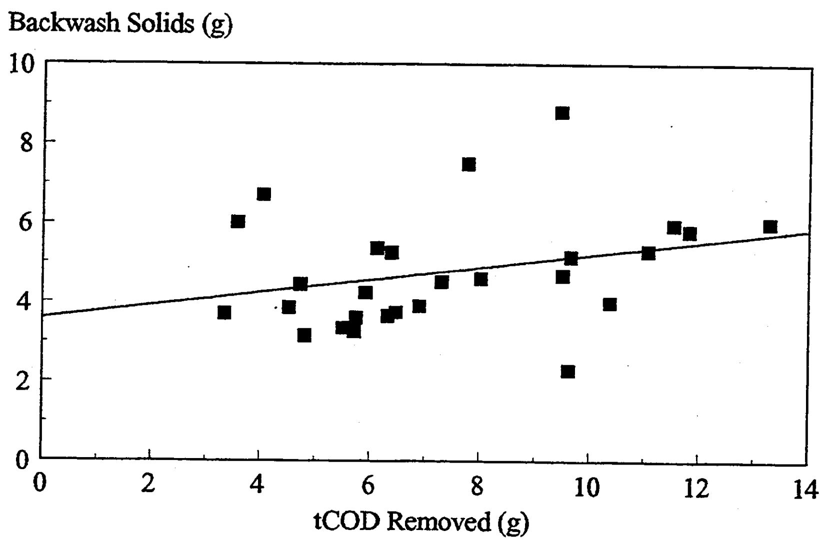

Figure 8. Solids Produced versus Total COD Removed During Steady-State (Floating Media).

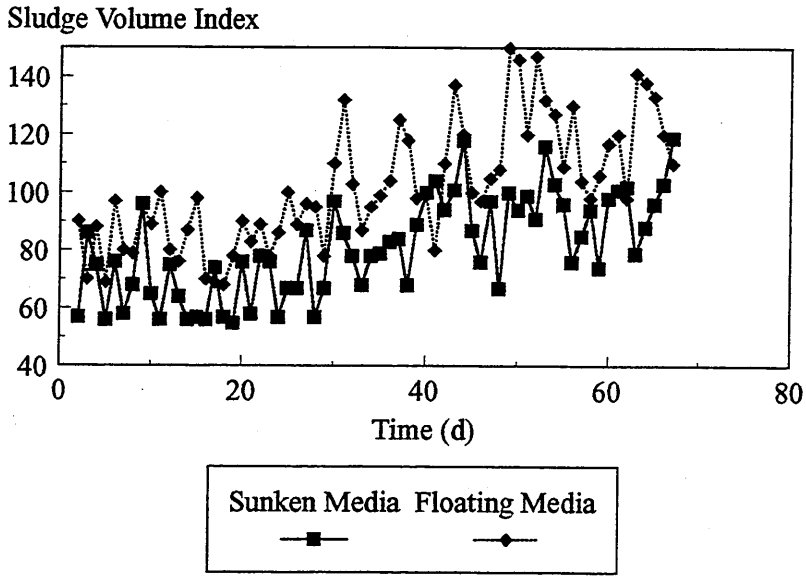

Figure 9. SVI Values for Backwashed Solids.

for other secondary treatment BAFs (Rusten, 1984; Jimenez *et al.*, 1987; Pujol *et al.*, 1992b). This indicated that a more rapidly growing but less stable biomass was formed during start-up than at steady state. The solids obtained from backwashing showed good settleabilty (Figure 9). During start-up the SVI of the solids obtained showed values of between 50 and 100. During steady-state the SVI values showed a marked increase. Though throughout the experimental period the backwashing solids were seen to fall rapidly the solids obtained during the start-up periods were observed to fall more rapidly than in the steady-state solids. This indicated that the solids obtained during the start-up period were primarily filtered solids with only small amounts of biomass whereas during the steady state period the proportion of biomass solids removed, increased.

Though the method of backwashing was simple and economical, it appeared that both medias tended to be overwashed at higher flowrates. Unlike underwashing which causes premature headloss and solids breakthrough, overwashing caused loss of effluent quality through reduced filtration capability and low biomass concentration (Robinson *et al.*, 1994). Thus unlike most reactors where recovery takes up to 2 h, the pilot scale reactors used in these experiments took up to 18 h to recover (Pujol et al., 1992; Robinson et al., 1994; Visvanathan and Nhien, 1995). It thus appeared that overwashing the pilot BAF reactors removed a large fraction of the active biomass, which reduced the filtration capability and tCOD removal efficiency. Subsequently a long lag time after backwashing was required for the growth of sufficient biomass to take place (Amar *et al.*, 1986). Generally only approximately 30% of the biomass is required to be removed without reducing reactor performance (Bacquet *et al.*, 1991).

The amount of liquid used in backwashing the pilot reactors was 10 l. This was equivalent to 3.5% of the treated water during the minimum flowrate of 0.2 l min^{-1} or 1.4% during the maximum flowrate of 0.5 l min^{-1}. This compared to 0.9 to 8.3% used in sand and anthracite filters and 2 to 15% used in BAFs (Ruffer and Roenwinkel, 1984: Paffoni *et al.*, 1990; Dillon and Thomas, 1990; Park and Ganczarczyk, 1994; Wheale and Cooper-Smith, 1995). Since no clogging occurred during the entire experimental period, it appeared that the method used was sufficient, but the use of only a small amount of partially treated water made this method very economical.

The overall results showed that when process liquid was passed upflow through sunken and floating granular media, the floating media performs better at suspended solids and total COD removal. By running floating media reactors upflow, the traditional gravity type filter was reversed. The liquid thus flowed in the direction of grain compression, unlike upflow reactors that use sunken media, thus the floating media performed better than the sunken media due to bed compression and showed higher total solids retention at higher flowrates (Rogalla and Bourbigot, 1990). The type of backwashing also affected the performance of the reactors. Backwashing downflow (draining) worked against the line of compression in the floating media and with the line of compression in the sunken media. Thus the floating media opened up during backwashing which allowed the solids to be removed more easily than in the sunken media (Jimenez *et al.*, 1987; Rogalla and Bourbigot, 1990).

In conclusion, ideal backwashing should be carried out on the basis of headloss, allowing optimum use of BAF reactors. Ideal solids filtration occurs with the direction of media compression, i.e.upflow in floating media and downflow in sunken media. Optimum backwashing of floating media reactors should be carried out downflow, against the line of bed compression and sunken media reactors should be backwashed in an upflow direction.

ACKNOWLEDGEMENTS

The author (A. Mann) is grateful to the Engineering and Physical Sciences Research Council (EPSRC) for a quota studentship.

REFERENCES

Amar, D., Partos, J., Granet, C., Faup, G.M., Audic, J.M. (1986). The Use of an Upflow Fixed Bed Reactor for Treatment of a Primary Settled Domestic Sewage. *Wat. Res.*, 20, 9-14.

American Public Health Association. (1992) Standard Methods for the Examination of Water and Wastewater. 18th Edition, American Public Health Organisation, Washington, D.C.

Anderson, B. (1989). Tentative Nitrogen Removal with Fixed Bed Processes in Malmo Sewage Treatment Plant. *Proc. Technical Advances in Biofilm Reactors,* Nice, April, 265-276.

Arvin, E., Harremoes, P. (1989). Concepts and Models for Biofilm Reactor Performance. *Proc. Technical Advances in Biofilm Reactors,* Nice, April, 191-212.

Bacquet, G., Joret, J.C., Rogalla, F., Bourbigot, M.M. (1991). Biofilm Start-Up and Control in Aerated Biofilter. *Environ. Tech.* 12, 747-756.

Boller, M. (1987). Nutrient Removal from Wastewater. *Proc. 7th EWPCA Symposium,* Munich, 253-278.

Buswell, A.M., Pearson, E.L. (1929). The Nidas (Nest) Rack, a Modern Development of the Travis Colloider. *Sew. Works Jour.*, 1, 187-195.

Carrand, G., Capon, B., Rasconi, A., Brenner, R. (1990). Elimination of Carbonaceous and Nitrogenous Pollutants by a Twin-Stage Fixed Growth Process. *Wat. Sci. Tech.*, 22, 1/2, 261-272.

Chudoba, J., Stakova, P., Kondo, M. (1991). Compartmentalized versus Completely-Mixed Biological Wastewater Treatment Systems. *Wat. Res.*, 25, 973-978.

Dillon, G.R., Thomas, V.K. (1990). A Pilot Scale Evaluation of the 'Biocarbone' Process for the Treatment of Settled Sewage and for Tertiary Nitrification of Secondary Effluent. *Wat. Sci. Tech.*, 22, 1/2, 305-316.

Faup, G.-M., Leprince, A., Pannier, M. (1982). Biological Nitrification in an Up Flow Fixed Bed Reactor (UFBR). *Wat. Sci. Tech.*, 14, 795-810.

Gray, T.W. (1993). Biological Aerated Filters: The Solution to Small Footprint Plant. *Int. Wat. Environ. Eng.*, Summer, 5-10.

Han-Chang, S., Thomas, V.K. (1993). A Study of Fibre Media Submerged Biological Aerated Filter - Activated Sludge Process (FM-AS) for Sewage Treatment. *Wat. Sci. Tech.*, 27, 5-6, 413-423.

Hoehn, R.C., Ray, A.D. (1973). Effects of Thickness on Bacterial Film. *J.WPCF.*, 45, 2302-2319.

Jepson, S.-E., Jansen, J.C. (1993). Biological Filters for Post-Denitrification. *Wat. Sci. Tech.*, 27, 5-6, 369-379.

Jimenez, B., Capedeville, B., Roques, H., Faup, G.M. (1987). Design Considerations for a Nitrification-Denitrification Process Using Two Fixed-Bed Reactors in Series. *Wat. Sci. Tech.* 19, Rio, 139-150.

Kirichenko, A.G., Drushlyak, O.G. (1988). Removal of Petroleum Products, SSAS and Biogenic Pollutants During Tertiary Treatment of Waste Water in Aerated Filters. 10, 151-154.

Kurbiel, J. (1987). Upflow Contact Filtration of Biologically Treated Wastewater for Water Reclamation. *Trib.Cebedeau,* 520, 33-37.

Mann, A., Fitzpatrick, C.S.B., Stephenson, T. (1995). A Comparison of Floating and Sunken Media Biological Aerated Filters Using Tracer Study Techniques. *Trans. IChemE., B,* 73, 137-143.

Nielson, P.H., Harremoes, P. (1995). Solids: Report of the Disscusion Session. *Wat. Sci. Tech.*, 32, 8, 273-275.

Paffoni, C., Gousailles, M., Rogalla, F., Gilles, P. (1990). Aerated Biofilters for Nitrification and Effluent Polishing. *Wat. Sci. Tech.*, 22, 7/8, 181-189.

Park, J.-W., Ganczarczyk, J.J. (1994). Gravity Separation of Biomass Washed Out from an Aerated Submerged Filter. *Environ. Tech.*, 15, 945-955.

Pujol, R., Canler, J.P., Vachon, A., Vidou, P. (1992a). Biological Aerated Filters: An Adapted Biological Process for Wastewater from Coastal Areas. *Wat. Sci. Tech.*, 25, 12, 175-184.

Pujol, R., Canler, J.P., Iwema, A. (1992b). Biological Aerated Filters: An Attractive and Alternative Biological Process. *Wat. Sci. Tech.*, 26, 3-4, 693-702.

Quickenden, J., Mittal, R., Gros, H. (1992). Efficient Nutrient Removal with Sulzer Biopur and Filtration Systems. *Europ. Conf. on Nutrient Removal from Wastewater*, Wakefield.

Richard, Y., Cyr, R. (1990). Les Possibilities de Traitment par Cultures Fixees Aerobies. *TSM.*, 7-8, 389-392.

Robinson, A.B., Brignal, W.J., Smith, A.J. (1994). Construction and Operation of a Submerged Aerated Filter Sewage Treatment Works. *J.IWEM.*, 8, 215-227.

Rogalla, F., Bourbigot, M.-M. (1990). New Developments in Complete Nitrogen Removal with Biological Aerated Filters. *Wat. Sci. Tech.* 22, 1/2, 273-280.

Ruffer, H., Rosenwinkel, K.-H. (1984). The Use of Biofiltration for Further Wastewater Treatment. *Wat. Sci.Tech.*, 16, Vienna, 241-260.

Rusten, B. (1984). Wastewater Treatment with Biological Aerated Filters. *J. WPCF.*, 56, 424-431.

Rusten, B., Odegaard, H. (1986). Treatment of Food Industry Effluents in Aerated Submerged Biological Filters. *Vatten*, 42, 187-193.

Ryhiner, G., Birou, B., Gros, H. (1994). The Use of Structured Packings in Biofilm Reactors for Wastewater Treatment. *Wat. Sci. Tech.*, 26, 3-4, 723-731.

Smith, A.J., Hardy, P.J. (1992). High-Rate Sewage Treatment Using Biological Aerated Filters. *J.IWEM.*, 6, 179-193.

Stevenson, D.G. (1994). The Specification of Filtering Materials for Rapid-Gravity Filtration. *J.IWEM*, 8, 527-533.

Valentis, G., Lesavre, J. (1989). Wastewater Treatment by Attached-Growth Micro-Organisms on a Geotextile Support. *Proc. Technical Advances in Biofilm Reactors*, Nice, April, 59-67.

Vedry, B., Paffoni, C., Gousailles, M., Bernard, C. (1994). First Months Operation of Two Biofilter Prototypes in the Waste Water Plant of Acheres. *Wat. Sci. Tech.* 29, 10-11, 39-46.

Visvanathan, C., Nhien, T.T.H. (1995). Study on Aerated Biofilter Process Under High Temperature Conditions. *Environ. Tech.*, 16, 301-314.

Wheale, G., Cooper-Smith, G.D. (1995). Operational Experience with Biological Aerated Filters. *J.CIWEM*, 9, 109-119.

Wilderer, P.A., Cunningham, A., Schindler, U. (1995). Hydrodynamics and Shear Stress: Report from the Discussion Session. *Wat. Sci. Tech.*, 32, 8, 271-272.

Chapter 8.

CONCLUSIONS AND RECOMMENDATIONS.

Conclusions.

1. In upflow BAF reactors solids removal was greatest using floating media with mean removal rates of 78 to 60% at loadings of 0.486 to 1.397 kg m^{-3} d^{-1} compared to 66 to 35% in the sunken media. Thus highest solids removal occurs when the liquid flow is in the direction of media compression. It can then be concluded that in downflow BAFs solids removal would be greater in sunken media than floating media reactors.

2. Floating media was shown to perform better than sunken media during shock loadings of up to 1.5 times the nominal flow. At these higher flowrates the reduction in suspended solids and soluble COD removal rates was less in the floating media than in the sunken media with a drop in removal rate of 6% in the sunken media and 3% in the floating media at low flowrates. This difference in performance appeared to be due to higher biomass washout in the sunken media than the floating media indicating weaker biomass attachment to the sunken media and thus a more unstable biomass.

3. Greater soluble COD and solids removal was seen in the upflow floating media BAF than in sunken media BAF. The floating media removed 75 to 40% and the sunken media 68 to 30% at loadings of 0.568 to 1.403 kg m^{-3} d^{-1}. This indicated that bed compression improves solids removal and promotes the growth of heterorotrophic bacteria.

4. During nitrification higher rates of ammonia removal were seen in the sunken media than in the floating media indicated that autotrophic bacteria growth is promoted using an open structured granular bed.

5. Low temperature shocks (4 °C from 8 °C) showed a greater effect on the sunken media than on the floating media indicating a more stable biofilm on the floating media though with increased temperatures (up to 10.2 °C) performance of the sunken media reactor showed a more rapid increase in performance. This indicated that the sunken media promoted rapid biofilm growth though the structure of the biofilm was more unstable than in the floating media reactor.

6. Backwashing was carried out on a daily basis during secondary treatment and weekly during tertiary treatment, and was more effective in the floating media than in the sunken media when carried out downflow, in opposition to the upflow of the process liquid flow direction.

7. Tracer profiles allowed the performance of BAFs in biological treatment to be predicted. The tracer studies indicated that downflow reactors should use sunken

media and upflow reactors should use floating media for the best treatment of wastewater.

8. Both upflow and downflow granular media BAFs were shown to have pseudo plug flow characteristics. The closeness to ideal plug flow depended on the rate of liquid flow and the aeration rate. At increased aeration rates the performance of reactors decreased through high shear and liquid bypassing but at low aeration rates oxygen transfer may be limited.

9. Soluble COD profiles along the length of BAF reactors have shown that in secondary treatment soluble COD removal was directly dependant on the influent soluble COD concentration. Subsequently the soluble COD rate of removal was found to be first order.

10. A simple first order empirical model was found directly relating soluble COD to effluent soluble COD and reactor height.

11. The start-up of BAFs using activated sludge seeding compared to using the process wastewater at the nominal flowrate showed a more rapid start-up and produced a more stable biofilm. Shock hydraulic loadings showed biofilm grown during low flowrates was unstable at higher flowrates.

12. During tertiary nitrification, ammonia removal was shown to be carried out at a reaction rate of between zero and half order. Thus ammonia removal was generally independent of ammonia concentration, though diffusion inhibition of ammonia could not be ruled out as a factor influencing nitrification.

13. The frequency and rate of backwashing had a great effect on reactor performance. Overwashing of reactors caused poor recovery and thus longer periods of time when the reactors are off-line especially during tertiary treatment..

Recommendations.

1. Direct comparisons are required between upflow and downflow reactors using identical media to access performance during biological treatment.

2. Different media dimensions such as size, shape and surface texture need to investigated and compared to access their effect on the biological performance and hydrodynamics of reactors.

3. Further investigations are needed using the empirical model produced in this work to look at different media types and larger reactor sizes .

4. An accurate mechanistic model is required that takes into account the different factors in reactor design such as flow direction, media type, aeration and backwashing rates.

5. Overall plant optimisation is needed to reduce costs especially in backwashing requirements and aeration rates.

Appendix 1.

ANALYTICAL METHODS.

Biochemical Oxygen Demand (BOD)
HMSO (1988). 5 Day Biochemical Oxygen Demand. Second Edition.

Conductivity
American Public Health Association. (1992). Standard Methods for the Examination of Water and Wastewater. 18th Edition, 2510B, 2-45.

Chemical Oxygen Demand (COD)
HACH (1992).Water Analysis Handbook, 2nd Edition, 494-505.

Dissolved Oxygen (DO)
American Public Health Association. (1992). Standard Methods for the Examination of Water and Wastewater. 18th Edition, 4500-OG, 4-103.

Nitrogen (Ammonia)
American Public Health Association. (1992). Standard Methods for the Examination of Water and Wastewater. 18th Edition, 4500-NH_3C, 4-78.

Nitrogen (Nitrate)
HMSO. (1981). Oxidised Nitrogen in Waters, Method E, 31-35..

Nitrogen (Nitrite)
HMSO. (1981). Oxidised Nitrogen in Waters, Method H, 49-53.

pH
American Public Health Association. (1992). Standard Methods for the Examination of Water and Wastewater. 18th Edition, 4500-H^+,B, 4-65.

Settled Sludge Volume
American Public Health Association. (1992). Standard Methods for the Examination of Water and Wastewater. 18th Edition, 2710D, 2-65.

Suspended Solids (SS).
American Public Health Association. (1992). Standard Methods for the Examination of Water and Wastewater. 18th Edition, 2540D, 2-56.

Temperature
American Public Health Association. (1992). Standard Methods for the Examination of Water and Wastewater. 18th Edition, 2550B, 2-59.

Appendix 2.

EXPERIMENTAL RIG DESIGN.

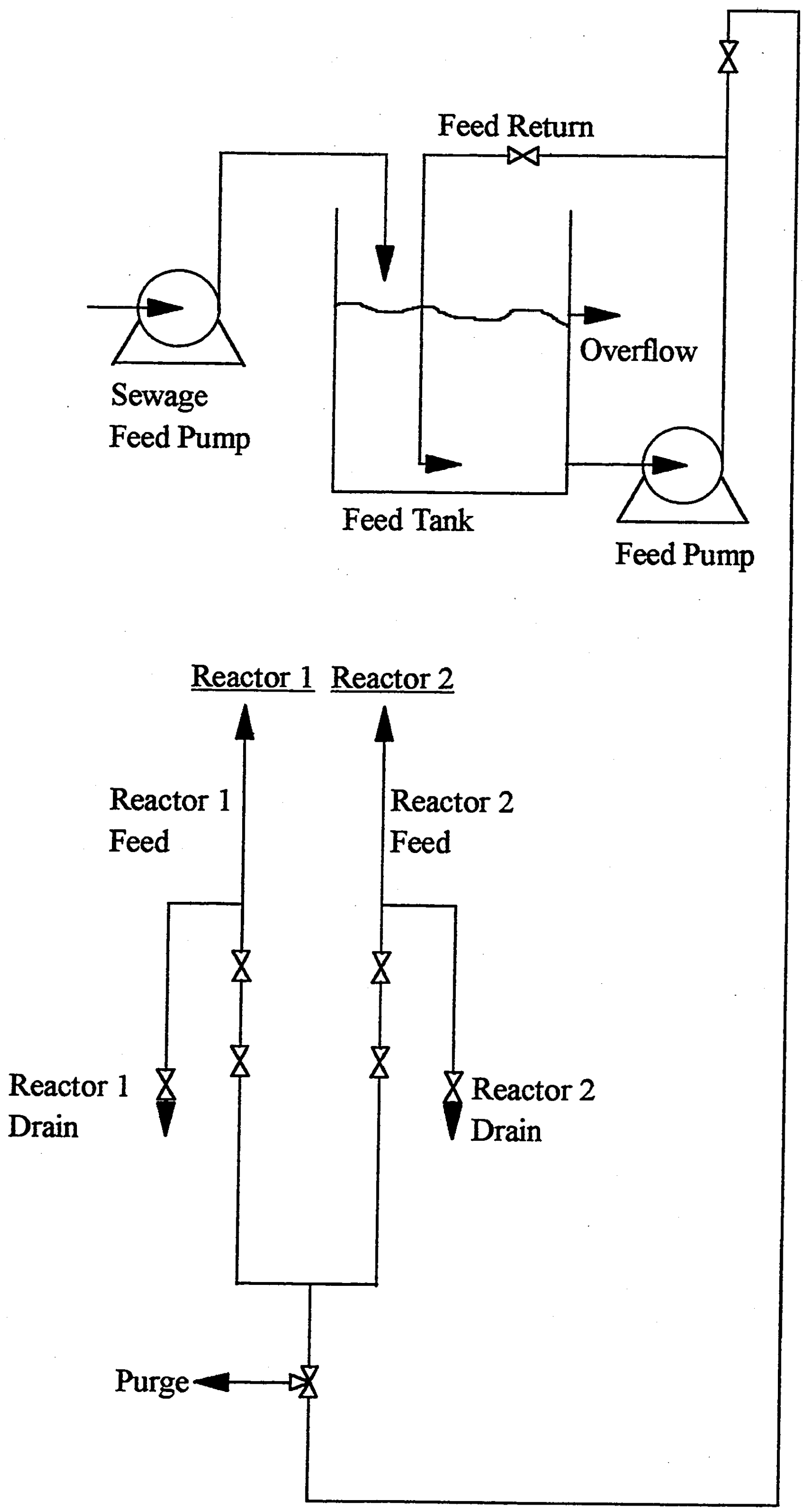

Figure 1. Process Feed System

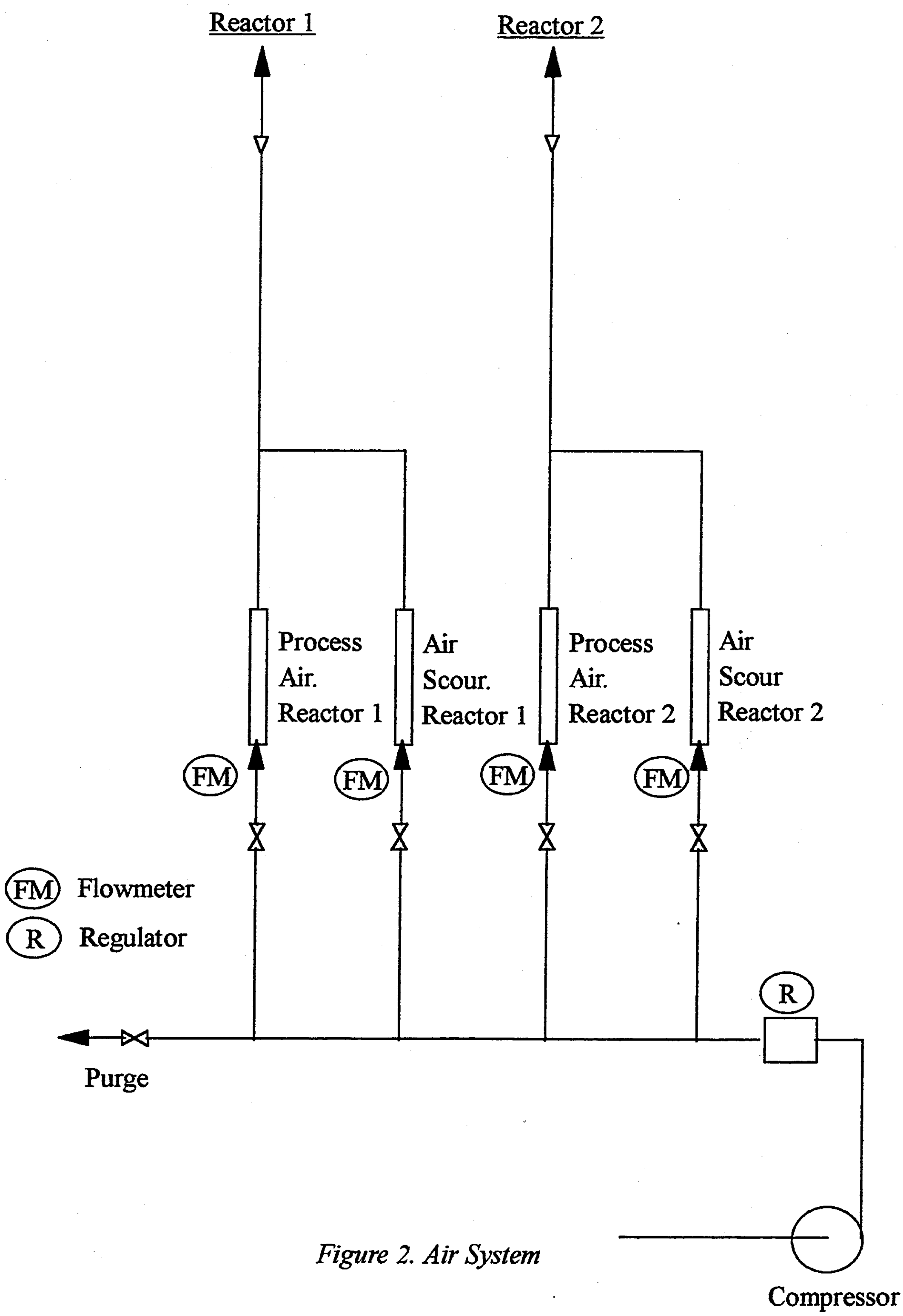

Figure 2. Air System

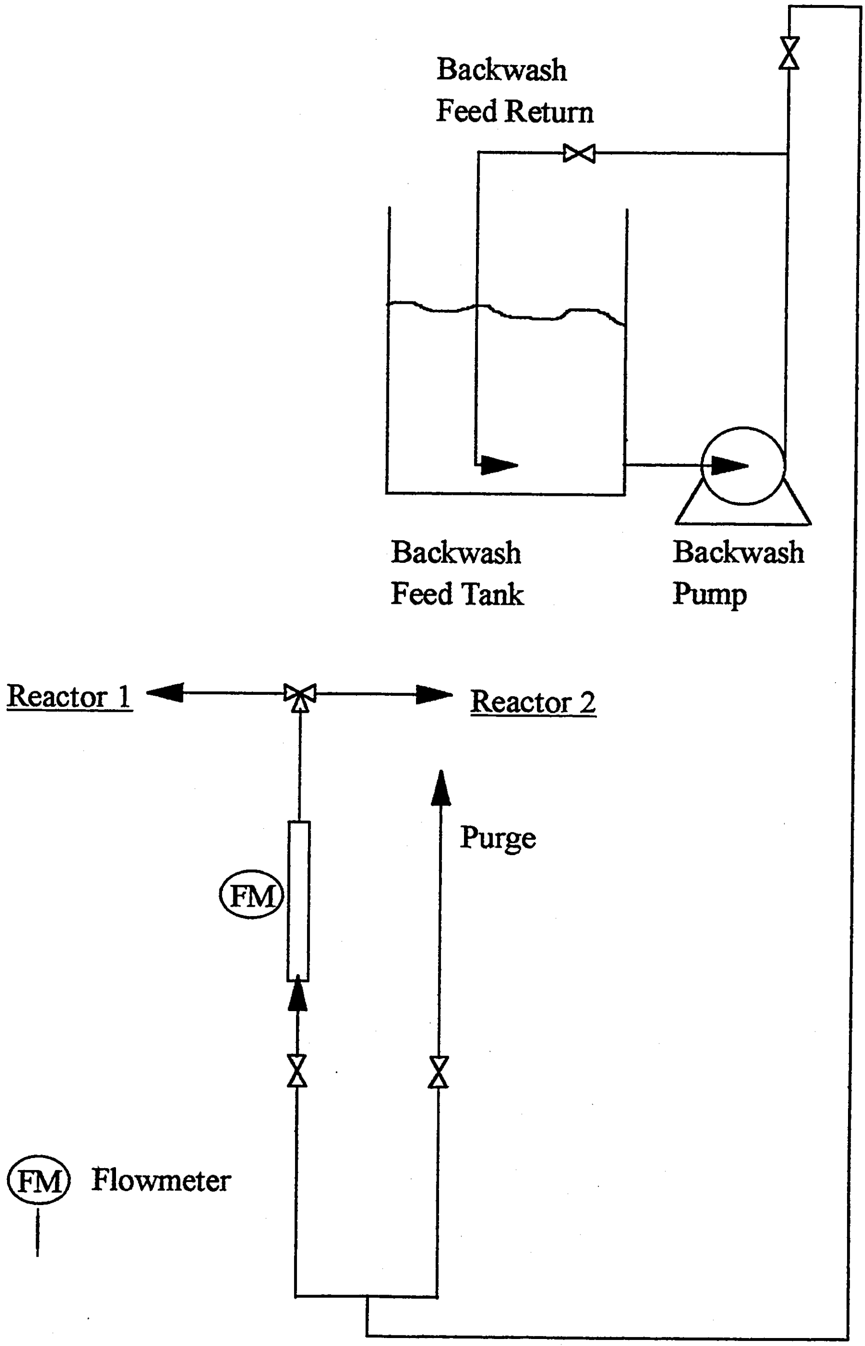

Figure 3. Backwash System

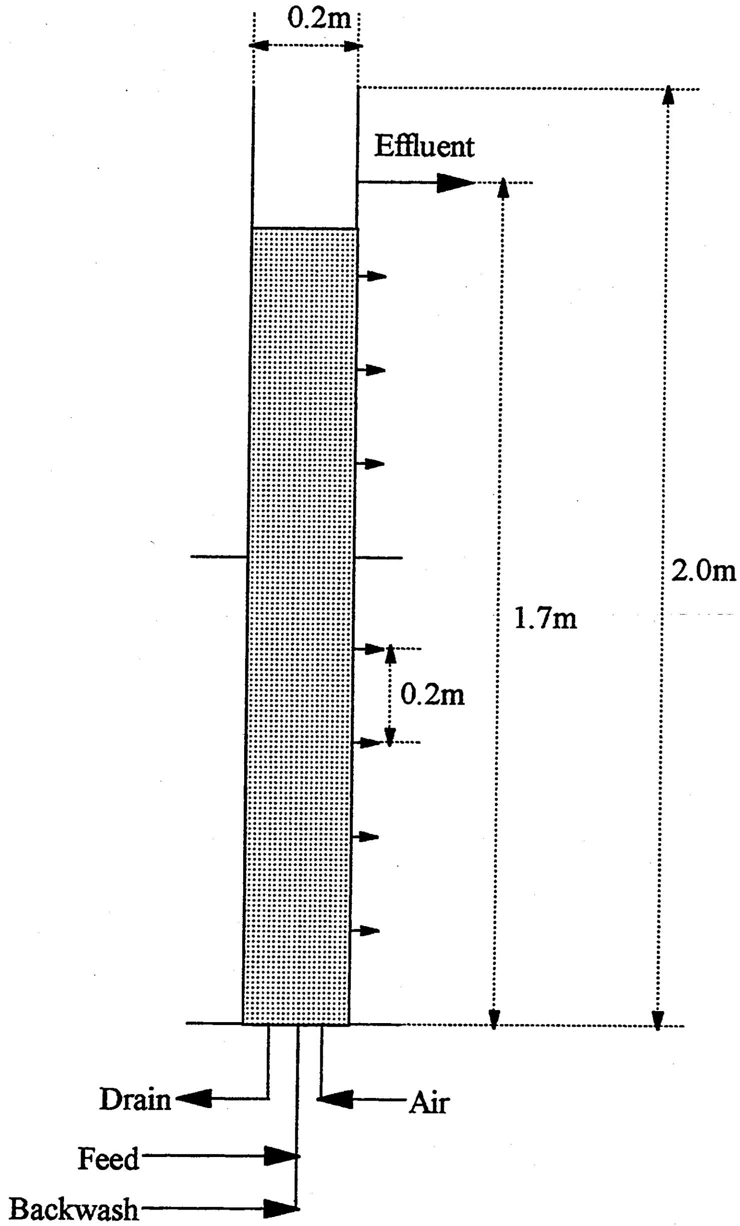

Figure 4. Upflow Reactor

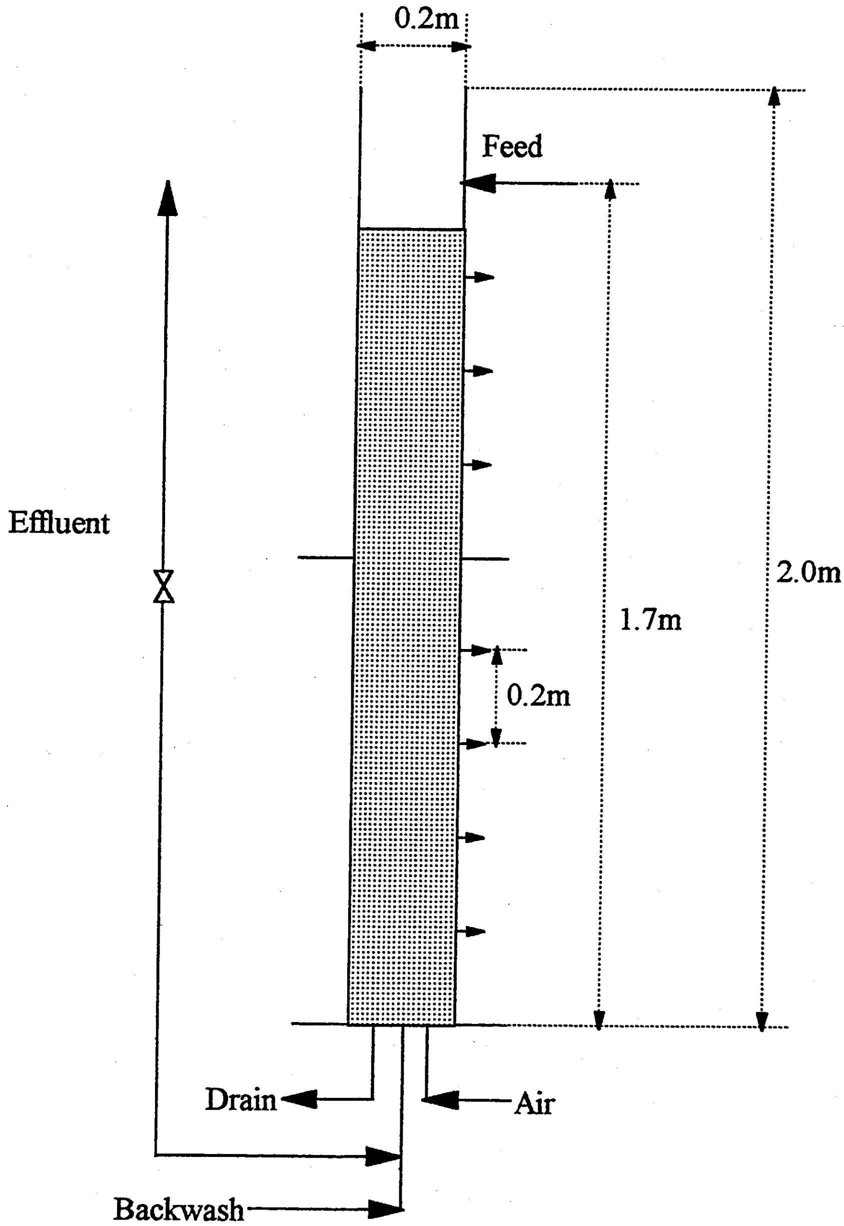

Figure 5. Downflow Reactor

Appendix 3.

OTHER PRESENTATIONS AND PUBLICATIONS.

Stephenson, T., Mann, A., Upton, J. (1993). The Small Footprint Wastewater Treatment Process. *Chem. Ind.*, 533-536.

Mann, A., Stephenson, T. (1993). Wastewater Treatment by Biological Aerated Filters - Latest Developments. *WET News*, May, 6-8.

Mann, A., Fitzpatrick, C., Stephenson, T. (1994). A Comparison of Floating and Sunken Media Support Biological Aerated Filters. *Proc 1994 IChemE Research Event*, London, 298-300.

Mann, A., Fitzpatrick, C., Stephenson, T. (1994). Performance of Floating and Sunken Media Upflow Biological Aerated Filters. *2nd Int. Symp. Environ Biotech.*, Brighton, UK..

Stephenson, T., Mann, A. (1995). Biological Aerated Filters - Here to Stay. *Wat. Waste Treat.*, 38, 25-26.

Mann, A., Fitzpatrick, C., Stephenson, T. (1995). The Biological Performance of Floating and Sunken Media Biological Aerated Filters. *Proc. 1995 IChemE Research Event*, Edinburgh, 387-389.

Mann, A. (1995). Can BAFs be Modelled for STOAT ? *1st STOAT User Group Meeting*, Dec.

Mann, A., Stephenson, T. (1996). Pilot Scale Comparisons of Floating / Sunken Media and Up / Downflow BAFs. *Proc. 2nd BAF Symp.*, Cranfield, UK.

Mann, A., L. Mendoza-Espinosa, Stephenson, T. (1996). A Comparison of Pilot-Scale Biological Aerated Filters Using Sunken and Floating Media. *3rd Conf. of IAWQ on Biofilm Systems*, Copenhagen.

Mann, A., Stephenson, T. (1996). Performance of Floating and Sunken Media Biological Aerated Filters Under Unsteady State Conditions. *18th Biennial Conf. of IAWQ*, Singapore.

Mendoza-Espinosa, L., Mann, A., Stephenson, T. (1997). Determination of Flow Pattern and Active Volume in Biological Aerated Filters Under Upflow and Downflow Conditions. *Proc 1997 IChemE Research Event*, Nottingham.