Cranfield UNIVERSITY

Biological Aerated Filters

BAF 2

Cranfield University
12th June 1996

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2nd Symposium on Biological Aerated Filters (BAF2)
12 June 1996

Following the success of the first BAF symposium held here in 1993, Cranfield University's School of Water Sciences is holding a second one day symposium on Biological Aerated Filters.

Over the last three years there has been a great deal of work on the development and optimisation of what has become one of the leading processes in wastewater treatment. The aim of this second symposium is to introduce recent work carried out in this field, bringing together many of the world's leading exponents of BAF technology and its application.

BAF2 represents an ideal opportunity to update your knowledge of these developments.

BAF2 Programme

9:30 Registration and coffee

10:25 Chairman's morning introduction

10:30 Trouble shooting and optimisation of BAF systems.
A Smith, Thames Water

11:00 Pilot scale comparisons of floating/sunken media and up/downflow BAFs.
A Mann, School of Water Sciences, Cranfield University.

11:20 Combined treatment of dairy and municipal wastewater in BAFs.
Howard Rundle, Tetra (Europe) Ltd.

11:40 North European experience of BAFs.
P Sagberg, Veas, Norway.
(to be confirmed)

12:00 The Poole Harbour wastewater treatment works.
P Brewer, Wessex Water Engineering.

12:30 Lunch

2:00 Chairman's afternoon introduction

2:10 The moving bed biological aerated filter
T Stephenson, School of Water Sciences, Cranfield University

2:30 Operational trials of different proprietary Lamella and BAF systems.
F Budge, Halcrow Consulting Engineers and D Gorrie, Grampian Regional Council.

3:00 Aeration optimisation in biological aerated filters.
P Pearce, Thames Water.

3:30 Operating performance and future development of the Biobed system.
A Cantwell, Brightwater Engineering.

4:00 *Close of Meeting and Tea*

The School of Water Sciences

The School of Water Sciences is the UK's only academic centre to specialise in process technologies for water and wastewater treatment. The school has considerable experience in research and development, working with many of the world's leading water companies and organisations concerned with water and effluent treatment. This experience ensures that the school if well positioned to offer consultancy and research and development related to these process technologies. The School has particular expertise associated with biotechnological applications including BAFs.

In addition to research and development and consultancy, the School of Water Sciences is recognised as a leading centre for the training of process technologies with funding from the EPSRC and approval of its programmes from the IChemE and CIWEM.

BAF and the biotechnology short courses have been developed to advance the understanding and implementation of these technologies.

School of Water Sciences

MODELLING BIOLOGICAL AERATED FILTERS FOR WASTEWATER TREATMENT.

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ABSTRACT

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

2. Two reactors were set-up to run in parallel treating settled domestic sewage using media identical in size and shape except one was less dense than water (relative density 0.92) and the other denser than water (relative density 1.05). The reactors were run upflow with liquid flowrates of 0.2 to 0.5 l min^{-1} and an air : liquid ratio of 10 : 1. After 4 weeks from start-up steady-state was reached. From this point, samples were taken at different heights along the reactors at timed intervals and profiles of sCOD removal against reactor height were produced. This analysis was repeated for a number of different flowrates and organic loadings (0.57 to 1.40 kg sCOD m^{-3} d^{-1}). The resulting data was then used with the empirical model, which was based on a first order reaction, to calculate the values of k* (overall process constant) and n (media factor). A much higher value of k* was found for the floating media (100) compared with the value found for the sunken media (33). This indicated the greater efficiency of sCOD removal in the floating media. The values of the media constants were similar showing the similarity in the media shape and size.

INTRODUCTION

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

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

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

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

7. In contrast to mechanistic models, empirical models merely relate input and output variables of a process to one another, i. e. relates hydraulic or organic loading rates to nutrient removal rate and does not require the measurement of parameters such as oxygen utilisation rate and mass transfer coefficients (Hamoda, 1989; Meunier and Williamson, 1981; Wu *et al.*, 1983). By measuring input and output variables the

assumptions that uniformity of biomass distribution and ideal plug flow do not need to be made (Eckenfelder and Barnhart, 1963; Wu *et al.*, 1983).

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

THEORY

9. Biological aerated filters and trickling filters show similar characteristics despite the fact that trickling filters are not submerged. Tracer studies carried out on trickling filters showed that the liquid undergoes plug flow movement through the beds (Sinkoff *et al.*, 1959). Similarly, tracer studies carried out to ascertain the hydrodynamic characteristics of BAF reactors showed a plug flow profile (Mann *et al.*, 1995). Thus the approach taken to design a model for BAF performance was similar to that used for trickling filters (Velz, 1948; Schulze, 1960; Eckenfelder *et al.*, 1961, 1963; Galler *et al.*, 1964; Logan *et al.*, 1987b). Following the design of a basic model, results obtained during biological treatment were used to examine the application of the model as in trickling filter studies (Harrison and Daigger, 1987; Logan *et al.*, 1987a; Matasci *et al.*, 1986).

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

$$\frac{dC}{dt} = kC \qquad (1)$$

which integrates to

$$\frac{C}{C_i} = e^{-kt} \qquad (2)$$

but t can be related to the volumetric loading, thus

$$t = \frac{k'}{B_V} \qquad (3)$$

where k' is the biomass constant which depends on the hydraulic characteristics of the reactor.

The volumetric loading can be re-written,

$$B_V = \frac{QC_i}{HA} \qquad (4)$$

thus

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

and

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

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

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

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

MATERIALS AND METHODS

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

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

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

Sampling took place each morning over this steady-state period, approximately 1 to 4 h prior to the daily backwashing.

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

RESULTS AND DISCUSSION

16. Start-up took four weeks of the total experimental period of 6 months. A maximum sCOD removal rate of approximately 90% was reached which dropped until steady state values of 66% sCOD removal for the sunken media and 78% sCOD removal for the floating media were attained. Over the steady state experimental period of 3 months these values varied from 60% to 75% for the sunken media and 71% to 85% for the floating media at the initial flowrate of 0.2 l min^{-1}. At increased flowrates the mean sCOD removal rate at steady state dropped, with only 36% sCOD removal for the sunken media and 40% sCOD removal for the floating media at a

Table 1 Influent and effluent process conditions.

	sCOD (mg l^{-1})	Suspended Solids (mg l^{-1})	Flowrate Q (m^3 d^{-1})	Dissolved Oxygen (mg l^{-1})	pH	Ammonia (mg l^{-1})
Influent						
Min.	80	80	0.29	0.8	6.4	15.2
Max.	210	120	0.58	1.6	8.1	30.6
Effluent						
Sunken						
Min.	16	19	0.29	5.5	6.2	14.4
Max.	148	104	0.58	8.6	7.9	30.4
Floating						
Min.	8	11	0.29	5.2	6	12.9
Max.	131	97	0.58	8.5	7.9	30.1

flowrate of 0.5 l min^{-1}. Over the experimental period influent conditions were maintained over a narrow range (Table 1). Effluent dissolved oxygen (DO) values remained at between 5.2 and 8.6 mg l^{-1} insuring oxygen availability did not become a limiting factor. Ammonia removal rates were low throughout indicating limited nitrification occurring. Though the model used was based on the sCOD values, the sBOD was also monitored over the experimental period. The BOD : COD ratio was found to be 0.45 for the influent and 0.20 for the effluent thus allowing the model constants to be found, if required, in terms of sBOD removal.

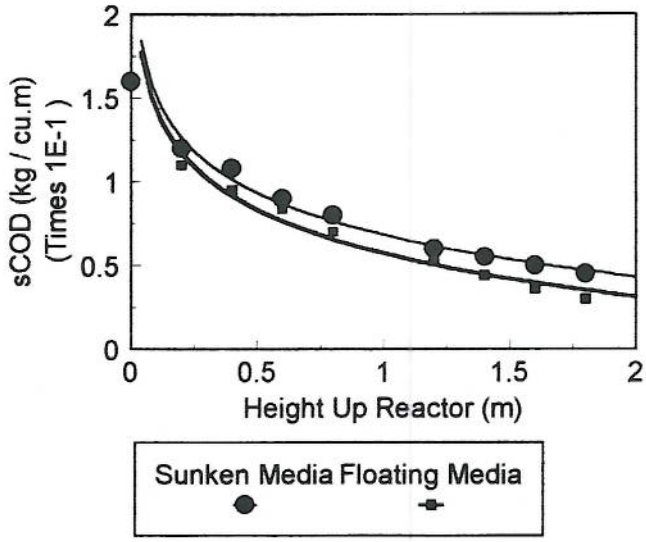

Figure 1 sCOD removal profile.

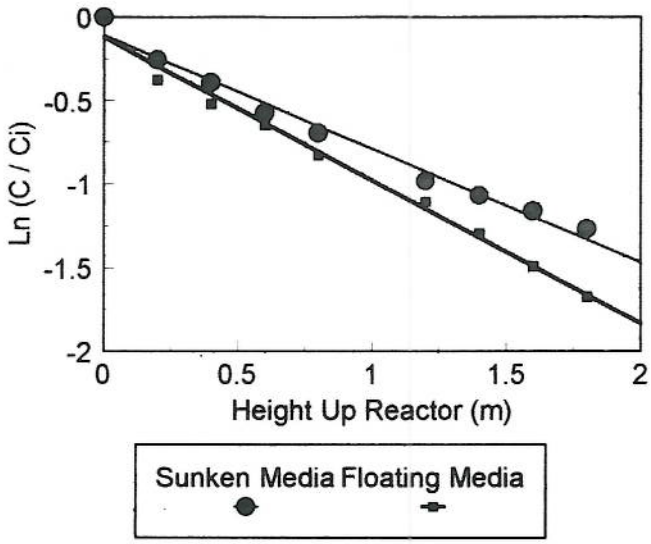

Figure 2 Logarithmic sCOD removal profile.

17. Figure 1 shows a typical sCOD profile obtained. To calculate values of n and k* nine sets of results were used with R^2 values between 0.978 and 0.993. R^2 indicates variation between the data points (dependant variables) and the regression or best fit line (independent variables). Values of R^2 range between 0 and 1 where values close to 1 indicate a close correlation between the points and the regression line. It was

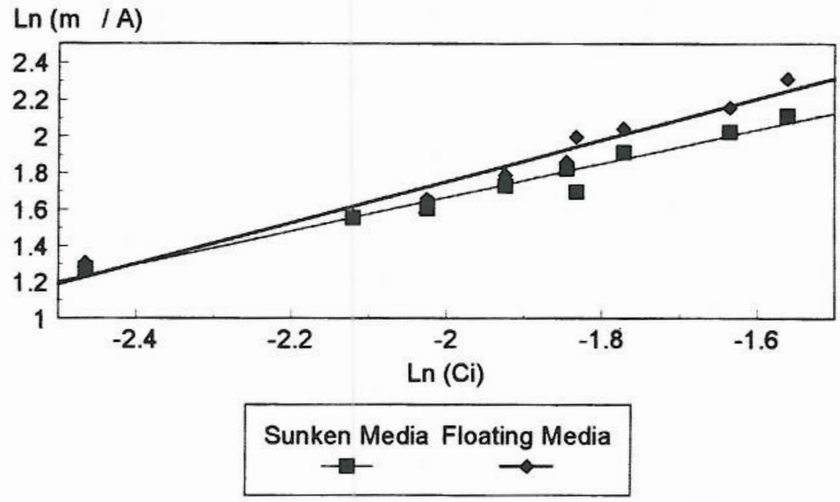

Figure 3 Graphical calculation of n and k*

demonstrated that the greatest removal rate occurred at the base of the reactors then decreased exponentially as the liquid moved up the reactors and the sCOD became a limiting factor in nutrient removal. The subsequent logarithmic graph of the same values (Figure 2) showed a close approximation to the 1st order equation (Equation 5). R^2 values for the logarithmic graphs varied from 0.981 to 0.995. This confirmed that the initial part of the model could be applied to this process. Analysis made at different organic loadings were then used to find the constants n and k^* (Figure 3) with R^2 values of 0.958 for the sunken media line and 0.964 for the floating media line.

	k*	n
Sunken Media	33	0.9231
Floating Media	100	1.1286

TABLE 2 Constant results.

18. The results from the model (Table 2) showed the difference in the constants between the two reactors. The k^* constant indicated the overall performance of the reactors in removing sCOD. The greater the value of k^* the lower the C/C_i values and thus a higher overall sCOD removal achieved. The large difference in values of k^* found from the experimental results indicated a better performance for the floating media than for the sunken media, which was confirmed by sCOD removal values monitored daily over the experimental period. In the case of the media constants, similar values were obtained which indicated the similarity in the media designs. Removal of sCOD fitted the model well and over the period of 6 to 24 h from backwashing there was only a slight variation in sCOD removal rates. After backwashing the effluent solids were high, often greater than the influent concentration, due to dislodged solids leaving the reactors. After this period the effluent solids concentration dropped as biomass growth caused increased solids

capture and the remaining dislodged solids left the reactors in the effluent. This continued until the concentration began to increase again, prior to backwashing, through sloughing of biomass (Lewandowski *et al.*, 1994; Cao and Alaerts, 1995). The mechanics of particulate matter removal differ from those of dissolved organic matter with indications that the two interact through particulate adsorption hindering the dissolved organic removal rate (Sarner, 1986). A great variation in the effluent solids over a 24 h period and only slight variation in the sCOD removal rate indicated that the suspended solids concentration did not affect sCOD removal efficiency. Thus no correction to the model was deemed necessary.

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

20. The difference in media types is shown by the value of n, the media constant. The values found for the sunken and floating media were similar (0.9231 and 1.1286) which would be expected for media similar in shape and surface texture. Since the only variation between the medias was the density, the difference in n values may be attributed to this and its consequential effects on the flow through the reactors.

21. The use of this model thus allows a direct comparison between media performance and gives a simple method to ascertain the ideal media type to be used for a process and minimum height needed for a required discharge concent. The values of k* and n are specific to the media, under specific conditions of flow velocity and direction, temperature and aeration rate. Work on different granular media types give an improved understanding of the effect that shape, size and surface texture have on the values of the constants k^* and n and on the comparative performance (Eckenfelder and Barnhart, 1963; Harrison and Daigger, 1987). In addition constants could also be found for the media types under different process conditions such as downflow and with different aeration rates as this affects the amount of mixing within granular reactors and subsequently the performance in nutrient removal (Mann *et al.*, 1995). Thus these experiments are required to obtain results that may be used to predict the ideal media required for individual plants. They would also allow the performance of media types to be predicted purely on their physical properties (Eckenfelder, 1961).

NOMENCLATURE

C	sCOD concentration	kg m^{-3}
C_i	Influent sCOD concentration	kg m^{-3}

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

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