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LEOPOLDO GUILLERMO MENDOZA-ESPINOSA

Upflow and downflow biological aerated filters (BAFs)
for wastewater treatment

SCHOOL OF WATER SCIENCES

PhD THESIS
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1999

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Upflow and downflow biological aerated filters (BAFs) for
wastewater treatment

Supervisor: Prof. Tom Stephenson

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

There's gotta be an easier way, man!

Bart Simpson

ABSTRACT

Investigations on two small pilot-scale biological aerated filters operating in upflow and downflow mode using identical media were undertaken in order to compare their performance for the removal of carbonaceous matter and ammonia, calibration of empirical models for suspended solids, soluble and total COD and their resistance to organic and hydraulic shock loadings. The initial existence of a plenum chamber under the media bed caused severe reduction in active volume in the downflow reactor. Elimination of the plenum chamber by means of the addition of media solved the problem and both reactors presented plug-flow conditions. Their performance for carbonaceous matter and ammonia removal was compared. With the exception of ammonia, the removal efficiencies for the parameters studied in both types of reactor were very similar. The downflow reactor achieved a greater removal of ammonia than the upflow BAF. Autotrophic activity was located at the bottom of the reactor in the downflow column and at the top in the upflow unit. Had the running conditions been set identically for both columns, the difference in performance between the two reactors may have been due to the impact of the backwashing on the biological population, effectively washing-out the autotrophs. The verification of a simple empirical model for soluble COD removal for a different type of media than the previously reported was conducted. The new values for the media constant and the overall process constant of the system suggested that overall, the performance of the system was dictated by its media type/liquid flow characteristics. The constants from 4 empirical models were used to predict the removal of SS, sCOD and tCOD in a small BAF (0.07 m³ empty bed volume) and a large BAF (0.69 m³ empty bed volume). The model for soluble COD that included parameters such as gas and liquid velocity provided values closer to the actual observed values. The model for suspended solids and tCOD removal (based on a rapid gravity filters model) was efficient in predicting the overall removal rates of SS and tCOD in both reactors. When the models were used for scaling-up results, only the simplest model that did not include parameters such as gas and liquid velocity was effective. It was demonstrated that a small pilot-scale BAF was not capable of absorbing high peak organic shock loads when the organic loads were increased from a normal load of 1.2-1.4 kg sCOD m⁻³ d⁻¹ to shock loads between 5.1- 7.3 kg sCOD m⁻³ d⁻¹, based on empty bed volumes. Nevertheless, the effect on the biomass was limited as normal performance resumed very quickly. Increases in the hydraulic velocity from 0.7-1.0 m h⁻¹ to 1.5-2.9 m h⁻¹ had little affect on BAF performance in terms of soluble COD removal. Soluble COD effluent concentrations of <40 mg l⁻¹ (~16 mg l⁻¹ BOD₅) were detected thus, meeting most discharge consents in UK sewage works. The recovery of the reactor in both cases was fast, resuming normal performance within 60 min after the application of the shock load.

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UNITS

atm	Atmosphere
s	Second
min	Minute
h	Hour
d	Day
g	Gram
μm	Micrometre ($1 \times 10^{-6} \text{ m}$)
mg	Miligram ($1 \times 10^{-3} \text{ g}$)
kg	kilogram ($1 \times 10^3 \text{ g}$)
m	Meter
mm	Milimeter ($1 \times 10^{-3} \text{ m}$)
cm	Centimeter ($1 \times 10^{-2} \text{ m}$)
m^2	Square meter
l	Litre
m^3	Cubic meter ($1 \times 10^3 \text{ l}$)
M	Molar solution

NOTATION

AF	Air flowrate
BAF	Biological aerated filter
BOD	Biochemical oxygen demand
BOD ₅	Five day biochemical oxygen demand test
CO ₂	Carbon dioxide
COD	Chemical oxygen demand
EBV	Empty bed volume
EOD	Estimated oxygen demand
IAWQ	International Association on Water Quality
LF	Liquid flowrate
MRE	Mean relative error
N ₂	Nitrogen
NH ₃ -N	Nitrogen as ammonia
NH ₄ -N	Nitrogen as ammonium
NO ₃ -N	Nitrogen as nitrate
O ₂	Oxygen
OTE	Oxygen transfer efficiency
PBR	Packed-bed bioreactor
pe	Population equivalent
pH	Potential hydrogen
PVC	Polyvinyl chloride
RBC	Rotating biological contactor
RTD	Residence time distribution
sCOD	Soluble chemical oxygen demand
SEM	Scanning electron micrographs
SS	Suspended solids
tCOD	Total chemical oxygen demand
tHRT	Theoretical hydraulic retention time
TKN	Total Kjeldahl nitrogen
UASB	Upflow anaerobic sludge bed reactor
UV	Ultra violet

SYMBOLS

Chapter 4.

c	Concentration of tracer (mg l^{-1} or mol l^{-1})
D	Longitudinal or axial dispersion coefficient ($\text{m}^2 \text{s}^{-1}$)
u	Liquid velocity (m s^{-1})
L	Length of reactor (m)
σ_θ^2	Variance
N	Number of tanks in series
t_m	Mean residence time
t_p	Peak residence time

Chapter 6.

C	COD concentration (mg l^{-1})
T	Time
C_{in}	Influent COD concentration (mg l^{-1})
k	Rate constant
k'	Biomass constant
B_V	Volumetric loading
Q	Volumetric flowrate ($\text{m}^3 \text{d}^{-1}$)
H	Height of reactor (m)
A	Cross sectional area of reactor (m^2)
m	Values of the slope dependent on C_{in}
n	Media constant
k^*	Process constant

Chapter 7.

C_{in}	Concentration of sCOD at influent (g m^{-3})
C_{out}	Concentration of sCOD at effluent (g m^{-3})
Q	Influent flowrate ($\text{m}^3 \text{d}^{-1}$)
A	Cross sectional area of reactor (m^2)
H	Effective depth of column (m)
m	A parameter dependent on the specific media used
k	A pseudo rate constant (m^{-1})
α	Heterogenous coefficient
γ	Correction factor
U_G	Superficial velocity of air (m s^{-1})
U_L	Superficial velocity of liquid (m s^{-1})
z	Parameter constant for height
l	Parameter constant for gas velocity
p	Parameter constant for liquid velocity
C_{in}	Weight percentage of SS and/or tCOD in the influent (dimensionless)
C_{out}	Weight percentage of SS and/or tCOD in the effluent (dimensionless)
V_L	Liquid superficial velocity (m s^{-1})
λ_0	Filter coefficient (m^{-1})
σ_s^{-1}	Ultimate specific deposit (dimensionless)
τ	Corrected time variable (s)
t	Filter running time (s)
ε_0	Porosity of media (0.4)

Chapter 8.

p	Absolute pressure (atm)
V	Gas volume (m^3)
N	Moles (gmol)
R	Ideal gas constant ($8.2057 \times 10^{-5} \text{ m}^3 \text{atm gmol}^{-1} \text{ }^\circ\text{K}^{-1}$)
T	Absolute temperature ($^\circ\text{K}$)
τ	Dimensionless time

PUBLICATIONS

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- Chapter 5* Mendoza-Espinosa, L. G. and Stephenson, T. (1997) "Nitrification comparison in a downflow and an upflow BAF - Preliminary results". *Proceedings of the International Conference on Advanced Wastewater Treatment Processes*, University of Leeds, 8-11 September 1997. (in press)
- Chapter 6* Mendoza-Espinosa, L. G. and Stephenson, T. (1998). "A process model to evaluate the performance of a biological aerated filter". *Biotechnology Techniques*, **12**(5), 373-376.
- Chapter 7* Mendoza-Espinosa, L. G. and Stephenson, T. "Process models for the removal of suspended solids and organic matter, and the scale-up of biological aerated filters". *IAWQ 4th International Conference on Biofilm Systems*, New York, USA. (submitted)
- Chapter 8* Mendoza-Espinosa, L. G. and Stephenson, T. "Organic and hydraulic shock loadings on a biological aerated filter". *Environmental Technology*. (submitted)

Chapter 1

INTRODUCTION AND SYNOPSIS

INTRODUCTION AND SYNOPSIS

Biological aerated filters (BAFs) are submerged fixed media wastewater treatment processes that combine aerobic biological treatment and biomass solids separation by depth filtration. The advantages of the BAF process over other secondary treatment processes are their high organic removal rates per volume of reactor and the high quality effluent produced. Organic removal rates of up to $4.1 \text{ kg BOD m}^{-3} \text{ d}^{-1}$ are normally reported, while trickling filters, oxidation ditches and activated sludge plants normally achieve removal rates of $0.06 \text{ kg BOD m}^{-3} \text{ d}^{-1}$, $0.35 \text{ kg BOD m}^{-3} \text{ d}^{-1}$ and $0.42 \text{ kg BOD m}^{-3} \text{ d}^{-1}$, respectively (Smith and Hardy, 1992). As a consequence, the area required for BAFs can be one fourth or one fifth of that needed for activated sludge and trickling filters, respectively. Hence, BAFs have become an attractive alternative to other biological processes in sewage works that need to be upgraded especially where land availability is limited like in big metropolis and coastal areas.

BAFs can be used to achieve simultaneous removal of carbonaceous BOD and solids (secondary treatment), ammonia and solids (tertiary treatment) and/or removal of solids, carbonaceous BOD and ammonia (combined secondary and tertiary treatment) in a single operating unit. Moreover, additional design alterations allow for denitrification and phosphorus removal. Biological activity within the reactor is maintained by injecting oxygen into the system via compressed air. Granular media are used as a biomass support media and solids filtration. Therefore, periodic backwashing is necessary to avoid blocking of the system (Condren, 1990).

The type of media used is one of the most important factors when designing BAFs. Media parameters such as density, size, shape and surface texture will affect the performance of the reactor.

BAFs can be operated at different hydraulic configurations, under upflow or downflow conditions. As the air is always injected at the bottom of the reactor, during upflow conditions, air and water are running co-currently while under downflow conditions air

and water are running counter-currently along the reactor. Downflow reactors are limited to a maximum flowrate, which is dependent on the height and pressure drop across the reactor bed. Upflow reactors do not have this restriction, though the flowrate will depend on the pump capacity.

Since the late 1980s, work on the development of BAFs has grown with a large number of proprietary system now available (Table 1.1). Due to the high energy consumption associated with aeration and backwashing of BAFs (Carrand *et al.*, 1990), application has been limited. In order to decrease costs associated with backwashing, investigations on recirculating media BAFs, similar to continuous filtration systems, have been undertaken (Stephenson, 1996; Daamen *et al.*, 1996). The process has similar benefits to the original BAF design, though it is capable of solids, carbonaceous and ammonia removal in a single reactor. However, these processes have yet to be optimised and commercialised.

Table 1.1. Proprietary biological aerated filters.

Process Name	Media	Liquid Flow	NH ₃ -N loading rate (kg m ⁻³ d ⁻¹)	BOD Loading rate (kg m ⁻³ d ⁻¹)	Supplier
Biobead	Floating	Upflow	0.3	3 - 5	Brightwater Engineering Ltd.
Biocarbene	Sunken	Downflow	0.3 (TKN)	2.5 - 3.5	OTV - Birwelco
Biofor	Sunken	Upflow	N/A	2 - 3, up to 6	Degremont Ltd.
Biostyr	Floating	Upflow	1.0	N/A	OTV - Birwelco
ColOX	Sunken	Up/Downflow	0.4-0.8	4	Capital Controls
SAFe	Sunken	Downflow	0.35	3 (95% rem.)	Paterson-Candy Ltd.
REBAF*	Floating	Up/Downflow	N/A	N/A	Comenco
Biogenerator*	Floating	Downflow	0.6 (design)	6 (design)	RCL (Heerenveen, Holland)

N/A. Not Available

*recirculating media systems

Biobead: Developed by Brightwater Engineering. The media consists of buoyant plastic beads, specially designed to have a rough surface, with bed heights of 2-3 m. The majority of the plants are in the south west of England.

Biocarbone: Developed by the French Omnium de Traitement et de Valorisation (OTV), it is the most established commercial process with >40 plants worldwide (Rogalla and Sibony, 1992). It uses expanded shale of 3-6 mm diameter as a media for BOD/COD removal or 2-5 mm diameter media for nitrification. Media depth is approximately 2.5 m.

Biofor: Degremont's Biofor has been used in more than 30 European sites, including a commercial unit on the south coast of England in Poole (Wessex Water). It uses 1-4 mm diameter expanded clay media called 'Biolite' with bed depths that vary between 3 to 4 m.

Biostyr: Devised by OTV-Birwelco, Biostyr uses floating expanded polystyrene beads of 2-4 mm diameter media called 'Biostyrene'. In the past five years the technology has been used in over 20 installations worldwide.

ColOX: developed by Tetra - The Capital Controls Group, an American company specialising in filtration. Granular sunken silica sand with a diameter of 3-7 mm is used as media with a bed depth of approximately 4 m. There is a full-scale plant in the UK, treating a mixed municipal/dairy effluent.

SAFe: Originating from Thames Water and Process Water Technology Ltd (PWT) in the mid 1980s and now marketed by Paterson Candy Ltd (PCL). SAFe Uses granular Pulverised Fuel Ash (PFA) media, with a diameter of 3-5 mm and bed depths of approximately 1.9 m.

ReBAF: Devised by Comenco, this recirculating media BAF uses small plastic floating media with a bed height of up to 4 m. There are currently no full-scale systems, with the process presently undergoing pilot-plant trials (Stephenson, 1996).

Biogenerator: Patented by RCL (Patent WO 93/01137) the process was developed in The Netherlands (Daamen *et al.*, 1996). The recirculating media BAF uses spherical floating media with a diameter greater than 18 mm in bed heights up to 10 m. It has been tested in domestic and industrial wastewater treatment applications.

Many different BAF configurations exist and a great number of papers about the performance of specific proprietary units are available (Rogalla *et al.*, 1990; Pujol *et al.*, 1994; Peladan *et al.*, 1997). However, few investigations make direct performance comparisons between these different configuration options.

Therefore, the overall objective was to undertake investigations on two small pilot-scale biological aerated filters operating in upflow and downflow mode. Specific objectives were to: assess the mixing characteristics of two BAFs under upflow and downflow conditions with the use of tracer studies; compare the performance of the upflow and downflow BAFs for combined soluble carbonaceous matter and ammonia removal; test the applicability of a simple empirical model for soluble chemical oxygen demand (sCOD) for the upflow BAF using sunken media; examine the suitability of empirical models for scaling-up purposes by obtaining data from a large pilot-scale BAF using identical media to the small pilot-scale columns and monitor the performance of the small pilot-scale BAF under organic and hydraulic shock loading conditions.

After reviewing the current state-of-the-art of design, performance and optimisation of biological aerated filters (Chapter 3), it was concluded that areas in which further study was required included mechanisms of solids removal, aeration optimisation by monitoring oxygen utilisation through off-gas analyses, optimisation of backwashing and media characteristics, monitoring resistance to shock loadings, comparison of upflow and downflow systems and evaluation of models for scaling-up purposes.

Tracer studies in a downflow BAF and an upflow BAF found that the hydrodynamic mixing was not greatly affected by high air velocities (Chapter 4). The retention times

did not decrease as expected, probably due to the high density of the media used (2500 kg m^{-3}). The existence of a plenum chamber under the media bed caused severe reduction in active volume in the downflow reactor due to the entrapment of air bubbles. Increases in the nominal liquid flowrate from 0.4 l min^{-1} (hydraulic velocity of 1.8 m h^{-1}) to 0.8 l min^{-1} (3.6 m h^{-1}) affected the hydrodynamic conditions of the reactors in a greater way than changes in the air flowrate. Both columns presented similar hydraulic conditions, which is important for scaling-up purposes.

Two pilot-scale BAFs were operated with identical media under downflow and upflow conditions in order to compare their performance during combined soluble COD and ammonia removal (Chapter 5). At loading rates of around $1.8 \text{ kg SS m}^{-3} \text{ d}^{-1}$, $2.4 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$, $5.2 \text{ kg tCOD m}^{-3} \text{ d}^{-1}$ and $0.51 \text{ kg NH}_3\text{-N m}^{-3} \text{ d}^{-1}$ (based on a working volume in the media section of 25 l in each reactor), removal rates for of SS, soluble and total COD for both reactors were very similar. The upflow reactor achieved average removal rates of $1.3 \text{ kg SS m}^{-3} \text{ d}^{-1}$, $1.9 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$ and $3.9 \text{ kg tCOD m}^{-3} \text{ d}^{-1}$. The downflow reactor reached average removal rates of $1.4 \text{ kg SS m}^{-3} \text{ d}^{-1}$, $1.6 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$ and $4.2 \text{ kg tCOD m}^{-3} \text{ d}^{-1}$ (based on a working volume in the media section of 25 l in each reactor). Plug-flow conditions and autotroph-heterotroph competition caused autotrophic activity to be located at the bottom of the reactor in the downflow column and at the top in the upflow unit. Co-current air and water backwashing could have washed-out autotrophs in the upflow BAF.

The verification of a simple empirical model for soluble COD removal for a different type of media than the previously reported by Mann and Stephenson (1997) was conducted (Chapter 6). The media constant (n) was found to have a value of 0.97 and the overall process constant of the system (k^*) was 46 suggesting that overall, the performance of the system was dictated by the liquid flow pattern and that upflow reactors with floating media are as effective as downflow reactors with sunken media.

The constants from 3 empirical models for soluble COD removal were used to predict the removal of sCOD in a small BAF ($0.07 \text{ m}^3 \text{ EBV}$) and a large BAF ($0.69 \text{ m}^3 \text{ EBV}$)

(Chapter 7). From the three models tested, the one including parameters such as gas and liquid velocity provided values closer to the actual observed values, when the corresponding constants were used for verifying the models. All three models were successful in predicting overall performance of the reactor based on organic loading rate and when predicting the profile removal of sCOD. A model previously used for rapid gravity filters was used for the prediction of SS and tCOD in a small and a large BAFs. The model was efficient in predicting the overall removal rates of SS and tCOD in both reactors. The model helped to prove that by reducing the height of the large BAF from 3.9 m to 2.9 m the efficiency of the reactor would not be adversely affected. For scaling-up purposes, model 1 was the most effective for predicting the removal of sCOD in the large pilot-scale BAF when using constants from the small pilot-scale BAF. The lack of kinematic similarity (differences in the organic and hydraulic rates) and geometric similarity (smaller media diameter/ reactor diameter ratio) between reactors seem to have limited the capacity for scaling-up of models 2, 3 and 4, which did take into account parameters like sCOD influent concentration, air and gas velocity.

The performance of a biological aerated filter (BAF) during organic and hydraulic shock loads was investigated. Off-gas analysis of O_2 was undertaken to try to establish a correlation between off-gas monitoring and BAF performance for the secondary treatment of wastewater. The off-gas analysis technique did not appear to be useful for monitoring the impact of organic and hydraulic shock loads on the biological population of the BAF. The operational oxygen transfer efficiencies (OTEs) of the system were calculated based on the oxygen off-gas concentrations. Increases in the organic and hydraulic loading to the BAF did not appear to have a significant impact on the OTE as other studies suggest. The BAF used for the present experiment was not capable of absorbing high peak organic shock loads when the organic loads were increased from a normal load of $1.2\text{--}1.4 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$ to shock loads between $5.1\text{--}7.3 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$, based on empty bed volumes. Nevertheless, the effect on the biomass was limited as normal performance resumed very quickly. Increases in the hydraulic velocity from $0.7\text{--}1.0 \text{ m h}^{-1}$ to $1.5\text{--}2.9 \text{ m h}^{-1}$ had little effect on BAF

performance in terms of soluble COD removal. Soluble COD effluent concentrations of $<40 \text{ mg l}^{-1}$ ($\sim 16 \text{ mg l}^{-1} \text{ BOD}_5$) were detected thus, meeting most discharge consents in UK sewage works. The recovery of the reactor in both cases was fast, resuming normal performance within 60 min after the application of the shock load.

Chapter 2

OBJECTIVES

OBJECTIVES

The overall objective was to undertake investigations on pilot-scale biological aerated filters operating in upflow and downflow mode. Specific objectives were to:

1. Review the current state-of-the-art of design, performance and optimisation of biological aerated filters (Chapter 3).
2. Assess the mixing characteristics of two BAFs under upflow and downflow conditions with the use of tracer studies (Chapter 4).
3. Compare the performance of the upflow and downflow BAFs for combined soluble carbonaceous biochemical oxygen demand (BOD) and ammonia removal (Chapter 5).
4. Test the applicability of a simple empirical model for soluble chemical oxygen demand (sCOD) for a BAF using sunken media (Chapter 6).
5. Examine the suitability of empirical models for scaling-up purposes by obtaining data from a large pilot-scale BAF using identical media to the small pilot-scale columns (Chapter 7).
6. Assess the performance of the small pilot-scale BAF under organic and hydraulic shock loading conditions (Chapter 8).

Chapter 3

A REVIEW OF BIOLOGICAL AERATED FILTERS
(BAFs) FOR WASTEWATER TREATMENT

A REVIEW OF BIOLOGICAL AERATED FILTERS (BAFs) FOR WASTEWATER TREATMENT

ABSTRACT

Biological aerated filters (BAFs) can combine ammonia, carbonaceous and solids removal in a single unit process. Biological nutrient removal (nitrogen and phosphorus) can also be accomplished. Removal rates based on reactor volume for carbonaceous BOD, ammonia and nitrates of up to $4.1 \text{ kg BOD m}^{-3} \text{ d}^{-1}$, $1.27 \text{ kg NH}_3\text{-N m}^{-3} \text{ d}^{-1}$ and $5 \text{ kg NO}_3\text{-N m}^{-3} \text{ d}^{-1}$ respectively, are normally reported. The small footprint and adaptability of BAFs allows them to be used in upgrading established works, especially those in built up areas where space is at a premium. Areas in which further study is required include mechanisms of solids removal, aeration optimisation by monitoring oxygen utilisation through off-gas analyses, optimisation of backwashing and media characteristics, monitoring resistance to shock loadings, comparison of upflow and downflow systems and evaluation of models for scaling-up purposes.

3.1 Introduction

Submerged aerated filters are biofilm systems used for secondary and tertiary biological treatment of wastewater (Stensel *et al.*, 1983). They consist of three phases: a solids phase that acts as the support media for microbial growth, a liquid phase in which the solid material is submerged and a gas phase created by the input of air into the reactor. The media in submerged aerated filters can be structured or granular. When granular media is used, the system is capable of removing organic matter and suspended solids from wastewater at the same time; thus, there is no need for a subsequent solids separation stage i.e., sedimentation tanks. Periodic backwashing is necessary to avoid blocking of the system due to solids entrapment and biofilm growth. This review focuses on granular media submerged biological aerated filters, also known as biological aerated filters (BAFs), which have the unique capability of achieving solids, carbonaceous organic matter and nutrient removal in a single reactor.

3.1.1 History and Origins

Although true BAFs, capable of removing solids and organic matter without requiring additional sedimentation, did not appear until the end of the 1980s (Pujol *et al.*, 1994) their origins can be traced back to contact aerators, which have been used for more than 50 years (Rusten, 1984). Probably the first report of these reactor types dates from 1913, where aerated tanks containing layers of slate supporting a biofilm were employed for sewage treatment (Clark, 1930). According to Buswell and Pearson (1929) many filters using fine grade natural material as media had been used in experimental and large-scale installations but by the end of the 1920s they had not come into general use due to the problem of clogging. Cotton, hemp, copper gauze, galvanised iron gauze, lath, brushwood, veneer and bucket wood were all tested as media and, according to Buswell and Pearson (1929), in 1925 work on an aerated sand filter was mentioned for the first time. In the 1930s work was carried out using the 'Tank Filter', otherwise known as the 'Emscher Filter'. These reactors used coarse slag instead of slate layers as the support medium which, according to Bach (1937), led to

an increase in the dissolved oxygen content of the sewage and hence better treatment. By 1939 the Hays Process Filter, also called the Submerged Contact Aeration Process, was developed (Wilford and Conlon, 1957). This introduced the concept of using two stages of biological treatment with rock as a medium. By 1943 submerged contact aeration units with sedimentation tanks were capable of achieving >80% removal of BOD₅ at 74 plants treating domestic sewage in the USA (Griffith, 1943).

According to the British Ministry of Technology (1968), in 1964 a patent was brought out by Albright in Wilson describing a submerged aerated process using coke as the biological support medium. This was one of the first times a process was described that combined the bioreactor with solids removal. In 1967 work was carried out using a support medium of short PVC tubes (Ministry of Technology, 1968). The use of a lower grade plastic medium was suggested to reduce costs but it was already highlighted that operating costs for the process were potentially high.

In the late 1960s and early 1970s work continued on submerged reactors using several types of media. Aerobic treatment through both direct and indirect pure oxygenation was applied to achieve nitrification (Haug and McCarty, 1972; McHarness *et al.*, 1975). An aerobic sand-medium “pulsed adsorption bed” process was described by Johnson and Baumann (1971) and by 1975 packed-bed reactors (PBRs) were being investigated. Aerobic PBRs used air instead of pure oxygen and were run using silica sand, anthracite coal and plastic media (Young *et al.*, 1975; McHarness *et al.*, 1975). Work was also being undertaken in Japan on submerged fixed media processes (Iwai *et al.*, 1977). The main objective of all these processes was to achieve organic matter or ammonia removal; suspended solids removal was also mentioned but without reference to backwashing for filter cleaning.

It was only in the late 1970s and early 1980s that the biological aerated filter (BAF) was developed in Europe to combine solids removal with the bioreactor (Pujol *et al.*, 1994). According to Condren (1990), patents for the process were granted in the USA in February 1978, followed by France (in October 1978 and 1979) and Canada (in

1981). BAFs have also been known as packed bed reactors, submerged aerated filters and immersed aero-filters (Faup *et al.*, 1982; Rusten, 1984); but they all have in common the fact that they are capable of removing organic matter and/or ammonia without requiring further solids separation units downstream. Work continued during the 1980s in the development of BAFs specifically for secondary and tertiary treatment in North America, Europe and Japan (Condren, 1990; Canler and Perret, 1994; Carrio *et al.*, 1995). Between the late 1980s and the present day, work on developing BAFs has continued with a large number of systems now available. At present, many full-scale BAFs are under operation, mainly in Europe, and can be used for denitrification and phosphorus removal as well as for solids and carbonaceous removal (Rogalla *et al.*, 1990).

3.2 Design and operation of BAFs

3.2.1 Liquid Flow and Hydraulics

Flow characteristics in BAFs are influenced by the liquid and air input, as well as by the media (Le Cloirec and Martin, 1984). At the same time, the hydrodynamics of reactors affect the physical and physiological structure of biofilms (Wilderer *et al.*, 1995). Although some researchers have found that BAFs act as completely mixed systems (Costa Reis and Sant'Anna, 1985) due to either channelling and backmixing (Boller *et al.*, 1994), in most cases, flow patterns within BAFs are close to plug flow (Mann *et al.*, 1995; Fdz-Polanco *et al.*, 1996). BAFs usually operate at water velocities of 1 to 10 m³m⁻² h⁻¹ (Stensel *et al.*, 1988; Paffoni *et al.*, 1990; Vedry *et al.*, 1994). However, recent upflow biofilters have been able to achieve 90% NH₄-N removal at a water velocity of 30 m³ m⁻² h⁻¹ when treating storm water and diluted wastewater (Peladan *et al.*, 1997).

Wastewater can be supplied to BAFs at either the top or the bottom of the reactor. Hence, BAFs are operated either in upflow or downflow mode (Figure 3.1). Downflow systems with counter-current air/water flow may offer the advantage of having the air

supplied in contact with the final effluent for a greater period of time. This is particularly important when carbonaceous matter and ammonia removal is required in a single reactor. In this mode of operation, nitrifying microorganisms, which inhabit the lower portion of the filter, receive air rich in oxygen, and therefore will not suffer oxygen limitation (Gonzalez-Martinez and Wilderer, 1991). As the wastewater to be treated enters at the top, blockage of influent nozzles is avoided (Desbos *et al.*, 1990).

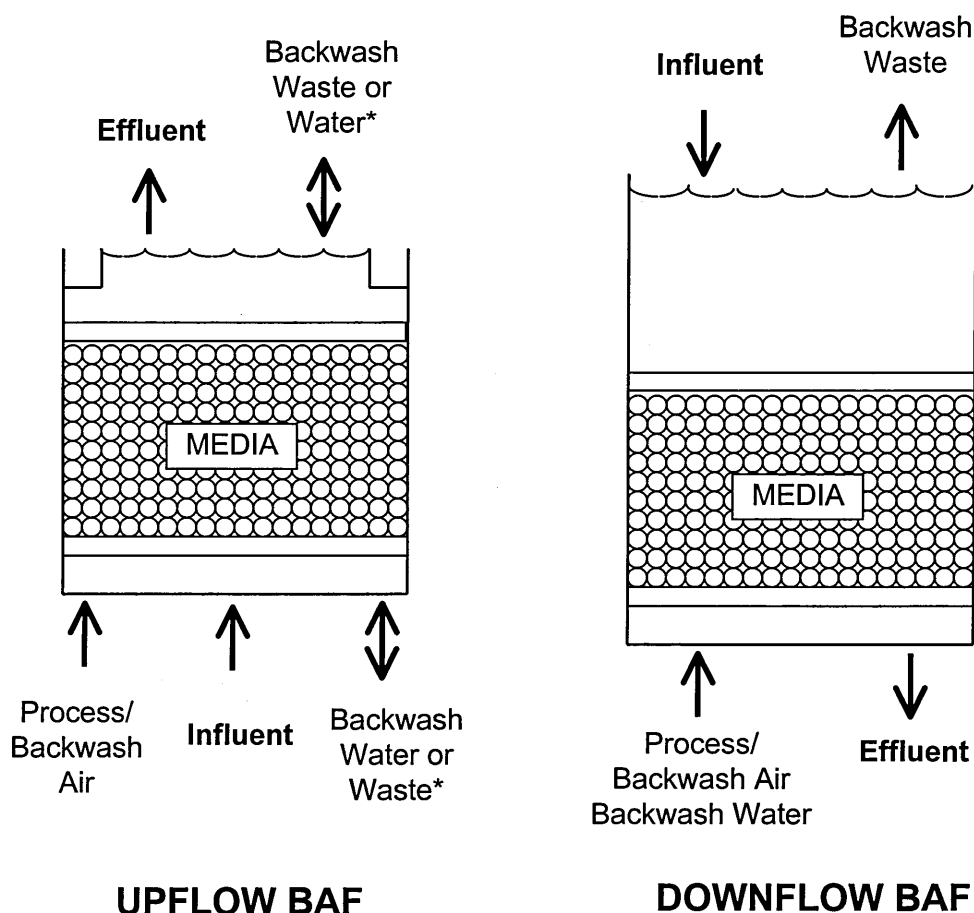


Figure 3.1. Schematic of upflow and downflow biological aerated filters. *Upflow BAFs can be washed by injecting backwash water either co-currently or counter-currently to the operating direction of the flow.

It has been claimed that upflow systems with co-current air/wastewater flow are able to cope with higher influent flow rates than downflow systems, the filter run time is increased by avoiding the formation of air pockets and more effective oxygenation occurs due to bubbles not coalescing and retaining their optimal surface/volume ratio (Iida and Teranishi, 1984). In upflow systems there is a reduced likelihood of odour

problems occurring due to air stripping of volatile compounds since the ambient air is only in contact with the treated effluent (Iida and Teranishi, 1984).

3.2.2 Media types

The packing media in biological reactors significantly influence the hydraulic characteristics and the oxygen-substrate mass transfer. The media should be resistant to attrition, chemically stable, with a high specific surface area and with a low apparent specific weight (Valentis and Lesavre, 1989; Kent *et al.*, 1996). Granular media for BAFs can be either denser than water to give sunken media or less dense than water to produce floating media (Table 2.1). Normally, denser media will require higher flowrates when backwashed and hence higher energy costs. Full-scale BAFs normally operate with a media bed depth of 2-4 m (Stensel *et al.*, 1988; Pujol *et al.*, 1994).

Table 3.1. Types of media used in biological aerated filters.

Media	Media Type	Media Size (mm)	Liquid Flow	Reference
Sunken	expanded clay	~ 1.3	upflow	Sagberg <i>et al.</i> , 1992
		~ 1.9		Smith and Hardy, 1992
		~ 3.5		Vedry <i>et al.</i> , 1994
		6 - 8		Fruhen <i>et al.</i> , 1994
	anthracite	3	downflow	Terayama <i>et al.</i> , 1997
	vitified clay	3 - 4	downflow	Clapp <i>et al.</i> , 1994
	expanded shale	3 - 6	downflow	Dillon and Thomas, 1990
		3 - 6		Bacquet <i>et al.</i> , 1991
		2 - 6		Robinson <i>et al.</i> , 1994
		2.5 - 3.5		Tschui <i>et al.</i> , 1994
	porous stone	20 - 35	upflow	Costa Reis and Sant'Anna, 1985
	slag	~ 40	downflow	Dee <i>et al.</i> , 1994
	granular	~ 2.8	upflow	Jimenez <i>et al.</i> , 1987
		3 - 6	downflow	Pujol <i>et al.</i> , 1994
Floating	polystyrene	2.5 - 3.5	upflow	Pujol <i>et al.</i> , 1994
		3 - 3.5	upflow	Tschui <i>et al.</i> , 1994
		-		Toettrup <i>et al.</i> , 1994
		3.5		Vedry <i>et al.</i> , 1994
		3.5		Andersen <i>et al.</i> , 1995
		2 - 3.5		Visvanathan and Nhien, 1995
	polypropylene	2.3 - 2.7	upflow	Mann and Stephenson, 1997
		5	upflow	Zeghal <i>et al.</i> , 1995
	recycled plastic			

Sunken media may be used both for downflow or upflow configurations, while floating media is used only in upflow systems. Sunken media tend to be made of natural products such as shale or phyllosilicates, e. g. baked clay or pozzolana (Ros and Mejac, 1991; Vedry *et al.*, 1994). Floating media tends to be produced from synthetic products such as polystyrene (Vedry *et al.*, 1994; Andersen *et al.*, 1995). The high specific surface area ($1000\text{--}1500\text{ m}^2\text{ m}^{-3}$) of the granular media allows for a high biomass concentration to be maintained within the system. The removal of solids in the media negates the need for secondary clarifiers, however, backwashing is required to remove excess biomass and captured solids (Dillon and Thomas, 1990; Smith and Hardy, 1992) which increases both the capital and operating costs of the system (Valentis and Lesavre, 1989).

The size of the granular media used affects the efficiency of treatment in terms of physical removal of solids and available surface area for biofilm attachment (Smith and Marsh, 1995). The use of large media ($>6\text{ mm}$) causes a reduction in nutrient and solids removal through high void spaces and a reduced area available for biofilm growth (Stensel *et al.*, 1988). Nevertheless, using large media reduces backwashing requirements and thus brings down operating costs (Costa Reis and Sant' Anna, 1985; Pujol *et al.*, 1994). Using small media ($<3\text{ mm}$) enables better filtration and a large surface area for biofilm growth, though more frequent and vigorous backwashing is required (Robinson *et al.*, 1994). Therefore, it has been suggested that large media ($>6\text{ mm}$) should be used for roughing, intermediate size media ($3\text{--}6\text{ mm}$) for general treatment and fine media ($<3\text{ mm}$) for effluent polishing and/or for tertiary treatment (Quickenden *et al.*, 1992). The shape of the media also affects reactor performance. Irregular grains have been found to improve performance of BAFs compared to spherically shaped media, 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) but it is likely that they will affect backwashing rates.

Roughness of the media also affects the performance of the reactor. Rough surface media provide more sites for biofilm attachment than smooth media. Therefore, smooth media restricts biofilm growth, as the microorganisms are unable to attach properly to the surface. Reactors that use such media could be unstable under variable air and liquid velocities causing biofilm sloughing (Harendranath *et al.*, 1996). Rougher media allow bacteria to firmly attach, producing more stable biofilms (Harendranath *et al.*, 1996) and have also been found to improve solids retention (Ruffer and Rosenwinkel, 1984).

3.2.3 Process Aeration and Oxygen Utilisation

The volume of air required for treatment is a function of the pollution load, the endogenous respiration rate of the biomass and the oxygen transfer efficiency (Robinson *et al.*, 1994). In BAFs adequate aeration is essential to reach the desired level of treatment (Vedry *et al.*, 1994). Condren (1990) reports that in certain sunken media BAFs 70 m³ of air are required to remove 1 kg of soluble BOD₅, 20 m³ of air to remove 1 kg of insoluble BOD₅ and 200 m³ of air to remove 1 kg of NH₄-N. When the aeration rate is too low, insufficient substrate removal will occur and anoxic zones may be created within the reactor. If the aeration rate is too high scouring of the biofilm and a reduction of the solids removal efficiency may occur (Pearce, 1996).

According to Stensel *et al.* (1988), oxygen utilisation efficiencies obtained in BAFs (5-9.2%) are higher than those for activated sludge in tanks of equal depth, probably due to an increase in the detention time of air bubbles by entrapment and tortuous paths caused by the media. Moreover, because bubbles can be in direct contact with the biofilm in BAFs, it has been suggested that 50-65% of the oxygen supplied is transferred directly (interfacially) to the microorganisms (Lee and Stensel, 1986). Any increase in the substrate loading results in an increase in oxygen utilisation rate, whilst bulk liquid oxygen concentrations show only slight changes. Thus, BAFs are not limited by bulk liquid oxygen concentrations and are consequently able to remove high substrate loadings (Reiber and Stensel, 1985). Although some investigators have found

that oxygen transfer efficiencies (OTEs) are largely unaffected by the type of aerator used (Canziani, 1988; Mann *et al.*, 1995; Harris *et al.*, 1996), coarse-bubble diffusers could be more efficient than fine-bubble diffusers for BAFs due to their superior aeration performance, reliability and comparatively lower energy requirements (Hodkinson *et al.*, 1998). Off-gas analysis has been proposed as a reliable means of monitoring and controlling aeration rates in activated sludge systems (Redmon *et al.*, 1983) and its potential application for BAFs has also been proposed (Harris *et al.*, 1996).

3.2.4 Backwashing / Air Scour

Backwashing is carried out to remove captured solids and excess biomass in BAFs (Faup *et al.*, 1982; Park and Ganczarczyk, 1994). Backwashing consists normally of high rate aeration (air scour) that dislodges filtered solids and excess biomass combined with high rate liquid flows to remove the solids from the reactor vessel through flushing (Canler and Perret, 1994). Backwashing must maximise the length of the purification cycles whilst minimising energy consumption and the quantities of air and wash water. The wash cycle must not damage the support media and the remaining biomass required for the rapid restart of the bioreactor. The time between washes usually has to be maximised in order to maintain a stable population of active biomass within the reactor and avoid media loss (Carrand *et al.*, 1990). Approximately 30% of the biomass can be removed by backwashing without reducing reactor performance to any great degree (Bacquet *et al.*, 1991). Underwashing can lead to premature headloss, filters queuing to wash at intermittent times and possible solids breakthrough, resulting in more frequent backwashing (Robinson *et al.*, 1994). On the other hand, overwashing can cause a loss of effluent clarity due to the reduced biological filtration of colloidal particles and the eventual deterioration of the treatment efficiency as the biomass concentration is reduced (Robinson *et al.*, 1994). Thus excessive biomass removal results in longer recovery times to allow sufficient biomass growth to take place (Amar *et al.*, 1986). Nevertheless, if the backwashing process is undertaken correctly it will have little influence on the effluent quality (Visvanathan and Nhien, 1995).

The frequency of backwashing will depend on the size, shape, density and voidage of the media and the characteristics of the wastewater to be treated since the biofilm growth will be influenced by the organic loading rates and biomass growth rate (Robinson *et al.*, 1994). Reactors with stable, rapidly growing biofilms involved in carbonaceous removal (secondary treatment) require more frequent and intense backwashing than reactors with less established and slower growing biofilms used for nitrification (tertiary treatment). Therefore, when used for secondary treatment, backwashing is carried out every 12 to 48 h while in tertiary treatment backwashing is normally carried out on a weekly basis (Dillon and Thomas, 1990; Bacquet *et al.*, 1991, Smith and Hardy, 1992).

Two methods are used to control the frequency of backwashing. The simplest method is to backwash on a regular basis (e. g. every 24h for carbonaceous matter removal). This method does not require additional monitoring equipment and does not allow the time period between washes to be optimised (Wheale and Cooper-Smith, 1995). The second method consists of carrying out backwashing when a pre-set head pressure loss is reached, which allows for variable strength wastewaters to be treated through optimisation of the filter run time (Paffoni *et al.*, 1990).

The type of backwashing carried out is determined largely by the process configuration. Reactors running in downflow mode are backwashed in an upflow direction, in opposition to the direction of the bed compression (Jimenez *et al.*, 1987; Rogalla and Sibony, 1992). Reactors using floating media backwash either in downflow (Andersen *et al.*, 1995; Visvanathan and Nhien, 1995) or in upflow direction (Mendoza-Espinosa *et al.*, 1997b).

Air and liquid flowrates used for backwashing normally vary between 0.43-0.52 m³ air per m³ media per min and 0.33-0.35 m³ water per m³ media per min, respectively (Condren, 1990). As mentioned earlier, the backwashing rate will depend on the amount of captured suspended solids and biofilm formation, which will depend on the

characteristics of the media used (i. e. roughness and density). Backwashing can account for up to 15-20% of the total daily energy consumption in full-scale BAFs (Rogalla *et al.*, 1990). Generally, the amount of liquid required for backwashing full scale BAF plants is 2% of the total treated water if used for tertiary nitrification and 12-35% when used for secondary treatment (Dillon and Thomas, 1990; Canler and Perret, 1994; Wheale and Cooper-Smith, 1995). In comparison, sand and anthracite filters require 0.9-8.3% of the total treated water (Ruffer and Rosenwinkel, 1984).

3.3 Biomass characteristics

Due to the configuration of BAFs, access to the biofilm is very restricted and its study remains limited. Biofilm thickness in BAFs is strongly related to the scouring effect during media backwashing (Boller *et al.*, 1994) as well as by the substrate loading rate (Ohashi *et al.*, 1995). Sludge from BAFs has been shown to have a greater number of flocs than those obtained from activated sludge, trickling filters and digested sludge particles (Park and Ganczarczyk, 1994). It has been suggested that biofilm thickness for simultaneous carbonaceous and nitrogenous removal should not exceed 120 μm and if so, the medium should be backwashed to avoid clogging (Tschui *et al.*, 1994; Ohashi *et al.*, 1995). In accordance with classic biofilm theory (Wanner and Gujer, 1986), excess biomass in BAFs does not contribute towards treatment performances (Bacquet *et al.*, 1991), and thus the thickness of the active biofilm layer is limited.

Biological aerated filters treating settled domestic wastewaters have been found to contain a variety of bacterial genera including *Acinetobacter*, *Enterobacteriaceae*, *Aeromonas*, *Flavobacterium*, *Alcaligenes*, *Pseudomonas* and *Moraxella* (Zinebi *et al.*, 1994). Zoning of particular species has not been found, though the concentration of each microbial species varies at different reactor heights (Giuliano and Joret, 1989; Villaverde *et al.*, 1997a). Predominant bacteria types also vary during acclimatisation to chemicals that are difficult to biodegrade (Fujie *et al.*, 1994). It was found that predominant bacteria species changed from *Proteobacteria* to *Mycobacterium* and

finally *Methylobacterium*. As well as bacteria, BAF biofilms contain higher organisms such as protozoa and worms that influence the bacterial population dynamics due to predation and consequently reactor performance (De Beer and Muyzer, 1995). Microscopic and enzymatic tests have concluded that backwashing has an important effect on the distribution of biomass along a nitrifying reactor (Gschlosl *et al.*, 1997). Microorganisms indicating a high load, *Mayorella sp.* and *Paramecium sp.* and partially low oxygen concentrations, *Vorticella microstoma*, were predominant before backwashing. The most common microorganisms in a nitrifying reactor were dispersed ciliated protozoes such as *Aspidia sp.* (Gschlosl *et al.*, 1997). However, from an engineering point of view detection of functional bacteria group is more important than a detailed description of which species are present (de Beer and Muyzer, 1995).

The production of sludge in BAFs used for secondary treatment under steady-state conditions has been reported to be 0.63-1.06 kg SS per kg BOD (Amar *et al.*, 1986; Dillon and Thomas, 1990; Rogalla *et al.*, 1990; Rogalla and Sibony, 1992) and 0.41 kg SS per kg COD eliminated (Canler and Perret, 1994). Sludge produced during combined secondary and tertiary treatment was found to be 0.13-0.32 kg SS per kg BOD removed (Iida and Teranishi, 1984). Sludge produced from BAFs has shown good settleability (Robinson *et al.*, 1994).

3.4 Substrate removal

3.4.1 Solids Filtration and Carbonaceous Removal

The most widespread use of BAFs has been for the removal of carbonaceous BOD combined with solids filtration (Table 3.2).

Table 3.2. Carbonaceous matter removal using biological aerated filters for wastewater treatment.

Reactor configuration	Reactor volume (m ³)	Influent concentration (mg l ⁻¹)	Organic loading rate (kg m ⁻³ d ⁻¹)	Removal (%)	HRT (h)	Reference
Upflow	0.0085	3000-3500	3.3 - 15.4	33-82	4.5-23	Costa Reis and Sant'Anna, 1985
Downflow	0.3	424 (average)	<9.2 COD	>90	0.4-0.76	Dillon and Thomas, 1990
Downflow	0.14	<200	10.5 COD	~55	0.5	Rogalla <i>et al.</i> , 1990a
Downflow	Full scale	354 (average)	7.5 COD	75	-	Rogalla <i>et al.</i> , 1990b
Downflow	6 cells, 54 m ³ each					
Downflow	0.2 - 0.3	324 (average)	<15	86	0.4-0.6	Rogalla and Bourbigot, 1990
Downflow	0.02	350	8 - 10	90	-	Bacquet <i>et al.</i> , 1991
Downflow	22	13.6 sCOD _{Mn}		12	-	Sakuma <i>et al.</i> , 1993
	0.1	13.6 sCOD _{Mn}		20	-	
	0.7	9 sCOD _{Cr}	2.3 COD _{Cr}	30	-	
Upflow & Downflow	Full scale plants	35 - 607	0.5 - 6.3	55 - 85	-	Pujol <i>et al.</i> , 1992
Downflow	31.5 - 90.3					
Downflow	Full scale	131	1.5 COD	93	1.3	Wheale and Cooper-Smith, 1995
	4 cells 143 m ³ each					
Upflow	Full scale, two units	25 - 43	~ 2.4 (per unit)	48 - 70	21	Peng <i>et al.</i> , 1995
	151.2 m ³ each					
Upflow	Full scale, 8 cells	109 - 250 (BOD)	4	> 93	-	Brewer, 1996
	219 m ³ each					

Solids removal in BAFs is achieved mainly through filtration (Stensel *et al.*, 1988; Ryhiner *et al.*, 1994) and is dependent on 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 Harremoës, 1989).

Removal of carbonaceous pollutants in BAFs is carried out by three mechanisms: solids filtration, absorption and oxidation (Stensel *et al.*, 1988). The advantages of BAF reactors over other secondary processes are their capability to reach high organic removal rates per volume of reactor and the production of high quality effluent. Organic removal rates of up to $4.1 \text{ kg BOD m}^{-3} \text{ d}^{-1}$ are normally reported in partially nitrifying BAFs (Table 3.2). In comparison, trickling filters, oxidation ditches and activated sludge plants normally achieve removal rates of $0.06 \text{ kg BOD m}^{-3} \text{ d}^{-1}$, $0.35 \text{ kg BOD m}^{-3} \text{ d}^{-1}$ and $0.42 \text{ kg BOD m}^{-3} \text{ d}^{-1}$, respectively (Smith and Hardy, 1992; Pujol *et al.*, 1994). However, as BAFs are operated as plug-flow reactors they can be affected by shock loads and toxic or inhibitory substances (Chen and Cheng, 1994).

BAFs have been shown to be more stable at variable and low temperatures than suspended growth processes (Koutsakos *et al.*, 1992) and other biofilm systems such as trickling filters and rotating biological contactors (Tschui *et al.*, 1994; Stensel *et al.*, 1988; Sakuma *et al.*, 1993). At $<5^{\circ}\text{C}$ there was a decrease in suspended solids removal, BOD removal and nitrification with evidence that prolonged very low temperatures caused loss of biofilm (Koutsakos *et al.*, 1992). A 75% decrease in the yield coefficient was observed between 5°C and 35°C (Fujie *et al.*, 1994). Studies of the impact of high temperatures on the performance of BAFs for carbonaceous removal have found that maximum growth rate and substrate removal occur at approximately 38°C , within the mesophilic bacteria range (Visvanathan and Nhien, 1995). Greater temperatures reduce the growth yield of the microorganisms responsible for organic removal. Above 41°C , thermophilic rather than mesophilic microorganisms predominate and thus alter biological growth and substrate removal rates (Visvanathan and Nhien, 1995).

3.4.2 Nitrification

BAFs can be used for simultaneous SS, COD and ammonia removal or for SS and ammonia removal (Table 3.3).

Secondary treatment BAFs used primarily to remove carbonaceous material can achieve partial nitrification at ammonia loadings of up to $1 \text{ kg NH}_3\text{-N m}^{-3} \text{ d}^{-1}$ (Table 3.3). In BAFs used for tertiary nitrification a small fraction (2-3%) of the ammonia is removed via cell growth and the rest is oxidised by autotrophic bacteria (Akunna *et al.*, 1994; Cecen and Gonenc, 1995). Once a stable population of nitrifiers is reached, effluent residuals are directly dependent on influent concentrations and hydraulic residence time. Nitrification at rates of up to $1.27 \text{ kg NH}_3\text{-N m}^{-3} \text{ d}^{-1}$ has been reached.

Being a biological fixed-film process, biological aerated filters share some types of nitrification inhibition with trickling filters and RBCs: through competition with non-nitrifying biological species, predation by higher organisms (Lee and Welander, 1994) and through the influence of solids, substrate concentration (Rogalla *et al.*, 1990) and toxic substances (Oslislo and Lewandowski, 1985; Beg and Hassan, 1987). Biological competition is predominantly dependent on the nutrient concentration of the wastewater to be treated (Boller *et al.*, 1994). Though autotrophic nitrifying bacteria require some carbonaceous material, at high concentrations the rapid rate of growth of heterotrophic bacteria results in limited autotrophic growth and subsequent nitrification reduction (Akunna *et al.*, 1994). It was found that when the COD loading was increased between 7 and 8 times in tertiary nitrifying BAFs, the efficiency of the reactors decreased between 30 and 50% (Boller *et al.*, 1997). High solids concentrations also inhibit nitrification through adsorption onto the biofilm surface which 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 rates decrease as organic loadings increase, even when the C: N ratio is constant (Al-Haddad *et al.*, 1991). High levels of

Table 3.3. Ammonia removal using biological aerated filters for wastewater treatment.

Reactor configuration	Reactor volume (m ³)	Influent concentration (mg l ⁻¹)	Organic loading rate (kg m ⁻³ d ⁻¹)	Removal (%)	HRT (h)	Reference
Upflow	0.55	22.7 - 37.5	0.11 mg NH ₄ -N mg ⁻¹ VM d ⁻¹ *	62 - 84	6 - 12.9	Faup <i>et al.</i> , 1982
Upflow (Intermittent aeration)	4.18	17.8	0.03 - 0.05	88.4	7 - 8	Iida and Teranishi, 1984
Downflow	1	13 - 20.9	0.39 - 0.84 (NH ₄ -N)	90 - 99	0.5 - 0.9	Rogalla and Payraudeau, 1988
Upflow	Full Scale		< 0.46			Carrand <i>et al.</i> , 1990
Downflow	0.3	40.7	< 0.58	65 - 100	0.32 - 0.83	Dillon and Thomas, 1990
Downflow	0.14	< 20	< 0.6	-	1	Rogalla <i>et al.</i> , 1990a
Downflow	0.2 - 0.3	40 (TKN)	< 1 (NH ₄ -N)	> 95	1	Rogalla and Bourbigot, 1990
Downflow	22	11	0.9* (NH ₄ -N)	57	-	Sakuma <i>et al.</i> , 1993
	0.1	23.4	1.2* (NH ₄ -N)	78	-	
Upflow	Full scale, 333	~ 22	~ 1.87 (NH ₄ -N)	89	-	Vedry <i>et al.</i> , 1994
Upflow	0.81 - 1.16		< 3 (NH ₄ -N)	80	-	Peladan <i>et al.</i> , 1996
Upflow	Full scale, 18 cells 292 m ³ each	13 - 28	0.15	92 - 96	-	Brewer, 1996
Upflow	0.79	~ 27	1.12 (NH ₄ -N)	~ 82	-	Gennens <i>et al.</i> , 1997

* removal rates

ammonia promote the growth of ammonia oxidisers but not of nitrite oxidisers resulting in a build up of nitrite (Hao and Chen, 1994; Villaverde *et al.*, 1997a).

It has been suggested that a bulk DO concentration of 4-5 mg l⁻¹ should be maintained within a BAF to avoid inhibition of nitrification (Chen and Cheng, 1994). The transition from ammonia limitation to oxygen limitation occurs when the ratio of bulk oxygen concentration to bulk ammonia concentration is in the range of 2-4.5 (Cecen and Gonenc, 1995). As this is not often achieved in full scale BAFs, the nitrification process is often oxygen limited (Cecen and Gonenc, 1995), especially at the reactor inlet, as most of the organic substrate is consumed in that part of the reactor (Beg *et al.*, 1995). Oxygen limitation can lead to nitrite accumulation and subsequent inhibition of *Nitrobacter* activity (Fdz-Polanco *et al.*, 1994). Nitrification rate is also affected by pH. It has been found that nitrification rate increases by 13% per pH unit between pH values of 5.0 to 8.5, which corresponds to the ammonium oxidisers maximum activity (Villaverde *et al.*, 1997b). Although temperature influences the nitrification rate in submerged bioreactors (Iida and Teranishi, 1984; Paffoni *et al.*, 1990), the same degree of nitrification can be maintained if the hydraulic retention time is increased as temperature decreases (McHarness *et al.*, 1992). Maximum nitrification activity occurs at 28-29°C (Fdz-Polanco *et al.*, 1994).

3.4.3 Nutrient Removal

BAFs in series have been used for pre- or post-denitrification. When used for post-denitrification an external carbon source is required after the main carbonaceous/nitrifying BAF through the addition of methanol (Zeghal *et al.*, 1995), isopropanol (Hwang *et al.*, 1994), molasses (Cecen and Gonenc, 1995), ethanol or hydrolysed sewage sludge (Aesoy *et al.*, 1998). The sewage is then fed to an anoxic denitrifying reactor. In pre-denitrification, in order to solve the problem of carbon availability, a mixture of recycled nitrified effluent and settled sewage can be fed to an anoxic pre-denitrification reactor situated prior to the main carbonaceous/nitrifying BAF (Dee *et al.*, 1994). In pre- and post-denitrification, BAFs have been used for

denitrification simply by being operated without aeration (Jepsen and Jansen, 1993; Janning *et al.*, 1995) running as anoxic sand filters with biomass still present.

Another alternative is to run an upflow BAF with the aerators placed part way up the floating media in order to allow for both an aerated and a non-aerated zone (Rogalla and Bourbigot, 1990) (Figure 3.2). The lower anoxic zone removes the soluble organic matter and convert organic nitrogen into ammonia while the upper aerated zone achieves nitrification and removes most of the remaining carbonaceous matter (Chui *et al.*, 1996). The denitrification efficiency of the reactor will depend on the loading and the recirculation ratio to the anoxic zone (Rogalla and Bourbigot, 1990). High denitrification rates of up to $5 \text{ kg NO}_3\text{-N m}^{-3} \text{ d}^{-1}$ are reached when a carbon source is present in excess (Jepsen and Jansen, 1993; Toettrup *et al.*, 1994) at an influent COD/NO_x-N ratio of 5 (Cecen and Gonenc, 1995; Chui *et al.*, 1996) or >6 (Rogalla and Bourbigot, 1990).

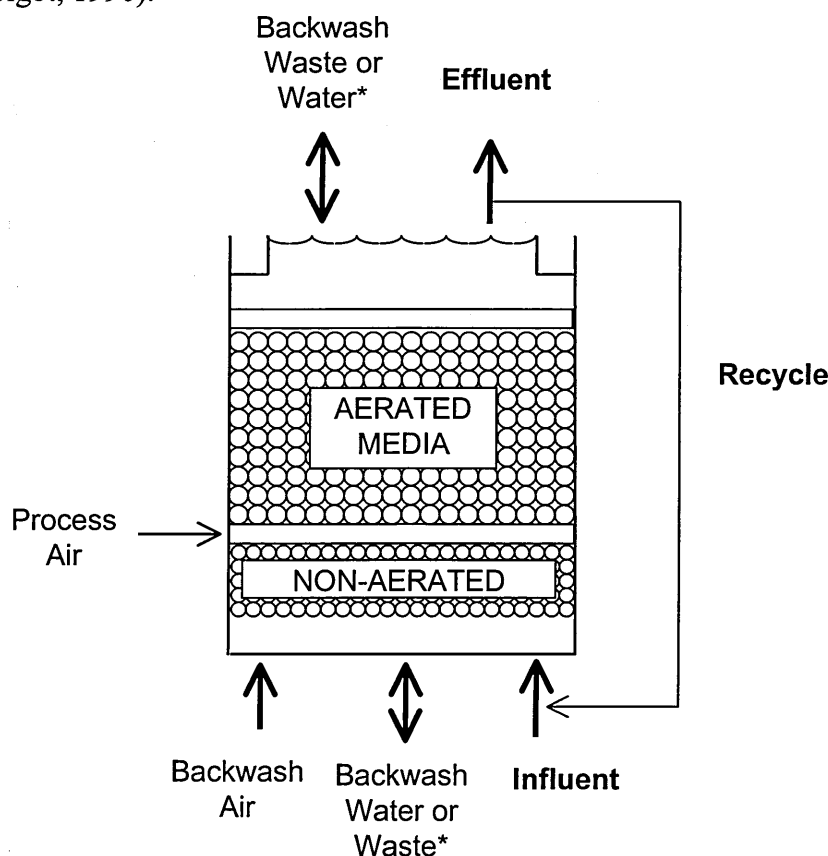


Figure 3.2. Upflow biological aerated filter for combined secondary and tertiary treatment and denitrification.

More recently, BAFs have been used for phosphorus removal by using chemical precipitation (Rogalla *et al.*, 1990; Goncalves *et al.*, 1994; Clark *et al.*, 1997). Biological aerated filters can achieve up to 35% removal of total phosphorus through physical filtration without any chemical dosing (Clark *et al.*, 1997). When a coagulant is used, phosphorus removal is mainly achieved by flocculation and sedimentation but biological filtration removed up to 50% of the residual phosphorus after settling (Rogalla *et al.*, 1990). When chemical dosing is applied, the average phosphorus removal efficiency at different dosing ratios was 85% and the general performance of the BAF with respect to SS, BOD, COD and total Kjeldahl nitrogen (TKN) removal was not affected (Clark *et al.*, 1997).

3.5 Industrial applications

Plants treating domestic waste have been able to treat wastewater containing petroleum products (hydrocarbons) with concentrations of up to 20 mg l⁻¹, at efficiencies of up to 95% (Drushlyak and Kirichenko, 1990). BAFs have also been used for treating high strength industrial wastewater (Kleiber *et al.*, 1994) but pretreatment to minimise concentrations of suspended solids, oil and grease entering the reactor and in some cases nutrient addition must be considered (Stensel and Reiber, 1983). Other applications include the treatment of wastewater from the paper mill industry (Isohata and Kawabata, 1984; Rovel *et al.*, 1994; Kantardjieff and Jones, 1997), fish meat processing plant and corn starch manufacturing plant (Isohata and Kawabata, 1984), aluminium rolling mill (Adachi and Fuchu, 1991), the distillery industry (Costa Reis and Sant'Anna, 1985) and industrial hazardous waste landfill leachate (Smith, 1995).

3.6 Modelling

Mechanistic modelling of BAFs is complex due to the many variables involved in such a biofilm process. Easy access to the interior is required without disturbing the remaining biofilm and reactor operation but in practice this is difficult to achieve

(Gjaltema and Griebbe, 1995). In contrast, empirical models might be less accurate than mechanistic models but are based on parameters that can be easily measured. Empirical models directly relate input variables such as organic loading or substrate concentration to output variables such as substrate removal rate; general assumptions like the uniformity of biomass distribution and ideal plug flow have to be made (Meunier and Williamson, 1981; Hamoda, 1989).

Costa-Reis and Sant'Anna (1985) proposed a mathematical model for BAF scale-up based on kinetics, mixing and biofilm growth. Recently, attempts to develop semi-empirical models for submerged filters have been undertaken. Le Tallec *et al.* (1997a) proposed a simple model based on the IAWQ Activated Sludge Model for a nitrifying/denitrifying filter (Henze *et al.*, 1987). Mann and Stephenson (1997) developed an empirical model for BAFs that related the influent soluble chemical oxygen demand (sCOD) to the effluent sCOD and reactor height. Mendoza-Espinosa and Stephenson (1998) used this model to compare the effect of media and the hydraulic configuration of BAFs on soluble carbonaceous matter removal. Jacob *et al.* (1997) developed a generalised model for carbon removal, denitrification and nitrification. As suspended solids removal was also included, the model was able to predict overall removal efficiencies and filter clogging. Beg *et al.* (1997) performed numerical simulations to analyse the performance of an upflow BAF under simultaneous carbon oxidation and nitrification reactions. They concluded that the presence of organic matter inhibited nitrification when simultaneous carbon oxidation and nitrification were undertaken and that sufficient oxygen should be supplied near the reactor inlet where most of the organic oxidation is taking place.

3.7 Discussion

According to Condren (1990) air requirements, electrical power consumption and manpower needs for biological aerated filters could be equivalent or slightly less than those associated with other more conventional wastewater treatment processes. However, it appears that other investigations have concluded that the energy consumption associated with aeration and backwashing of BAFs has limited their application. The mechanical and electrical costs are considerably higher than the civil engineering costs (Wheale and Cooper-Smith, 1995). Power consumption of a full scale BAF plant averaged £0.01 m⁻³ of water treated, equivalent to £50 per tonne of effective oxygen demand (EOD) removed, where EOD is the average BOD load plus 4.5 times the average NH₃-N load (Robinson *et al.*, 1994). The specific average daily consumption measured in 12 full-scale BAF plants was 1.16 kWh per kg COD eliminated per day (Canler and Perret, 1994). The minimum power consumption of a pilot-scale BAF was found to be 0.75 kWh per kg BOD + NH₃-N removed, which compares favourably with a nitrifying activated sludge plant (Lilly *et al.*, 1991). Average power consumption in a package BAF plant was 0.35 kWh per m⁻³ of sewage treated based on an average daily flow of 420 m³ d⁻¹ which, according to the authors, was very high (Smith and Hardy, 1992). Dillon and Thomas (1990) stated that the cost of sewage treatment using a BAF process was comparable to that of the activated sludge process for sewage treatment works >50,000 population equivalent (pe); for plants <10,000 pe the BAF was the most expensive compared to activated sludge and trickling filters.

Biological aerated filters can achieve high removal rates of carbonaceous organic matter and ammonia (Tables 3.2 and 3.3) due to their high biomass concentration (Smith and Hardy, 1992) which also allows for short hydraulic retention times of <1 h. Although loadings of up to 8 kg BOD m⁻³ d⁻¹ are achievable, after this point there is a drop in removal efficiency by 10-20% compared to the maximum removal achieved (Rogalla *et al.*, 1994). Recent high velocity BAFs have been reported to be capable of removing up to 18 kg COD m⁻³ d⁻¹ (Peladan *et al.*, 1997).

It has been suggested that nitrification is enhanced at high airflow velocities, when the laminar boundary layer of the biofilm decreases and the mass flow increases (Tschui *et al.*, 1994). The turbulence may also reduce hydraulic channelling and advective transport in the granular media. This would result in better utilisation of the total attached biomass and thus increase nitrification rates. Despite the reduction in hydraulic retention times, the increase in the water velocity might have a positive effect on the overall oxygen mass transfer rate and therefore, on the maximum capacity of the filter (Peladan *et al.*, 1997).

The mechanisms of solids removal in BAFs has been an area that has not received too much attention. According to Stensel *et al.* (1988) only a small fraction of the suspended solids (SS) removed in BAFs are degraded since most are washed out of the system after only a relatively short time. Experiments with solids tracers like fluorescent microbeads (Reichert and Wanner, 1997) would be useful to determine the residence time of solids in BAFs.

In order to optimise removal rates in BAFs future research should focus on enhancing oxygen utilisation. With reduced energy consumption BAFs would certainly become a viable alternative for nitrification, especially when the additional benefit of SS removal is taken into account (Boller *et al.*, 1994). Aeration could be optimised by the use of direct off-gas measurements and direct control of aeration rate by the measurement of oxygen utilised (Harris *et al.*, 1996). Another method for aeration optimisation is to use on-line sensors based on UV absorption (Rogalla *et al.*, 1990). UV absorption can be directly correlated to COD loading and automatic adjustment of the air flowrate to the BAF undertaken. New BAF plants have been built with on-line measurement of ammonia and nitrogen to allow for oxygen adjustment and carbon dosing, respectively, as well as the optimum recirculation ratio (Borregaard, 1997). By these means, the effluent quality as well as the energy consumption are optimised.

Although backwashing has long been highlighted as an important parameter affecting BAFs performance (Stensel *et al.*, 1988) it appears that little work has been undertaken for its optimisation. Recently investigations on recirculating media BAFs, similar to continuous filtration systems, have been undertaken (Kramer and Wouters, 1993; Stephenson, 1996; Daamen *et al.*, 1996). The process has the same benefits as the original BAF design and can remove solids, carbonaceous matter and ammonia in a single reactor. By keeping constant media recirculation backwashing is avoided and thus footprint is even smaller. However, these processes have yet to be optimised.

It has been suggested that in order to find the most suitable media for BAFs, physical parameters alone, such as specific area and attrition resistance, are not sufficient. Biomass studies with scanning electron micrographs (SEM), as well as backwashing and performance studies should also be undertaken (Kent *et al.*, 1996). The direct comparison of different process configurations like sunken and floating media, upflow and downflow reactors is necessary to assess the full capability of BAFs. Tracer studies undertaken to compare floating and sunken media suggested that floating media would be best for high solids wastewater and sunken media for low solids wastewater (Mann *et al.*, 1995). Nitrification in upflow and downflow BAFs has been compared. Grasmick *et al.* (1984) found that the downflow reactor performed better than the upflow unit at increasing ammonia loading rates. Moreover, the downflow and upflow BAFs were equally affected by increases in the liquid and air velocities though the downflow reactor achieved better ammonia removal. Newbigging *et al.* (1995) found that an upflow BAF achieved better nitrification than a downflow reactor and that different media bed heights were required in each reactor to achieve the same performance. Hence, according to Newbigging *et al.* (1995), the preference of configuration should be based on the capital and running cost of each process.

According to Hamoda and Al-Ghusain (1998) properly designed fixed-film processes are generally reliable due to the large amount of biomass present which allows them to withstand hydraulic and organic shock loads effectively. Yet, investigations of the impact of peak flows and the response to shock loading in a pilot-scale three-stage

BAF plant have concluded that short-term fluctuations in the wastewater flowrate directly affected the removal of organic carbon (Chudoba and Pujol, 1998). Moreover, the effect of shock loads and toxic or inhibitory substances in BAFs operated as plug-flow reactors are slowly dispersed and high peak loadings may result in organic breakthrough due to limited treatment efficiency (Ruffer and Rosenwinkel, 1984; Chen and Cheng, 1994). Therefore, there is not a general consent about the capability of BAFs for withstanding organic and hydraulic shock loads. Further work is necessary to clarify the subject.

Only recently has there been attempts to try to incorporate the removal of suspended solids into the modelling of biological aerated filters (Jacob *et al.*, 1997). The use of kinetics parameters from the International Association on Water Quality (IAWQ) Activated Sludge Model 1 (Henze *et al.*, 1987) caused errors in the validation of the biological part of the model developed by Jacob *et al.* (1997). Hence, although it is possible to include several approaches when modelling BAFs, the complexity involved in the validation and application of such models make them unlikely to be used. In order to avoid this, well-designed simple empirical models would have the potential for being used as useful tools for BAF design.

Some new wastewater treatment plants have used BAFs in preference to other secondary and tertiary processes (Carrio *et al.*, 1995). Due to their reliability to meet effluent criteria, BAFs are being considered as a viable process to conventional activated sludge treatment for some municipal wastewater treatment works in Canada (Parker *et al.*, 1998). Recently, BAFs have been used downstream of upflow anaerobic sludge bed reactors (UASB), resulting in a compact, efficient and low energy consumption system that could be a solution for treating domestic wastewater in Brazilian metropolitan areas (Goncalves *et al.*, 1998). The adaptability of BAFs and their small footprint allows them to be used in upgrading established works (Rogalla *et al.*, 1994), especially those in built up areas where there is limited space for expansion (Pujol *et al.*, 1994). Moreover, the compact design and the ability to conceal BAF plants underground makes them ideal for environmentally sensitive sites such as

coastal areas (Gilles, 1990; Sampa and Tanaka, 1995). The high effluent qualities that can be achieved with BAFs has allowed for effective reclamation and reuse of the domestic wastewater in buildings (Adachi and Fuchu, 1991). Due to the low volumes of wastewater treated and hence the low operating costs involved, BAFs have also been used as part of small package plants and have proved to be reliable and easy and economic to construct, maintain and operate (Watanabe *et al.*, 1993).

Design of biological aerated filters has thus reached a point where the basic concept is well understood but further optimisation is required for the process to be competitive with traditional solutions in more situations.

3.8 Conclusions

- Granular media submerged aerated biofilters can combine ammonia, carbonaceous matter and solids removal in a single unit process. Biological nutrient removal (N and P) can also be accomplished.
- Removal rates based on reactor volume for carbonaceous BOD, ammonia and nitrates of up to $4.1 \text{ kg BOD m}^{-3} \text{ d}^{-1}$, $1.27 \text{ kg NH}_3\text{-N m}^{-3} \text{ d}^{-1}$ and $5 \text{ kg NO}_3\text{-N m}^{-3} \text{ d}^{-1}$ respectively, are normally reported.
- The small footprint and adaptability of BAFs allows them to be used in upgrading established works, especially those in built up areas where space is at a premium.
- Areas in which further study is required include: mechanisms of solids removal, aeration optimisation by monitoring oxygen utilisation through off-gas analyses, optimisation of backwashing and media characteristics, monitoring resistance to shock loadings, comparison of upflow and downflow systems and evaluation of models for scaling-up purposes.

Chapter 4

DETERMINATION OF FLOW PATTERN AND ACTIVE
VOLUME IN BIOLOGICAL AERATED FILTERS UNDER
UPFLOW AND DOWNFLOW CONDITIONS

DETERMINATION OF FLOW PATTERN AND ACTIVE VOLUME IN BIOLOGICAL AERATED FILTERS UNDER UPFLOW AND DOWNFLOW CONDITIONS.

ABSTRACT

Although there is currently a great number of papers concerned with the performance of biological aerated filters, very few make a direct comparison between upflow/downflow conditions. The aim of this study was to compare the flow pattern and the active volume of two pilot-scale biological aerated filters using identical media under downflow and upflow conditions. Within the ranges studied, the hydrodynamic mixing was found to not be greatly affected by high air velocities and the retention times did not decrease as expected, probably due to the high density of the media used (2500 kg m^{-3}). The existence of a plenum chamber under the media bed caused severe reduction in active volume in the downflow reactor due to the entrapment of air bubbles. Elimination of the plenum chamber by means of the addition of media solved the problem. Increases in the nominal liquid flowrate from 0.4 l min^{-1} (hydraulic velocity of 1.8 m h^{-1}) to 0.8 l min^{-1} (3.6 m h^{-1}) affected the hydrodynamic conditions of the reactors in a greater way than changes in the air flowrate. Both columns presented similar hydraulic conditions, which is important for scaling-up purposes.

4.1 Introduction

Biological aerated filters (BAFs) are three-phase reactors for wastewater treatment using granular media for biomass support and solids removal by filtration. Developed over the last 20 years (Faup *et al.*, 1982), they are capable of achieving carbonaceous matter removal, solids filtration, nitrification and/or denitrification in a single reactor. Backwashing is required to remove the excess biofilm that grows on the media and the solids entrapped. The effluent normally exits to a storage tank where some of the effluent is re-used for backwashing the filter. There is a range of designs with variations in flow directions and the type of media used. In downflow systems the influent is introduced to the top of the filter medium and aeration is applied in counter-current mode. In upflow systems the influent is introduced at a greater pressure than the filter depth at the base of the filter near the point where air is injected. The air/sewage mixture flows co-currently through the medium and the effluent exits at the top of the filter. Although there are a number of papers about biological filters performance (Stensel *et al.*, 1988; Rogalla and Sibony, 1992), very few make a direct comparison between upflow/downflow conditions.

Hydrodynamics play an important role in bioreactor performance. Flow characteristics in BAFs are influenced by the liquid and air inputs, as well as by the media (Le Cloirec and Martin, 1984). The microbial population is greatly affected by the hydraulic behaviour of the system (Iida and Teranishi, 1984) as well as by other factors such as temperature and pH (Roustan *et al.*, 1990).

According to Costa Reis and Sant'Anna (1985), BAFs act as completely mixed systems, due to either channelling and/or backmixing (Boller *et al.*, 1994). However, in most cases flow patterns within BAFs are close to plug-flow (Mann *et al.*, 1995; Fdz-Polanco *et al.*, 1996). Plug-flow operation produces zones of different organic and nutrient strengths, which encourages the growth of different bacterial species (Fdz-Polanco *et al.*, 1996) and is particularly beneficial in tertiary treatment, to avoid direct competition between heterotrophic and autotrophic bacteria.

According to Eckenfelder and Quirk (1983), in order to up-scale results from pilot-scale reactors to full-scale systems, four aspects should remain constant:

- hydraulic loading
- film thickness
- surface area of packing medium
- hydraulic characteristics of the system

Therefore, the aim of this study was to characterise the hydraulics of the system by comparing the flow pattern and the active volume of two pilot-scale biological aerated filters using identical media under downflow and upflow conditions. The study of the hydrodynamic characteristics of the BAFs will be essential to scaling-up any future results obtained.

4.1.1 Theory

The flow patterns in a constant flow reactor (like BAFs) range between two ideal extreme conditions: plug-flow and completely mixed (Levenspiel, 1972). Plug-flow, is defined as an orderly flow through the reactor with no element of fluid overtaking or mixing with any other ahead or behind. There might be some lateral mixing but there is no mixing or diffusion along the flow path. In a completely mixed reactor, all its contents are stirred and uniform throughout. Hence, the exit stream would have the same composition as the fluid within the reactor (Levenspiel, 1972).

In plug-flow reactors the concentration of reactant decreases progressively through the system. In completely mixed flow reactors the concentration drops immediately to a low level. A system of completely mixed reactors connected in series could give an approximation to plug-flow conditions (Levenspiel, 1972). The reactant concentration would be uniform in individual reactors yet it would change progressively from reactor to reactor. Therefore, the greater the number of completely mixed reactors in series, the closer the behaviour the system approaches plug-flow. Most reactors however, are somewhere between plug-flow or completely mixed flow tending towards one or the

other due to the channelling of fluid, the creation of stagnant zones and recycling of the fluid. Deviations from idealised flow adversely affect reactor performance, and make the process difficult to model, and should be avoided.

In order to determine the flow pattern of a reactor the stimulus-response experiment is commonly used (Levenspiel, 1972). This gives an indication of the time that individual molecules stay in the vessel. In other words, the distribution of residence times of the flowing liquid. In a reactor the fluid elements require different lengths of time to exit the reactor as they travel along different routes through the reactor. The distribution of the recorded time that these elements of fluid take to exit the reactor is called the residence time distribution (RTD) of fluid.

RTD curves are obtained by plotting time against the concentration of the tracer. Time is usually in a dimensionless form, calculated by dividing the theoretical hydraulic time (t_{HRT}) by the actual running time (t) of the tracer. The t_{HRT} is calculated by dividing the volume of the reactor by the hydraulic flowrate. Other time measurements are the peak residence time (t_p) and the mean residence time (t_m). The peak residence time is the time at which the tracer reaches its highest concentration and can be obtained graphically. The mean residence time is calculated using the following expression:

$$t_m = \frac{\sum tC}{\sum C} \quad (4.1)$$

where t is time (min) and C is the tracer concentration of tracer (i.e. mg l⁻¹; mol l⁻¹).

If $t_m \neq t_{HRT}$ the reactor would present channelling or stagnant zones

If $t_m \neq t_p$ the system would present a degree of mixing.

The percentage of actual active volume used was calculated using the following expression:

$$\% \text{ active volume} = \frac{t_m}{tHRT} \times 100 \quad (4.2)$$

Tracer information can be used in conjunction with flow models to predict the performance of real reactors when ideal plug-flow or completely mixed conditions are not achieved. The method used depends on whether the reactor can be considered as a linear system or not. A process is linear when any change in magnitude of the stimulus results in a proportional change in magnitude of the response. Therefore, biological aerated filters can be considered as a linear process.

Dispersion models are used to characterise non-ideal flow within reactors. The dispersed plug-flow model or dispersed model considers the plug-flow of a fluid that has a certain degree of backmixing or intermixing, the magnitude of which is independent of position within the vessel. Depending on the degree of turbulence or intermixing the predictions from this model will range from plug-flow to completely mixed flow. The degree of backmixing is characterised by the parameter D , which is called the longitudinal or axial dispersion coefficient ($\text{m}^2 \text{s}^{-1}$). The dimensionless group $\left(\frac{D}{uL}\right)$ or reactor dispersion number is the parameter that measures the extent of axial dispersion. Where u is the liquid velocity (m s^{-1}) and L is the length of the reactor (m).

RTD curves obtained from reactors similar to BAFs normally have a pronounced tail (Fdz-Polanco *et al.*, 1996) indicating a large extent of dispersion. In such cases, the dispersion model can be fitted if the appropriate boundary conditions are assumed. In order to simplify calculations, the closed vessel boundary condition is generally used for BAFs (Mann *et al.*, 1995; Fdz-Polanco *et al.*, 1996). Thus, the dispersion number can be calculated by iteration using the following equation;

$$\sigma_{\theta}^2 = 2 \frac{D}{uL} - 2 \left[\frac{D}{uL} \right]^2 \left(1 - e^{(1 - uL/D)} \right) \quad (4.3)$$

Where σ_{θ}^2 is the variance of the values of the output tracer concentration. If $\frac{D}{uL}$ tends to zero, there is negligible dispersion, hence plug-flow condition will prevail. If $\frac{D}{uL}$ tends to infinity, a large degree of dispersion is occurring, therefore completely mixed condition will prevail.

Additionally, the equivalent number of tanks in series (N) can be found from the following expression;

$$N = \frac{1}{\sigma_{\theta}^2} \quad (4.4)$$

The value of N will elucidate the degree of backmixing and can be considered as a good approximation to obtain the flow pattern of submerged biofilters (Fdz-Polanco *et al.*, 1996). If $N > 4$, plug-flow conditions would prevail (Shar *et al.*, 1978; Grobicki and Stuckey, 1992).

4.2 Methods

Two identical PVC reactors were used. Each reactor was 0.2 m in diameter, 2.1 m in height and contained approximately 54 litres (2 m bed height) of 2-3 mm diameter silica sand medium (Figure 4.1). The medium was supplied by Tetra-The Capital Control Group. Rigid polypropylene meshes with 1.5 mm diameter holes were placed at the bottom and at the top of the beds to keep the media in place. Air was introduced at the bottom of the reactors by means of three 5 cm filter nozzles (Degremont, UK) around a centrally placed liquid distributor.

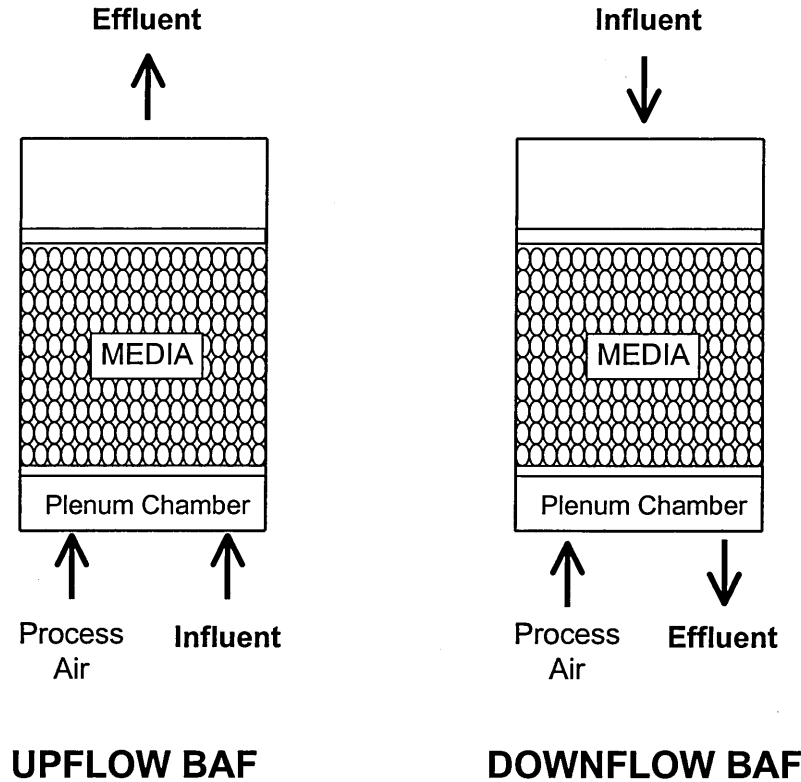


Figure 4.1. Schematic diagram of the pilot scale biological aerated filters.

Tracer studies were carried out at two liquid flowrates and four air flowrates to determine their influence over the performance of the reactors. The flowrates studied were those to be used as nominal flowrates during further operational tests. The liquid flowrates studied were 0.4 l min^{-1} (equivalent to hydraulic velocity of 1.8 m h^{-1}) and 0.6 l min^{-1} (2.7 m h^{-1}), while the air flowrates were $0, 2 \text{ l min}^{-1}$ (9 m h^{-1}), 3 l min^{-1} (13.4 m h^{-1}) and 4 l min^{-1} (17.9 m h^{-1}). A 0.1 M solution of sodium chloride (GPR, Merck, Poole, UK) was used as the tracer for all experiments. Previous work concluded that low molecular weight substances such as salts can be used as a representative tracer for aqueous solutions (Mann *et al.*, 1995; Fdz-Polanco *et al.*, 1996). In all experiments, the tracer was injected as a pulse input, this is, the volume applied was smaller than 1% of the reactor's volume and time of injection was less than 5% of the theoretical hydraulic residence time (Fdz-Polanco *et al.*, 1996). Since conductivity is directly related to

molar concentration, the output tracer concentration was measured using a conductivity meter (LTH Electronics Ltd, Luton, UK). During these experiments, the air and liquid distributors converged in the plenum chamber (10 cm height) located below the filter bed (Figure 4.1).

In all experiments the residence time distribution (RTD) curves were monitored over a period of 2 times the actual time/theoretical hydraulic retention time (tHRT) or until >90% of the tracer injected had been recovered.

4.3 Results and Discussion

The RTD curves for upflow and downflow BAFs at different air and liquid flowrates are presented in Figures 4.2, 4.3 and 4.4.

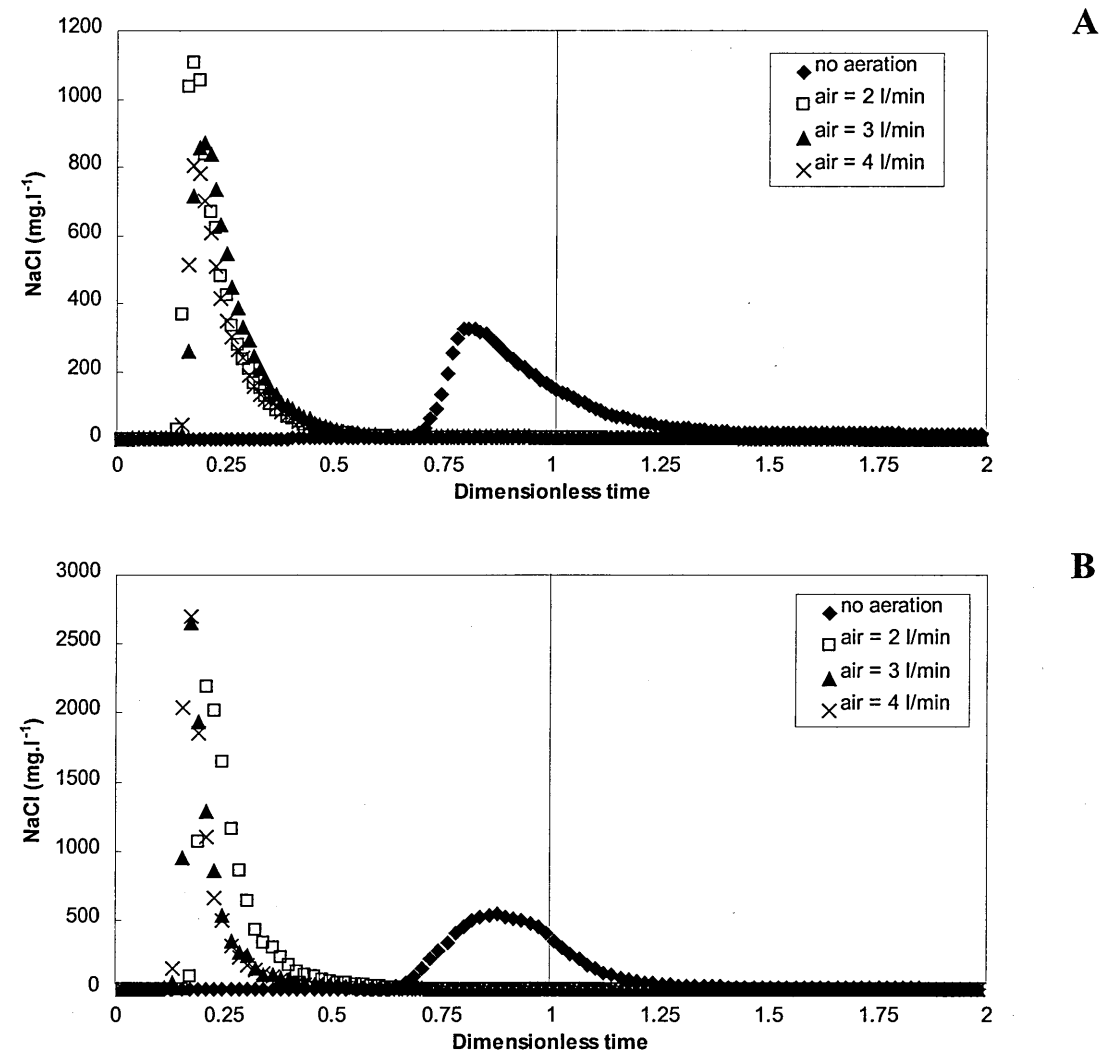


Figure 4.2. RTD curves for the downflow BAF running at liquid flowrates of 0.4 l min⁻¹ (A) and 0.6 l min⁻¹ (B), and at four different aeration rates.

It is clear that during experiments with the downflow reactor at hydraulic flowrates of 0.4 and 0.6 l min⁻¹ the aeration had an adverse effect on the operating volume of the reactor. When no air was present, the downflow reactor presented conditions very close to plug-flow. However, when air was supplied the peak time (t_p) was much smaller

than the theoretical hydraulic retention time (t_{HRT}) suggesting only a small fraction of the volume of the reactor was being used (Figure 4.2).

The upflow reactor did not appear to suffer the same problem at any of the hydraulic flowrates used (Figures 4.3 and 4.4). At a hydraulic flowrate of 0.4 l min^{-1} and when no air was supplied, the reactor ran under conditions very close to plug-flow. When air was injected there were difference again between the t_p and the t_{HRT} though not as severe as in the downflow reactor. As the air flowrate was increased, the conditions in the upflow reactor shifted slightly from completely mixed to plug-flow, as can be seen in the way the t_p for each air flowrate gets closer to the t_p without aeration.

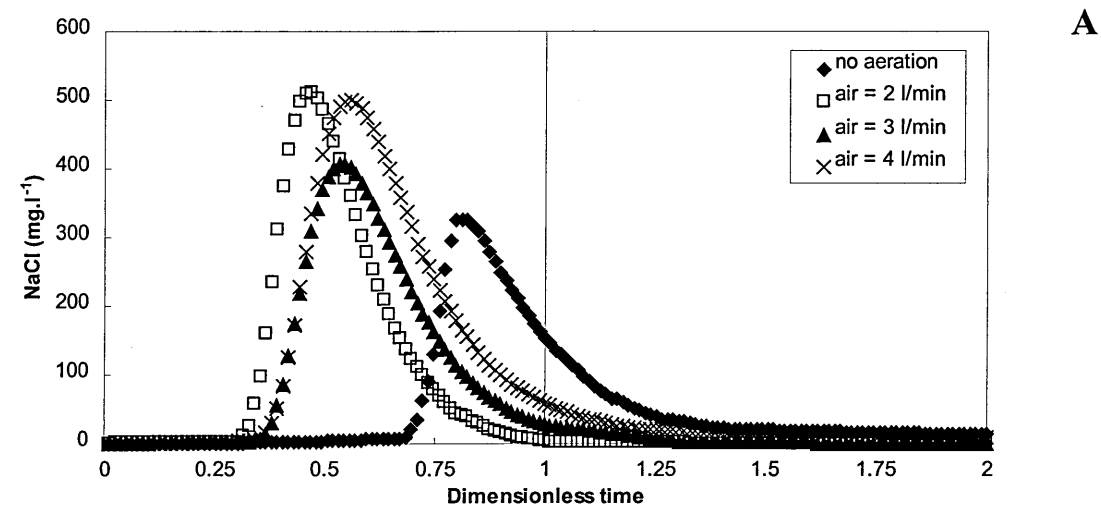


Figure 4.3. RTD curves for the upflow BAF running at a liquid flowrate of 0.4 l.min^{-1} and at four different aeration rates.

When the hydraulic flowrate was increased to 0.6 l min^{-1} the hydraulic pattern of the upflow reactor did not seem to change too much (Figure 4.3). With air there was greater degree of mixing. However, it appears that there was a greater degree of mixing at the low aeration rates and the mixing decreased as the aeration flowrate increased.

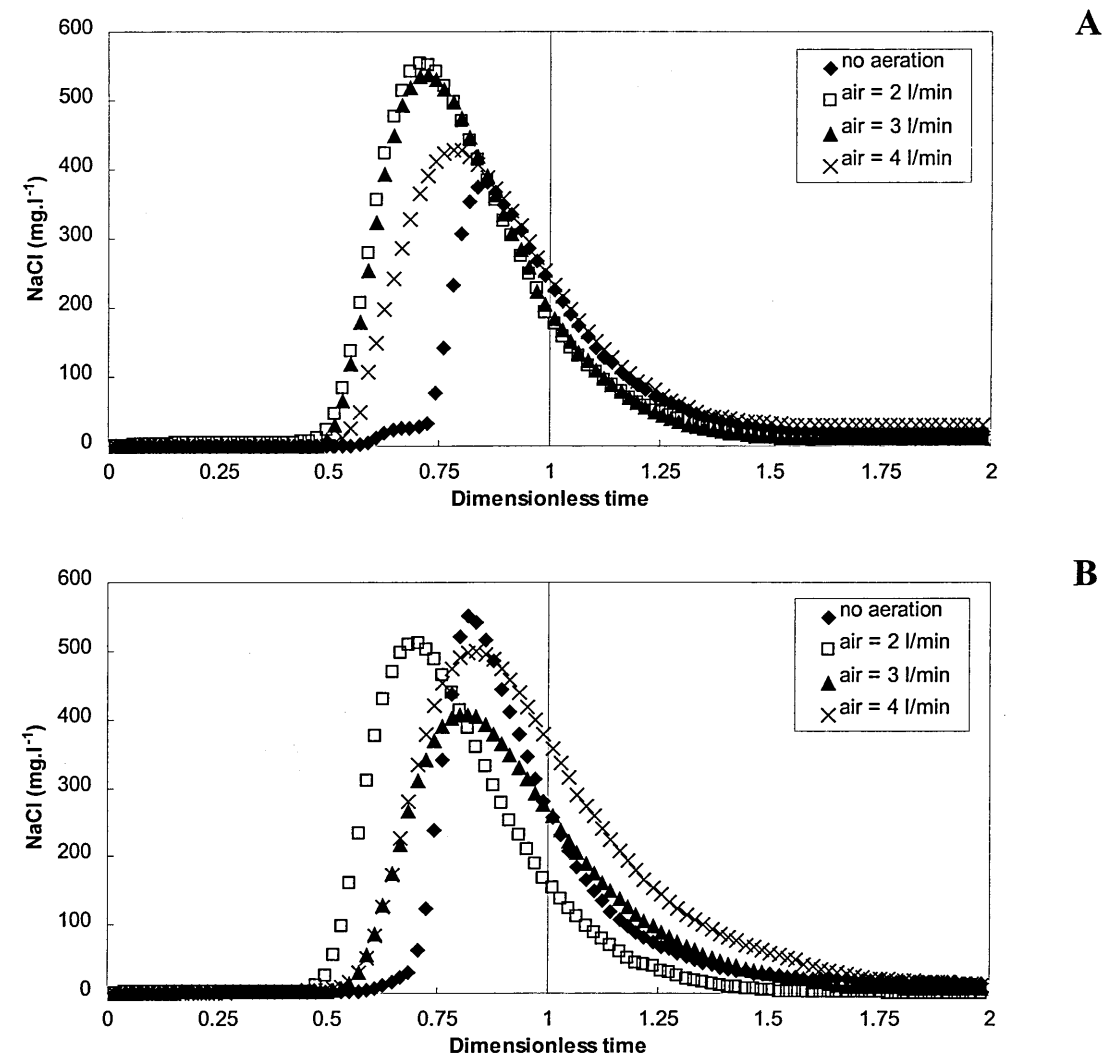


Figure 4.4. Two examples of RTD curves for the upflow BAF running at a liquid flowrate of 0.6 l.min^{-1} and at four different aeration rates.

The dispersion number and number of reactors in series were calculated (Table 4.1). The differences between the t_m and the $tHRT$ confirmed that severe channelling and stagnant zones were present in the downflow reactor.

Table 4.1. Theoretical hydraulic retention time (*tHRT*), mean residence time (*t_m*), peak time (*t_p*), dispersion number (*D/uL*) and number of reactors in series (*N*) for the upflow and downflow reactors. LF = liquid flowrate; AF = air flowrate.

LF l min ⁻¹	AF l min ⁻¹	<i>tHRT</i> min	Upflow Reactor				Downflow Reactor			
			<i>t_m</i> min	<i>t_p</i> min	<i>D/uL</i>	<i>N</i>	<i>t_m</i> min	<i>t_p</i> min	<i>D/uL</i>	<i>N</i>
0.4	0	79	72	64	0.011	43	18	13.5	0.072	7
0.4	2	79	70	64	0.020	26	18	13.5	0.065	8
0.4	3	79	67	60	0.025	20	20	15.5	0.056	10
0.4	4	79	71	64	0.024	22	20	14	0.060	9
0.6	0	52.5	51	45	0.022	25	11	9	0.049	11
0.6	2	52.5	43	37	0.028	18	13	11	0.047	11
0.6	3	52.5	46	40	0.025	20	12	9	0.060	10
0.6	4	52.5	48	43	0.026	20	12	10	0.044	12

In the upflow BAF at both liquid flowrates, and during all three air flowrates, the mean residence time was very similar to the *tHRT*, indicating the non-existence of channelling or stagnant zones in the reactor. The active volume of the upflow reactor fluctuated between 85-91% and 82-97% at hydraulic flowrates of 0.4 l min⁻¹ (1.8 m h⁻¹) and 0.6 l min⁻¹ (2.7 m h⁻¹), respectively. In both reactors the residence time in all cases was larger than the peak residence time of the tracer, indicating that a certain degree of mixing occurred (Table 4.1). It has been highlighted that a certain degree of mixing is needed in BAFs to allow the biomass to remove the maximum amount of nutrients from the liquid and to distribute nutrients along the biofilm (Mann *et al.*, 1995). Hence, it was thought that the conditions in the upflow BAF were favourable for biological treatment.

Dispersion numbers in the upflow reactor ranged from 0.011 to 0.026 under all conditions, which confirmed that a small amount of mixing occurred within the reactors. Analysis of the peak residence times revealed that aeration had no real effect on the hydrodynamics of the system. These results are contrary to what was expected. It has been reported that aeration flowrates greatly affected the mean and peak times of sunken and floating media BAFs (Mann *et al.*, 1995). However, recycled

polypropylene with densities smaller than 1000 kg m^{-3} was used as the floating media and 1050 kg m^{-3} as sunken media. The media used in the present experiment had a density of 2500 kg m^{-3} and remained relatively undisturbed by the different air flowrates tested. The upflow reactor showed a clear plug-flow pattern with a small amount of mixing which might result in good dissolved nutrient removal but poor solids filtration (Mann *et al.*, 1995). The smaller values of N obtained in the upflow reactor without aeration suggests that the upflow reactor was close to plug-flow conditions when no air was injected (Table 4.1). At higher air flowrates there was a small greater degree of mixing however, plug-flow conditions prevailed.

In the downflow BAF the active volume (as calculated by equation 4.2) was only 23-25% and 21-25% of the total volume at hydraulic flowrates of 0.4 l min^{-1} (1.8 m h^{-1}) and 0.6 l min^{-1} (2.7 m h^{-1}), respectively. The peak time was always less than the mean residence time, indicating a great degree of mixing. Without aeration, plug-flow conditions prevailed in the reactor. However, when air was applied a great degree of channelling and bypassing occurred. It was thought that the plenum chamber located at the bottom of the reactor was entrapping bubbles and caused the observed channelling. The media over the plenum chamber acted as a barrier that prevented the continuous flow of air through the reactor. The same phenomenon was not apparent in the upflow BAF due to the co-current flow of water and air. It was therefore decided to eliminate the plenum chamber by filling it with media. Since the upflow and the downflow reactors had to be running under identical conditions, the plenum chambers of both reactors were removed. Tracer studies were then undertaken at a constant aeration rate of 2 l min^{-1} (9 m h^{-1}) and at three different liquid flowrates 0.4 l min^{-1} (1.8 m h^{-1}), 0.6 l min^{-1} (2.7 m h^{-1}) and 0.8 l min^{-1} (3.6 m h^{-1}) (Figure 4.5).

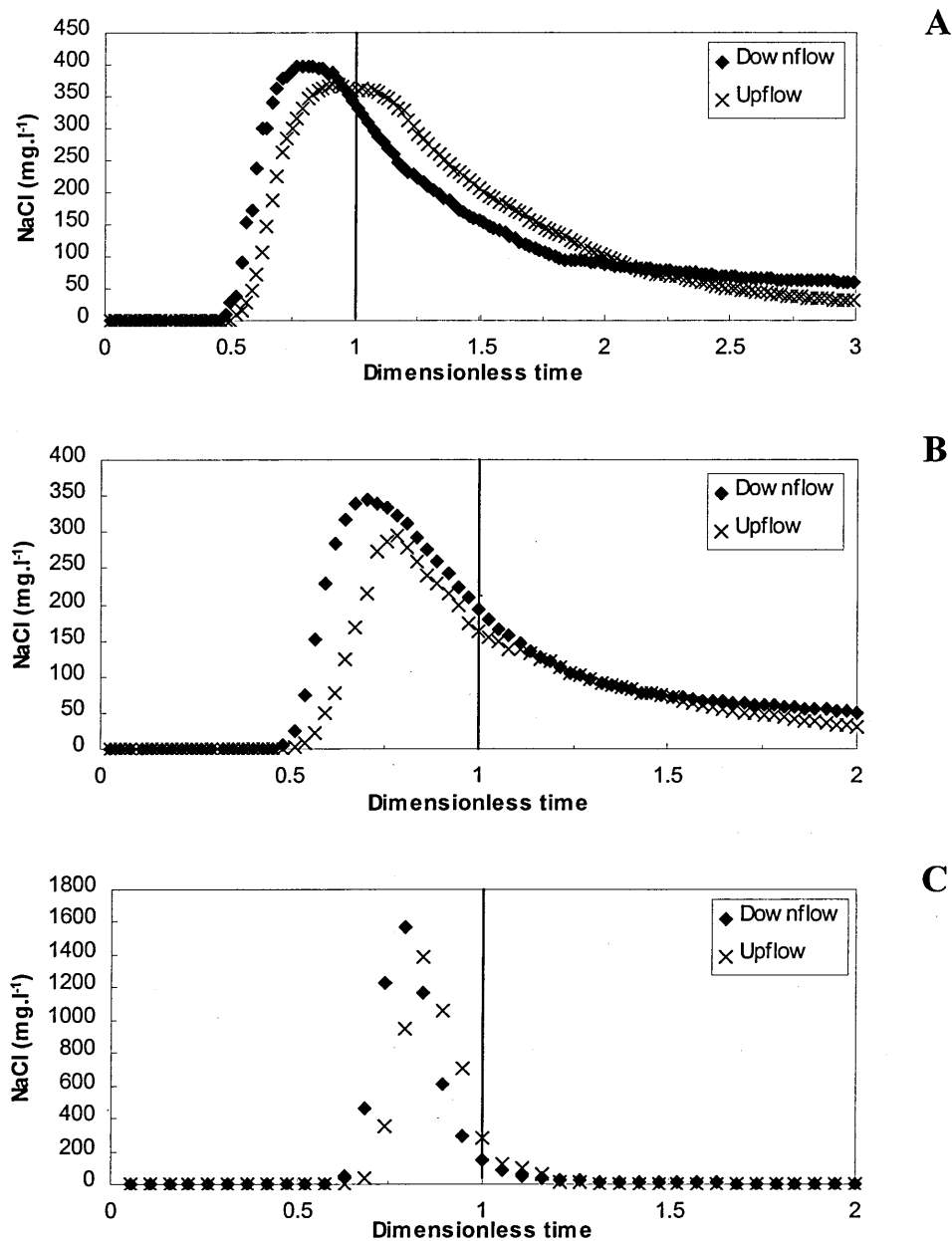


Figure 4.5. RTD curves for the downflow and upflow BAF after modifications running at an air flowrate of 2 l min⁻¹ and at liquid flowrates of 0.4 (A), 0.6 (B) and 0.8 (C) l min⁻¹.

Changes in the reactors increased the mean residence time, the peak residence time and the dispersion number in the upflow reactor (Table 4.2). The larger dispersion number and smaller N highlight the fact that a greater degree of mixing was achieved.

Table 4.2. Theoretical hydraulic retention time (*tHRT*), mean residence time (*t_m*), peak time (*t_p*), dispersion number (*D/uL*) and number of reactors in series (*N*) for the upflow and downflow reactors after modifications. LF = liquid flowrate; AF = air flowrate.

LF l.min ⁻¹	AF l.min ⁻¹	<i>tHRT</i> min	Upflow Reactor				Downflow Reactor			
			<i>t_m</i> min	<i>t_p</i> min	<i>D/uL</i>	<i>N</i>	<i>t_m</i> min	<i>t_p</i> min	<i>D/uL</i>	<i>N</i>
0.4	2	74	72	67	0.045	11	68	58	0.047	11
0.6	2	50	42	39	0.055	9.6	39	35	0.071	7.5
0.8	2	37	32	31	0.007	65	31	29	0.014	31

In the downflow reactor, the hydraulic conditions improved dramatically. In both reactors more than 85% of the theoretical volume was used. No channelling or stagnant zones existed, though the peak time was always less than the mean residence time, indicating a certain degree of mixing, which decreased at higher flowrates. At a hydraulic flowrate of 0.4 l min⁻¹ (1.8 m h⁻¹) there appeared to be a great degree of mixing in both reactors. When the flowrate was increased to 0.6 l min⁻¹ (2.7 m h⁻¹) the reactors approximated plug-flow conditions more closely but there was a degree of mixing. At 0.8 l min⁻¹ (3.6 m h⁻¹) plug-flow conditions prevailed in both reactors. According to the dispersion number and *N*, there was a significant amount of mixing at liquid flowrates of 0.4 and 0.6 l min⁻¹ and at 0.8 l min⁻¹ plug-flow conditions prevailed. After modifications a greater degree of tailing of the RTD curves occurred (Figure 4.5) which, according to Fdz-Polanco *et al.* (1996), is typical in these types of reactors.

4.4 Conclusions

- An increase in air velocities ($0 - 17.9 \text{ m h}^{-1}$) did not significantly affect liquid mixing within the reactors and retention times did not decrease as expected, probably due to the high density (2500 kg m^{-3}) of the media used.
- The existence of a plenum chamber under the media bed caused severe reductions in the active volume in the downflow reactor probably due to the entrapment of air bubbles. Elimination of the plenum chamber by means of the addition of media solved the problem.
- After modifications, the active volume of the reactors was $>85\%$ of the theoretical volume hence, channelling or stagnant zones were practically non-existent. Increases in the nominal liquid flowrate from 0.4 l min^{-1} (hydraulic velocity of 1.8 m h^{-1}) to 0.8 l min^{-1} (3.6 m h^{-1}) affected the hydrodynamic conditions of the reactors in a greater way than changes in the air flowrate
- A degree of mixing was observed within the reactors (as shown by differences between the peak time and mean residence time values). However plug-flow conditions prevailed. Both columns presented similar hydraulic conditions, which is important for scaling-up purposes.

Chapter 5

COMBINED SOLUBLE COD AND AMMONIA
REMOVAL IN A DOWNFLOW AND AN UPFLOW
BIOLOGICAL AERATED FILTER

COMBINED SOLUBLE COD AND AMMONIA REMOVAL IN A DOWNFLOW AND AN UPFLOW BIOLOGICAL AERATED FILTER

ABSTRACT

In order to compare the performance of two biological aerated filters during combined soluble COD and ammonia removal, two pilot-scale reactors were operated with identical media and under downflow and upflow conditions, using domestic settled sewage as a feed. Loading rates to both reactors were around $1.8 \text{ kg SS m}^{-3} \text{ d}^{-1}$, $2.4 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$, $5.2 \text{ kg tCOD m}^{-3} \text{ d}^{-1}$ and $0.51 \text{ kg NH}_3\text{-N m}^{-3} \text{ d}^{-1}$ (based on a working volume in the media section of 25 l in each reactor). Removal rates of SS, soluble and total COD for both reactors were very similar. The upflow reactor achieved average removal rates of $1.3 \text{ kg SS m}^{-3} \text{ d}^{-1}$, $1.9 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$ and $3.9 \text{ kg tCOD m}^{-3} \text{ d}^{-1}$. The downflow reactor reached average removal rates of $1.4 \text{ kg SS m}^{-3} \text{ d}^{-1}$, $1.6 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$ and $4.2 \text{ kg tCOD m}^{-3} \text{ d}^{-1}$ (based on a working volume in the media section of 25 l in each reactor). Plug-flow conditions and autotrophs-heterotrophs competition caused autotrophic activity to be located at the bottom of the reactor in the downflow column and at the top in the upflow unit. Co-current air and water backwashing could have washed-out autotrophs in the upflow BAF..

5.1 Introduction

Biological aerated filters (BAFs) are three-phase reactors for wastewater treatment that use granular media for biomass support. BAFs are capable of achieving carbonaceous matter removal, solids filtration, nitrification and/or denitrification in a single reactor. There is a range of designs with variations in flow directions and the type of media used. In downflow systems the influent is introduced to the top of the filter medium and aeration is applied in counter-current mode. The effluent exits via underdrains, normally to a storage tank with an overflow discharge, where a proportion of it is re-used for backwashing the filter. In upflow systems the influent is introduced at the base of the filter near the point where the process air is injected. The air and wastewater flow co-currently through the medium and the effluent exits at the top of the filter (Iida and Teranishi, 1984). Both configurations have benefits and disadvantages. Downflow reactors are limited to a maximum flowrate that will depend on the height and pressure drop across the reactor bed; on the other hand, upflow reactors are limited by the pump capacity. Downflow systems with counter-current air/water flow may offer the advantage of having the air supplied in contact with the final effluent for a greater period of time (González-Martínez and Wilderer, 1991). In upflow reactors, suspended solids and the biomass within the reactors are more uniformly distributed, which allows higher influent flowrates to be used (Peladan *et al.*, 1997). However, due to the fact that the untreated wastewater is supplied at the bottom of the reactor along with the process air, air nozzles blockage is possible (Desbos *et al.*, 1990).

Biological aerated filters can be used for simultaneous suspended solids (SS), chemical oxygen demand (COD) and ammonia removal (combined secondary and tertiary treatment) or for SS and ammonia removal (tertiary treatment). When used for combined secondary and tertiary treatment, the ammonia removal rates tend to be lower than those achieved during tertiary treatment due to the presence of carbonaceous matter; ammonia removal rates of up to $1 \text{ kg NH}_3\text{-N m}^{-3} \text{ d}^{-1}$ have been

obtained (Table 5.1). Biological aerated filters used for tertiary treatment are able to reach higher nitrification rates of up to $1.2 \text{ kg NH}_3\text{-N m}^{-3} \text{ d}^{-1}$ (Table 5.1).

Table 5.1. Nitrification rates obtained in biological aerated filters used for combined secondary and tertiary treatment, and for tertiary treatment only.

Reactor configuration	Type of Treatment	Maximum nitrification rate ($\text{kg NH}_4\text{-N m}^{-3} \text{ d}^{-1}$)	Reference
Upflow	combined	1.0	Rogalla and Bourbigot (1990)
Upflow	combined	0.19	Carrand <i>et al.</i> (1990)
Downflow	tertiary	$0.39\text{--}0.84 \ddagger$ (at 22°C)	Rogalla and Payraudeau (1988)
Downflow	tertiary	$0.6 \dagger$	Schlegel (1988)
Downflow	tertiary	$0.9\text{--}1.2$	Sakuma <i>et al.</i> (1993)
Downflow	tertiary	0.5 (at 13°C)	Paffoni <i>et al.</i> (1990)
Upflow	tertiary	$0.12\text{--}0.54$	Fruhen <i>et al.</i> (1994)
Upflow	tertiary	0.55	Fdz-Polanco <i>et al.</i> (1994)
Downflow	tertiary	0.22	Terayama <i>et al.</i> (1997)

$\dagger \text{ kg TKN m}^{-3} \text{ d}^{-1}$

$\ddagger \text{ kg NH}_4\text{-N m}^{-3} \text{ media d}^{-1}$

As in other fixed-film processes, nitrification in BAFs can be inhibited in a number of ways. Nitrifying biomass in BAFs can be adversely affected by toxic substances (Beg and Hassan, 1987), temperature, low oxygen concentrations (Fdz-Polanco *et al.*, 1994), pH (Villaverde *et al.*, 1997b) and substrate concentration (Rogalla and Bourbigot, 1990). Moreover, direct competition with non-nitrifying biological species is one of the main reasons for the inhibition of nitrification (Lee and Welander, 1994).

The aim of this study was to compare the performance of two biological aerated filters during combined soluble COD and ammonia removal. The two pilot-scale biological aerated filters were operated with identical media under downflow and upflow conditions.

5.2 Materials and Methods

Experiments were undertaken in two small pilot-scale PVC reactors, of 0.2 m diameter and with a total bed height of 2 m (Figures 5.1 and 5.2). Primary settled sewage from Cranfield University's sewage treatment works was used as a feed by means of a centrifugal pump (Mono Pumps Ltd. Type CML 223). Compressed air was supplied at a liquid:air ratio of 1:10. Airflow was controlled via a Platon glass tube VA flowmeter with a 1-10 l min⁻¹ capacity (CT Platon Ltd, Basingstoke, UK) during normal operation and a Platon glass tube VA flowmeter with a 2-25 l min⁻¹ capacity (CT Platon Ltd, Basingstoke, UK) during backwashing. Liquid flowrate was controlled with the use of a stop cock valve (Durapipe S&LP, Staffordshire, UK) and a bulb valve (Durapipe S&LP, Staffordshire, UK) for each reactor.

The upflow column had a height of 2.1 m while the downflow reactor required a height of 3.1 m to provide enough pressure to overcome resistance in the filter. The extra section at the top of the downflow reactor was added after tracer studies were undertaken to characterise the hydraulic characteristics of both reactors. It was demonstrated that plug-flow conditions prevailed within the reactors when their overall height was 2.1 m (Chapter 4). In order to determine if the extra section would provide any changes in the removal capacity of the downflow BAF, wastewater samples were taken at heights of 2 and 3 m. It was found that the concentration of all the parameters analysed (SS, sCOD, tCOD, NH₃-N) were the same at both heights. Hence, the section above the media was found to be completely mixed and did not provide any further improvement in terms of wastewater treatment. Therefore, in order to be able to compare directly the results from both reactors, the theoretical hydraulic retention time and working volume for each reactor were calculated based on a total column height of 2.1 m and a media height of 2 m. (Table 5.2).

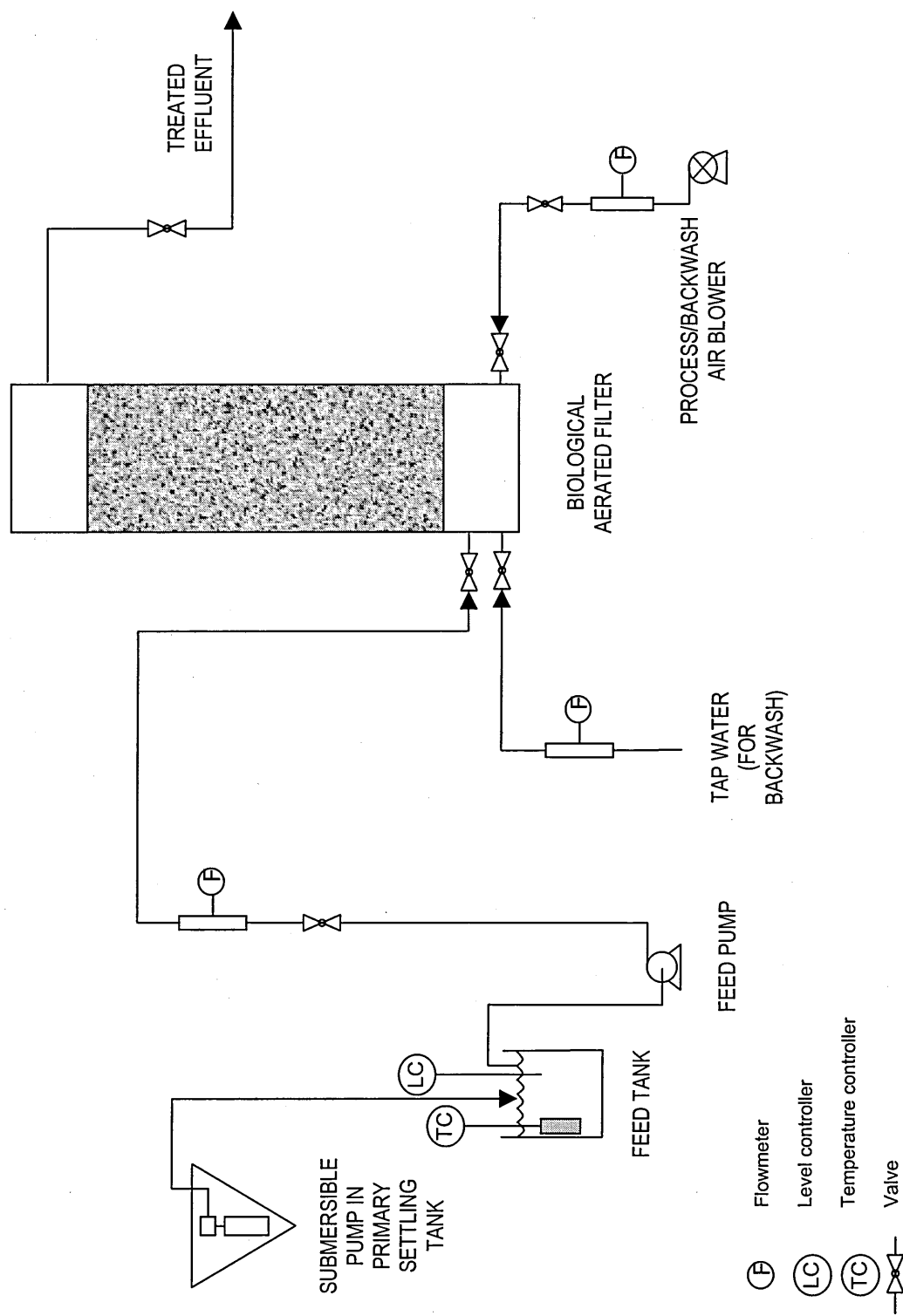


Figure 5.1. Schematic diagram of the small pilot-scale upflow BAF.

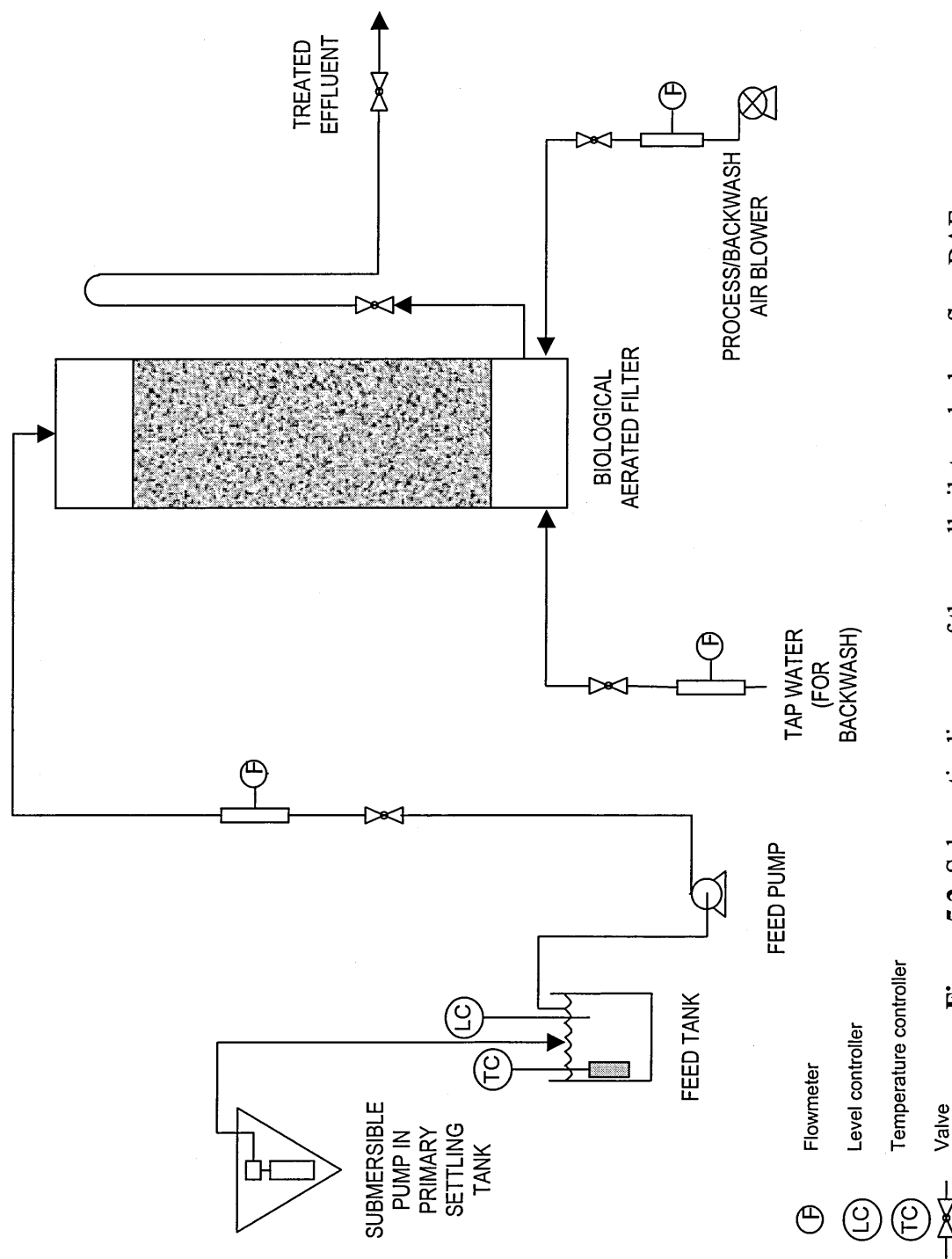


Figure 5.2. Schematic diagram of the small pilot-scale downflow BAF.

Table 5.2. Dimensions of the small pilot-scale BAFs running in upflow and downflow mode.

	Upflow mode	Downflow mode
Column diameter (m)	0.2	0.2
Sampling ports location (m)*	0.2, 0.4, 0.6, 0.8, 1.2, 1.4, 1.6, 1.8	0.2, 0.4, 0.6, 0.8, 1.2, 1.4, 1.6, 1.8
Column height (m)	2.1	3.1
Media bed height (m)	2	2
Empty bed volume - EBV (m ³)	6.6×10^{-3}	9.7×10^{-3}
Media volume (m ³)	3.8×10^{-3}	3.8×10^{-3}
Working volume (m ³)	2.6×10^{-3}	5.7×10^{-3}
Working volume in media section (m ³)	2.5×10^{-3}	2.5×10^{-3}

The small pilot-scale reactors were backwashed manually, using tap water to clean the media bed. The backwash cycle in the small pilot-scale columns consisted of:

1. feed pump and air switched off
2. five min delay
3. backwashing air comes on; runs for 5 min
4. backwash water supplied for 10 min (along with backwash air)
5. backwash air off; backwash water on for 5 min
6. backwash water off
7. five min delay
8. feed pump and air restored; normal operation resumed

The diameter of the reactors was at least 50 mean particle diameters in size to limit wall effects (Lang *et al.*, 1993). Each reactor contained 38 l of media (Table 5.3).

Table 5.3. Media properties

Media Characteristic	Value
Shape	Spherical
Size	2.00 - 3.15 mm
Material	Sand quartz
Voidage	40 %
Density	2500 kg m ⁻³
Specific Surface Area	1500 m ² m ⁻³

Backwashing was carried out on a daily basis. Suspended solids (SS) were analysed using standard methods (Greenberg *et al.*, 1992). Soluble COD (sCOD), total COD (tCOD) and ammonia (NH₃-N) were analysed using the HACH test-n-tube methods (Camlab, Nuffield, UK). Dissolved oxygen was measured using a Jenway DO2-9071 meter (Merck Ltd., Leicester, UK); pH and temperature were measured with a Merck 3071 pH meter (Merck Ltd., Leicester, UK).

5.3 Results and Discussion

The hydraulic flowrate throughout the experiments was $0.43 \text{ m}^3 \text{ d}^{-1}$ which corresponded to a theoretical hydraulic retention time (tHRT) of 93 min based on a working volume of 25 l in each reactor. Experiments in the downflow and upflow column lasted for approximately 85 and 100 d, respectively. Steady-state for SS, sCOD and tCOD removal was reached in approximately 20 d, when the difference between influent and effluent concentration of the parameters was constant, thus reaching a steady removal (Figures 5.3 to 5.8). However, removal of $\text{NH}_3\text{-N}$ varied along the whole experiment (Figures 5.9 to 5.10). Nitrification was never fully achieved so it was not possible to confirm that steady-state was reached.

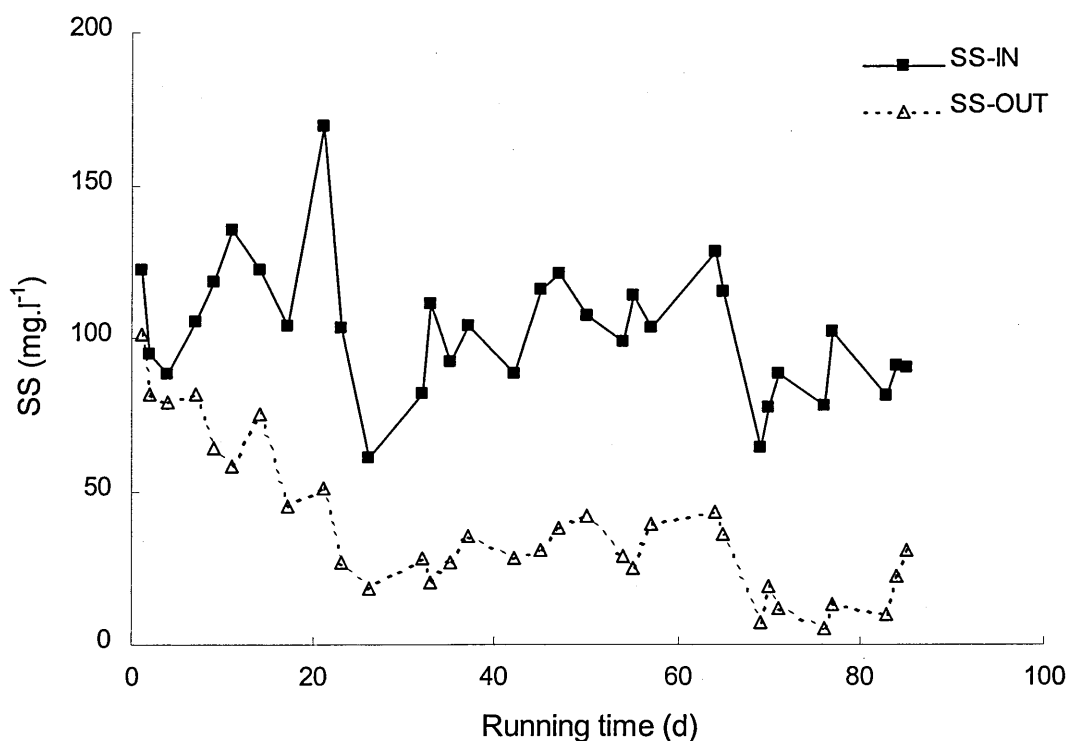


Figure 5.3. Influent and effluent concentration of suspended solids in the upflow BAF.

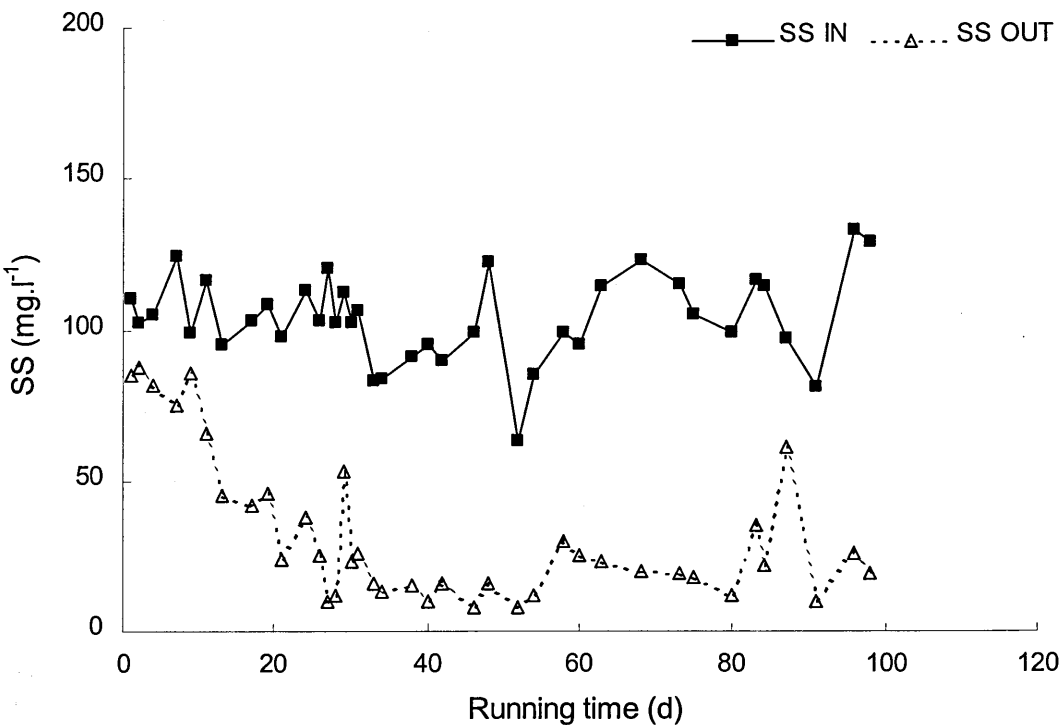


Figure 5.4. Influent and effluent concentration of suspended solids in the downflow BAF.

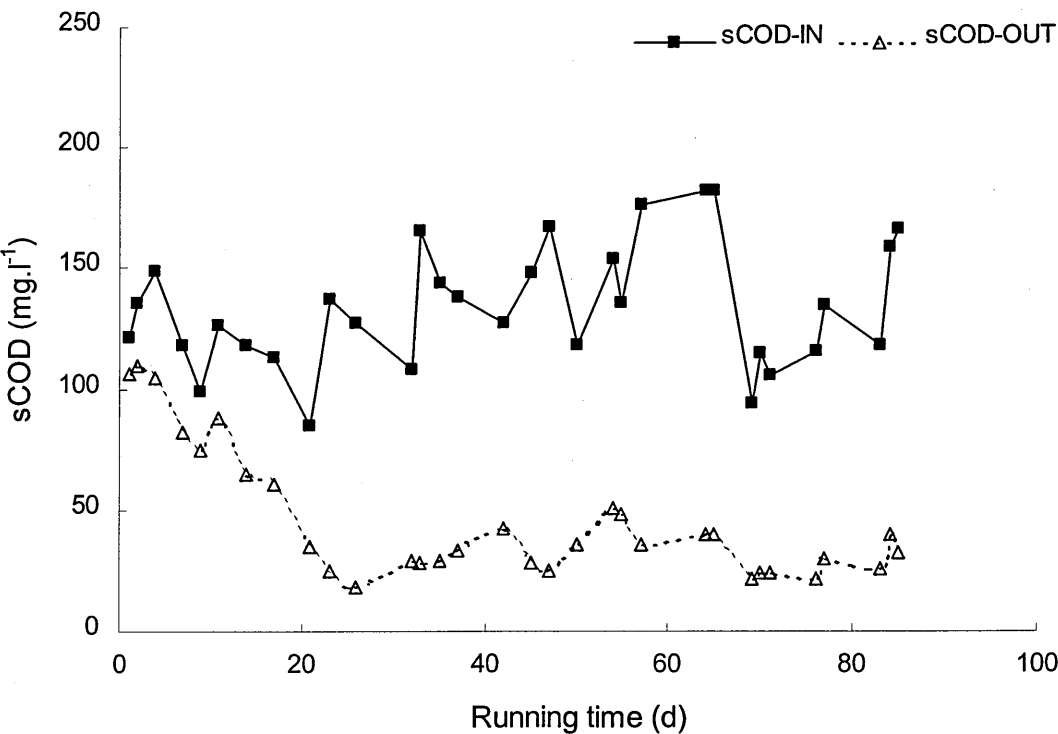


Figure 5.5. Influent and effluent concentration of soluble COD in the upflow BAF.

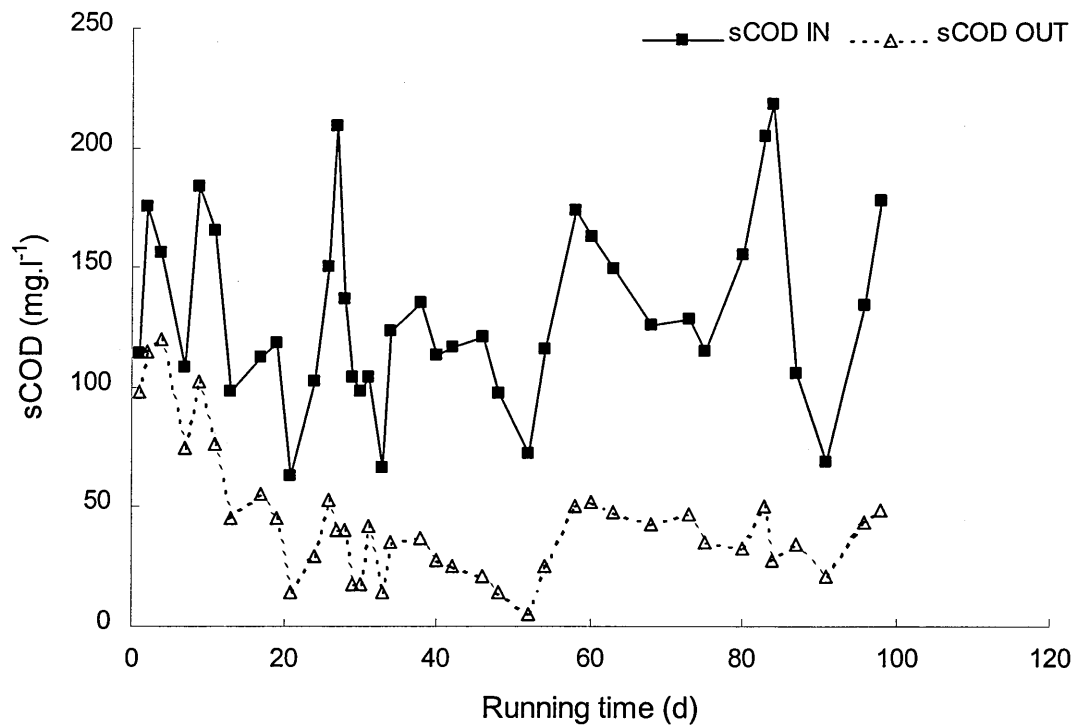


Figure 5.6. Influent and effluent concentration of soluble COD in the downflow BAF.

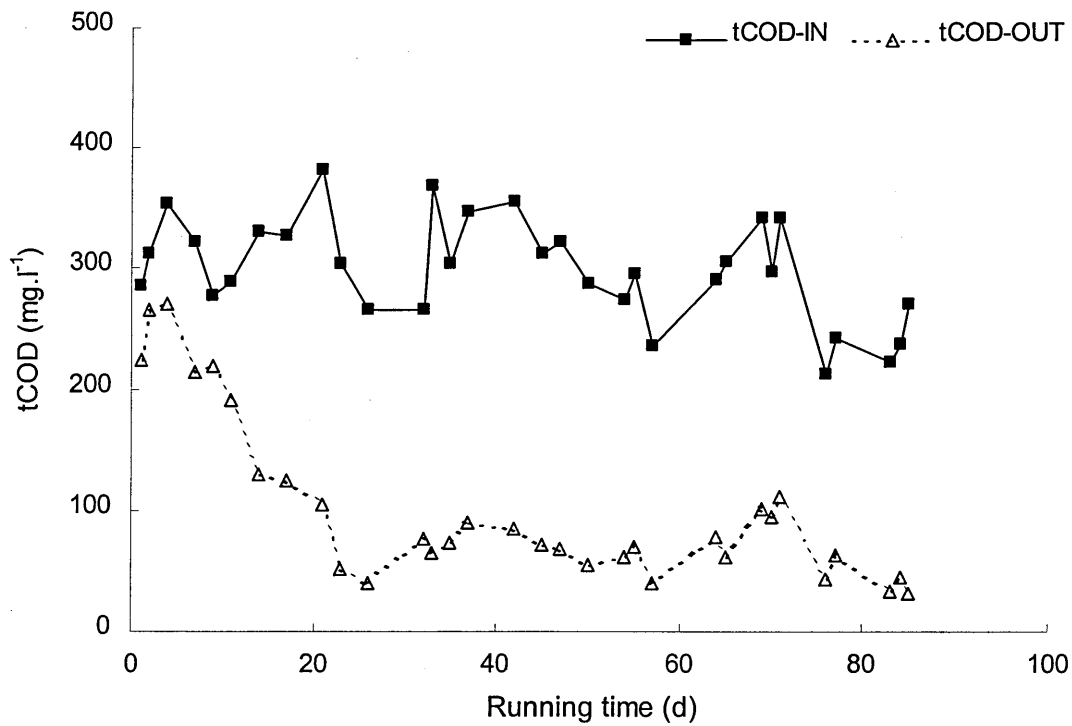


Figure 5.7. Influent and effluent concentration of total COD in the upflow BAF.

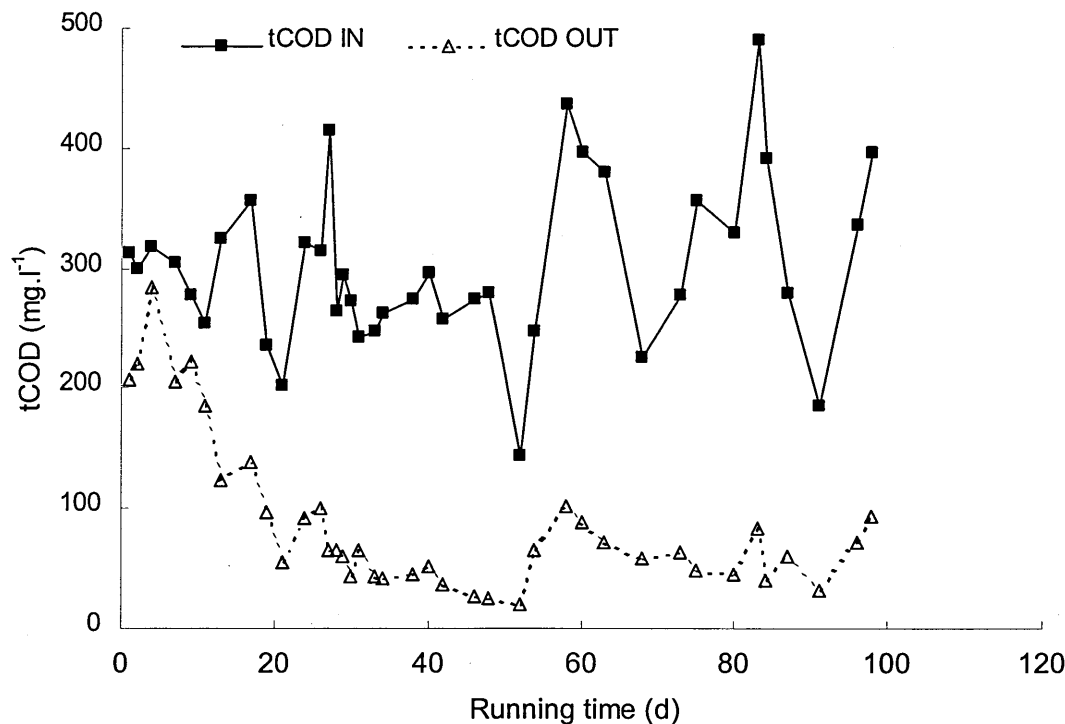


Figure 5.8. Influent and effluent concentration of total COD in the downflow BAF.

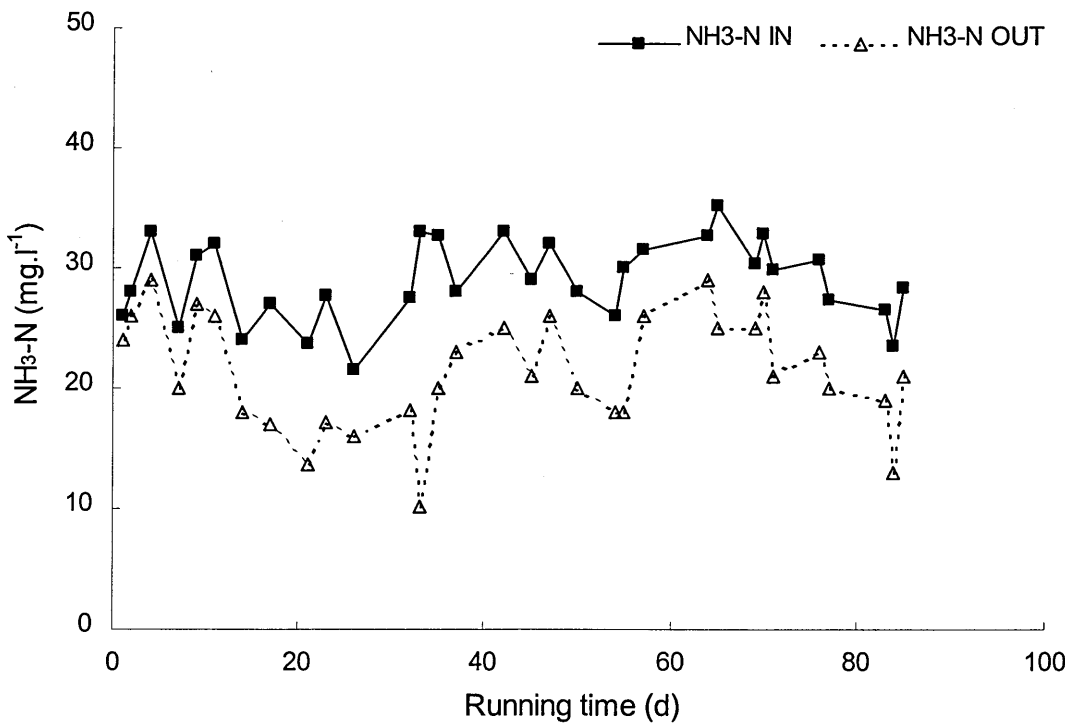


Figure 5.9. Influent and effluent concentration of ammonia in the upflow BAF.

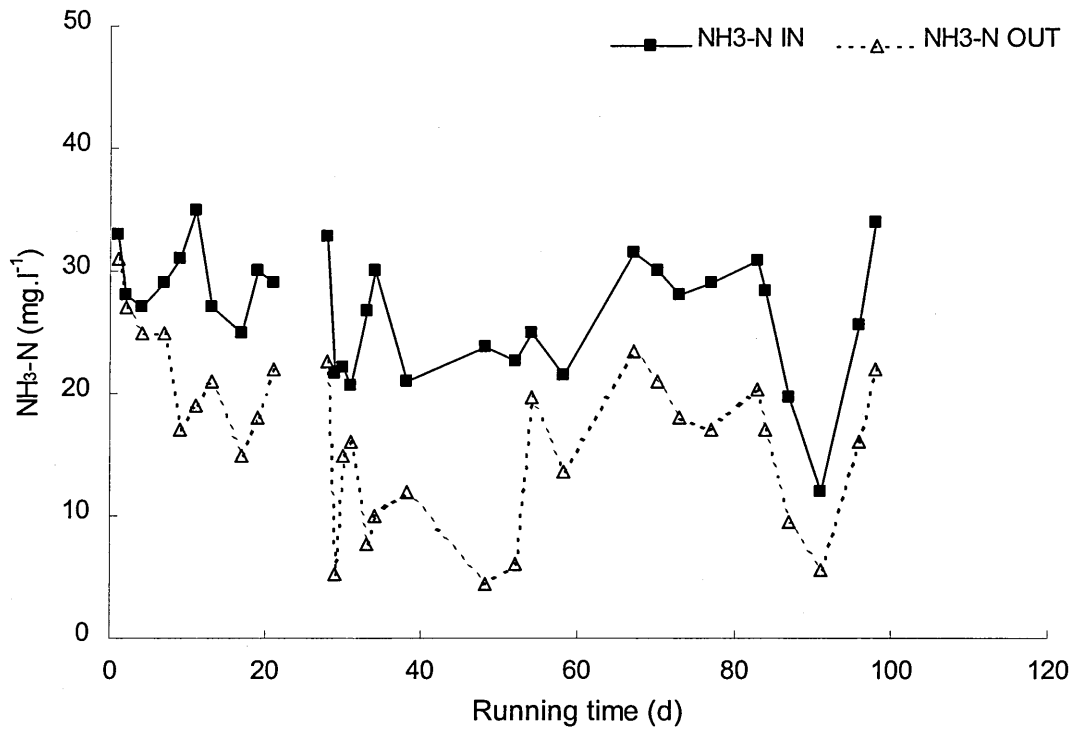


Figure 5.10. Influent and effluent concentration of ammonia in the downflow BAF.

Temperature ranged from 20 to 25°C, pH decreased from a value of 7.8 at the influent to pH 7.0 at the outlet; the dissolved oxygen concentration within the reactor was maintained at >4 mg l⁻¹. According to Gonzalez-Martinez and Wilderer (1991), in downflow reactors where carbonaceous matter and ammonia are removed simultaneously, microorganisms living in the lower portion of the filter receive air rich in oxygen, and therefore will not suffer oxygen limitation. During the present experiments air was supplied in such a way that oxygen limitation did not occur throughout the whole height of the reactors (upflow and downflow).

According to Figures 5.3 to 5.8, both upflow and downflow reactors gave stable effluent concentrations of SS, sCOD and tCOD, even when the influent concentrations varied greatly. Although the experiments were undertaken simultaneously, influent and effluent samples were taken at different times and the steady-state operating periods were different for each reactor. Therefore, the concentration of influent parameters varied slightly from one reactor the other (Tables 5.4 and 5.5).

Table 5.4. Average and range of influent and effluent concentrations of SS, sCOD, tCOD and NH₃-N for the upflow BAF under steady-state conditions (24 samples).

	Influent (mg l ⁻¹)		Effluent (mg l ⁻¹)		Removal (%)	
	Average	Range	Average	Range	Average	Range
Suspended Solids	99	61-169	27	5-51	74	61-94
Soluble COD	137	85-182	32	18-51	76	59-86
Total COD	295	212-382	67	32-112	78	68-88
Ammonia, NH ₃ -N [†]	29	22-35	21	10-29	29	11-69

[†] unsteady state

The percentages of removal were calculated after the reactors reached steady-state. As explained earlier, steady-state did not seem to have been reached in terms of ammonia removal. However, in order to maintain consistency, the percentage of removal for ammonia presented in Tables 5.4 and 5.5 were calculated from samples taken from running day 21 onwards (when steady-state was reached for SS, sCOD and tCOD).

Table 5.5. Average and range of influent and effluent concentrations of SS, sCOD, tCOD and NH₃-N for the downflow BAF under steady-state conditions (30 samples).

	Influent (mg l ⁻¹)		Effluent (mg l ⁻¹)		Removal (%)	
	Average	Range	Average	Range	Average	Range
Suspended Solids	103	63-133	22	8-61	79	37-92
Soluble COD	128	63-128	33	5-53	74	60-93
Total COD	303	144-490	58	20-102	81	68-91
Ammonia, NH ₃ -N [†]	26	12-34	15	4-24	42	9-81

[†] unsteady state

The downflow reactor was more efficient in removing SS, sCOD and NH₃-N than the upflow BAF. However, the latter achieved a slightly better removal of sCOD. The main difference in treatment was in terms of ammonia removal. Although steady-state for ammonia removal was not reached in any of the reactors, it can be highlighted that the downflow reactor achieved an average ammonia removal of 42% against 29% from the upflow BAF (Tables 5.4 and 5.5).

The organic loadings to both reactors were very similar (Table 5.6) and fell within the ranges of loadings previously reported for BAFs (Rogalla and Bourbigot, 1990; Pujol *et al.*, 1994). The removal rates are also very similar for the upflow and the downflow reactors. The downflow reactor achieved a larger removal rate for SS, total COD and NH₃-N, while the upflow reactor reached a higher removal for soluble COD.

Table 5.6. Loading and removal rates of SS, sCOD, tCOD and NH₃-N for the upflow and downflow BAFs based on working volumes of 25 l for each reactor. Average loading (range of loading).

	Upflow reactor		Downflow reactor	
	Loading rate (kg m ⁻³ d ⁻¹)	Removal rate (kg m ⁻³ d ⁻¹)	Loading rate (kg m ⁻³ d ⁻¹)	Removal rate (kg m ⁻³ d ⁻¹)
Suspended solids	1.7 (1.1-2.9)	1.3 (0.7-2.0)	1.8 (1.1-2.3)	1.4 (0.6-1.9)
Soluble COD	2.4 (1.6-3.1)	1.9 (1.2-2.5)	2.2 (1.1-3.8)	1.6 (0.8-3.3)
Total COD	5.1 (3.7-6.6)	3.9 (2.9-5.3)	5.2 (2.5-8.5)	4.2 (2.1 –7.0)
Ammonia, NH ₃ -N [†]	0.51 (0.37-0.61)	0.15 (0.06-0.40)	0.44(0.21-0.59)	0.19 (0.08-0.35)

[†] unsteady state

The removal of NH₃-N was the main difference in performance between the reactors. The downflow unit reached a nitrification rate approximately 50% greater than that reached in the upflow reactor. Rogalla and Bourbigot (1990) found that BAFs used primarily to remove carbonaceous matter have been able to treat ammonia loadings up to 1 kg NH₃-N m⁻³ d⁻¹. However, work by Carrand *et al.* (1990) reported a maximum nitrification rate of 0.19 kg NH₃-N m⁻³ d⁻¹, which is very close to the nitrification rates reached in the present experiment. In the present experiments both BAFs proved to be efficient in the removal of SS, sCOD and tCOD at the loading applied. Nevertheless,

nitrification was limited. Boller *et al.*, 1994 demonstrated that the efficiency of reactors undertaking simultaneous secondary and tertiary treatment was dependent on the amount of solids and carbonaceous material present and that nitrification could be inhibited by the presence of a high concentrations of solids. Although in the present experiment the loading of solids was not particularly high, the presence of solids and carbonaceous matter could have adversely affected nitrification.

Previous work by Grasmick *et al.* (1984) and Newbigging *et al.* (1995) have directly compared the nitrification capacity of BAFs under upflow and downflow conditions but have reached different conclusions. Using synthetic sewage as a feed, Grasmick *et al.* (1984) reported that nitrification was achieved at loading rates of up to $0.6 \text{ kg NH}_4^+ \text{-N m}^{-3} \text{ d}^{-1}$. The downflow reactor performed better than the upflow BAF when the ammonia loading rate increased but the difference was minimal. The downflow and upflow BAFs were equally affected by increases in the liquid and the air velocity but overall the downflow reactor achieved better ammonia removal. Newbigging *et al.* (1995) found that the downflow BAF achieved better solids separation than the upflow BAF; however, the latter achieved higher nitrification removal rates when primary settled sewage was used as a feed. Though the upflow BAF required a longer period of acclimatisation (5 weeks), once it reached steady-state, nitrification remained constant even during cold weather operation, whilst the downflow reactors performance was reduced. Removal efficiencies for both reactors were greater than 70% for SS and ammonia and 80% for biochemical oxygen demand (BOD). The downflow BAF required twice as much air than the upflow reactor on an air rate per surface area basis. The two types of reactor required different media bed heights to achieve the desired performance. The downflow BAF operated at a media bed height of 2.5 m while the upflow BAF operated at 3.9 m indicating that a downflow reactor plant footprint would be twice as large as required by an upflow unit. Newbigging *et al.* (1995) conclude that both reactors achieved consistently high effluent qualities, such that the preference between the two mode of operation should be based on the capital and running costs of each process.

In the present experiment, the downflow column achieved greater ammonia removal rates than the upflow unit; thus, verifying the results obtained by Grasmick *et al.* (1984). As mentioned earlier (Faup *et al.*, 1982), high concentrations of carbonaceous material can promote the growth of heterotrophic bacteria and thus limit the growth of autotrophic bacteria, reducing nitrification. Therefore, under such circumstances heterotrophic organisms always out-compete autotrophic bacteria. Only when the concentration of organic matter decreases, do the autotrophs have a real chance to grow. Thus, it would be expected that in upflow columns the heterotrophs would tend to be located at the bottom of the reactor while in downflow columns they would be located at the top. Conversely, autotrophs should be found at the bottom in downflow columns and at the top in upflow units. Figures 5.11 and 5.12 confirm such results.

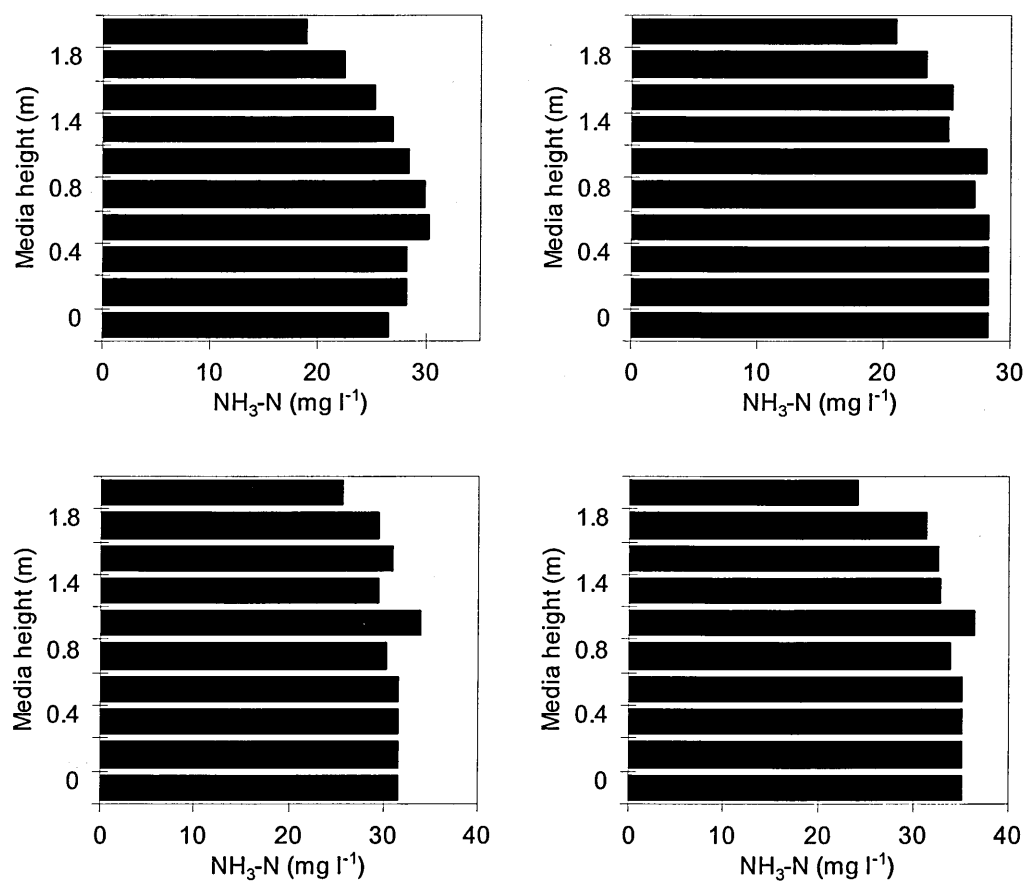


Figure 5.11. Profiles for the removal of ammonia in the upflow reactor (running days 71, 77, 83 and 85, respectively).

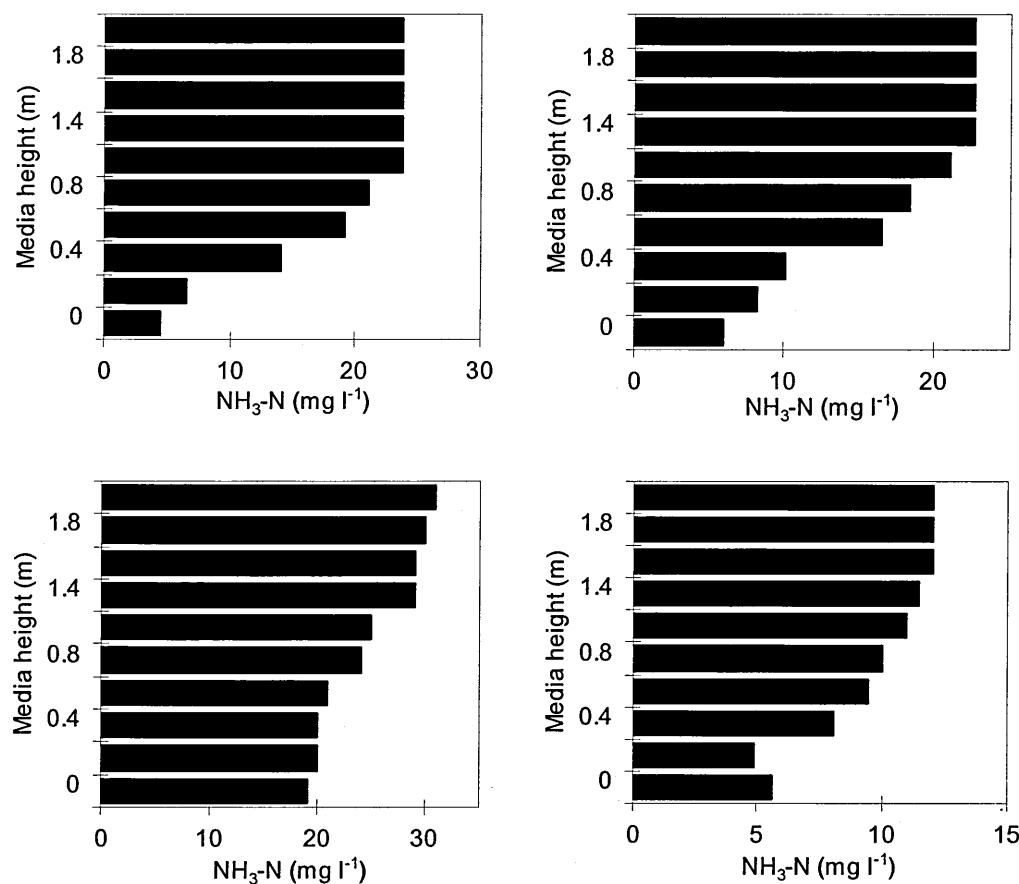


Figure 5.12. Profiles for the removal of ammonia in the downflow reactor (running days 48, 52, 83 and 91, respectively).

It can be seen that in upflow conditions, the greater removal of ammonia is undertaken at the top of the reactor while in the downflow reactor, nitrification occurred at the bottom of the system. These findings suggest that autotrophs were located exactly at those zones.

This pattern can be explained by the fact that, regardless of the flow direction, the backwashing of most sunken media BAFs is carried out by introducing air and water at high velocities to the bottom of the reactors (Jimenez *et al.*, 1987; Pujol *et al.*, 1992). In downflow reactors this action would promote the even distribution of autotrophs along the column. Moreover, according to Faup *et al.* (1982) and Newbigging *et al.* (1995), downflow systems reduce intermixing which subsequently reduce competition

between autotrophs and heterotrophs micro-organisms thus, allowing for more efficient BOD and ammonia removal single-step reactors. In upflow units however, the mixing of the autotrophic and heterotrophic populations would cause a reduction in the treatment efficiency (Wilderer *et al.*, 1995). Due to the co-current action of the operation and backwashing flow, autotrophic bacteria would be prone to be washed out. Furthermore, according to Akkuna *et al.* (1994), during a backwash in downflow reactors the heterotrophic bacteria, which are slightly dense, move to the top of the reactor. At the same time, the autotrophic bacteria, being more dense, move to the bottom of the reactor, promoting the even distribution of bacteria populations along the system.

Had the running conditions been set identically for both columns, the difference in performance between the two reactors in this experiment may have been entirely due to the impact of the backwashing on the biological population as explained earlier. In order to limit the effect of backwashing on the performance of the upflow BAF, some systems use a counter-current backwash (Andersen *et al.*, 1995). A backwashing cycle would consist of applying air scour for a certain time and the backwashing water would be introduced at the top. The backwashing water would simply flow by gravity to the bottom of the filter. However, because there is no pressurised water to wash the filter, a large volume of backwashing water should be available over the filter bed in order to allow for enough pressure head along the bed and avoid blockage.

5.4 Conclusions

- Loading rates to both reactors were around $1.8 \text{ kg SS m}^{-3} \text{ d}^{-1}$, $2.4 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$, $5.2 \text{ kg tCOD m}^{-3} \text{ d}^{-1}$ and $0.51 \text{ kg NH}_3\text{-N m}^{-3} \text{ d}^{-1}$ (based on a working volume in the media section of 25 l in each reactor).
- Removal rates for of SS, soluble and total COD for both reactors were very similar. The upflow reactor achieved average removal rates of $1.3 \text{ kg SS m}^{-3} \text{ d}^{-1}$, $1.9 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$ and $3.9 \text{ kg tCOD m}^{-3} \text{ d}^{-1}$. The downflow reactor reached average removal rates of $1.4 \text{ kg SS m}^{-3} \text{ d}^{-1}$, $1.6 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$ and $4.2 \text{ kg tCOD m}^{-3} \text{ d}^{-1}$ (based on a working volume in the media section of 25 l in each reactor)
- Steady-state was not reached in terms of ammonia removal. The average percent of removal of ammonia in the downflow reactor was 42%, whilst the upflow unit reached only 29%, corresponding to nitrification rates of 0.15 and 0.19 $\text{kg NH}_3\text{-N m}^{-3} \text{ d}^{-1}$ in the upflow and downflow reactors, respectively.
- Autotrophic activity was found at the bottom of the reactor in the downflow column and at the top in the upflow unit due to plug-flow conditions of the reactor and autotrophs-heterotrophs competition. Co-current air and water backwashing could have washed-out autotrophs in the upflow BAF causing the decrease in the removal of ammonia.

Chapter 6

THE USE OF A PROCESS MODEL FOR EVALUATING
THE PERFORMANCE OF BIOLOGICAL AERATED
FILTERS

THE USE OF A PROCESS MODEL FOR EVALUATING THE PERFORMANCE OF BIOLOGICAL AERATED FILTERS

ABSTRACT

Biological aerated filters are three-phase reactors for wastewater treatment that use granular media for biomass support. The verification of a simple empirical model for soluble COD removal for a different type of media than that previously reported was conducted. The media constant (n) was found to have a value of 0.97 and the overall process constant of the system (k^*) was 46 suggesting that overall, the performance of a BAF is greatly influenced the media/liquid flow pattern. Ideally, upflow reactors should operate with floating media and downflow reactors with sunken media.

6.1 Introduction

Models for biological fixed-film systems can be based on mechanistic phenomena (Jennings *et al.*, 1976; Hamoda, 1989) or on empirical relationships (Rusten, 1984; Peng *et al.*, 1995). Process models need to rely on empirical relationships (Germain, 1966), correlating substrate removal to parameters such as filter depth, flowrate and contact time. Thus, process models simply relate input and output variables of a process to one another and, by doing so, the assumptions of uniformity of biomass distribution and ideal plug flow do not need to be made.

Disagreement exists concerning the most representative substrate to use for modelling. Leglise *et al.* (1980) used total organic loading as the design criterion to size the tank volume of BAFs. Moreover, Rusten (1984) found that organic removal rate vs. organic loading rate of a submerged biofilm reactor followed a Monod type equation for an influent total chemical oxygen demand (tCOD) concentration of 129-579 mg l⁻¹. He stipulated that filter design should be based on total organic loadings since it appears to give much better correlation between rates and loads than a design based on soluble organic matter. However, according to Stensel *et al.* (1988), since different wastewaters have different fractions of soluble biochemical oxygen demand (sBOD) and soluble chemical oxygen demand (sCOD), the use of total organic loading may be too simplistic. Therefore, the impact of soluble organic loading must be considered for BAF design. The authors found that treatment performance of the system was better correlated with sCOD volumetric loading rates.

Further disagreement is found about the suitability of using biochemical oxygen demand (BOD) for kinetic studies. Dillon and Thomas (1990) obtained a linear relationship between BOD removal rates and BOD loading rates in BAFs. However, Paffoni *et al.* (1990) and Jaques (1992) could not find any linear relationship between effluent BOD and BOD loading rate for BAFs. Recently Mann and Stephenson (1997) have produced an empirical model by relating the influent soluble chemical oxygen

demand (sCOD) to the effluent sCOD and the reactor height. With the use of simple equations, two constants are found, the media constant (n) which indicates the stability of the process, and the overall process constant (k^*) that gives an approximation of the performance of the reactor in removing sCOD.

The objective of the present work was to verify the suitability of the simple empirical model developed by Mann and Stephenson (1997) for a different type of media by obtaining the media constant (n) and the overall process constant (k^*) of the system.

6.2 Materials and Methods

Experiments were undertaken in a small pilot-scale PVC column, 2.1 m in height and 0.2 m diameter, with a total bed height of 2 m running in upflow mode. Sampling ports were located at 0.2, 0.4, 0.6, 0.8, 1.2, 1.4, 1.6 and 1.8 m from the bottom of the reactor (see Chapter 4). The diameter of the reactor was at least 50 mean particle diameter in size to limit wall effects (Lang *et al.*, 1993).

The reactor contained 38 l of quartz sand spherical media, with a diameter of 2-3.2 mm, a specific surface area of $1500 \text{ m}^2 \text{ m}^{-3}$ and 40% voidage. Air was supplied at an air:water ratio of 1:10. Primary settled sewage from Cranfield University's sewage treatment works was used as a feed. The reactor was backwashed twice a day (for backwashing procedure, see Chapter 4).

Start-up of the reactor consisted in applying the process liquid at the nominal process flowrate ($0.29 - 0.43 \text{ m}^3 \text{ d}^{-1}$) until steady-state was reached. Variation in the nominal flowrate was mostly due to limitations in the accuracy of the valves employed (stop cock and bulba valve, Durapipe S&LP, Staffordshire, UK) and not due to step increases that would cause non steady-state conditions within the reactor. In order to monitor reactor performance, sampling of the influent and effluent of the reactor was undertaken every three days approximately for the whole experimental period. Once

steady-state was reached, soluble chemical oxygen demand (sCOD) profiles of the reactors were taken at staggered time intervals from the influent, sampling ports and effluent.

Suspended solids were analysed according to standard methods (Greenberg *et al.*, 1992). Soluble COD and total COD (tCOD) were analysed using the HACH test-n-tube methods (Camlab, Nuffield, UK). Dissolved oxygen was measured using a Jenway DO₂-9071 meter (Merck Ltd., Leicester, UK), and pH and temperature were measured with a Merck 3071 pH meter (Merck Ltd., Leicester, UK).

The model used in the present study was developed by Mann and Stephenson (1997) and was modified from a trickling filter model (Velz, 1948, Eckenfelder *et al.*, 1961). Despite the fact that BAFs operate with submerged media, they share common characteristics with trickling filters;; in particular, both operate under plug-flow conditions (Sinkoff *et al.*, 1959; Mann *et al.*, 1995).

6.2.1 Model description

The reaction rate for substrate utilisation (i.e. soluble COD) in a biofilm can be described as first order

$$\frac{dC}{dt} = kC \quad (6.1)$$

where C is the soluble COD concentration and k is the rate constant at a certain time and sCOD concentration. By integration

$$\frac{C}{C_i} = e^{-kt} \quad (6.2)$$

where C_i is the influent sCOD concentration (mg l⁻¹)

Time (t) can be related to the volumetric loading by means of a constant

$$t = \frac{k'}{B_V} \quad (6.3)$$

where k' is the biomass constant which depends on the hydraulic characteristics of the reactor and B_V is the volumetric loading which depends on the flowrate (Q , m³ d⁻¹), the initial concentration of sCOD (C_i , mg l⁻¹), the height of the reactor (H , m) and the cross sectional area of the reactor (A , m²)

$$B_V = \frac{QC_i}{HA} \quad (6.4)$$

Hence equation 6.3 can be rearranged as

$$t = \frac{k' HA}{QC_i} \quad (6.5)$$

As the reactor runs under plug-flow conditions, the time (t) and height of the reactor (H) are directly related. Therefore, by substituting t with H and k with m in equation 6.2 we have

$$\frac{C}{C_i} = e^{-mH} \quad (6.6)$$

where

$$m = \frac{k^* A}{QC_i^n} \quad (6.7)$$

Substituting 6.6 into 6.7

$$\frac{C}{C_i} = e^{\frac{k^* A}{Q C_i^n} H} \quad (6.8)$$

By applying natural logarithms to 6.8

$$\ln\left(\frac{C}{C_i}\right) = \frac{-k^* A}{Q C_i^n} H \quad (6.9)$$

By plotting the height of the reactor against the natural logarithm of (C/C_i) the slope (m) for each run can be obtained for different values of C_i .

Equation 6.7 can be rearranged into

$$\frac{Qm}{A} = \frac{k^*}{C_i^n} \quad (6.10)$$

and rearranging again,

$$\ln\left(\frac{Qm}{A}\right) = -n \ln C_i + \ln k^* \quad (6.11)$$

Hence, from Equation 6.11

$y = \ln\left(\frac{Qm}{A}\right)$ where m is the values of the slope obtained from equation 10

$m = -n$ (the media constant)

$x = \ln C_i$ (C_i of the several runs)

$c = \ln k^*$ (the process constant)

and by plotting x and y , the constants n and k^* were obtained.

6.3 Results and Discussion

The upflow biological aerated filters were kept running for approximately 60 days (20 days of start-up and 40 days of steady-state) (Figures 6.1 and 6.2). Temperature ranged between 15 and 18°C and pH decreased from pH 7.8 at the influent to pH 7.0 at the outlet. The concentration of dissolved oxygen along the reactor was kept above 4 mg l⁻¹. Primary settled sewage was applied to the reactors at different hydraulic flowrates that ranged between 0.29 and 0.43 m³ d⁻¹, that corresponded to theoretical hydraulic retention times (tHRT's) between 125 and 84 min, respectively. The organic loading rates applied were between 1.1 and 2.5 kg sCOD m⁻³ d⁻¹. Theoretical hydraulic retention times and loading rates calculations are based on a working volume of 25 l (see Chapter 5).

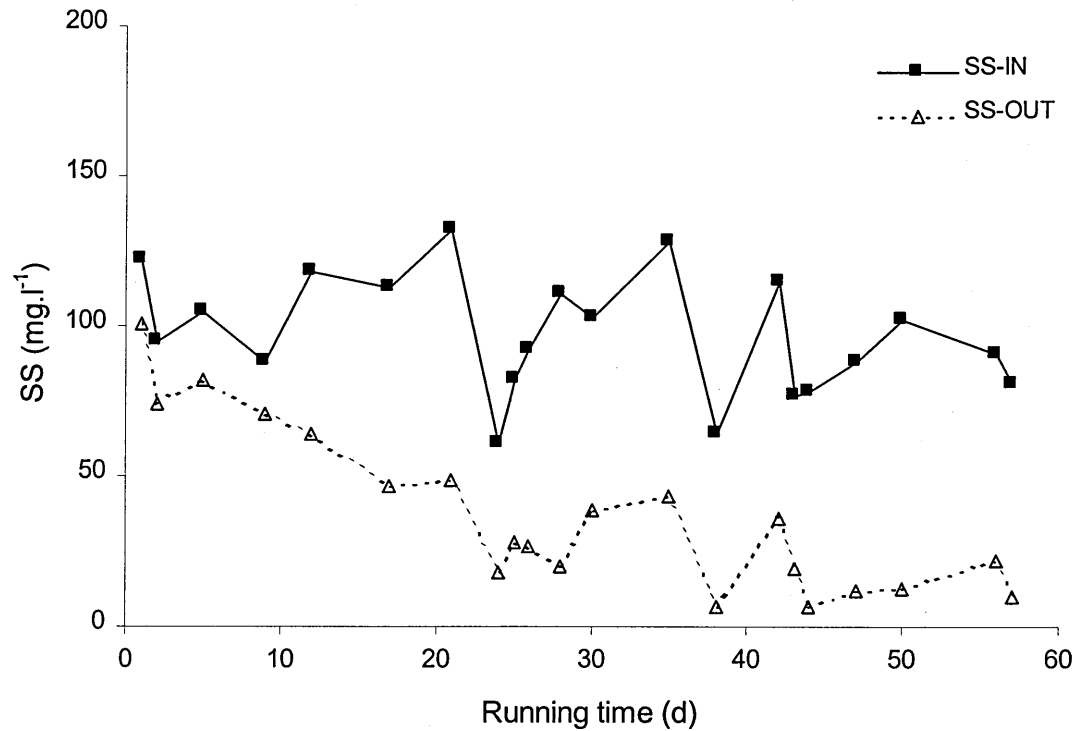


Figure 6.1. Influent and effluent concentration of suspended solids in the BAF.

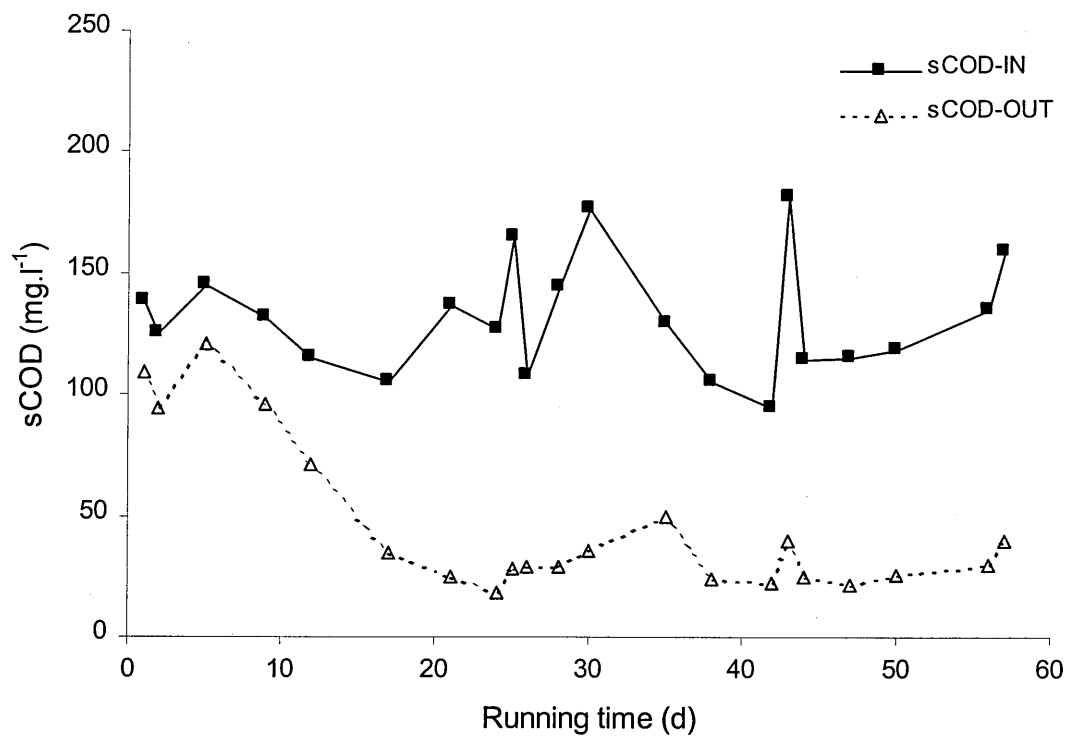


Figure 6.2. Influent and effluent concentration of soluble COD in the BAF.

Start-up took approximately 20 days and during steady-state the average removal efficiencies reached were 72% for SS and 78% for soluble COD (Table 6.1).

Table 6.1. Average and range of influent and effluent concentration of SS and sCOD for the BAF under steady-state conditions (40 days, 16 samples).

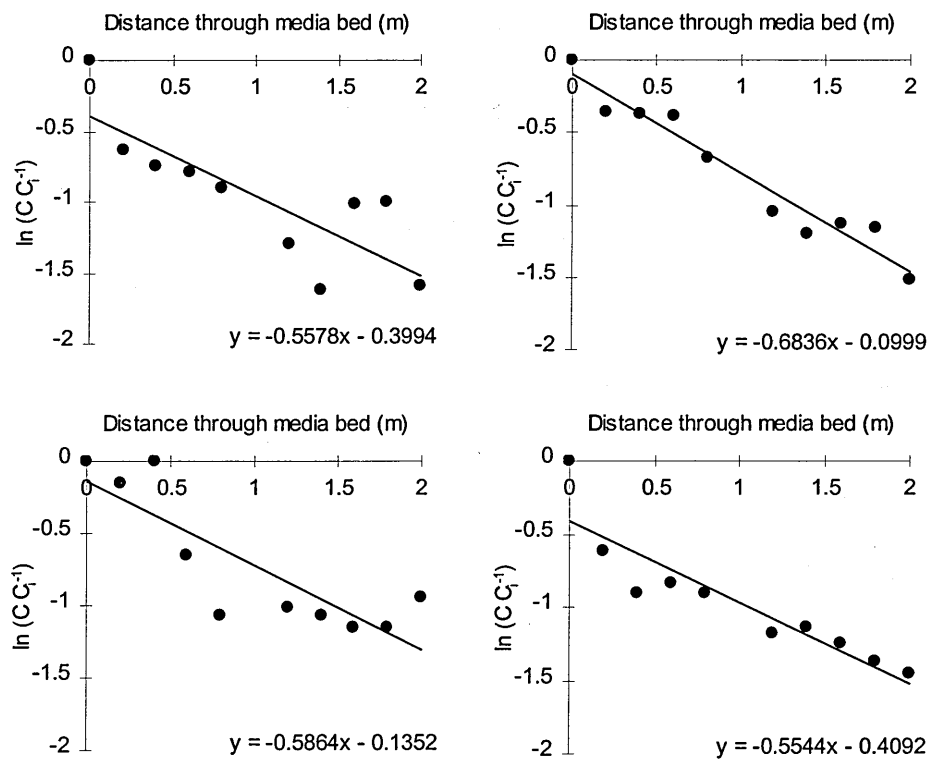
	Influent (mg l ⁻¹)		Effluent (mg l ⁻¹)		Removal (%)	
	Average	Range	Average	Range	Average	Range
Suspended Solids	94	65-132	23	7-49	72	59-94
Soluble COD	134	94-182	30	18-50	78	61-86

In terms of organic loading and removal rates, those reached in the present experiment (Table 6.2) compare well to those obtained elsewhere in full-scale BAFS treating domestic wastewater (Wheale and Cooper-Smith, 1995).

Table 6.2. Loading and removal rates for SS and sCOD to the BAFs during steady-state based on a working volume of 25 l). Average rate (range of rates).

	Loading rate (kg m ⁻³ d ⁻¹)	Removal rate (kg m ⁻³ d ⁻¹)
Suspended Solids	1.2 (0.7-1.9)	0.8 (0.4-1.5)
Soluble COD	1.7 (1.1-2.5)	1.3 (0.8-2.0)

A series of eight logarithmic profiles for soluble COD removal (Figure 6.3) were used to calculate the values of the slope (*m*) at the different organic loadings (*C_i*) and volumetric flowrates (*Q*).



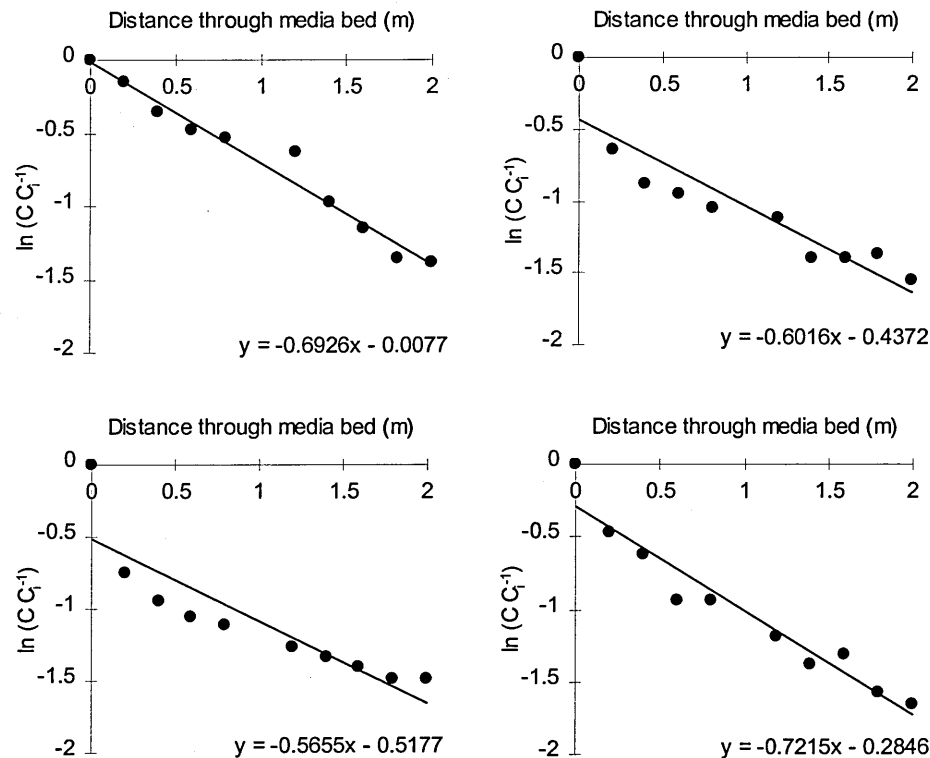


Figure 6.3. Logarithmic sCOD removal profiles.

By plotting the logarithmic value of C_i against the relationship between flowrate (Q), slope (m) and cross-sectional area (A), the constants k^* and n were obtained (Figure 6.4).

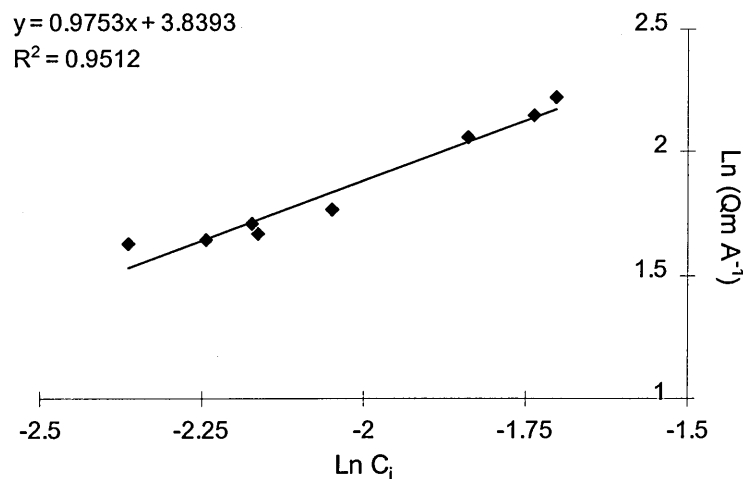


Figure 6.4. Calculation of the process constant (k^*) and the media constant (n) for the BAF.

The new constants are presented in Table 6.3.

Table 6.3. Process constants (k^*) and media constants (n) for the sunken media reactor operated under upflow conditions and those obtained elsewhere.

Type of reactor	k^*	n
Sunken media, upflow reactor †	33	0.92
Floating media, upflow reactor †	55	1.13
Sunken media, upflow reactor ‡	46	0.97

† Mann and Stephenson (1997)

‡ This study

Temperature, pH, dissolved oxygen concentration and organic loadings in Mann and Stephenson (1997) and this investigation were almost identical. Moreover, in both investigations plug-flow conditions prevailed (Chapter 4; Mann *et al.*, 1995) and backwashing frequencies were the same. As mentioned earlier, k^* indicates the overall performance of the reactor in removing sCOD and hence, it is a rate of reaction. High values of k^* indicate high sCOD removal and vice versa. The rate of reaction can be affected by temperature, pH, dissolved oxygen concentration, biomass, organic loadings and the hydraulic characteristics of the system. The value of k^* found for the new media shows that it performed better than the sunken media used in previous work. However, it was outperformed by the floating media investigated previously. The difference in the values of k^* between both sunken media reactors could be due to the biomass characteristics and concentration in the system.

The media constant (n) indicated the stability of the process in other words, the variation of sCOD removal rates over a range of sCOD concentrations. Larger values of n would indicate greater stability of the system at high loadings, whilst smaller values would show a greater variation of sCOD over a small range of influent sCOD concentrations. Again, the hydraulic conditions of the systems in the two investigations were almost identical. It appeared that the floating media reactor used previously was more resistant to shock loadings than the reactor with new media. The media constant would vary depending on size, surface texture, surface area, density (floating or sunken

media) and the hydraulics of the system. It has been demonstrated that spherical media was less efficient than angular media for biofilm formation probably due to the higher surface area available and the better oxygen transfer achieved with angular media (Sagberg and Berg, 1996). The media used in the present experiment was spherical and this would have affected biofilm formation.

Clearly, biomass was the responsible for difference in k^* values, so it can be assumed that the properties of the new media were not optimum for biomass establishment. However, although shape, roughness, size and density of the media affect biofilm formation and detachment, flow direction and aeration rates also influence biofilm growth by affecting the compactness of the bed and the liquid flow through it (Cao and Alerts, 1995). From the comparison of n and k^* values, it was evident that the media used for this experiment performed better than the sunken used previously during upflow configuration though was less efficient than the floating media (Mann and Stephenson, 1997). Thus, it can be assumed that the media used in the present study was more suitable for biofilm formation than the plastic media used previously. The flow pattern also dictated the overall performance of the system. The floating media in upflow configuration was more efficient in removing COD than the same configuration with sunken media. It appeared that the compression of the bed due to the buoyancy force of the media and the flow of air and water acting upwards improved solids retention and promoted the growth of bacteria. This could be due to the natural bed compression caused by the hydraulic flowrate (Mann *et al.*, 1998). This simple model, unlike mechanistic models, does not require the measurement of a great number of variables though its application should be limited to obtaining a fast prediction of soluble COD removal in individual plants.

6.4 Conclusion

- A simple empirical model was used to obtain the media constant ($n = 0.97$) and the overall performance constant ($k^* = 46$) of a small pilot-scale BAF running under upflow conditions.
- By maintaining constant conditions, the results were directly compared to other investigations, revealing which process condition was responsible for the overall reactor performance.
- It was found that the media type/flow pattern was the main condition affecting the overall performance of the system. Therefore, the best BAF configuration would be upflow reactors with floating media and downflow reactors with sunken media.

Chapter 7

PROCESS MODELS FOR THE REMOVAL OF
SUSPENDED SOLIDS AND ORGANIC MATTER, AND
THE SCALE-UP OF BIOLOGICAL AERATED FILTERS

PROCESS MODELS FOR THE REMOVAL OF SUSPENDED SOLIDS AND ORGANIC MATTER, AND THE SCALE-UP OF BIOLOGICAL AERATED FILTERS.

ABSTRACT

The constants from 3 empirical models for soluble COD (sCOD) removal were used to predict its removal in a small BAF (0.07 m³ EBV) and a large BAF (0.69 m³ EBV). From the three models tested, the one including parameters such as gas and liquid velocity provided close to the actual observed values, when the corresponding constants were used for verifying the models. All three models were successful in predicting overall performance of the reactor based on organic loading rate and when predicting the profile removal of sCOD. A model previously used for rapid gravity filters was used for the prediction of SS and tCOD in a small and a large BAF. The model was efficient in predicting the overall removal rates of SS and tCOD in both reactors. The model helped to prove that by reducing the height of the large BAF from 3.9 m to 2.9 m the efficiency of the reactor would not be adversely affected. For scaling-up purposes, model 1 was the most effective for predicting the removal of sCOD in the large pilot-scale BAF when using constants from the small pilot-scale BAF. The lack of kinematic similarity (differences in the organic and hydraulic rates) between reactors seem to have limited the capacity for scaling-up of models 2, 3 and 4 since these models did take into account parameters like sCOD influent concentration, air and gas velocity.

7.1 Introduction

Scaling up is one of the most frequent problems in the design of water and wastewater treatment systems. Systems of different scale can be related by the so-called theory of similarity (Horvath and Schmidtke, 1983) which consists of six types:

Geometric similarity exists when there is a correspondent point from one body (or system) to the other; the distance between any two points on the model must bear a constant ratio to the distance between the corresponding two points in the full-scale system.

Kinematic similarity is reached when the velocity at a point in the model present a constant ratio to the velocity at the corresponding point in the other system. Since fluids and solids are in motion, in geometrically similar systems particles need to have corresponding velocities.

Dynamic similarity occurs when forces of the same kind act upon corresponding particles at corresponding times in geometrically similar systems; it is particularly important in power consumption or pressure loss prediction.

Chemical, biological and/or thermal are the fourth, fifth and sixth types similarities. They exist when there is a fixed relation between the concentration, the biological population and/or temperature at corresponding points between the systems, respectively.

Although a total similarity between different scale systems cannot be reached, approximate modelling (or partial modelling) can yield acceptable results especially when the modelling problem is broken into several components (Schmidtke and Horvath, 1977; Horvath and Schmidtke, 1983). In order to establish similitude criteria, it is also necessary to establish initial and boundary conditions.

Boundary conditions are particularly important to establish in reactor design as the wall effect can modify the hydraulic conditions of the system. Moreover, the physical, chemical and biological conditions of the system are also affected. It has been implied that wall effects are more important in smaller units and that they should be taken into account (Coulson and Richardson, 1978). Particles in a packed bed do not pack as closely in the region near the wall as they do in the centre of the bed. Therefore the actual resistance to flow in a bed with a small diameter is less than it would be in an infinite container for the same flowrate per unit area of bed cross section. However, it has been stated that wall effects are not a problem when the ratio of the particle diameter to the diameter of the column is >20 (Manem and Rittmann, 1989), or >50 (Lang *et al.*, 1993).

According to Eckenfelder and Quirk (1983), in order to scale-up results from pilot-scale reactors to full-scale systems, four aspects should remain constant:

- hydraulic loading
- film thickness
- surface area of packing medium
- hydraulic characteristics of the system

Moreover, according to Manem and Rittmann (1989) the scaling procedure for biofilm reactors should be based on the similitude principle and the following conditions should apply:

- the same type of reactions and microorganisms must be occurring in both sizes of reactor
- the same biofilm configuration must be used for both processes (e. g. it is not possible to scale-up from an RBC to a trickling filter)
- the same effluent concentration of the rate limiting substrate must be obtained from both processes
- the temperature in both processes must be the same

Surface concentration of substrate, biofilm shear loss and mass balance are the three criteria that define similitude for scaling up. In terms of biofilm theory, the kinetic parameters that can change between small and large-scale fixed-film systems are the depth of the effective diffusion layer, the density of microorganisms within the biofilm and the thickness of the biofilm.

Models of the performance of BAFs have been very limited. Although mechanistic models have been developed, they tend to be valid under limited conditions and stable over only a small range (Hamoda, 1989). On the other hand, empirical models simply relate the influent and effluent concentrations of the parameter to be analysed. The application of these models for the design of biofilm systems has more or less worked for the removal of BOD, SS and $\text{NH}_3\text{-N}$ yet, according to Arvin and Harremoës (1989), it could cause errors when used for scale-up. Recently, attempts to develop semi-empirical models for submerged filters have been undertaken. Le Tallec *et al.* (1997) proposed a simple model based on the International Association on Water Quality Activated Sludge Model (Henze *et al.*, 1987) for a nitrification/denitrification filter. Mann and Stephenson (1997) developed an empirical model for BAFs that related the influent soluble chemical oxygen demand (sCOD) to the effluent sCOD and reactor height. However, these simple models should only be used as a tool for obtaining a fast prediction of soluble COD removal in individual plants. Jacob *et al.* (1997) developed a more complex dynamic model for BAFs.

Therefore, the present experiment had the objective of testing two previously developed models for soluble chemical oxygen demand (sCOD) removal and a modified version of a model that included the input of parameters such as liquid and gas velocity. Due to the unsuitability of these models to predict suspended solids (SS) and total chemical oxygen demand (tCOD) removal, a fourth model based on previous work on rapid gravity filters was used for this purpose. The models were tested in BAFs of two sizes and the suitability of the models for scale-up purposes was also tested.

7.2 Materials and methods

Experiments were undertaken in two pilot-scale reactors of different dimensions, both running in upflow mode. The small pilot-scale BAF had a height of 2.1 m and 0.2 m diameter, and a total media bed height of 2 m (Table 7.1). Please refer to Chapter 3 for further description. The large pilot-scale BAF was 0.5 m wide and 3.9 m high, with a total media height of 3 m (Figure 7.1). Primary settled sewage was fed to the large pilot-scale BAF with the use of a Watson-Marlow 700 series peristaltic pump with a capacity $0\text{--}100\text{ l min}^{-1}$ (Watson-Marlow Ltd, Cornwall, UK) at an average volumetric loading of 6 l min^{-1} . Backwashing water was supplied by means of a Grundfos peristaltic pump (model A006422) of $60\text{ m}^3\text{ h}^{-1}$ capacity (Grundfos Pumps Ltd, Leighton Buzzard, UK). Air and liquid flow to the large BAF were monitored by means of a Platon plastic tube VA flowmeter with a $40\text{--}440\text{ l min}^{-1}$ range and a Platon plastic tube VA flowmeter with a $300\text{--}3000\text{ l h}^{-1}$ range, respectively (CT Platon Ltd, Basingstoke, UK). Bulb valves (Durapipe S&LP, Staffordshire, UK) were used to control both hydraulic and gas flowrates. An automatic motorised valve (Ultravalve Ltd, Halsowen, UK) was used to prevent backwash water from syphoning back through the effluent pipe. Sampling ports were located at 0.6, 1.2, 1.7, 2.3 and 2.9 m from the bottom of the reactor (Table 7.1). The media in both reactors consisted of quartz spherical sand, with a diameter of 2–3 mm, a specific surface area of $1500\text{ m}^2\text{ m}^{-3}$, density of 2500 kg m^{-3} and 40% voidage.

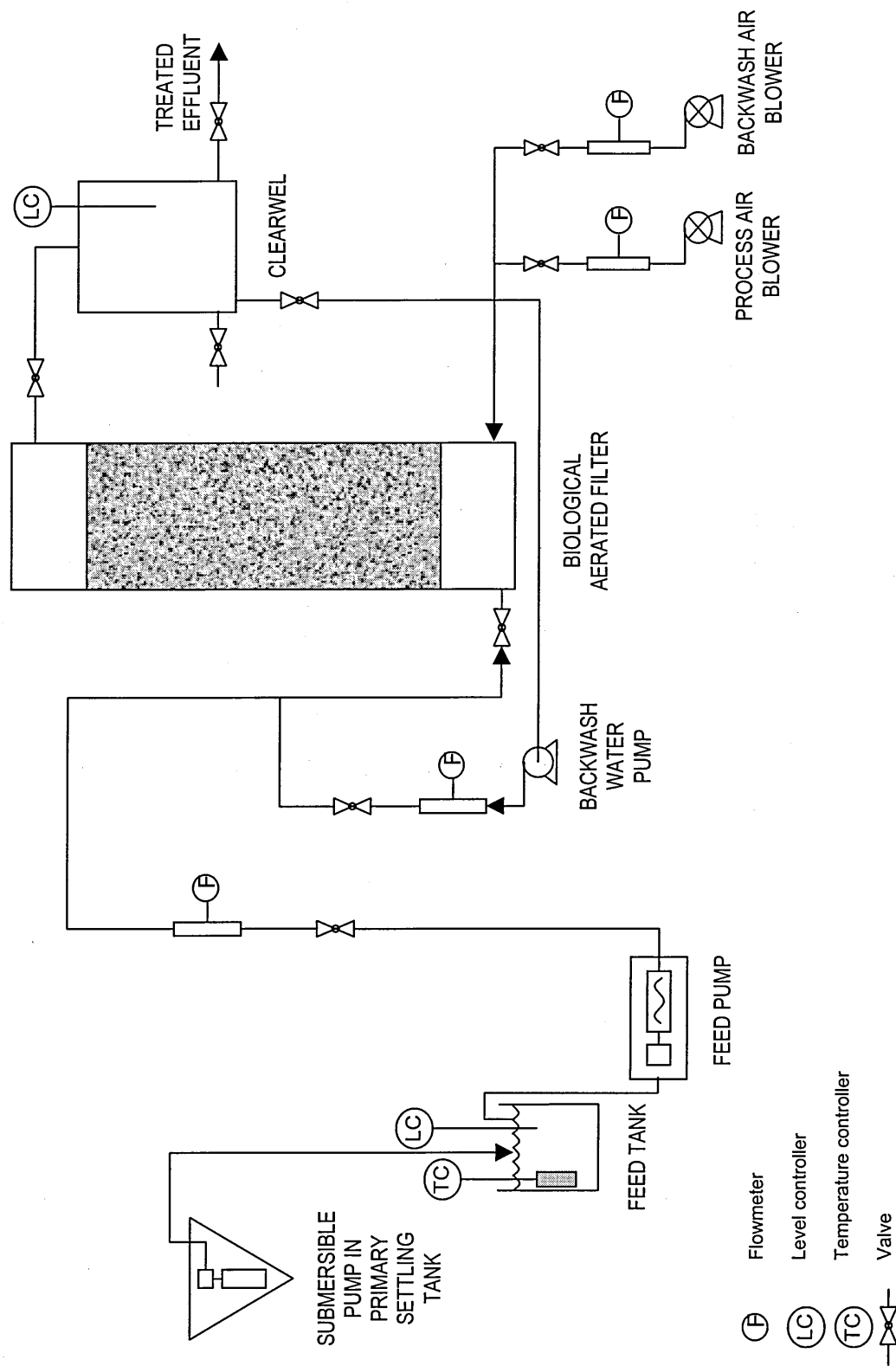


Figure 7.1. Schematic diagram of the large pilot-scale BAF running in upflow mode.

Table 7.1. Dimensions and operating conditions of the small and large pilot-scale BAFs

	Small pilot-scale BAF	Large pilot-scale BAF
Column diameter (m)	0.2	0.5
Column height (m)	2.1	3.9
Media bed height (m)	2	3
Empty bed volume - EBV (m ³)	0.07	0.69
Media volume (m ³)	0.04	0.39
Working volume (m ³)	0.03	0.3
Media diameter (m)	2-3 x 10 ⁻³	2-3 x 10 ⁻³
Media voidage	0.4	0.4
Theoretical HRT* (h)	1.1 - 1.5	0.8
Hydraulic flowrate (l h ⁻¹)	21 - 27	360
Hydraulic velocity (m h ⁻¹)	0.7 - 0.9	1.8

*based on working volumes

† based on empty bed volumes (EBV)

Backwashing in the large pilot-scale reactor was undertaken automatically by means of a timer. The backwash cycle consisted of the following sequence (adapted from the Tetra Colox Pilot Column Process Operation Manual)

1. feed pump off, the outlet valve closes
2. backwash air comes on; runs for five min
3. backwash water on; air and water run simultaneously for 15 min
4. backwash air stops; backwash water still runs for further 5 min (or is tripped off by low level in clearwell)
5. backwash water off
6. two minute delay
7. feed pump comes on; normal operation is resumed

Primary settled sewage from Cranfield University’s sewage treatment works was used as a feed. Initial trails with both reactors demonstrated that, in order to avoid clogging, the small reactor should be backwashed once a day and the large reactor three times a

day. In both reactors, start-up was undertaken by running primary settled sewage at the nominal flowrate until steady-state was reached. Sampling of the influent and effluent of the reactors was undertaken approximately every 3 days to assess their removal performance. Temperature was checked daily by means of a mercury thermometer (-5 to 100 °C) range (Camlab, Nuffield, UK). Once steady-state was reached, suspended solids (SS), soluble chemical oxygen demand (sCOD) and total chemical oxygen demand (tCOD) profiles of the reactors were taken at staggered time intervals from the influent pipe, sampling ports and effluent pipe. Suspended solids were analysed using standard methods (Greenberg *et al.*, 1992); soluble COD and total COD were analysed using the HACH test-n-tube methods (Camlab, Nuffield, UK).

7.2.1 Models for soluble chemical oxygen demand (sCOD)

Three empirical models for sCOD were tested

Model 1 was initially developed for trickling filters design (Eckenfelder, 1961) and has been also used successfully on a biofilter with aeration in an external tank (Chung *et al.*, 1996). The model consists of the following expression

$$\frac{C_{out}}{C_{in}} = \exp \left[\frac{-kH}{\left(\frac{Q}{A}\right)^n} \right] \quad (7.1)$$

where

C_{in} = concentration of sCOD at influent [g m^{-3}]

C_{out} = concentration of sCOD at the specific point of the reactor corresponding to the height [g m^{-3}]

Q = influent flowrate [$\text{m}^3 \text{d}^{-1}$]

A = cross sectional area of biofilter (A) [m^2]

H = effective depth of the column [m]

n = a parameter dependent on the specific media used

k = a pseudo rate constant [m^{-1}]

According to James (1984), in order to develop a model, it is necessary to look for the various parameters that affect the system. Hence, Model 2 was based on work by Eckenfelder (1961) but three constants, incorporating a factor that depended on the influent concentration of COD, instead of two are obtained by the following expression (Mann and Stephenson, 1997):

$$\frac{C_{out}}{C_{in}} = \exp \left[\frac{-kH}{\left(\frac{Q}{A}\right)^n C_{in}^m} \right] \quad (7.2)$$

m = the value of the slope depending on the influent concentration of sCOD (C_{in})

n = the media constant

k = the process constant

Model 3 is also based on the model for trickling filters developed by (Eckenfelder, 1961). Model 3 has not been presented elsewhere previously and incorporated a heterogeneous coefficient α that accounts for changes in the diameter of the filter, the effective height of the media bed and the superficial velocity of air and liquid, parameters that affect the performance of BAFs (Chapter 3):

$$\frac{(1 + \alpha)C_{out}}{C_{in}} = \exp \left[-\frac{kH}{\left(\frac{Q}{A}\right)^n} \right] \quad (7.3)$$

As mentioned before, α is a function of the inlet concentration of the parameter to be removed, the effective height of the filter, the superficial velocity of air [UG , m s^{-1}] and the superficial velocity of the liquid [UL , m s^{-1}].

Thus

$$\alpha = \gamma H^z UG^l UL^p \quad (7.4)$$

Where

γ = correcting factor, and

z, l, p = parameter constants for reactor height, gas velocity and liquid velocity, respectively.

Equation 7.3 can be rewritten as

$$\frac{C_{out}}{C_{in}} = \frac{\exp\left[-\frac{kH}{\left(\frac{Q}{A}\right)^n}\right]}{1 + \gamma H^z UG^l UL^p} \quad (7.5)$$

7.2.2 Model for suspended solids (SS) and total chemical oxygen demand (tCOD)

SS and tCOD were calculated using a model for rapid sand filters (Gimbel, 1989).

$$\frac{C_{out}}{C_{in}} = \frac{\exp(C_{in} V_L \lambda_0 \sigma_s^{-1} \tau)}{\exp(C_{in} V_L \lambda_0 \sigma_s^{-1} \tau) + \exp(\lambda_0 H) - 1} \quad (7.6)$$

where

C_{in} = weight percent of SS and/or tCOD in the influent (dimensionless)

C_{out} = weight percent of SS and/or tCOD in the effluent (dimensionless)

V_L = liquid superficial velocity (m s^{-1})

λ_0 = filter coefficient (m^{-1})

σ_s^{-1} = ultimate specific deposit (dimensionless)

H = effective depth of filter (m)

τ = corrected time variable (s), calculated as follows

$$\tau = t - \frac{\varepsilon_0 H}{V_L} \quad (7.7)$$

ε_0 = porosity of media = 0.4

t = filter running time (s)

The constants for all models were calculated using programs written in QBASIC, included in Appendix 1. The calculated constants were then incorporated into the models. Results from the models and the actual observed values of each parameter studied were compared using spreadsheets in Microsoft Excel for Windows version 7.0.

The program also calculated the mean relative error (MRE) for each run as follows

$$MRE = \frac{\sum_0^n \left[\left(\frac{\text{observed data}}{\text{calculated data}} - 1 \right) * 100 \right]}{n} \quad (7.8)$$

Where n is the number of observed (and calculated) points for each run. If $MRE \approx 0$ the difference between the observed and the calculated data would be small hence, the model would be more accurate.

7.2.3 Scaling-up results

Initially the four models were used to obtain the empirical constants with the data from the small pilot-scale plant. In order to test the ability of the models for scaling-up, the constants from the small pilot-scale BAF were used for simulating the removal in the large pilot-scale BAF. The resulting simulated removal profiles were compared to the actual data obtained experimentally and the graphs were analysed to find out which model gave the closest approximation to the experimental data.

7.3 Results and Discussion

7.3.1 Small pilot-scale reactor

7.3.1.1 Overall performance

Suspended solids and sCOD data for the small pilot-scale column has already been presented in Chapter 6. Data for total COD had not been presented previously. Overall performance of the reactor during steady-state conditions is presented below (Table 7.2).

Table 7.2. Average and range of influent and effluent concentrations of SS, sCOD and tCOD for the small pilot-scale BAF under steady-state conditions (16 samples).

	Influent (mg l ⁻¹)		Effluent (mg l ⁻¹)		Removal (%)	
	Average	Range	Average	Range	Average	Range
Suspended Solids	94	65-132	23	7-49	72	59-94
Soluble COD	134	94-182	30	18-50	78	61-86
Total COD	273	212-343	64	32-115	77	68-88

Temperature of the feed water ranged from 13 to 19°C with an average of 16°C. The dissolved oxygen (DO) concentration along the reactor was kept above 4 mg l⁻¹. The small BAF achieved average percentages of removal > 72% for SS, sCOD and tCOD. Loading rates applied and removed are presented below (Table 7.3)

Table 7.3. Loading and removal rates of SS, sCOD and tCOD in the small pilot-scale BAF (based on empty bed volumes – EBV).

	Small pilot-scale BAF	
	Loading rate (kg m ⁻³ d ⁻¹)	Removal rate (kg m ⁻³ d ⁻¹)
Suspended solids	0.4 (0.3-0.7)	0.3 (0.2-0.4)
Soluble COD	0.7 (0.4-1.1)	0.5 (0.3-0.9)
Total COD	1.3 (0.9-2.2)	1.0 (0.7-1.5)

The average removal rates reached in the small pilot-scale BAF were 0.3 kg SS m⁻³ d⁻¹, 0.5 kg sCOD m⁻³ d⁻¹ and 1.0 kg tCOD m⁻³ d⁻¹ which were lower than those reported elsewhere (Rogalla and Bourbigot, 1990; Pujol *et al.*, 1994). Substrate removal was very good (Table 7.2) so there did not seemed to be deficiencies in the treatment. These relatively low removal rates could have been caused by the loading rates applied to the reactor which resulted from the characteristics of the wastewater used and the hydraulic loadings applied.

In order to calculate the constants for the models, a total of eleven profiles were undertaken in the small column (Appendix 2). In order to optimise the results, not all the profiles were used for calculating the model constants. Hence, in the small pilot-scale BAF six of these profiles were used for calculating the constants for sCOD, seven for tCOD and nine for SS removal, respectively. Constants for each model are listed below in Table 7.4.

Table 7.4. Calculated constants for the small pilot-scale BAF.

	Small BAF
Model 1	
<i>k</i>	-1.41 x 10 ⁻²
<i>n</i>	0.467
Model 2	
<i>k</i>	42.5
<i>n</i>	0.037
<i>m</i>	0.839
Model 3	
<i>K</i>	-0.029
<i>n</i>	0.18
<i>γ</i>	1.03 x 10 ⁻⁴
<i>z</i>	0.894
<i>p</i>	-0.803
<i>l</i>	-0.330
Model 4	
λ ₀ (SS)	0.883
σ _s ⁻¹ (SS)	1734.8
λ ₀ (tCOD)	0.84
σ _s ⁻¹ (tCOD)	6 x 10 ⁵

Once the constants were calculated, the models were tested for their suitability at predicting SS and organic matter removal.

7.3.1.2 Soluble COD removal

The models were initially tested for their overall accuracy in predicting the removal of sCOD based on organic loading rates. Of the three models used for sCOD removal, Model 3 incorporating the factor α proved to be give the closest predictions to the actual observed values of sCOD in the small scale BAF (Figure 7.2). The graph was obtained by plotting the observed sCOD loading (from the influent concentration) and the observed sCOD removal (from the effluent concentration) and those rates calculated by the models.

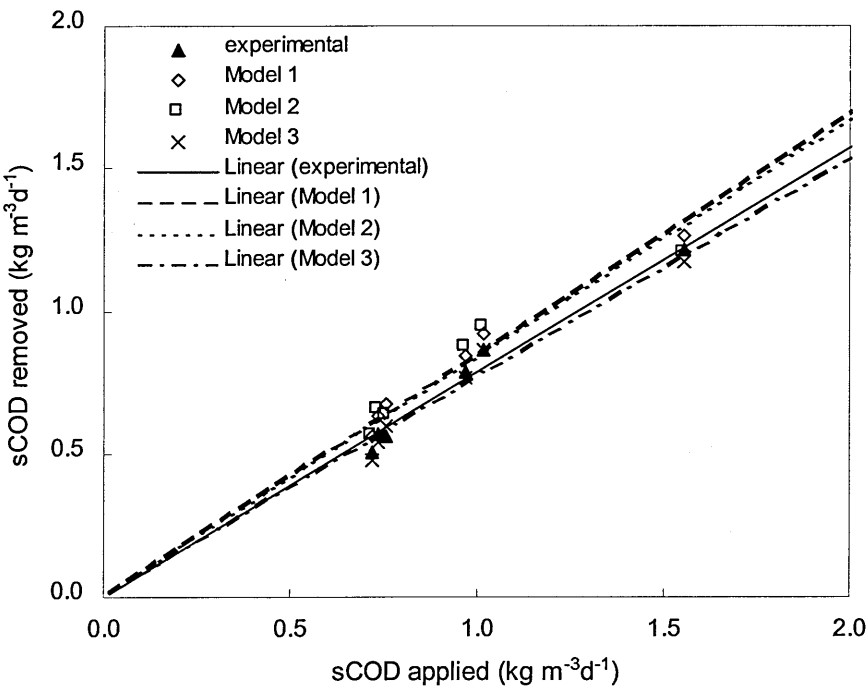


Figure 7.2. Rates of soluble COD applied and removed in the small BAF.

Next, each model was tested to find its suitability for predicting sCOD along the height of the reactor. Results from Model 1, 2 and 3 are presented in Figure 7.3. All three models gave very good approximations to the actual sCOD removal profile but encountered the greatest problems at the bottom section of the reactor (<1 m height), where several observed sCOD values were smaller than those calculated. Model 1 and 2 have almost the same mean relative error (MRE) but Model 3 gave the lowest MRE.

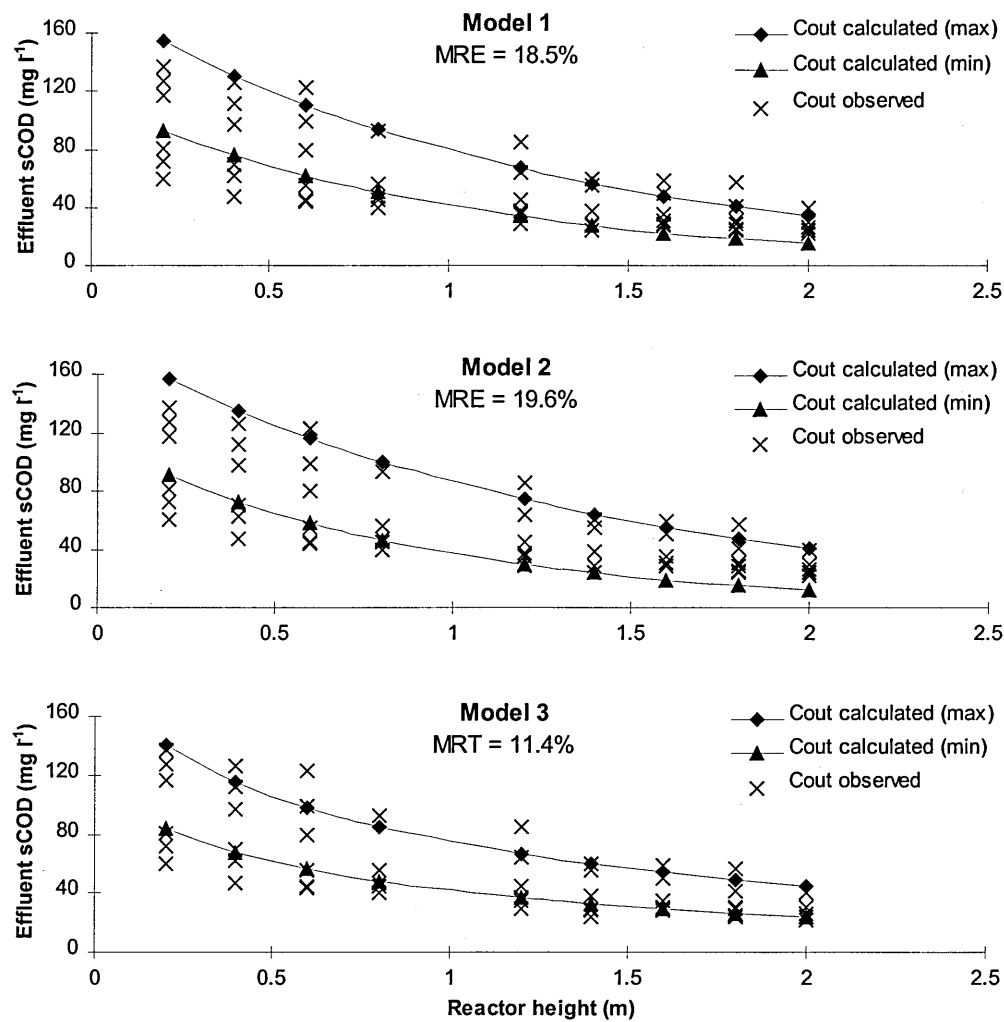


Figure 7.3. Results from six soluble COD removal profiles for Model 1, 2 and 3 undertaken in the small BAF. The lines represent the profiles obtained from the models when applying the maximum and minimum influent concentration.

Model 3 gives the lowest number of observed points outside the ranges predicted which confirms the fact that it was the most efficient in simulating the sCOD removal profile in the small reactor (Figure 7.3).

7.3.1.3 Suspended solids and total COD removal

Model 4 was used for predicting the removal of suspended solids and total COD and seemed very effective in predicting SS removal in the small pilot-scale BAF (Figure 7.4). There is a certain amount of scattering of the data obtained experimentally in the small reactor. The predicted data from model is very close to the experimental, slightly overestimating the real performance of the small BAF.

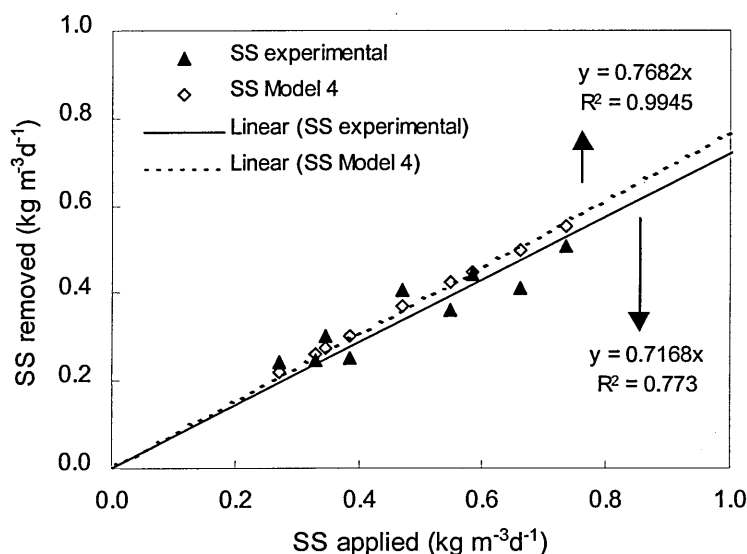


Figure 7.4. Rates of suspended solids applied and removed in the small BAF.

In terms of the tCOD, Model 4 again appeared to overestimate the removal capacity of the small BAF (Figure 7.5); the actual observed rates of tCOD removal were smaller than those predicted by the model.

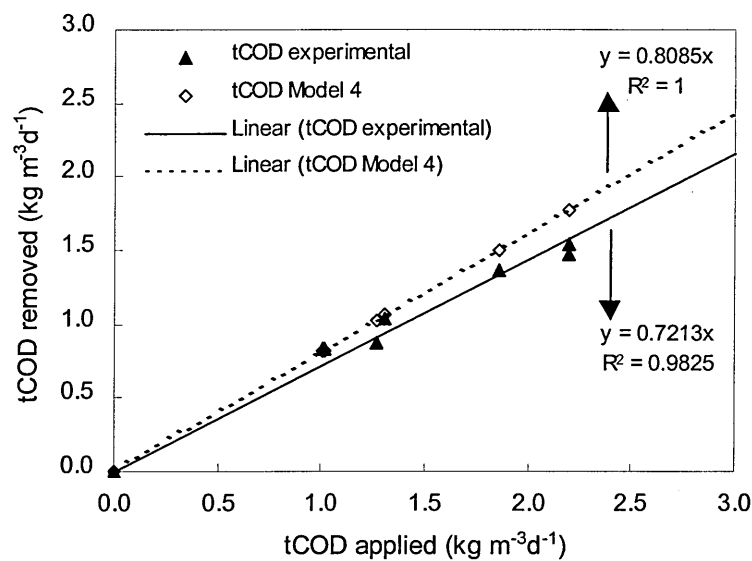


Figure 7.5. Rates of total COD applied and removed in the small BAF.

From the profiles for the removal of SS (Figure 7.6) and tCOD (Figure 7.7), it was clear that although the model is efficient in predicting the removal in the lower section of the reactor ($<1.5\text{ m}$), it is not as efficient in the upper section ($>1.5\text{ m}$).

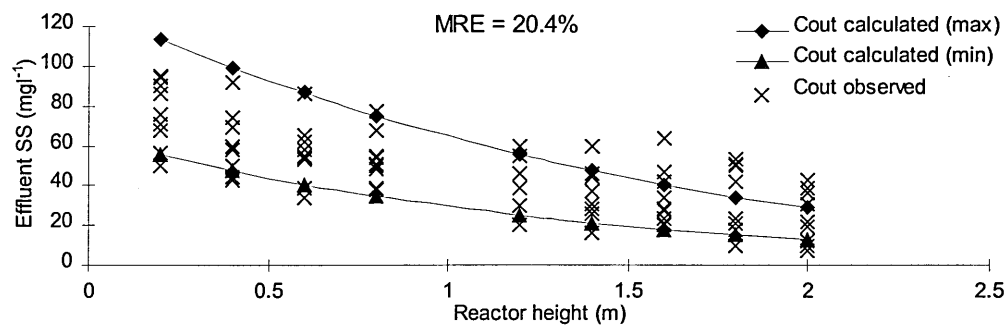


Figure 7.6. Results from nine suspended solids removal profiles for Model 4 undertaken in the small BAF. The lines represent the profiles obtained from the model when applying the maximum and minimum influent concentration.

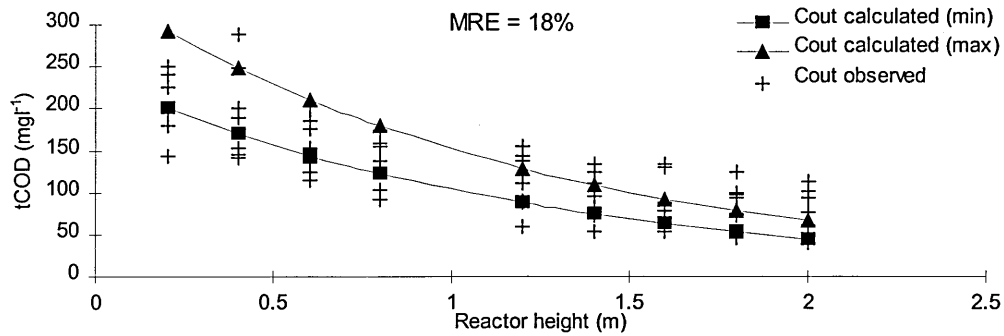


Figure 7.7. Results from seven total COD removal profiles for Model 4 undertaken in the small BAF. The lines represent the profiles obtained from the model when applying the maximum and minimum influent concentration.

This was more evident at the very top of the reactor, where the effluent samples were taken. The model was less accurate in predicting the tCOD concentration in this section, resulting in the relatively high values of MRE.

7.3.2 Large pilot-scale reactor

7.3.2.1 Overall performance

Data for the large pilot-scale BAF was obtained over an experimental period of two months and 30 days under steady-state conditions (Table 7.5).

Table 7.5. Average and range of influent and effluent concentrations of SS, sCOD and tCOD for the large pilot-scale BAF under steady-state conditions (13 samples).

	Influent (mg l ⁻¹)		Effluent (mg l ⁻¹)		Removal (%)	
	Average	Range	Average	Range	Average	Range
Suspended Solids	117	97-137	24	10-44	80	65-90
Soluble COD	134	86-166	42	33-54	68	57-72
Total COD	316	226-368	72	48-108	78	70-85

Although the hydraulic loading along with the film thickness, surface area of packing medium and the hydraulic characteristics of the system should remain constant in order to scale-up results from pilot scale reactors to full-scale systems (Eckenfelder and Quirk, 1983), this was not possible to reach in the present experiments. Due to the characteristics of the mechanical equipment used (i.e. pump capacity, pipe dimensions), the large column could not run at a lower flowrate and the small BAF could not run at a higher flowrate. Hence, the actual hydraulic velocity and, consequently, the suspended solids and organic loading rates applied to the large column (Table 7.5) were roughly twice than those applied to the small column (Table 7.3). In terms of organic removal rate, the large column was more efficient than the small column (Tables 7.3 and 7.6).

Table 7.6. Loading and removal rates of SS, sCOD and tCOD in the large pilot-scale BAFs based on empty bed volumes – EBV.

	Large pilot-scale BAF	
	Loading rate (kg m ⁻³ d ⁻¹)	Removal rate (kg m ⁻³ d ⁻¹)
Suspended solids	1.5 (1.2-1.7)	1.2 (1.0-1.5)
Soluble COD	1.7 (1.1-2.1)	1.1 (0.6-1.5)
Total COD	4.0 (2.8-4.6)	3.0 (2.2-3.4)

Percentages of removal of substrate in the large pilot-scale BAF were satisfactory (Table 7.5) leading to average removal rates of 1.2 kg SS m⁻³ d⁻¹, 1.1 kg sCOD m⁻³ d⁻¹ and 3.0 kg tCOD m⁻³ d⁻¹ which are higher than those reached in the small column (Table 7.3). Thus, the large pilot-scale BAF achieved greater removal rates than those reported by Wheale and Cooper-Smith (1995) in a full-scale downflow system (1.5 kg COD m⁻³ d⁻¹) and by Peng *et al.* (1995) in a full-scale upflow system (2.4 kg COD m⁻³ d⁻¹).

Six profiles were undertaken in the large BAF (Appendix 3) from which data from five profiles were used for calculating the constants for sCOD, six for tCOD and five for SS removal presented in Table 7.7.

Table 7.7. Calculated constants for the large pilot-scale BAF.

	Large BAF
Model 1	
k	-50.4×10^{-2}
n	-0.049
Model 2	
k	13.7
n	-0.81
m	-0.525
Model 3	
K	-0.001
n	0.41
γ	0.59
z	1.256
P	0.050
L	-
Model 4	
λ_0 (SS)	1.772
σ_s^{-1} (SS)	1149.4
λ_0 (tCOD)	0.543
σ_s^{-1} (tCOD)	6322.5

Due to the differences in dimensions between the two reactors, as well as the loading rates and running conditions, the constants obtained for the large reactor were different to those obtained for the small BAF (Table 7.4).

7.3.2.2 Soluble COD removal

Similar results to those obtained in the small BAF are obtained in the large BAF for the prediction of the overall sCOD removal based on influent and effluent concentrations (Figure 7.8). Model 2, 1 and 3 appears to be, in that order, more efficient in predicting the overall performance of the BAF.

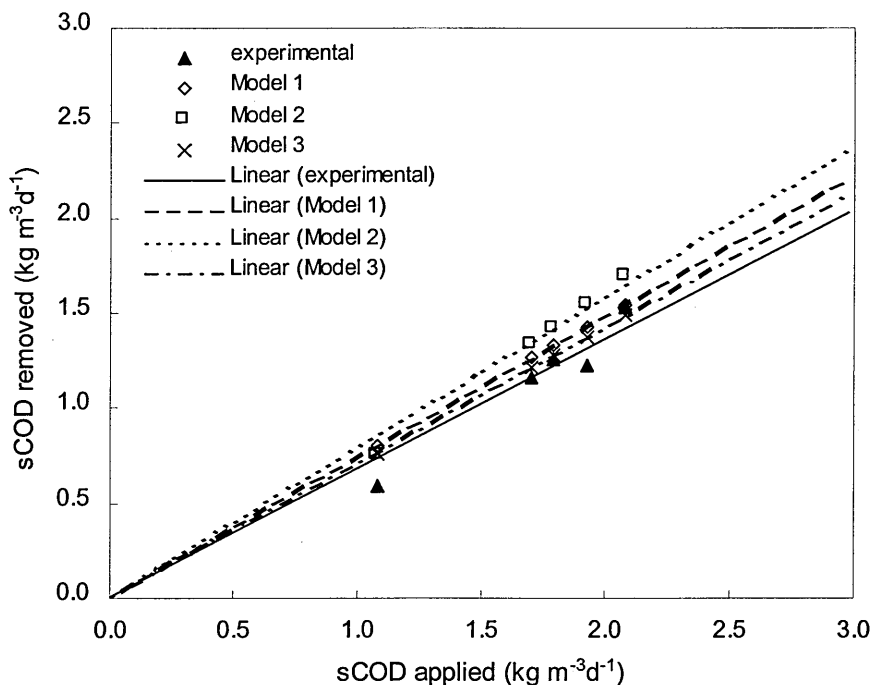


Figure 7.8. Rates of soluble COD applied and removed in the large BAF.

In all three cases, the models overestimated the performance of the reactor yet model 3 gave the closest approximation to the experimental results.

The profiles of sCOD removal in the large column show that all models appeared to be efficient in predicting sCOD removal but in all cases the models failed to predict with greater accuracy the sCOD concentration at the top section of the BAF (Figure 7.9). From the actual observed points it can be highlighted that the concentration of sCOD at a height of 3.9 m was almost the same as that encountered at 2.9 m. With the exception of one point in Model 2, all the observed points at 2.9 m fell within the range predicted by the models. At a height of 3.9 m only Model 1 and 3 seem able to predict accurately the actual sCOD concentration which can explain the fact that both models had a very similar MRE. Consequently, Model 2 presented the highest value of MRE.

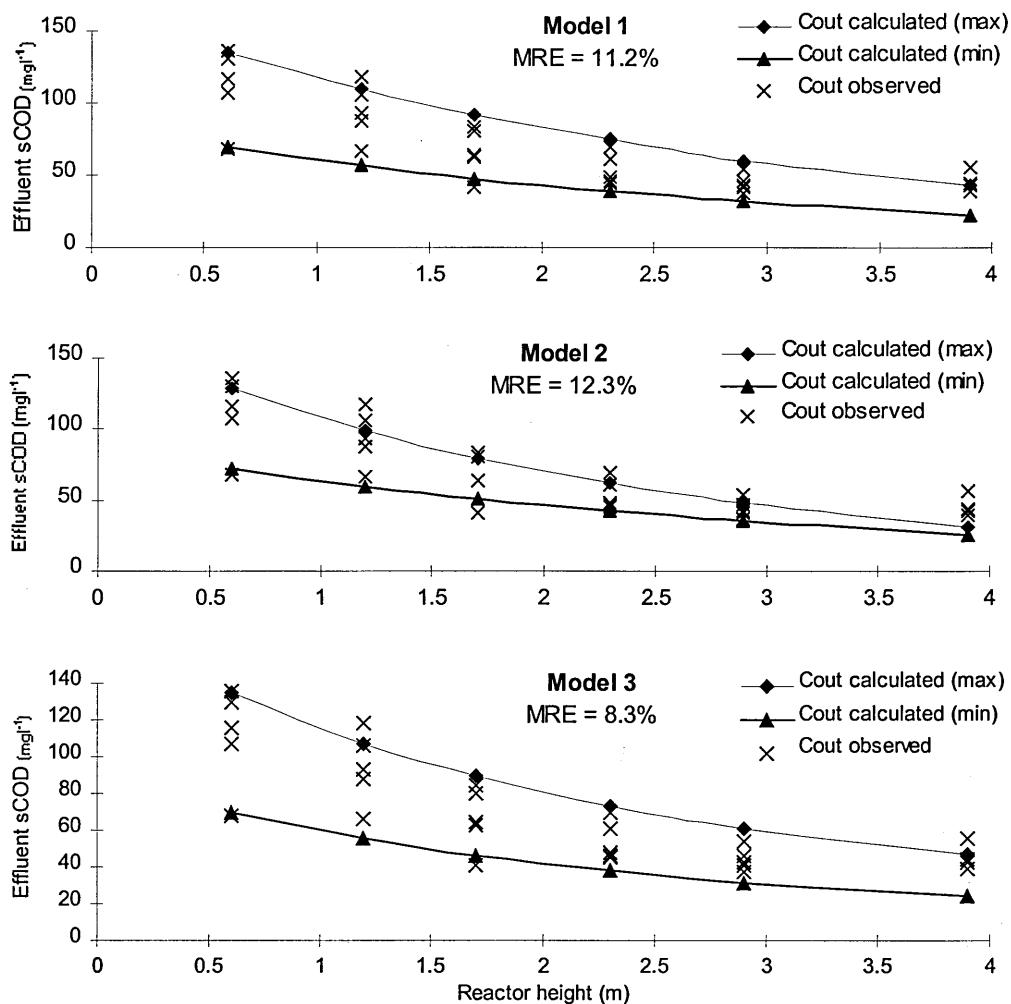


Figure 7.9. Results from five soluble COD removal profiles for Model 1, 2 and 3 undertaken in the large BAF. The lines represent the profiles obtained from the models when applying the maximum and minimum influent concentration.

7.3.2.3 Suspended solids and total COD removal

There was some variation in the effluent concentration of SS and tCOD which caused scattering of the removal rates of the variables when plotted against the loading rates in the large BAF (Figures 7.10 and 7.11).

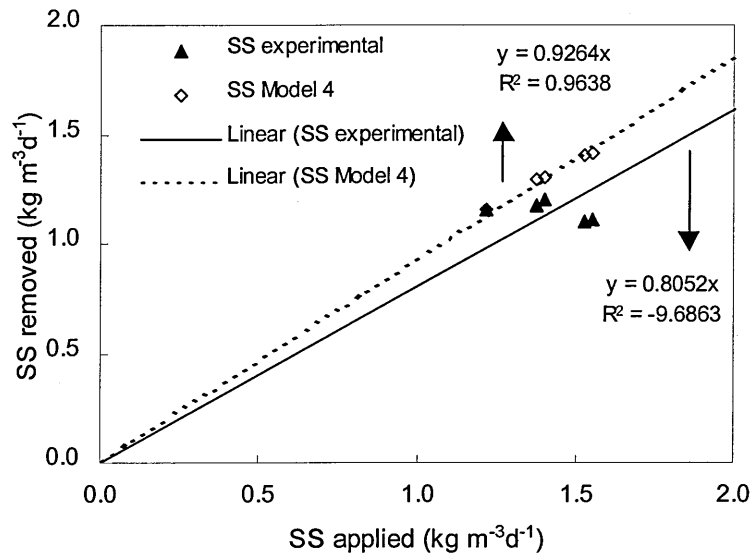


Figure 7.10. Rates of suspended solids (SS) applied and removed in the large BAF.

The values obtained from the model were higher than the actual observed values for both SS and tCOD. The difference between the observed and predicted data was greater for SS than for tCOD.

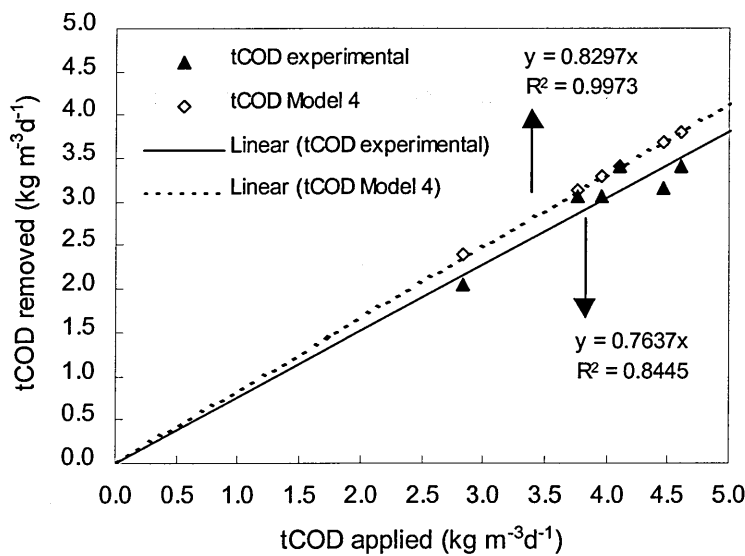


Figure 7.11. Rate of total COD applied and removed in the large BAF.

The profiles of SS and tCOD removal in the large column demonstrated that the model was efficient in estimating the variables concentration at the middle section of the reactor (1.5-3 m) but underestimated the concentration at the bottom and top (Figures 7.12 and 7.13).

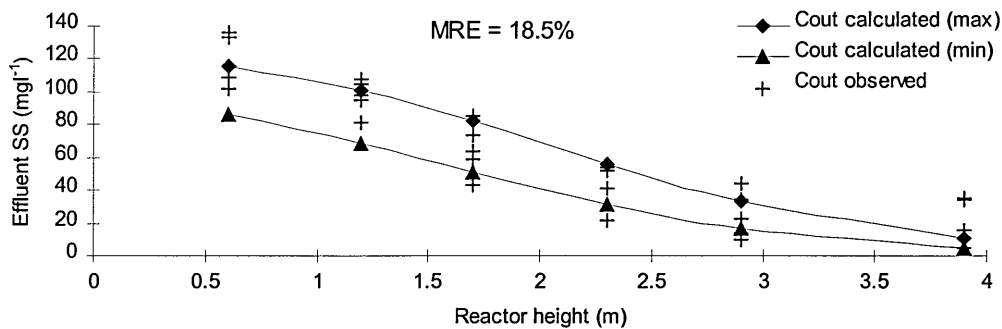


Figure 7.12. Results from five suspended solids (SS) removal profiles for Model 4 undertaken in the large BAF. The lines represent the profiles obtained from the model when applying the maximum and minimum influent concentration.

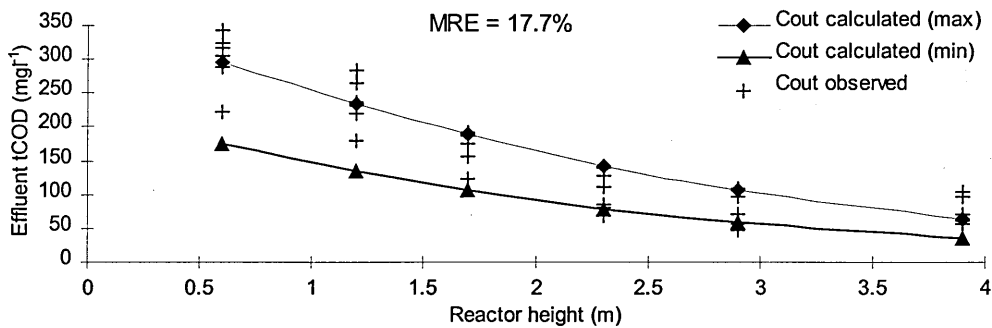


Figure 7.13. Results from six total COD (tCOD) removal profiles for Model 4 undertaken in the large BAF. The lines represent the profiles obtained from the model when applying the maximum and minimum influent concentration.

By comparing the observed concentrations of SS and tCOD at the very top of the reactor (3.9 m) with the sampling port immediately below (2.9 m) it seemed that the top of the reactor (final 1 m) did not achieve further removal of any kind. Solids removal studies by means of particle distribution profiles in sunken media BAFs operated in downflow mode acted as a deep bed filter with decreasing particles removal along the depth for all particle sizes (Boller *et al.*, 1997). The same study

pointed out that floating media BAFs operated in upflow mode showed an increase in the number of particles towards the outlet of the reactor that might indicate solids washout due to bubble induced turbulence. This might be the explanation for the higher suspended solids values found in the top section of the large pilot-scale reactor in the present study. It appeared that suspended solids were being dislodged mainly due to the effect of the air at the top of the filter bed causing deterioration in the quality of the effluent. This same pattern is observed in the small pilot-scale BAF (section 7.3.1.3). The airflow does not seem to have affected the port located at 2.9 m. Therefore, it presented higher removal efficiencies than the port at 3.9 m. It was thus suggested to run the model again excluding the data obtained in that section (data from the sampling port at 3.9 m) (Figures 7.14 and 7.15).

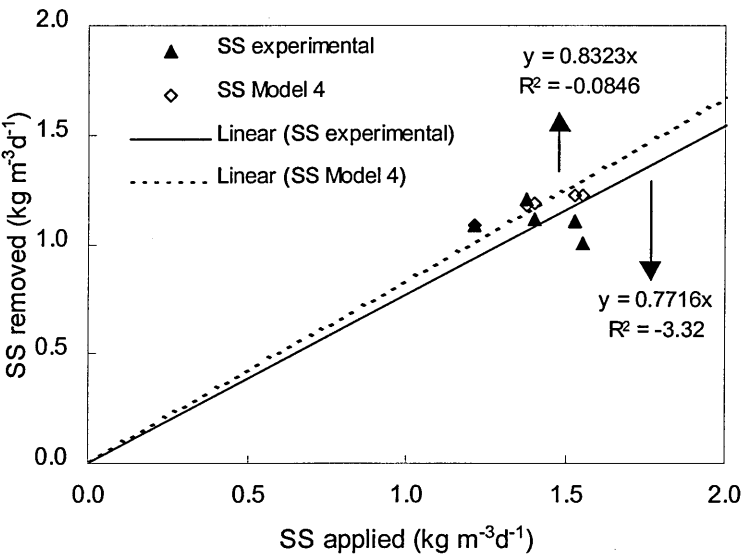


Figure 7.14. Rates of suspended solids applied and removed in the large BAF. Total height of the reactor = 2.9 m.

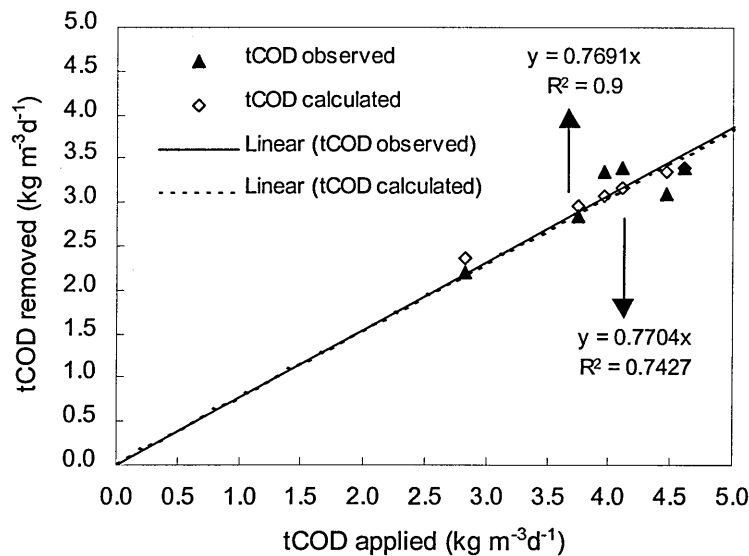


Figure 7.15. Rates of total COD applied and removed in the large BAF. Total height of the reactor = 2.9 m.

It is evident that, by excluding the observed data from the top of the reactor, the values predicted by the model were closer to the trend obtained by the observed data. The variation in the concentration of SS in the effluent still caused inaccuracies in the predicted values by the model however they are very close to the general trend (Figure 7.14). On the other hand, three of the five predicted tCOD values were actually within the trend obtained by the observed data. These results indicate the suitability of the model in predicting the performance of the large BAF in terms of SS and tCOD removal.

The profiles of SS and tCOD removal when the total height of the reactor was assumed to be 2.9 m showed that the values predicted by the model were much closer to the actual values observed (Figures 7.16 and 7.17). By removing the data at the top of the reactor, the concentration of suspended solids in the effluent was very close to the observed data. A couple of observed values at the bottom of the reactor (0.6 m) and one value at the top (2.9 m) were still over the values predicted by the model (Figure 7.16). Nevertheless, the number of observed points outside the model prediction was lower than those obtained previously (Figure 7.12) and the MRE decreased from 18.5% to 10% indicating that the accuracy of the model was enhanced.

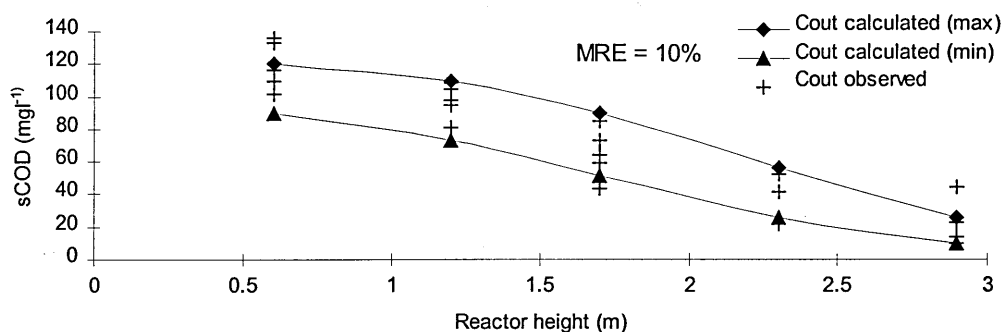


Figure 7.16. Results from five suspended solids removal profiles for Model 4 undertaken in the large BAF (2.9 m height). The lines represent the profiles obtained from the model when applying the maximum and minimum influent concentration.

In terms of tCOD removal, the new predicted values were much closer to the observed concentrations (Figure 7.17). When compared to the values obtained previously at a height of 3.9 m (Figure 7.14), it can be observed that by removing the last observed data the new predicted values were close to those observed. This was more evident in the lower section of the reactor, where the previous calculations had failed to allow for such high concentrations of tCOD. The MRE also decreased from 17.7 to 13.3 indicating a greater accuracy of the model.

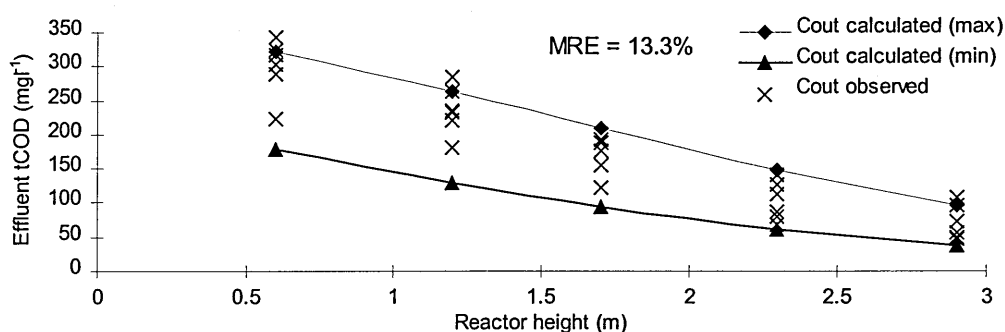


Figure 7.17. Results from six total COD (tCOD) removal profiles for Model 4 undertaken in the large BAF (2.9 m height). The lines represent the profiles obtained from the model when applying the maximum and minimum influent concentration.

These results confirm that in the upper section (2.9-3.9 m), the reactor was not achieving any significant removal of SS and tCOD. By running the model again without the data from the upper section, a greater correlation between observed and

predicted data was reached. This would mean that if the height of the reactor would be decreased from 3.9 to 2.9 m, the degree of treatment would not be adversely affected. This could save money in terms of less pressure headloss, which would cause a reduction in the requirements for aeration, thus energy consumption and costs. It was thus suggested that the model was not only efficient in predicting the removal of SS and tCOD but by comparing the observed and predicted data it was possible to make some decisions regarding the optimum reactor height.

7.3.3 Scaling-up results

7.3.3.1 Soluble COD removal

Constants from the small pilot-scale plant were used to run the models again with the data obtained from the large-scale reactor. Graphic comparison between the experimental sCOD and those predicted by the models are presented below (Figures 7.18-7.23).

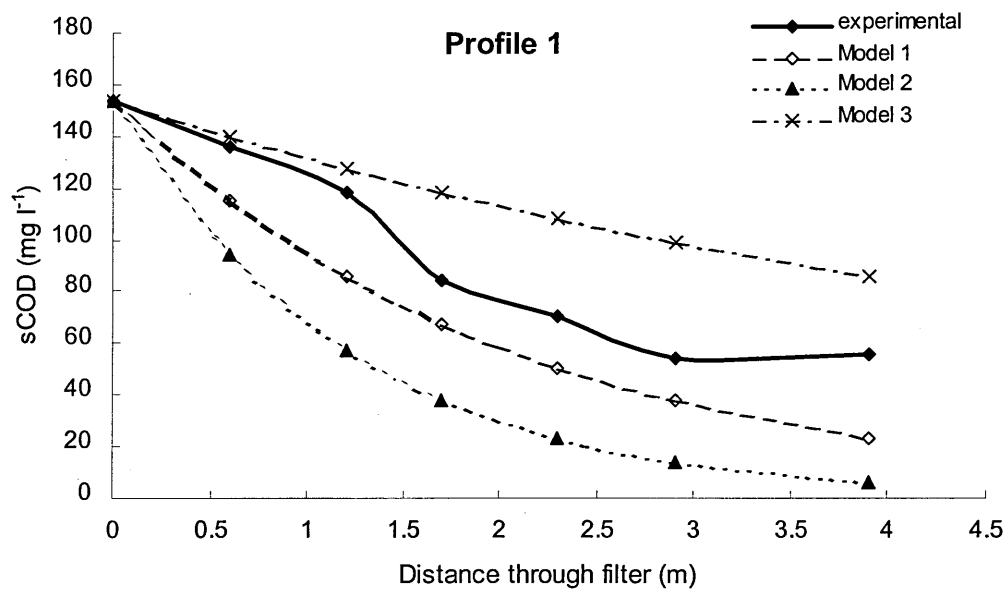


Figure 7.18. Experimental and predicted removal of soluble COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 1.

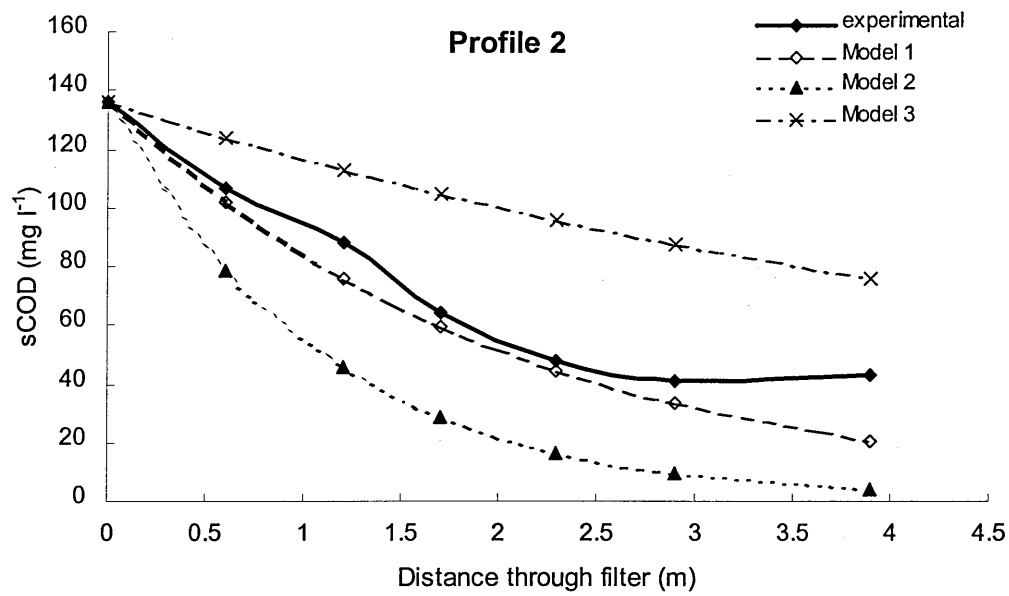


Figure 7.19. Experimental and predicted removal of soluble COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 2.

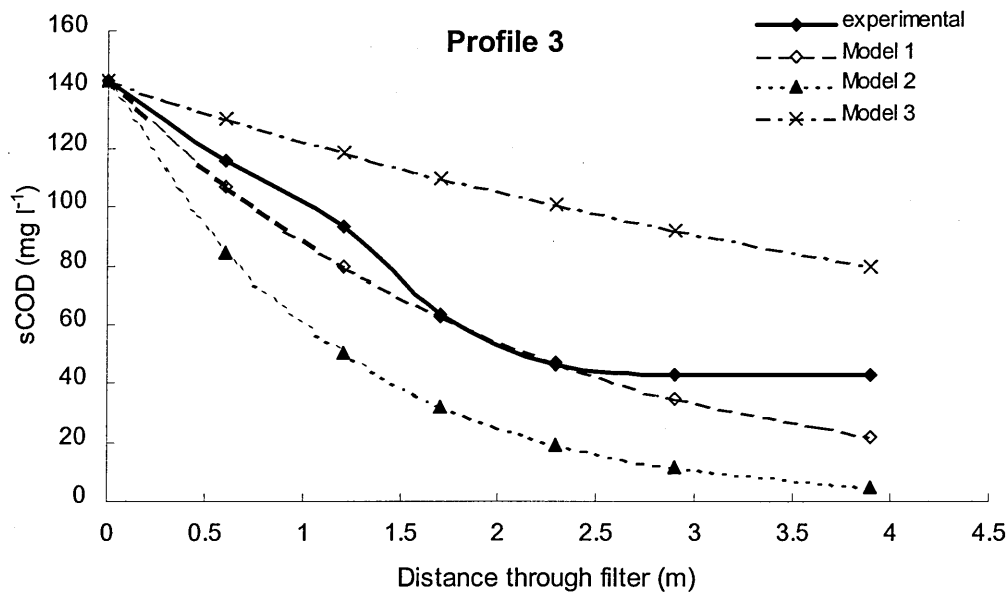


Figure 7.20. Experimental and predicted removal of soluble COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 3.

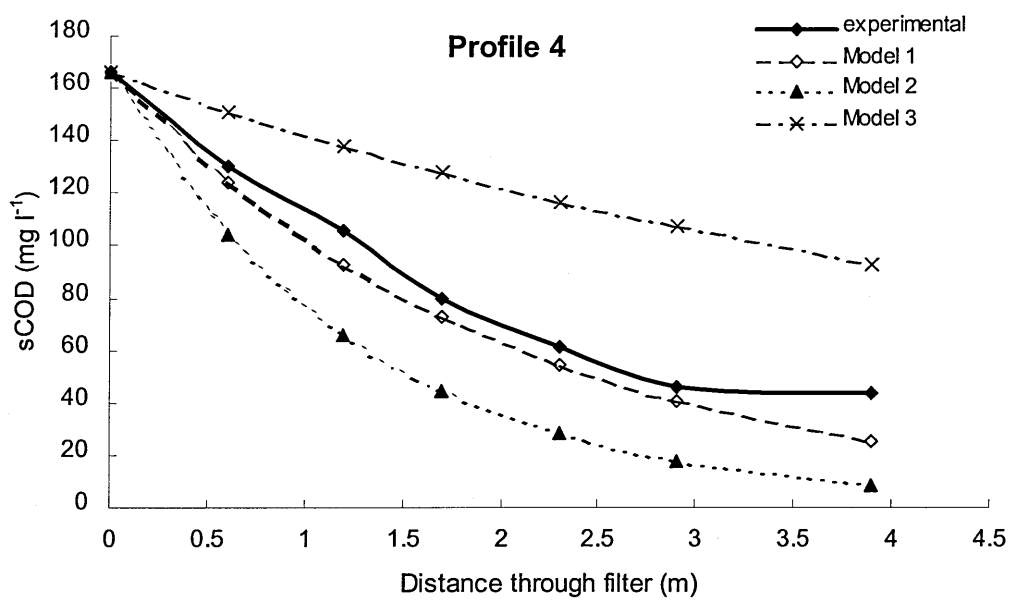


Figure 7.21. Experimental and predicted removal of soluble COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 4.

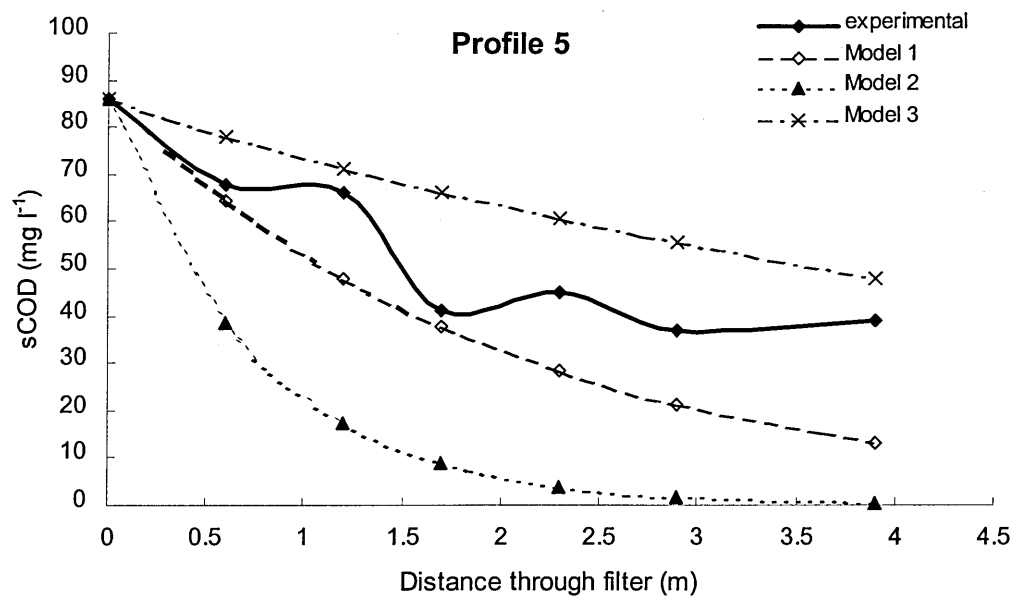


Figure 7.22. Experimental and predicted removal of soluble COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 5.

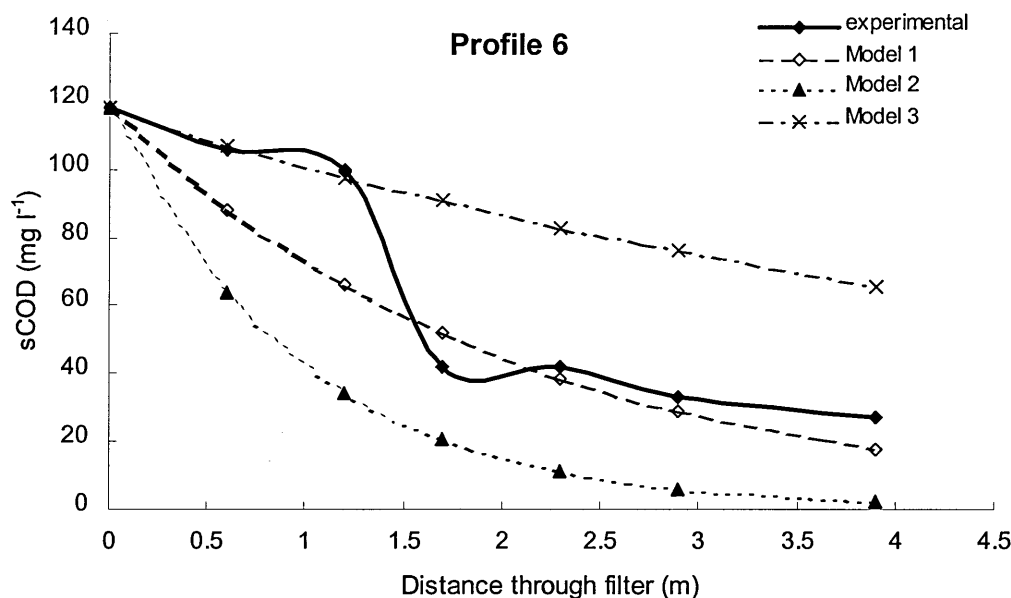


Figure 7.23. Experimental and predicted removal of soluble COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 6.

Model 1 gave the closest approximation to the experimental data in all profiles. Model 2 overestimated the removal capacity of the large pilot-scale BAF and Model 3 underestimated it. By far, Model 3 appears to be the least accurate of the three models for predicting soluble COD removal. Model 3 had already proved to be effective in predicting the removal of sCOD when the corresponding constants from the large BAF were used to verify the model (Section 7.3.2.2) but failed to do so when the constants from the small reactor were used to run the model. Model 1, effectively the Eckenfelder model devised for trickling filters performance (Eckenfelder, 1961), is a very simple empirical model that only takes into account the influent concentration, height of reactor, influent flowrate and cross sectional area of the filter (see equation 7.1). The model makes the assumption that soluble COD removal is a first order reaction thus, is a function of time. Due to the fact that plug-flow conditions prevailed within the reactor (Chapter 4), the removal of sCOD was proportional to the depth of the reactor (James, 1984). Models 2 and 3 incorporated other parameters into Eckenfelder's model such as media and process constants (equation 7.2) and hydraulic and gas velocity to the reactor (equation 7.3). The incorporation of such parameters

proved to have made the models more effective when predicting removal in the large column (with constants from the large column). Air and hydraulic flowrate to the reactors would have a direct effect on the hydraulic configuration of the system, affecting the physical and physiological structure of the biofilm (Arvin and Harremoes, 1990; Wilderer *et al.*, 1995) and thus affecting its removal efficiency. Hence, the models incorporating such parameters were not as effective for scaling-up purposes as the simple model that only took into account first-order reaction kinetics.

7.3.3.2 Suspended solids and total COD removal

Likewise, constants from the small pilot-scale plant were used to run the models again with the data obtained from the large-scale reactor. Graphic comparison between the experimental suspended solids and those predicted by the models are presented below (Figures 7.24 7.29).

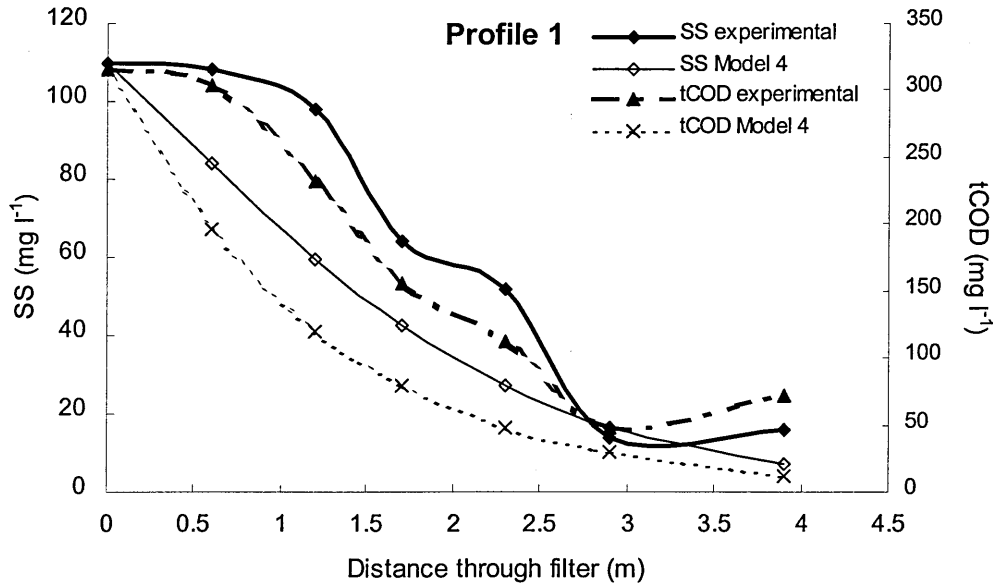


Figure 7.24. Experimental and predicted removal of suspended solids and total COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 1.

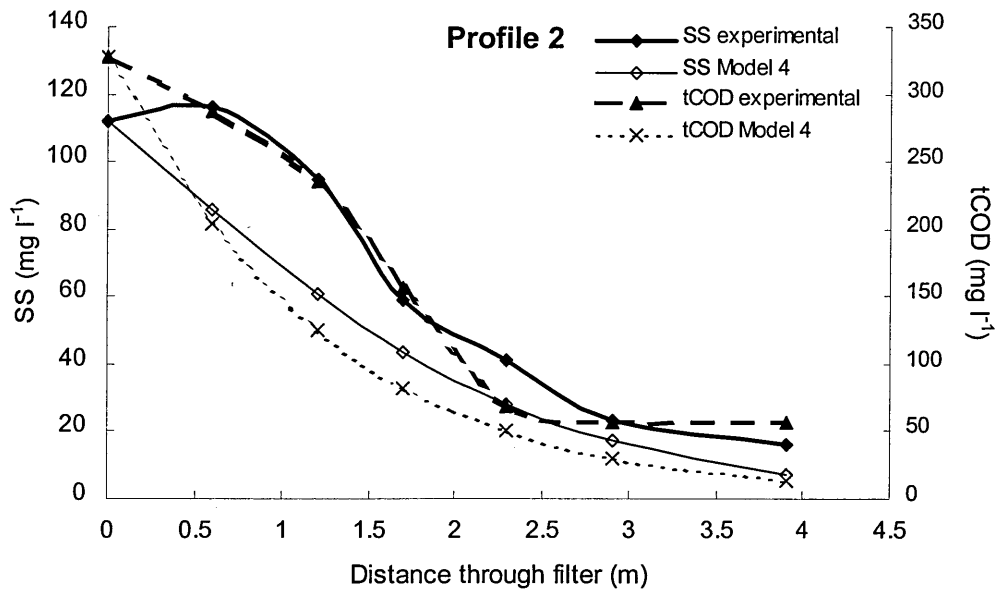


Figure 7.25. Experimental and predicted removal of suspended solids and total COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 2.

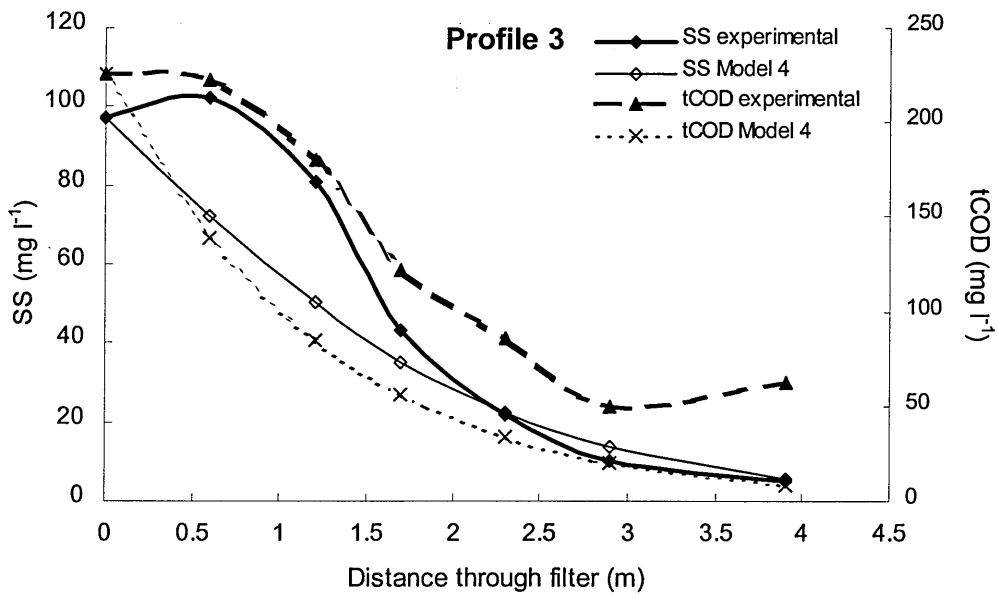


Figure 7.26. Experimental and predicted removal of suspended solids and total COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 3.

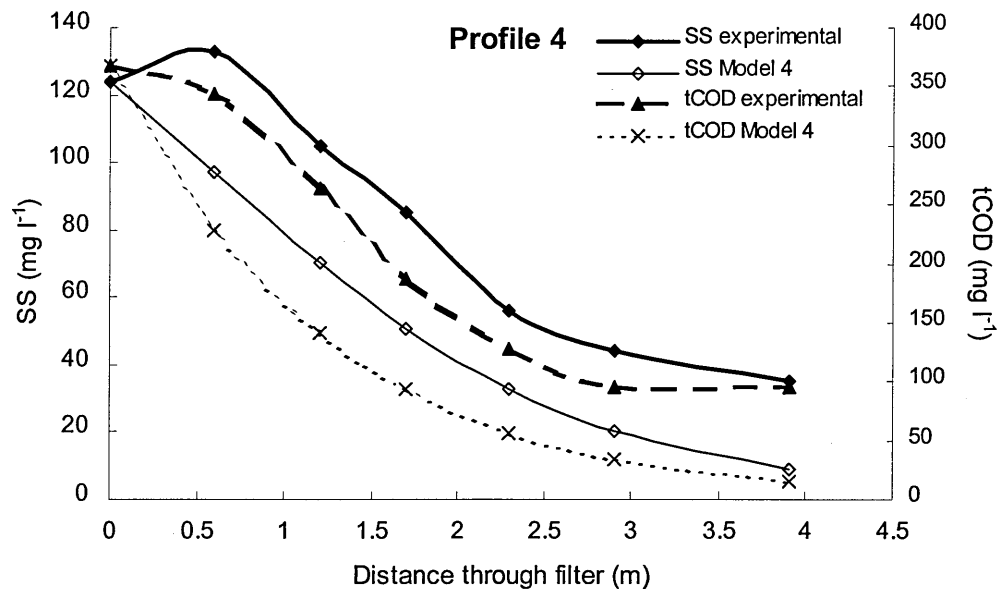


Figure 7.27. Experimental and predicted removal of suspended solids and total COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 4.

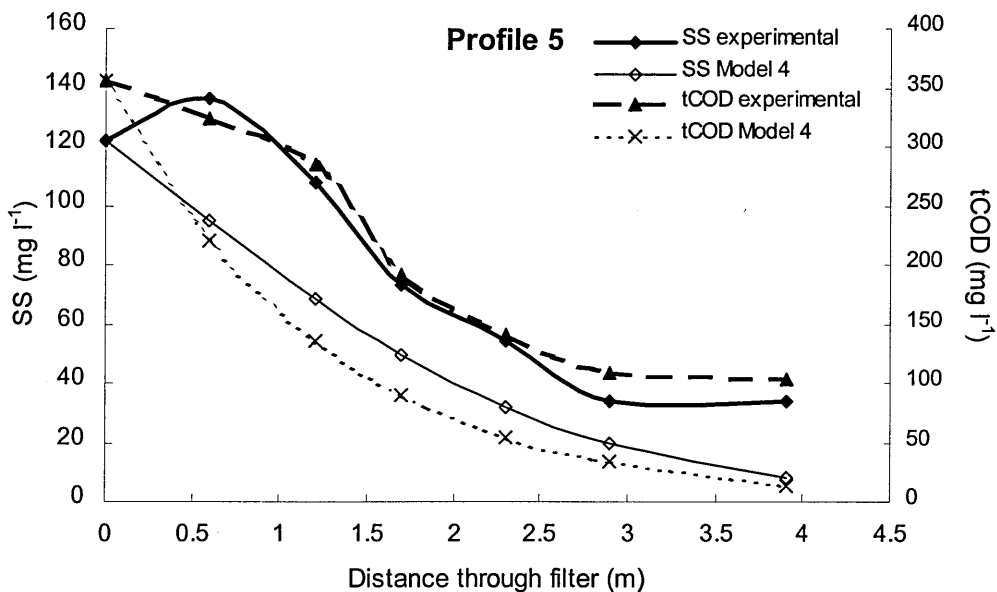


Figure 7.28. Experimental and predicted removal of suspended solids and total COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 5.

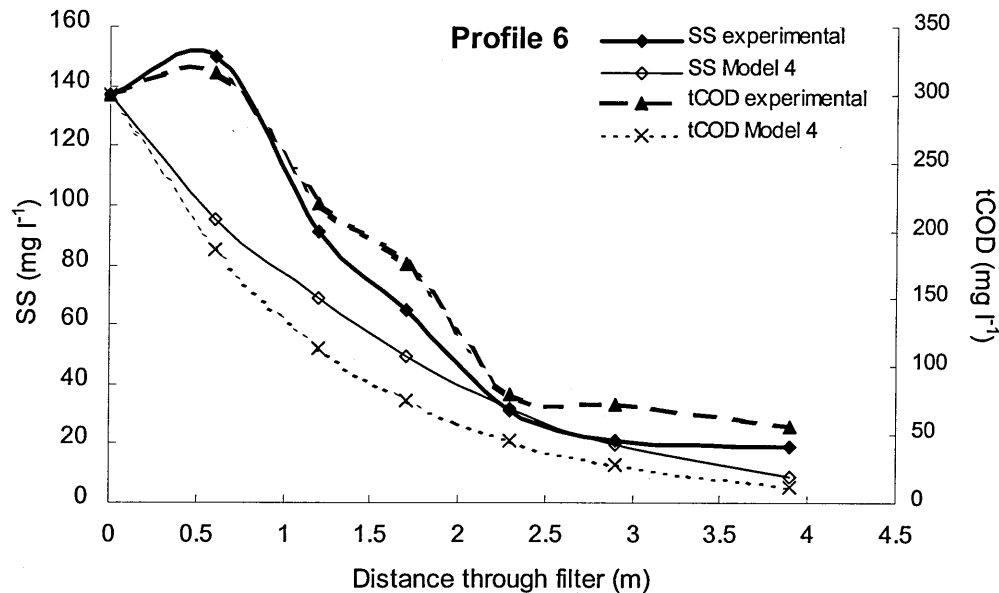


Figure 7.29. Experimental and predicted removal of suspended solids and total COD through the large pilot-scale BAF using constants from the small pilot-scale reactor. Profile 6.

Model 4 appears to have been more effective in predicting the removal of suspended solids than the removal of total COD when the constants from the small pilot-scale BAF were used in the large pilot-scale column. Total COD consists of soluble and particulate COD (Henze *et al.*, 1988). Suspended solids account for a large fraction of the particulate COD hence, in an indirect way, tCOD removal will depend on the removal of SS which in BAFs is mostly undertaken by physical means (Stensel, 1988; Ryhiner *et al.*, 1994). Total COD removal will also depend on the removal of the soluble fraction. In BAFs, soluble COD removal is undertaken by biological means hence, tCOD removal will depend on both physical and biological mechanisms. Model 4, being a model for rapid sand filters (Gimbel, 1989) did not take into account any biological parameters. Hence, the prediction for the removal of tCOD was not as accurate as the prediction for the removal of suspended solids.

The fact that the model in general had little success when used for scaling-up, suggests that there were not many similarities between the systems to allow for such a direct comparison. The theory of similarity stipulates that geometrical, kinematic, dynamic,

chemical, biological and/or thermal must exist between the systems of different dimensions (Horvath and Schmidtke, 1983). The hydraulic loading, film thickness, surface area of packing material and hydraulic characteristics of the system should remain constant (Eckenfelder and Quirk, 1983). By using the same wastewater as a feed, the same medium for biofilm formation and by running both reactors at the same temperature, it was believed that dynamic, chemical, biological and thermal similarities between reactors were being satisfied. Geometric similarity could have been reached, considering that the empty bed volume (EBV) for the large pilot-scale column was almost exactly 10 times larger than that for the small pilot-scale column (Table 7.1). Unfortunately, due to limitations in the equipment as explained earlier, the hydraulic and organic loading rates to the large reactor were considerably higher than those applied to the small BAF (Tables 7.3 and 7.6). Therefore, kinematic similarity could not be reached between the reactors. Parameters like differences in the concentration of the substrate in the inlet and/or the hydraulic loading to the reactors (in terms of hydraulic velocity) played an important role in limiting the kinematic similarity between reactors. This would explain why Models 2, 3 and 4 that do incorporate such parameters were not successful in scaling-up the results.

7.4 Conclusions

- In terms of soluble COD removal, all three models tested were successful in predicting overall performance of the reactor based on organic loading rates and when predicting the profile removal of sCOD. Model 3, which included gas and liquid velocity parameters, provided values closest to the actual observed values when the corresponding constants were used for verifying the models.
- Model 4, previously used for rapid gravity filter, was efficient in predicting the overall removal rates of SS and tCOD in both reactors. The model helped to prove that by reducing the height of the large BAF from 3.9 m to 2.9 m the efficiency of the reactor would not be adversely affected.
- For scaling-up purposes, Model 1 was the most effective for predicting the removal of sCOD in the large pilot-scale BAF when using constants from the small pilot-scale BAF. The lack of kinematic similarity (differences in the organic and hydraulic rates) between reactors seem to have limited the capacity for scaling-up of Models 2, 3 and 4 since these models took into account parameters such as sCOD influent concentration, air and gas velocity.

Chapter 8

ORGANIC AND HYDRAULIC SHOCK LOADS ON A
BIOLOGICAL AERATED FILTERS

ORGANIC AND HYDRAULIC SHOCK LOADINGS ON A BIOLOGICAL AERATED FILTER

ABSTRACT

The performance of a biological aerated filter (BAF) during short term (60 min) organic and hydraulic shock loads was investigated. Off-gas analysis of O_2 was undertaken to try to establish a correlation between off-gas monitoring and BAF performance for the secondary treatment of wastewater. The off-gas analysis technique did not appear to be useful for monitoring the impact of organic and hydraulic shock loads on the biological population of the BAF. The operational oxygen transfer efficiencies (OTEs) of the system were calculated based on the oxygen off-gas concentrations. Increases in the organic and hydraulic loading to the BAF did not appear to have a significant impact on the OTE as other studies suggest. The BAF used for the present experiment was not capable of absorbing high peak organic shock loads when the organic loads were increased from a normal load of $1.2\text{--}1.4 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$ to shock loads between $5.1\text{--}7.3 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$, based on empty bed volumes. Nevertheless, the effect on the biomass was limited as normal performance resumed very quickly. Increases in the hydraulic velocity from $0.7\text{--}1.0 \text{ m h}^{-1}$ to $1.5\text{--}2.9 \text{ m h}^{-1}$ had little affect on BAF performance in terms of soluble COD removal. Yet, sCOD effluent concentrations of $<55 \text{ mg l}^{-1}$ ($\sim 22 \text{ mg l}^{-1} \text{ BOD}_5$) were always achieved thus, marginally failing to meet the discharge consents in most UK sewage. The recovery of the reactor in both cases was fast, resuming normal performance within 60 min after the application of the shock load.

8.1 Introduction

Wastewater treatment works have to be able to cope with changes in the composition of the wastewater and with daily fluctuation in influent flow. Although the installation of flow-equalization, flow-splitting and bypass facilities should be able to deal with peak flowrates, it is essential that the process units are sized to cope with variable flowrates. Sudden increases in substrate concentrations and peak hydraulic flowrates are often the cause for failing discharge consents (Droste, 1997).

It has been suggested that shock loads in continuous flow biofilm reactors such as biological aerated filters (BAFs) cause stratification and uneven distribution of the biomass, thus, limiting the ability of the reactor to respond (Woolard, 1997). However, there is no general consensus about the resistance of BAFs to shock loadings. Biological aerated filters tend to operate under plug-flow conditions (Fdz-Polanco *et al.*, 1996) and treat organic shock loadings through a gradual reduction in concentration (Chen and Cheng, 1994). According to some authors (Wilderer *et al.*, 1993), plug-flow fixed-bed biofilm reactors tend to have species, population size and activity stratification along the reactor length in the direction of the flow. Therefore, if the feed concentration or the flowrate is increased, some components of the wastewater are transported to zones of the reactor where the required microorganisms are not present in adequate numbers. Moreover, high peak loadings may cause breakthrough of organic matter (Ruffer and Rosenwinkel, 1984). However, although it has been suggested that high concentrations of toxic matter, acids and alkaline have a negative impact on BAF performance, under normal circumstances recovery can be quick due to the adaptability of the biofilm (Adachi and Fuchū, 1991). This adaptability could make BAFs resistant to shock hydraulic and organic loads when treating wastewater with hydrocarbon compounds (Holladay *et al.*, 1978).

In many wastewater treatment plants, peak influent flowrates as well as biochemical oxygen demand (BOD) and suspended solids (SS) loadings can reach two or more

times average values (Droste, 1997). Previous studies have found that BAFs are capable of treating seasonal changes in the influent characteristics (Gilles, 1990). It has also been found that BAFs are capable of performing satisfactorily under unsteady-state conditions (Mann *et al.*, 1999). However, the study of the impact of hourly short-term variations on BAFs, e. g. those caused by the habits of community residents or industrial activities, are limited (Desbos *et al.*, 1990).

Although off-gas analysis has been used for monitoring aeration efficiencies in BAFs (Harris *et al.*, 1996), their use has not been widespread. The objective of this work was to monitor the performance of a biological aerated filter during organic and hydraulic shock loads. Moreover, off-gas analysis of O_2 was undertaken to try to establish a correlation between off-gas monitoring and BAF performance during the secondary treatment of wastewater.

8.2 Materials and Methods

Experiments were undertaken in a cylindrical polyvinylchloride (PVC) reactor with an overall height of 3 m and diameter of 0.2 m (see Chapter 4). The reactor contained 38 l of quartz sand spherical media, with a diameter of 2-3.2 mm, a specific surface area of $1500 \text{ m}^2 \text{ m}^{-3}$ and 40% voidage. The reactor was operated in downflow mode through a media bed of 2 m. Primary settled sewage was introduced at the top of the column at a flowrate of 0.4 l min^{-1} ($0.576 \text{ m}^3 \text{ d}^{-1}$). The overall column of water was 3 m and was kept constant by means of an overflow at the top of the reactor. Compressed air was supplied via three aerators (Degremont UK, Dunstable, UK) at a flowrate of 4 l min^{-1} . The reactor was backwashed daily to avoid clogging (see Chapter 4).

Previous tracer studies with sodium chloride have demonstrated that plug-flow conditions prevailed within the reactor where the media was present (Chapter 4). The section on top of the media was found to be completely mixed. Percentages of removal were calculated based on parameter concentration at the inlet and the outlet.

8.2.1 Organic shock load application

During organic shock loadings the hydraulic flowrate remained constant and the organic loading was increased by adding artificial sewage to the primary settled sewage until it reached the desired strength. Due to the composition of the artificial sewage (Table 8.1), the chemical oxygen demand (COD) of the wastewater was increased but the concentration of suspended solids (SS) was not affected.

Table 8.1. Artificial sewage composition (from Sterritt and Lester, 1981).

Reagent	g l ⁻¹
Neutralised Bacterial Peptone (L34) †	7.80
Lab Lemco (L29) †	5.20
NH ₄ Cl ‡	2.40
NaCl ‡	0.30
KH ₂ PO ₄ ‡	0.20
CaCl ₂ ‡	0.16
MgSO ₄ · 7H ₂ O ‡	0.05

† Oxoid Ltd, Milton Keynes, UK

‡ Merck Ltd, Leicester, UK

Five organic shock loadings were applied (OSL 1, OSL 2, OSL 3, OSL 4, OSL 5). Due to the variable characteristics of the wastewater, a constant organic loading for all shock loads was not reached but fell within a range (Table 8.2). Shock load conditions lasted for 60 minutes after which normal operating conditions were resumed.

Table 8.2. Organic loadings during normal operation and organic shock loads. (hydraulic flowrate 0.58 m³ d⁻¹: hydraulic velocity 1.8 m h⁻¹: HRT section with media 63 min).

	Normal operation		Shock load	
	Organic loading based on total EBV* (kg m ⁻³ d ⁻¹)	Organic loading based on media volume (kg m ⁻³ d ⁻¹)	Organic loading based on total EBV* (kg m ⁻³ d ⁻¹)	Organic loading based on media volume (kg m ⁻³ d ⁻¹)
OSL 1	1.53	2.26	3.14	4.62
OSL 2	1.59	2.35	4.29	7.06
OSL 3	1.76	2.59	5.26	7.75
OSL 4	1.83	2.70	5.91	8.72
OSL 5	1.82	2.68	6.21	9.16

*empty bed volume

8.2.2 Hydraulic shock load application

Four hydraulic shock loads were undertaken at 2 times the normal hydraulic flowrate for a period of 60 min (HSL 1, HSL 2, HSL 3, HSL 4). A fifth hydraulic shock load

was applied at 3 times the normal operating hydraulic flowrate of the BAF (HSL 5). Tap water was used to dilute the primary sewage so that the volumetric organic loading remained similar (Table 8.3).

Table 8.3. Organic loadings during normal operation and hydraulic shock loads.

	Normal operation				Shock load			
	Hydraulic flowrate ($\text{m}^3 \text{d}^{-1}$)	Hydraulic velocity (m h^{-1})	Organic loading based on total EBV* ($\text{kg m}^{-3} \text{d}^{-1}$)	Organic loading based on total EBV* ($\text{kg m}^{-3} \text{d}^{-1}$)	Hydraulic flowrate ($\text{m}^3 \text{d}^{-1}$)	Hydraulic velocity (m h^{-1})	Organic loading based on media volume ($\text{kg m}^{-3} \text{d}^{-1}$)	Organic loading based on media volume ($\text{kg m}^{-3} \text{d}^{-1}$)
HSL 1	0.58	1.8	1.03	1.52	1.25	3.7	0.67	0.99
HSL 2	0.72	2.2	1.67	2.46	1.34	4.0	1.67	2.47
HSL 3	0.72	2.2	1.16	1.71	1.22	3.7	1.09	1.61
HSL 4	0.50	1.6	0.68	0.93	1.15	3.6	0.82	1.21
HSL 5	0.50	1.6	0.81	1.19	2.16	7.4	2.7	3.9

*empty bed volume

8.2.3 Calculation of oxygen transfer efficiencies (OTEs) using off-gas analysis

Oxygen mass balances were undertaken to determine the aeration efficiencies in the BAF. The analysis of O_2 was undertaken using an ADC7000 series analyser with paramagnetic cell (Analytical Development Company Ltd, Herts., UK) which gave the O_2 concentration as a percentage, corrected for temperature, humidity and pressure. The reactor had a collection hood at the top with a small orifice through which a polypropylene pipe with an internal diameter of 5 mm was connected to the oxygen analyser. The analyser had an internal pump set to such a flowrate that the time between the actual gas sampling and the reading was 15 seconds. The oxygen transfer efficiency (OTE), defined as the ratio of the mass of oxygen transferred to the mass of oxygen supplied, was determined using off-gas analysis. Gas flow measurements could be omitted in the calculation of the OTE of the system by using the molar ratios of inlet and outlet oxygen to the inert gas fraction (Redmon *et al.*, 1983). The OTE was calculated as follows.

First, the off-gas analyser provided a reading of oxygen in volume percent. Therefore, the volume of oxygen in 1 m³ of air was calculated as:

$$\text{m}^3 \text{ O}_2 = \frac{\% \text{ O}_2 \text{ offgas}}{100} \quad (8.1)$$

The relationship between volume of gas and its moles at 1 atm and at a certain temperature is determined by using the following equation (Doran, 1995):

$$pV = nRT \quad (8.2)$$

where p = absolute pressure (atm)

V = gas volume (m³)

n = moles (gmol)

R = ideal gas constant (8.2057 x 10⁻⁵ m³ atm gmol⁻¹ °K⁻¹)

T = absolute temperature (°K)

Hence:

$$n = \frac{pV}{RT} \quad (8.3)$$

The amount of O₂ that each cubic metre of off-gas contains is calculated by:

$$(\text{gmol O}_2)(32) = \text{g O}_2 \text{ per m}^3 \text{ metre of offgas} \quad (8.4)$$

Multiplied by with the flowrate provided the amount of O₂ supplied or consumed in the system:

$$\text{mass of O}_2 \text{ supplied or consumed} = \frac{\text{g O}_2}{\text{m}^3} \times Q_{\text{air}} \quad (8.5)$$

where Q_{air} is the air flowrate (m³ d⁻¹).

Finally, the OTE was derived by the following expression:

$$\text{OTE} = \frac{\text{mass O}_2 \text{ in} - \text{mass O}_2 \text{ out}}{\text{mass O}_2 \text{ in}} \quad (8.6)$$

8.2.4 Wastewater sampling and analysis

During organic shock loadings application, samples of the influent and effluent of the reactor were undertaken every 20 min. After the application of the organic shock loads, wastewater sampling was undertaken every 10 min until steady-state conditions were resumed. Influent and effluent samples were taken every 15 min during hydraulic shock loadings.

All wastewater samples were analysed for suspended solids (SS) and soluble chemical oxygen demand (sCOD) using standard methods (Greenberg *et al.*, 1992). Dissolved oxygen (DO) and temperature were measured using a Jenway DO-9071 meter (Merck Ltd, Leicester, UK).

8.3 Results and Discussion

8.3.1 Off-gas analyses during organic shock loads

All time values were transformed into dimensionless time (τ) by dividing the actual time by the theoretical hydraulic retention time (Table 8.2); hence, a dimensionless value of 1 is equal to 63 min. During all the shock loads, the concentration of sCOD in the effluent started to increase between τ values of 0.6 and 1.3 with the maximum at τ values between 1 and 1.7 (Figures 8.1-8.5). In all cases the effluent concentrations of SS and sCOD returned to normal within 2τ .

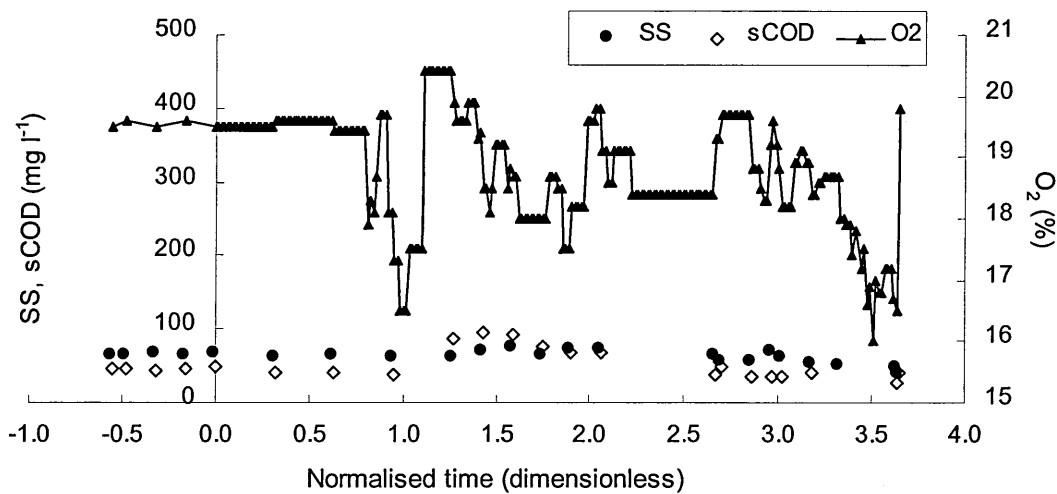


Figure 8.1. Off-gas oxygen, effluent SS and effluent sCOD variation during organic shock load 1 (OSL 1).

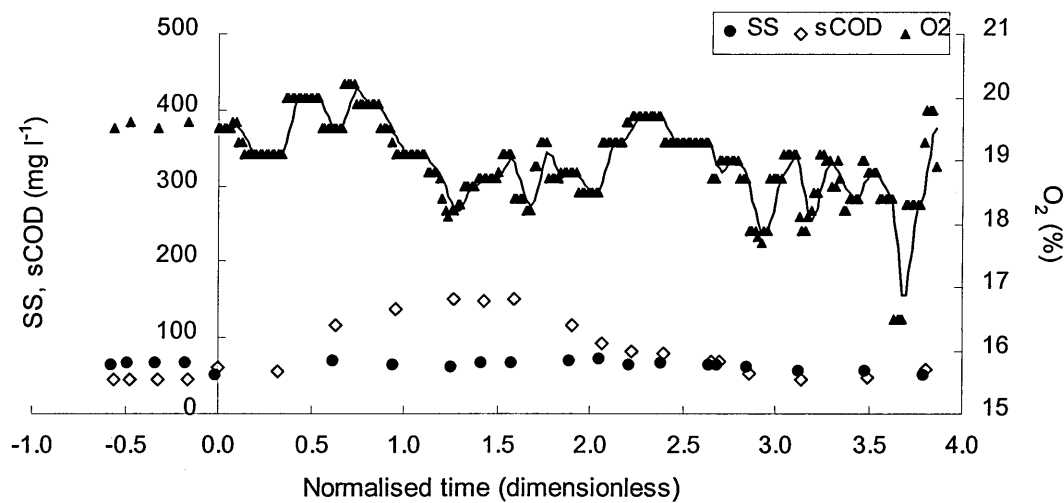


Figure 8.2. Off-gas oxygen, effluent SS and effluent sCOD variation during organic shock load 2 (OSL 2).

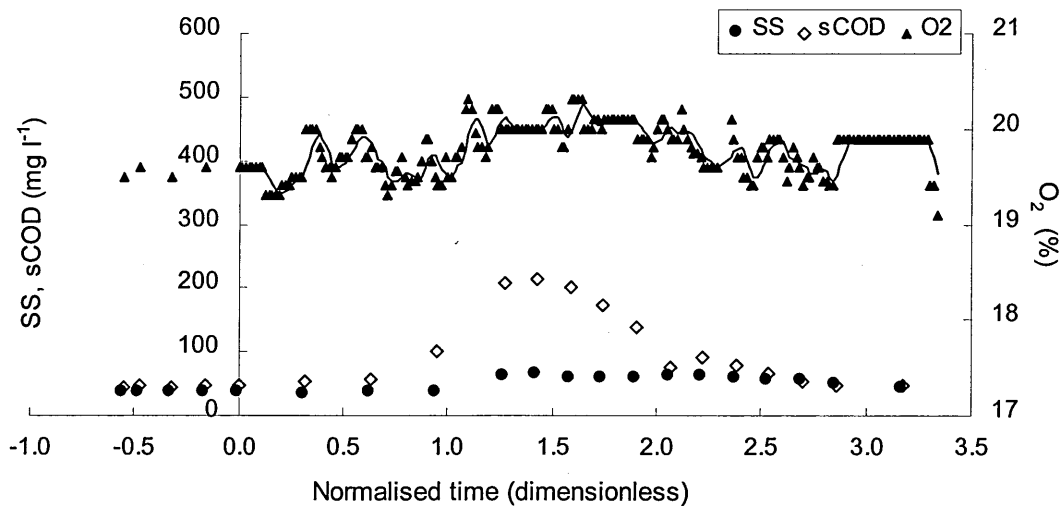


Figure 8.3. Off-gas oxygen, effluent SS and effluent sCOD variation during organic shock load 3 (OSL 3).

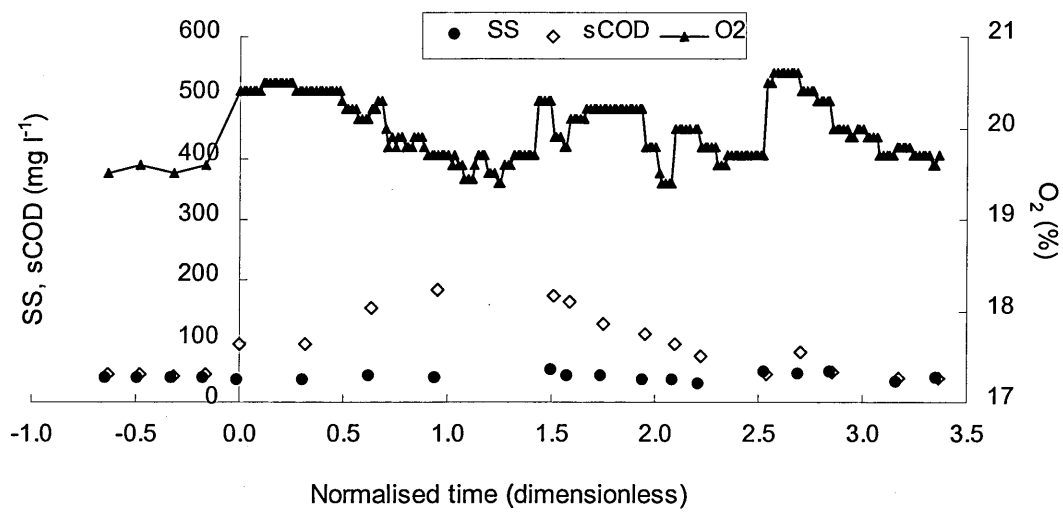


Figure 8.4. Off-gas oxygen, effluent SS and effluent sCOD variation during organic shock load 4.

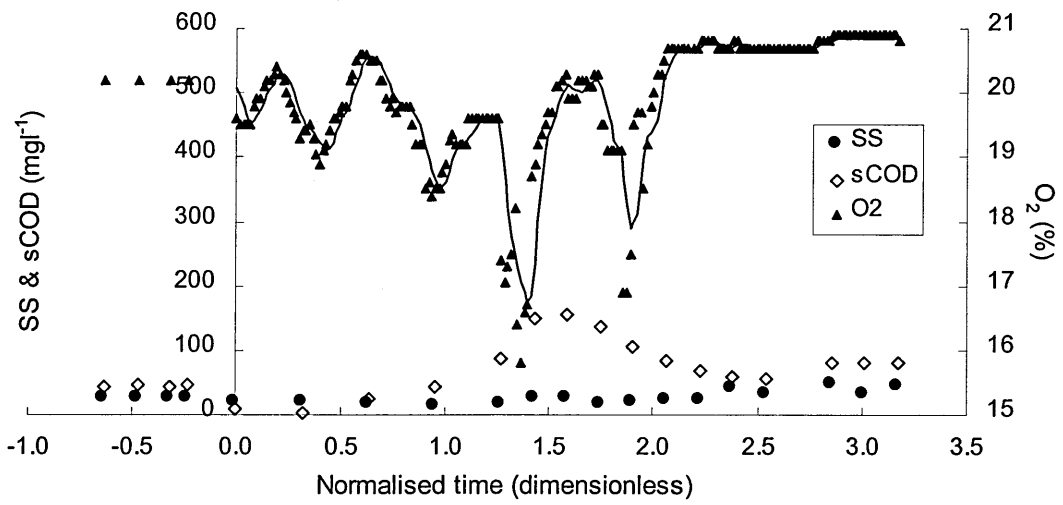


Figure 8.5. Off-gas oxygen, effluent SS and effluent sCOD variation during organic shock load 5 (OSL 5).

It was expected that the use of off-gas analysis would monitor the response of the biological system to the increase in organic substrate during the organic shock loads. However, Figures 8.1–8.5 do not seem to show this. In order to try to find a correlation between reactor performance and O_2 off-gas production, the concentration of sCOD in the effluent and off-gas O_2 were plotted (Figure 8.6).

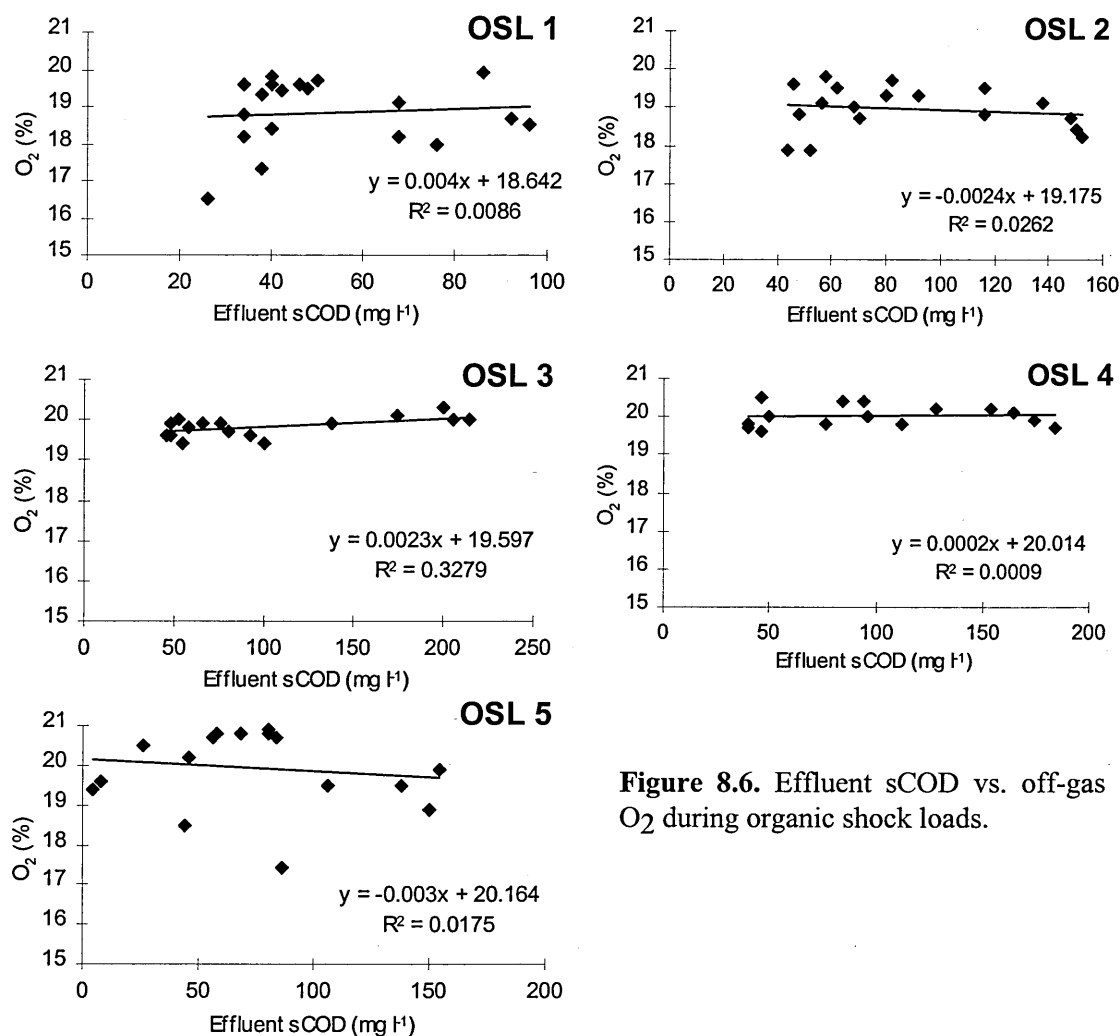


Figure 8.6. Effluent sCOD vs. off-gas O_2 during organic shock loads.

Clearly, the effluent sCOD and off-gas O_2 do not have any type of correlation. In all five organic shock loads there is no clear linear pattern between the two variables.

Thus, it was thought that due to the reactor configuration (with the overflow at the top), the flowrate could have had a greater effect on off-gas O_2 than the actual organic matter removed (Figure 8.7).

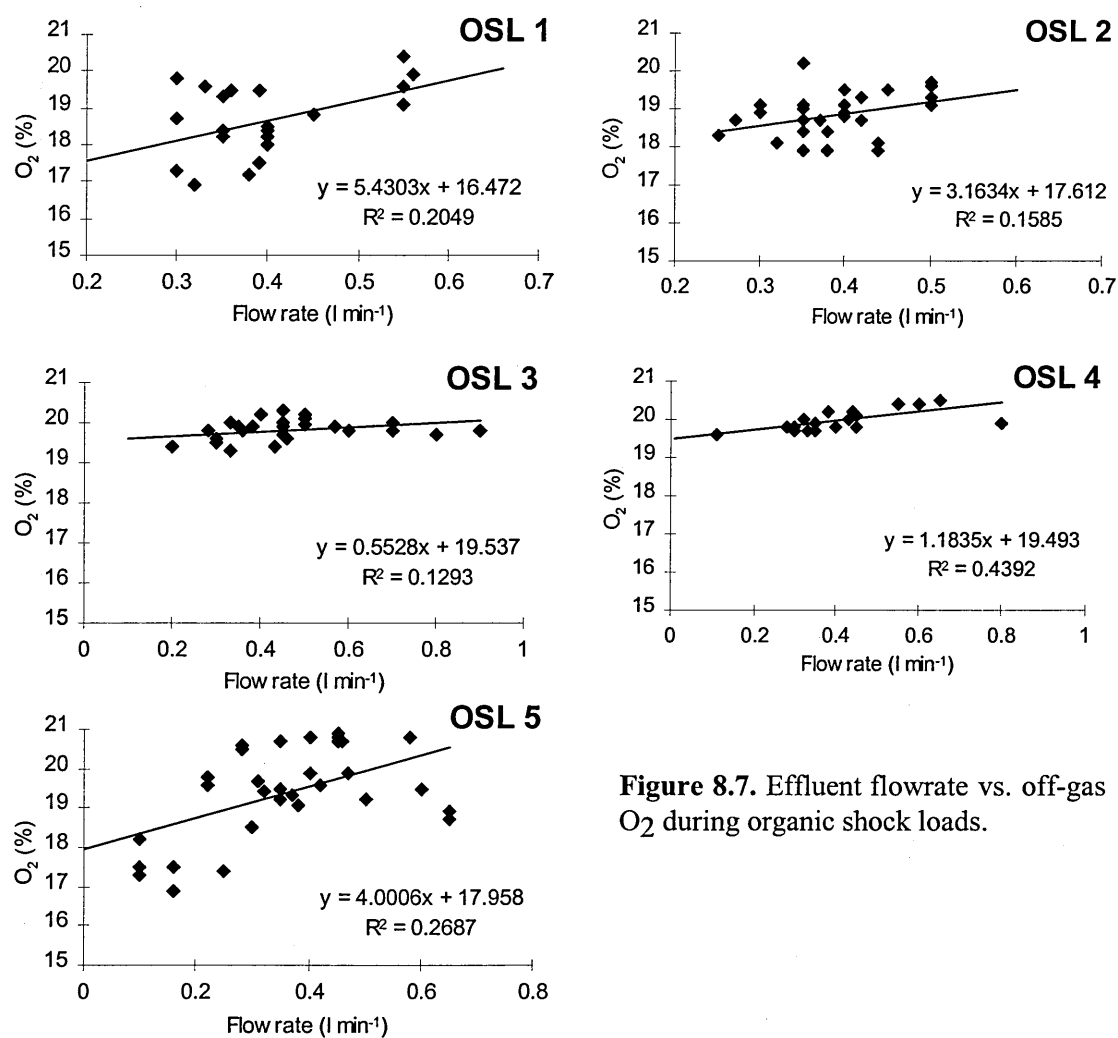


Figure 8.7. Effluent flowrate vs. off-gas O₂ during organic shock loads.

Correlation between off-gas O₂ and the flowrate was very low. Hence, it was not possible to demonstrate that the off-gas measurement was affected by the presence of the overflow at the top of the reactor not by the removal of soluble COD.

8.3.2 Off-gas analysis during hydraulic shock loads

Off-gas O₂ appeared to follow a more consistent pattern during hydraulic shock loads (Figures 8.8-8.12) than during organic shock loads previously presented. When the flowrate was increased by a factor of 2 (Figures 8.8-8.11), the lowest percentages of O₂ were always detected around 70-80 min after the start of the shock load, 10-20 after its end. When the flowrate was further increased (Figure 8.12) the lowest off-gas readings were detected 120 min after the start of the shock load, 60 min after its end.

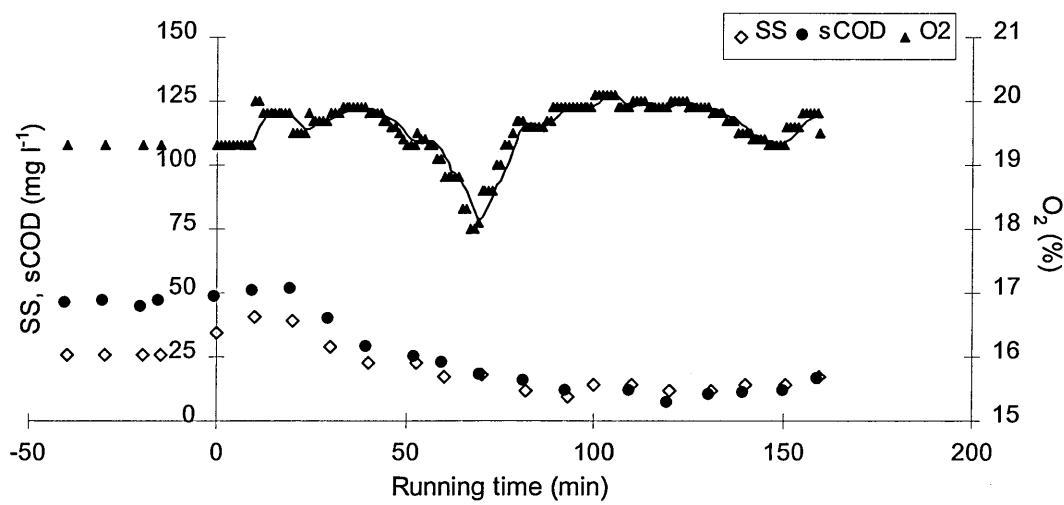


Figure 8.8. Off-gas oxygen, effluent SS and effluent sCOD variation during hydraulic shock load 1 (HSL 1).

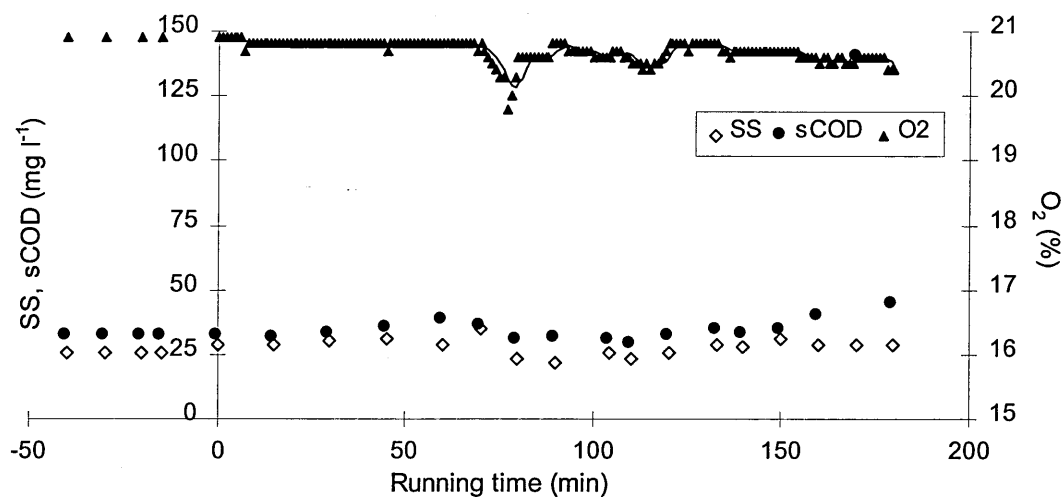


Figure 8.9. Off-gas oxygen, effluent SS and effluent sCOD variation during hydraulic shock load 2 (HSL 2).

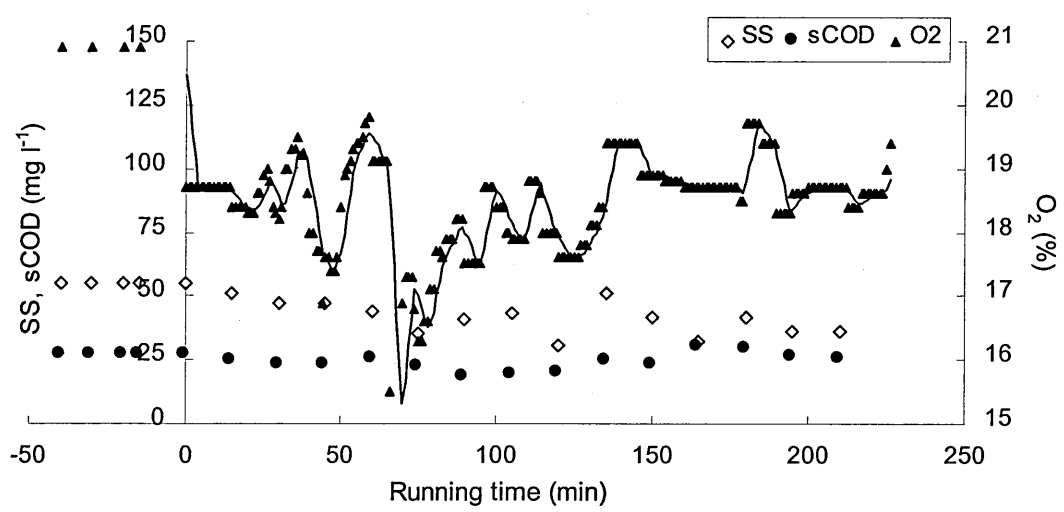


Figure 8.10. Off-gas oxygen, effluent SS and effluent sCOD variation during hydraulic shock load 3 (HSL 3).

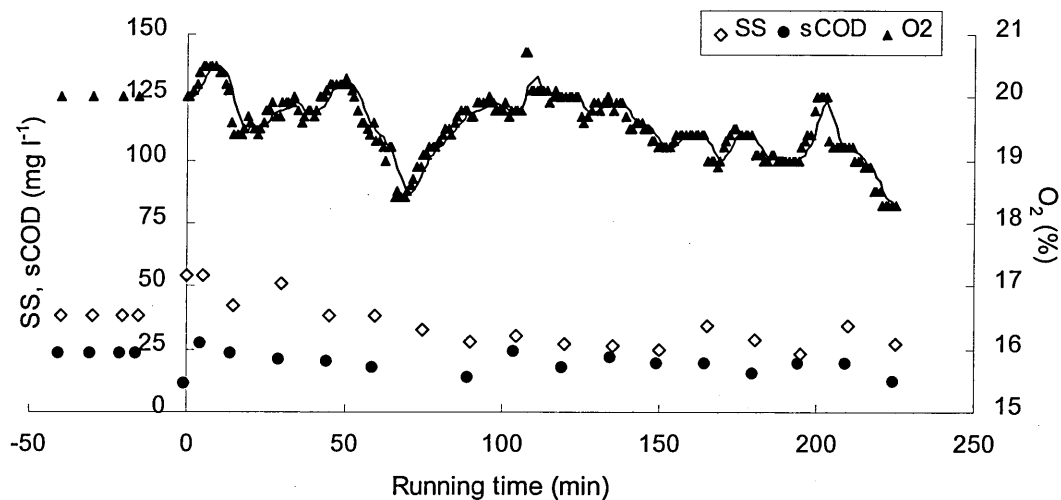


Figure 8.11. Off-gas oxygen, effluent SS and effluent sCOD variation during hydraulic shock load 4 (HSL 4).

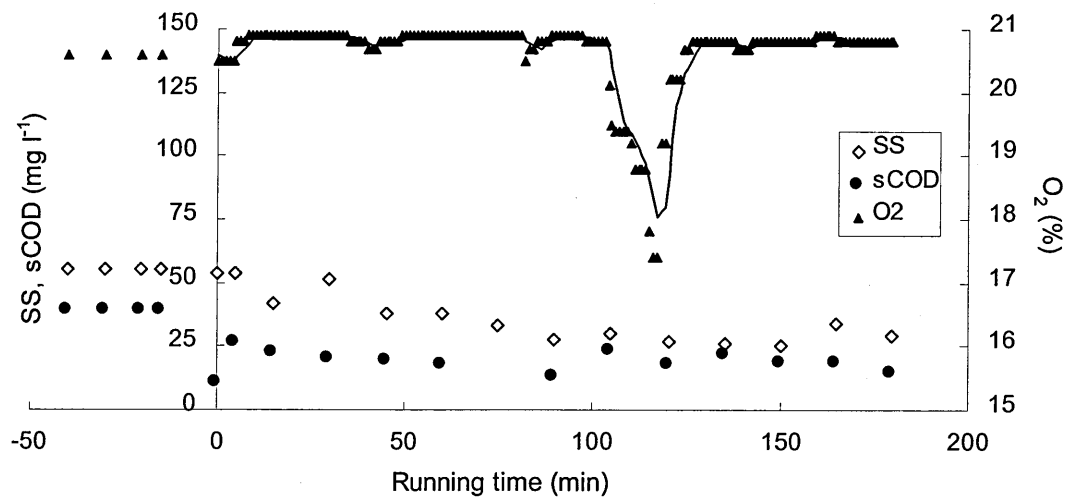


Figure 8.12. Off-gas oxygen, effluent SS and effluent sCOD variation during hydraulic shock load 5 (HSL 5).

In order to find out if the variation of the off-gas O₂ was being caused by the removal of soluble organic matter in the BAF, the effluent sCOD and off-gas O₂ were plotted (Figure 8.13).

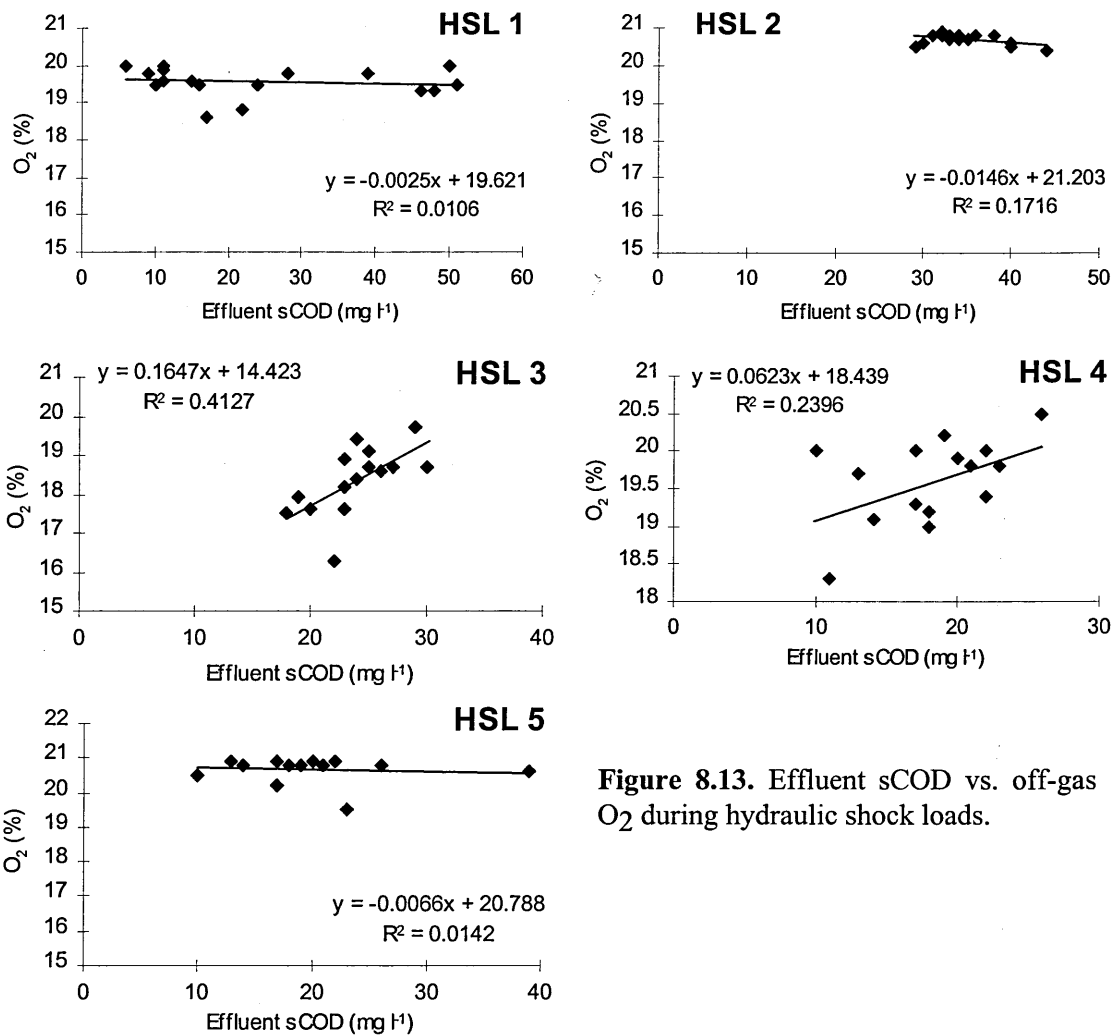


Figure 8.13. Effluent sCOD vs. off-gas O₂ during hydraulic shock loads.

In general, there was no correlation between the off-gas oxygen and effluents sCOD. Only during hydraulic shock load 3 (HSL 3) there was a positive correlation between the two parameters. The off-gas O₂ results from HSL 3 showed the greater amount of scattering of all the hydraulic shock loads.

Hence, hydraulic flowrate was plotted against off-gas O_2 to try to find an explanation for the variation of the former (Figure 8.14).

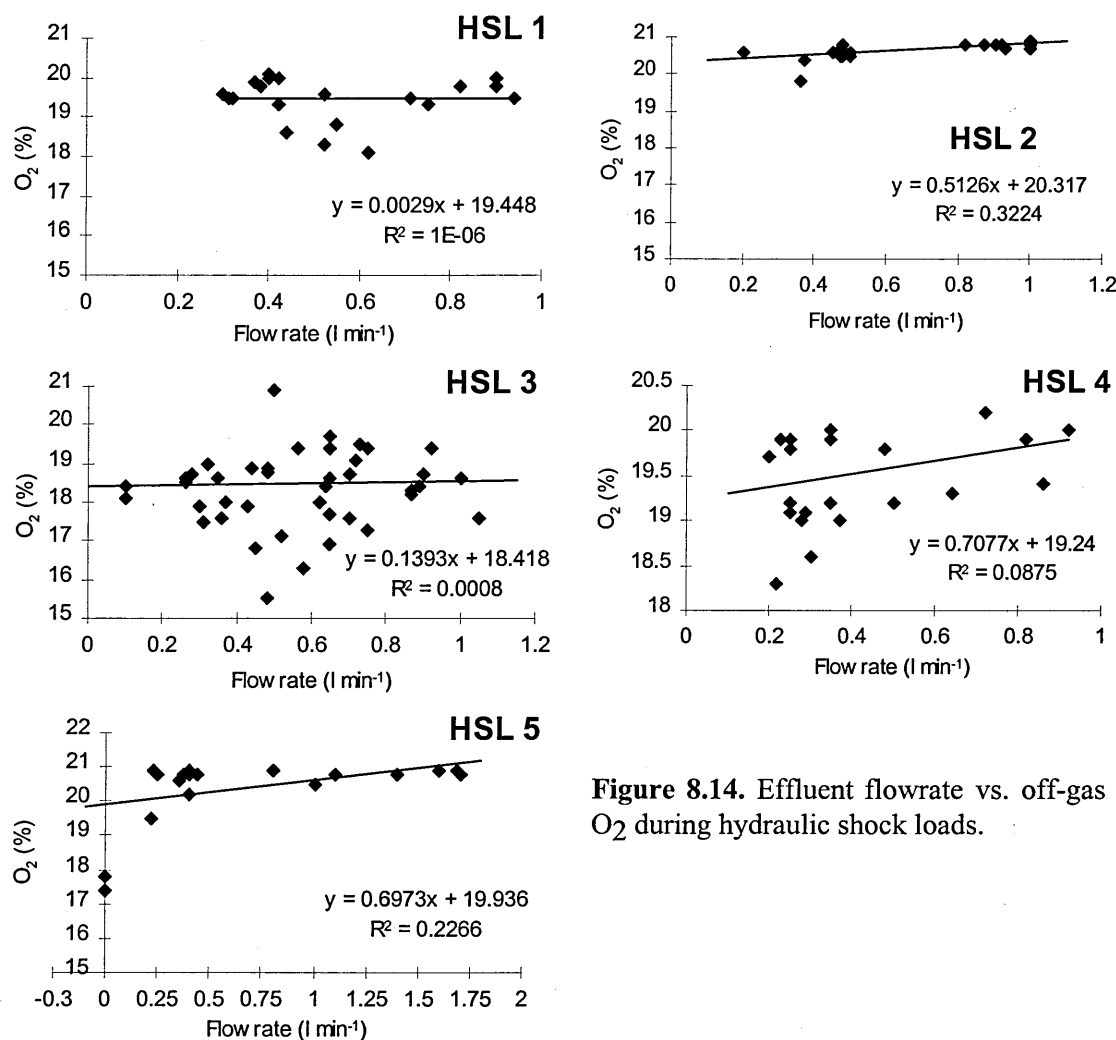


Figure 8.14. Effluent flowrate vs. off-gas O_2 during hydraulic shock loads.

It was not possible to demonstrate that the off-gas concentration was being influenced by the variation in the influent flowrate since no direct correlation between the two parameters was observed.

8.3.3 Oxygen transfer efficiency

It has previously been highlighted that steady-state conditions should prevail during off-gas analysis (Redmon *et al.*, 1983 Harris *et al.*, 1996) and that the reactor has to reach equilibrium after each new air flow. One problem associated with the use of the off-gas analysis during the shock loading period in the present experiment is the lack of steady-state conditions which could have caused unreliable readings. However, off-gas analysis have been forwarded as an on-line method for calculating an observed respiratory coefficient that indicates the COD/TOC ratio of the converted substrate for carbonaceous matter removal in activated sludge plants (Hellings *et al.*, 1996). Using off-gas analysis in full-scale installations would imply that off-gas readings would not be affected by normal fluctuations in the characteristics of the influent. Off-gas readings in the present experiment were undertaken at both steady (before shock load) and non-steady state (during shock load) conditions (Table 8.4).

Table 8.4. Oxygen transfer efficiencies (OTEs) during organic (OSL) and hydraulic shock loads (HSL).

	OTEs before shock load (%)	OTEs during shock load (%)	OTEs after shock load (%)	OTEs during entire run (%)
OSL 1	15.2	20.2	20.2	19.1
OSL 2	15.1	16.6	19.8	18.3
OSL 3	15.4	14.3	13.2	13.8
OSL 4	15.1	11.6	12.4	12.8
OSL 5	11.4	15.6	12.5	12.9
HSL 1	16.6	15.9	14.3	15.4
HSL 2	7.6	8.0	9.0	8.5
HSL 3	20.3	22.2	23.1	22.3
HSL 4	12.5	13.1	16.5	14.1
HSL 5	8.2	7.7	9.6	8.5

There appears to be very little difference between OTE values before, during and after the application of the shock loads. In some cases, the OTE increases during the shock load (OS L 1, OSL 2, OSL 5, HSL 2, HSL 3 and HSL) yet, it decreases during the other shock loads. After the application of the shock loads the same OTE results are

obtained, the OTE values decrease or increase depending on the experiment but it is not possible to find a clear pattern for these findings.

Although high organic loading rates increase the OTE in BAFs due to the increased oxygen demand (Harris *et al.*, 1996), the OTEs found in the present experiment differed only slightly during organic shock loads. The organic loading to the BAF increased slightly but progressively from OSL 1 to OSL 5 (Figures 8.15-8.19) and this could have been reflected in the fact that the OTE during the entire run tended to decrease from 19.1 to 12.9 (Table 8.3). However, the OTE values presented by Harris *et al.* (1996) were obtained in a BAF achieving both BOD and ammonia removal. When the organic loading was increased, ammonia removal stopped causing a decrease in the overall oxygen consumption and thus increasing the OTE. The reactor in the present experiment was achieving a constant yet small degree of ammonia removal (10%) during the entire organic shock loadings. A similar decrease in the overall oxygen consumption as explained by Harris *et al.* (1996) did not happen and the OTE values remained similar. Therefore, in the present experiment it seems that by increasing the organic loading rate the OTE values decreased, contrary to what has been reported before (Harris *et al.*, 1996). Yet, the fact the OTE for the whole runs decreased could not be associated to this and could have been caused by other means as discussed below.

During the hydraulic shock loading experiments there was a greater variation in the OTE measurement as the overall percentage ranged from 8.5 to 22.3%. It is not known why these differences were detected. The OTE observed during the actual hydraulic shock load period and the overall OTE for each run was very similar, suggesting that it was not affected by an increase in hydraulic velocity. It is not possible to fully explain the differences in the OTE values from between shock loads, in particular the low values observed during hydraulic shock loads 2 and 5. In both cases it can be observed (Figures 8.4b and e) that the percentage of O₂ in the off-gas during the shock load was less scattered than during the other hydraulic shock loads (Figures 8.4a, c and d). During hydraulic shock load 2 the influent concentration of sCOD was the highest of

all runs but the percentage of removal remained at similar levels than other runs (Table 8.4). As mentioned earlier, greater organic loads increase microbial activity within BAFs, causing an increase in the consumption of oxygen and thus improving transfer efficiency. However, during hydraulic shock load 2 the OTE did not increase but decreased. The OTE values for the present experiment were larger than those found previously in BAFs which ranged between 4 - 13% for low and high loadings (Harris *et al.*, 1996).

During the hydraulic shock loads once again the fraction of oxygen consumption associated with the DO increase was very small (Figure 8.6). Although it has been suggested that nitrification can be enhanced at high air flow velocities (Tschui *et al.*, 1994) and that increases in the water velocity might have a positive effect on the overall oxygen mass transfer rate in BAFs (Peladan *et al.*, 1997), the short term application of high velocity in the present experiment adversely affected the nitrifying population of the BAF and no $\text{NH}_3\text{-N}$ removal was achieved. This is in agreement with previous work on three stage BAFs that concluded that short-term flowrate fluctuations directly affected the removal of organic carbon which indirectly affected nitrification as the breakthrough COD entered the nitrification reactor (Chudoba *et al.*, 1998). Due to the dilution of the wastewater during the hydraulic shock load period, the oxygen consumption rate associated to BOD_5 removal remained fairly constant during the whole experimentation period. Unfortunately, once again it was not possible to find any clear pattern between the oxygen consumption registered by the off-gas analysis and that associated with the increase in DO and the removal of ammonia and biodegradable carbonaceous matter.

8.3.4 BAF recovery during organic shock loads

During all the shock loads, the concentration of sCOD in the effluent started to increase between τ values of 0.6 and 1.3 with the maximum at τ values between 1 and 1.7 (Figure 8.15-8.19). Maximum concentration breakthrough appeared during shock load 3 (Figure 8.17), when the effluent concentration reached more than 200 mg l⁻¹. However, in most cases the maximum effluent sCOD concentration did not exceed 150 mg l⁻¹. In all cases the effluent concentrations of SS and sCOD returned to normal within 2τ .

These results show that the reactor suffered an increase in effluent concentration during an approximate 60 min period. Due to the fact that the section of the reactor above the media was completely mixed, the high organic concentrations during shock loading (although diluted) were present immediately at the beginning of the media section. In general, the greater effect of the organic shock load was detected at τ approximately 1 (63 min) and it took another 63 min for the effects of the shock load to dissipate. The performance recovery took the same time as the shock loading.

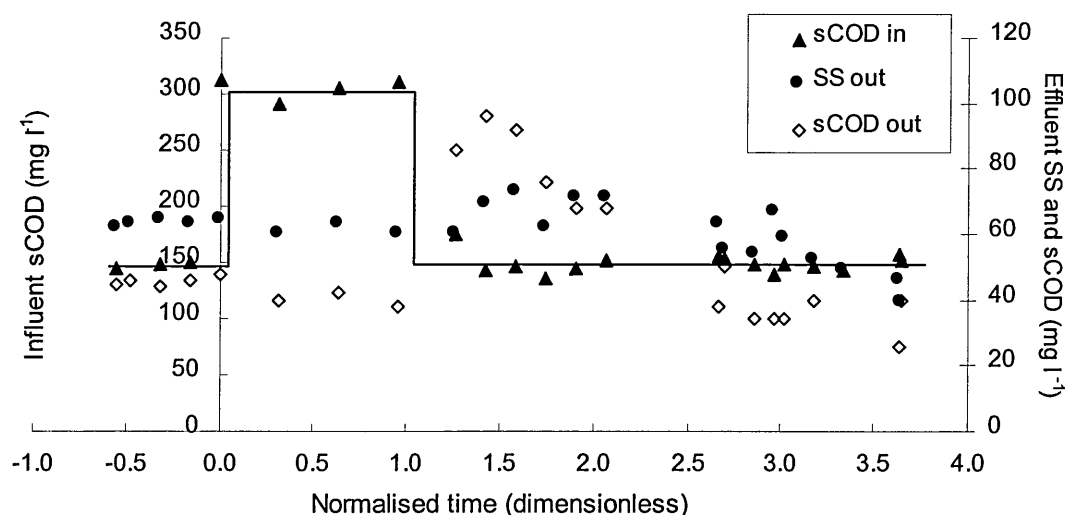


Figure 8.15. Influent and effluent soluble COD and effluent SS variation during organic shock load 1 (OSL 1).

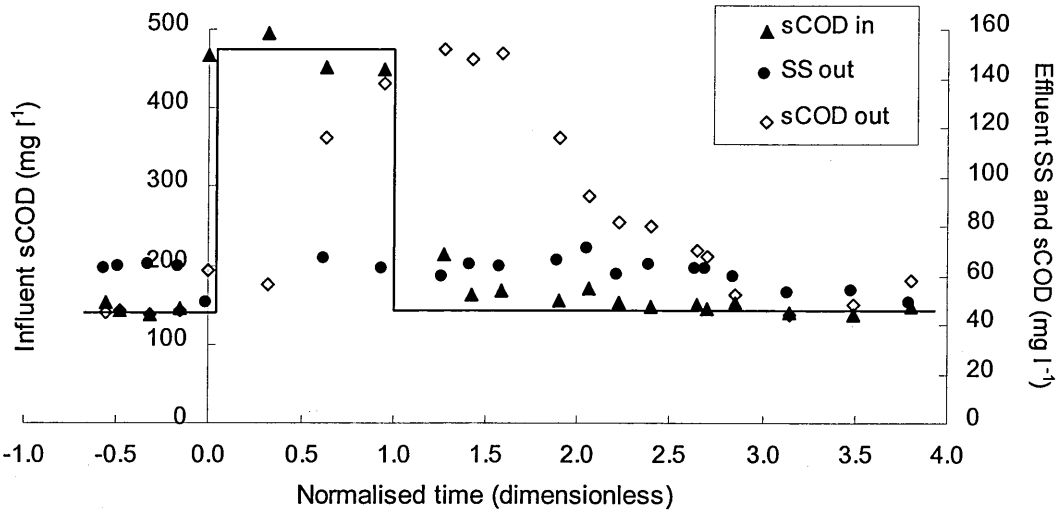


Figure 8.16. Influent and effluent soluble COD and effluent SS variation during organic shock load 2 (OSL 2).

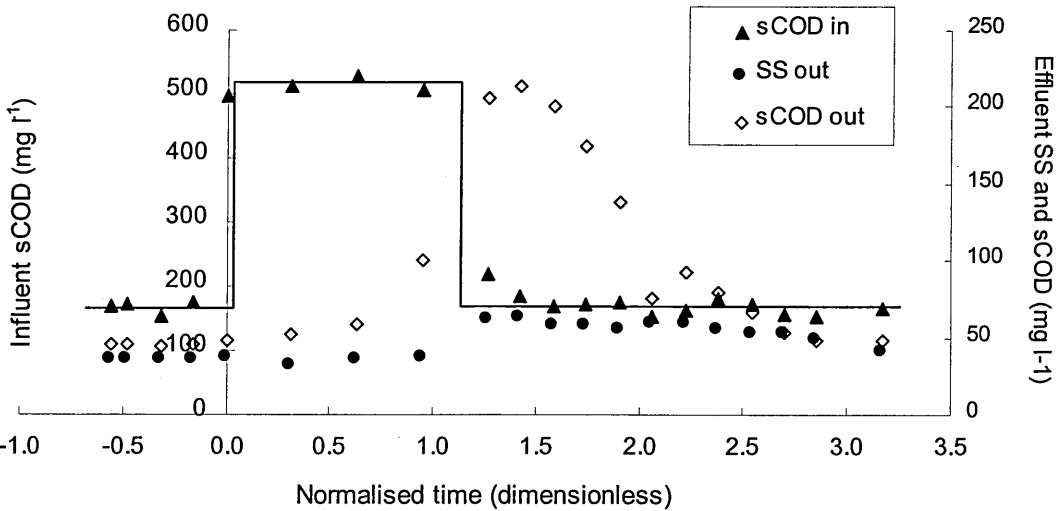


Figure 8.17. Influent and effluent soluble COD and effluent SS variation during organic shock load 3 (OSL 3).

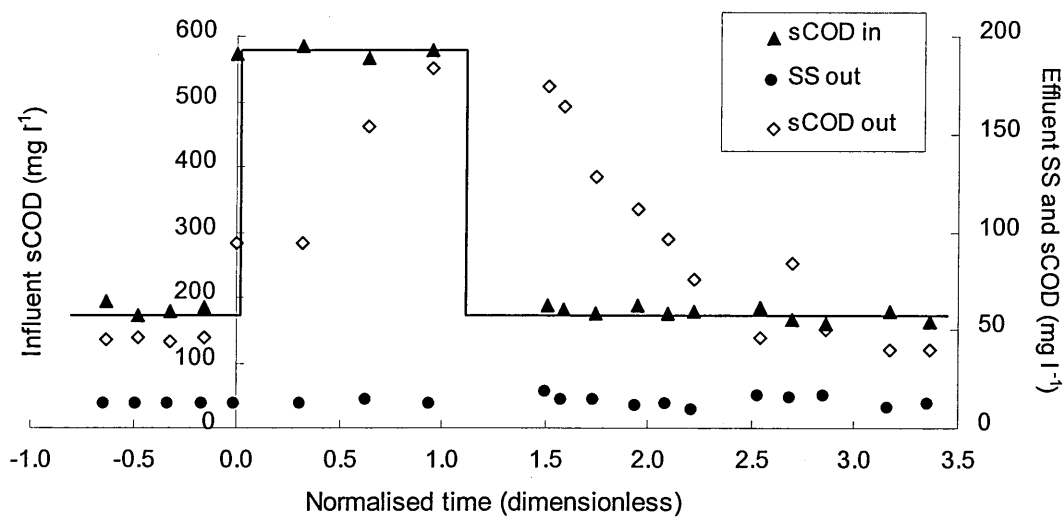


Figure 8.18. Influent and effluent soluble COD and effluent SS variation during organic shock load 4 (OSL 4).

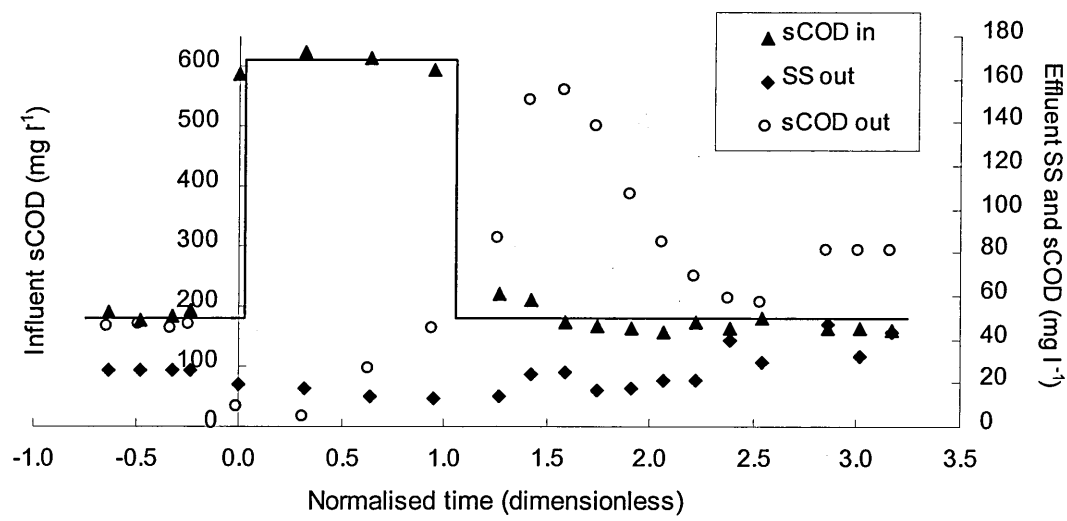


Figure 8.19. Influent and effluent soluble COD and effluent SS variation during organic shock load 5 (OSL 5).

In all runs, the reactor achieving very similar percentages of removal of sCOD before the shock loads were applied, when $\tau < 0$, (Table 8.5). During the actual application of the shock loads, the removal appeared to increase but once the high organic loading traveled along the reactor ($1 < \tau < 2$) the removal dropped dramatically. Between τ values 2 and 3 the BAF was still recovering and when $\tau > 3$ it returned to normal removal values.

Table 8.5. Amount of sCOD in mg l^{-1} being removed and efficiency of removal (in brackets and italics) before, during and after organic shock loads (OSL).

	OSL 1	OSL 2	OSL 3	OSL 4	OSL 5
$\tau < 0$	103 (71)	100 (71)	122 (70)	137 (72)	140 (72)
$0 < \tau < 1$	263 (87)	373 (81)	446 (88)	444 (77)	584 (97)
$1 < \tau < 2$	65 (46)	34 (9)	-3 (-24)	39 (10)	61 (21)
$2 < \tau < 3$	105 (69)	80 (52)	94 (54)	102 (56)	98 (57)
$\tau > 3$	121 (78)	92 (69)	117 (68)	131 (75)	83 (50)

The reactor appeared to present a limited resistance to organic shock loads yet its recovery was fast. It has been suggested that the response of the fixed-film systems to organic shock loadings will depend on the growth rate and yield of the microorganisms involved, the organic loading variations and the duration of the shock load (Wilderer *et al.*, 1993). The quick recovery of the BAF during organic shock loads could suggest the effect of the shock loads on the biomass was limited, probably due to its good acclimatisation to the substrate. In order to achieve continuously efficient secondary treatment of wastewater under variable nutrient loading in BAFs, buffer tanks or variations in the number of operating filters should be considered as highlighted previously by Le Tallec *et al.* (1997).

8.3.5 BAF recovery during hydraulic shock loads

The influent was diluted in order to keep a constant organic loading during the hydraulic shock load experiments, thus simulating storm conditions. The increase in the influent flowrate did not affect the performance of the system (Figures 8.20-8.24) the way increases in the organic loading did. In Figures 8.20 to 8.24, time was not converted into dimensionless time (t^*) because tHRT would have varied once the hydraulic flowrate was increased during the shock load application. It was thought that by presenting the variation of SS and sCOD in the effluent against the actual running time of the experiment the performance of the BAF would be clearer.

When the hydraulic flowrate was increased at 3 times greater than normal (HSL 5), the performance of the BAF did not deteriorate; the effluent concentrations of SS and sCOD reached values of 60 and 30 mg l⁻¹, respectively (Figure 8.24). Once the shock load was over, effluent values returned to the levels lower to those observed previous to the shock load application.

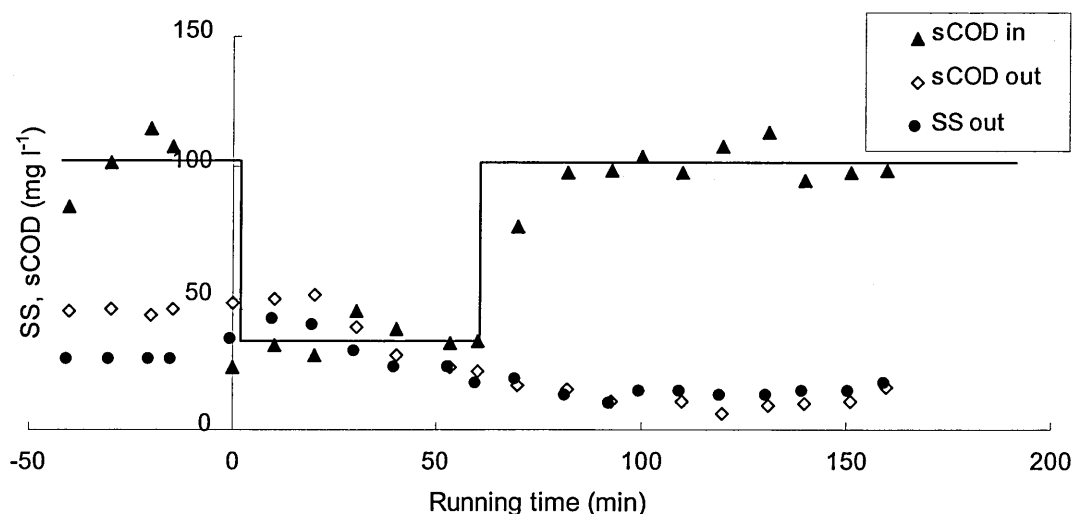


Figure 8.20. Influent and effluent soluble COD and effluent SS variation during hydraulic shock load 1 (HSL 1).

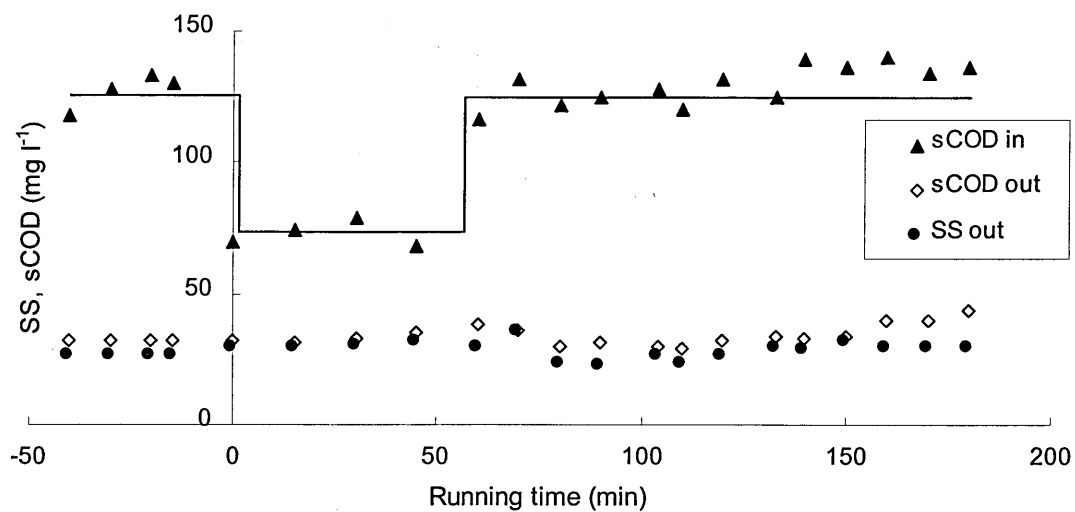


Figure 8.21. Influent and effluent soluble COD and effluent SS variation during hydraulic shock load 2 (HSL 2).

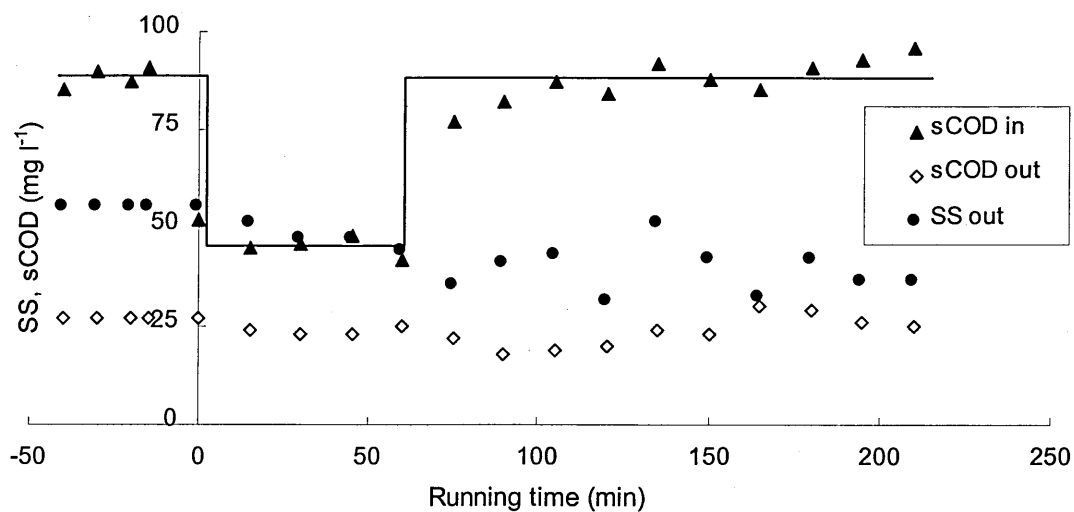


Figure 8.22. Influent and effluent soluble COD and effluent SS variation during hydraulic shock load 3 (HSL 3).

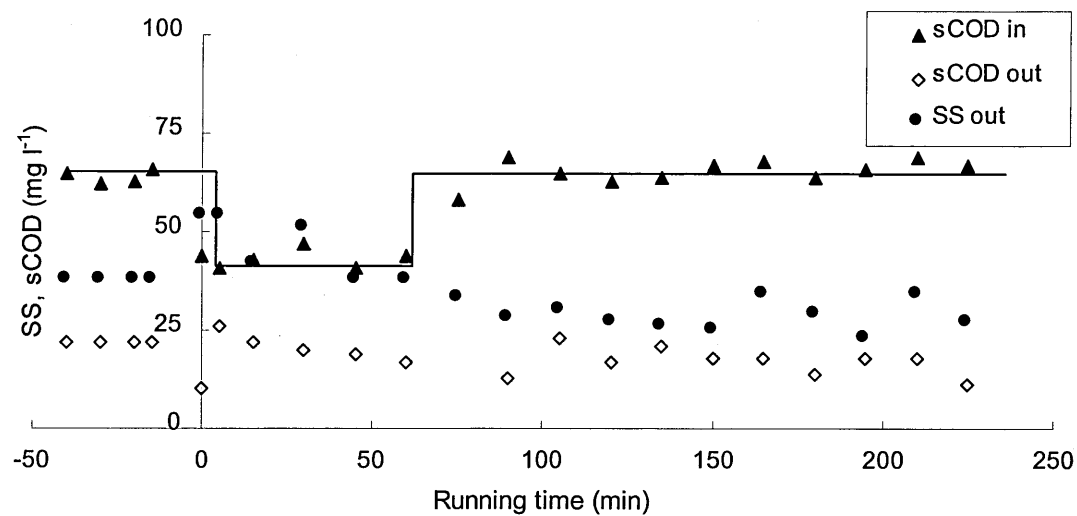


Figure 8.23. Influent and effluent soluble COD and effluent SS variation during hydraulic shock load 4 (HSL 4).

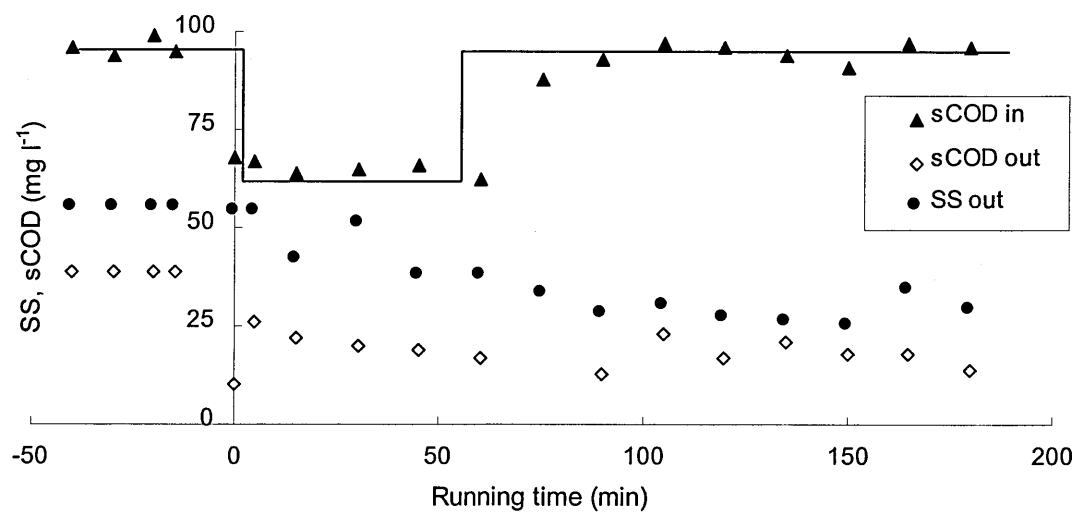


Figure 8.24. Influent and effluent soluble COD and effluent SS variation during hydraulic shock load 5 (HSL 5).

Due to the fact the organic loading during the hydraulic shock loads was not increased, the percentage of sCOD removal was only slightly reduced during the actual application of the hydraulic shock loads (Table 8.6). After each shock loads the percentage of removal was greater than obtained prior to the change in hydraulic conditions. The effluent concentration of sCOD did not exceed 55 mg l⁻¹. Investigations on ratio between BOD₅ and COD of the settled domestic sewage used during the present experiments revealed that the sCOD:tCOD after biological treatment was 1:2. Likewise, the tCOD:BOD₅ ratio after biological treatment was 5:1. Hence, the maximum effluent concentration of BOD₅ in the effluent would be approximately 22 mg l⁻¹. According to Droste (1997), most of the sewage treatment plants in the UK need to meet a discharge consent 11-20 mg l⁻¹. Therefore, the BAF in the present study would have marginally failed the discharge consents when hydraulic shock loadings were applied. When the hydraulic shock load was 3 times the normal operating flowrate the sCOD removal was not adversely affected (Figure 8.24). In terms of suspended solids removal during the application of the shock loads the effluent concentration increased though it returned to concentrations similar to those obtained prior to the shock load almost immediately after normal flowrates were resumed.

Table 8.6. Amount of sCOD in mg l⁻¹ being removed and efficiency of removal (in brackets and italics) before, during and after hydraulic shock loads (HSL).

	HSL 1	HSL 2	HSL 3	HSL 4	HSL 5
Before shock load	57 (36)	95 (75)	61 (69)	42 (63)	57 (55)
During shock load	1- (-56)	48 (54)	22 (46)	24 (58)	46 (72)
0-60 min after shock load	87 (83)	95 (76)	63 (77)	52 (70)	80 (80)
61-120 min after shock load	90 (84)	98 (71)	63 (70)	48 (70)	77 (79)
>121 min after shock load	-	-	69 (71)	52 (73)	-

Previous BAF experiments using wastewater with low suspended solids concentration as a feed have demonstrated that high liquid velocities increase mass transfer rates, resulting in greater percentages of suspended solids, carbonaceous matter and ammonia removal (Peladan *et al.*, 1997). Although high water velocities promote the even

distribution of solids across the media bed (Peladan *et al.*, 1997), at suspended solids concentrations generally encountered in domestic primary settled sewage (150 mg l^{-1}) (Droste, 1997) there is a potential for the media bed to block. In the present study this problem did not occur because the actual hydraulic shock load was applied for a short period of time (60 min). Hence, the reactor was not affected by an increase in the operating hydraulic velocity, though it could be possible that an increase in the period of application of such conditions would cause deterioration of BAF performance leading to eventual blockage if the backwashing regime for the reactor is not modified.

8.4 Conclusions

- The off-gas analysis technique did not appear to be useful for monitoring the impact of organic and hydraulic shock loads on the biological population of the BAF.
- The operational oxygen transfer efficiencies (OTEs) of the system were calculated based on the oxygen off-gas concentrations. Increases in the organic and hydraulic loading to the BAF did not appear to have a significant impact on the OTE as other studies suggest.
- The BAF used for the present experiment was not capable of absorbing high peak organic shock loads when the organic loads were increased from a normal load of 1.2-1.4 kg sCOD m⁻³ d⁻¹ to shock loads between 5.1- 7.3 kg sCOD m⁻³ d⁻¹, based on empty bed volumes. Nevertheless, the effect on the biomass was limited as normal performance resumed very quickly.
- Increases in the hydraulic velocity from 0.7-1.0 m h⁻¹ to 1.5-2.9 m h⁻¹ had little affect on BAF performance in terms of soluble COD removal. Soluble COD effluent concentrations of <40 mg l⁻¹ (~16 mg l⁻¹ BOD₅) were detected thus, meeting most discharge consents in UK sewage works. The recovery of the reactor in both cases was fast, resuming normal performance within 60 min after the application of the shock load.

Chapter 9

DISCUSSION

DISCUSSION

The study of biological aerated filters (BAFs) has reached a point in which the nature of the process is mostly understood. Despite the great number of proprietary systems in the market, there are certain areas in which it appears that few published results are available (Chapter 3). It has been several years since BAFs were proposed as an alternative process to conventional systems such as trickling filters and activated sludge (Faup *et al.*, 1982). Since then, BAFs have been optimised in several ways, mostly in terms of organic loading and removal capacity (Smith and Hardy, 1992). The recent reports on so-called high-velocity operating BAFs (Peladan *et al.*, 1997) could imply that in terms of organic and hydraulic loadings, BAFs are reaching their full potential. The normal hydraulic operating conditions of these reactors ($30 \text{ m}^3 \text{ m}^{-2} \text{ h}^{-1}$) are now very close to those hydraulic flowrates used only during backwashing in normal BAFs (Wheale and Cooper-Smith, 1995). Peladan *et al.* (1997) and Pujol *et al.* (1998) suggested that high water velocities promote better distribution of substrate and biomass within BAFs and that there is a reduction in the overall mass-transfer coefficient between the biofilm and the substrate (ammonia), yet this point needs confirmation. Boller *et al.* (1997) concluded that high rate floating media BAFs operated in upflow mode cannot be considered as an effective final effluent polishing step while sunken media BAFs operated in downflow mode can. If sunken media was to be used, the high velocities would cause fluidisation of the bed and therefore the reactor would be operating more as a fluidised-bed reactor than as a fixed-media (Chung *et al.*, 1996) system. Nevertheless, enhancement of nitrification at high hydraulic velocity is still to be confirmed by other studies and an explanation for such increase to be found. It is therefore possible that the increase of the loading to BAFs has reached saturation point and that in order to enhance substrate removal performances other aspects such as reactor configuration, modelling of the process and resistance to shock loadings have to be optimised.

One of the most important aspects in chemical and biological reactors is the characterisation of their flow pattern (Boller *et al.*, 1997). The type of flow in the

reactor and the flowrate will determine the actual hydraulic retention time (HRT) of the system. This means that it will dictate the actual time that the liquid is inside the reactor and thus, the time in which it is potentially treatable. Moreover, in fixed-film systems the hydrodynamics of reactors affect the physical and physiological structure of the biofilm (Arvin and Harremoes, 1990; Wilderer *et al.*, 1995). High increases in the water velocities can cause high erosion of the biofilm by increasing the shear stress on the biofilm and thus reducing reactor efficiency (Lewandowski and Walser, 1991). On the other hand, at low liquid velocities there is little erosion of the biofilm but the resulting nutrient limitation could cause low growth rate of the biofilm (Lewandowski and Walser, 1991). Moreover, in order to scale-up results from pilot-scale to full-scale systems, characterisation of the hydraulic characteristics of the systems is essential (Eckenfelder and Quirk, 1983).

The finding that under normal operating conditions plug-flow prevailed in granular media upflow and downflow BAFs (Chapter 4) agreed with previous work by Mann *et al.* (1995) and Fdz-Polanco *et al.* (1996). However, Boller *et al.* (1997) found that BAFs could present different degrees of mixing depending on the amount of channelling and backmixing. The initial presence of a plenum chamber at the bottom of the downflow BAF in the present study adversely affected the hydraulic characteristics of the reactor and almost produced completely mixed conditions. This is in agreement with a study by Seguret and Racault (1998) who suggested that aeration had a significant role in the mixing pattern in the lower part of a full-scale floating media BAF running in upflow mode. The rise of air bubbles created convection currents in the lower part of the filter, affecting the transport of particles from the centre to the walls. According to Boller *et al.* (1994), the air flow through the filter media in BAFs could lead to channel preferential flow and backmixing causing heterogeneous concentration of substrate in a given cross-sectional area and subsequent variability of the conditions at the biofilm surface. Bubble entrapment within the media can reduce the working volume in BAFs (Seguret and Racault, 1998). According to Pearce (1996), in upflow BAFs liquid and air flow in the same direction so variation of a parameter does not necessarily affect the other. In downflow systems

however, at high liquid flowrates and low aeration rates there could be a hold-up of the air (Pearce, 1996). Boller *et al.* (1997) found that in downflow BAFs used for secondary treatment of wastewater, the retention of solids and air by the media increased the head-loss along the reactor during the filter run. There was partial clogging of the top section of the reactor which caused higher resistance to the flow of air, thus causing greater gas hold-up at the top. Floating media reactors operated in upflow mode have a more moderate head-loss increase throughout the whole filter run even at higher hydraulic loading rates than sunken media reactors (Boller *et al.*, 1997). In the present experiment by eliminating the plenum chamber, the channelling and stagnant zones in the downflow reactor disappeared and plug-flow conditions prevailed.

Tracer studies indicated that after modifications to the reactors, the aeration rate had no real effect on the hydrodynamic characteristic of the system. It seemed that due to its high density (2500 kg m^{-3}) the media used in the present experiment remained relatively undisturbed by the different air flowrates tested. Studies with low density media have found the contrary (Mann *et al.*, 1994). The upflow and downflow BAF presented the same flow pattern and HRT, which was necessary to allow for the direct comparison of operational performances of the BAFs.

The performance of two sunken media pilot-scale biological aerated filters for soluble carbonaceous matter and ammonia removal were compared (Chapter 5). The BAFs used identical media and primary settled sewage was used as a feed. The upflow and downflow reactors achieved almost identical efficiencies of removal and removal rates for SS, soluble COD and total COD. There was a small difference in terms of ammonia removal, where the downflow reactor achieved 42% removal and the upflow reactor reached 29%. Ammonia concentration profiles (Figures 5.11 and 5.12) were undertaken in both reactors in order to indirectly locate the biological population with autotrophic activity. Fdz-Polanco *et al.* (1996) indicated that BAFs operated under plug-flow conditions could produce zones of different nutrient strengths and thus allowed for the growth of different bacterial species which are necessary conditions for

combined soluble carbonaceous matter and ammonia removal. It appeared that this was the case in the present experiment as autotrophic bacteria were found at the bottom of the reactor in the downflow column and at the top in the upflow unit. There are several explanations for this as nitrification (and the activity of autotrophic bacteria) can be inhibited by a number of ways, mostly due to substrate and/or oxygen limitation (Boller *et al.*, 1994). Cecen and Gonenc (1995) highlighted that the transition between ammonia (substrate) limitation to oxygen limitation occurred when the ratio of bulk oxygen concentration to bulk ammonia concentration was in the range of 2:4.5. Beg *et al.* (1997) stipulated that, as most of the organic substrate is consumed in the inlet of the reactor, it is likely that this section of the BAFs would be oxygen limited (thus, inhibiting nitrification). However, dissolved oxygen (DO) concentration along both reactors in the present experiment was always $>4 \text{ mg l}^{-1}$ thus indicating that oxygen was not a limiting factor for nitrification (Chen and Cheng, 1994). It is more likely that inhibition of nitrification in the present experiment would have been due to competition with non-nitrifying species, which has been demonstrated by Lee and Welandar (1994). Autotrophic bacteria require some carbonaceous matter for metabolism (Akunna *et al.*, 1994). However, in a system with a dynamic population of heterotrophic and autotrophic bacteria, the fast-growing heterotrophic bacteria will always outcompete the slow-growing autotrophic population at high concentrations of carbonaceous matter (Al-Haddad *et al.*, 1991; Akunna *et al.*, 1994). In the present experiment autotrophic bacteria were found at the bottom of the reactor in the downflow column and at the top in the upflow reactor in sections away from the inlets, where the concentration of carbonaceous matter would have decreased due to prior heterotrophic activity. Wilderer *et al.* (1993) found this same distribution of autotrophic population in his studies on an upflow continuous flow biofilm reactor, indicating that plug-flow conditions favoured the stratification of species.

The fact that in the present experiment the downflow reactor achieved a slightly better removal of ammonia than the upflow BAF is in accordance to finding by Newbigging *et al.* (1995). The authors suggested that the operation of plug-flow BAFs in downflow mode reduced intermixing which subsequently reduced competition between

autotrophic and heterotrophic microorganisms; thus allowing for more efficient BOD and ammonia removal single-step reactors. There is, however, another possible explanation for such a difference. In the present experiment, the backwashing of the sunken media reactors was carried out by introducing air and water and high velocities at the bottom of the reactors, regardless of the direction of the flow at which the influent was introduced. As the autotrophs were located at the top of the upflow BAF, they could have been washed out. Moreover, it is likely that the backwashing sequence would have pushed some heterotrophs to the top of the reactor thus promoting the mixing of the autotrophic and heterotrophic populations which has caused reductions in the treatment efficiency in other studies (Wilderer *et al.*, 1995). In order to avoid direct competition between heterotrophic and autotrophic bacteria, a series of reactors in a multi-stage system could be introduced. Each reactor would be substrate specific with either heterotrophic or autotrophic bacteria resulting in optimised growth for each population, leading to enhanced reaction kinetics and reactor performance (Chudoba and Pujol, 1998).

Mechanistic modelling of BAFs is very complex due to the many variables involved in the system. Sampling of the biofilm without disturbing reactor performance and the rest of the biofilm is extremely difficult to achieve (Gjaltema and Griebe, 1995). In contrast, empirical models might be less accurate than mechanistic models but are based on parameters that can be easily measured, directly relating input variables such as organic loading or substrate concentration to output variables such as substrate removal rate. Due to similarities in their hydraulic profile (plug-flow conditions), empirical models developed for trickling filters have been successfully adapted for BAFs (Mann and Stephenson, 1997). In Chapter 6 the development and calibration of a simple empirical model for BAFs was undertaken. The media and process constants for the BAF were compared to those obtained previously (Mann and Stephenson, 1997). This comparison allowed to determine that the media and the hydraulic configuration of the system should be taken into account and are important for the performance of BAFs. The soluble COD removal efficiency in the present experiment was lower than that reported by Mann and Stephenson (1997) for the floating media

BAF running in upflow mode. In the latter, it appeared that the compression of the bed due to the buoyancy force of the media in combination with the flow and air acting upward promoted the growth of bacteria and improved the removal of carbonaceous matter (Mann *et al.*, 1998).

Only recently has there been attempts to try to incorporate the removal of suspended solids into the modelling of biological aerated filters (Jacob *et al.*, 1997). The generalised model by Jacob *et al.* (1997) took into account the classic biological processes (biofilm characteristics) and chemical engineering (three-phase reactor) approaches as well as a physical processes approach by means of incorporating a depth filtration term. However, this last term could not be validated, as the feed water used for such purposes did not have any suspended solids. Although Toettrup *et al.* (1994) successfully applied kinetic expressions developed in rotating biological contactors (RBCs) onto small-scale ($0.2 \times 10^{-3} \text{ m}^3$ of media) and large-scale (30,000 population equivalent) nitrifying and denitrifying BAFs, the use of kinetic parameters from the International Association on Water Quality (IAWQ) Activated Sludge Model 1 (Henze *et al.*, 1987) caused errors in the validation of the biological part of the model developed by Jacob *et al.* (1997). Hence, it was demonstrated that although it is possible to include several approaches when modelling BAFs, the complexity involved in the validation and application of such models make them unlikely to be used.

Therefore, once again simple empirical models have the potential for being used as useful tools for BAF design. A model initially developed for deep bed filters (Gimbel 1989) was successfully used for modelling the removal of SS and tCOD in biological aerated filters of different sizes when the appropriate constants were used (Chapter 7). The total COD of a wastewater is composed of the soluble and the particulate fraction (Henze *et al.*, 1988). Suspended solids comprises a large fraction of the particulate fraction of primary settled sewage thus, suspended solids and total chemical oxygen demand removal are strongly associated. As the model for deep bed filters was effective in predicting SS removal, it was also effective in predicting in tCOD due to the association between these two parameters.

As different wastewaters may have different fractions of particulate and soluble carbonaceous matter, the use of total COD for design purposes might be misleading. Therefore, soluble carbonaceous matter (sCOD) should be used instead (Stensel *et al.*, (1988). Thus, in the present experiment (Chapter 7) the removal of soluble COD was modelled using two empirical models based on first-order rate models for trickling filters and the Eckenfelder model itself (Eckenfelder, 1961). All three models gave good approximations to the overall performance of the reactors when the corresponding constants were used that is, when the constants from the small column were used to model the small column and when those obtained for the large BAF were used to predict removal in the large BAF. Model 3 gave the closest values to those observed and took into account parameters such as gas and liquid velocity. This indicated that the empirical models although simple, are very effective if the adequate constants are included.

Scaling-up of biofilm systems can be undertaken in a number of ways. According to Eckenfelder and Quirk (1983), the hydraulic loading, film thickness, surface of packing medium and the hydraulic characteristics of the system to be scaled-up should remain constant. Horvath and Schmidtke (1983) proposed the theory of similarity for scaling-up water and wastewater treatment systems. The theory stipulated that geometric, kinematic, dynamic, chemical, biological and/or thermal similarity should exist between systems of different scales. Later, Manem and Rittmann (1989) agreed that the scaling-up procedure for biofilm reactors should be based on the principle of similitude. Advantages of using the similitude criteria are that when these are made equal, the reactors of different scale can have the same performance even if all mechanistic parameters are not known and full kinetic models have not been solved (Manem and Rittmann, 1989). According to the authors, the same type of reactions and microorganisms, the same effluent concentration of the rate limiting substrate and the same temperature must exist between both reactors. From the three models for sCOD, only the simplest of them all, Model 1 (the Eckenfelder model, Eckenfelder, 1961) was effective for scaling-up purposes. The model assumed that soluble COD removal

followed first-order kinetics but overlooked other parameters that influence reactor performance like media characteristics, gas and liquid flowrates (velocities). Model 4 appeared to have been more effective in predicting the removal of suspended solids than the removal of total COD when the constants from the small pilot-scale BAF were used in the large pilot-scale column. Nevertheless, the prediction of suspended solids removal did not appear to be as accurate as the prediction for sCOD removal using Model 1. Suspended solids removal, being undertaken primarily by physical mechanisms in BAFs (Ryhiner *et al.*, 1994), could have been more affected by the hydraulic configuration of the system than sCOD removal, which is more dependent on biological mechanisms. Although the diameter of the reactors was at least 50 mean particle diameters in size to limit wall effects (Lang *et al.*, 1993), due to the smaller dimensions of the small BAF, wall effects could have had a greater impact on its SS removal efficiency. Therefore, the application of the constants from the small BAF onto the large BAF could have been unsuccessful due to lack of geometric similarity, as mentioned by Horvath and Schmidtke (1983). In terms of soluble COD prediction, Models 2, 3 and 4 that did take into account parameters such as differences in the concentration of the substrate in the inlet and/or the hydraulic loading to the reactors (in terms of hydraulic velocity) were not successful in scaling-up the results. These parameters would directly affect the hydraulic and organic loading rates to the reactors. As the loadings to the large and small BAF were different (Tables 7.3 and 7.6), there was a lack of kinematic similarity between the systems which could have affected negatively the prediction of Model 2, 3 and 4.

The performance of a biological aerated filter (BAF) during organic and hydraulic shock loads was monitored and off-gas analysis of O_2 was undertaken to try to establish a correlation between off-gas monitoring and BAF performance (Chapter 8). Aeration efficiency optimisation has been highlighted as an area in which further study should be undertaken in order to decrease operational costs of BAFs (Chapter 3). Off-gas analysis has been proposed as a reliable means of monitoring and controlling aeration rates in activated sludge systems (Redmon *et al.*, 1983) and its potential for BAF application was mentioned by Stensel *et al.* (1988) and Harris *et al.* (1996).

Stensel *et al.* (1988) found that oxygen consumption determined from off-gas analysis were close to those results obtained from a simple empirical equation relating oxygen utilisation and sludge production with influent soluble and total BOD and influent SS/total BOD ratios. According to Hellinga *et al.* (1996), off-gas analysis has the advantage over measurements in the liquid phase in providing a reading of a well-mixed sample of gas that leaves the reactor, so the reading provides a measurement of the overall reactor performance while a probe in the liquid phase is only a “point measurement”.

Off-gas analysis of oxygen (O_2) was undertaken in the present work to try to establish a correlation between off-gas monitoring and BAF performance for the secondary treatment of wastewater during unsteady-state conditions. The operational oxygen transfer efficiencies (OTEs) of the system were calculated based on the O_2 off-gas concentrations. Increases in the organic and hydraulic loading to the BAF did not appear to have a significant impact on the OTE as other studies suggest (Harris *et al.*, 1996). Pearce (1996) reported that OTEs in a small pilot-scale BAF increased at increasing organic loadings. This pattern was attributed to interfacial transfer directly between air bubbles and the biofilm (biomass). Pearce (1996) found that increases in the hydraulic loading to a downflow BAF did not increase the OTEs and attributed this phenomenon to possible channelling of air through preferential paths across the media and the reactor.

Although some authors mention that during off-gas analysis steady-state conditions should prevail (Redmon *et al.*, 1983 Harris *et al.*, 1996), Hellinga *et al.* (1996) highlighted the potential of the use of off-gas measurements for monitoring overall oxidation activity and therefore their ability for being used as on-line sensors. However, off-gas analysis did not appear to be a useful tool for monitoring the impact of organic and hydraulic shock loads on the biological population of the BAF in the present experiment. The oxygen transfer efficiencies (OTEs) obtained in the present study are higher than those previously obtained in BAFs (Stensel *et al.*, 1988) and close to those obtained for aeration systems using air aeration ceramic dome diffusers

(25-37 %), jet aerators (15-24 %) and non-porous diffusers (12-13 %) (Huibregtse *et al.*, 1983; Yunt *et al.*, 1980). The values obtained in the present study fit within those values reported by Paffoni *et al.* (1990) and Rogalla and Sibony (1992) (7 to 15% during tertiary nitrification). The authors stipulated that OTEs in BAFs were higher than those found in activated sludge tanks with similar liquid depths (Benefield and Randall, 1980) probably due to an increase in the air bubble detention time caused by the media. Moreover, the time of contact between bubbles and biofilm in a BAF could be increased due to the physical retention by the media thus, improving OTEs (Lee and Stensel, 1986). In the present study the OTE increases during some hydraulic and organic shock loads and during others it decreases, so there does not seem to be a pattern for the variation of OTE.

According to Young and Stewart (1979), fixed-film reactors perform better than suspended growth reactors during shock loadings of primary settled sewage. Stable operation, long retention of biomass solids and ability to cope with toxic substances are some advantages of using fixed-film reactors (Hamoda and Al-Ghusain, 1998). According to Young and Stewart (1979), in plug-flow reactors organic shock loads were treated through gradual reduction in concentration. Hamoda and Al-Ghusain (1998) stipulated that properly designed fixed-film processes are generally reliable and resistant to hydraulic and shock loadings effectively due to the large amount of biomass that they can sustain.

However, because BAFs are operated as plug-flow reactors they can be affected by shock loads and toxic or inhibitory substances, as their effects are slowly dispersed (Chen and Cheng, 1994) and high peak loadings may result in organic breakthrough due to limited treatment efficiency. Wilderer *et al.* (1993) highlighted that increases in the feed concentration of plug-flow reactors would cause deterioration in the quality of the effluent due to breakthrough.

In the present investigation, a sunken media pilot-scale BAF running in upflow mode was not capable of absorbing organic shock loads. It was found that there was

breakthrough of soluble organic matter through the system when the organic loadings were increased, which is in agreement with conclusions by Chen and Cheng (1994). Such results could be due to the plug-flow characteristics of the system in the present study (Chapter 4). The response to shock loads in plug-flow reactors would depend on the growth rate and yield of the biological population involved, on the mass loading variations (i.e. how much is being applied) and on the duration of the shock loading conditions (Wilderer *et al.*, 1993). Thus, it appeared that the effect of the shock loads on the biomass was limited probably due to its good acclimatisation to the substrate, the fact that the biomass was not washed out during the high loadings and the hydraulic configuration of the reactor.

When hydraulic shock loads were applied, the BAF performance in terms of soluble chemical oxygen demand (sCOD) removal was not adversely affected. Soluble COD effluent concentrations of $<40 \text{ mg l}^{-1}$ ($\sim 16 \text{ mg l}^{-1} \text{ BOD}_5$) were detected thus, meeting most discharge consents in UK sewage works. Although suspended solids removal deteriorated, recovery of the reactor was fast. Performance of BAFs has been found to be adversely affected by step-increases to their hydraulic (and organic) loading (Mann *et al.*, 1999). Investigations of the impact of peak flows and the response to shock loading in a pilot-scale three-stage BAF plant have concluded that short-term fluctuations in the wastewater flowrate directly affected the removal of organic carbon (Chudoba and Pujol, 1998). Step increases on the hydraulic loading caused breakthrough of organic carbon from the carbon-removal filter to the ammonia-removal unit. The increase in carbon in the nitrifying unit caused competition between heterotrophic and autotrophic bacteria, leading to a decrease in the performance of the filter. Hydraulic shock loads to the system caused an overloading of the first unit (carbon-removal unit) but did not affect the final effluent concentration of SS and COD of the system due to the removal capacity of the final 2 units. In other studies, high hydraulic flowrates application had a negative effect on the COD removal efficiency of a full-scale sunken media BAF running in upflow mode (Pujol *et al.*, 1998).

As mentioned earlier, suspended solids removal was adversely affected by the application of hydraulic shock loads in the present study. Pujol *et al.*, 1998) suggested that high hydraulic conditions allowed for good distribution of solids and substrate and biomass along the filter bed in BAFs. Yet it is more likely that increases in the hydraulic flowrate would have caused an increment in the shear stress, which subsequently caused an increase in biofilm detachment as other studies suggested (Mann *et al.*, 1999), increasing the amount of suspended solids in the effluent.

Chapter 10

CONCLUSIONS

CONCLUSIONS

1. Two pilot-scale BAFs running in upflow and downflow configuration with identical media were found to have plug-flow conditions with a small degree of mixing. Increases in the nominal air (from 0 to 17.9 m h^{-1}) and hydraulic velocities ($1.8\text{--}2.7 \text{ m h}^{-1}$) did not affect the hydrodynamic characteristics of the systems until hydraulic velocities of 3.6 m h^{-1} were applied.
2. At loading rates of approximately $1.8 \text{ kg SS m}^{-3} \text{ d}^{-1}$, $2.4 \text{ kg sCOD m}^{-3} \text{ d}^{-1}$, $5.2 \text{ kg tCOD m}^{-3} \text{ d}^{-1}$ and $0.51 \text{ kg NH}_3\text{-N m}^{-3} \text{ d}^{-1}$ (based on working volumes of 25 l in each reactor), removal efficiencies for SS, soluble and total COD in the upflow and downflow small pilot-scale BAFs were 76, 75 and 79% removal, respectively. The downflow BAF achieved better removal of ammonia than the upflow reactor. Plug-flow conditions and autotroph-heterotroph competition caused autotrophic activity to be located at the bottom of the reactor in the downflow column and at the top in the upflow unit. Co-current air and water backwashing could have washed-out autotrophs in the upflow BAF.
3. A first-order kinetic empirical model for soluble COD removal was used to find media and process constants for a sunken media BAF operated in upflow mode. Comparison of the media constant (n) and the overall process constant of the system (k^*) to previous work with floating and sunken media BAF suggested that the performance of a BAF was greatly influenced the media/liquid flow pattern. Ideally, upflow reactors should operate with floating media and downflow reactors with sunken media.
4. Three empirical models for soluble COD removal and one for SS and tCOD removal were used to predict the removals in a small BAF (0.07 m^3 empty bed volume) and a large BAF (0.69 m^3 empty bed volume). All four models were successful in predicting overall performance of the reactors when the corresponding constants from each BAF were used for prediction (i.e. constants

from small BAF tested on small BAF). For scaling-up purposes, model 1 (a first-order kinetic model developed for trickling filters performance) was the most effective for predicting the removal of sCOD in the large pilot-scale BAF when using constants from the small pilot-scale BAF. The model overlooked parameters that influence reactor performance such as gas and liquid velocities. Models that did take into account such parameters did not produce such good results when used for scaling-up purposes probably due to the lack of kinematic similarity (differences in the organic and hydraulic rates) between the small and large reactors.

5. The off-gas analysis technique did not appear to be useful for monitoring the impact of organic and hydraulic shock loads on the biological population of the BAF. Increases in the organic and hydraulic loading to the BAF did not appear to have a significant impact on the OTE as other studies suggest. The BAF used for the present experiment was not capable of absorbing high peak organic shock loads. when the organic loads were increased from a normal load of 1.2-1.4 kg sCOD m⁻³ d⁻¹ to shock loads between 5.1- 7.3 kg sCOD m⁻³ d⁻¹, based on empty bed volumes. Nevertheless, the effect on the biomass was limited as normal performance resumed very quickly. Increases in the hydraulic velocity from 0.7-1.0 m h⁻¹ to 1.5-2.9 m h⁻¹ had little effect on BAF performance in terms of soluble COD removal. Soluble COD effluent concentrations of <40 mg l⁻¹ (~16 mg l⁻¹ BOD₅) were detected thus, meeting most discharge consents in UK sewage works. The recovery of the reactor in both cases was fast, resuming normal performance within 60 min after the application of the shock load.

Chapter 11

RECOMMENDATIONS FOR FURTHER WORK

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1. The optimisation of aeration by means of monitoring oxygen utilisation through off-gas analyses is an area that requires further work. The present study undertook off-gas analyses on a small pilot-scale BAF operating in downflow mode. It appeared that the measurement of off-gas oxygen was very sensitive and that reliable readings were difficult to obtain. Although the experimental set-up was constantly checked for faults, these were not evident, yet the values of oxygen transfer efficiency (OTE) obtained with the off-gas analysis did not provide useful data to enhance the results from the present study. Further work has to be undertaken in order to establish if this technique could be a useful way of monitoring biomass activity in BAFs.
2. It is recommended to obtain an empirical model that could integrate the removal of suspended solids with organic matter (as BOD or COD). For this purpose, further work on suspended solids removal profiles should be undertaken simultaneously with the monitoring of head-loss development. Moreover, it could be useful to undertake profiles on the particle size distribution in BAFs and also relate them to pressure data in order to have a greater understanding of the solids removal mechanisms in BAFs.

Chapter 12

REFERENCES

REFERENCES

- Adachi S. and Fuchu Y. **Reclamation and reuse of wastewater by biological aerated filter process.** *Water Science and Technology*, 24(9), 1991, 195-204.
- Aesoy, A., Odegaard, H., Bach, K., Pujol, R. and Hamon, M. **Denitrification in a packed bed biofilm reactor (BIOFOR) - Experiments with different carbon sources.** *Water Research*, 32, 1998, 1463-1470.
- Akunna, J., Bizeau, C., Moletta, R., Bernet, N. and Heduit, A. **Combined organic carbon and complete nitrogen removal using anaerobic and aerobic upflow filters.** *Water Science and Technology*, 30(12), 1994, 297-306.
- Al-Haddad, A. A., Zeidan, M. O. and Hamoda, M. F. **Nitrification in the aerated submerged fixed-film (ASFF) bioreactor.** *Journal of Biotechnology*, 18, 1991, 115-128.
- Amar, D., Partos, J., Granet, C., Faup, G. M. and Audic, J. M. **The use of an upflow mixed bed reactor for treatment of a primary settled domestic sewage.** *Water Research*, 20, 1986, 9-14.
- Andersen, K. L., Bundgaard, E., Andersen, V. R., Hong, S. N. and Heist, J. F. **Nutrient removal in fixed-film systems.** IN: Proceedings of the Water Environment Federation 68th Annual Conference, 1995, Miami, USA, 581-590.
- Arvin, E. and Harremoes, P. **Concepts and models for biofilm reactors performance.** IN: Proceedings of the Conference on Technical Advances in Biofilm Reactors, 4-6 April 1989, Nice, France, 191-212.
- Bacquet, G., Joret, J. C., Rogalla, F. and Bourbigot, M. M. **Biofilm start-up and control in aerated biofilter.** *Environmental Technology*, 12, 1991, 747-756.
- Bach, H. **The 'Tank Filter' for the purification of sewage and trade wastes.** *Water Works Sewage*, 1937, 389-393.

- Beg, S. A. and Hassan, M. M. **Effects of inhibitors on nitrification in a packed-bed biological flow reactor.** *Water Research*, 21, 1987, 191-198.
- Beg, S. A., Hassan, M. M. and Chaudhry, M. A. S. **Multi-substrate analysis of carbon oxidation and nitrification in an upflow packed-bed biofilm reactor.** *Journal of Chemical Technology and Biotechnology*, 64, 1995, 367-378.
- Beg, S. A., Hassan, M. M. and Chaudhry, M. A. S. **Effect of fasting condition on multi-substrate carbon oxidation and nitrification system in an upflow packed-bed biofilm reactor.** *Chemical Engineering Technology*, 20, 1997, 162-170.
- Benefield, L. D. and Randall, C. W. Biological process design for wastewater treatment. New Jersey: Prentice Hall, 1980.
- Boller, M., Gujer, W. and Tsuchi, M. **Parameters affecting nitrifying biofilm reactors.** *Water Science and Technology*, 19(10-11), 1994, 1-11.
- Boller, M., Tsuchi, M. and Gujer, W. **Effects of transient nutrient concentrations in tertiary biofilm reactors.** *Water Science and Technology*, 36(1), 1997, 101-109.
- Borregaard, V. R. **Experiences with nutrient removal in a fixed-film system at full-scale wastewater treatment plants.** *Water Science and Technology*, 36(1), 1997, 129-137.
- Brewer, P. **Poole waste water treatment BAF plant.** IN: Proceedings of the 2nd BAF Symposium, June 1996, Cranfield, UK, 85 pp.
- Buswell, A. M. and Pearson, E. L. **The Nidus (Nest) rack, a modern development of the travis colloider.** *Sewage Works Journal*, 1, 1929, 187-195.
- Canler, J. P. and Perret, J. M. **Biological aerated filters: assessment of the process based on 12 sewage treatment plants.** *Water Science and Technology*.

- 29(10-11), 1994, 13-22.
- Canziani, R. **Submerged Aerated Biofilters. IV - Aeration Characteristics.** *Ingegneria Ambientale*, 17, 1988, 627-636.
- Cao, Y. S. and Alaerts, G. J. **Influence of reactor type and shear stress on aerobic biofilm morphology, population and kinetics.** *Water Research*, 29, 1995, 107-118.
- Carrand, G., Capon, B., Rasconi, A. and Brenner, R. **Elimination of carbonaceous and nitrogenous pollutants by a twin stage fixed growth process.** *Water Science and Technology*, 22(1/2), 1990, 261-272.
- Carrio, L. A., Zhang, J., Pujnn, J. and Mitchell, A. **Biological aerated filtration; New York City's experience.** IN: Proceedings of the Water Environment Federation 68th Annual Conference, 1995, Miami, USA, 805-811.
- Cecen, F. and Gonenc, I. E. **Criteria for nitrification and denitrification of high-strength wastes in two upflow submerged filters.** *Water Environment Research*, 67, 1995, 132-142.
- Chen S. and Cheng S. **The enhancement of nitrification by indirect aeration and kinetic control in a submerged biofilm reactor.** *Water Science and Technology*, 30(11), 1994, 131-142.
- Chudoba, P. and Pujol, R. **A three-stage biofiltration process: performance of a pilot plant.** *Water Science and Technology*, 38(8-9), 1998, 257-265.
- Chui, P. C., Terashima, Y., Tay, J. H. and Ozaki, H. **Performance of a partly aerated biofilter in the removal of nitrogen.** *Water Science and Technology*, 34(1-2), 1996, 187-194.
- Chung, P. -G., Kwak, D. -H. and Kim, Y. **Application of a simple empirical fixed-film model to the fluidized biofilter process.** *Water Science and Technology*, 34(3-4), 1996, 427-433.

- Clapp, L. W., Talarczyk, M. R., Park, J. K. and Boyle, W. C. **Performance comparison between activated sludge and fixed-film processes for priority pollutant removals.** *Water Environment Research*, 66, 1994, 153-160.
- Clark, H.W. **Past and present developments in sewage disposal and purification..** *Sewage Works Journal*, 2, 1930, 561-571.
- Clark, T.-A., Stephenson, T. and Pearce, S **Phosphorus removal by chemical precipitation in a biological aerated filter.** *Water Research*, 31, 1997, 2557-2563.
- Condren, A. J. *Technology Assessment of the Biological Aerated Filter.* US Environmental Protection Agency, Report USEPA/600/2-90/015, 1990.
- Costa Reis, L. G. and Sant'Anna, G. L. **Aerobic treatment of concentrated wastewater in a submerged bed reactor.** *Water Research*, 19, 1985, 1341-1345.
- Coulson, J. M. and Richardson, J. F. **Flow of fluids through granular beds and packed columns.** IN: Chemical Engineering volume 2, 3rd ed. Oxford: Pergamon Press, 1978, 125-171.
- Daamen, E. J., Delgorge, F. H., van Loosdrecht, M. C. M. and Heijnen, J. J. **The Biogenerator: a new biofilter reactor concept without backwashing requirement.** IN: Proceedings of the International Association on Water Quality 3rd International Special Conference on Biofilm Systems, Copenhagen, Denmark, 27-30 August 1996.
- De Beer, D. and Muyzer, G. **Multispecies biofilms report from the discussion session.** *Water Science and Technology*, 32(8), 1995, 269-270.
- Dee, A., James, N., Jones, I., Strickland, J., Upton, J. and Cooper, P. **Pre- or post-denitrification at biological filter works? A case study.** *Water Science and Technology*, 29(10-11), 1994, 145-155.

- Desbos, G., Rogala, F., Sibony, J. and Bourbigot, M. -M. **Biofiltration as a compact technique for small wastewater treatment plants.** *Water Science and Technology*, 22(3-4), 1990, 145-152.
- Dillon, G. R. and Thomas, V. K. **A pilot-scale evaluation of the 'Biocarbhone process' for the treatment of settled sewage and for tertiary nitrification of secondary effluent.** *Water Science and Technology*, 22(1/2), 1990, 305-316.
- Doran, P. M. Bioprocess Engineering Principles. UK: Academic Press, 1995, 439.
- Droste, R. L. Theory and Practice of Water and Wastewater Treatment. USA: John Wiley & Sons, 1997, 800 pp.
- Drushlyak, O. S. and Kirichenko, A. G. **Final treatment of effluents with petroleum products on aerated filters.** *Khimiya i Tekhnologiya Vod*, 12, 1990, 1121-1122.
- Eckenfelder, W. W. **Trickling filtration design and performance.** *Journal of the Sanitary Engineering Division, Proc. of the ASCE*, 87(SA4), 1961, 33-45.
- Eckenfelder, W. W. and Quirk, T. P. **Scale-up of activated sludge and trickling filter processes.** IN: Scale-up of water and wastewater treatment processes, Schmidtke, N. W. and Smith, D. W. (Eds.), Woburn MA: Butterworth Publishers, 1983, 447 pp.
- Faup, G.- M., Leprince, A. and Pannier, M. **Biological nitrification in an upflow fixed bed reactor (UFBR).** *Water Science and Technology*, 14, 1982, 795-810.
- Fdz-Polanco, F., Garcia, P. and Villaverde, S. **Adsorption and diffusion effects on the residence time distribution of submerged biofilters.** *Environmental Technology*, 17, 1996, 687-696.
- Fdz-Polanco, F., Villaverde, S. and Garcia, P. A. **Temperature effect on nitrifying**

- bacteria activity in biofilters: activation and free ammonia inhibition.** *Water Science and Technology*, 30(11), 1994, 121-130.
- Fruhen, M., Bocker, K., Eidens, S., Haaf, D., Liebeskind, M. and Schmidt, F. **Tertiary nitrification in pilot-plant plug-flow fixed film reactors with long term ammonium deficiency.** *Water Science and Technology*, 29(10-11), 1994, 61-67.
- Fujie, K., Hu, H. Y., Tanaka, H. and Urano, K. **Ecological studies of aerobic submerged biofilters on the basis of respiratory quinone profiles.** *Water Science and Technology*, 29(7), 1994, 373-376.
- Germain J. E. **Economical treatment of domestic waste by plastic-medium trickling filters.** *Journal of the Water Pollution Control Federation*, 38, 1966, 192-203.
- Gilles, P. **Industrial scale applications of fixed biomass on the mediterranean seaboard. Design operating results.** *Water Science and Technology*, 22(1/2), 1990, 281-292.
- Gimbel, R. **Theoretical approach to deep bed filtration.** IN: Vigneswaran, S. and Ben Aim, R. (eds.) Water, Wastewater and Sludge Filtration, Florida: CRC Press, 1989, 17-56.
- Giuliano, C. and Joret, J. C. **Distribution, characterisation and activity of microbial biomass of an aerobic fixed-bed reactor.** *Water Science and Technology*, 20(11/12), 1988, 455-457.
- Gjaltema, A. and Griebe, T. **Laboratory biofilm reactors and on-line monitoring: report of the discussion session.** *Water Science and Technology*, 32(8), 1995, 257-261
- Goncalves, R. F., Araujo, V. L. and Chernicharo, C. L. A. **Association of a UASB reactor and a submerged aerated biofilter for domestic sewage treatment.**

- Water Science and Technology*, 38(8-9), 1998, 189-195.
- Goncalves, R. F., Nogueira, F. N., Le Grand, L. and Rogalla, F. **Nitrogen and biological phosphorus removal in submerged biofilters.** *Water Science and Technology*, 30(11), 1994, 1-12
- Gonzalez-Martinez, S. and Wilderer, P. A. **Phosphate removal in a biofilm reactor.** *Water Science and Technology*, 23, 1991, 1405-1415.
- Grasmick, A., Elmaleh, S. and Yahi, H. **Nitrification by attached cell reactors aerated at co- or counter-current. Experimental data and modelling.** *Water Research*, 18, 1984, 885-891.
- Gray, T. W. **Biological aerated filters: the solution to small footprint plant.** *International Water Environmental Engineer*, Summer 1993, 5-10.
- Greenberg, A. E., Clesceri, L. S. and Eaton, A. D. (Eds.) Standard Methods for the Examination of Water and Wastewater 18th edn, Washington: APHA/AWWA/WEF, 1992.
- Griffith, L. B. **Contact aeration for sewage treatment.** *Engineering News Record*, 28, 1943, 60-64.
- Grobicki, A. and Stuckey, D. C. **Hydrodynamic characteristics of the anaerobic baffled reactor.** *Water Research*, 26(3), 1992, 371-378.
- Gschlosl, T., Michel, I., Heiter, M., Nerger, C. and Rehbein, V. **Microscopic and enzymatic investigation on biofilms of wastewater treatment systems.** *Water Science and Technology*, 36(1), 1997, 21-30
- Hamoda, M. F. **Kinetic analysis of aerated submerged fixed-film (ASFF) bioreactors.** *Water Research*, 23, 1989, 1147-1154.
- Hamoda, M. F. and Al-Ghusain, I. A. **Analysis of organic removal rates in the aerated submerged fixed film process.** *Water Science and Technology*,

- 38(8-9), 1998, 213-221.
- Hao, O. J. and Chen, J. M. **Factors affecting nitrite build-up in submerged filter system.** *Journal of Environmental Engineering*, 120, 1994, 1298-1307.
- Harendranath, C. S., Anuja, K., Singh, A., Gunaseelan, A., Satish, K. And Lala, K. **Immobilization in fixed film reactors: an ultrastructural approach.** *Water Science and Technology*, 33(8), 1996, 7-15
- Harris, S. L., Stephenson, T. and Pearce, P. **Aeration investigation of biological aerated filters using off-gas analysis.** *Water Science and Technology*, 34(3-4), 1996, 307-314.
- Haug, R. T. and McCarty, P. L. **Nitrification with submerged filters.** *Journal of the Water Pollution Control Federation*, 44, 1972, 2086-2102.
- Hellinga, C., Vanrolleghem, P., van Loosdrecht, M. C. M. and Heijnen, J. J. **The potential of off-gas analyses for monitoring wastewater treatment plants.** *Water Science and Technology*, 33(1), 1996, 13-23.
- Henze, M., Grady Jr., C. P. L., Gujer, W., Marais, G. V. R. and Matsuo, T. **Activated sludge model 1.** IAWPRC Scientific and Technical Reports No 1, Elsevier Science, London, 1987.
- Hodkinson, B. J., Williams, J. B. and Ha, T. N. **Effects of plastic support media on the diffusion of air in a submerged aerated filter.** *Journal of the Chartered Institution of Water and Environmental Management*, 12(June), 1998, 188-190.
- Holladay D. W., Hancher C. W., Scott C. D. and Chilcote D. D. **Biodegradation of phenolic waste liquors in stirred-tank, packed-bed and fluidised-bed reactors.** *Journal of the Water Pollution Control Federation*, 50, 1978, 2573-2589.

- Horvath, I. and Schmidtke, N. W. **Scale-up and scale-down concepts and problems.** IN: Scale-up of water and wastewater treatment processes, Schmidtke, N. W. and Smith, D. W. (Eds.), Woburn MA: Butterworth Publishers, 1983, 447 pp.
- Huibregtse, G. L., Rooney, T. C. and Rasmussen, D. C. **Factors affecting fine bubble diffused aeration.** *Journal of the Water Pollution Control Federation*, 55(8), 1983, 1057-1064.
- Hwang, Y., Sakuma, H and Tanaka, T. **Denitrification with isopropanol as a carbon source in a biofilm system.** *Water Science and Technology*, 30(11), 1994, 69-78.
- Iida, Y. and Teranishi, A. **Nitrogen removal from municipal wastewater by a single submerged filter.** *Journal of the Water Pollution Control Federation*, 56, 1984, 251-258.
- Isohata, Y. and Kawabata, M. **Development and field experience of anticontact packed-bed reactor for wastewater treatment.** Proceedings of the 2nd International Conference on Fixed-Film Biological Processes, 1994, Arlington, VA, 989-1013.
- Iwai, S., Ohmori, H. and Tanaka, T. **An advanced sewage treatment process-combination of aerated submerged biological filtration and ultra-filtration with pulverised activated carbon.** *Desalination*, 23, 1977, 29-36.
- Jacob, J., Le Lann, J. -M., Pingaud, H. and Capdeville, B. **A generalised approach for dynamic modelling and simulation of biofilters: application to wastewater denitrification.** *Chemical Engineering Journal*, 65, 1997, 133-143.
- James, A. (Ed). An Introduction to Water Quality Modelling. Norwich, UK: John Wiley & Sons, 1984, 234 pp.
- Janning, K. F., Harremoes, P. and Nielsen, M. **Evaluation and modelling the kinetics in a full scale submerged denitrification filter.** *Water Science and*

- Technology*, 32(8), 1995, 115-123.
- Jaques, A. ***Treatment of municipal and industrial wastewater using biological aerated flooded filters***. Cranfield University, School of Water Sciences, MSc Thesis, 1992.
- Jennings P. A., Snoeyink V. L. and Chian E. V. K. **Theoretical model for a submerged biological filter**. *Biotechnology and Bioengineering*, 18, 1976, 1249-1273.
- Jepsen, S. E. and La Cour Jansen, J. **Biological filters for post-denitrification**. *Water Science and Technology*, 27(5-6), 1993, 369-379.
- Jimenez, B., Capdeville, B., Roques, H. and Faup, G. M. **Design considerations for a nitrification-denitrification process using two fixed-bed reactors in series**. *Water Science and Technology*, 19, 1987, 139-150.
- Johnson, R. L. and Baumann, E. R. **Advanced organics removal by pulsed adsorption beds**. *Journal of the Water Pollution Control Federation*, 43, 1971, 1640-1657.
- Kantardjieff, A. and Jones, J. P. **Practical experiences with aerobic biofilters in TMP (thermomechanical pulping), sulfite and fine paper mills in Canada**. *Water Science and Technology*, 35(2-3), 1997, 227-234.
- Kent, T. D., Fitzpatrick, C. S. B. and Williams, S. C. **Testing of biological aerated filter (BAF) media**. *Water Science and Technology*, 34(3-4), 1996, 363-370.
- Kleiber, B., Roudon, G., Bigot, B. and Sibony, J. **Assessment of aerated biofiltration at industrial scale**. *Water Science and Technology*, 29(10-11), 1994, 197-208.
- Koutsakos, E., Smith, A. J., and Brignal, W. J. **Temperature effects on the performance of a submerged aerated filter process**. IN: Proceedings of the 15th International Symposium on Wastewater Treatment, 1992, Montreal,

- Canada, 227-236.
- Kramer, J. P. and Wouters, J. W. **Dynasand filtration, a continuous type of rapid filtration.** IN: Proceedings of the 6th World Filtration Congress, 1993, Nagoya, Japan, 841-845.
- Lang, J. S., Giron, J. J., Hansen, A. T., Trussell, R. R. and Hodges, W. E. **Investigating filter performance as a function of the ratio of filter size to media size.** *Journal of the American Water Works Association*, 85(10), 1993, 122-130.
- Le Cloirec, P. and Martin, G. **The mean residential time application in an aerated immersed biological filter.** *Environmental Technology Letters*, 5, 1984, 275-282.
- Le Tallec, X., Lesouef, A., Payraudeau, M. and Zeghal, S. **Simple modelling of biological aerated filter.** IN: Verachtert, H. and Verstraete, W. (eds) Proceedings of the International Symposium on Environmental Biotechnology, Part1, 21-23 April 1997a, Oostende, Belgium, 195-198
- Le Tallec, X., Zeghal, S., Vidal, A. and Lesouef, A. **Effect of influent quality variability on biofilter operation.** *Water Science and Technology*, 36(1), 1997, 111-117.
- Lee, K. M. and Stensel, H. D. **Aeration and substrate utilisation in a sparged packed-bed biofilm reactor.** *Journal of the Water Pollution Control Federation*, 58, 1986, 1066-1072.
- Lee, N. M. and Welander, T. **Influence of predators on nitrification in aerobic biofilm processes.** *Water Science and Technology*, 29(7), 1994, 355-363.
- Leglise, J. P., Gilles, P. and Mureaud, H. **A new development in the biological aerated filter technology.** Proceedings of the 53rd Annual Water Pollution Control Federation Conference, 1980, Las Vegas, USA.

- Levenspiel, O. Chemical Reaction Engineering, 2nd ed. New York: John Wiley & Sons, 1972.
- Lewandowski, Z. and Walser, G. **Influence of hydrodynamics on biofilm accumulation.** *Environmental Engineering*, 11, 1991, 619-624.
- Lilly, W., Bourn, G., Crabtree, H., Upton, J. and Thomas, V. **The production of high quality effluents in sewage treatment using the biocarbone process.** *Journal of the Institution on Water and Environmental Management*, 5, 1991, 123-133.
- Manem, J. A. and Rittmann, B. E. **Scaling procedure for biofilm processes.** *Water Science and Technology*, 22(1/2), 1990, 329-346.
- Mann, A. T., Fitzpatrick, C. S. B. and Stephenson, T. **A comparison of floating and sunken media biological aerated filters using tracer study techniques.** *Transaction Institution of Chemical Engineers Pt. B*, 73, 1995, 137-143.
- Mann, A., Mendoza-Espinosa, L. G. and Stephenson, T. **A comparison of floating and sunken media biological aerated filters for nitrification.** *Journal of Chemical Technology and Biotechnology*, 72, 1998, 273-279.
- Mann, A., Mendoza-Espinosa, L. G. and Stephenson, T. **Performance of floating and sunken media biological aerated filters under unsteady state conditions.** *Water Research*, 33(4), 1999, 1108-1113.
- Mann A. T. and Stephenson, T. **Modelling biological aerated filters for wastewater treatment.** *Water Research*, 31, 1997, 2443-2448.
- McHarness, D. D., Haug, R. T. and McCarty, P. L. **Field studies of nitrification with submerged filters.** *Journal of the Water Pollution Control Federation*, 47, 1975, 291-309.
- Mendoza-Espinosa, L. G., Mann, A. and Stephenson, T. **Determination of flow pattern and active volume in biological aerated filters under upflow and**

- downflow conditions.** IN: Proceedings of the Institution of Chemical Engineers Research Event - Volume 1, 7-9 April 1997a, Nottingham, UK, 121-124.
- Mendoza-Espinosa L. G., Mann A. and Stephenson T. **Nitrification in an upflow and a downflow BAF: preliminary results.** IN: *Proc. of the Int. Conf. on Advanced Wastewater Treatment Processes*, September 1997b, Leeds, UK.
- Mendoza-Espinosa, L. G. and Stephenson, T. **A process model to evaluate the performance of a biological aerated filter.** *Biotechnology Techniques*, 12, 1998, 373-376.
- Metcalf & Eddy (1991) Wastewater Engineering, Treatment, Disposal and Reuse 3rd ed. Singapore: McGraw-Hill, 1991, 1334.
- Meunier, A. D. and Williamson, K. L. **Packed bed biofilm reactors: simplified model.** *Journal of the Environmental Engineering Division*, 107, 1981, 1727-1735.
- Ministry of Technology. Water Pollution Research 1967. London: Her Majesty's Stationery Office, 1968, 100-103.
- Newbigging, M. L., Stephenson, J. P. and Romano, L. S. **Upflow or downflow BAFs - which provides the best overall performance.** IN: Proceedings of the Water Environment Federation 68th Annual Conference, 1995, Miami, USA, 783-794.
- Ohashi, A., Viraj de Silva, D. G., Mobarry, B., Manem, J. A., Stahl, D. A. and Rittmann, B. E. **Influence of substrate C/N ratio on the structure of multi-species biofilms consisting of nitrifiers and heterotrophs.** *Water Science and Technology*, 32(8), 1995, 75-84
- Oslislo, A. and Lewandowski, Z. **Inhibition of nitrification in packed bed reactors by selected organic compounds.** *Water Research*, 19, 1985, 423-426.

- Paffoni, C., Gousailles, M., Rogalla, F. and Gilles, P. **Aerated biofilters for nitrification and effluent polishing.** *Water Science and Technology*, 22(7/8), 1990, 181-189.
- Park, J. W. and Ganczarczyk, J. J. **Gravity separation of biomass washed out from an aerated submerged filter.** *Environmental Technology*, 15, 1994, 945-955.
- Parker, D. S., Romano, L. S. and Horneck, H. S. **Making a trickling filter/solids contact process work for cold weather nitrification and phosphorus removal.** *Water Environment Research*, 70, 1998, 181-188.
- Pearce, P. **Aeration optimisation of biological aerated filters.** IN: Proceedings of the 2nd BAF Symposium, June 1996, Cranfield, UK, pp 85.
- Peladan, J. -G., Lemmel, H. and Pujol, R. **High nitrification rate with upflow biofiltration.** *Water Science and Technology*, 34(1-2), 1996, 347-353.
- Peladan, J. -G., Lemmel, H., Tarallo, S., Tattersall, S. and Pujol, R. **A new generation of upflow biofilters with high water velocities.** IN: Proceedings of the International Conference on Advanced Wastewater Treatment Processes, 8-11 September 1997, Leeds, UK (in press)
- Peng, J., Stevens, D. K. and Yiang, X. **A pioneer project of wastewater reuse in China.** *Water Research*, 29, 1995, 357-363.
- Pujol, R., Canler, J. P. and Iwema, A. **Biological aerated filters: an attractive and alternative biological process.** *Water Science and Technology*, 26(3-4), 1992, 693-702.
- Pujol, R., Hamon, M., Kandel, X. and Lemmel, H. **Biofilters: flexible, reliable biological reactors.** *Water Science and Technology*, 29(10-11), 1994, 33-38.
- Pujol, R., Lemmel, H. and Gousailles, M. **A keypoint of nitrification in an upflow biofiltration reactor.** *Water Science and Technology*, 38(3), 1998, 43-49.

- Quickenden, J., Mittal, R. and Gros, H. **Effluent nutrient removal with Sulzer Biopur and filtration systems.** IN: Proceedings of the European Conference on Nutrient Removal from Wastewater, September 1992, Wakefield, UK.
- Redmon, D., Boyle, W. C. And Ewing, L. **Oxygen transfer efficiency measurements in mixed liquor using off-gas techniques.** *Journal of the Water Pollution Control Federation*, 55, 1983, 1338-1347.
- Reiber, S. H. and Stensel, H. D. **Oxygen transfer in fixed film systems.** *Journal of the Water Pollution Control Federation*, 57, 1985, 135-140.
- Reichert, P. and Wanner, O. **Movement of solids in biofilms: significance of liquid phase transport.** *Water Science and Technology*, 36(1), 1997, 321-328.
- Robinson, A. B., Brignal, W. J. and Smith, A. J. **Construction and operation of a submerged aerated filter sewage treatment works.** *Journal of the Institution of Water and Environmental Management*, 8, 1994, 215-227.
- Rogalla, F. and Bourbigot, M. -M. **New developments in complete nitrogen removal with biological aerated filters.** *Water Science and Technology*. 22(1/2), 1990, 273-280.
- Rogalla, F., Lamouche, A., Specht, W. and Kleiber, B. **High rate aerated biofilters for plant upgrading.** *Water Science and Technology*, 29(12), 1994, 207-216.
- Rogalla, F. and Payraudeau, M. **Tertiary nitrification with fixed biomass reactors.** *Water Supply*, 6, 1988, 347-354.
- Rogalla, F, Payraudeau, M., Bacquet, G., Bourbigot, M. -M., Sibony, J. and Gilles, P. **Nitrification and phosphorus precipitation with biological aerated filters.** *Journal of the Water Pollution Control Federation*, 62, 1990, 169-176.
- Rogalla, F. and Sibony, J. **Biocarbone aerated filters - ten years after: past, present and plenty of potential.** *Water Science and Technology*, 26(9-11), 1992, 2043-2048.

- Rogalla, F., Townshend, A. and Sibony, J. **Innovative wastewater treatment technology for environmentally-sensitive sites.** IN: Haigh, M. D. F. and James, C. P. (eds.), Water and Environmental Management - Design and Construction of Works, 1991.
- Ros, M. and Mejac, B. **Treatment of wastewater in an upflow packed-bed reactor.** *Water Science and Technology*, 24(7), 1991, 81-88.
- Roustan, M., Metral, C., Capdeville, B., Audic, J. M. and Marty, A. **Hydrodynamics study of three phase reactor with bubble supported particles,** IN: Proceedings of the 5th Mediterranean Congress on Chemical Engineering, 5-7 November 1990, Barcelona, Spain, 291-292.
- Rovel, J. M., Trudel, J. P., Lavallee, P. and Schroeter, I. **Paper mill effluent treatment using biofiltration.** *Water Science and Technology*, 29(10-11), 1994, 217-222.
- Ruffer, H. and Rosenwinkel, K. H. **The use of biofiltration for further wastewater treatment.** *Water Science and Technology*, 16, 1984, 241-260.
- Rusten, B. **Wastewater treatment with aerated submerged biological filters.** *Journal of the Water Pollution Control Federation*, 56, 1984, 424-431.
- Ryhiner, G., Sorensen, K., Birou, B. and Gros, H. **Biofilm reactors configuration for advanced nutrient removal.** *Water Science and Technology*, 29(10-11), 1994, 111-117.
- Sagberg, P., Dauthuille, P. and Hamon, M. **Biofilm reactors: a compact solution for the upgrading of wastewater treatment plants.** *Water Science and Technology*, 26(3-4), 1992, 733-742,
- Sagberg, P. and Berg, K. G. **Experiences with Biofor reactors at VEAS, Norway.** IN: Proceedings of the 2nd BAF Symposium, June 1996, Cranfield, UK, pp 85.

- Sakuma, H., Tanaka, T. and Maki, Y. **Studies on nitrification and COD removal using biological aerated filter.** IN: Proceedings of the 6th World Filtration Congress, 1993, Nagoya, Japan, 743-746.
- Sampa H. and Tanaka, T. **Pilot-plant study of a new wastewater treatment system** *Journal of the Chartered Institution of Water and Environmental Management*, 9, 1995, 564-572.
- Schlegel, S. **The use of submerged biological filters for nitrification.** *Water Science and Technology*, 20(4/5), 1988, 177-187.
- Schmidtke, N. W. and Horvath, I. **Scale-up methodology for surface aerated reactors.** *Progress in Water Technology*, 2, 1977, 477-493.
- Seguret, F. and Racault, Y. **Hydrodynamic behaviour of a full-scale submerged biofilter and its possible influence on performances.** *Water Science and Technology*, 38(8-9), 1998, 249-256.
- Shar, Y. T., Stiegel, G. J. And Sharma, M. M. **Backmixing in gas-liquid reactors.** *American Institute of Chemical Engineers Journal*, 24(3), 1978, 369-401.
- Sinkoff, M. D., Porges, R. and McDermott, J. H. **Mean residence time of a liquid in a trickling filter.** *Journal of the Sanitary Engineering Division, Proceedings of the ASCE*, 85(SA6), 1959, 51-60.
- Smith, D. P. **Oxygen flux limitation in aerobic fixed-film biotreatment of a hazardous landfill leachate.** *Journal of Hazardous Materials*, 44, 1995, 77-91.
- Smith, A. J. and Hardy, P. J. **High-rate sewage treatment using biological aerated filters.** *Journal of the Institution of Water and Environmental Management*, 6, 1992, 179-193.
- Smith, A. J. and Marsh, P. **Enhanced wastewater treatment with lamella tube settlers, submerged aerated filters, and ultra-violet radiation.** IN:

- Proceedings of the Water Environment Federation 68th Annual Conference, 1995, Miami, USA, 627-637.
- Stensel, H. D., Brenner, R. C., Lee, K. M., Melcer, H. and Rakness, K. **Biological aerated filter evaluation**. *Journal of Environmental Engineering*, 114, 1988, 655-671.
- Stensel, H. D. and Reiber, S. **Industrial wastewater treatment with a new biological fixed-film system**. *Environmental Progress*, 2, 1983, 110-115.
- Stephenson, T. **Pilot-scale investigation of a recirculating plastic media biological aerated filter (REBAF)**. IN: Proceedings of the International Association on Water Quality 18th Biennial Conference - Preprints 10, Singapore, 23-28 June 1996, 57.
- Sterrit, R. M. and Lester, J. N. **The influence of sludge on heavy-metal removal in the activated sludge process**. *Water Research*, 15(1), 1981, 59-65.
- Sundararajan, A. and Lu, L.- K. **Biological oxygen transfer enhancement in wastewater treatment systems**. *Water Environment Research*, 67, 1995, 848-854.
- Terayama, Y., Nishimura, T., Isono, M. and Hosoya, M. **Feasibility study on nitrogen removal process by biological aerated filter**. IN: Proceedings of the Water Environment Federation 70th Annual Conference, 1997, Chicago, USA, 641-652.
- Toettrup, H., Rogalla, F., Vidal, A. and Harremoes, P. **The treatment trilogy of floating filters: from pilot to prototype to plant**. *Water Science and Technology*, 29(10-11), 1994, 23-32.
- Tschui, M., Boller, M., Gujer, W., Eugster, J., Mader, C. and Stengel, C. **Tertiary nitrification in aerated pilot biofilters**. *Water Science and Technology*, 29(10-11), 1994, 53-60.

- Valentis, G. and Lesavre, J. **Waste-water treatment by attached-growth microorganisms on a geotextile support.** *Water Science and Technology*, 22(1-2), 1989, 43-51.
- Vedry, B., Paffoni, C., Gousailles, M. and Bernard, C. **First months operation of two biofilter prototypes in the waste water plant of Acheres.** *Water Science and Technology*, 29(10-11), 1994, 39-46.
- Velz, C. J. **A basic law for the performance of biological filters.** *Sewage Works Journal*, 20, 1948, 607-617.
- Villaverde, S., Fernandez, M. T., Urueña, M. A. and Fdz-Polanco, F. **Influence of substrate concentration on the growth and activity of a nitrifying biofilm in a submerged biofilter.** *Environmental Technology*, 18, 1997a, 921-928.
- Villaverde, S., Garcia-Encina, P. A. and Fdz-Polanco, F. **Influence of pH over nitrifying biofilm activity in submerged biofilters.** *Water Research*, 31(5), 1997b, 1180-1186.
- Visvanathan, C. and Nhien, T. T. H. **Study on aerated biofilter process under high temperature conditions.** *Environmental Technology*, 16, 1995, 301-314.
- Wanner, O. and Gujer, W. **A multispecies biofilm model.** *Biotechnology and Bioengineering*, 28, 1986, 314-328.
- Watanabe, T., Kuniyasu, K. and Ohmori, H. **Anaerobic and aerobic submerged biofilter system for small scale on-site domestic sewage treatment.** *Water Science and Technology*, 27(1), 1993, 51-57
- Wheale, G. and Cooper-Smith, G. D. **Operational experience with biological aerated filters.** *Journal of the Chartered Institution of Water and Environment Management*, 9, 1995, 109-118.
- Wilderer, P. A., Cunningham, A. and Schlindler, U. **Hydrodynamics and shear stress: report from the discussion session.** *Water Science and Technology*.

- 32(8), 1995, 271-272.
- Wilderer, P. A., Roske, I., Ueberschar, A, and Davids, L. **Continuous flow and sequence batch operation of biofilm reactors: a comparative study of shock loading responses.** *Biofouling*, 6, 1993, 295-304.
- Wilford, J. and Conlon, T. P. **Contact aeration sewage treatment plants in New Jersey.** *Sewage and Industrial Wastes*, 29, 1957, 845-855.
- Woolard, C. R. **The advantages of periodically operated biofilm reactors for the treatment of highly variable wastewater.** *Water Science and Technology*, 35(1), 1997, 199-206.
- Young, J. C., Baumann, E. R. and Wall, D. J. **Packed bed reactors for secondary effluent BOD and ammonia removal.** *Journal of the Water Pollution Control Federation*, 47, 1975, 46-56.
- Young, J. C. and Stewart, M. C. **PBR – A new addition to the AWT family.** *Water Wastes Engineering*, 1979, 20-25.
- Yunt, F., Hancuff, T., Brenner, R and Shell, G. **An evaluation of submerged aeration equipment – Clean water test results.** IN: Proceedings of the WWEMA Industrial Pollution Conference, 1980, Houston, USA
- Zeghal, S., Nagem Nogueira, F., Salzer, C. and Rogalla, F. **Tertiary denitrification using an upflow floating biofilter.** IN: Proceedings of the Water Environment Federation 68th Annual Conference, 1995, Miami, USA, 813-821.
- Zinebi, S., Henriette, C., Petitdemange, E. and Joret, J. C. **Identification and characterisation of bacterial activities involved in wastewater treatment by aerobic fixed-bed reactor.** *Water Research*, 28, 1994, 2575-2582.

APPENDICES

APPENDIX 1. QBASIC PROGRAMS FOR THE CALCULATION OF THE MODEL CONSTANTS.

Small BAF. Model 1 for sCOD

```

REM MODEL NO.1 FOR BAFS (BAFSUI-1)
DIM C(20, 20), CCAL(20, 20), K(20, 20), Y(20, 20), X(20, 20), H(20), Q(20),
P(20, 20), W(20, 20)
INPUT M, N
FOR I = 1 TO M
FOR J = 0 TO N
READ C(I, J)
PRINT C(I, J);
NEXT J
PRINT "*****"
NEXT I
FOR J = 1 TO N
READ H(J)
PRINT H(J);
NEXT J
PRINT "*****"
FOR I = 1 TO M
READ Q(I)
PRINT Q(I);
Q(I) = Q(I) / 1000 / 60
NEXT I
PRINT "*****"
A = 3.1416 / 4 * .2 * .2
SY = 0
SX = 0
SXY = 0
SXX = 0
FOR I = 1 TO M
FOR J = 1 TO N
P(I, J) = LOG(C(I, J) / C(I, 0))
K(I, J) = -P(I, J) / H(J)
Y(I, J) = LOG(K(I, J))
X(I, J) = LOG(Q(I) / A)
SY = SY + Y(I, J)
SX = SX + X(I, J)
SXY = SXY + X(I, J) * Y(I, J)
SXX = SXX + X(I, J) * X(I, J)
NEXT J
NEXT I
N1 = (SXY - SX * SY / (M * N)) / (SXX - SX * SX / (M * N))
B0 = SY / (M * N) - N1 * SX / (M * N)
K1 = EXP(B0)
PRINT "N="; -N1, "K="; K1
SUM = 0
MAXE = .08
MINE = .02
ME = 0
WMAX = 10
WMIN = .8
FOR I = 1 TO M
FOR J = 1 TO N
CCAL(I, J) = C(I, 0) * EXP(-K1 * H(J) / ((Q(I) / A) ^ (-N1)))
W(I, J) = ABS(CCAL(I, J) - C(I, J))

```

```

SUM = SUM + W(I, J)
E = W(I, J) / C(I, J)
ME = ME + E
IF E < MAXE THEN 250
MAXE = E
250 IF E > MINE THEN 255
MINE = E
255 IF W(I, J) < WMAX THEN 260
WMAX = W(I, J)
260 IF W(I, J) > WMIN THEN 270
WMIN = W(I, J)
270 NEXT J
NEXT I
ME = ME / M / (N + 1)
PRINT "WMAX="; WMAX, "WMIN="; WMIN, "SUM="; SUM,
PRINT "MAX RELATIVE ERROR="; MAXE, "MIN RELATIVE ERROR="; MINE, "MEAN
RELATIVE ERROR="; ME
PRINT "DIMENSION IN THE MODEL:C--mg/l      H--m      Q--m3/s      A--m2      "
PRINT "k---OBSERVED REACTION-RATE,m/s      n---EXPIRICAL CONSTANT"
END
DATA 182,127,126,123,93,64,55,59,57,40
DATA 114,60,47,44,40,37,28,28,29,24
DATA 115,72,62,45,45,35,29,31,24,22
DATA 118,81,70,55,47,45,38,35,31,26
DATA 159,137,112,99,93,85,60,50,41,40
DATA 135,117,97,80,56,29,24,29,25,30
DATA 0.2,0.4,0.6,0.8,1.2,1.4,1.6,1.8,2
DATA 0.3,0.2,.2,.2,.3,.15

```

Small BAF. Model 2 for sCOD

```

DIM C(20, 20), k(20, 20), Y(20, 20), X1(20, 20), H(20), UG(20), UL(20), P(20,
20), W(20, 20)
DIM X2(20, 20), A(20, 20), B(20), Q(20)
INPUT M, N, P
FOR I = 1 TO M
FOR J = 0 TO N
READ C(I, J)
PRINT C(I, J);
NEXT J
PRINT
"*****"
NEXT I
FOR J = 1 TO N
READ H(J)
PRINT H(J);
NEXT J
PRINT
"*****"
FOR I = 1 TO M
READ Q(I)
PRINT Q(I);
UL(I) = Q(I) / 1000 / 60 / (3.1416 / 4 * .2 * .2)
Q(I) = Q(I) / 1000 / 60
NEXT I
PRINT
"*****"
A = 3.1416 / 4 * .2 * .2
FOR I = 1 TO P
FOR J = 1 TO P

```

```

A(I, J) = 0
NEXT J
B(I) = 0
NEXT I

FOR I = 1 TO M
FOR J = 1 TO N
P(I, J) = LOG(C(I, J) / C(I, 0))
k(I, J) = -P(I, J) / H(J)
Y(I, J) = LOG(k(I, J))
X1(I, J) = LOG(Q(I) / A)
X2(I, J) = LOG(C(I, 0))
A(1, 1) = A(1, 1) + 1
A(1, 2) = A(1, 2) + X1(I, J)
A(1, 3) = A(1, 3) + X2(I, J)
A(2, 1) = A(2, 1) + X1(I, J)
A(2, 2) = A(2, 2) + X1(I, J) * X1(I, J)
A(2, 3) = A(2, 3) + X1(I, J) * X2(I, J)
A(3, 1) = A(3, 1) + X2(I, J)
A(3, 3) = A(3, 3) + X2(I, J) * X2(I, J)
A(3, 2) = A(3, 2) + X2(I, J) * X1(I, J)
B(1) = B(1) + Y(I, J)
B(2) = B(2) + X1(I, J) * Y(I, J)
B(3) = B(3) + X2(I, J) * Y(I, J)
NEXT J
NEXT I

FOR I = 1 TO P
FOR J = 1 TO P
PRINT A(I, J);
NEXT J
PRINT B(I)
NEXT I

FOR k = 1 TO P - 1
IF ABS(A(k, k)) < .000001 THEN 240
FOR I = k + 1 TO P
X = A(I, k) / A(k, k)
FOR J = 1 TO P
A(I, J) = A(I, J) - A(k, J) * X
NEXT J
B(I) = B(I) - B(k) * X
NEXT I
240 NEXT k

FOR I = 1 TO P
FOR J = 1 TO P
PRINT A(I, J);
NEXT J
PRINT B(I);
NEXT I

FOR k = P TO 1 STEP -1
SUM = B(k)
IF k > P THEN 245
FOR R = k + 1 TO P
SUM = SUM - A(k, R) * B(R)
NEXT R
245 B(k) = SUM / A(k, k)
PRINT k; B(k),
NEXT k

```

```

K1 = EXP(B(1))
A1 = -B(2)
A2 = -B(3)
PRINT "K="; K1, "N="; A1, "M="; A2
ME = 0
MAXE = .08
MINE = .02
SUM1 = 0
WMAX = 10
WMIN = 1.5
FOR I = 1 TO M
FOR J = 1 TO N
Q = EXP(-K1 * H(J) / ((Q(I) / A) ^ A1) / (C(I, 0) ^ A2))
CCAL(I, J) = C(I, 0) * Q
W(I, J) = ABS(CCAL(I, J) - C(I, J))
SUM1 = SUM1 + W(I, J)
E = W(I, J) / C(I, J)
ME = ME + E
IF E < MAXE THEN 250
MAXE = E
250 IF E > MINE THEN 255
MINE = E
255 IF W(I, J) < WMAX THEN 260
WMAX = W(I, J)
260 IF W(I, J) > WMIN THEN 270
WMIN = W(I, J)
270 NEXT J
NEXT I
ME = ME / M / (N + 1)
PRINT "WMAX="; WMAX, "WMIN="; WMIN, "SUM="; SUM1,
PRINT "MAX RELATIVE ERROR="; MAXE, "MIN RELATIVE ERROR="; MINE, "MEAN
RELATIVE ERROR="; ME
PRINT "DIMENSION:C--mg/l      H--m      Q--m3/d      A--m2      "
PRINT "k--reaction-rate constant,m/d  m,n--empirical constants"
END
DATA 182,127,126,123,93,64,55,59,57,40
DATA 114,60,47,44,40,37,28,28,29,24
DATA 115,72,62,45,45,35,29,31,24,22
DATA 118,81,70,55,47,45,38,35,31,26
DATA 159,137,112,99,93,85,60,50,41,40
DATA 135,117,97,80,56,29,24,29,25,30
DATA 0.2,0.4,0.6,0.8,1.2,1.4,1.6,1.8,2
DATA 0.3,.2,.2,.2,.3,.15,0.2,.25,.2,.2

```

Small BAF. Model 3 for sCOD

```

DIM C(20, 20), K(20, 20), Y(20, 20), X1(20, 20), H(20), UG(20), UL(20), P(20,
20), W(20, 20)
DIM X2(20, 20), X3(20, 20), X4(20, 20), X5(20, 20), D(20), G(20, 20), F(20,
20)
DIM A(6, 6), B(6), Q(20), QG(20), SUM(100, 100), E(20, 20), PE(20, 20),
ME(100, 100)
INPUT M, N, P
FOR I = 1 TO M
FOR J = 0 TO N
READ C(I, J)
PRINT C(I, J);
NEXT J

```

```

PRINT
"*****"
NEXT I
FOR J = 1 TO N
READ H(J)
PRINT H(J);
NEXT J
PRINT
"*****"
FOR I = 1 TO M
READ Q(I)
PRINT Q(I);
UL(I) = Q(I) / 1000 / 60 / (3.1416 / 4 * .2 * .2)
Q(I) = Q(I) / 1000 / 60
NEXT I
PRINT
"*****"
A = 3.1416 / 4 * .2 * .2
FOR I = 1 TO M
READ QG(I)
PRINT QG(I);
UG(I) = QG(I) / 1000 / 60 / (3.1416 / 4 * .2 * .2)
NEXT I
PRINT
"*****"
MEMIN = .1179831
N1 = .179
FOR S = 1 TO 10
N1 = N1 + .001
K1 = .025
FOR T = 1 TO 10
K1 = K1 + .001

FOR L = 1 TO P
B(L) = 0
FOR K = 1 TO P
A(L, K) = 0
NEXT K
NEXT L
SUMMIN = 100

FOR I = 1 TO M
FOR J = 1 TO N
F(I, J) = -K1 * H(J) / ((Q(I) / A) ^ N1)
E(I, J) = EXP(F(I, J))
PE(I, J) = (C(I, 0) / C(I, J) * E(I, J)) - 1
IF PE(I, J) <= 0 GOTO 300
IF K1 > 1 GOTO 330
P(I, J) = LOG(PE(I, J))
X1(I, J) = LOG(H(J))
X2(I, J) = LOG(UL(I))
X3(I, J) = LOG(UG(I))
A(1, 1) = A(1, 1) + 1
A(1, 2) = A(1, 2) + X1(I, J)
A(1, 3) = A(1, 3) + X2(I, J)
A(1, 4) = A(1, 4) + X3(I, J)
A(2, 1) = A(2, 1) + X1(I, J)
A(2, 2) = A(2, 2) + X1(I, J) * X1(I, J)
A(2, 3) = A(2, 3) + X2(I, J) * X1(I, J)
A(2, 4) = A(2, 4) + X3(I, J) * X1(I, J)
A(3, 1) = A(3, 1) + X2(I, J)

```

```

A(3, 2) = A(3, 2) + X1(I, J) * X2(I, J)
A(3, 3) = A(3, 3) + X2(I, J) * X2(I, J)
A(3, 4) = A(3, 4) + X3(I, J) * X2(I, J)
A(4, 1) = A(4, 1) + X3(I, J)
A(4, 2) = A(4, 2) + X1(I, J) * X3(I, J)
A(4, 3) = A(4, 3) + X2(I, J) * X3(I, J)
A(4, 4) = A(4, 4) + X3(I, J) * X3(I, J)
B(1) = B(1) + P(I, J)
B(2) = B(2) + X1(I, J) * P(I, J)
B(3) = B(3) + X2(I, J) * P(I, J)
B(4) = B(4) + X3(I, J) * P(I, J)
NEXT J
NEXT I
FOR K = 1 TO P - 1
IF ABS(A(K, K)) < .000001 THEN 240
FOR I = K + 1 TO P
X = A(I, K) / A(K, K)
FOR J = 1 TO P
A(I, J) = A(I, J) - A(K, J) * X
NEXT J
B(I) = B(I) - B(K) * X
NEXT I
240 NEXT K

FOR K = P TO 1 STEP -1
SUM = B(K)
IF K > P THEN 245
FOR R = K + 1 TO P
SUM = SUM - A(K, R) * B(R)
NEXT R
245 B(K) = SUM / A(K, K)

NEXT K
ME = 0
MAXE = .08
MINE = .02
WMAX = 10
WMIN = .8
A0 = B(1)
A1 = B(2)
A2 = B(3)
A3 = B(4)
SUM(S, T) = 0
FOR I = 1 TO M
FOR J = 1 TO N
Q = EXP(-K1 * H(J) / ((Q(I) / A) ^ N1))
E = EXP(A0)
IF A4 > -10 THEN 100
F = 0
GOTO 101
100 F = H(J) ^ A1 * UL(I) ^ A2 * UG(I) ^ A3
101 Z = Q / (1 + E * F)
G(I, J) = C(I, 0) * Z
W(I, J) = ABS(G(I, J) - C(I, J))
250 SUM(S, T) = SUM(S, T) + W(I, J)
ER = W(I, J) / C(I, J)
ME = ME + ER
IF ER < MAXE THEN 251
MAXE = ER
251 IF ER > MINE THEN 255
MINE = ER

```

```

255 IF W(I, J) < WMAX THEN 260
WMAX = W(I, J)
260 IF W(I, J) > WMIN THEN 270
WMIN = W(I, J)
270 NEXT J
NEXT I
ME(S, T) = ME / M / (N + 1)
IF ME(S, T) > MEMIN THEN 290
MEMIN = ME(S, T)
PRINT N1; K1; E; A1; A2; A3; MAXE; MINE; MEMIN; WMAX; WMIN; SUM(S, T);
PRINT "*****"
290 IF SUM(S, T) > SUMMIN THEN 300
IF ABS(SUM(S, T) - SUMMIN) < 100 THEN 350
SUMMIN = SUM(S, T)
300 NEXT T
IF N1 > 1 GOTO 350
330 NEXT S

350 PRINT N1; A0; A1; A2; A3; A4; A5; SUMMIN
END
DATA 114,60,47,44,40,37,28,28,29,24
DATA 115,72,62,45,45,35,29,31,24,22
DATA 118,81,70,55,47,45,38,35,31,26

DATA 182,127,126,123,93,64,55,59,57,40
DATA 159,137,112,99,93,85,60,50,41,40
DATA 135,117,97,80,56,29,24,29,25,30

DATA 0.2,0.4,0.6,0.8,1.2,1.4,1.6,1.8,2
DATA 0.2,.2,.2,.3,.3,.15
DATA 3,3,3,2,4,3

```

Small BAF. Model 4 for tCOD

```

DIM C(20, 20), K(20, 20), Y(20, 20), X1(20, 20), H(20), UG(20), UL(20), P(20,
20), W(20, 20)
DIM X2(20, 20), X3(20, 20), X4(20, 20), X5(20, 20), D(20), G(20, 20), F(20,
20)
DIM A(6, 6), B(6), Q(20), QG(20), SUM(100, 100), E(20, 20), PE(20, 20),
ME(100, 100)
INPUT M, N
FOR I = 1 TO M
FOR J = 0 TO N
READ C(I, J)
PRINT C(I, J);
NEXT J
PRINT
PRINT "*****"
NEXT I
FOR J = 1 TO N
READ H(J)
PRINT H(J);
NEXT J
PRINT
PRINT "*****"
FOR I = 1 TO M
READ Q(I)
PRINT Q(I);
UL(I) = Q(I) / 1000 / 60 / (3.1416 / 4 * .2 * .2)
Q(I) = Q(I) / 1000 / 60

```

```

NEXT I
PRINT
"*****"
A = 3.1416 / 4 * .2 * .2

MEMIN = .4
K1 = 13
FOR S = 1 TO 1
K1 = K1 + 1
K2 = .83
FOR T = 1 TO 10
K2 = K2 + .01
Q1 = 0
ME = 0
MAXE = .08
MINE = .02
WMAX = .0000001
WMIN = .8
SUM(S, T) = 0
FOR I = 1 TO M
FOR J = 1 TO N
T(J) = 57600 - .4 * H(J) / UL(I)
Q1 = EXP(K1 * UL(I) * C(I, 0) / 1000000 * T(J))
Q2 = Q1 + EXP(K2 * H(J)) - 1
Z = Q1 / Q2
G(I, J) = C(I, 0) * Z
W(I, J) = ABS(G(I, J) - C(I, J))
250 SUM(S, T) = SUM(S, T) + W(I, J)
ER = W(I, J) / C(I, J)
251 ME = ME + ER
IF ER < MAXE THEN 252
MAXE = ER

252 IF ER > MINE THEN 255
MINE = ER
255 IF W(I, J) < WMAX THEN 260
WMAX = W(I, J)
260 IF W(I, J) > WMIN THEN 270
WMIN = W(I, J)
270 NEXT J
NEXT I
ME(S, T) = ME / M / (N + 1)
IF ME(S, T) > MEMIN THEN 290
MEMIN = ME(S, T)
PRINT K1; K2; MAXE; MINE; MEMIN; WMAX; WMIN; SUM(S, T);
PRINT "*****"
290 IF SUM(S, T) > SUMMIN THEN 300
IF ABS(SUM(S, T) - SUMMIN) < 100 THEN 350
SUMMIN = SUM(S, T)
300 NEXT T
IF N1 > 1 GOTO 350
330 NEXT S

350 PRINT K1; K2
END
DATA 343,194,189,153,154,138,124,133,124,102
DATA 297,225,288,175,138,110,95,86,94,94
DATA 343,251,249,203,173,154,134,129,99,112

DATA 238,179,146,136,103,59,53,79,74,44
DATA 236,144,142,114,103,88,73,53,45,41

```

```
DATA 291,240,201,186,159,144,111,88,72,77
DATA 306,225,153,125,92,90,78,78,98,62
```

```
DATA .2,.4,.6,.8,1.2,1.4,1.6,1.8,2
DATA .3,.2,.3,.2,.2,.3,.2
```

Small BAF. Model 4 for SS

```
DIM C(20, 20), K(20, 20), Y(20, 20), X1(20, 20), H(20), UG(20), UL(20), P(20,
20), W(20, 20)
DIM X2(20, 20), X3(20, 20), X4(20, 20), X5(20, 20), D(20), G(20, 20), F(20,
20)
DIM A(6, 6), B(6), Q(20), QG(20), SUM(100, 100), E(20, 20), PE(20, 20),
ME(100, 100)
INPUT M, N
FOR I = 1 TO M
FOR J = 0 TO N
READ C(I, J)
PRINT C(I, J);
NEXT J
PRINT
"*****"
NEXT I
FOR J = 1 TO N
READ H(J)
PRINT H(J);
NEXT J
PRINT
"*****"
FOR I = 1 TO M
READ Q(I)
PRINT Q(I);
UL(I) = Q(I) / 1000 / 60 / (3.1416 / 4 * .2 * .2)
Q(I) = Q(I) / 1000 / 60
NEXT I
PRINT
"*****"
A = 3.1416 / 4 * .2 * .2

MEMIN = .4
K1 = 508
FOR S = 1 TO 10
K1 = K1 + 1
K2 = .882
FOR T = 1 TO 10
K2 = K2 + .001
Q1 = 0
ME = 0
MAXE = .08
MINE = .02
WMAX = .0000001
WMIN = .8
SUM(S, T) = 0
FOR I = 1 TO M
FOR J = 1 TO N
T(J) = 57600 - .4 * H(J) / UL(I)
Q1 = EXP(K1 * UL(I) * C(I, 0) / 1000000 * T(J))
Q2 = Q1 + EXP(K2 * H(J)) - 1
Z = Q1 / Q2
G(I, J) = C(I, 0) * Z
```

```

W(I, J) = ABS(G(I, J) - C(I, J))
250 SUM(S, T) = SUM(S, T) + W(I, J)
ER = W(I, J) / C(I, J)
251 ME = ME + ER
IF ER < MAXE THEN 252
MAXE = ER

252 IF ER > MINE THEN 255
MINE = ER
255 IF W(I, J) < WMAX THEN 260
WMAX = W(I, J)
260 IF W(I, J) > WMIN THEN 270
WMIN = W(I, J)
270 NEXT J
NEXT I
ME(S, T) = ME / M / (N + 1)
IF ME(S, T) > MEMIN THEN 290
MEMIN = ME(S, T)
PRINT K1; K2; MAXE; MINE; MEMIN; WMAX; WMIN; SUM(S, T);
PRINT "*****"
290 IF SUM(S, T) > SUMMIN THEN 300
IF ABS(SUM(S, T) - SUMMIN) < 100 THEN 350
SUMMIN = SUM(S, T)
300 NEXT T
IF N1 > 1 GOTO 350
330 NEXT S

350 PRINT K1; K2
END
DATA 103,71,60,55,54,55,44,47,50,39
DATA 128,90,74,86,77,39,46,42,42,43
DATA 115,94,92,65,68,60,60,64,51,36
DATA 64,76,44,34,48,30,28,21,10,7
DATA 77,86,50,39,39,30,16,27,18,19
DATA 88,50,92,54,38,20,37,20,23,12

DATA 81,56,43,53,51,39,31,23,21,10
DATA 91,68,58,58,50,46,26,34,23,22
DATA 90,95,69,62,55,60,37,28,53,31

DATA .2,.4,.6,.8,1.2,1.4,1.6,1.8,2
DATA .3,.2,.3,.2,.2,.25,.2,.3,.2

```

Large BAF. Model 1 for sCOD

```

DIM C(20, 20), CCAL(20, 20), K(20, 20), Y(20, 20), X(20, 20), H(20), Q(20),
P(20, 20), W(20, 20)
INPUT M, N
FOR I = 1 TO M
FOR J = 0 TO N
READ C(I, J)
PRINT C(I, J);
NEXT J
PRINT "*****"
NEXT I
FOR J = 1 TO N
READ H(J)
PRINT H(J);
NEXT J

```

```

PRINT "*****"
FOR I = 1 TO M
  READ Q(I)
  PRINT Q(I);
  Q(I) = Q(I) / 1000 / 60
NEXT I
PRINT "*****"
A = 3.1416 / 4 * .5 * .5
SY = 0
SX = 0
SXY = 0
SXX = 0
FOR I = 1 TO M
  FOR J = 1 TO N
    P(I, J) = LOG(C(I, J) / C(I, 0))
    K(I, J) = -P(I, J) / H(J)
    Y(I, J) = LOG(K(I, J))
    X(I, J) = LOG(Q(I) / A)
    SY = SY + Y(I, J)
    SX = SX + X(I, J)
    SXY = SXY + X(I, J) * Y(I, J)
    SXX = SXX + X(I, J) * X(I, J)
  NEXT J
NEXT I
N1 = (SXY - SX * SY / (M * N)) / (SXX - SX * SX / (M * N))
B0 = SY / (M * N) - N1 * SX / (M * N)
K1 = EXP(B0)
PRINT "N="; -N1, "K="; K1
SUM = 0
MAXE = .08
MINE = .02
ME = 0
WMAX = 10
WMIN = .8
FOR I = 1 TO M
  FOR J = 1 TO N
    CCAL(I, J) = C(I, 0) * EXP(-K1 * H(J) / ((Q(I) / A) ^ (-N1)))
    W(I, J) = ABS(CCAL(I, J) - C(I, J))
    SUM = SUM + W(I, J)
    E = W(I, J) / C(I, J)
    ME = ME + E
    IF E < MAXE THEN 250
    MAXE = E
    250 IF E > MINE THEN 255
    MINE = E
    255 IF W(I, J) < WMAX THEN 260
    WMAX = W(I, J)
    260 IF W(I, J) > WMIN THEN 270
    WMIN = W(I, J)
  270 NEXT J
NEXT I
ME = ME / M / (N + 1)
PRINT "WMAX="; WMAX, "WMIN="; WMIN, "SUM="; SUM,
PRINT "MAX RELATIVE ERROR="; MAXE, "MIN RELATIVE ERROR="; MINE, "MEAN
RELATIVE ERROR="; ME
PRINT "DIMENSION IN THE MODEL:C--mg/l      H--m      Q--m3/s      A--m2      "
PRINT "k---OBSERVED REACTION-RATE,m/s      n---EXPIRICAL CONSTANT"
END
DATA 154,136,118,84,70,54,56
DATA 136,107,88,64,48,41,43
DATA 143,116,93,63,46,43,43

```

DATA 166,130,106,80,61,46,44

DATA 86,68,66,41,45,37,39

DATA 0.6,1.2,1.7,2.3,2.9,3.9

DATA 6,6,6,6,6

Large BAF. Model 2 for sCOD

```
DIM C(20, 20), K(20, 20), Y(20, 20), X1(20, 20), H(20), UG(20), UL(20), P(20, 20), W(20, 20)
```

```
DIM X2(20, 20), A(20, 20), B(20), Q(20)
```

```
INPUT M, N, P
```

```
FOR I = 1 TO M
```

```
FOR J = 0 TO N
```

```
READ C(I, J)
```

```
LPRINT C(I, J);
```

```
NEXT J
```

```
LPRINT
```

```
"*****"
```

```
NEXT I
```

```
FOR J = 1 TO N
```

```
READ H(J)
```

```
LPRINT H(J);
```

```
NEXT J
```

```
LPRINT
```

```
"*****"
```

```
FOR I = 1 TO M
```

```
READ Q(I)
```

```
LPRINT Q(I);
```

```
UL(I) = Q(I) / 1000 / 60 / (3.1416 / 4 * .5 * .5)
```

```
Q(I) = Q(I) / 1000 / 60
```

```
NEXT I
```

```
LPRINT
```

```
"*****"
```

```
A = 3.1416 / 4 * .5 * .5
```

```
FOR I = 1 TO P
```

```
FOR J = 1 TO P
```

```
A(I, J) = 0
```

```
NEXT J
```

```
B(I) = 0
```

```
NEXT I
```

```
FOR I = 1 TO M
```

```
FOR J = 1 TO N
```

```
P(I, J) = LOG(C(I, J) / C(I, 0))
```

```
K(I, J) = -P(I, J) / H(J)
```

```
Y(I, J) = LOG(K(I, J))
```

```
X1(I, J) = LOG(Q(I) / A)
```

```
X2(I, J) = LOG(C(I, 0))
```

```
A(1, 1) = A(1, 1) + 1
```

```
A(1, 2) = A(1, 2) + X1(I, J)
```

```
A(1, 3) = A(1, 3) + X2(I, J)
```

```
A(2, 1) = A(2, 1) + X1(I, J)
```

```
A(2, 2) = A(2, 2) + X1(I, J) * X1(I, J)
```

```
A(2, 3) = A(2, 3) + X1(I, J) * X2(I, J)
```

```
A(3, 1) = A(3, 1) + X2(I, J)
```

```
A(3, 3) = A(3, 3) + X2(I, J) * X2(I, J)
```

```
A(3, 2) = A(3, 2) + X2(I, J) * X1(I, J)
```

```
B(1) = B(1) + Y(I, J)
```

```
B(2) = B(2) + X1(I, J) * Y(I, J)
```

```

B(3) = B(3) + X2(I, J) * Y(I, J)
NEXT J
NEXT I

FOR I = 1 TO P
FOR J = 1 TO P
PRINT A(I, J);
NEXT J
PRINT B(I)
NEXT I

FOR K = 1 TO P - 1
IF ABS(A(K, K)) < .000001 THEN 240
FOR I = K + 1 TO P
X = A(I, K) / A(K, K)
FOR J = 1 TO P
A(I, J) = A(I, J) - A(K, J) * X
NEXT J
B(I) = B(I) - B(K) * X
NEXT I
240 NEXT K

FOR I = 1 TO P
FOR J = 1 TO P
PRINT A(I, J);
NEXT J
PRINT B(I);
NEXT I

FOR K = P TO 1 STEP -1
SUM = B(K)
IF K > P THEN 245
FOR R = K + 1 TO P
SUM = SUM - A(K, R) * B(R)
NEXT R
245 B(K) = SUM / A(K, K)
PRINT K; B(K),
NEXT K

K1 = EXP(B(1))
A1 = -B(2)
A2 = -B(3)
LPRINT "K="; K1, "N="; A1, "M="; A2
ME = 0
MAXE = .08
MINE = .02
SUM1 = 0
WMAX = 10
WMIN = 1.5
FOR I = 1 TO M
FOR J = 1 TO N
Q = EXP(-K1 * H(J) / ((Q(I) / A) ^ A1) / (C(I, 0) ^ A2))
CCAL(I, J) = C(I, 0) * Q
W(I, J) = ABS(CCAL(I, J) - C(I, J))
SUM1 = SUM1 + W(I, J)
E = W(I, J) / C(I, J)
ME = ME + E
IF E < MAXE THEN 250
MAXE = E
250 IF E > MINE THEN 255
MINE = E

```

```

255 IF W(I, J) < WMAX THEN 260
WMAX = W(I, J)
260 IF W(I, J) > WMIN THEN 270
WMIN = W(I, J)
270 NEXT J
NEXT I
ME = ME / M / (N + 1)
LPRINT "WMAX="; WMAX, "WMIN="; WMIN, "SUM="; SUM1,
LPRINT "MAX RELATIVE ERROR="; MAXE, "MIN RELATIVE ERROR="; MINE, "MEAN
RELATIVE ERROR="; ME
LPRINT "DIMENSION:C--mg/l      H--m      Q--m3/s      A--m2      "
LPRINT "k--reaction-rate constant,m/s  m,n--expirical constants"
END
DATA 154,136,118,84,70,54,56
DATA 136,107,88,64,48,41,43
DATA 143,116,93,63,46,43,43
DATA 166,130,106,80,61,46,44
DATA 86,68,66,41,45,37,39
DATA .6,1.2,1.7,2.3,2.9,3.5,3.9
DATA 6,6,6,6,6

```

Large BAF. Model 3 for sCOD

```

DIM C(20, 20), K(20, 20), Y(20, 20), X1(20, 20), H(20), UG(20), UL(20), P(20,
20), W(20, 20)
DIM X2(20, 20), X3(20, 20), X4(20, 20), X5(20, 20), D(20), G(20, 20), F(20,
20)
DIM A(6, 6), B(6), Q(20), QG(20), SUM(100, 100), E(20, 20), PE(20, 20),
ME(100, 100)
INPUT M, N, P
FOR I = 1 TO M
FOR J = 0 TO N
READ C(I, J)
PRINT C(I, J);
NEXT J
PRINT
"*****"
NEXT I
FOR J = 1 TO N
READ H(J)
PRINT H(J);
NEXT J
PRINT
"*****"
FOR I = 1 TO M
READ Q(I)
PRINT Q(I);
UL(I) = Q(I) / 1000 / 60 / (3.1416 / 4 * .5 * .5)
Q(I) = Q(I) / 1000 / 60
NEXT I
PRINT
"*****"
A = 3.1416 / 4 * .5 * .5

MEMIN = .4
N1 = .41
FOR S = 1 TO 10
N1 = N1 + .001
K1 = 0!
FOR T = 1 TO 10

```

```

K1 = K1 + .001

FOR L = 1 TO P
  B(L) = 0
  FOR K = 1 TO P
    A(L, K) = 0
  NEXT K
NEXT L
SUMMIN = 100

FOR I = 1 TO M
  FOR J = 1 TO N
    F(I, J) = -K1 * H(J) / ((Q(I) / A) ^ N1)
    E(I, J) = EXP(F(I, J))
    PE(I, J) = (C(I, 0) / C(I, J) * E(I, J)) - 1
    IF PE(I, J) <= 0 GOTO 300
    IF K1 > 1 GOTO 330
    P(I, J) = LOG(PE(I, J))
    X1(I, J) = LOG(H(J))
    X2(I, J) = LOG(UL(I))
    A(1, 1) = A(1, 1) + 1
    A(1, 2) = A(1, 2) + X1(I, J)
    A(1, 3) = A(1, 3) + X2(I, J)

    A(2, 1) = A(2, 1) + X1(I, J)
    A(2, 2) = A(2, 2) + X1(I, J) * X1(I, J)
    A(2, 3) = A(2, 3) + X2(I, J) * X1(I, J)

    A(3, 1) = A(3, 1) + X2(I, J)
    A(3, 2) = A(3, 2) + X1(I, J) * X2(I, J)
    A(3, 3) = A(3, 3) + X2(I, J) * X2(I, J)

    B(1) = B(1) + P(I, J)
    B(2) = B(2) + X1(I, J) * P(I, J)
    B(3) = B(3) + X2(I, J) * P(I, J)
    B(4) = B(4) + X3(I, J) * P(I, J)
  NEXT J
NEXT I
FOR K = 1 TO P - 1
  IF ABS(A(K, K)) < .000001 THEN 240
  FOR I = K + 1 TO P
    X = A(I, K) / A(K, K)
    FOR J = 1 TO P
      A(I, J) = A(I, J) - A(K, J) * X
    NEXT J
    B(I) = B(I) - B(K) * X
  NEXT I
  240 NEXT K

FOR K = P TO 1 STEP -1
  SUM = B(K)
  IF K > P THEN 245
  FOR R = K + 1 TO P
    SUM = SUM - A(K, R) * B(R)
  NEXT R
  245 B(K) = SUM / A(K, K)

NEXT K
ME = 0
MAXE = .08
MINE = .02

```

```

WMAX = 10
WMIN = .8
A0 = B(1)
A1 = B(2)
A2 = B(3)
SUM(S, T) = 0
FOR I = 1 TO M
  FOR J = 1 TO N
    Q = EXP(-K1 * H(J) / ((Q(I) / A) ^ N1))
    E = EXP(A0)
    F = H(J) ^ A1 * UL(I) ^ A2
    Z = Q / (1 + E * F)
    G(I, J) = C(I, 0) * Z
    W(I, J) = ABS(G(I, J) - C(I, J))
    250 SUM(S, T) = SUM(S, T) + W(I, J)
    ER = W(I, J) / C(I, J)
    ME = ME + ER
    IF ER < MAXE THEN 251
    MAXE = ER
    251 IF ER > MINE THEN 255
    MINE = ER
    255 IF W(I, J) < WMAX THEN 260
    WMAX = W(I, J)
    260 IF W(I, J) > WMIN THEN 270
    WMIN = W(I, J)
    270 NEXT J
  NEXT I
  ME(S, T) = ME / M / (N + 1)
  IF ME(S, T) > .18994 THEN 290
  MEMIN = ME(S, T)
  PRINT N1; K1; E; A1; A2; A3; MAXE; MINE; MEMIN; WMAX; WMIN; SUM(S, T);
  PRINT "*****"
  290 IF SUM(S, T) > SUMMIN THEN 300
  IF ABS(SUM(S, T) - SUMMIN) < 100 THEN 350
  SUMMIN = SUM(S, T)
  300 NEXT T
  IF N1 > 1 GOTO 350
  330 NEXT S

350 PRINT N1; A0; A1; A2; A3; A4; A5; SUMMIN
END
DATA 154,136,118,84,70,54,56
DATA 136,107,88,64,48,41,43
DATA 143,116,93,63,46,43,43
DATA 166,130,106,80,61,46,44
DATA 86,68,66,41,45,37,39

DATA .6,1.2,1.7,2.3,2.9,3.9
DATA 6,6,6,6,6

```

Large BAF.Model 4 for tCOD

```

DIM C(20, 20), K(20, 20), Y(20, 20), X1(20, 20), H(20), UG(20), UL(20), P(20,
20), W(20, 20)
DIM X2(20, 20), X3(20, 20), X4(20, 20), X5(20, 20), D(20), G(20, 20), F(20,
20)
DIM A(6, 6), B(6), Q(20), QG(20), SUM(100, 100), E(20, 20), PE(20, 20),
ME(100, 100)
INPUT M, N
FOR I = 1 TO M

```

```

FOR J = 0 TO N
READ C(I, J)
PRINT C(I, J);
NEXT J
PRINT
"*****"
NEXT I
FOR J = 1 TO N
READ H(J)
PRINT H(J);
NEXT J
PRINT
"*****"
FOR I = 1 TO M
READ Q(I)
PRINT Q(I);
UL(I) = Q(I) / 1000 / 60 / (3.1416 / 4 * .5 * .5)
Q(I) = Q(I) / 1000 / 60
NEXT I
PRINT
"*****"
A = 3.1416 / 4 * .5 * .5

MEMIN = .4
K1 = 84.45
FOR S = 1 TO 10
K1 = K1 + .01
K2 = .533
FOR T = 1 TO 10
K2 = K2 + .001
Q1 = 0
ME = 0
MAXE = .08
MINE = .02
WMAX = .0000001
WMIN = .8
SUM(S, T) = 0
FOR I = 1 TO M
FOR J = 1 TO N
T(J) = 28800 - .4 * H(J) / UL(I)
Q1 = EXP(K1 * UL(I) * C(I, 0) / 1000000 * T(J))
Q2 = Q1 + EXP(K2 * H(J)) - 1
Z = Q1 / Q2
G(I, J) = C(I, 0) * Z
W(I, J) = ABS(G(I, J) - C(I, J))
250 SUM(S, T) = SUM(S, T) + W(I, J)
ER = W(I, J) / C(I, J)

251 ME = ME + ER
IF ER < MAXE THEN 252
MAXE = ER

252 IF ER > MINE THEN 255
MINE = ER
255 IF W(I, J) < WMAX THEN 260
WMAX = W(I, J)
260 IF W(I, J) > WMIN THEN 270
WMIN = W(I, J)
270 NEXT J
NEXT I
ME(S, T) = ME / M / (N + 1)

```

```

IF ME(S, T) > MEMIN THEN 290
MEMIN = ME(S, T)
PRINT K1; K2; MAXE; MINE; MEMIN; WMAX; WMIN; SUM(S, T);
PRINT "*****"
290 IF SUM(S, T) > SUMMIN THEN 300
IF ABS(SUM(S, T) - SUMMIN) < 100 THEN 350
SUMMIN = SUM(S, T)
300 NEXT T
IF N1 > 1 GOTO 350
330 NEXT S

350 PRINT K1; K2
END
DATA 316,304,232,156,112,48,72
DATA 328,288,236,156,68,56,56
DATA 226,222,180,122,86,50,62
DATA 300,316,220,176,80,72,56
DATA 368,344,264,188,128,96,96
DATA 356,324,284,192,140,108,104

DATA .6,1.2,1.7,2.3,2.9,3.9
DATA 6,6,6,6,6,6

```

Large BAF. Model 4 for SS

```

DIM C(20, 20), K(20, 20), Y(20, 20), X1(20, 20), H(20), UG(20), UL(20), P(20,
20), W(20, 20)
DIM X2(20, 20), X3(20, 20), X4(20, 20), X5(20, 20), D(20), G(20, 20), F(20,
20)
DIM A(6, 6), B(6), Q(20), QG(20), SUM(100, 100), E(20, 20), PE(20, 20),
ME(100, 100)
INPUT M, N
FOR I = 1 TO M
FOR J = 0 TO N
READ C(I, J)
PRINT C(I, J);
NEXT J
PRINT
"*****"
NEXT I
FOR J = 1 TO N
READ H(J)
PRINT H(J);
NEXT J
PRINT
"*****"
FOR I = 1 TO M
READ Q(I)
PRINT Q(I);
UL(I) = Q(I) / 1000 / 60 / (3.1416 / 4 * .5 * .5)
Q(I) = Q(I) / 1000 / 60
NEXT I
PRINT
"*****"
A = 3.1416 / 4 * .5 * .5

MEMIN = .4
K1 = 1539
FOR S = 1 TO 10
K1 = K1 + 1

```

```

K2 = 1.25
FOR T = 1 TO 10
K2 = K2 + .001
Q1 = 0
ME = 0
MAXE = .08
MINE = .02
WMAX = .0000001
WMIN = .8
SUM(S, T) = 0
FOR I = 1 TO M
FOR J = 1 TO N
T(J) = 28800 - .4 * H(J) / UL(I)
Q1 = EXP(K1 * UL(I) * C(I, 0) / 1000000 * T(J))
Q2 = Q1 + EXP(K2 * H(J)) - 1
Z = Q1 / Q2
G(I, J) = C(I, 0) * Z
W(I, J) = ABS(G(I, J) - C(I, J))
250 SUM(S, T) = SUM(S, T) + W(I, J)
ER = W(I, J) / C(I, J)

251 ME = ME + ER
IF ER < MAXE THEN 252
MAXE = ER

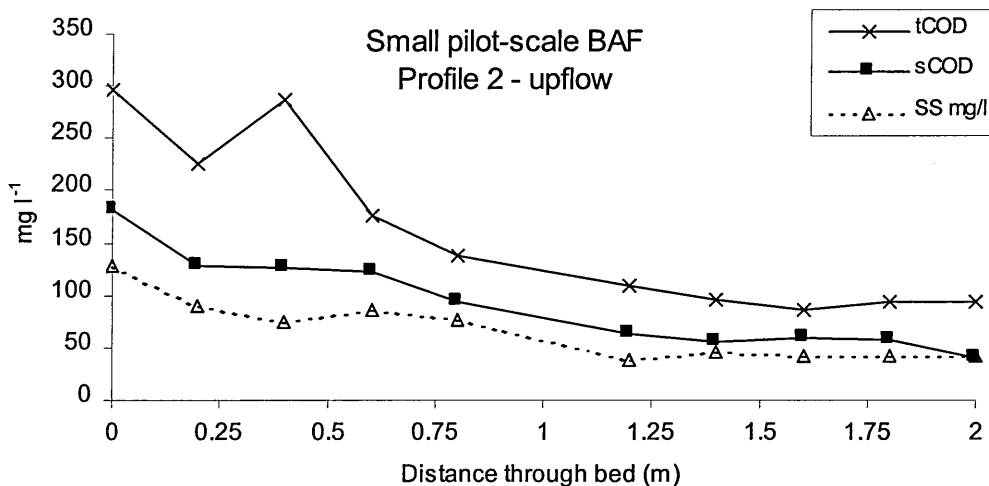
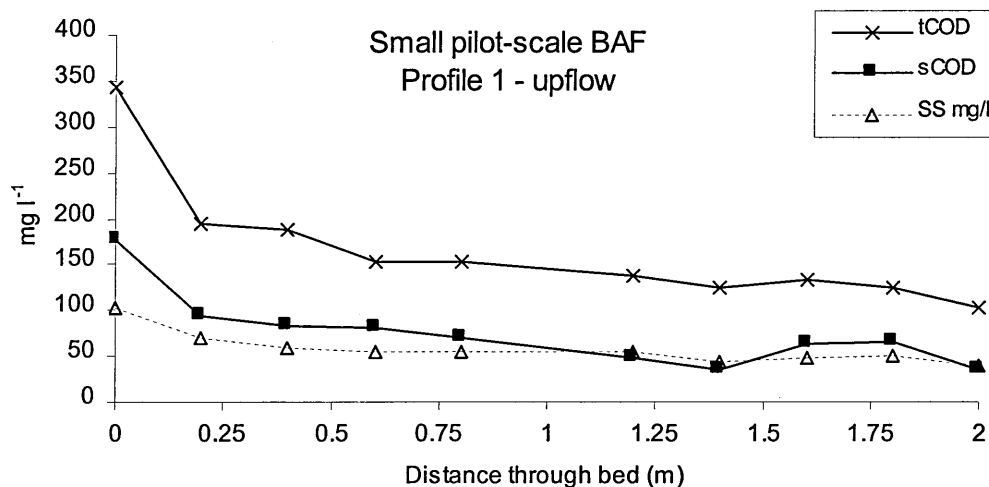
252 IF ER > MINE THEN 255
MINE = ER
255 IF W(I, J) < WMAX THEN 260
WMAX = W(I, J)
260 IF W(I, J) > WMIN THEN 270
WMIN = W(I, J)
270 NEXT J
NEXT I
ME(S, T) = ME / M / (N + 1)
IF ME(S, T) > MEMIN THEN 290
MEMIN = ME(S, T)
PRINT K1; K2; MAXE; MINE; MEMIN; WMAX; WMIN; SUM(S, T);
PRINT "*****"
290 IF SUM(S, T) > SUMMIN THEN 300
IF ABS(SUM(S, T) - SUMMIN) < 100 THEN 350
SUMMIN = SUM(S, T)
300 NEXT T
IF N1 > 1 GOTO 350
330 NEXT S

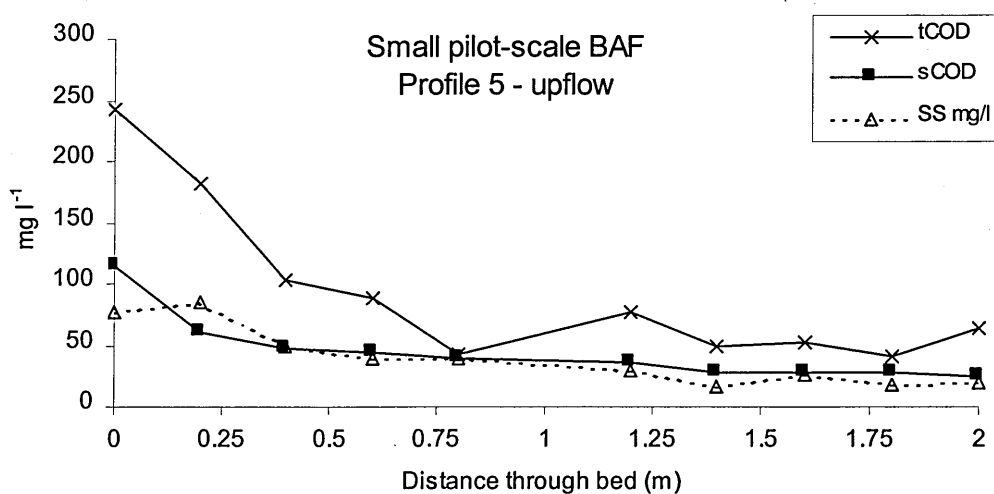
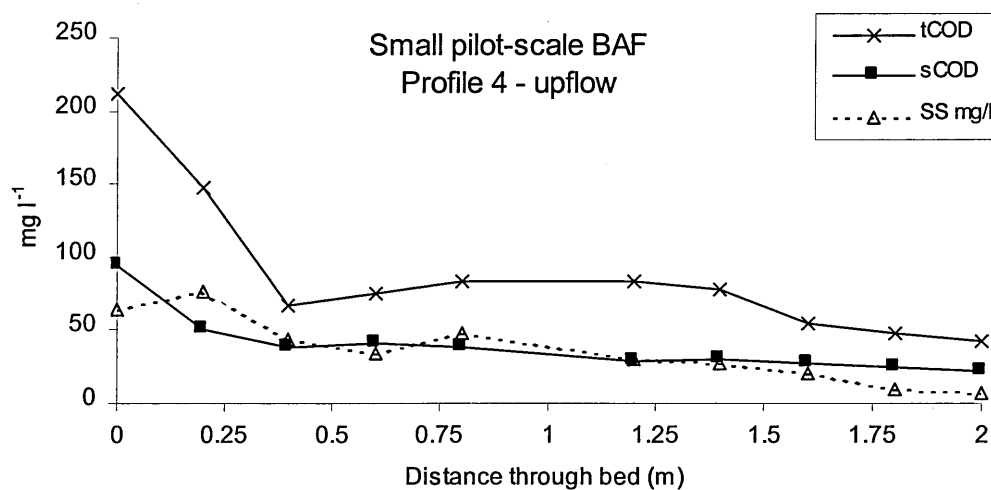
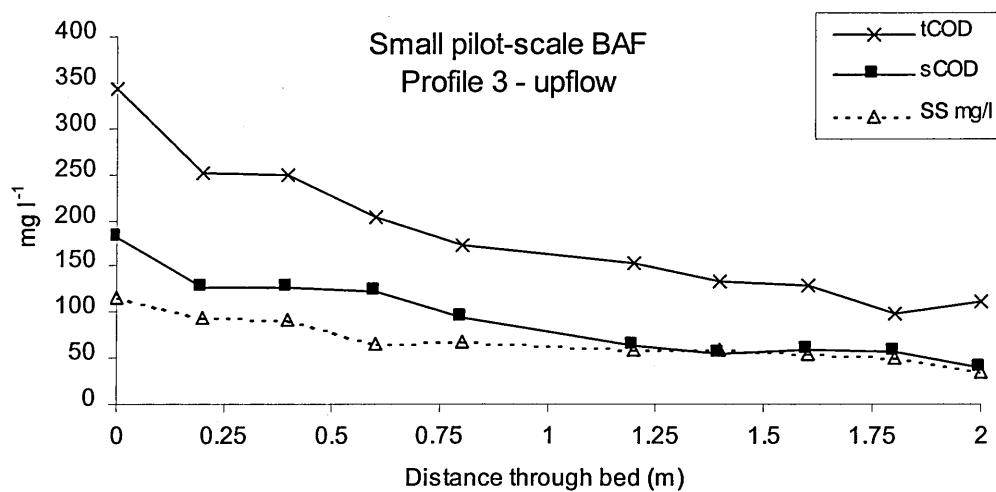
350 PRINT K1; K2
END
DATA 110,108,98,64,52,14,16
DATA 112,116,95,59,41,23,16
DATA 97,102,81,43,22,10,5
DATA 124,133,105,85,56,44,35
DATA 122,136,108,73,54,34,34

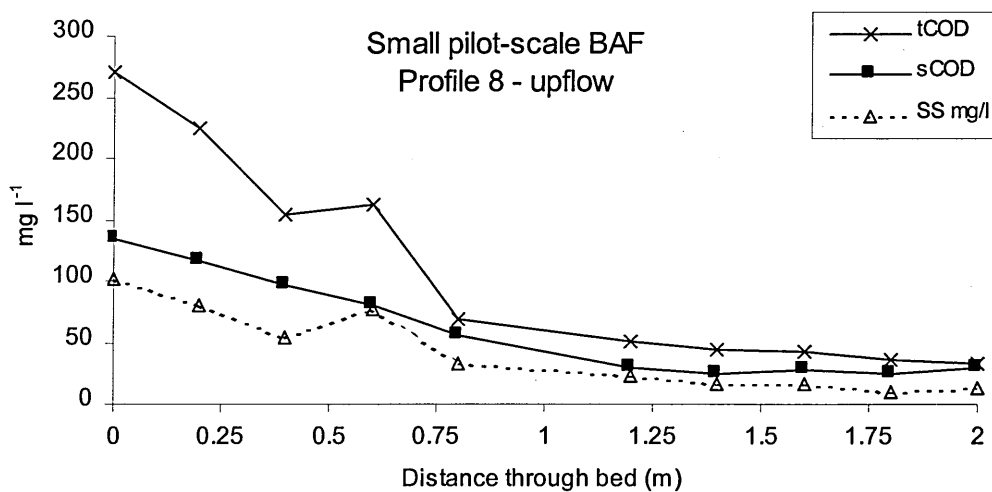
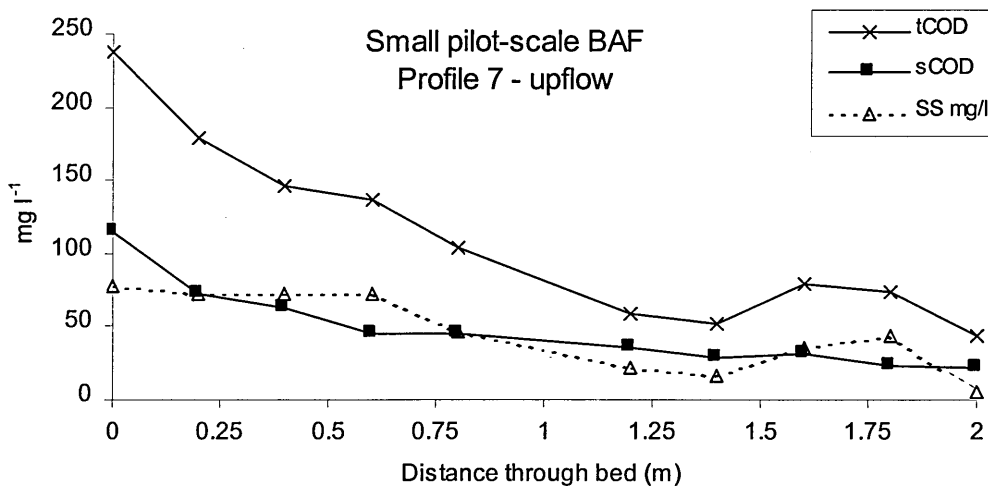
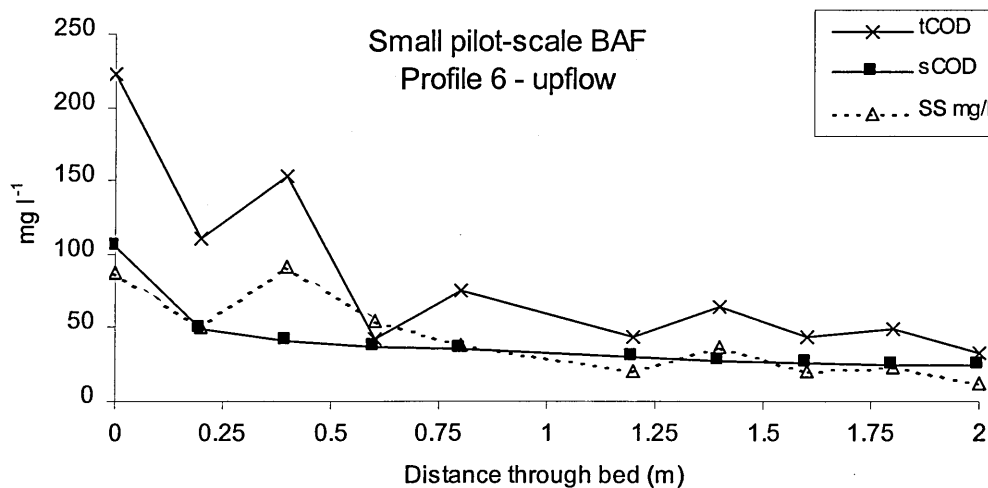
DATA .6,1.2,1.7,2.3,2.9,3.9
DATA 6,6,6,6,6,6

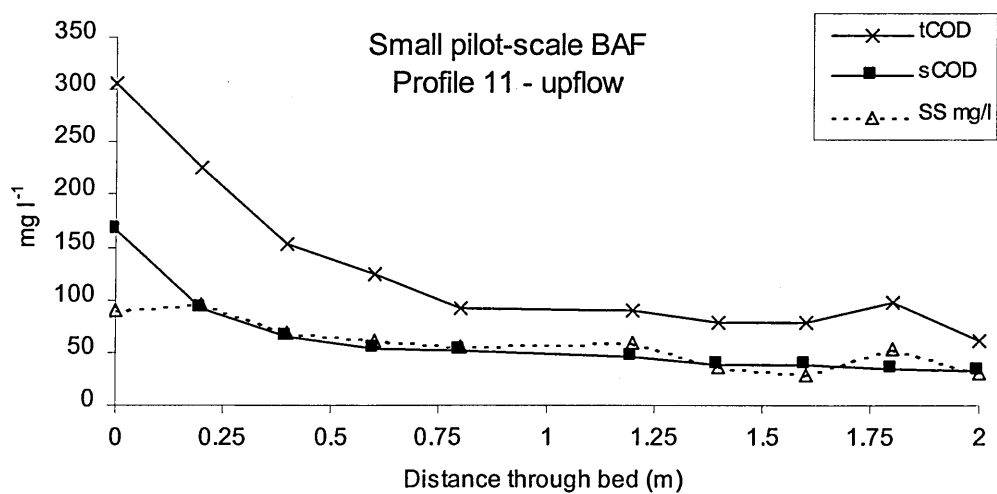
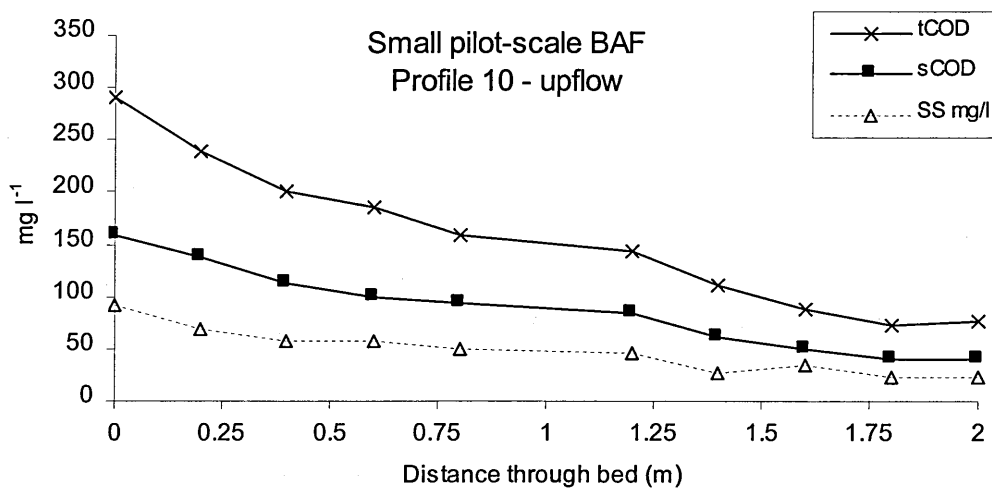
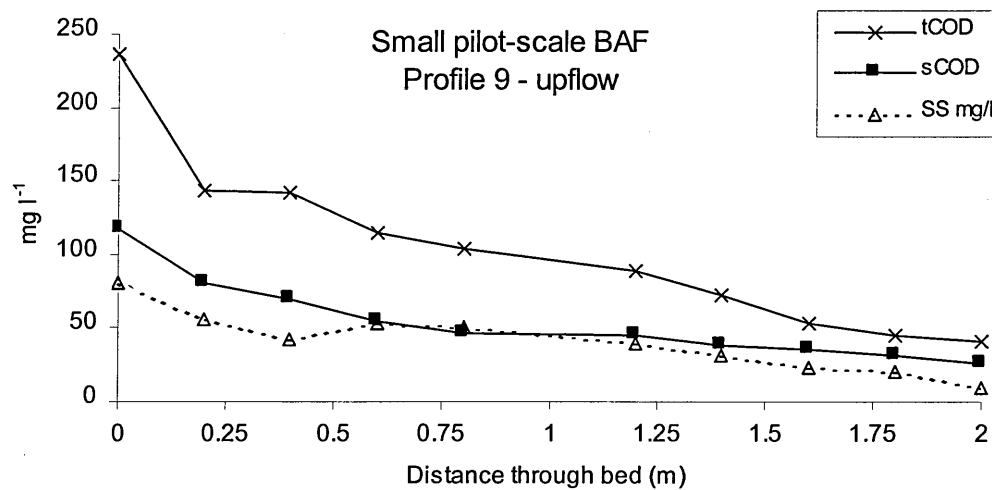
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APPENDIX 2. SUSPENDED SOLIDS, SOLUBLE COD AND TOTAL COD PROFILES FOR THE SMALL PILOT-SCALE BAF RUNNING IN UPFLOW MODE (CHAPTER 7).









APPENDIX 3. SUSPENDED SOLIDS, SOLUBLE COD AND TOTAL COD PROFILES FOR THE LARGE PILOT-SCALE BAF RUNNING IN UPFLOW MODE (CHAPTER 7).

