CRANFIELD UNIVERSITY

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PROCESS MODELLING OF THE MBBR / AS HYBRID PROCESS

SCHOOL OF WATER SCIENCES

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Process modelling of the MBBR / AS hybrid process

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Dedicated to Ciarán Andrew Berry - a source of more sleepless nights than this thesis ever was

ABSTRACT

With the UK water companies under increasing financial and environmental pressures they are having to look for innovative options to provide lower capital and operational cost solutions to asset management in order to maintain profits.

Process modelling provides a tool which can be used to help identify more cost effective options through a better understanding of processes and their limitations. Although modelling can be used in this way, sound judgement needs to be applied in using models as misinterpretation of simulated results could lead to the wrong option being selected.

Processes which combine fixed and suspended biomass have been used to improve the performance of existing wastewater treatment works which previously had either a fixed or a suspended growth process only. In addition to improved efficiency, the processes have been found to be more robust.

Full scale trials of a fixed film / activated sludge hybrid process, supported by model simulation, have shown that the hybrid process would enable a 120 000 PE sewage treatment works to meet a new, stricter effluent discharge consent without the construction of new tanks.

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ABBREVIATIONS

AAO - Anaerobic/anoxic/oxic

ALPHA- Alternate phase - step feed process

ASP - Activated sludge plant

ATU - allyl thiourea

BAF - Biological aerated filter

BNR - Biological nutrient removal

BOD_5 - 5 Day Biochemical oxygen demand

CAPEX - Capital expenditure

COD - Chemical oxygen demand

COPI - Construction price index

CSTR - Completely stirred tank reactor

DO - Dissolved oxygen

F: M ratio - Food : micro-organism ratio

FST - Final settling tanks

HRT -Hydraulic retention time

MBBR - Moving bed biofilm reactor

MBBR-AS - Moving bed biofilm reactor - activated sludge

MBBR-SCR -Moving bed biofilm reactor - solids contact reaeration

MLSS - Mixed liquor suspended solids

NDBEPR - Nitrification/denitrification biological enhanced phosphorus removal

NO_x-N - Oxidised nitrogen species

OPEX - Operating expenditure

OUR - Oxygen uptake rate

Plc - Programmable logic controller

RAS - Return activated sludge

RBC - Rotating biological contactor

RDN - Reaeration denitrification nitrification process

RNC - River needs consent

RTD - Residence time distribution

SOTE - Standard oxygen transfer efficiency

SRT - Sludge retention time

SSVI - Stirred specific volume index

STW - Sewage treatment works

SVI - Sludge volume index

TF/SC - Trickling filter/solids contact

TKN - Total Kjeldahl nitrogen

TOC - Total organic carbon

TSS - Total suspended solids

UPM - Urban pollution management

UWWTD - Urban wastewater treatment directive

VSS - Volatile suspended solids

WOC - Water only companies

WSC - Water and sewerage companies

NOMENCLATURE

μ_{max-A} - Maximum specific growth rate - autotrophs

μ_{max-H} - Maximum specific growth rate - heterotrophs

S_A - Volatile acids/fermentation products

S_I - Soluble inert non-biodegradable organics

S_{ND} - Soluble organic nitrogen

S_{NH} - Biodegradable ammoniacal nitrogen

S_s - Readily biodegradable substrate

V_{ml} - Volume of activated sludge mixed liquor

V_{ww} - Volume of wastewater

X_{AUT} - Autotrophic nitrifying biomass

X_H - Heterotrophic biomass

X_I - Particulate inert non-biodegradable organics

X_{ND} - Particulate organic nitrogen

X_{PAO}- Phosphorus accumulating organisms

X_{PHA} - Stored polyhydroxyalkanoate

X_S - Slowly biodegradable substrate

Y_A - Autotrophic yield coefficient

Y_H - Heterotrophic yield coefficient

CHAPTER 1 - INTRODUCTION AND OBJECTIVES

1.1 INTRODUCTION

1.1.1 The Moving Bed Biofilm Reactor (MBBR) process

Suspended growth systems have been used at wastewater treatment works for treatment of organic carbon and nutrients but they are not without their problems. Problems have been encountered with settling of the suspended biomass, the reactors are large and there are costs associated with recycling the biomass.

Traditional biofilm processes such as trickling filters have a high footprint and although nitrification has been achieved the process security for meeting very low effluent ammonia concentrations is not significant, especially under low temperature conditions. Another problem associated with these processes is seasonal sloughing of biomass which can cause non-compliance with effluent suspended solids consents.

Recent advances with biofilm processes such as aerated submerged biological filters and biological aerated filters have enabled nitrification to be achieved in a small footprint processes. Nitrification rates of up to 66 mg NH^4-N m^{-2} d^{-1} have been reported in two stage submerged aerated filters (Rusten, 1984) although it was noted that the organic load varied too much to achieve optimal nitrification conditions.

Industries such as dairies produce high strength wastewaters which can overload municipal wastewater treatment plants if discharged to the sewer. Submerged aerated filters have been used as compact high rate processes treating high organic loads (up to 70 g COD m^{-2} d^{-1}) to achieve COD reductions of around 80 % (Ødegaard and Rusten, 1990).

Problems associated with high rate biofilm processes include the poor settling of the biomass produced, the head losses across the process and the control required can be quite complex especially if backwashing of filter beds is required. High maintenance requirements and large volumes of backwash water requiring treatment are also associated with these processes. The use of a 2 stage process has been found to improve the settling characteristics of the solids produced (Rusten, 1984).

The moving bed biofilm process developed by Kaldnes Milijøteknologi uses biofilm carrier elements which move freely within the reactor, being kept in suspension by the process air or a mixer. Coarse bubble diffused aeration is used to provide the process air. The carriers are retained in the reactor by sieves on the outlet. The process has low head losses, filter bed channelling does not occur and backwashing is not required. Excessive biofilm growth is controlled by the air bubbles eroding the biofilm, preventing clogging of the media and caused turbulence, ensuring good contact between the substrate and the biomass (Rusten *et al.*, 1994a). A further advantage of the process is that it is relatively simple to retrofit into an existing activated sludge plant to upgrade processes within the existing reactor volumes.

Different volumetric fills of the biofilm carrier can be used to provide different process capacities up to a maximum fill of 70% Ødegaard *et al.* (1994). The carriers have an effective specific surface area of 500 m^2 m^{-3} giving a maximum reactor specific surface area of 350 m^2 m^{-3} for a 70% fill or 250 m^2 m^{-3} for a 50% fill.

The process can be used in different configurations to achieve organic carbon removal, nitrification and denitrification of both municipal and industrial wastewaters.

Organic Carbon Removal

In pilot tests, Pastorelli *et al.* (1997) looked at carbon removal and nitrification in MBBRs , finding a linear relationship between filtered COD removal and filtered COD loading for loading rates up to 8 gCOD m^{-2} (media surface area) d^{-1} and dissolved oxygen concentrations greater than 2 mg l^{-1}. In aerated submerged biological filters, COD removal of up to75 % has been reported for a load of 5 g COD m^{-2} d^{-1} (Rusten, 1984). The removal rate decreased with increasing load, only 57 % removal being achieved at a load of 120 g COD m^{-2} d^{-1}. The same author reported little influence of temperature in the range 10 - 20 °C concurring with the findings of Ødegaard *et al.* (1994), however in experiments to identify the true temperature dependence of the process, a temperature coefficient similar to that for ASPs was observed (Rusten *et al.*, 1994a).

At small plants in Norway MBBRs have been used with chemical precipitation to achieve removal of organics and phosphorus. For organic loads of up to 6.4 g COD m^{-2} d^{-1}, COD removals of greater than 90 % were achieved with total sludge production of 0.55 g TS (g COD)$^{-1}$ (Rusten *et al.*, 1997; Ødegaard *et al.*, 1993). Sludge production of between 0.35 and 0.55 g TS (g COD)$^{-1}$ removed was observed for single stage and 2 stage reactors, the solids production increasing with increased COD removal (Rusten *et al.*, 1994a).

Ødegaard (1994) reported the use of MBBRs to treat high strength industrial wastewaters such as those from the dairy (85% COD removal at 500 g COD m^{-3} d^{-1}), potato chips (97.5% COD removal at 4 kg COD m^{-3} d^{-1}) and pulp and paper industries (up to 70% COD removal at 50 kg COD m^{-3} d^{-1}). Although loads of up to 50 kg COD m^{-3} d^{-1} were applied successfully to MBBRs, it was found that at very high loads (>60 kg COD m^{-3} d^{-1}) inhibition occurred. The effective F:M achieved in some plants was up to 1.6 kg COD kg^{-1} TS d^{-1}, a higher loading than usually applied to activated sludge plants. Two MBBR reactors in series following an equalisation basin was found to achieve up to 95 % removal of soluble COD from poultry processing wastewater with an overall organic loading of 60 g COD m^{-2} d^{-1} (Rusten *et al.*, 1998).

Very high organic loads are also a characteristic of pulp and paper industry wastewaters, COD concentrations of up to 27 000 mg l^{-1} being observed. At four different pulp and paper processing plants pilot trials were carried out to assess the effectiveness of MBBRs for treating the wastewaters (Rusten *et al.*, 1994b). COD removal efficiencies between 38 and 78 % were observed at volumetric loading rates of up to 40 kg COD m^{-3} d^{-1}. The success of the pilot trials has led to implementation of full scale plants to treat these wastewaters. Dalentoft and Thulin (1997) reported the use of a MBBR in conjunction an activated sludge plant in a 2 stage process to treat forestry industry wastewaters. In a full scale plant the 1st stage MBBR was loaded at 10-15 kg COD m^{-3} d^{-1} and the following ASP at up to 0.4 kg BOD_5 (kg biomass COD)$^{-1}$ d^{-1}. It was found that the use of this configuration improved the settleability over a conventional ASP and was a stable process, recovering quickly from shock upsets.

Nitrification

Pastorelli *et al.* (1997) found nitrification in MBBRs to be nearly first order with respect to DO, implying liquid film diffusion limitation (Hem *et al.*, 1994). No nitrification took place with a DO concentration less than 2.08 mg l^{-1}. This finding concurs with those of Hem *et al.* (1994) who found that nitrification rate was dependent on the oxygen concentration for oxygen to ammonium ratios below 2 g O_2 (g NH_4-N)$^{-1}$. At oxygen to ammonium ratios above 5 g O_2 (g NH_4-N)$^{-1}$ nitrification rates was dependent only on the ammonium concentration. Ødegaard *et al.* (1994) also reported 2-3 mg l^{-1} as the lowest DO concentration at which nitrification would occur in MBBRs.

The linear relationship between nitrification and DO concentration can lead to high energy costs but has the advantage that DO concentration can be used to control the process (Pastorelli *et al.*, 1997). A further advantage is the reduction of the adverse effects of low temperatures on nitrification. Nitrification rate decreases with temperature but oxygen solubility increases with decreasing temperature, the increased DO off setting the temperature effects. As a result of this small temperature influences have been reported for temperatures between 8 and 16 °C (Ødegaard *et al.*, 1994).

Hem *et al.* (1994) also found that nitrification was affected by the organic load on the MBBR. At high organic loads, > 5 g total BOD_7 m^{-2} d^{-1} there was no nitrification observed due to the growth of heterotrophs outcompeting the nitrifiers. At organic loads of 1-2 g total BOD_7 m^{-2} d^{-1} the nitrification rate observed was 0.7-1.2 g NO_x-N m^{-2} d^{-1}, decreasing to 0.3-0.8 g NO_x-N m^{-2} d^{-1} for loads of 2-3 g total BOD_7 m^{-2} d^{-1}. Ødegaard *et al.* (1994) found that for loads below 10 g COD m^{-2} d^{-1} nitrification occurred but that it was inhibited by the presence of organic matter.

The MBBR process has been used to upgrade existing activated sludge plants for nitrogen removal. A reported 80 - 90 % total N removal was achieved for an empty bed HRT of 2.6 hours with a specific nitrification rate of 1 - 1.3 g NH_4-N m^{-2} d^{-1} being observed (Rusten *et al.*, 1994a). The sludge production observed in this plant was 0.36 kg TS (kg COD)$^{-1}$.

Denitrification

Denitrification rates of 2.2 g NO_3^--N m^{-2} d^{-1} have been achieved in MBBRs which needed very little acclimatisation to the denitrification (Pastorelli *et al.*, 1997). The short acclimatisation period required was attributed to simultaneous denitrification occurring in the deeper regions of the biofilm. A similar denitrification rate, 2.5 g NO_3^--N m^{-2} d^{-1}, was reported by Rusten *et al.* (1994a) at a site where an existing conventional activated sludge plant with insufficient retention for nitrification had been upgraded using the MBBR process. Ødegaard *et al.* (1993) report a denitrification rate of only 1.25 g NO_3^--N m^{-2} d^{-1} although it was suggested that denitrification might have been biomass limited - highlighted by the presence of nitrite. The denitrification rate could be improved by increasing the fill of media and by improved hydraulic control of the plant.

In a comparison of pre and post-denitrification using MBBRs, specific denitrification rates of < 0.4 g NO_x-N m^{-2} d^{-1} and 2.0 - 2.2 g NO_x-N m^{-2} d^{-1} were measured for pre-denitrification and post-denitrification, respectively (Rusten *et al.*, 1995). The low rate for pre-denitrification was attributed to the low concentration of readily biodegradable COD in the influent wastewater. This was not a problem for the post-denitrification as an external carbon source was used with the rate of addition controlled to maintain an optimum C:N ratio in the anoxic reactor.

1.1.2 MBBR / AS plants

A number of pilot trials have found promising results by combining the MBBR with an activated sludge process to treat municipal wastewaters. Nitrification has been achieved with sludge ages as low as 4 d for the AS stage (Jaouen, 1996; Lievre, 1997; Michel, 1998). These pilot results and those from full scale applications of other processes combining fixed and suspended growth indicate that the use of the MBBR / AS plant at full scale could be used to upgrade plants with smaller reactors than would be required with conventional activated sludge plants (ASPs).

1.2 OBJECTIVES

1. To investigate the use of combined fixed film / activated sludge process for upgrading existing wastewater treatment plants.

2. Assess the benefits to be obtained through the application of wastewater treatment process models.

3. Through the use of a full scale trial, assess the suitability of the MBBR / AS hybrid process for upgrading an existing carbonaceous treatment plant to ammonia removal within existing tank volumes.

4. Assess whether a recently developed process model of the MBBR process sufficiently describes process performance for it to be used in process design.

5. Simulate future process performance of a full scale hybrid retrofit at Colchester STW using the calibrated process model in order to support the findings of the trial.

6. Investigate the major business issues within the regulated water industry in the UK, identifying how process modelling and the development of hybrid processes can help a company meet the business challenges it faces.

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CHAPTER 2 - LITERATURE REVIEW

2.1 HYBRID FIXED FILM AND SUSPENDED GROWTH SYSTEMS FOR WASTEWATER TREATMENT

2.1.1 Introduction

Hybrid systems have been developed to utilise the benefits of both fixed film and suspended biomass. Taking a modelling approach, Nicol *et al.* (1987) predicted that the hybrid systems maintained several advantages over conventional activated sludge systems. The advantages included resistance to failure under large hydraulic surges when suspended biomass would be washed out of the system; maintained nitrification in periods of inhibition, e.g. due to low temperatures or toxic pollutants; stable carbon oxidation and nitrification at shorter HRTs and low SRTs for the suspended biomass..

Other benefits of hybrid systems that have been put forward include improved settleability of the activated sludge, improved plant performance within existing volumes, denitrification within existing aerobic volumes, reduced loads on final clarifiers, improved treatment of slowly biodegradable pollutants such as insecticides.

2.1.2 Upgrading of activated sludge plants

The performance of activated sludge plants are related to the F:M ratio of the plant. This ratio can be decreased by increasing the biomass concentration within the process. This can be done by maintaining the same MLSS concentration and increasing the volume of the plant or by increasing the MLSS concentration for the same size plant. Increasing the size of the plant is an expensive option in terms of the capital expenditure and there may also be space limitations. The extent to which the MLSS concentration can be increased is limited by the capacity of the final settling tanks. If the MLSS concentration is too high the biomass will not settle in the FSTs and will be washed out with the effluent.

By incorporating fixed biomass into the activated sludge plant the total biomass concentration is increased. However, as there is no increase in suspended biomass the FSTs are not overloaded. In pilot scale trials, Hegemann (1984) found that by introducing porous plastic carriers into the activated sludge tank (the Linpor process) the overall biomass dry solids could be increased from around 2.5 kg/m^3 in a system with no carriers to up to 8 kg/m^3. The fixed biofilm concentration achieved was reported to be as high as 15 kg/m^3. The amount of carrier varied up to 40 % of the reactor volume. The increased biomass in the system resulted in lower F:M ratios and hence improved process performance. Similar results have been reported for the Linpor process in full scale applications (Morper and Wildmoser, 1990). In a 13 100 m^3 plant adding the biomass carrier up to 30 % of the volume achieved a F:M ratio less than 50 % of that without fixed biomass for a similar volumetric loading. As a result the final effluent BOD_5 and COD were 47 % and 30 % lower respectively for the combined fixed film / suspended growth system. Gebara (1999) found that the introduction of plastic nets as a biomass support medium into a 28 l pilot activated sludge tank resulted in improved BOD removal and plant stability. A mathematical model of the process was developed and verified against experimental data. Applying the model led to the conclusions that the use of the plastic nets could lead to plant upgrades within existing reactor sizes and that smaller tanks could be used at new plants.

2.1.3 Nitrification in hybrid plants

As discharge consents are tightened many wastewater treatment plants are having to introduce nutrient removal, i.e. nitrification, denitrification and phosphorus removal. In order to upgrade an existing activated sludge plant (ASP) to nitrification the sludge retention time (sludge age) has to be increased. This can be achieved by increasing the mixed liquor suspended solids concentration or by extending the plant volume. Increasing the MLSS can be limited by the capacity of the final settlement tanks since the mass loading will increase with increasing MLSS. Increasing the volume of the aeration tanks can entail very high capital costs which make it a less favourable option.

By incorporating a fixed film biomass in the aeration tanks the total biomass and therefore sludge age, is increased. However because the additional biomass is retained in the aeration tank there is no additional load on the final settlement tanks. The capital cost of retrofitting biofilm carriers compares favourably with that of building additional plant volume.

The observation of Nicol *et al.* (1988) that hybrid systems should maintain improved nitrification under transient inhibitory conditions has been supported by the findings at some hybrid plants- both full scale and pilot scale.

In a full scale trial to achieve nitrification within existing reactor volumes, Bonhomme *et al.* (1990) installed Biofix media in 20% of a 650 m^3 activated sludge tank. It was found that although there were no nitrifiers in the biofilm ammonia conversion was achieved at rates of 0.12 - 0.15 kg m^{-3} d^{-1} . The nitrification was attributed to the increased total biomass in the system. Barnes (1986) reported on trials of the Captor process in which biofilm carrying pads were installed in the first 25% of the activated sludge reactor. It was found that the fixed film stage, by reducing the organic load by up to 80%, made the rest of the tank available for nitrification. Pilot scale systems incorporating different biomass carriers were tested against a control by Sen *et al.* (1994). It was found that the hybrid systems achieved up to 50% greater nitrification than the control. Under steady state conditions all systems achieved complete nitrification at sludge ages of 3.4 d at temperatures of 12 °C. However, when transient ammonia spikes were introduced, the systems with fixed biomass achieved better nitrification. After pilot testing of 3 biomass carriers Ringlace was selected for installation at a full scale plant, 475 m m^{-3} being installed (Lessel, 1994). Nitrifiers were found in both the fixed and the suspended biomass and up to 97% ammonia removal was achieved at temperatures down to 8 °C. The fixed film was found to offer a buffer against process disturbances at 10 °C. Jones *et al.* (1998) reported installation of Ringlace and Biomatrix media in a 225 m^3 reactor, the plant achieving better nitrification than the control plant at winter temperatures. The nitrification rate was found to be 0.42 kg d^{-1} per 1000m of media. A higher nitrification rate of 1.7 kg d^{-1} per 1000m of Ringlace media was observed by Randall and Sen (1996) in full scale trials to achieve nitrification within an existing low HRT activated sludge plant. It was estimated that the use of biofilm carriers would reduce the capital cost of achieving nitrogen removal from US $24M to $9.2M. Lab scale work by Wanner *et al.* (1988) found that >70% ammonia removal could be achieved in hybrid systems and that the nitrification was independent of SRT due to the predominance of nitrifiers in the biofilm. Contrary to the findings of Bonhomme *et al.* (1990), Wanner *et al.* (1988) found that there was no nitrification in the suspended biomass but 40 mg g^{-1} h^{-1} was achieved in the biofilm. An existing activated sludge plant has been upgraded to achieve 50 % nitrification by converting one of the two reactors to a Linpor biofilm process resulting in a 2 stage hybrid system with fixed film nitrification (Morper and Wildmoser, 1990).

2.1.4 Denitrification in hybrid plants

The use of biofilms in activated sludge systems has also been put forward for achieving denitrification. Barnes *et al.* (1988) proposed that the oxygen transfer limitation in thick biofilms, resulting in anoxic conditions within the fixed film, could result in denitrification in existing nitrification reactors. Lab scale experiments with flyscreen meshes as biofilm carriers supported this with > 60 % total Nitrogen removal being achieved in an aerobic reactor. The benefits of this process are upgrading to total N removal within existing reactor volumes and recovery of alkalinity within the nitrification reactor with no increase in operational costs. Hao *et al.* (1995) investigated the used of fibrous biofilm carriers for use in denitrification. The denitrification rates achieved, up to 2.4 mg N g^{-1} VSS d^{-1}, were significantly greater than in control experiments with no fixed biomass. The higher rates were explained by the significantly higher total biomass in the system, estimated to be between 30 and 60 times greater than in the control experiments. In lab scale experiments to evaluate different media for hybrid nitrification reactors, Sen *et al.* (1994) also found that some denitrification was achieved within the fixed film under aerobic conditions. Similar observations were made by Randall and Sen (1996) in full scale trials of integrated fixed film activated sludge systems where 88 % of the nitrate formed in the media section was also denitrified within this zone. This would enable a reduction in size of the anoxic zone. Runart (1986) reported > 75 % total N removal in summer conditions in the Stählermatic system.

2.1.5 Biological phosphorus removal in hybrid plants

Biofilms in the form of RBCs have been incorporated in pilot scale biological nutrient removal plants. Chuang *et al.* (1997) installed an RBC in the aerobic zone of a BNR plant with the aim of maintaining nitrifiers within the aerobic zone at lower SRTs, enabling more phosphorus to be removed in the surplus activated sludge. Partial nitrification (42 %) was achieved at a sludge age of 5 days. Su and Ouyang (1996) also installed a RBC in the aerobic zone to maintain nitrification at reduced sludge ages.

However, they also used fully submerged rotating discs in the anaerobic and anoxic zones. It was hypothesised that the anaerobic biofilm would hydrolyse slowly biodegradable COD to readily biodegradable COD and short chain fatty acids resulting in improved P release in the anaerobic zone and denitrification. Removal rates achieved in the pilot plant were 97 % and 70 % for total phosphorus and total nitrogen respectively.

By using a hydrophilic gel biofilm support media in the aerobic zone of an anaerobic anoxic oxic (AAO) BNR plant, Mishima *et al.* (1996) aimed to achieve biological nutrient removal in a bioreactor half the size of a conventional AAO plant. At an aerobic sludge age of 2.7 - 5.7 d final effluent concentrations of 8.3 mg l^{-1} total nitrogen and 0.6 mg l^{-1} total P were achieved. Due to the low aerobic sludge age and reduced nitrification below 15°C observed in batch tests, it was estimated that 93 % of the nitrification occurred within the biofilm.

Other research has focused on using fixed film carriers in the anoxic zone of a BNR plant in order to maximise denitrification, > 90 % removal being achieved when the NO_X-N load was less than 0.4 kg m^{-3} d^{-1} (Liu *et al.*, 1996). As well as providing optimum conditions for denitrification, by feeding most of the influent wastewater to the anoxic zone, around 52 % of the COD was removed in the anoxic zone, reducing the aeration requirements in the aerobic zone. Total P removal was also greater than 90 %.

2.1.6 Sludge settleability in hybrid plants

Some authors have reported improved settling of activated sludge which also has biofilm carriers. Hegemann (1984) reported improved SVIs with the use of Linpor in an ASP which meant higher MLSS could be used without overloading the clarifiers. Similar results with Linpor were reported by Morper and Wildmoser (1990). In experiments with different levels of fixed biomass in an activated sludge reactor, Gebara (1999) found that the SVI of the suspended biomass decreased with increasing fixed biomass. Use of Ringlace in full scale trials was found to improve the SVI from 198 ml g^{-1} to 42 ml g^{-1} enabling the use of a higher MLSS concentration (Lessel, 1994). Jones *et al.* (1998) made similar observations at a plant also using Ringlace but also with Biomatrix. A sludge volume index of < 80 ml g^{-1} was reported for the Stählermatic process, improved sludge quality being highlighted as one of the benefits of the system (Runart, 1986). Wanner *et al.* (1988) carried out experiments to explain the improved settling observed in integrated fixed film activated sludge systems. It was found when the fixed biomass exceeded 50 % of the total biomass in the system there were less filamentous bacteria present which led to lower SVIs - the fixed film effectively acted as a selector. Although some authors have reported that the settling improved with hybrid systems the same was not found by Bonhomme *et al.* (1990).

2.1.7 Problems encountered in hybrid plants

Some problems have been encountered with combined fixed film activated sludge systems. Some authors have reported the presence of worms in systems using Ringlace (Jones *et al.*, 1998 ; Lessel, 1994). Remedial measures have had to be implemented to control the worms. The higher biomass concentrations in smaller volumes will result in higher volumetric oxygen requirements and therefore the possibility of oxygen transfer problems (Nicol *et al.*, 1988). This needs to be considered when designing aeration systems for hybrid processes. When installing biomass carriers in activated sludge plants it is important to get even distribution of wastewater to all of the carriers. This is less of a problem for free moving carriers such as Linpor and Kaldnes media. In systems with stationary media support such as Ringlace and Biofix, double spiral roll aeration systems have been employed to prevent short circuiting and to ensure good air distribution to the biofilm (Bonhomme *et al.*, 1990 ; Randall and Sen, 1996).

2.1.8 Hybrid plants for treatment of persistent toxic chemicals

Nicol *et al.* (1988) predicted that activated sludge plants incorporating fixed biomass would be more resistant to toxic chemicals than a conventional ASP. Supporting this theory is the use of hybrid systems to treat toxic slowly biodegradable chemicals (Chen *et al.*, 1998). The work of Chen *et al.* (1998) proposed such a hybrid system for the treatment of trichlorfon, an organophosphate insecticide. Glucose had to be added as it was required by the trichlorfon degrading bacteria. It was thought that the slowly biodegradable chemical would be removed mainly in the fixed biomass but it was found that the suspended biomass was responsible for trichlorfon degradation. However, a model developed by the same authors was used to predict that for very slowly biodegradable chemical with a high diffusivity fixed biomass would give an advantage to the treatment system.

2.1.9 Two stage hybrid plants

If additional volume has to be built, it can be reduced by using a biofilm process. A number of plants have been built where a biofilm process has been installed ahead of the ASP to enable the ASP to achieve nitrification or where a small ASP has been built as a nitrification stage after a filter plant.. Different hypotheses have been put forward for how these two stage processes achieve nitrification. Daigger *et al.* (1991) proposed that nitrifiers which started to grow in the first stage filters of a trickling filter - solids contact (TF-SC) process seeded the activated sludge enhancing the nitrification in the second stage.

Other work has found no presence of nitrifiers in the first stage but hypothesises that the biofilm process reduces the organic load on the activated sludge stage which increases its nitrification potential. Sunner *et al.* (1998) reported the use of a two stage MBBR-AS process at 2 sewage treatment works with a high percentage of industrial effluent. It was found that removing 75 - 80% of the BOD in a first stage MBBR reactor reduced the load a subsequent ASP to an F:M ratio of 0.11 kg BOD (kg TSS)$^{-1}$ d^{-1} allowing it to nitrify. Successful pilot studies resulted in full scale plants being built. The design loads to the MBBRs were 4 and 5 kg BOD m^{-3} d^{-1} (equivalent to 16 and 20 g BOD m^{-2} d^{-1}) respectively which is significantly higher than the 0.5 kg BOD m^{-3} d^{-1} that overloaded the existing 1st stage trickling filters at one of the sites. The process used an intermediate settling stage between the two biological stages to reduce the solids load to the ASP stage. Pilot trials of an MBBR / AS hybrid plant found that no nitrification was achieved in the MBBR but that the nitrification rates in the subsequent ASP were greater than those observed in conventional ASPs (Lievre, 1997; Michel, 1998)

The 2 stage MBBR-AS plant compared favourably in cost terms to an ASP with reported lower capital and operational costs and a significantly smaller footprint. Sludge production was reported to be 33 higher than for a conventional ASP system but good settling properties were reported with average SSVIs as low as 52 ml g^{-1} (Sunner *et al.*, 1998). Dalentoft and Thulin (1997) reported similar higher sludge production for a 2 stage MBBR-AS relative to that of a single stage ASP but indicated that the capital cost for the 2 stage process was similar to that of an ASP and the operating costs were higher for the 2 stage process. The difference in OPEX comparisons could be due to the nutrients required to be added in the latter example and the different geographical location. The plant described by Dalentoft and Thulin (1997) did not have an intermediate settling stage between the MBBR and the ASP.

Detailed comparisons of three processes, oxygen activated sludge, air activated sludge and trickling filter / solids contact (TF/SC) resulted in the TF/SC process being selected for upgrading a large works in Canada (Parker *et al.*, 1994). The process was selected due to it being a robust process that was easy to operate and yet it had the lowest total present value of the three options, due mainly to lower operating and maintenance costs. Pilot scale testing of an MBBR solids contact reaeration (MBBR/SCR) process gave results comparable to those of a TF/SC process but with much higher volumetric loading on the first stage, 5 kg BOD_5 m^{-3} d^{-1} compared with 1.6 kg BOD_5 m^{-3} d^{-1} (Rusten *et al.*, 1996). This would enable a stable process similar to the TF/SC plant but with a smaller footprint for the first fixed film stage.

2.1.10 Summary

The relative performance of some plants that have utilised fixed and suspended growth are summarised in Table 1. The use of such systems can give a robust option to upgrade wastewater treatment plants with minimum additional build, whether it be improving carbonaceous treatment plants or upgrading to nitrogen and phosphorus removal. Significant savings in capital expenditure have been identified through the use of fixed film activated sludge systems and in many cases more stable operation has been reported.

The sludge ages and F:M ratios reported in Table 1 are based on the suspended biomass only due to the difficulty of quantifying the fixed biomass in some systems.

Table 1. Performance summary of different plants combining fixed and suspended growth systems.

	Biofilm carrier in AS tank				2 stage fixed growth - suspended growth processes			
	RBC - AS[1]	Ringlace[2]	Linpor[3]	FM-AS[4]	TF-SC[5]	TF-SC[6]	MBBR-AS[7]	MBBR-AS[8]
Scale	Pilot scale	Full scale	Full scale	Pilot scale	Full scale	Full scale	Pilot scale	Pilot scale
Biofilm loading (kgBOD m^{-3} d^{-1})		...	...	6.7	1.42	0.43	3.96	5.0
Media fill (%)		26	20		...	...	...	50
F:M ratio (kgBOD kg^{-1}MLSS d^{-1})	0.25 (VSS)	0.09	0.19	0.28	...	...	0.21	0.16
Sludge age (MLSS only) (d)	14.5	...	...	...	...	2.2	2.7	4
SVI (ml g^{-1})	...	56	...	...	...	72	...	...
$SSVI_{3.5}$ (ml g^{-1})	...	...	...	...	...	...	...	173
BOD removal (%)	95 (sBOD)	94	3 mg l^{-1}	95	98	96	99	96
COD removal (%)	91	87	85	88	...	...	...	...
TSS removal (%)	...	...	...	88	96	92	96	91
NH_4-N removal (%)	99	83	60	95	...	98	94	99

[1]Su and Ouyang (1996); [2]Lessel (1994); [3]Morper and Wildmoser (1990); [4]Han-Chang and Thomas (1993); [5]Parker *et al.* (1993); [6]Daigger *et al.* (1991); [7]Sunner *et al.* (1998); [8]Michel, (1998).

2.2 APPLICATION OF PROCESS MODELS

2.2.1 Introduction

As wastewater treatment processes have been developed, mathematical models of the processes have been proposed in order to describe process performance and to gain a better understanding of the processes involved. Since 1970 when Lawrence and McCarty (1970) proposed using Monod kinetics to model biological processes, activated sludge modelling has been the focus of much research. The Lawrence and McCarty model has been modified and expanded, notably by Marais and Ekama (1976), Ekama and Marais (1979) and Dold *et al.* (1980), culminating in the publication of the IAWPRC Activated Sludge Model No.1 (Henze *et al.*, 1987). The IAWPRC (now IAWQ) model aimed to review existing models and select the simplest model for realistic prediction of single sludge processes. It was also intended as a focus for activated sludge modelling and a starting point for further model development. Since its publication the IAWQ model has been extended to biological phosphorus removal (Henze *et al.*, 1995) and modified to model different processes, e.g. sequencing batch reactors (SBRs) (Oles and Wilderer, 1991) and pure oxygen activated sludge (Yuan *et al.*, 1994).

The complex mathematical models which have helped in understanding how processes work but if they are to be applied by engineers to real life situations they need to be in a format which makes them easy to use. Many of the models have been incorporated into custom built software packages for the modelling and simulation of treatment plants. Examples of these packages include STOAT (Dudley, 1996), GPS-X (Hydromantis, 1996) and EFOR (EFOR, 1998). These packages allow the user to build models of plant layouts and simulate different conditions on desktop computers.

Some of the areas in which wastewater treatment models have been applied are design of plant; upgrading plants to treat higher loads; optimisation of existing plants; control of plants and training of plant managers, operators, etc.

2.2.2 Design

Traditional design of wastewater treatment plants is based on unit loading rates, e.g. kg BOD m^{-3} d^{-1}. Average flow and load data are used with peaking factors and safety factors added to account for diurnal variation and wet weather flows. This method for designing has often led to oversized plants, costing more to build than was necessary (Parker *et al.*, 1992). Parker *et al.* (1992) also reflected that since wastewater is dynamic in nature, it makes sense to use dynamic models when identifying processes to treat it. By using dynamic simulation in the design process it is possible to design plants to produce the required final effluent quality without being oversized. Examples include a UK water utility which made construction cost savings of approximately £70 000 at one plant, modelling having resulted in the oxidation ditch being sized 23% smaller than it would have been with manual calculations (Monro, 1996).

Models have not only shown promise in designing 'typical' plants but Daigger and Nolasco (1995) used them at different stages of the design process. Modelling led to the selection of a novel process, of which a pilot plant was designed using the model. Trial results were similar to those predicted by the initial model which was then recalibrated using the pilot plant data for use in designing the full scale plant.

Modified versions of the IAWQ activated sludge model no. 1 have been used in the design of sequencing batch reactors (Andreottola *et al.*, 1997; Oles and Wilderer, 1991). The models were then used to optimise the anoxic / aerobic cycle times and reactor volumes to achieve a set effluent quality of lab and pilot scale plants.

Oversizing of plants can be avoided because models can give greater confidence in processes due to the risks being quantified. This in turn allows the safety factors traditionally applied to be reduced to more realistic values.

If models are to be used to design a new wastewater treatment plant, a little caution should be applied when interpreting the results as wastewater characteristics can be unique to a specific plant. Hence the absence of analytical data for the wastewater can lead to uncertainty in model predictions.

2.2.3 Upgrading

The introduction of tighter effluent discharge limits for wastewater treatment plants as well as population growth has led to many plants not having sufficient treatment capacity to treat higher loads and therefore require upgrading. The first stage of any upgrading work is an assessment of the available capacity of the plant (Parker *et al.*, 1992). Dynamic simulation is a useful tool in assessing plant capacity under different load conditions, e.g. simulation of storm conditions. Smith and Dudley (1997) used the STOAT simulation software to assess the capacity of two activated sludge plants for treating predicted higher future flows and loads. In both cases it was found that the plants needed extending. The capacity of a treatment plant in Toronto, Canada has been assessed using activated sludge models in GPS-X, (Daigger and Nolasco, 1995).

Once it has been identified that the plant needs upgrading, there are usually a number of different options which could be implemented. Dynamic simulation has been used as a tool to compare the different options. The IAWQ (no. 1) model was used to compare three different configurations for nitrogen removal (Leseouf *et al.*, 1992). The simulation indicated that a 10% volume saving could be made by using a sludge reaeration, denitrification, nitrification (RDN) process instead of a conventional two tank anoxic / aerated process. A further 10% volume saving could be made by using an alternate phase - step feed (ALPHA) process. Alternatively, if the HRT was the same for all three processes, the RDN and ALPHA processes would have higher sludge ages by 2 days and 4 days respectively.

In order to reduce civil engineering work and hence capital costs the ideal upgrade is one which can be implemented within the existing plant layout. WEST simulation software has been used to evaluate different upgrading scenarios (Coen *et al.*, 1996). It was identified that increased nitrogen removal could be achieved within existing reactor volumes by introducing anoxic zones and step feeding. Kurata *et al.* (1996) modelled injection of primary sludge into the biological stage of a biological phosphorus plant and found that increased phosphorus removal efficiency could be achieved with no adverse effect on nitrification.

It is important to consider the works as a whole as upgrading one process at a wastewater treatment plant can have knock on effects on other processes. Göhle *et al.* (1996) used EFOR to evaluate the effect on sludge blanket levels in the final clarifiers of increasing the mixed liquor concentration in the aeration basin at a treatment plant. Critical solids loading rates for the final tanks at different values of $SSVI_{3.5}$ were identified.

2.2.4 Optimisation

Often the most cost effective approach to upgrading is the optimisation of an existing facility (Daigger *et al.*, 1992), e.g. activated sludge plants with a low F:M loading can often achieve better effluent standards without extension if they are optimised (Leeuw *et al.*, 1996). Optimising existing plant can reduce the need for costly capital schemes.

Traditional practice in wastewater treatment has been to optimise each individual unit process, a practice which will seldom result in the optimal design of the total system (Mishra *et al.*, 1975). In order to optimise the works as a whole it is necessary to consider not only the individual processes but also the interactions between processes. Dynamic models have been used to assess different methods of optimisation for individual processes and plants as a whole and have been used successfully on full scale and pilot scale plants. Examples include the use of calibrated computer models to optimise a pilot scale nitrification plant (Dupont and Sinkjaer, 1994). Once calibrated, the model was used to predict optimum values for several operational parameters. The predicted optimum values were implemented on the pilot plant, giving improvements of 30% for nitrate and 15% for total nitrogen in the final effluent, corresponding well with the model predictions. Leeuw *et al.* (1996) predicted 40% improvements in effluent total nitrogen could be achieved at a full scale nitrification / denitrification plant through the implementation of six process modifications which had been simulated with a calibrated activated sludge model.

A major operating cost of an activated sludge plant is the energy costs of aeration. Activated sludge models have been used to reduce the aeration requirements in activated sludge plants (Horan and Chen, 1998; Monro, 1996 and Healey, 1989), SBRs

(Demuynck *et al.*, 1994) and oxidation ditches (Parker *et al.*, 1992). In addition to reducing the aeration requirements, application of the Nitrification Denitrification Biological Excess Phosphorus Removal (NDBEPR) (Wentzel *et al.*, 1992) to SBRs by Demuynck *et al.* (1994) led to predictions of a 50% reduction of the external carbon source requirements for total nitrogen removal.

Urban drainage systems consist of three components, i.e. the sewer system, the wastewater treatment plant and the receiving water, all of which interact with one another. By use of models of the three components, the system as a whole can be considered. Gall *et al.* (1995) adapted existing suspended growth models, to consider the biological processes occurring within the sewer systems and their effect on the subsequent treatment plant. It was found that reactions taking place in the sewers could have a significant impact on the treatment plant.

In the UK the Urban Pollution Management (UPM) programme, co-ordinated by the Water Research Centre (WRc), has made use of linked models for sewer systems (MOSQITO), wastewater treatment (STOAT) and river quality (MIKE 11) to examine the interactions between the three components (Clifforde, 1992). The UPM method has been proved feasible with its application in the city of Derby catchment area (Crabtree *et al.*, 1996). The use of UPM in this case showed that a proposed upgrading option would meet river quality requirements at a lower cost than other options arrived at through traditional planning approaches. Linked sewer and treatment models have also been used to assess the possibility of using the surplus capacity in the sewer system as storage to smooth the flow to the treatment plant (Mark *et al.*, 1998).

The combination of sewer system, treatment process and receiving water models can be considered sequentially, the output of one model being stored in a data file to be used as the input to the next model. However, if control strategies or real time control are being considered the three models need to be run simultaneously as an integrated model (Gall *et al.*, 1995; Fronteau *et al.*, 1997).

A problem that can be encountered when using linked models for sewer system, sewage treatment and receiving water quality is that each models uses different state variables which are often incompatible. Standardisation of the variables used in each of the three models is recommended if the are to be used to model the drainage system as a whole, (Fronteau *et al.*, 1997).

2.2.5 Control

The ability of dynamic models to simulate changing plant performance with time makes them powerful tools for assessing different control strategies at wastewater treatment plants (Barnett *et al.*, 1995).

Models have been used to optimise control strategies at wastewater treatment plants, resulting in energy cost savings. Comparison of two aeration control strategies by Pederson (1992) resulted in predicted cost savings through the implementation of a DO controlled aeration strategy. Brouwer *et al.* (1994) estimated that energy savings between 11 and 21% could be realised at an industrial wastewater treatment plant by controlling oxygen uptake rate (OUR) after simulation (in STEM) to assess different control strategies. Similar energy savings were predicted by Thornberg and Thomsen (1994) who used EFOR to test control strategies on a recirculating activated sludge plant for nitrogen removal. The model identified that by introducing anoxic phases in the aerobic zones, the total nitrogen in the effluent could be reduced without affecting the effluent ammonia concentration. In addition to the energy savings it was predicted that further savings would be made on chemical addition.

As well as being used to assess different control strategies, models can be used directly in operational control of plants. Barnett *et al.* (1995) report a plant where dynamic models are being used for two purposes, firstly to design a closed loop control system using multiple measurements and control variables and secondly, to develop an open loop advisory system in which the model parameters are updated in real time to give an indication of the state of the process.

Models such as the IAWQ models for activated sludge are not always the most suitable for use in the day to day operation and control of a wastewater treatment works due to their complexity (Novotny *et al.*, 1992). Jeppsson and Olsson (1993) have also recognised that the complexity of some models makes them unsuitable for control and have developed reduced order models, requiring fewer input parameters and states, for use in control systems. Their reduced order model retains the main features of the IAWQ model and the parameters can be updated with data from on-line monitors.

2.2.6 Training

Calibrated models can be used as tools to enhance operators' understanding of individual wastewater treatment processes and how they behave and interact dynamically. Through improved understanding, '.... models provide us with the power to be more correct more often, better prepared and more often in control, to plan, predict and design for contingencies.' (Lindrea *et al.*, 1994).

Schreiber *et al.* (1992) described a training simulator for operators which made use of a graphical display of the plant. This enables the operators to relate to the real plant and see what effects changing various control parameters will have on the plant. Use of the simulator allowed instructors to examine the trainees' responses to different situations. Another system which makes use of a graphical user interface has been implemented using GPS-X in the city of San Diego (Barnett *et al.*, 1995). Using this system, trainees can alter various control variables such as wastage rate, and receive feedback on the effects of their actions on display screens.

When upset events occur, such as the appearance of toxic substances at a wastewater treatment plant, there are usually a number of options available to plant operators to deal with the problem. Metzger (1994) developed a simulator that allowed users to check the consequences of different operator actions to deal with appearance of cyanide at an industrial wastewater treatment plant. It was found that several operator actions could lead to a dangerous build up of cyanide. By making use of the simulator during training, operators would know the correct action to take in such circumstances.

2.2.7 Summary

Despite the research effort that has gone into the development of complex mechanistic models for wastewater treatment processes, they are only of use to industry if some benefit is obtained through their application - usually in terms of financial savings. The examples discussed have shown that there are considerable benefits to be obtained from the successful application of models, whether it be operational cost savings, capital cost savings or improved plant performance. However, there are costs associated with modelling which have to be considered.

Calibration of the complex dynamic models of wastewater treatment requires a great deal of initial data, including flow and water quality data, kinetic information and stoichiometric ratios. The IAWQ model no.1, for example, requires the wastewater to be characterised into 12 components and a total of 5 stoichiometric and 14 kinetic parameters, although 4 of the stoichiometric and 4 of the kinetic parameters can be considered as constants as they vary little from one wastewater to another (Gostelow, 1997). Not all of the required data is measured routinely at wastewater treatment plants and to obtain all of the data required to calibrate a dynamic model can be costly both in terms of money and time.

2.3 WASTEWATER CHARACTERISATION

An accurate characterisation of the influent is critical if any model predictions are to be used with any confidence. If carbon, nitrogen and phosphorus removal are to be considered then these components need to be defined in the influent.

2.3.1 Organic carbon

When defining the organic carbon in a wastewater there are several methods which can be used, including: biochemical oxygen demand (BOD); chemical oxygen demand (COD) and total organic carbon (TOC). Models have been developed based on BOD and COD and each method has advantages and disadvantages.

Biochemical oxygen demand (BOD) is routinely measured at most wastewater treatment works and therefore the systems are in place for sample collection and analysis and there is usually a reasonable amount of historical data available. However, BOD cannot be measured on-line and the duration of the test (usually 5 days in the UK) can hinder prompt action in the event of process disturbances. The main disadvantage of BOD for modelling purposes is that it is not possible to carry out a mass balance on BOD.

Chemical oxygen demand (COD) is the only one of these three methods for organic carbon measurement which allows a mass balance across the substrate, biomass and oxygen consumption to be carried out. The mass balance is based on electron equivalents with oxygen as the terminal electron acceptor (Henze *et al.*, 1987). It is for this reason that the IAWPRC (now IAWQ) Task Group on Mathematical Modelling for design and Operation of Biological Wastewater adopted COD as the measure of organic material in their Activated Sludge Model No. 1 (Henze *et al.*, 1987).

Whichever measure of organic carbon is used, both BOD and COD have been applied successfully in modelling work (Smith and Dudley, 1997), it can be subdivided further into fractions based upon their biodegradability. The fractions are often defined using their physical state, i.e. whether they are soluble or particulate to simplify the model structure although in reality the fractionation is more concerned with the biodegradability than the physical state.

2.3.2 Organic fractions used in models

For the Jones (1978) model the BOD fractions are soluble and particulate BOD with BOD_5 being the most commonly used BOD measurement (Smith and Dudley, 1997). Soluble BOD can be used directly by the bacteria but the particulate BOD has to be hydrolysed to soluble BOD before uptake.

For the Activated Sludge Model no.1 of the IAWQ the influent COD is divided into four fractions, readily biodegradable substrate (S_S), soluble inert COD (S_I), slowly biodegradable substrate (X_S) and particulate inert COD (X_I) (Henze *et al.*, 1987). Although the biodegradable fractions are assigned as being soluble and particulate, the fractionation is more concerned with the biodegradability than the physical state. The biomass in the influent was neglected in the Activated Sludge Model No.1 even though it has been noted that this fraction may account for up to 15 % of the influent total COD (Henze *et al.*, 1995). The biomass can be included in X_S without affecting modelling but a higher value must be used for the heterotrophic yield coefficient, Y_H , (Henze., 1992).

Other authors have found that better results could be obtained if the slowly biodegradable fraction was further divided into rapidly hydrolysable and slowly hydrolysable fractions (Sollfrank and Gujer, 1991) with hydrolysis being modelled as first order kinetics. Lesouef *et al.* (1992) found that a better model fit could be obtained if the 'soluble' biodegradable COD was divided into three fractions - readily biodegradable, slowly biodegradable and adsorbable COD.

The IAWQ Activated Sludge Model No.2 contains COD fractions which were redefined to accurately predict the phosphorus removal systems (Henze *et al.*, 1995). The COD fractions used in the Activated Sludge Model No. 2 are: readily (fermentable) biodegradable substrate (S_F); volatile acids / fermentation products (acetate) (S_A); inert, non-biodegradable organics (S_I); inert, non-biodegradable organics (X_I); slowly biodegradable substrate (X_S); heterotrophic biomass (X_H); phosphorus accumulating organisms (X_{PAO}); stored poly-hydroxy-alkanoate (X_{PHA}) and autotrophic, nitrifying biomass (X_{AUT}). The total influent COD is then defined by equation 1

$$C_{TCOD} = S_A + S_F + S_I + X_I + X_S + X_H + X_{PAO} + X_{PHA} + X_{AUT} \qquad \text{Equation 1}$$

In practice it has been found that X_{PAO} , X_{PHA} and X_{AUT} are very small and X_H can be included with X_S without significantly affecting modelling. In this case the total influent COD can be simplified to equation 2 (Henze *et al.*, 1995)

$$C_{TCOD} = S_A + S_F + S_I + X_I + X_S \qquad \text{Equation 2}$$

2.3.3 Methods for defining COD fractions for use in activated sludge models

Readily Biodegradable Substrate (S_S)

The readily biodegradable substrate (S_S) is the only fraction used directly by the biomass. The method proposed by the IAWQ Task Group for evaluating this fraction of the influent organics is that described by Ekama *et al.* (1986). A continuous flow through activated sludge plant is operated at a sludge age of 2 - 3 days and is fed cyclically, 12 hours on 12 hours off. The oxygen uptake rate (OUR) is measured in the aeration tank. When the feed is stopped there is an immediate drop in the OUR as the biomass is then using only the S_S produced by hydrolysis of the slowly biodegradable organics (X_S), a slower process than the biodegradation of the S_S in the influent. The readily biodegradable substrate can be calculated from the magnitude of the change in OUR (ΔOUR) using equation 3

$$S_{S0} = \left(\frac{1}{1-Y_H}\right) \cdot \Delta OUR \cdot \frac{V \cdot 24}{Q} \qquad \text{Equation 3}$$

where, Y_H is the heterotrophic yield coefficient (mg COD mg COD^{-1})

ΔOUR is the change in OUR (mg l^{-1} h^{-1})

V is the reactor volume (l)

Q is the feed flow rate immediately before it is stopped (l d^{-1})

24 is a conversion from days to hours

Sollfrank and Gujer (1991) used a lab scale activated sludge plant to measure some of the stoichiometric and kinetic parameters as well as the readily biodegradable fraction of the influent COD. The plant was fed with different concentration COD spikes and the OUR was measured. The sudden drop in OUR following the end of the addition of the COD spikes was used to measure S_{S0} using equation 3. The average readily biodegradable fraction of the wastewater tested was found to be 15.5 ± 1.8 % of the influent COD. This compares to 20 % for unsettled wastewater and 28 % for settled wastewater in South Africa (Ekama *et al.*, 1986).

Running a continuous plant to calculate S_{S0} can be an expensive and time consuming method for practical application to different plants and other methods have been proposed as alternatives. In addition to the continuous method described, Ekama *et al.* (1986) also used aerobic and anoxic batch tests to evaluate S_{S0}. For the aerobic method, measured volumes of wastewater and activated sludge mixed liquor were mixed in an aerated and stirred batch reactor. The OUR was measured every 5 - 10 minutes from the start of the test for up to 5 hours. Plotting OUR against time gave a graph of the form shown in figure 1 was obtained.

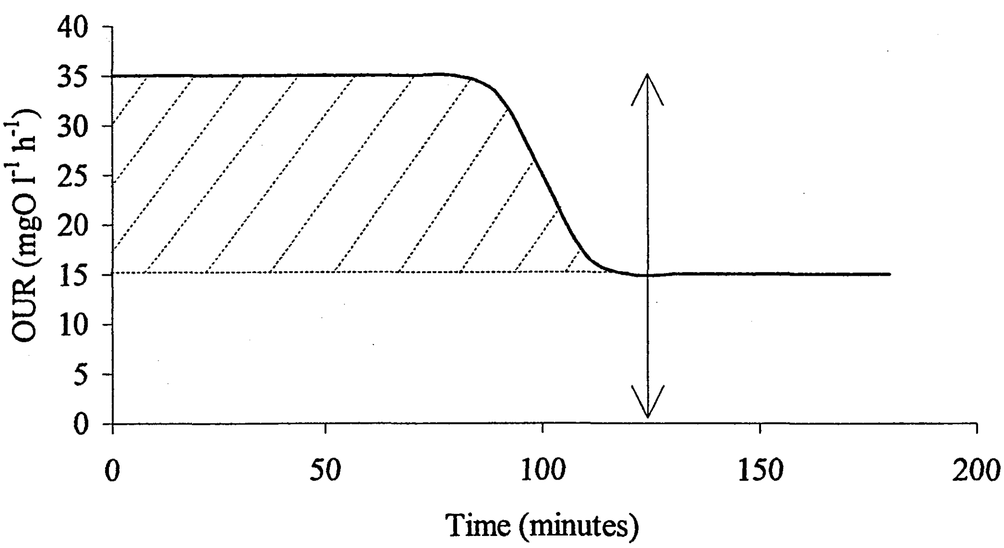

Figure 1. Variation of oxygen uptake rate with time in a batch test to determine the readily biodegradable fraction of the COD in a wastewater sample.

The shaded area in figure 1 is proportional to S_{S0} and the initial high OUR is proportional to the maximum specific growth rate of the heterotrophs ($\mu_{max\ H}$). The anoxic batch test involved adding nitrate to the wastewater and activated sludge mixture and measuring the nitrate concentration for the period of the test - up to 5 hours. Plotting the nitrate concentration against time gave a graph similar to the one in figure 2.

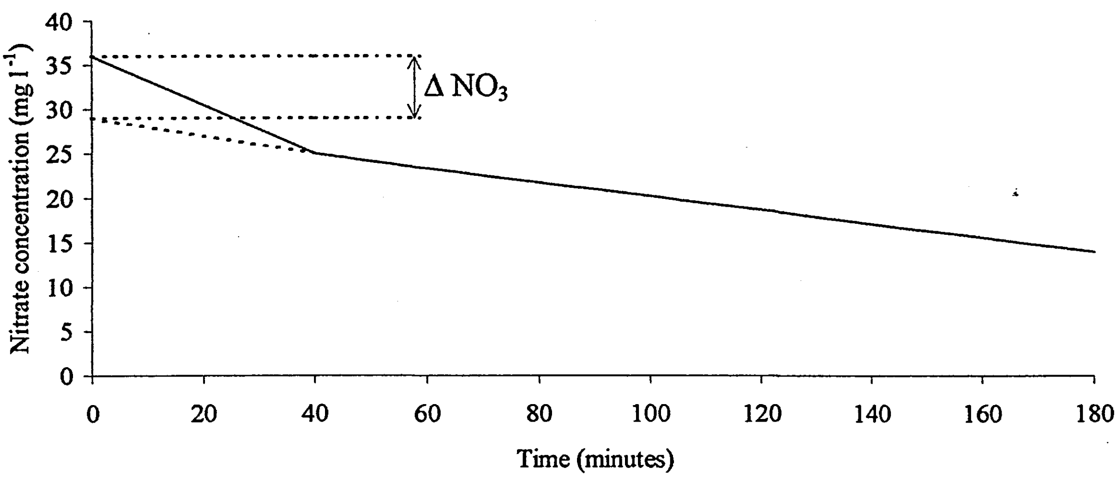

Figure 2. Variation of nitrate concentration with time in an anoxic batch test to measure the readily biodegradable substrate in a wastewater sample.

The readily biodegradable substrate can be calculated from $\Delta\ NO_3$ in figure 2 by applying equation 4.

Readily biodegradable substrate,

$$S_{S0} = \frac{2.86}{(1 - Y_H)} \cdot \Delta NO_3 \cdot \frac{(V_{ww} + V_{ml})}{V_{ww}} \qquad \text{Equation 4}$$

where, V_{ww} is the volume of wastewater (l)

V_{ml} is the volume of activated sludge mixed liquor (l)

2.86 is the electron acceptor capacity of NO_3 relative to oxygen as O

Y_H is the heterotrophic yield coefficient (mg COD mg COD^{-1})

$\Delta\ NO_3$ is the change in NO_3 concentration during degradation of the S_{S0}

A series of batch OUR tests was used by Kappeler and Gujer (1992) to evaluate several COD fractions, stoichiometric and kinetic parameters. Three tests were carried out as follows:

1. A test with centrifuged wastewater and a very small amount of activated sludge biomass to determine the maximum specific growth rate and the coefficient for endogenous respiration.

2. A test with just wastewater to determine the concentration of heterotrophic biomass in the influent.

3. A test with activated sludge biomass and wastewater in a ratio of approximately 1 to 2 to determine the readily biodegradable COD in the wastewater.

Fitting the model to the OUR curves obtained enabled other stoichiometric and kinetic parameters to be estimated. Using these methods S_{S0} was found to be 7 % and 8 % of the influent total COD at two different works in Switzerland (Kappeler and Gujer, 1992) and 24 % at pulp and paper plant in the UK (Chen and Horan, 1998).

Although good results have been obtained through the use of batch tests, it has been shown that different growth rates can be observed for tests on the same wastewater when comparing continuous and batch procedures (Novák *et al.*, 1994). This observation was explained by the difference in F:M ratios between the two tests. In a batch test the initial F:M ratio is higher than that in a continuous test, favouring faster growing micro-organisms at the start of the batch test. As a result of the differences between the two methods, it was concluded that care should be exercised when applying the results of batch tests to continuous systems.

A simplified batch test was proposed by Xu and Hasselblad (1996) which involved a single OUR measurement. Activated sludge and wastewater were mixed and aerated tap water was then added to increase the dissolved oxygen (DO) concentration. The

DO concentration was then logged over time, giving a graph of the form shown in figure 3.

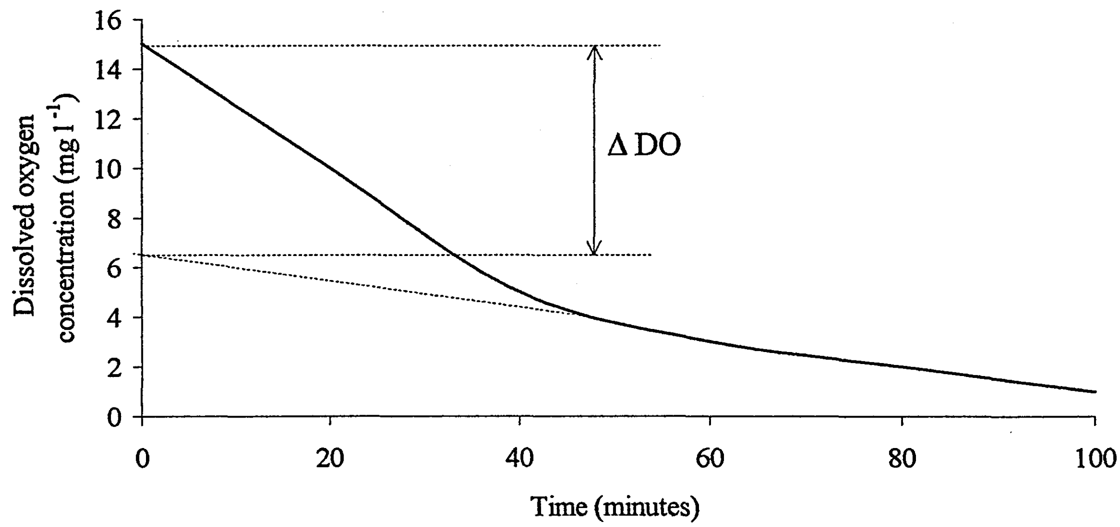

Figure 3. Variation of dissolved oxygen concentration with time in a single oxygen uptake test to measure the readily biodegradable fraction of wastewater.

The readily biodegradable COD is assumed to be equivalent to acetic acid and a calibration curve obtained of oxygen consumption versus acetic acid concentration. By comparing Δ DO from figure 3 to the calibration curve the equivalent concentration of acetic acid can be found and equated to S_{S0}. The results obtained using this method were found to be similar to those found in parallel tests using the batch procedures of Kappeler and Gujer (1992). The method has been used in the calibration of a model of a nutrient removal plant (Xu and Hultman, 1996).

When carrying out the batch tests it is important to select the correct F:M ratio so that the correct shape of curve is obtained that is suitable for extracting the required information. The tests should also be carried out using biomass which has adapted to the wastewater being tested (Ekama *et al.*, 1986 ; Henze *et al.*, 1987 ; Kappeler and Gujer, 1992). Several authors have noted that when carrying out batch tests nitrification should be inhibited with allyl thiourea (ATU) as the OUR due to nitrification can affect calculation of $\mu_{max\ H}$ (Ekama *et al.*, 1986; Kappeler and Gujer, 1992). It has been found, however, that as well as inhibiting nitrification, adding ATU can also inhibit the action of the heterotrophic biomass being used in the test leading to errors in measuring S_{S0} (Novák *et al.*, 1994). The same authors also noted that the OUR due to nitrification had a negligible effect on the exponential shape of the graph used to calculate $\mu_{max\,H}$.

Due to the difference in transport phenomena between fixed film and suspended growth systems it has been recommended that methods specific to biofilm systems are developed to evaluate the COD fractions available to biofilm biomass (Nogueira *et al.*, 1998). The same author put forward two methods, one a batch test and the other continuous, relating the nitrate uptake rate to the readily biodegradable substrate.

Although batch tests are less time consuming than the continuous methods, quicker methods have been proposed which are based on physical or physical-chemical methods. Mamais *et al.* (1993) stated that the readily biodegradable soluble COD equates to the difference between the truly soluble COD and the inert soluble COD in the influent. It was proposed that the truly soluble COD is the filtrate of a wastewater which has been flocculated to remove the colloidal particles. Although the flocculation step is relatively simple and quick to carry out, the method relies on the soluble inert fraction being known. Results obtained using the method were comparable to those found in parallel tests using the continuous flow method of Ekama *et al.* (1986).

Ultrafiltration is another method that has been proposed for measuring S_{S0} (Dold *et al.*, 1986). Torrijos *et al.* (1994) found that filtration at a pore size of 0.1 µm produced a fraction of the influent COD that contained only rapidly biodegradable and inert fractions. The absence of a soluble slowly biodegradable fraction differs from the findings of Sollfrank and Gujer (1991). The presence or absence of slowly biodegradable compounds within the soluble fraction is likely to be site specific and can be affected by factors such as the sewers through which the wastewater is transported (Henze, 1992). Ultrafiltration with membranes of 100 000, 50 000 and 3 000 Daltons of wastewater before and after biodegradation showed that below 0.1 µm molecular weight did not significantly affect biodegradability (Torrijos *et al.*, 1994).

Inert soluble organics (S_I)

In ASM1 the soluble inert organics are modelled as passing through the plant unchanged, being discharged in the effluent. However, it has been shown that in reality this fraction increases through the plant due to the production of inert soluble compounds in the activated sludge process (Orhon *et al.*, 1989; Germirli *et al.*, 1991). Instead of modelling the production of S_I in ASM1 a simplified method was used whereby the S_I term in the model input incorporates the actual initial soluble inert fraction and the produced soluble inerts.

Soluble inert COD can be estimated as the soluble COD after a 20 day BOD measurement (Henze *et al.*, 1987) or as the soluble COD in the effluent of an activated sludge plant operating under low load conditions (Ekama *et al.*, 1986). As a further improvement Henze (1992) noted that it would be better to use soluble COD minus soluble BOD in the effluent from such a plant.

In defining a flocculation method for measuring S_S Mamais *et al.* (1993) defined S_I as the truly soluble effluent COD from a low loaded activated sludge plant. It was suggested that for existing plants the COD of a flocculated and filtered sample of final effluent gave the required value of S_I.

Batch test methods have been described to evaluate both S_I and X_I in which a sample of the wastewater was aerated until no further change in COD occurred. The soluble inert COD was taken as the soluble COD remaining in the sample after the biodegradation. In a similar method Henze *et al.* (1992) defined the soluble inert fraction as the dissolved COD after 20 - 30 days oxidation of wastewater mixed with activated sludge.

For practical purposes, S_I has been assumed to be 90 % of the soluble COD in the secondary effluent of an activated sludge plant (Siegrist and Tschui, 1992; Xu and Hultman, 1996).

Slowly biodegradable substrate (X_S)

In ASM1 the slowly biodegradable fraction of the influent COD is not used directly by the biomass. It is modelled as first being hydrolysed to S_S which is then used by the biomass.

This organic fraction is usually calculated as the difference between the influent total COD and the other fractions (Henze *et al.*, 1987; Henze *et al.*, 1995). In the case of ASM1 this would simplify to

$$X_S = C_{TCOD} - S_I - S_S - X_I \qquad \text{Equation 5}$$

assuming that X_B is negligible.

Since it has been shown that X_B in the influent is not negligible and can represent up to 25% of the influent COD, the term X_S in equation 5 also incorporates any biomass present in the influent (Henze, 1992).

Oxygen uptake rate measurements in a batch test which was substrate limited have been used together with iterative curve fitting to determine X_S (Kappeler and Gujer, 1992). The same authors also suggested a batch test with just wastewater to measure X_S.

Inert particulate organics (X_I)

The inert particulate organics are not degraded in the activated sludge model but the fraction is important as it contributes to the sludge production. It is assumed that all of the particulate inerts are enmeshed in the activated sludge flocs and removed from the system with the waste sludge.

Calibration of the modelled sludge production against the observed sludge production is suggested as a method of obtaining the particulate inert fraction of the influent COD (Ekama *et al.*, 1986; Henze *et al.*, 1987). An alternative procedure was outlined by Lesouef *et al.* (1992) in which the wastewater sample and a filtered sample were aerated in parallel until no further change in COD occurred. After aeration the particulate COD in the unfiltered sample was the sum of the biomass created and X_I , assuming that all the degradable fractions have been degraded. The yield was obtained from the filtered sample, enabling calculation of the biomass created and hence X_I.

Having proposed methods to measure S_S, X_s and X_B, by curve fitting to a batch OUR test and S_I and COD_{TOTAL} by usual analysis, Kappeler and Gujer (1992) suggested obtaining X_I from equation 6.

$$X_I = C_{TCOD} - S_S - S_I - X_S - X_B \qquad \text{Equation 6}$$

2.3.4 Nitrogen fractions

The Jones BOD model (Jones, 1978) only has ammoniacal nitrogen as a user defined input. In the IAWQ models nitrogenous material is also divide into non-biodegradable and degradable fractions (Henze *et al.*, 1987). The non-biodegradable fraction is further divided into particulate (associated with particulate, non-biodegradable COD) and soluble (assumed to be negligible and not included in the model) sub-fractions. The biodegradable fraction consists of ammoniacal nitrogen (S_{NH}), soluble organic nitrogen (S_{ND}) and particulate organic nitrogen (X_{ND}).

The processes described by the model are hydrolysis of X_{ND} to S_{ND}, conversion of S_{ND} to S_{NH} by the action of heterotrophic bacteria and ammonia uptake as a nitrogen source for synthesis by heterotrophic and autotrophic bacteria and as an energy source for the growth of autotrophic bacteria.

2.3.5 Methods for defining nitrogen fractions for use in activated sludge models

Ammoniacal nitrogen can be measured directly using the appropriate laboratory analysis. The soluble inert organic nitrogen can be found by measuring the total Kjeldahl nitrogen (TKN) of the filtered sample used to determine S_I. By carrying out the TKN measurement on a filtered sample of the influent and subtracting the ammoniacal nitrogen and the soluble inert organic nitrogen the soluble biodegradable organic nitrogen can be found. The soluble biodegradable and the particulate biodegradable organic nitrogen are assumed to be in the same ratio as the readily and the slowly biodegradable COD. Hence X_{ND} can be found using equation 7 in which it is the only unknown.

$$\frac{S_{NDI}}{X_{NDI} + S_{NDI}} = \frac{S_{SI}}{X_{SI} + S_{SI}} \qquad \text{Equation 7}$$

There is no need to determine the particulate inert nitrogen (Henze *et al.*, 1987).

2.3.6 Practical application

Although ideally each component should be assessed individually this is not always practical as it can be very time consuming and therefore expensive. A practical solution to this is to measure commonly used wastewater characteristics such as BOD_5 , total suspended solids (TSS) and total Kjeldahl nitrogen. These characteristics are then approximated to the required model components using a number of influent ratios (Smith and Dudley, 1997). An example of this method is the BOD based influent model in GPS-X which takes BOD_5 and TSS as its input and then estimates the required COD fractions as shown in figure 4.

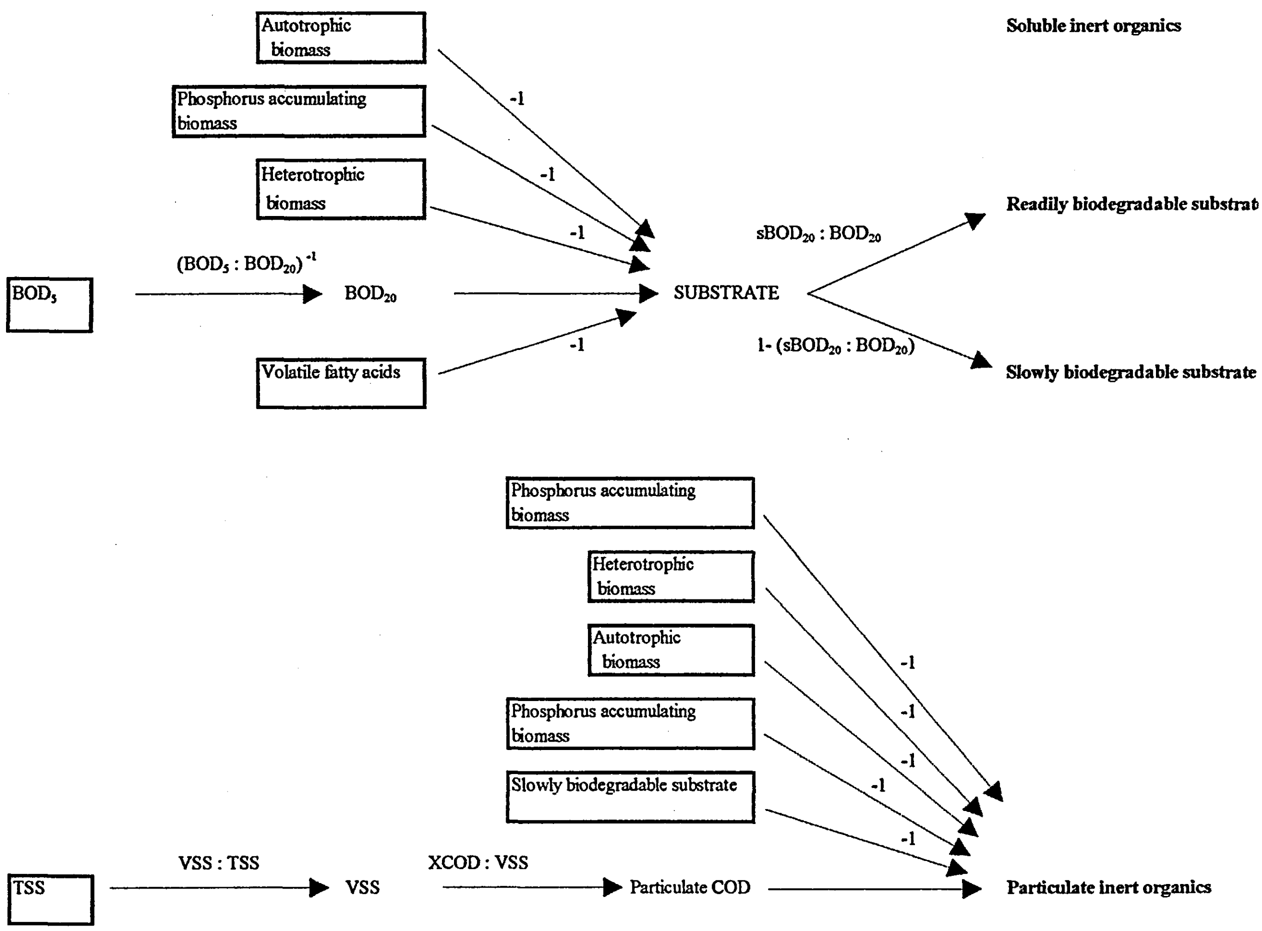

Figure 4. COD fractionation within GPS-X BOD based influent model. (Adapted from GPS-X Technical Reference).

Although this method allows for the COD states to be approximated it is not without its problems, e.g. BOD_{20} results are often unreliable. Smith and Dudley (1997) reported erroneous BOD_{20} results from 5 NAMAS accredited laboratories in the UK. Use of such results would lead to inaccurate predictions of the influent COD. If the influent COD is not described correctly the model predictions can not be used with any confidence.

The nitrogen fractions can be estimated by measuring one characteristic, e.g. TKN for every sample and then applying a number of pre-determined ratios (figure 5).

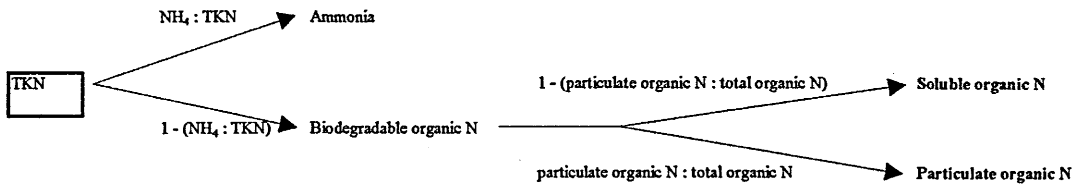

Figure 5. Nitrogen fractionation within GPS-X BOD based influent model. (Adapted from GPS-X Technical Reference).

The time and cost of data acquisition has been addressed by Bingley *et al.* (1996) as being a potential problem with modelling. As a result methods of making increased use of historical data have been developed which focus on using characteristics routinely measured at plants. Calibration of models using on-line data has also been tried with mixed results. Stokes and Coyle (1994) found that problems with drift in the instrumentation led to differences between modelled and actual results but models have been implemented with on-line updating of parameters (Barnett *et al.*, 1995).

There are approaches that have been used to try and reduce the sampling required to calibrate models and hence the cost of modelling. Reduced order models have been developed which use fewer input parameters and states (Jeppsson and Olsson, 1993; Gostelow and Parsons, 1998). An alternative method of simplifying models is to 'lump together' similar process units (Watson *et al.*, 1994). This approach does not reduce the number of parameters and states that have to be measured but does reduce the number of units that have to be calibrated and the total amount of data that has to be collected.

2.3.7 Cost savings possible through use of process modelling

The financial benefits should outweigh the costs of model calibration and application. The cost of the analysis required to develop a calibrated model of a wastewater treatment plant increases with the size and complexity of the works. Even using regularly measured parameters and ratios to approximate the wastewater characteristics, the cost of analysis can exceed £10 000 even for a conventional ASP. Add to this the cost of the modeller's time and the cost of a calibrated model can exceed £20 000. To make the modelling exercise pay for itself, the savings in opex or capex must be greater than these costs.

Many of the examples of where models have been applied reported reduced plant volumes being required as a result of the modelling. In these cases, the cost of the modelling is easily recouped as shown by the £70 000 saving at one plant due to the oxidation ditch being sized 23% smaller than it would have been with manual calculations (Monro, 1996).

A further disadvantage of relying on models is that there can be some temptation to use modelling for the sake of it when other methods might be more suitable. If a model is to be used it should only be as complicated as is necessary. The information obtained from simulations is only as good as the model and the input data used, with poor models or initial data giving misleading results (Andrews, 1993). Once a model has been calibrated and simulations run, the results of the simulations still need interpreting and it should be remembered that although dynamic models are a useful tool, they should only be used in conjunction with sound engineering judgement (Bingley *et al.*, 1996).

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CHAPTER 3 - FULL SCALE TRIALS OF THE MBBR / AS HYBRID PROCESS

3.1 INTRODUCTION

Colchester STW is a 120 000 PE works which needs to be upgraded in order to achieve an effluent quality within a new consent which comes into force from the end of 1999. The works currently has no effluent ammonia limit but under the new consent it will have to meet a 15 mg l^{-1} limit on a 95 percentile basis. Incorporation of fixed film carriers in activated sludge plants and the use of 2 stage fixed film - activated sludge processes have been proposed for upgrading to ammonia removal (Bonhomme *et al.*, 1990; Parker *et al.*, 1994). One of the options considered for the upgrade was a Kaldnes hybrid plant consisting of a moving bed biofilm reactor (MBBR) followed by an activated sludge stage. The aim of this process is to degrade most of the carbonaceous load in the biofilm reactor leaving the activated sludge zone for nitrification. Pilot scale work carried out at Anglian Water's Wastewater Innovation Centre (Jaouen, 1996; Lievre, 1997) indicated that nitrification could be achieved using the hybrid process at sludge ages shorter (as low as 4 days based on the suspended biomass only) than those normally used for nitrification in a single stage activated sludge plant (typically 8 days or more). The results also implied that retrofitting a biofilm reactor into the front end of an existing activated sludge plant would enable nitrification to be achieved within the existing tank volumes. This solution would lead to significantly lower capital expense than other options such as extending the existing ASP or adding a tertiary nitrification process, e.g. a biological aerated filter (BAF). Before proceeding with the upgrade for the entire plant, an empty existing aeration tank at Colchester STW was fitted out as a Kaldnes hybrid plant in order to test the process at full scale.

3.2 PLANT DESCRIPTION

3.2.1 Existing secondary treatment at Colchester STW

Prior to the hybrid lane being commissioned the secondary treatment was provided by three mechanically aerated activated sludge lanes and one fine bubble diffused aeration activated sludge lane. The volume of each of the surface aerated lanes was 2436 m^3 and the volume of the diffused aeration lane was 2655 m^3. Settled sewage from four primary sedimentation tanks mixed in a chamber before entering the four activated sludge lanes through a common channel. The diffused aeration lane treated approximately 40 % of the influent. The existing plant was not operated to nitrify.

3.2.2 Current and future consents

The current and future final effluent consent limits for Colchester are summarised in table 1. The future River Needs Consent (RNC), set by the Environment Agency comes into force in the year 2000 and in addition the works must also comply to the Urban Waste Water Directive (UWWTD) consent by the year 2001.

Table 1. Current and future final effluent consent limits for Colchester STW. (All consents as 95 percentiles.)

	Current consent	Future consent	
		River Needs Consent (Spot samples)	UWWTD consent (24 hour composite)
BOD	25 mg l^{-1}	35 mg l^{-1}	25 mg l^{-1}
Suspended solids	50 mg l^{-1}	70 mg l^{-1}	...
Ammonia	...	15 mg l^{-1}	...
COD	...	...	125 mg l^{-1}

3.2.3 Kaldnes Hybrid Plant

The hybrid plant at Colchester consisted of three zones, an anoxic zone, a MBBR zone and an activated sludge zone. The three zones were all built into one existing tank with dividing walls constructed between each of the zones. The sewage flowed through gaps in the dividing walls, baffles on each wall preventing short circuiting.

The **anoxic zone** had a volume of 171 m^3. There was a Flygt mixer mounted on one wall to prevent the solids from settling in the anoxic zone.

The **Kaldnes MBBR reactor** had a volume of 487 m^3 with a 50 % fill of Kaldnes media (Kaldnes Miljoteknologi AS, Norway). The Kaldnes media elements were cylinders 9 mm wide and 7 mm long with two cross pieces on the inside and longitudinal fins on the outside. They were made from high density polyethylene and had a specific density between 0.92 and 0.95 g cm^{-3}. The 50 % fill by volume resulted in a specific surface area of 250 m^2 m^{-3}. The media was retained in the Kaldnes zone by 5 mm wedgewire flat sieves between the anoxic zone and the Kaldnes zone and by 5 mm pipe sieves between the Kaldnes zone and the activated sludge zone. Coarse bubble aeration provided the process air requirements in the Kaldnes zone and also kept the media mixed. A single Holmes Rootes Dresser blower (55 kW) was used, the speed of which was manually adjustable.

The **activated sludge zone** had a volume of 1970 m^3 with aeration provided by fine bubble membrane diffusers on lift out grids. The dissolved oxygen in the activated sludge zone was controlled to a set point by the programmable logic controller (plc) altering the blower speed. The two blowers supplying air to the activated sludge zone were variable speed Holmes Rootes Dresser blowers (55 kW) operating as duty and assist.

A similar blower to that used for the Kaldnes zone was used as standby to both the Kaldnes zone and the activated sludge zone.

It was possible to feed the return activated sludge (RAS) either to the anoxic zone or direct to the front of the activated sludge zone although throughout the trial the plant was run with all the RAS returned to the activated sludge zone, making the anoxic zone redundant. Settled sewage was fed to the anoxic zone and there was a facility for bypassing some settled sewage direct to the activated sludge zone which was not used during the trial.

3.2.4 Design load for Kaldnes hybrid plant

The Kaldnes plant was designed to treat an average flow of 14 400 m^3 d^{-1} with an average influent settled sewage BOD_5 and ammonia concentrations of 173 mg l^{-1} and 32 mg l^{-1} respectively. The resultant design load for BOD_5 was 5.1 kg m^{-1} d^{-1} (or 20 gBOD_5 m^{-2} d^{-1}) on the Kaldnes zone, similar loading rates to those reported at other sites (Rusten *et al.*, 1996; Sunner *et al.*, 1998).

The trial took place in two phases.

3.3 FIRST PHASE (28TH APRIL 1997 TO 19TH MARCH 1998)

3.3.1 Plant layout and operation

During the first phase the hybrid plant was operated in parallel with the existing secondary treatment processes, as shown in figure 1. All five lanes received settled sewage from the same channel and the mixed liquor from all five lanes was mixed before the final sedimentation tanks. Four final tanks were used, 3 with a surface area of 578 m^2 and the other with a surface area of 616 m^2. The RAS to the four existing lanes was returned via one screw lift pump whilst a second screw lift pump took RAS from a common RAS channel to the hybrid lane. Surplus activated sludge was wasted from the RAS channel, giving no control of sludge age for the individual lanes.

Two months into the trial one of the surface aerated lanes was taken out of service as the plant had surplus capacity with up to 80 % of the flow being treated in the hybrid and diffused aeration lanes. The tank was isolated and drained down.

During the first phase of the trial the plant was operated with the intention of achieving full nitrification in all of the activated sludge lanes.

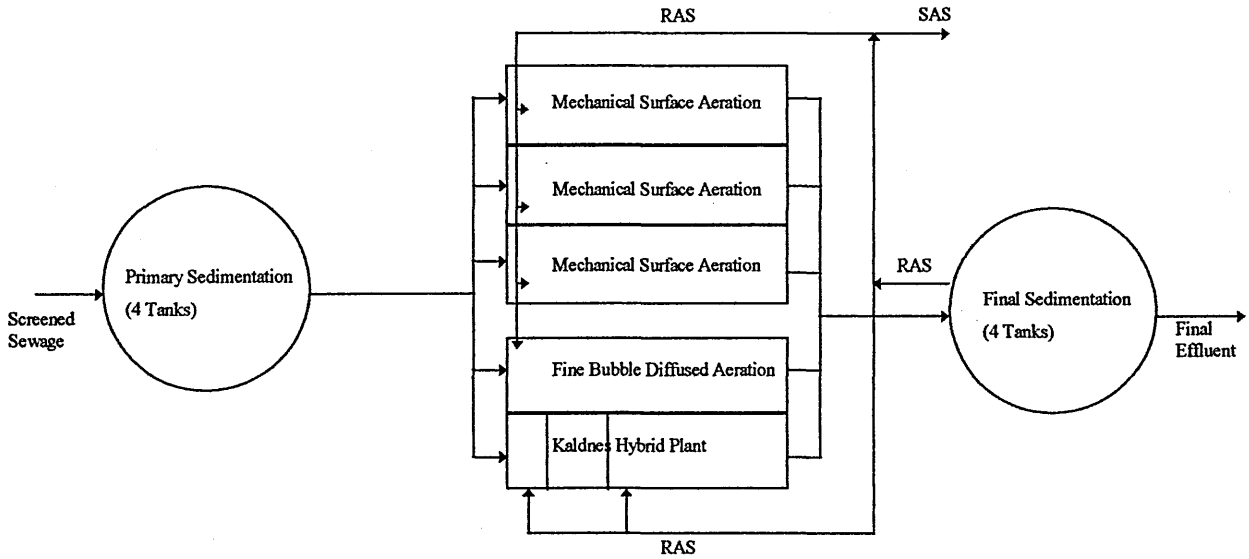

Figure 1. Secondary treatment configuration at Colchester STW during the first phase of the Kaldnes hybrid trial.

3.3.2 Sampling and analysis

During the first phase of the trial spot samples of settled sewage, MBBR effluent, mixed liquor from each lane and works final effluent were collected. All samples were tested on site for suspended solids and ammonia and the mixed liquor samples were tested for stirred specific volume index at a concentration of 3.5 g/l ($SSVI_{3.5}$). Some samples were analysed for nitrate, chemical oxygen demand (COD), filtered COD and total nitrogen. Suspended solids analysis was carried out by drying the samples in a microwave oven. The SSVI tests were carried out using WRc SSVI apparatus. On site analysis of ammonia, nitrate, total nitrogen and COD concentrations were carried out using Dr Lange cuvette tests and a Dr Lange LASA 20 photometer.

3.3.3 Results and discussion

Ammonia removal

During the first phase, the aim of the trial was to give sufficient capacity at the works to have full nitrification in all of the aeration lanes. In addition to the extra capacity it was thought that if the hybrid lane fully nitrified then it would seed the other aeration lanes with nitrifiers, enabling the whole plant to nitrify. A rough estimate only of the overall sludge age (5 days) could be made. The main parameters considered during this phase were the ammonia, the mixed liquor suspended solids (to ensure the works settling capacity was not exceeded) and the $SSVI_{3.5}$ to evaluate the effects of the biofilm process on the settleability of the activated sludge. The ammonia concentration in the hybrid plant effluent and the ammonia removal in the hybrid plant are shown in figures 2 and 3. It can be seen that during the first month of the trial the ammonia removal was very low. During this period the biofilm was only just starting to develop on the Kaldnes media but it had not developed significantly until a month into the trial. Other authors have reported periods of several weeks to a few months for biofilm concentration to stabilise (Hegemann, 1984; Bonhomme *et al.*, 1990) As the biofilm started to grow, the ammonia removal in the hybrid increased but full nitrification was still not achieved until 3 months into the trial.

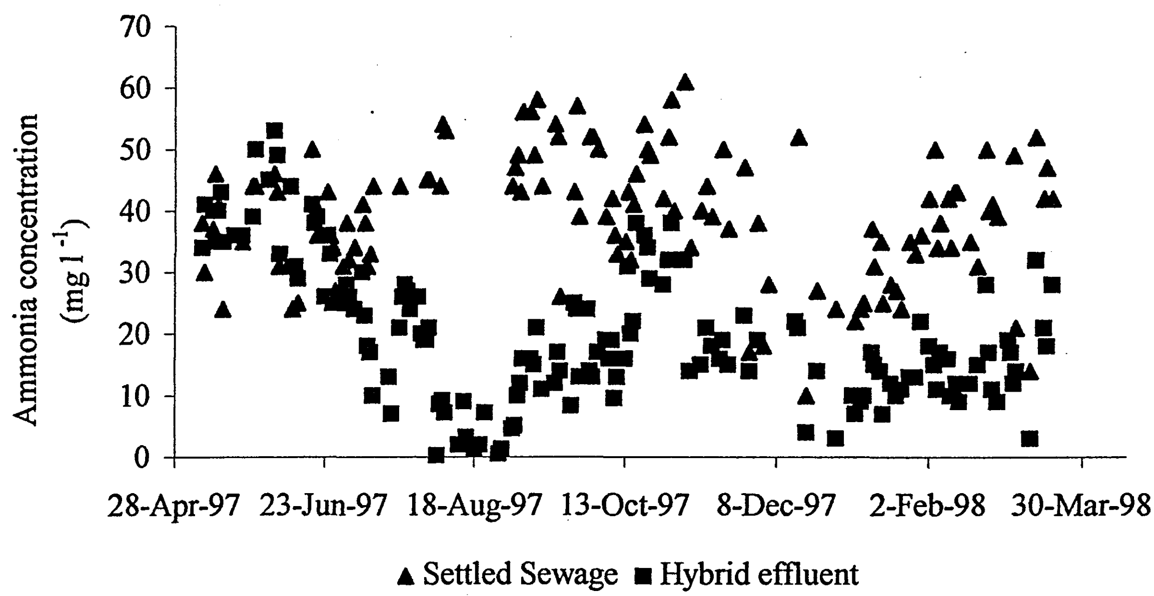

Figure 2. Variation of ammonia concentration (mg l^{-1}) in the settled sewage and hybrid effluent during first phase of the trial.

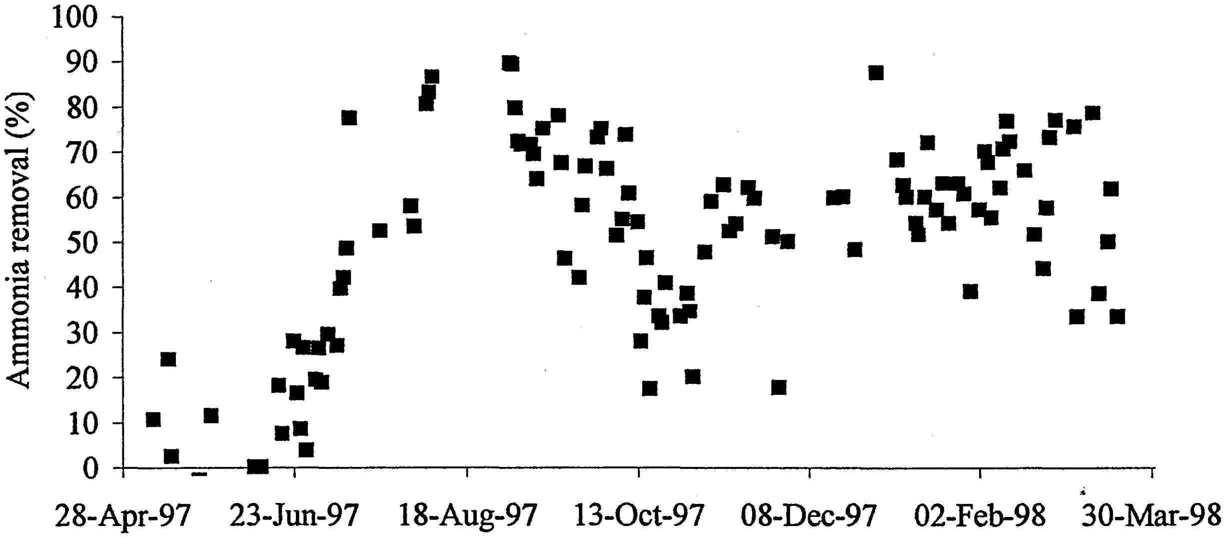

Figure 3. Ammonia removal in the hybrid plant during the first phase of the trial.

Three factors were judged to have prevented the plant from fully nitrifying. Firstly, the available final settlement capacity at the works limited the overall mixed liquors that could be maintained in the aeration tanks to approximately 2800 mg l^{-1} for the $SSVI_{3.5}$ observed. Keeping the mixed liquor below the allowable maximum to avoid solids washout resulted in an average overall sludge age of approximately 4 days between 1st August 1997 and 19th March 1998. This is below the usual sludge age used in nitrifying activated sludge plants which is usually in the range of 8-20 days, (Metcalf and Eddy, 1991). Also, based on the design BOD load and suspended biomass only, the design F:M ratio was 0.43 kg BOD kg^{-1} MLSS d^{-1}. Although Lessel (1994) reported nitrification at an F:M of 0.4 kg BOD kg^{-1} MLSS d^{-1} (based on the suspended biomass) it would not be expected to achieve nitrification at this load. A F:M ratio of < 0.15 kg BOD kg^{-1} MLSS d^{-1} is generally considered to be required for nitrification in single sludge AS systems (Bonhomme *et al.*, 1990). For separate stage nitrification systems F:M ratios in the range 0.04 - 0.16 kg BOD kg^{-1} MLSS d^{-1} are required [adapted from Metcalf and Eddy (1991) assuming a ratio of MLVSS:MLSS of 0.8 kg kg^{-1}].

The second factor was that although the hybrid plant started to nitrify, any nitrifiers produced were diluted when the mixed liquors were combined. This would have seeded the other lanes with nitrifiers but although the diffused aeration lane partially nitrified, the aeration capacity in the mechanically aerated lane was not sufficient to maintain complete nitrification, the third factor contributing to the plant not fully nitrifying. The dissolved oxygen concentration in each zone of the two surface aerated lanes was measured and is summarised in table 2, a zone being the area in which an aerator was positioned.

Table 2. Dissolved oxygen concentrations in surface aerated activated sludge lanes at Colchester STW.

	Zone of aeration tank					
	1	2	3	4	5	6
Aeration tank 1	✓ (0.6)	✓ (1.5)	✓ (1.7)	✓ (2.0)	✓ (1.4)	✗ (1.2)
Aeration tank 2	✓ (0.5)	✓ (0.6)	✗ (0.6)	✓ (1.4)	✓ (1.0)	✓ (1.1)

✓ Aerator working ✗ Aerator not working ( ) Dissolved oxygen concentration (mg l^{-1})

During September and October the ammonia removal in the hybrid plant decreased before picking up again in November. The loss of nitrification was caused by a problem with the primary sedimentation tanks which resulted in an increased load to the biological treatment stage. and hence sludge production in the aeration tanks. To avoid solids washout from the final sedimentation tanks more sludge had to be wasted, reducing the overall sludge age and hence the nitrification. Once the problems had been solved the nitrification increased again, during November, and partial nitrification was maintained until the end of the first phase of the trial.

Settling

The $SSVI_{3.5}$ at the works during the first stage of the trial is summarised in table 3. The average $SSVI_{3.5}$ above 100 ml g^{-1} indicates that the activated sludge was not settling well during the first trial period although it would probably be acceptable. A $SSVI_{3.5}$ of 60 - 75 ml g^{-1} is typical of a sludge with good settleability, (Ekama *et al.*, 1997), whilst a $SSVI_{3.5}$ of greater than 120 ml g^{-1} would be attributed to a poor settling sludge (The Institute of Water Pollution Control, 1987).

The poor settling was probably caused by the plant running in a state of partial nitrification, leading to denitrification, release of nitrogen gas and hence poor settleability and rising sludge in the final settling tanks. Ekama *et al.*, (1997) reported cases of denitrification occurring in the secondary settling tanks of plants which were designed to nitrify but not denitrify and in plants which were not designed for nitrification but which achieved it unintentionally due to operational conditions. Poor sludge condition due to the inadequate aeration in the surface aerated lanes and due to the plant running in a state of partial nitrification could also have contributed to the poor settling results. There was little difference observed between the settling of the sludge from the existing plant and that of the hybrid process.

Table 3. Average, maximum and minimum $SSVI_{3.5}$ during first phase of Kaldnes hybrid trials at Colchester STW.

	$SSVI_{3.5}$ (ml g^{-1})	
	Hybrid lane	Combined mixed liquor
Average	109	113
Maximum	208	172
Minimum	69	82

3.3.4 Summary

The aims of the trial were not achieved during the first stage as neither the hybrid lane or the works as a whole fully nitrified. It was thought that there was still potential in the hybrid process since despite all of the operational difficulties encountered the plant still partially nitrified. It was therefore decided to separate the hybrid plant from the rest of the secondary treatment as previously described. On 19th March 1998, the hybrid lane was separated and the second phase of the trial started.

3.4 SECOND PHASE (19TH MARCH 1998 TO 1ST MAY 1998)

3.4.1 Plant layout and operation

For the second phase of the trial the hybrid lane was isolated from the rest of the secondary treatment. By making use of two reconditioned humus tanks, surface area 422 m^2, for final settling and an over-pumping operation for returning sludges, the mixed liquor from the hybrid lane was kept separate from the rest of the mixed liquors (figure 4). Mixed liquor from the end of the activated sludge zone was pumped to the two humus tanks with two low speed centrifugal pumps. Once settled the RAS was pumped back to the screw lift pump with two submersible centrifugal pumps. For the hybrid plant surplus activated sludge was wasted directly from the aeration tank using a submersible pump so that the sludge age could be controlled.

During the second phase of the trial the hybrid lane was operated to fully nitrify whilst the rest of the secondary treatment was operated for carbonaceous removal only.

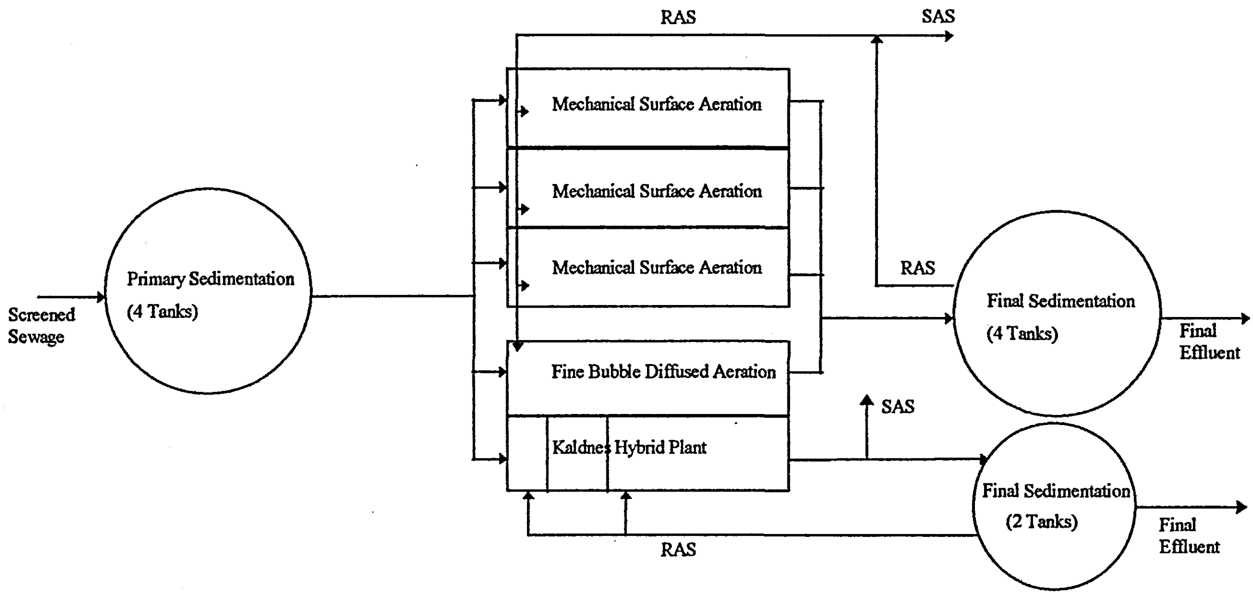

Figure 4. Secondary treatment configuration at Colchester STW during the second phase of the Kaldnes hybrid trial.

3.4.2 Sampling and analysis

During the second phase of the trial daily composite samples were collected, using Epic autosamplers, from the inlet to the anoxic zone, the outlet sieves from the Kaldnes zone, the effluent weir of the activated sludge zone and the combined effluent channel from the two humus tanks. These samples were analysed on site for suspended solids, total nitrogen, filtered total nitrogen, ammonia, nitrate, nitrite, total COD and filtered COD. A part of these samples was also sent to the Anglian Water laboratory at Whitlingham to be analysed for total 5 day biochemical oxygen demand (BOD_5) and filtered BOD_5 for all samples and for volatile suspended solids, BOD_{20} and filtered BOD_{20} for the settled sewage, Kaldnes effluent and mixed liquor samples. All analysis at Whitlingham was carried out using standard methods.

In addition to the daily composite samples, spot samples were taken to monitor overall performance and to check that storage in the autosampler did not affect the results significantly. Spot samples were taken of the settled sewage, Kaldnes effluent, mixed liquor from all four aeration tanks, combined mixed liquor from the distribution chamber to the final sedimentation tanks, final effluent from the hybrid plant and works final effluent. All of the spot samples were analysed for suspended solids, ammonia and nitrate. The hybrid activated sludge and the combined activated sludge from the other aeration lanes were also tested for $SSVI_{3.5}$.

On site analysis of ammonia, nitrate, nitrite, total nitrogen and COD concentrations were carried out using Dr Lange cuvette tests and a Dr Lange LASA 20 photometer.

3.4.3 Results and Discussion

Process loading

From 19th March to 22nd April the sludge age in the hybrid lane was nominally 11 days and from 22nd April to 1st May the sludge age was 5 days. However, when the mixed liquor transfer pumps were beaten by the incoming flow, the excess flow weired over into the existing plant, reducing the sludge age in an uncontrolled manner. The average flows and loads to the hybrid plant are summarised in table 4.

Table 4. Average flows and loads on the Kaldnes hybrid plant from 19th March to 1st May 1998.

	Influent flow rate (m^3 d^{-1})	Total BOD load (kg d^{-1})	Total BOD load on MBBR (kg m^{-3} d^{-1})	Ammonia load (kg d^{-1})
Average	6500	395	0.8	194
Maximum	8700	842	1.8	294
Minimum	3100	119	0.2	92

It can be seen from table 4 that the loads on the hybrid during the second phase of the trial are very low, an average of 0.8 kg m^{-3} d^{-1} compared to a design load of 5 kg m^{-3} d^{-1}. This was due to the extraordinary amount of rainfall experienced in the UK during April 1998 which resulted in low BOD_5 and ammonia concentrations in the influent. The flow to the hybrid during this period was limited by the capacity of the pumps taking the mixed liquor to the final sedimentation tanks.

Carbonaceous removal

The BOD_5 and COD removals were considered for each stage of the hybrid plant based on the daily composite samples. The average removals obtained after the hybrid plant had been separated are summarised in table 5.

Table 5. COD and BOD concentrations and removals in Kaldnes hybrid plant during second phase of trial.

	Inlet	MBBR effluent	Removal in MBBR (%)	Final effluent	Removal in AS zone (%)	Overall removal (%)
Total COD (mg l^{-1})	458	696	-52	67	90	85
Filtered COD (mg l^{-1})	126	69	45	43	38	66
Total BOD (mg l^{-1})	73	286	-292	10	97	86
Filtered BOD (mg l^{-1})	33	12	64	3	75	91

The average total BOD_5 and COD concentrations increased across the Kaldnes zone, this was probably caused by the high suspended solids concentrations in the Kaldnes reactor observed during periods of low influent flow to the works, (figure 5). These peaks were caused by the physical layout of the plant which resulted in mixed liquor from the surface aerated lanes to flow back into the 2 diffused aeration lanes. It is also possible that some mixed liquor from the activated sludge zone of the hybrid plant mixed back into the MBBR zone, increasing the solids concentration in samples. The variation of BOD_5 with suspended solids is shown in figure 6 which shows how BOD_5 increases with suspended solids in the Kaldnes effluent.

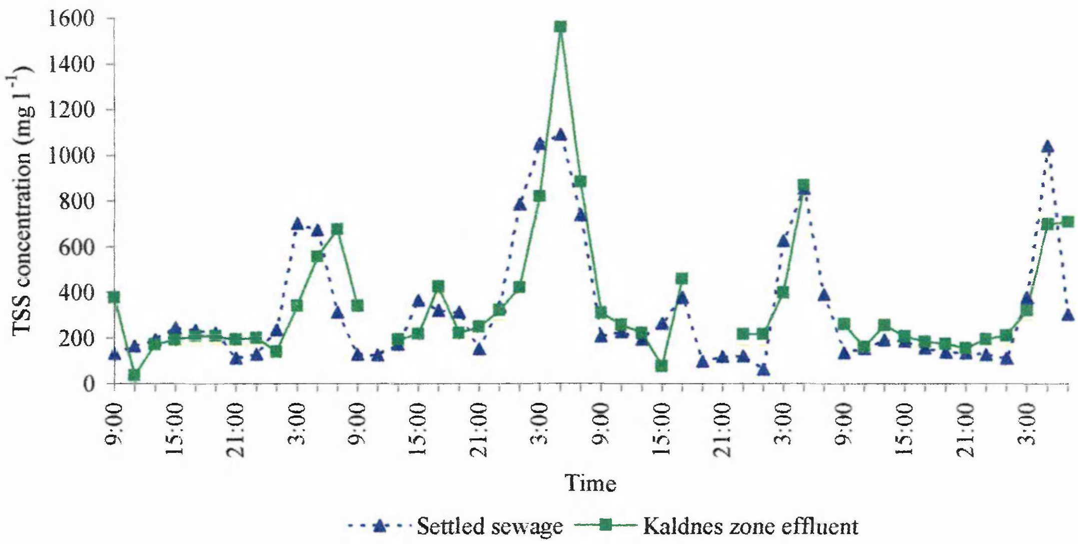

Figure 5. Diurnal variation of influent and Kaldnes zone effluent suspended solids concentration (mg l^{-1}) showing peaks during periods of low flow.

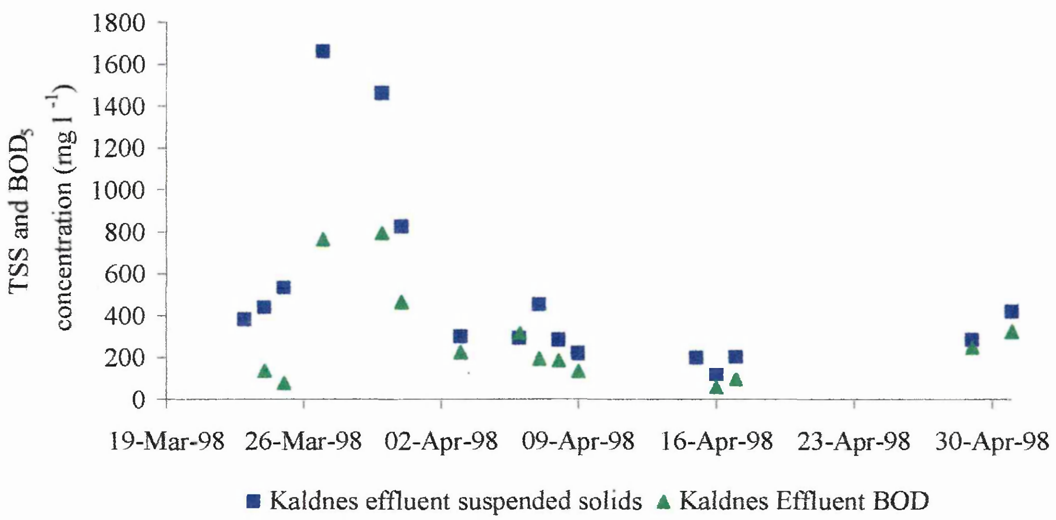

Figure 6. Variation of total BOD_5 with suspended solids for Kaldnes effluent.

It can be seen that from figure 6 that the BOD_5 concentration appears to follow the suspended solids concentration and a straight line relationship ($R^2 = 0.85$) is indicated by figure 7. The correlation coefficient for the two data sets is 0.94, indicating a strong relationship between the total BOD_5 and TSS.

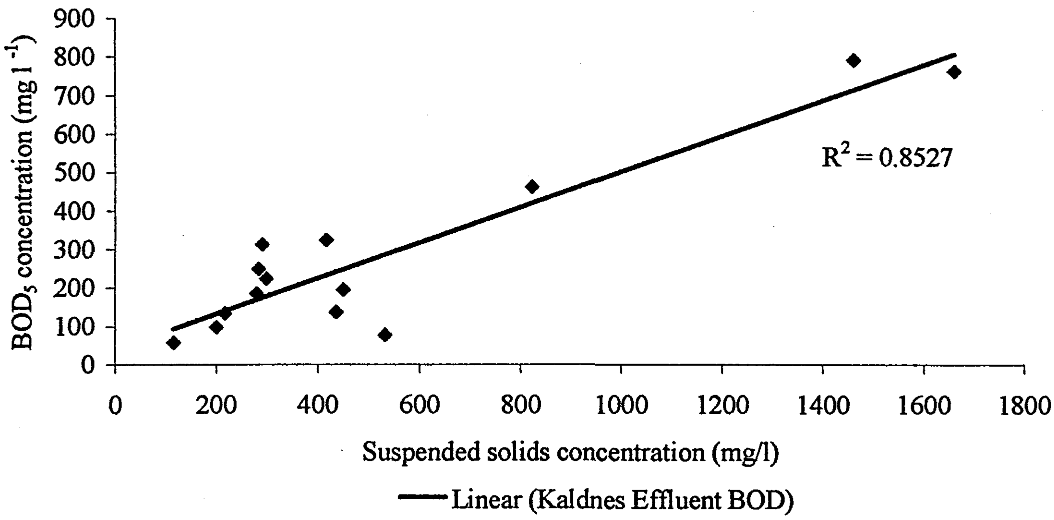

Figure 7. Relationship between BOD$_5$ and suspended solids for the effluent from the Kaldnes zone.

The soluble BOD$_5$ and soluble COD concentrations were reduced by an average of 64 % and 45 %, respectively in the Kaldnes zone as the biofilm degraded the readily available carbon. These results are similar to those achieved in a pilot scale hybrid plant during trials carried out at the Anglian Water Wastewater Innovation Centre in Cambridge. Results from the pilot plant showed average soluble BOD$_5$ removals between 21 and 63 % in the Kaldnes zone although the organic load was higher, 5 kg BOD$_5$ m^{-3} d^{-1}, (Michel, 1998). It was observed that soluble COD removal increased with surface loading rate (figure 8). The correlation coefficient for the two data sets is 0.81, indicating a strong relationship between the soluble COD removal and the surface loading rate. This relationship has been reported for other MBBR applications (Pastorelli *et al.*, 1997) and for submerged aerated filters (Rusten, 1984). The removal rates are similar to those observed by Pastorelli *et al.* (1997).

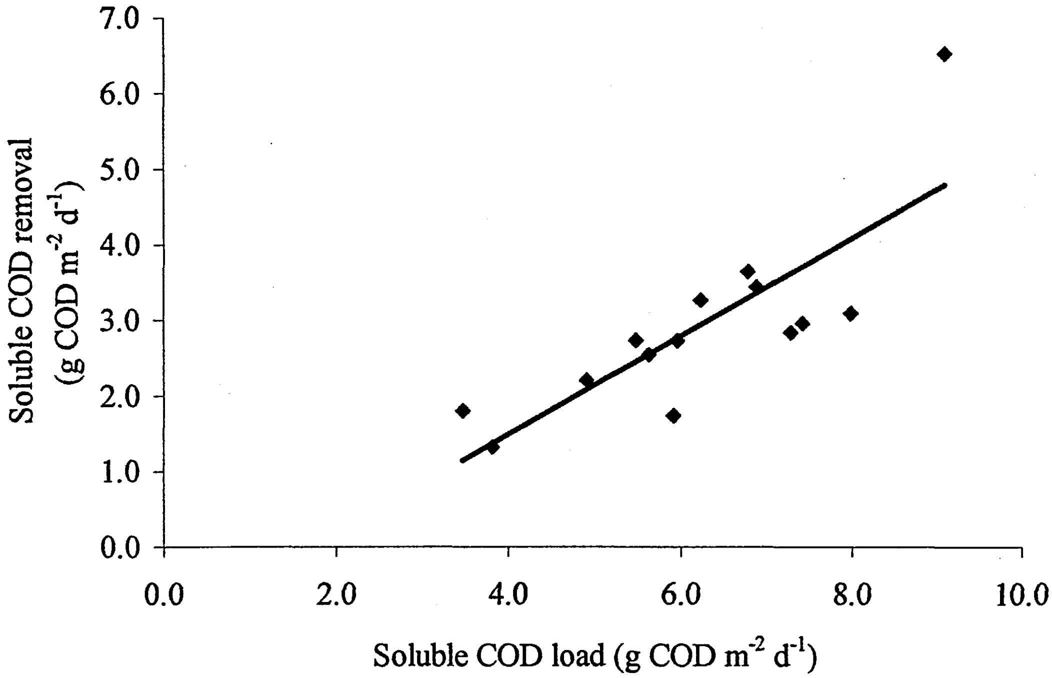

Figure 8. Soluble COD removal v soluble COD load for the MBBR zone of an MBBR/ASP hybrid plant.

The average removal of total BOD_5 through the whole plant was 86 % which is slightly lower than the best removal achieved in the pilot scale trials at Cambridge which was 94 %, (Michel, 1998). The hybrid plant final effluent total BOD_5 and total COD concentrations are shown in figure 9, from which it can be seen that except for one day, 25th March, both the BOD_5 and COD were within the future consent levels of 25 mg l^{-1} and 125 mg l^{-1} respectively.

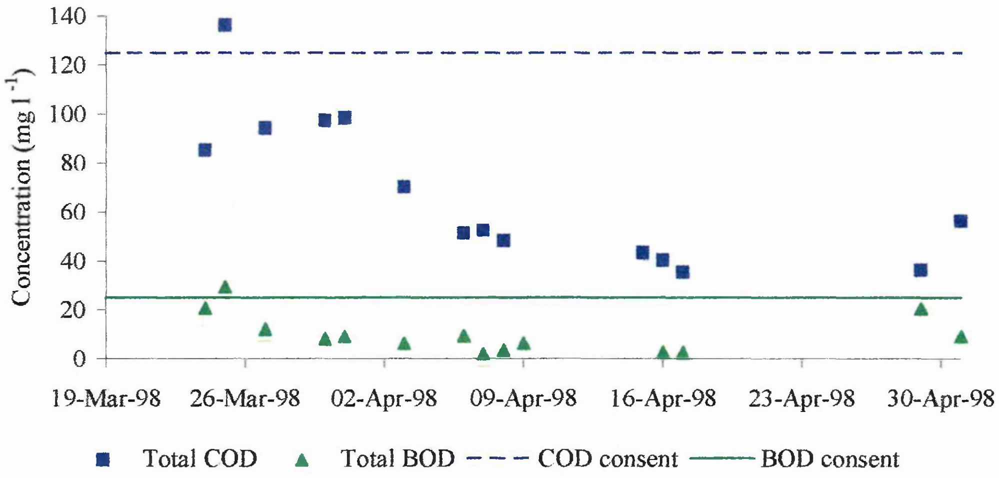

Figure 9. Final effluent total COD and total BOD_5 concentrations with consent limits for the Kaldnes hybrid plant at Colchester STW.

During the first week of the second phase the plant went through a period of partial nitrification when it was found that some denitrification was taking place in the final sedimentation tanks. The denitrification led to solids rising and it was these solids which caused the BOD_5 and COD concentrations to exceed the consent limit and future consent limit, respectively.

If the spot samples of the final effluent BOD_5 concentrations for the Kaldnes hybrid plant and the existing plant are compared for the second phase of the trial, (figure 10), it can be seen that during the first few days there were some samples that exceeded the consent level. This was probably due to the rising solids in the hybrid plant settling tanks in the first instance but the mixing of nitrified and non nitrified effluents may also have contributed.

The existing consent at Colchester STW was for BOD_5 and suspended solids only, meaning that the BOD_5 test was carried out using only 0.5 mg l^{-1} ATU for nitrification inhibition. However, in the presence of nitrifying bacteria this level of ATU can be insufficient to fully inhibit nitrification resulting in the BOD_5 test including the oxygen uptake for nitrification and hence giving an incorrectly high BOD_5 concentration. For plants with nitrification it is recommended that 2.0 mg l^{-1} ATU be used for nitrification inhibition in the BOD_5 test and if this is still not sufficient, the blue book method allows for the use of 5.0 mg l^{-1} ATU, (DoE - National Water Council, 1988).

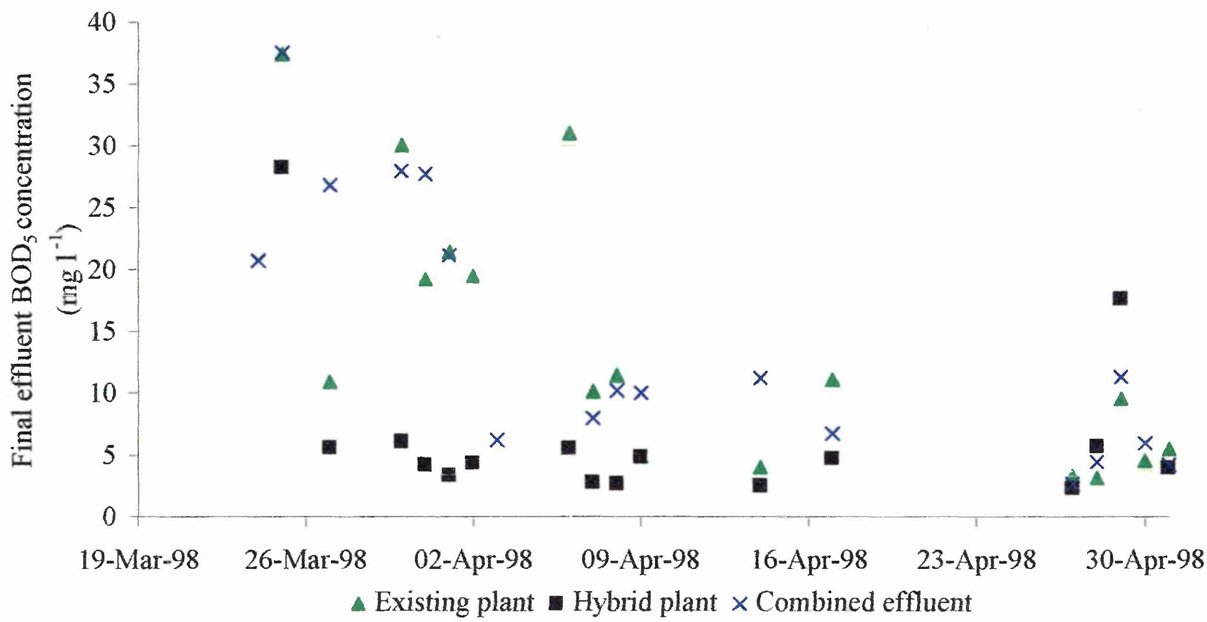

Figure 10. Final effluent BOD_5 concentrations for the Kaldnes hybrid, existing plant and combined effluents.

Nitrogen removal

The fate of nitrogen in the hybrid plant is summarised in table 6, from which it can be seen that no nitrification occurs in the Kaldnes zone and all of the ammonia degradation takes place in the activated sludge zone, as expected.

Table 6. Fate of nitrogen in the Kaldnes hybrid plant at Colchester STW.

	Average concentrations (mg l^{-1})		
	Influent	Kaldnes zone effluent	Final effluent
Ammonia - N	33	28	4
Nitrate - N	0.9	2.0	27.3
Nitrite - N	0.3	0.4	1.2
Total N	50	60	45
Filtered total N	41	37	36

Nitrification in the hybrid plant developed quickly once the lane had been separated from the rest of the biological process. This is highlighted by looking at the influent and effluent ammonia concentrations and the percentage ammonia removal in the hybrid plant (figure 11).

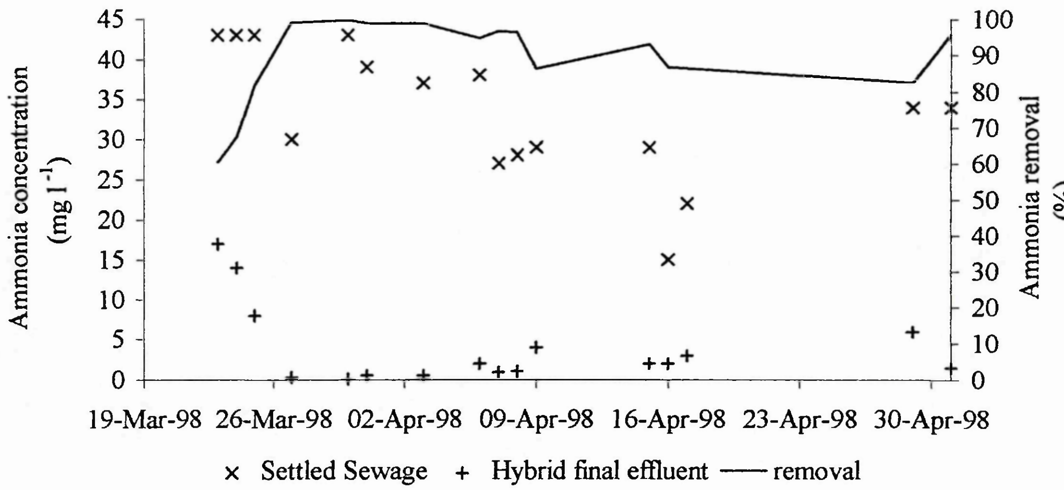

Figure 11. Settled sewage and final effluent ammonia concentrations and ammonia removal for the Kaldnes hybrid plant at Colchester STW.

A comparison of the nitrification achieved by each of the aeration lanes during the first and second phases has been made and is shown in terms of average effluent ammonia and nitrate concentrations, based on spot samples, (figure 12). As expected, once the hybrid lane had been separated the nitrification increased. This was due to the sludge age being increased and controlled to a greater extent and because the nitrifying bacteria from the hybrid were no longer diluted by mixing the different mixed liquors. A further factor in the improved ammonia conversion was probably the lower loading during the second phase. The F:M ratio during this period, based on the influent BOD_5 and MLSS only was 0.09 kg BOD_5 kg^{-1} MLSS d^{-1}, a load at which nitrification would be expected in a conventional ASP. During the second phase there was less nitrification in the existing diffused and surface aerated activated sludge lanes. This was expected as these lanes were operated for carbonaceous removal only.

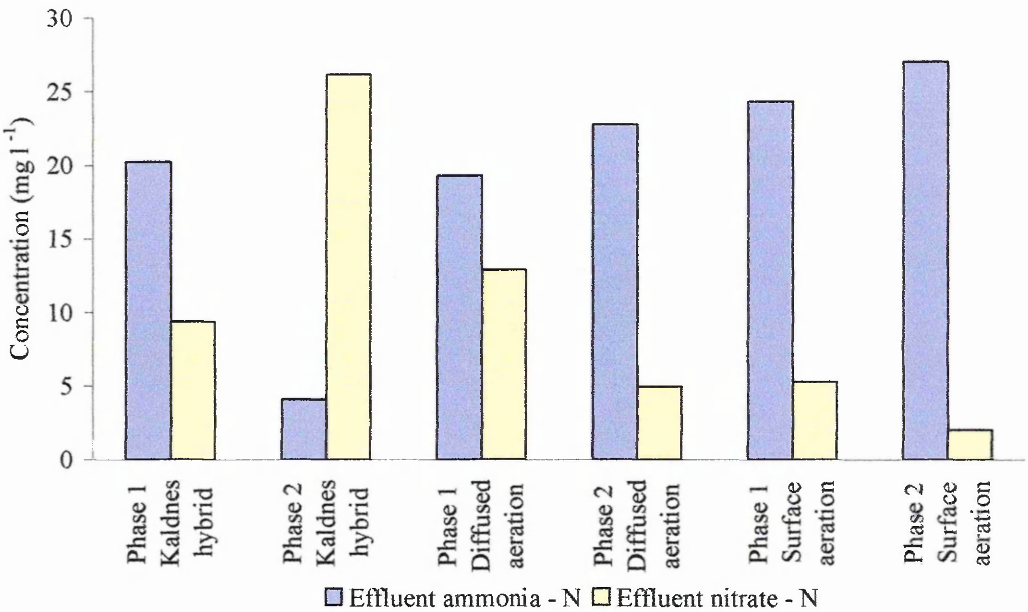

Figure 12. A comparison of the average effluent ammonia and nitrate concentrations in each of the secondary treatment processes during the 1st and 2nd phases of the Kaldnes hybrid trials at Colchester STW.

Settleability

The average $SSVI_{3.5}$ of the mixed liquor from the hybrid plant was 98 ml g^{-1} during the second phase of the trial. This shows a slight improvement (10%) in the sludge settling characteristics from the first phase of the trial. The slight improvement seen was not comparable to that reported for combinations of fixed and suspended growth elsewhere (Jones *et al.*, 1998; Parker *et al.*, 1993; Sunner *et al.*, 1998) where significant improvements in settleability were reported for the hybrid processes. The $SSVI_{3.5}$ is still not within the quoted range for a good settling sludge. This could be due to some release of nitrogen gas in the secondary settling process as there is no denitrification in the plant with the RAS being returned directly into the activated sludge tank. Bubbles of nitrogen gas were observed during the $SSVI_{3.5}$ test and occasionally caused the sludge to rise in the test cylinder. Poor settleability could also be due to poor floc formation caused by the low soluble substrate concentration entering the activated sludge zone.

Solids production

The total and volatile suspended solids during this period are summarised in table 7 from which it can be seen that the average suspended solids concentration increased in the Kaldnes zone as explained previously. Based on the average flows, suspended solids and BOD_5 data the sludge production and yield for the hybrid process as a whole were 593 kg TSS d^{-1} and 1.45 kg TSS (kg $BOD_5)^{-1}$ respectively. The calculated value for the sludge yield is very high, typical values being in the range 0.33-0.88 kg TSS (kg $BOD_5)^{-1}$ for plants with primary sedimentation, (Droste, 1997). Other MBBR hybrid plants have shown solids production from 0.54 kg TSS (kg $BOD_5)^{-1}$ (Rusten *et al.*, 1996) to 1 kg TSS (kg $BOD_5)^{-1}$ (Sunner *et al.*,1998). The high value obtained during the trial was probably due to the very low influent BOD_5 concentrations observed during the trial which resulted in low mass removals of BOD_5 and the increased solids due to mixed liquor transferring to the hybrid under low flow conditions as previously discussed.

Another factor could be the high particulate to total BOD_5 ratio observed during the trial period. It has been found that plants without primary sedimentation and therefore a higher particulate to total BOD_5 ratio going to secondary treatment have observed yields in the range 0.60 - 1.22 kg TSS (kg $BOD_5)^{-1}$, (Droste, 1997).

Table 7. Variation of total suspended solids (TSS) and volatile suspended solids (VSS) through the Kaldnes plant during the second phase of the Kaldnes hybrid trial.

		Settled Sewage	Kaldnes effluent	Mixed liquor	Final effluent
TSS (mgl^{-1})	Average	174	502	2290	25
	Maximum	286	1660	3600	68
	Minimum	29	116	1280	10
VSS : TSS	Average	0.69	0.82	0.78	...

3.4.4 Summary

After the Kaldnes hybrid plant had been separated from the other activated sludge lanes the performance improved with full nitrification and a final effluent within the future consent being achieved consistently. During April 1998 very heavy rainfall in the UK meant that the influent sewage was very dilute. This factor and the capacity limitations of the pumps used for mixed liquor transfer resulted in the hybrid plant being under loaded during the second phase of the trial.

A further complication was caused by the mixing of nitrified and non-nitrified effluents in the final effluent discharge channel which resulted in the works exceeding its final effluent consent on a number of occasions. As works compliance was critical it was decided to stop the trial on 1st May 1998 and run the whole plant for BOD_5 removal only in order to meet the existing final effluent consent.

Although the initial results from the second phase of the trial looked promising, a better judgement of the hybrid process could have been made after a further month or two of running the hybrid separately from the rest of the plant. This would have allowed the load to increase to normal levels and the activated sludge to stabilise so that the effects of the biofilm process on the sludge settleability could have been evaluated more fully. There is also very little data from operation of the hybrid at a sludge age of 5 days and more would be required to fully assess the performance of the hybrid process at this sludge age.

3.5 CONCLUSIONS

- The full plant could not nitrify with mixed sludges and the low sludge age achieved during the first phase of the trial.

- After the hybrid had been separated from the other activated sludge lanes, a nitrified effluent was achieved for sludge ages of 11 days and 5 days and an average organic load of 0.8 kg BOD_5 m^{-3} d^{-1}. The average effluent ammonia concentration during this period being 4 mg l^{-1}. This was a promising indication that the hybrid process would enable the Colchester plant to meet its new ammonia consent.

- The settleability of the sludge after the Kaldnes process was not very good, however with the process upsets which occurred this was perhaps to be expected. The hybrid sludge was no worse than the other sludges on site which indicates that it is the site rather than the specific process which causes the poor settleability.

3.6 FUTURE WORK

- The hybrid process needs to be tested at a higher load, up to 5 kg BOD_5 m^{-3} d^{-1}, for a longer period and higher F:M ratios.

- The hybrid process needs to be tested at a sludge age of 5 days for a longer period.

- More information is required on the effect of the hybrid process on the activated sludge settleability.

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CHAPTER 4 - PROCESS MODELLING OF A FULL SCALE MBBR/AS PLANT

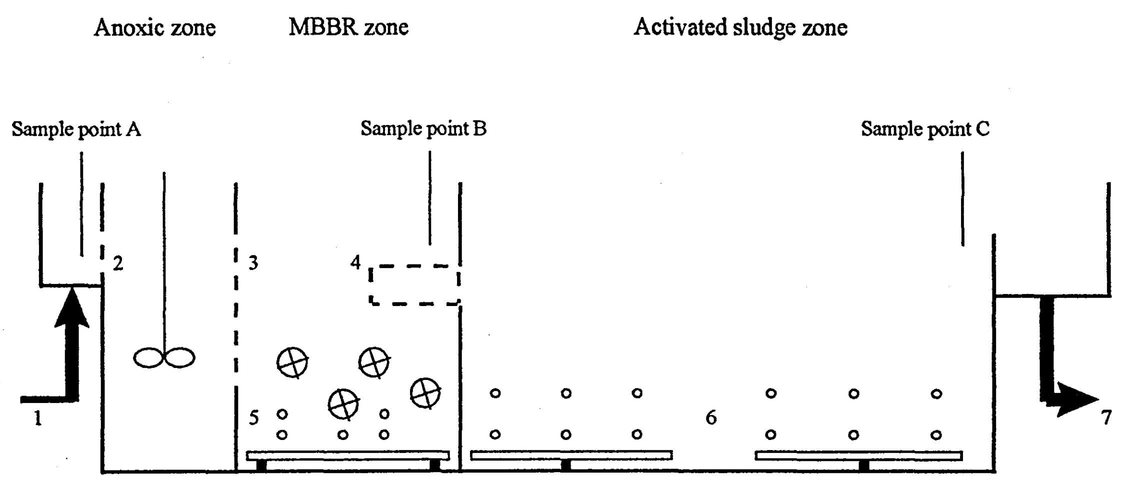
4.1 INTRODUCTION

Combined fixed film : activated sludge plants have been investigated as process options for upgrading existing wastewater treatment plants in order to treat increased flows and loads and to meet tighter legislated discharge consents. Bonhomme *et al.* (1990) observed that by incorporating fixed biomass in the activated sludge reactor the F:M ratio could be lowered and the sludge age increased, resulting in increased carbon and nitrogen removal. Trickling filters and activated sludge plants have been used in combination for secondary treatment of municipal wastewater (Daigger *et al.*, 1993). The moving bed biofilm reactor (MBBR) has been found to be a reliable biofilm process for organic and nitrogen removal which has low head losses and has no need for backwashing, (Pastorelli *et al.*, 1997). Rusten *et al.* (1996) developed a pilot scale moving bed biofilm reactor - solids contact reaeration process (MBBR/SCR) which achieved final effluent BOD_5 concentrations below 10 mg l^{-1} when treating settled wastewater at loads up to 5.0 kg m^{-3} d^{-1}. A moving bed biofilm reactor : activated sludge (MBBR/AS) process has been investigated as a possible solution for upgrading an existing activated sludge plant for ammonia removal within the existing tank volumes. A full scale trial plant was constructed by incorporating an anoxic zone and a MBBR zone into an existing aeration tank. In order to further investigate the process, a model of the MBBR process has been developed by Hydromantis, Inc., Canada and Anglian Water, UK. This paper reports the application of the model, in use for the first time, to a full scale MBBR/AS process in order to test its suitability for upgrading the plant and different operating scenarios.

4.2 MATERIALS AND METHODS

4.2.1 MBBR/AS hybrid plant

The plant consisted of three zones, a 171 m^3 anoxic zone (not used during these trials), a MBBR zone with a volume of 487 m^3 and an activated sludge zone with a volume of 1970 m^3 (figure 1). The MBBR had a 50 % fill of Kaldnes media (Kaldnes Miljoteknologi AS, Norway) with a specific surface area of 500 m^2 / m^3 and coarse bubble diffused aeration. The activated sludge plant had fine bubble diffused aeration. Initial work focused on the MBBR and once the model had been calibrated for this zone daily variations through the plant as a whole were used to calibrate the AS zone and final sedimentation tanks.



Key: 1. Settled sewage from primary sedimentation tanks; 2. Copa sacks in inlet chamber; 3. Flat sieves in wall between anoxic and MBBR zones; 4. Pipe sieves between MBBR and activated sludge zones; 5. Coarse bubble aeration in MBBR zone; 6. Fine bubble aeration in activated sludge zone; 7. Mixed liquor to final sedimentation tanks.

Figure 1. Full scale MBBR : activated sludge hybrid trial plant.

4.2.2 Sampling procedure

For modelling the whole process, daily composite samples collected during process proving were used to characterise the influent and for model calibration. Sampling was carried out at points A, B, C and also from the overflow to the final sedimentation tanks.

For the more detailed investigation of the performance of the MBBR zone, a four day diurnal sampling survey was carried out at the works. Using autosamplers (Bühler Montec, Salford, UK), samples were taken every 15 minutes and combined to obtain 2 hourly and 6 hourly composite samples. The influent sample was taken from the inlet chamber to the anoxic zone and the effluent from the MBBR reactor by the outlet pipe sieves (sample points A and B respectively in figure 1). Ice packs were used in the autosamplers to keep the samples cool until collection.

Flow data for the sampling period was collected through the site telemetry systems. Sewage temperature was measured each day using a mercury thermometer and the dissolved oxygen (DO) concentration in the MBBR zone was monitored using a portable DO probe (pHOX).

4.2.3 Sample analysis

Daily composite samples were analysed on site for suspended solids, total nitrogen, filtered total nitrogen, ammonia, nitrate, nitrite, total COD and filtered COD using Dr Lange test kits. A sub-sample was also sent to the Anglian Water laboratory at Whitlingham to be analysed for total 5 day biochemical oxygen demand (BOD_5) and filtered BOD_5 for all samples and for volatile suspended solids, BOD_{20} and filtered BOD_{20} for the settled sewage, Kaldnes effluent and mixed liquor samples.

The 2 hourly samples were analysed at Anglian Water's Whitlingham laboratory for BOD_5, total suspended solids (TSS), total Kjeldahl nitrogen (TKN) and ammonia in order to obtain the diurnal profiles. The 6 hourly composite samples were analysed for COD, BOD_{20}, TSS and volatile suspended solids (VSS). Sub-samples of the 6 hourly composite samples were filtered on site using glass fibre filters (Whatman GF/C) and were analysed for filtered BOD_5, filtered COD, filtered BOD_{20} and total oxidised nitrogen. The analysis of the 6 hourly composite samples was used to calculate the influent ratios required for the model input. BOD samples were analysed using the blue book method (DoE - National Water Council, 1988) and all other samples were analysed using standard methods (APHA, 1992).

4.2.4 Data analysis

The 50 percentile values of each of the wastewater characteristics were calculated and used for calculating the influent ratios for the model input and as the influent and effluent values for the steady state calibration. The influent ratios which had to be calculated for the model input were: filtered BOD_{20} : BOD_{20} ; ammonia : TKN; particulate organic nitrogen : total organic nitrogen; particulate COD : VSS; VSS : TSS and BOD_5 : BOD_{20}.

In addition to the influent ratios, the following ratios in the Kaldnes reactor also had to be calculated: particulate COD : VSS; VSS : TSS and BOD_5 : BOD_{20}.

4.2.5 The MBBR model

The MBBR model developed by Hydromantis and Anglian Water is based on the existing biofilm and Mantis activated sludge models in GPS-X. The biofilm is treated as a single plane area with six layers. The first layer is a liquid film layer at the boundary of the bulk liquid and the biofilm and the remaining five layers are layers of biofilm. The model simulates diffusion into the biofilm and between biofilm layers. Diffusion is based on Fick's law (equations 1 and 2), with the default diffusion rate values being those of the states in water. Reduction of diffusion due to the biofilm is allowed for by the use of a constant factor.

Accumulation in the liquid film

$$A_a\delta_L\left(\frac{dS^L}{dt}\right) = Q_L\left(S^L_{j-1} - S^L_j\right) - K_M A_a\left(S^L_j - S^{BLi}_j\right) + K_{ML}A_a\left(S^0 - S^L_j\right) \quad \text{Equation 1}$$

where, A_a surface area of biofilm through which transport is occurring (m^2)

δ_L thickness of the attached liquid layer (m)

S^L_j substrate concentration in liquid film horizontal section j (mg L^{-1})

t time (days)

S^{BLi}_j substrate concentration at biofilm liquid interface section j (mg L^{-1})

S^0 saturated liquid-film substrate concentration (mg L^{-1})

Q_L volumetric flow rate of attached liquid layer (L d^{-1})

K_M mass transfer coefficient from liquid to biofilm (m d^{-1})

K_{ML} oxygen transfer coefficient from air to liquid film (m d^{-1})

Accumulation in the biofilm

$$\frac{\partial S}{\partial t} = -D_S\frac{d^2S}{dy^2} + \left(S^B_{j-1} - S^B_j\right)\left(\frac{Q_B}{A\delta_B}\right) - R_S \quad \text{Equation 2}$$

where, A surface area of attached micro-organisms (m^2)

D_S state variable diffusion coefficient (m^2 d^{-1})

Q_B volumetric flow rate of attached biofilm layer (L d^{-1})

R_S substrate utilisation rate (mg L^{-1} d^{-1})

S state variable concentration in layer (mg L^{-1})

S^B_j state variable concentration in attached biofilm layer j (mg L^{-1})

t time (days)

y thickness of biofilm layer (m)

δ_B attached biofilm thickness in layer (m)

Substrate utilisation is due to the biological processes taking place in each of the biofilm layers and is described by the Mantis activated sludge model which is given in a matrix form in the GPS-X Technical Reference (Hydromantis, 1997). The Mantis model is based on the IAWQ activated sludge model number 1 (Henze *et al.*, 1987) but incorporates temperature dependency for the kinetic parameters, two extra growth processes for heterotrophic and autotrophic bacteria and does not model alkalinity (Hydromantis, 1997).

The biological model has been incorporated into a plug flow tank icon within a generic MBBR model library in GPS-X. This allows the MBBR to be incorporated into any plant layout that uses the other process models available within the software. The MBBR tank can have up to ten zones. The influent, return sludge (if any) and the air flow can be fed into any of the zones. Internal recycles can be set up between the zones, with or without the biofilm support media. A second input line enables methanol addition to be modelled, e.g. for biological nutrient removal applications. The percentage media fill in each reactor and the specific surface area of the media are defined by the user.

The settling model used for the final sedimentation tanks was the 1 dimensional settling model described by Takacs *et al.* (1991). No biological processes were simulated in the final sedimentation tanks.

4.2.6 Model plant layout

In the model layout a flow split was used on the RAS line in order to try to model a phenomenon observed on the full scale plant. At periods of low flow, due to the plant hydraulics, mixed liquor was observed to flow from the surface aeration lanes into the hybrid plant and from the aeration zone into the MBBR zone. As it was not possible to characterise these flows an attempt to approximate them was made by diverting some of the RAS flow direct to the MBBR zone. This was found to be necessary for the daily composite samples only, i.e. for steady state calibration.

4.2.7 Influent model

A number of different influent models are defined within GPS-X, including BOD based, COD based, BODCOD based and states models. For this work the BOD based influent model was used. The BOD based influent takes BOD_5, TSS and TKN as the influent variables and uses the defined influent ratios to calculate the state variables used in the model as shown in figures 2 and 3.

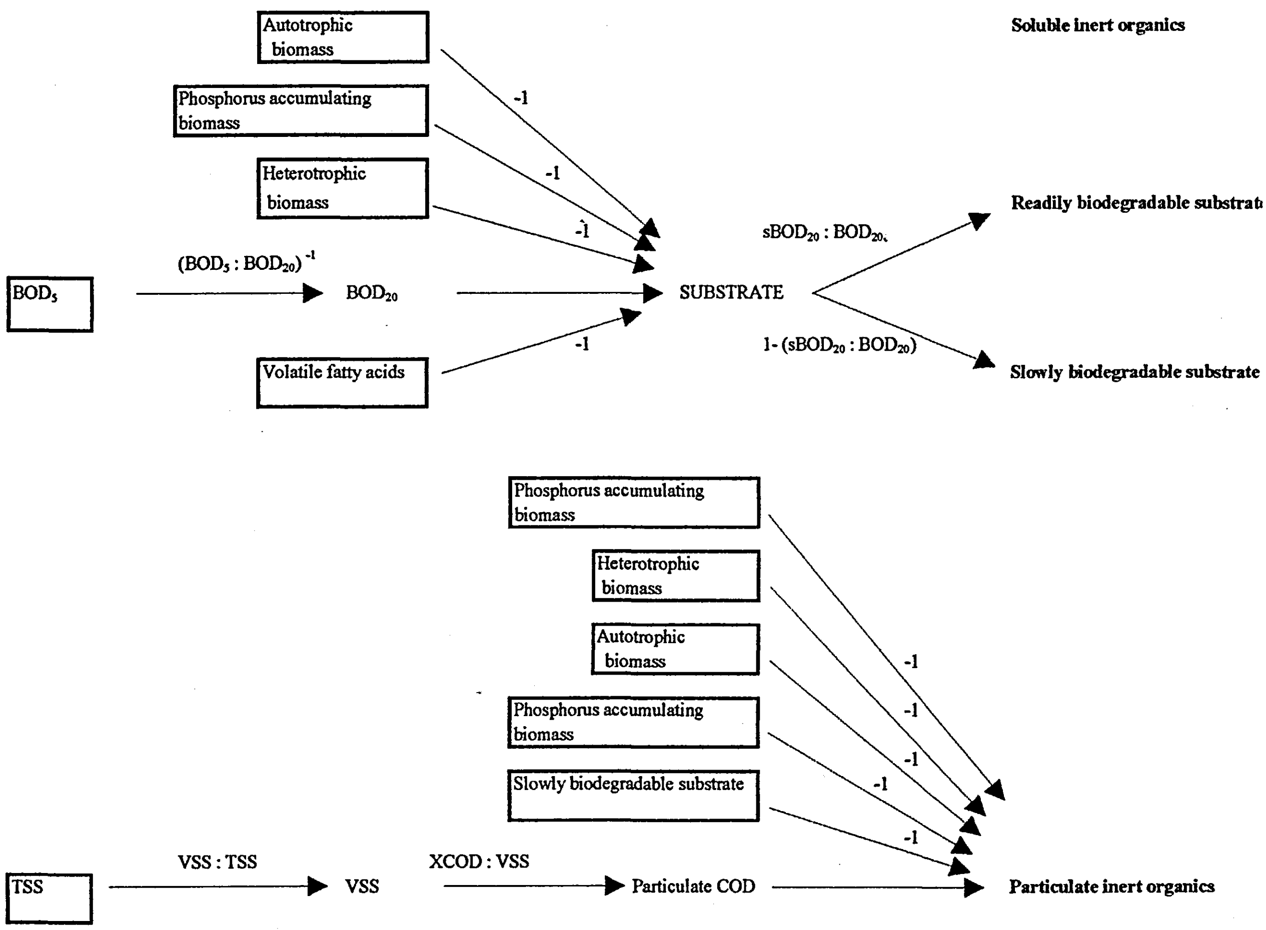


Figure 2. COD fractionation within GPS-X BOD based influent model.

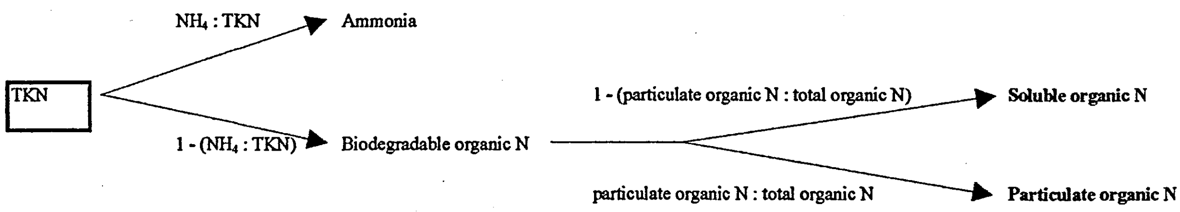


Figure 3. Nitrogen fractionation within GPS-X BOD based influent model.

4.2.8 Model calibration

The first stage of calibration involved the use of 50 percentile values of the wastewater characteristics to calibrate the model in steady state conditions. The kinetic and stoichiometric parameters were initially kept the same as the default values. They were then adjusted until the model results were within 20% of the measured data.

Once a good fit had been achieved, the model was re-calibrated using the data from the period of discrete sampling. Once the model of the MBBR had been calibrated the daily composite data was used to calibrate the activated sludge zone.

For dynamic calibration, data from one day of the sampling program was selected and replicated 4 times so that a 5 day simulation could be run, allowing the model to settle to a repeated output. The kinetic and operational parameters were then adjusted again to fit the model output to the actual dynamic data. Model fitting was carried out by visually comparing the model prediction with the actual data. Once a good visual fit had been obtained the model was validated using the full set of sampled data. No parameters were changed during validation.

Once the MBBR zone had been calibrated and validated the predicted effluent from the hybrid plant as a whole could be assessed. Because the trials were brought to a close before settling tests could be carried out, the model for the final sedimentation tanks could only be crudely calibrated using MLSS, final effluent TSS and RAS solids results from earlier in the trials.

Once the model had been calibrated, an influent model based on the design flows and loads for the full scale upgrade to a hybrid plant was set up and the possibility of using a hybrid plant for the upgrade was assessed. The average results and results for a typical diurnal variation were investigated.

4.3 RESULTS

4.3.1 Data analysis

The average concentrations (Table 1) and influent ratios derived from them (Table 2) are summarised in table 1 for the period of daily composite sampling across the whole plant. It can be seen in tables 1 and 2 that there are some errors in the analysis, particularly the BOD analysis, with some BOD_5 values being greater than the BOD_{20} value for the same sample. In these cases where incorrect ratios have been given by the analysis, e.g. the BOD_5:BOD_{20} ratio greater than 1, the default value was used for the modelling.

The results of the diurnal sampling program across the MBBR showed a number of problems with the analysis of the samples. The most obvious errors were the values of BOD_{20} being lower than those for BOD_5 and the ammonia values the same as the TKN values. There is no way of telling which results are incorrect although it was considered most likely to be the two tests which are carried out less frequently, i.e. BOD_{20} and TKN. The BOD_{20} and TKN values are important since they are used in the influent model to define the influent COD and nitrogen states which are required for using the biological process model. Hence if the BOD_{20} and TKN values are incorrect, the influent COD and nitrogen states will not be correct.

Table 1. Average concentrations of wastewater characteristics across the hybrid plant (19/3/98 - 1/5/98).

	50 percentile concentrations (mg l^{-1})			
	Influent	Kaldnes zone	Activated sludge zone	Final effluent
TSS	163	355	2079	19
VSS : TSS	0.72	0.82	0.78	...
BOD_5	59	195	1431	7
Filtered BOD_5	21	7	9	3
BOD_{20}	90	187	856	...
Filtered BOD_{20}	15	13	10	...
COD	414	607	2853	58
Filtered COD	118	64	43	41
Total N	45	...	...	...
Filtered total N	38	...	...	...
NH_4-N	31	26	3	2
NO_3-N	0	0	...	25

Table 2. Stoichiometric ratios required for GPS-X BOD based influent model.

	Stoichiometric ratios		
	Influent	Kaldnes zone	Activated sludge zone
$BOD_5 : BOD_{20}$	0.66	1.04	1.67
Filtered $BOD_{20} : BOD_{20}$	0.17	0.07	0.01
Particulate COD : VSS	2.52	1.87	1.55
BOD_5 : COD	0.14	...	...
NH_4-N : TKN	0.69	...	...
Particulate organic N : total organic N	0.50	...	...

(Shaded area indicates error with analysis as $BOD_5 : BOD_{20}$ cannot be greater than 1)

Figure 4 highlights a further factor which contributed to errors in analysis of the diurnal sampling data. It can be seen that during periods of very low flow to the works there are very high TSS and BOD_5 concentrations in the influent to the hybrid plant. This is explained by site hydraulics which meant that during low flow periods mixed liquor from the rest of the plant flowed back into the hybrid plant.

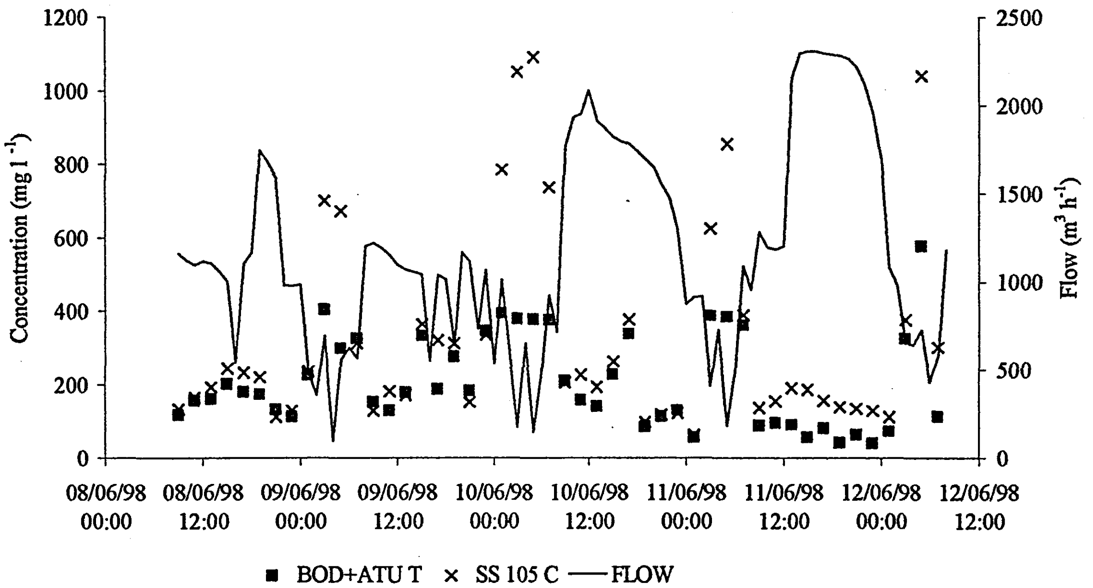


Figure 4. Variation of hybrid influent TSS and BOD_5 with total flow to the works.

In order to have a more realistic assessment of the average influent concentrations and stoichiometric ratios, the average values were recalculated with the results for the low flow periods omitted. This improved the results for the BOD_5:BOD_{20} ratio but the ammonia : TKN ratio was still incorrect.

The soluble COD and soluble BOD_5 in the influent and hence the effluent will also be predicted incorrectly as they are calculated from the influent BOD_5. When the influent BOD_5 peaks, the modelled influent soluble BOD_5 and soluble COD will also peak. However, in reality when the BOD_5 peaks at very low influent flow rates, the soluble BOD_5 and soluble COD are low as the low flow periods are during the night, corresponding to low load periods.

To enable the influent model to be defined the ratio from the composite sampling period was used and the influent TKN for the model calculated from the measured ammonia concentrations. The average influent and effluent concentrations and the stoichiometric ratios calculated for the diurnal sampling period are summarised in tables 3 and 4.

Table 3. Characterisation of settled sewage and Kaldnes reactor effluent at Colchester STW. Summary of analysis of data collected 8th - 12th June 1998.

Wastewater characteristic	Sampling regime	50 percentile concentration (mg l^{-1})	
		Settled sewage (influent)	Kaldnes reactor (effluent)
TSS	2 hr composite	163	207
BOD_5	2 hr composite	125	163
TKN	2 hr composite	46*	...
Ammonia	2 hr composite	32	31
Filtered BOD_5	6 hr composite	26	12
COD	6 hr composite	371	418
Filtered COD	6 hr composite	104	76
BOD_{20}	6 hr composite	188	228
Filtered BOD_{20}	6 hr composite	33	38
Filtered TKN	6 hr composite	39*	...
TSS	6 hr composite	255	275
VSS	6 hr composite	209	227

(* values calculated using ratios from daily composite sampling period).

Table 4. Stoichiometric ratios calculated for diurnal sampling period.

Ratio	Settled sewage (influent)		Kaldnes reactor	
	Calculated	Default	Calculated	Default
filtered BOD_{20} : BOD_{20}	0.18	0.10	...	...
ammonia : TKN	0.69*	0.70	...	...
particulate organic nitrogen : total organic nitrogen	0.5*	0.33	...	...
particulate COD : VSS	2.00	2.20	1.99	1.48
VSS : TSS	0.82	0.60	0.83	0.75
BOD_5 : BOD_{20}	0.66	0.66	0.71	0.66

(* These values taken from daily composite sampling period)

The wastewater temperature during the sampling period was fairly constant at 12 °C and the average DO concentration was 0.8 mg l^{-1} in the Kaldnes reactor.

4.3.2 Flow data

During the period of daily composite sampling daily flows were read manually from the inlet and RAS flowmeters. The average influent flow rate was 5891 m^3 d^{-1} and the average RAS rate was 6573 m^3 d^{-1}. The flow rates to the hybrid plant during the intensive sampling period were monitored by the site telemetry system but a fault with this system meant that flow data was not recorded for most of the sampling period. The relationship between the flow to the hybrid and the total flow to the works was obtained for existing data from the telemetry system. The relationship was then used to estimate the flow to the hybrid from the known flow to the works as a whole during the sampling period. The works inlet flow and the flow to the hybrid plant are shown in figure 5 with the estimated values indicated. Although not an ideal solution this procedure allowed some use to be made of the wastewater quality data collected. Once the hourly flow rates had been calculated as described, the 50 percentile value (15 000 m^3 d^{-1}) for the flow rate during sampling was calculated for use as the steady state flow rate during model calibration.

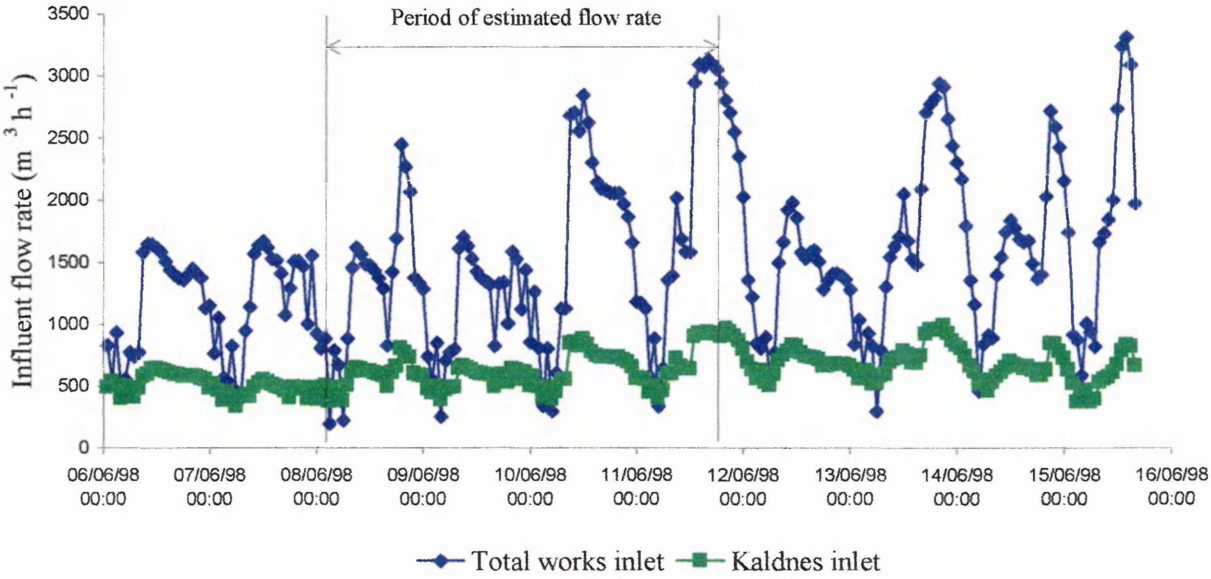


Figure 5. Total works inlet flow and flow to hybrid plant for the period just prior to and just after the sampling was carried out.

4.3.3 Steady state calibration

The steady state model predictions are compared to the observed results in table 5. The aim of the steady state calibration was for the model prediction to be within 20% of the actual data. The level of 20% was selected to correspond with the anticipated margin of error in the sample analysis. The only parameters changed from the default were the heterotrophic yield, Y_H, which was increased by 10% since the concentration of heterotrophs in the influent was taken to be 0 mg COD l^{-1} (Henze *et al.*, 1987) and the reduction of diffusion through the biofilm which was set to 80 % of the values in water (Characklis and Marshall, 1989). Although the filtered BOD_5 was outside of the range intended the model was not changed any further as this would have moved other determinands out of calibration.

Table 5. Model v actual data for steady state calibration.

Parameter	Model prediction ($mg\ l^{-1}$)	Observed results ($mg\ l^{-1}$)	Difference (%)
TSS	175	207	15
COD	351	418	16
Filtered COD	63	76	17
BOD_5	124	163	24
Filtered BOD_5	15	12	25
Ammonia	32	31	3

4.3.4 Dynamic calibration

During dynamic calibration no further parameter changes were required. From figure 6 it can be seen that the model prediction follows the trends in the measured data but does not match all of the highest and lowest points. The predicted solids concentration does not predict the highest peak but this is considered to be an error in either sampling or analysis as it is so far outside the range of the other values. The soluble BOD is over predicted by the model which is due to the influent soluble BOD also being over predicted because of the way in which the BOD based influent model is defined. Adjusting the influent to decrease the filtered BOD would also reduce the soluble COD. Hence for this data set it is not possible to have both the soluble BOD and the soluble COD model predictions fitting the measured data.

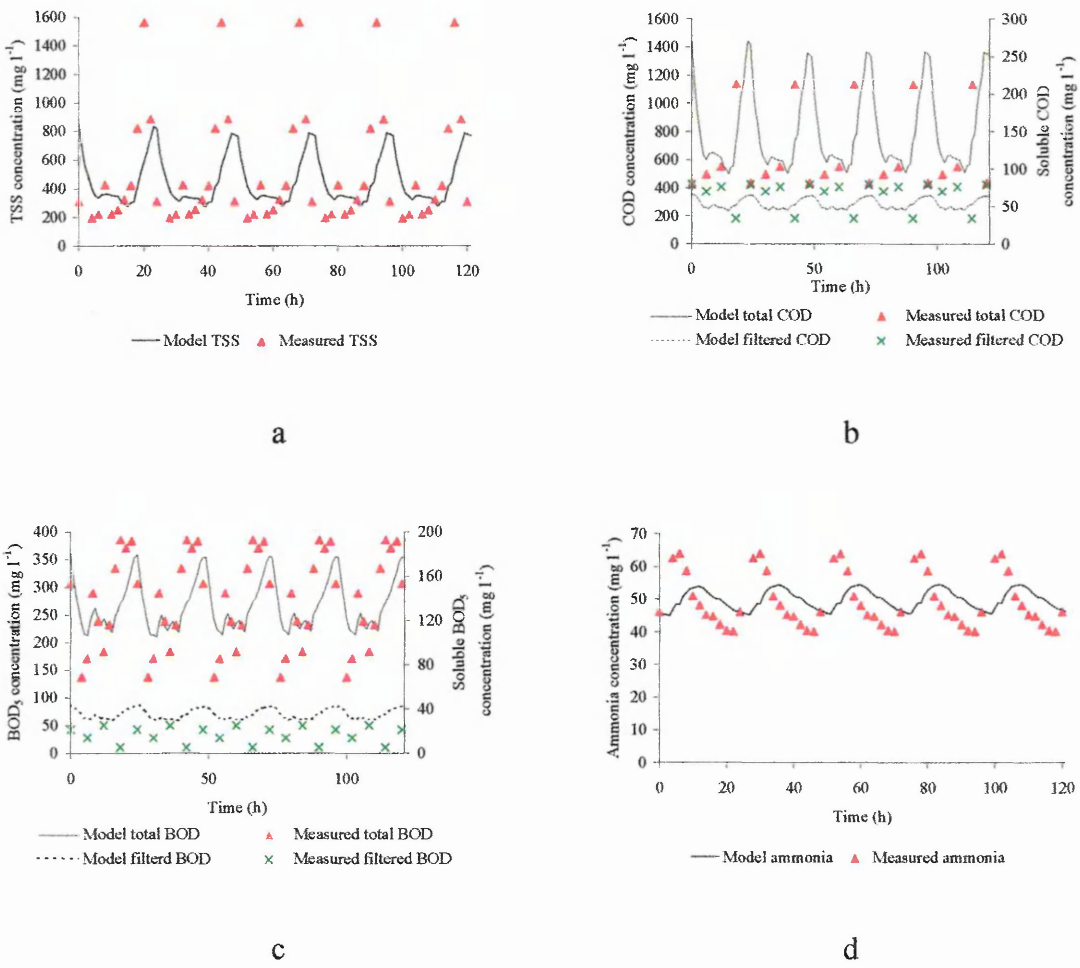
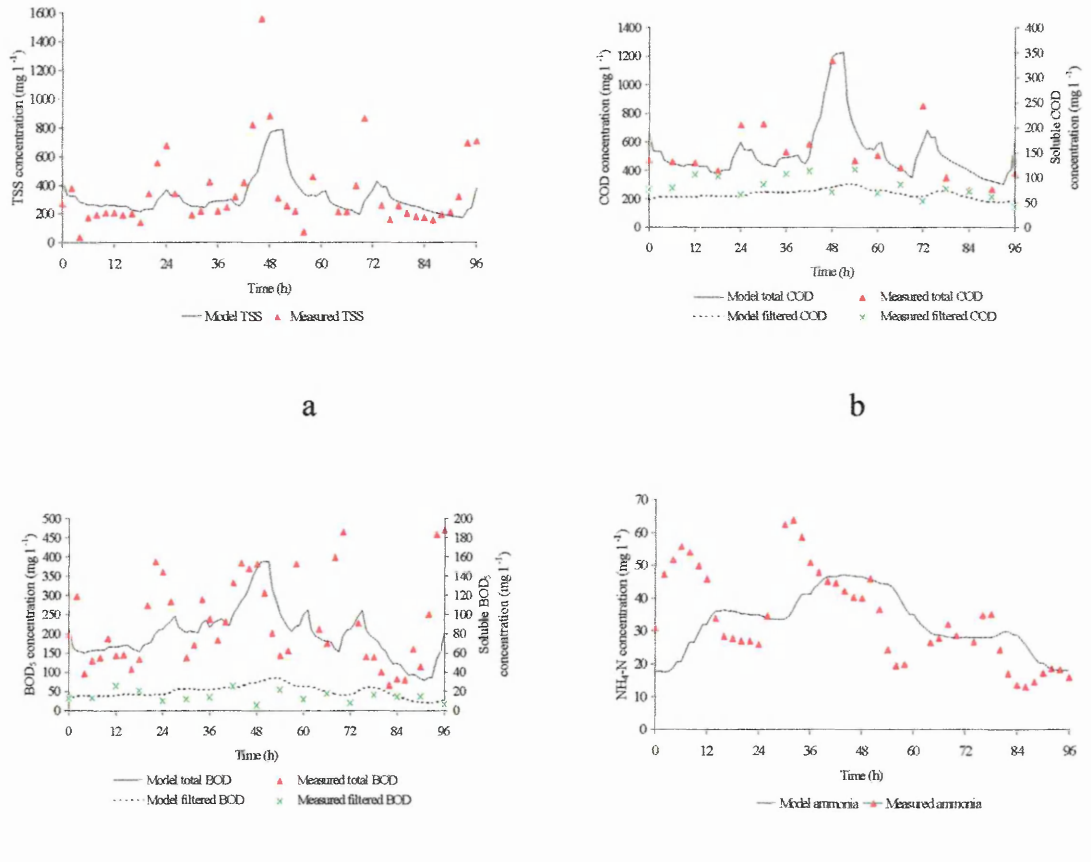


Figure 6. Comparison of model prediction and measured data after dynamic calibration.

(a modelled v measured effluent COD. b modelled v measured effluent BOD_5. c modelled v measured effluent total suspended solids. d modelled v measured effluent ammonia.)

4.3.5 Model validation

Once the dynamic calibration was considered acceptable the model was validated against all four days data. No changes were made to any of the kinetic or stoichiometric parameters during validation. The results of the validation are shown in figure 7.



a b

c d

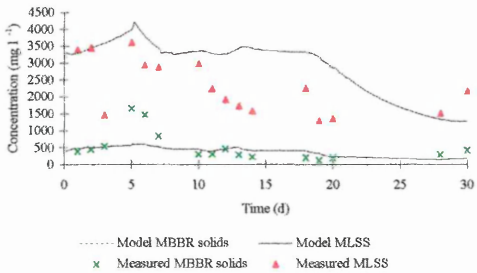
Figure 7. Comparison of model prediction and measured data after dynamic validation.

(a modelled v measured effluent COD. b modelled v measured effluent BOD_5. c modelled v measured effluent total suspended solids. d modelled v measured effluent ammonia.)

Given the problems identified with the influent analysis, the fit to the data was considered acceptable. All the major trends are predicted by the model although the some of the extreme values are not predicted by the model. The parameters used for the dynamic calibration of the MBBR were then used along with the daily composite sampling data model the performance of the plant as a whole.

4.3.6 Model calibration for complete hybrid plant

Once the MBBR had been calibrated, the daily composite data was used to try to calibrate the activated sludge stage of the process. Figure 8 shows the results of this calibration.



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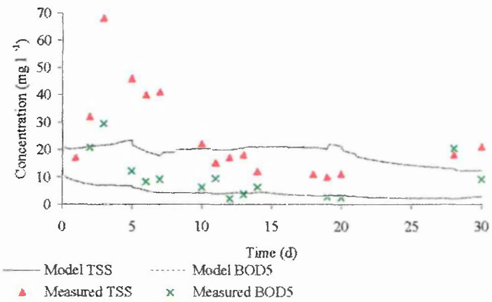

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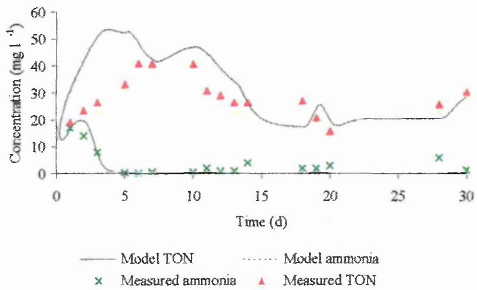

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Figure 8. Comparison of model prediction and measured data for the hybrid plant. (a MLSS in MBBR and ASP, b Final effluent TSS and BOD_5, c Final effluent ammonia and TON)

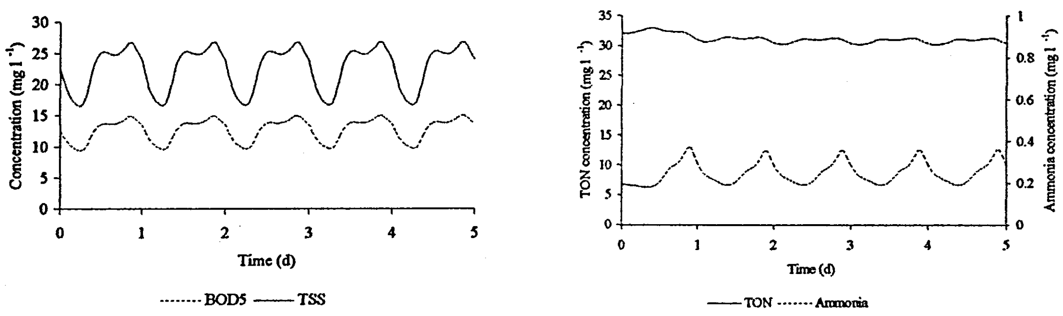
The model prediction of solids in the MBBR and the aeration tank are of a similar magnitude to the measured data but it can be seen that between day 10 and 15 the MLSS in the aeration tank decreased significantly and that the model does not predict this decrease. The reason for the decrease in MLSS was the failure of one of the mixed liquor transfer pumps which resulted in mixed liquor being washed out of the aeration lane over the stop log in the outlet channel. Since this additional sludge "wastage" was not an input to the model it is not unexpected that the model did not predict the reduction of MLSS. During the first few days of sampling rising sludge was observed in the final tanks. This is a mechanism not included in the model and hence the high final effluent TSS were not predicted by the model. The model gives a reasonable prediction of the increase in nitrification over the first few days and of the ammonia and TON concentrations generally. The nitrification kinetics had to be adjusted in order to prevent over-prediction of TON in the final effluent and some of the settling parameters had to be changed to calibrate the solids and BOD_5 in the final effluent. All parameters that were changed are summarised in table 6.

Table 6. Parameters adjusted during model calibration.

Parameter	Value used	Default value	Literature values
$\mu_{max, A}$ (d^{-1})	0.55	0.75	0.15 - 0.24 (Henze *et al.*,1997)
Y_A (g COD (g N)$^{-1}$)	0.18	0.15	0.15-0.2 (Henze *et al.*,1997)
Reduction of diffusion in biofilm	0.8	0.5	0.8 (Characklis and Marshall, 1989)
Flocculent zone settling parameter (m^3 g^{-1})	0.00125	0.0025	10^{-2} (WRc, 1999)
Non-settleable fraction of solids	0.005	0.001	$(1.23\text{-}2.59)\times10^{-3}$ (Takacs *et al.*,1991)

4.3.7 Scenario analysis

A model layout of the full scale hybrid upgrade was constructed and the stoichiometric ratios used in this work were applied to the design loads and flows. The predicted final effluent from the plant under average conditions was 22 mg l^{-1} TSS, 12 mg l^{-1} BOD_5, 0.2 mg l^{-1} NH_4-N and 26 mg l^{-1} TON. This predicted effluent is within the new consent obligation for Colchester STW which is 35/35/15 mg l^{-1} TSS/BOD_5/ammonia based on spot samples and a 95 %ile basis with an additional UWWTD consent of 25 mg l^{-1} BOD and 125 mg l^{-1} COD based on daily composite samples again on a 95 %ile basis. The typical diurnal variation in flow and load observed during the intensive sampling period was then applied to the average values to give an indication how the plant would perform under daily fluctuations. The results of this are summarised in figure 9 which shows that at all times during the day, the effluent quality is within the new consent for the works.



a b

Figure 9. Predicted effluent quality for full scale upgrade at Colchester with typical diurnal variations applied.

Although there are reservations over the confidence in the model, which will be considered in the discussion, figure 9 backs up experimental evidence that the use of the MBBR/AS hybrid process would enable a new ammonia consent to be met at Colchester STW.

4.4 DISCUSSION

The problems encountered with the analysis of some of the wastewater quality characteristics have been reported by other authors. Smith and Dudley (1997) found that 5 NAMAS accredited laboratories in the UK had errors in measured values of BOD_{20}. As BOD_{20} and filtered BOD_{20} are the main characteristics used to define the influent COD fractions when using the BOD based influent in GPS-X (figure 2) any errors in its measurement will cause a significant reduction in the accuracy of the model.

The COD fractions, as a percentage of the total influent COD, found using the relationships shown in figure 2 and the measured influent ratios are compared to previously reported values in table 7.

Table 7. Comparison of COD fractions calculated with values reported in the literature (Henze, 1992)

Fraction	BOD model (diurnal data)	BOD model (composite data)	COD model (composite data)	Literature values
S_I	11 %	11%	9%	3 - 10 %
S_S	10 %	4%	20%	16 - 33 %
X_I	33 %	64%	69%	4 - 13 %
X_S	46 %	21%	2%	40 - 60%

The soluble inert COD, S_I, was calculated as 90 % of the works final effluent filtered COD (Siegrist and Tschui, 1992 and Xu and Hultman, 1996). It can be seen that the model influents over predict the particulate inert COD and in the case of the BOD based models the influent COD is not predicted correctly. This could be due to errors in measurement of one or more of the wastewater characteristics used to calculate the ratios used to fractionate the influent. On site factors could affect this, e.g. solids carry over from the primary tanks could lead to a higher solids : BOD ratio which would lead to over prediction of X_I.

Generally it is assumed that the COD fractions remain constant as a proportion of the total influent COD over time (Henze, 1992). For the plant considered during this work it can be seen that there are distinct periods of unusually high influent suspended solids, BOD_5 and COD which correspond to low influent flow rates. During these periods the influent ratios are different to those of the rest of the time and so using a single average ratio for the whole period will have resulted in errors in the prediction of the influent COD. The problem can be overcome by using different influent files with the different influent ratios. However, there is a trade off in terms of the complexity of the model.

The kinetic parameters were left at the default values since the model predicted the observed trends without adjusting them. Varying the parameters within reasonable ranges did not significantly improve the model fit to the measured data (figure 10). As there was no nitrification in the MBBR zone, the model output is less sensitive to changes in kinetic parameters, the model being relatively insensitive to variations in heterotrophic parameters (Henze *et al.*, 1987). The largest effect is on the filtered BOD_5 as would be expected since the particulate substrate has to be hydrolysed before it diffuses into the biofilm, a slower process. The parameter which has the greatest effect is μ_{max}, and it is this parameter that Henze *et al.* (1995) recommend adjusting first during calibration.

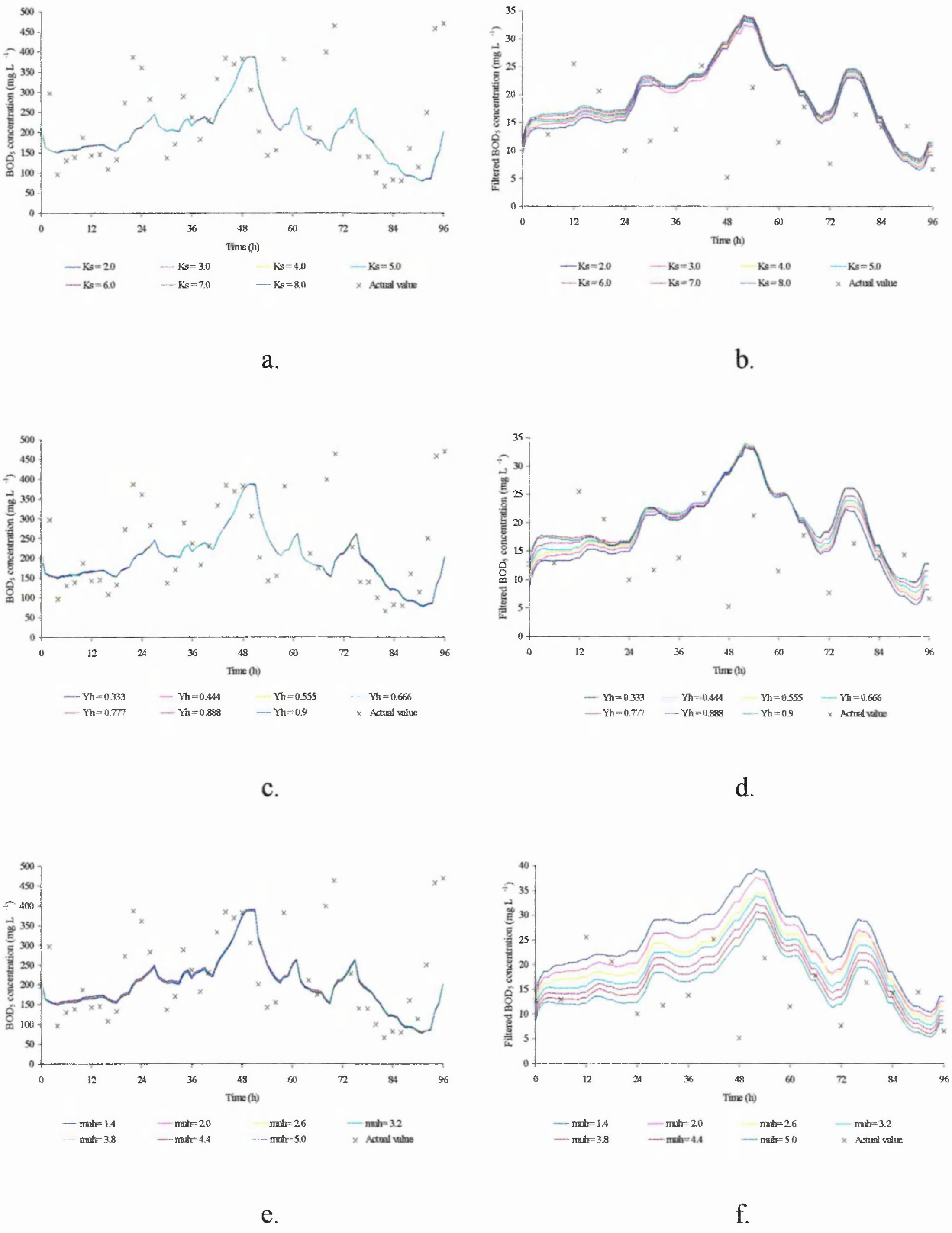


Figure 10. Variation of modelled effluent BOD_5 and filtered BOD_5 with K_s, Y_h and $\mu_{max\ h}$.

(a. and b. K_s ; c. and d. Y_h ; e. and f. $\mu_{max\ h}$)

The modelled effluent ammonia did not follow the same shape as the actual data although the cumulative effluent ammonia mass predicted was similar to the actual data (figure 11). This implies that the hydraulics in the model have not been defined correctly. A completely mixed reactor was assumed but the difference between the predicted and actual data suggest that the hydraulics should be more plug flow and defined by a number of completely stirred tank reactors (CSTRs) in series. This could be confirmed using tracer tests.

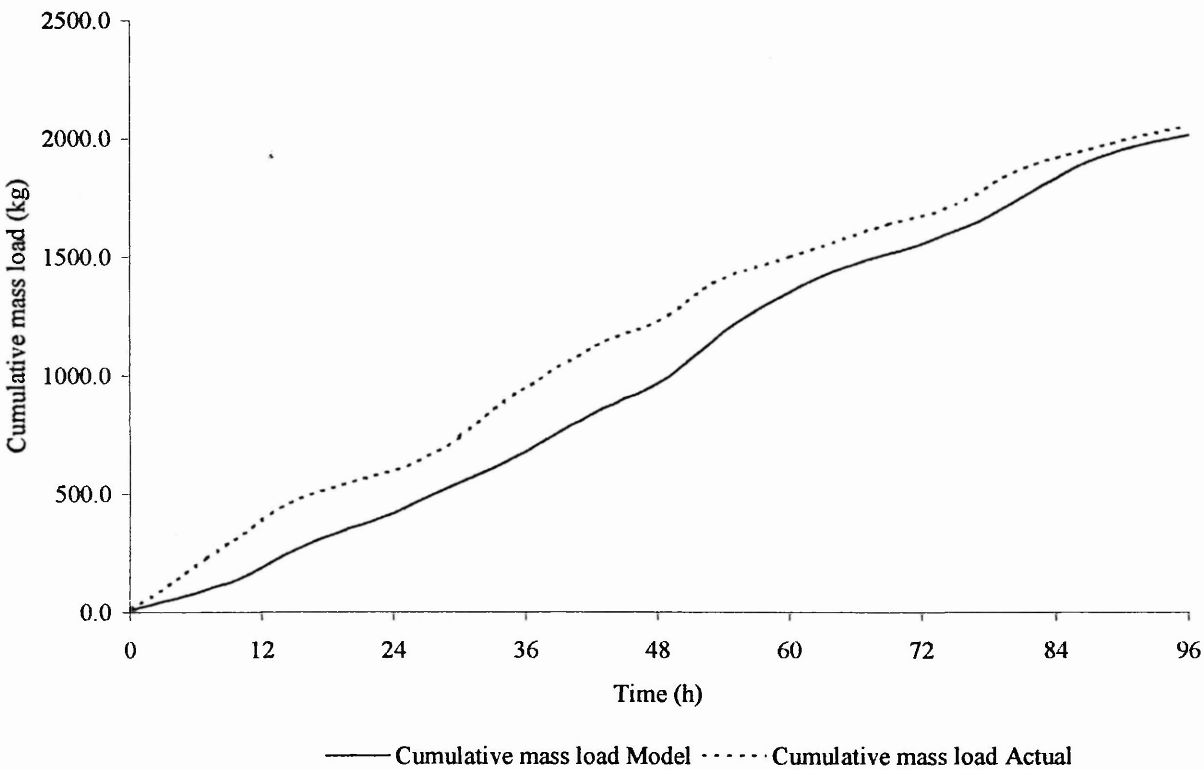


Figure 11. Cumulative mass load of ammonia in the effluent.

The simulations were repeated with different numbers of CSTRs in series (figure 12). It can be seen that use of more CSTRs resulted in the model prediction showing more response to influent variations but it still did not quite match the shape of the actual data. Some difficulties were encountered with the model in that it could not be resolved to steady state with any more than 3 CSTRs in series.

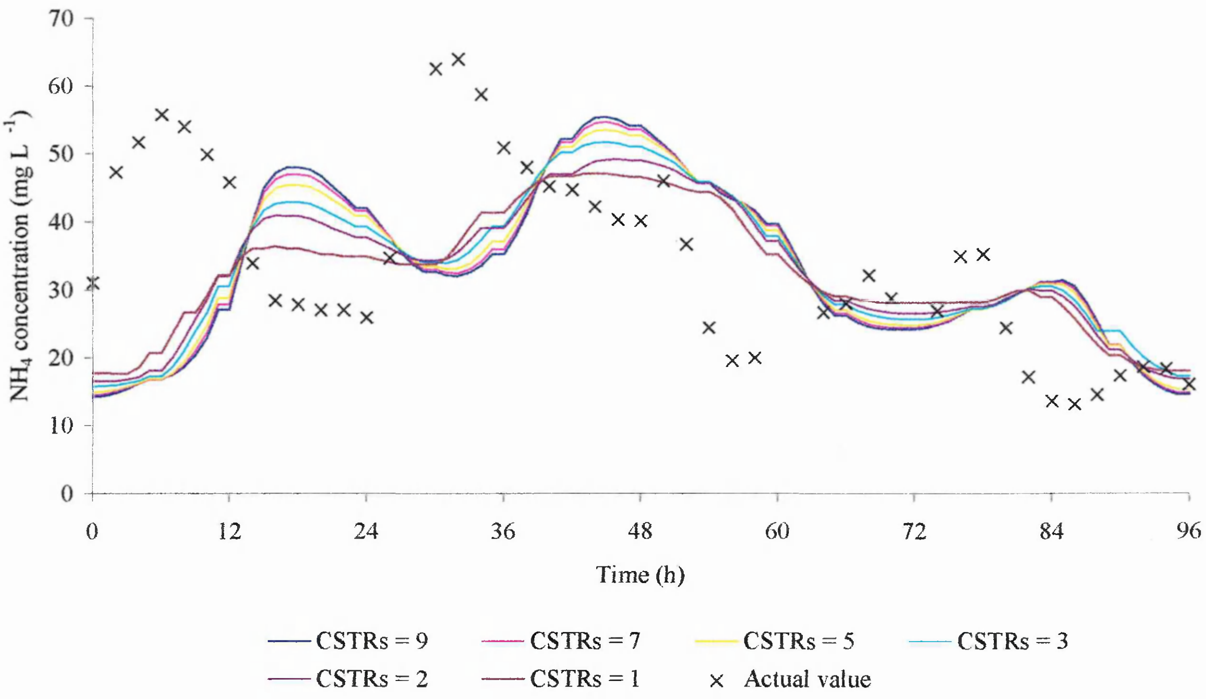


Figure 12. Variation of modelled effluent ammonia concentration with number of CSTRs in series.

A further attempt to improve the effluent ammonia prediction was carried out by bypassing some of the flow past the treatment process to try to account for any short circuiting (figure 13). Again, although a faster response was obtained the model still did not match the actual ammonia data.

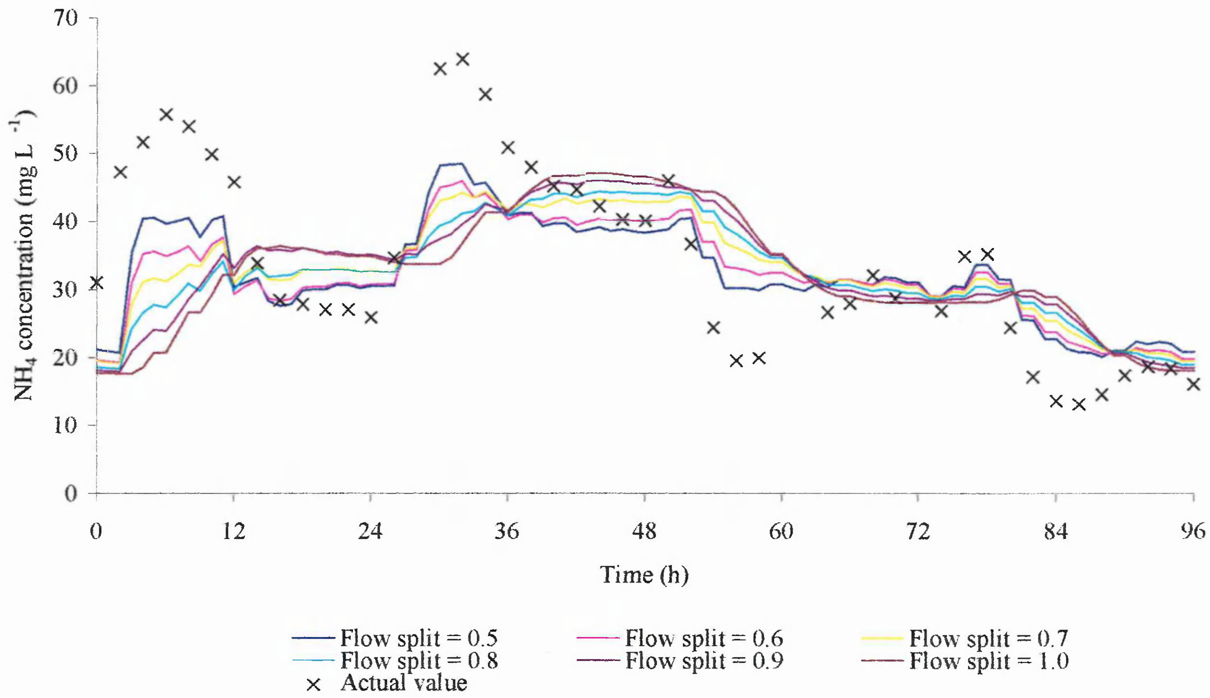


Figure 13. Variation of modelled effluent ammonia concentration with percentage flow split.

Given the presence of anoxic zones in the full scale upgrade and the recycle of mixed liquor to them, it was surprising that the model predicted an average final effluent TON concentration of 26 mg l^{-1}. Factors affecting denitrification include the C/N ratio. Henze *et al.* (1997) give optimum C/N ratios for denitrification as 3-3.5 kg BOD kg^{-1} N and 4-5 kg COD kg^{-1} N. During the intensive sampling programme the measured ratios were 2.7 kg BOD kg^{-1} N and 8 kg COD kg^{-1} N in the settled sewage but the high removal of soluble BOD and COD in the MBBR could make carbon limiting for denitrification. Henze *et al.* (1997) gave a maximum denitrification rate of approximately 1 gNO_3-N kg^{-1} VSS h^{-1} at 15°C using carbon from the influent wastewater and Metcalf & Eddy (1991) quote a range of 3 - 5 g NO_3-N kg^{-1} VSS h^{-1} at 20°C for wastewaters where carbon is not limiting. Given a total anoxic volume of 508 m^3, a MLSS of 2800 mg l^{-1} and a VSS to MLSS ratio of 0.87, the total VSS in the anoxic zones is 1237 kg. Hence the maximum denitrification expected would be between 1.2 and 6.2 kg h^{-1}. The recycled NO_3-N is approximately 40 kg h^{-1} so only 3 - 16 % denitrification would be expected.

4.5 CONCLUSIONS

- Steady state calibration was achieved with predicted effluent quality within 20 % of the actual data.

- The dynamically calibrated model gave a good prediction of effluent quality during validation, predicting all of the trends but not meeting all of the extreme high and low effluent concentrations.

- Application of the calibrated model indicates that retrofitting a MBBR/AS hybrid process would enable the new ammonia consent at Colchester to be achieved.

- Influent wastewater characterisation, hydraulic characterisation and plant operation were found to be more critical than the kinetic parameters to the model predictions.

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CHAPTER 5 - MODELLING OF MOVING BED BIOFILM REACTOR PILOT PLANT

5.1 INTRODUCTION

The moving bed biofilm reactor (MBBR) process is a compact low head loss biofilm process that has been used alone and in conjunction with activated sludge plants (ASPs) to upgrade existing wastewater treatment plants to achieve nitrification

Process models are increasingly being used during the design process to give optimised designs both in terms of capital and operational expenses and savings through the use of dynamic modelling have been reported.

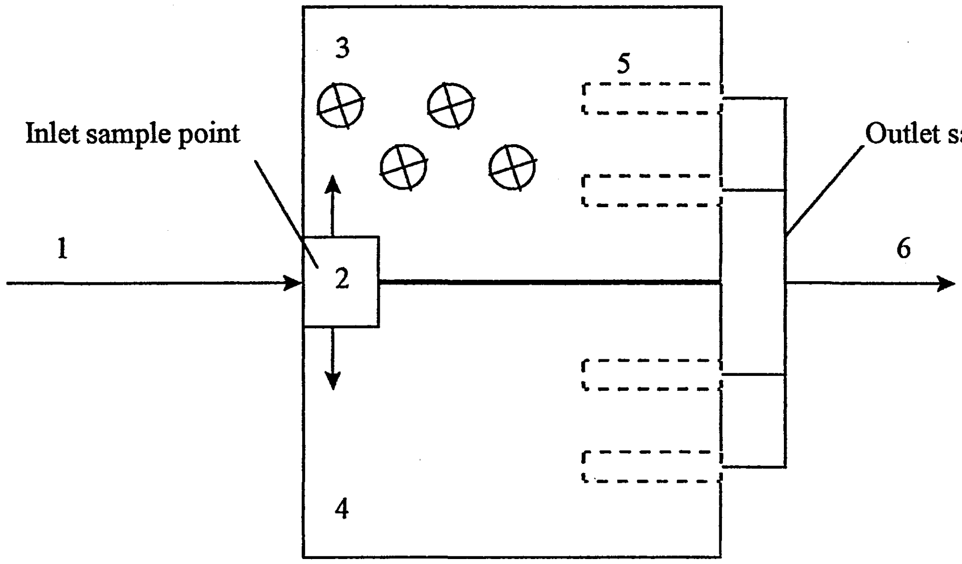
In order to fully assess the potential of the MBBR process, a model has been developed by Hydromantis in conjunction with Anglian Water. Initial attempts at calibrating the model for an MBBR/AS hybrid plant were limited due to errors in the wastewater analysis and the way in which the model influent COD fractions were calculated. This meant that although the model was calibrated to predict actual effluent data, the confidence in the model was not very high.

In order to gain more confidence in the model it was calibrated for a pilot MBBR. For this study the influent COD fractions and the effluent readily biodegradable substrate were estimated directly. This approach alleviated some of the problems associated with COD fractionation based on BOD measurement that were experienced in previous work.

5.2 MATERIALS AND METHODS

5.2.1 Moving bed biofilm reactor pilot plant

The pilot plant had two zones (figure 1), each with a volume of 27 m^3 and a 50 % fill of biofilm support media. Two different media were used, on in each reactor. The first was Kaldnes media (Kaldnes Miljoteknologi AS, Norway) which were cylinders 9 mm wide and 7 mm long with two cross pieces on the inside and longitudinal fins on the outside. They were made from high density polyethylene and had a specific density between 0.92 and 0.95 g/cm^3 and an effective specific surface area of 500 m^2 m^{-3} of media (K1 media). The second media was also a Kaldnes media but with an effective specific surface area of 310 m^2 m^{-3} (K2 media).



1 Settled sewage from STW; 2 Flow splitting chamber; 3 K1 reactor; 4 K2 reactor; 5 Outlet sieves; 6 Treated effluent returned to STW.

Figure 1. Pilot plant schematic.

Each reactor received the same influent, settled sewage, which was pumped at a constant flow rate of 15 l s^{-1} to a distribution chamber on the plant. The flow was split equally between the two reactors using V notch weirs. Allowing for the displacement of water by the media, 0.18 m^3 m^{-3} media, this gives a retention time in the reactor of 49 minutes, compared to a design retention of 48 minutes in the Colchester plant discussed in previous chapters.

Coarse bubble aeration was provided at a constant air flow rate to each reactor of 115 m^{3}h^{-1}. Clean water aeration tests carried out previously had shown the standard oxygen transfer efficiency (SOTE) to be 17 % for the K1 media and 7 % for the K2 media. Pipe sieves were used to retain the media in the reactors.

5.2.2 Sampling procedure

Grab samples were taken from the influent distribution chamber and the effluent weir of each reactor three times per week over a period of 5 weeks in order to estimate the COD fractions required for the model input. For calibration purposes two hourly time composite samples were collected from the same sampling points using autosamplers and sent to the Anglian Water laboratory at Whitlingham for analysis.

Samples of the media were taken from each reactor so that the total suspended solids of the biofilm could be measured.

The dissolved oxygen (DO) concentration and temperature in each reactor were recorded using portable DO probes (Orbisphere, Chesterfield, UK) and a data logger (Grant Instruments, Cambridge, UK).

5.2.3 Sample analysis

Grab samples were analysed for readily biodegradable COD, soluble inert COD, particulate inert COD, total COD, filtered COD, BOD_5, filtered BOD_5, total suspended solids, ammonia, total nitrogen and filtered total nitrogen.

Readily biodegradable substrate, S_S, was estimated using the procedure outlined by Mamais *et al.* (1993). The particulate and soluble inert fractions, X_I and S_I, were estimated using the methods proposed by Lesouef *et al.* (1992) which enabled the slowly biodegradable substrate, X_S, (incorporating any biomass) to be calculated by subtraction from the total COD. All COD measurement in these procedures and for the other grab sample analysis was carried out using the Hach test-n-tube method (Camlab, Cambridge, UK). Analysis for ammonia and total nitrogen was carried out using Dr Lange test kits. All other analysis was carried out using standard methods (APHA, 1992).

The two hourly composite samples were all analysed at the laboratory for COD, TSS and ammonia using standard methods.

To measure the dry solids content of the biofilm a method similar to that used by Nogueira *et al.* (1998), 10 elements of the carrier media were dried in an oven and weighed periodically until no further change in the mass of the carrier was observed. The biomass was then removed from the carrier elements by soaking them in an acid wash (5 % HCl solution). An ultrasonic bath was used to aid removal of the biomass. Once all of the biomass had been removed the carrier elements were dried and weighed again. The dry mass of the biomass was then calculated as:

Biomass = mass (carrier + biomass) - mass carrier Equation 1

Knowing the number of carriers per unit volume and the percentage reactor fill then allowed the biomass per m^3 of reactor to be calculated.

Biomass / m^3 reactor = (biomass / carrier) x (carriers / m^3) x % fill Equation 2

5.2.4 Data analysis

The 50 percentile values of each of the wastewater characteristics were calculated and used for calculating the influent and effluent values for the steady state calibration. One day's data was selected for the dynamic calibration, the data chosen being representative of typical conditions at the works and as complete as possible. Model validation was carried out against the complete 4 day data set without altering any parameters after the dynamic calibration.

5.2.5 Tracer test

During previous work modelling an MBBR, the model output corresponded to completely mixed reactor smoothing out the peaks observed in the influent. To check that the pilot reactor was a complete mix reactor a tracer test was carried out. Lithium chloride was used as the tracer, 1930 ml of 35 % w/w lithium chloride solution being added to the influent chamber as a pulse input. An initial effluent sample was taken before addition of the tracer. Further samples were then taken every 30 s for the first 30 minutes, every minute for the next 30 minutes and then every 5 minutes for a further hour after addition of the lithium chloride. The samples were filtered and acidified with 1M nitric acid. The samples were then analysed for lithium using an ICP-AES

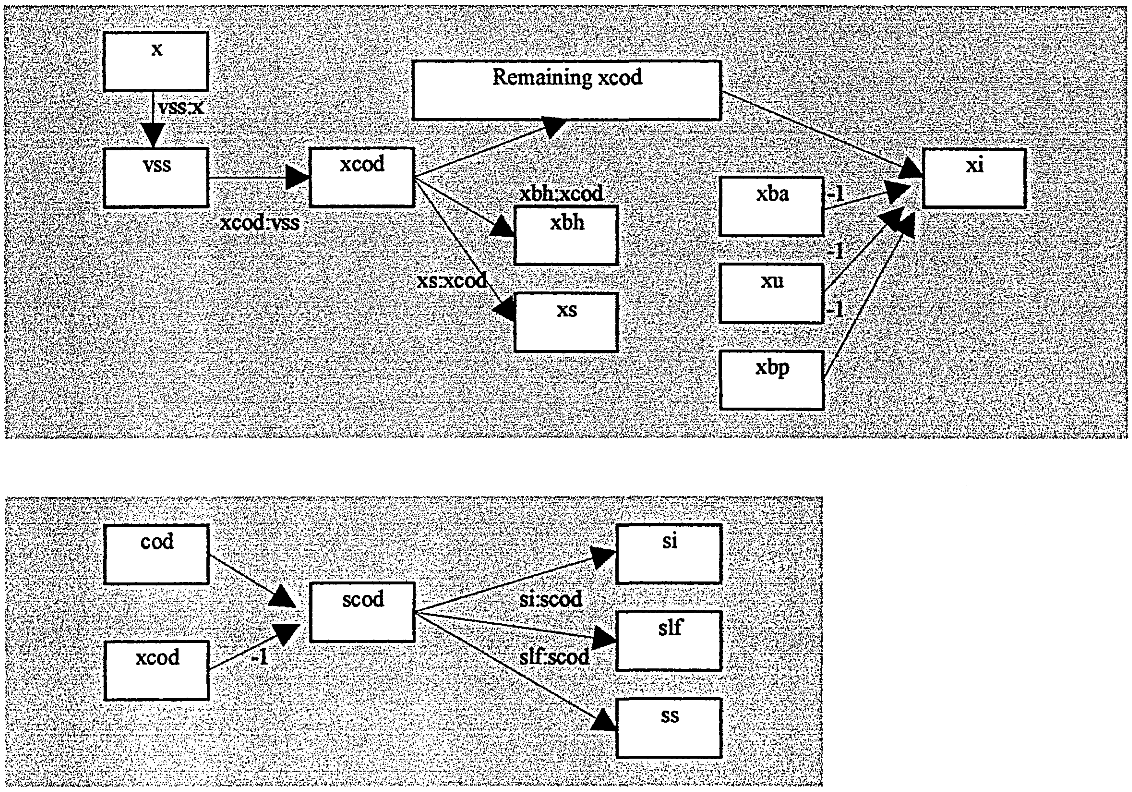
(Inductively coupled plasma - atomic emission spectrophotometer, Atom scan 16, Thermos Jarrell Ash, Sci-Tek, Bucks). The residence time distribution was plotted and compared to the tanks in series model as described by Levenspiel (1972) and Clark (1996) to confirm whether the tank was completely mixed as it was designed to be.

5.2.6 Model plant layout

The reactor containing the 500 m^2 m^{-3} media was modelled with a layout which consisted of the influent, a two way flow split, the MBBR and a two way flow mixer. The flow split means that a user defined proportion of the influent can bypass the MBBR, allowing for any short circuiting which may occur. The MBBR was modelled as 1 zone, assuming a completely mixed reactor, a 50 % media fill of the relevant media and diffused aeration with the air flow rates and SOTE as described..

5.2.7 Influent model

The ASM 2 influent model in GPS-X was selected. This model takes measured COD, TSS and TKN as the influent drivers and uses a number of stoichiometric ratios and state variables to calculate the COD fractions required for the model as shown in figure 2 (Hydromantis, 1997). The COD fractions measured were used to derive the stoichiometric ratios.



where	x	total suspended solids	vss	volatile suspended solids
	xcod	particulate COD	xbh	heterotrophic biomass
	xs	slowly biodegradable substrate	xba	autotrophic biomass
	xbp	phosphorus accumulating biomass	xi	particulate inert COD
	xu	unbiodegradable decay products	scod	soluble COD
	si	soluble inert COD	slf	volatile fatty acids
	ss	readily biodegradable substrate		

Figure 2. COD fractions in the GPS-X ASM 2 influent model (Hydromantis, 1997).

It was assumed that any biomass in the influent is incorporated into the term for slowly biodegradable substrate, hence the ratio *xbh:xcod* and the terms *xba, xbp* and *xu* were all set to zero. The fraction *si* was calculated but the MBBR model in GPS-X does not have a term for *si*. To allow for this the value of si was subtracted from the influent COD and the value in the model set to 0. Because the MBBR model is set up in a P removal library the influent has a term for VFAs. In this case it was assumed that any VFAs present were part of the readily biodegradable substrate and so the term slf was set to 0. The measured ratio of VSS:TSS was used but the measured particulate COD (total COD - filtered COD) was not used to calculate the ratio *xcod:vss*. The term *xcod* is split to give the slowly biodegradable substrate and the particulate inerts. As the slowly biodegradable substrate also includes some of the measured filtered COD, *xcod* was calculated from the measured COD fractions as $X_I + X_S$ and this figure was used to calculate *xcod:vss*.

5.2.8 Model calibration

The first stage of calibration involved the use of 50 percentile values of the wastewater characteristics to calibrate the model in steady state conditions. Having obtained a good fit between the model and the measured data (within 20 %), the model was re-calibrated using the diurnal data from one day of the sampling program. The data was replicated 4 times so that a 5 day simulation could be run, allowing the model to settle to a repeated output. The kinetic parameters and flow split were then adjusted again to fit the model output to the actual dynamic data. Model fitting was carried out by visually comparing the model prediction with the actual data. Once a good visual fit had been obtained the model was validated using the full set of sampled data. No parameters were changed during validation.

5.3 RESULTS

5.3.1 COD fractions

The COD fractions found using the methods of Mamais *et al.* (1993)and Lesouef *et al.* (1992) are compared to literature values in table 1. The estimated fractions are of a similar value to those previously reported. The slowly biodegradable substrate is at the high end of the range as it includes any biomass in the influent too.

Table 1. Comparison of estimated COD fractions with literature values.

	COD fractions as a percentage of influent total COD (%)			
	S_I	S_S	X_I	X_S
Influent	14	16	3	67
Effluent	19	4	...	...
Henze (1992)	3 - 10	16 - 33	4 - 13	40 - 60

5.3.2 Influent wastewater characteristics

The wastewater characteristics at the time of sampling for COD fractionation are summarised in table 2. There appears to be very little removal of filtered BOD_5 across the plant but this does not correspond with filtered COD removal which is 46 %. The BOD loading on the reactor was 1.8 kg BOD_5 m^{-3} d^{-1}, lower than the design load used in the previous work at Colchester which was up to 5 kg BOD_5 m^{-3} d^{-1}. No ammonia removal occurred in the MBBR, as expected.

Table 2. Average wastewater characteristics during COD fractionation.

	50 percentile values (mg l^{-1})		
	Influent	K1 effluent	K2 effluent
BOD_5	76	56	56
Filtered BOD_5	14	13	12
COD	222	197	206
Filtered COD	110	59	62
TSS	67	83	92
VSS	26	38	27
Total N	40	...	...
Filtered total N	36	...	...
NH_4-N	30	...	...

The average influent wastewater characteristics during the diurnal sampling period were 258 mg l^{-1} COD, 89 mg l^{-1} TSS and 30 mg l^{-1} ammonia. The wastewater temperature during the sampling period was fairly constant at 14 °C.

5.3.3 Tracer tests

During the tracer test only 72 % of the tracer was recovered, limiting the confidence in the results. The residence time distribution (RTD) of lithium in the MBBR shown in figure 3 fits best to the 1 CSTR model (figure 4) described by Levenspiel (1972) meaning the tank was completely mixed. However, the number of tanks when calculated was 1.6. The mean RTD calculated was 38 minutes against a theoretical retention time of 49 minutes. This implies that there is some short circuiting within the reactor and the RTD curve is similar to that for a completely mixed tank with short circuiting (Levenspiel, 1972).

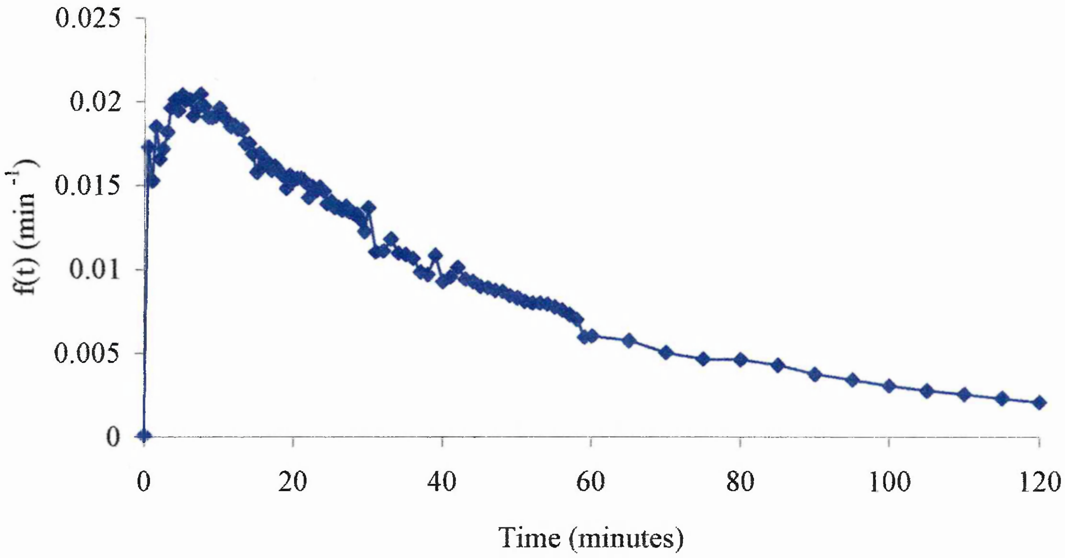


Figure 3. Residence time distribution of lithium tracer in MBBR.

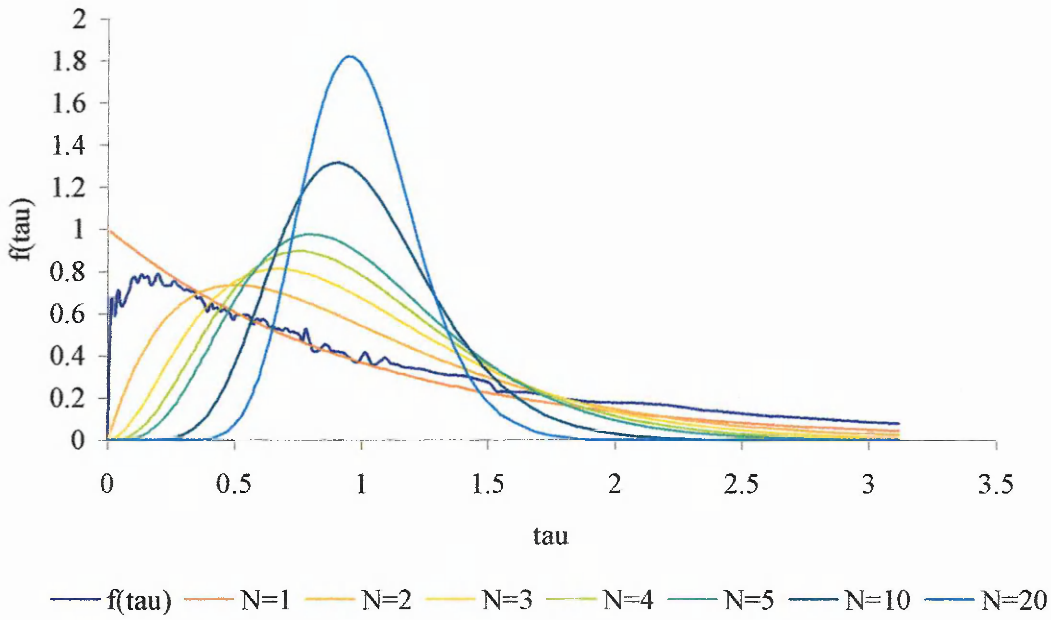


Figure 4. Comparison of dimensionless residence time distribution for the experimental data and the N tanks in series model.

5.3.4 Steady state calibration

The first steady state simulation was carried out using the influent ratios from table 2 and all of the default kinetic and stoichiometric parameters of the model. The molecular diffusion coefficients in the biofilm were set to 80 % of those in water. There is limited information on actual values of molecular diffusion coefficients in biofilms but Characklis and Marshall (1989) report 80 % as being an acceptable value. The value of μ_{max} for the heterotrophs was then adjusted until the model prediction was within 20 % of the measured data. Table 3 compares the modelled and measured data for steady state conditions for a μ_{max} of 8 d^{-1}.

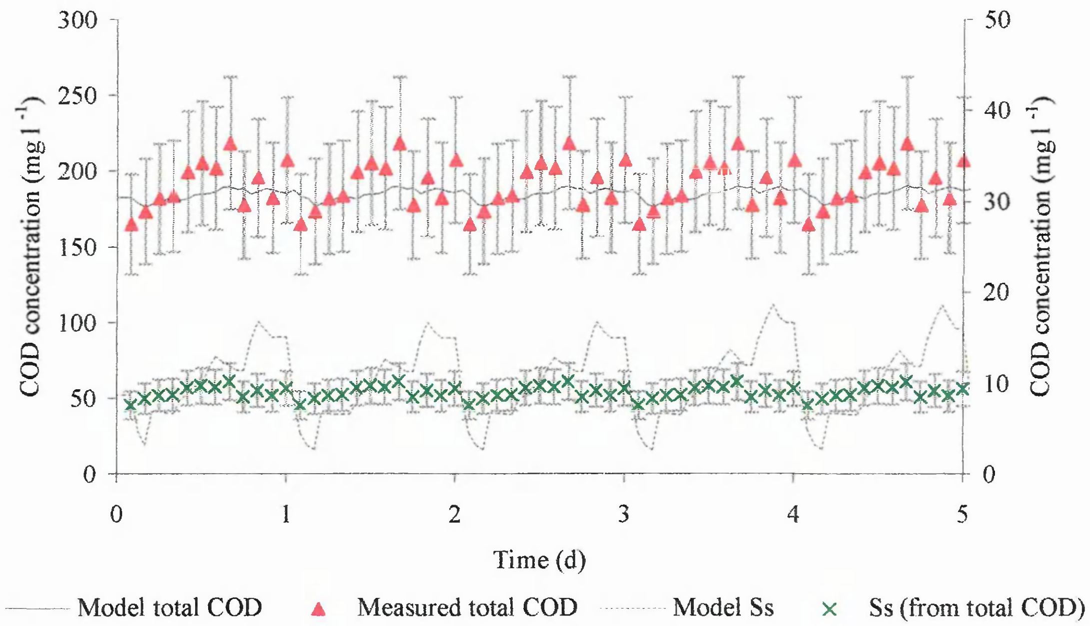
Table 3. Comparison of measured and modelled effluent quality for steady state calibration.

	Measured data ($mg\ l^{-1}$)	Model prediction ($mg\ l^{-1}$)	Difference (%)
TSS	108	111	3
COD ($-S_I$)	183	174	5
S_S	9	8	1
Ammonia	29	29	0
Biomass	5820	5624	3

5.3.5 Dynamic calibration

For dynamic calibration of the model, a 24 hour data set was selected which was representative of the works influent. The data set was then repeated for 5 days so that the model could be run until a repeating pattern was observed in the predicted effluent quality graphs. Steady state conditions were used as the initial conditions for dynamic simulation. The first dynamic simulation was carried out using the same kinetic and stoichiometric parameters as used for the steady state calibration. The results are shown in figure 5. The measured value for S_S is calculated as a percentage of the effluent total COD.

The model prediction does not follow all of the variations in the measured effluent quality. However the average values are similar and for the majority of the measured data points, the model prediction is within 20%. The model is behaving as a CSTR and therefore smoothes out the peaks and troughs. It appears that the actual plant does not behave in the same way which could be due to the short circuiting in the reactor discussed previously. The differences were considered acceptable and validation was carried out for the same parameter values.



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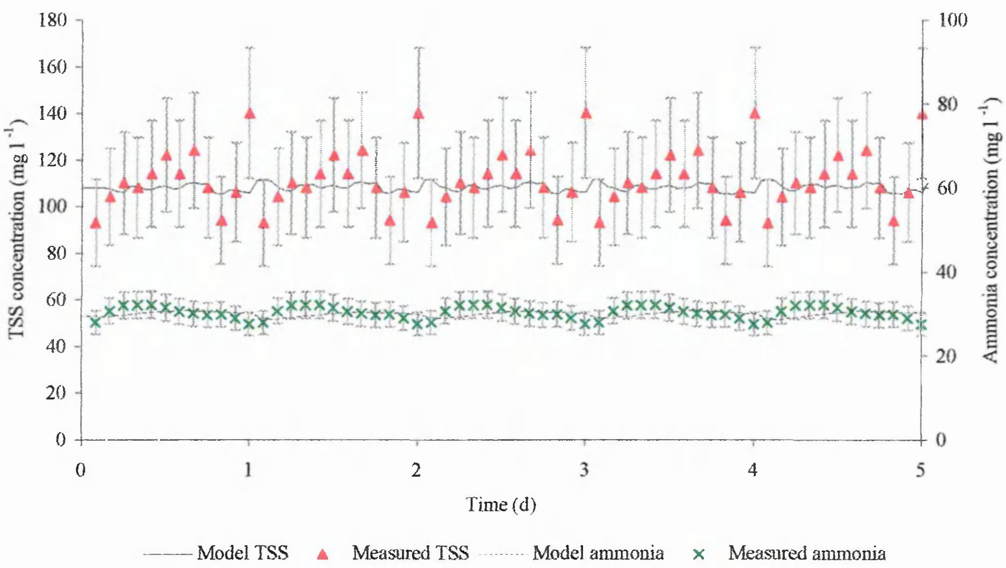

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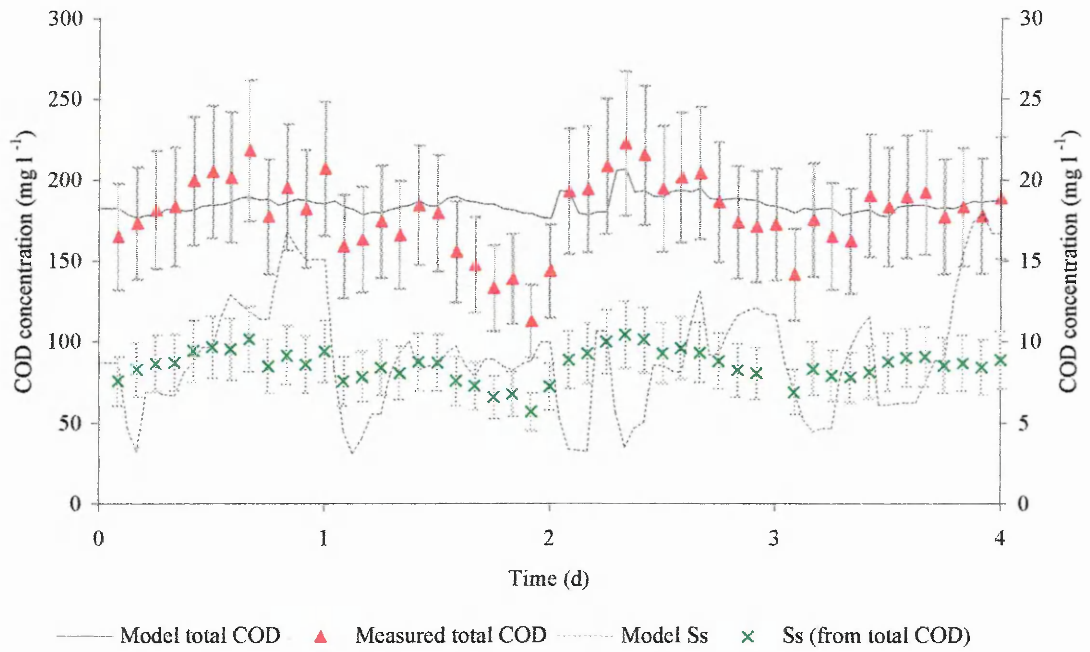
Figure 5. Comparison of measured and model data for dynamic calibration (20 % error bars shown), a) total COD and S_S, b) TSS and ammonia.

5.3.6 Model validation

The mean values of the model prediction and the measured data were compared (table 4) and found to be within 5 % of each other. The data in table 4 shows the model to be predicting the right magnitude of effluent quality. However, the model does not predict all of the diurnal variations observed in the measured data. From figure 6 it can be seen that the modelled COD does not follow the same downward slope as the measured data during the second day of sampling. In addition to this the model prediction for all parameters shows less variation than the measured data. This is probably due to the model being set up and behaving as an ideal completely mixed reactor, attenuating the peaks observed in the influent.

Table 4. Comparison of average measured and modelled effluent quality for dynamic validation.

	Measured data (mg l^{-1})	Model prediction (mg l^{-1})	Difference (%)
TSS	115	109	5
COD ($-S_I$)	179	185	3
S_S	9	9	0
Ammonia	29	29	0



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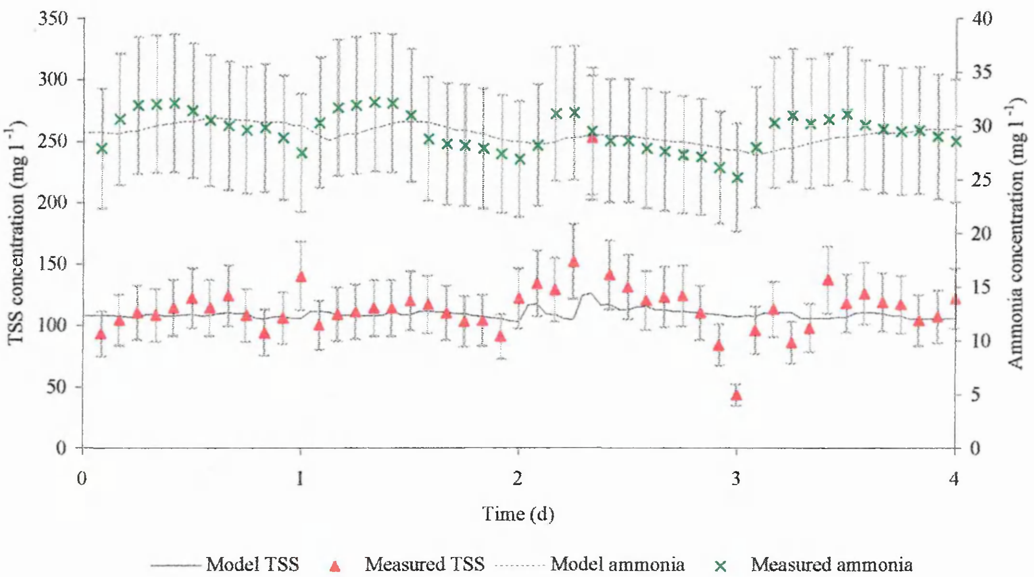

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Figure 6. Comparison of measured and model data for dynamic validation (20% error bars shown), a) total COD and S_S, b) TSS and ammonia.

To investigate why the model did not predict the large decrease in observed total COD during the second day of sampling the measured influent and effluent results were compared. The model responds only to the data it is given and as there is no large drop in influent COD concentration it is not unexpected that the model does not predict a drop in the effluent COD (figure 7).

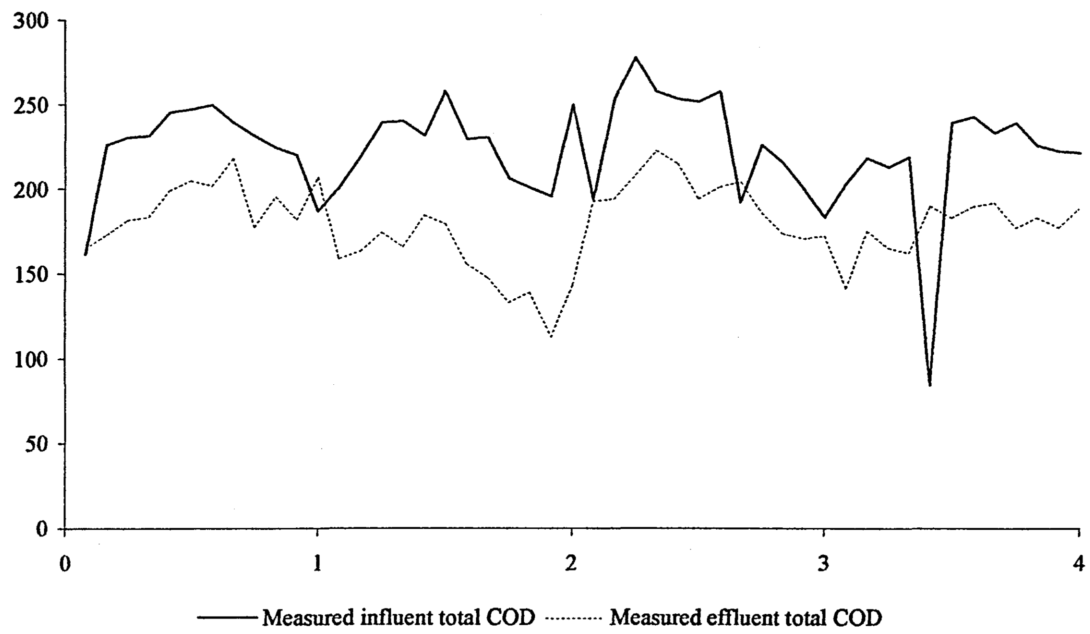

Figure 7. Variation of influent and effluent total COD during the diurnal sampling.

A further factor to consider is that although the influent total COD did vary over the sampling period the magnitude of the variation was not significant with most of the samples being within 20 % of the average.

The parameters that were changed during model calibration are summarised in table 5.

Table 5. Parameters changed from the default during model calibration.

Parameter	Default	Value used	Literature
$\mu_{max,H}$ (d^{-1})	3.2	8	4 - 8 (Henze *et al.*, 1997)
Reduction of diffusion in biofilm	0.5	0.8	0.8 (Characklis and Marshall,1989)

5.4 DISCUSSION

The procedures used to estimate the COD fractions during this work yielded values that were in line with values reported in literature (table 1). The flocculated COD method was relatively straightforward and quick to perform giving an estimate of S_S and S_I similar to the values reported by Henze (1992). The method for estimating X_S and X_I was more time consuming to carry out with the samples having to be aerated for up to 2 weeks. Although only COD measurements are required, 8 measurements were taken for each sample which could make the analysis very expensive if the fractionation is carried out on a significant number of samples.

For the average COD and TSS values stoichiometric ratios were selected that matched the influent model COD fractions to the measured average fractions using the ASM2 input model in GPS-X. This enabled the model input to be defined in terms of COD, TSS and TKN only. Generally it is assumed that the COD fractions remain constant relative to the influent total COD. However, using an influent model such as the ASM2 influent model yields gives influent COD fractions that vary as the ratio between influent COD and TSS vary. In the case of the model used for this work, the influent model was really only valid for (total COD - S_I)>2.02 x TSS. If the influent COD (less the soluble inert fraction for the model) became less than 2.02 x TSS the model influent S_S became less than zero which is not possible.

The main parameter change during calibration was the maximum specific growth rate for the heterotrophs, this is in line with the recommendations of the authors of the IAWQ activated sludge model no.2 (Henze *et al.*, 1997). The default value of μ_{max} for the heterotrophic bacteria was 3.2 d^{-1} so the use of 8 d^{-1} represented a significant change. It was however still in the ranges quoted in literature for μ_{max}, with Henze *et al.* (1995) quoting a typical figure of 6 d^{-1} and Henze *et al.* (1997) a range of 4-8 d^{-1}. As no nitrification was occurring the autotrophic parameters were left unchanged.

The model predicted the biomass in the reactor to within 3 % using the default parameters for the biofilm composition. These were not varied as parameters such as the biofilm density and water content were not measured. The default values for these parameters were 1.02 g cm^{-3} for the density and 0.1 for the dry content of the biofilm, giving a density of 102 kg (dry mass) m^{-3} (wet volume). Characklis and Marshall (1989) reported a range of 10 - 105 kg m^{-3} for a number of different types of biofilm including heterotrophic mixed population biofilms. The maximum biofilm thickness during modelling was set to 0.5 mm which is within the range of 0 - 1300 μm reported by Characklis and Marshall (1989).

5.5 CONCLUSIONS

The methods described by Mamais *et al.*(1993) and Leseouf *et al.* (1992) are simple methods for estimating COD fractions for use in modelling. Use of these methods was a better approach for defining the model influent than the use of the BOD based model in GPS-X had been found to over estimate X_I and under estimate S_S in previous work.

The COD fractions in the settled sewage at Cambridge STW were similar to reported values from other sites.

The GPS-X model of the MBBR was found to predict the average performance of the pilot plant to within 5 % in terms of total COD, S_S, TSS, ammonia and biomass solids. To achieve this level of fit μ_{max} was set to 8 d^{-1} and the reduction of molecular diffusion coefficients in the biofilm to 80% of those in water in line with reported values.

The model prediction was found not to show the same variation with time as the measured data. This was attributed to the reactor hydraulics and errors in the analysis.

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CHAPTER 6 - BUSINESS IMPLICATIONS

6.1 INTRODUCTION

The water industry in England and Wales has been through considerable change during the last 30 years, with the industry being completely restructured in 1973 and then changing from public to private ownership in 1989. The industry and the companies within it continue to develop, adapting to the new structures.

The aim of the work was to develop the business implications for the technical work carried out. This was achieved by analysing the market (environment) using tools such as PEST analysis and Porter's five forces model to find out the important issues in the industry. Once these environmental issues had been identified the strategies available to a firm in order to compete in the industry were developed. Ways in which the technical research work carried out can be applied to give the firm some competitive advantage and help to implement its competitive strategy were then outlined.

6.1.1 The Water Industry in England and Wales since 1973

Prior to 1973 there were approximately 1 400 sewerage operators, 187 water undertakers and 29 river boards. The Water Act of 1973 led to the creation of 10 Regional Water Authorities, based on river basins, that were responsible for the whole water cycle, including abstraction, drinking water treatment and supply, collection and treatment of waste waters, discharge to the environment and environmental water quality control. In addition to the 10 Water Authorities, 29 statutory water companies were allowed to remain, providing drinking water treatment and supply.

The restructuring was carried out with the intention of co-ordinating the use of water resources and hence improving efficiency and allowing economies of scale to be achieved (Parker and Sewell, 1988). An example of this was in the Black Country where 12 of 14 old overloaded sewage treatment works were closed down (Hassan *et al.*, 1996). Efficiency gains were increased further by the implementation improved organisational structures by the regional water authorities.

In 1989, as part of the Conservative Government's utility privatisation programme, the 10 Regional Water Authorities were privatised as water and sewerage companies (WSCs), with uncontested private monopolies, as part of the 1989 Water Act which also led to the creation of financial and environmental regulators for the industry. The 29 statutory water only companies (WOCs) became public limited companies in 1989 since when several have been amalgamated or taken over such that in 1998 only 17 remain (Water UK, 1998).

In return for the regional monopoly conditions, the water companies had to accept outside regulation from public watchdogs with powers to intervene in the interests of competition and the consumer (The Office of Water Services, Ofwat) and the environment (The National Rivers Authority, NRA). The settlement that was imposed included (Hassan *et al.*, 1996):

- The water companies can only raise prices by the retail price index plus a factor, K, the K factor to set by the regulator.
- The water industry's £ 5 billion debt was wiped out.
- There was a payment from the government to the water industry of £ 1.1 billion.
- A 10 year capital spending programme (Asset Management Programme (AMP)) was outlined for each of the companies (now reduced to a 5 year programme).

The 10 privatised water companies comprise of two parts, the core business and enterprise businesses. The core business is the water utility part and is regulated by Ofwat with strict controls on the level of service provided and the prices which can be charged to customers, taking into account the required investment in capital schemes. The enterprise businesses are outside of regulatory control.

6.2 BUSINESS ENVIRONMENT

The regulated core business of the 10 water and sewerage companies is to provide drinking water to customers in their region and to collect and treat wastewater from domestic and commercial properties prior to discharge to the environment. The 17 remaining water only companies have a role to provide potable water to their customers. In 1997/8 the total turnover of water and sewerage service provision in England and Wales was £6 460.93 million.

6.2.1 Water Supply

Water supply in England and Wales had a turnover of £2 996.08 million in 1997/8 (Water UK, 1998) earned by the 10 WSCs and 17 WOCs. Although turnover has increased over the last 5 years it can be seen from table 1 that the number of connected properties has increased very little since 1994/5. Figure 1 shows that the volume of water supplied has not increased significantly since 1973 - a trend that is likely to continue due to increased environmental awareness leading to less water consumption per capita. Industrial water users also seek to reduce water consumption as it can constitute a major cost to firms, a further factor that will limit the volume of water supplied in future. Given that the volume of water supplied hasn't increased significantly, the increased turnover is due to price increases.

Drinking water quality is regulated by the Drinking Water Inspectorate (DWI) which enforces that the companies comply with quality standards. Revisions to the Drinking Water Directive will require investment of more than £3 billion in order to meet new standards including reduced concentrations of lead, trihalomethanes, bromates and trihaloethanes in drinking water. The proposed Water Framework Directive will also have major implications for the water industry in the UK (Smith, 1997).

Table 1. Number of water and sewerage connections for the 10 water and sewerage companies in England and Wales.

	Resident connected populations (000s)							
	Water services				Sewerage services			
Company	94/95	95/96	96/97	97/98	94/95	95/96	96/97	97/98
Anglian	3 976	3 983	3 990	4 022	5 264	5 280	5 282	5 430
Dwr Cymru	2 778	2 786	2 788	2 818	2 932	2 940	2 943	2 952
Northumbrian	1 187	2 523	2 527	2 531	2 554	2 580	2 580	2 584
North West	6 805	6 843	6 856	6 866	6 756	6 791	6 791	6 782
Severn Trent	7 224	7 250	7 280	7 305	8 262	8 280	8 300	8 200
Southern	2 180	2 190	2 200	2 216	4 140	4 161	3 920	3 946
South West	1 493	1 506	1 516	1 506	1 362	1 384	1 394	1 402
Thames	7 329	7 358	7 381	7 589	11 565	11 600	11 635	11 937
Wessex	1 136	1 149	1 162	1 162	2 356	2 380	2 397	2 399
Yorkshire	4 407	4 451	4 463	4 508	4 805	4 805	4 738	4 738
Total	38 515	40 039	40 162	40 522	49 939	50 201	49 980	50 371

(Data from WSA, 1995; WSA, 1996; WSA, 1997; Water UK, 1998)

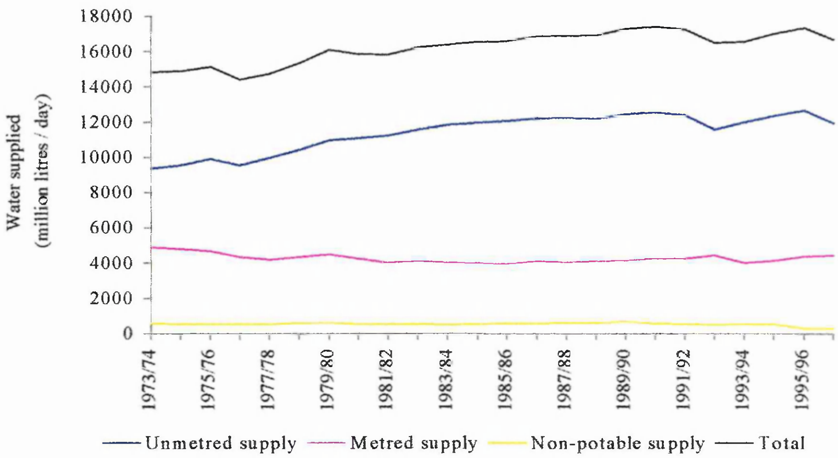

Figure 1. Water supplied in England and Wales 1973/74 to 1996/97 (WSA, 1997).

6.2.2 Sewerage Services

In England and Wales sewage collection and treatment is carried out by the 10 WSCs. The income to the 10 companies from sewerage services was £3 808.59 million in 1997/8 (Water UK, 1998). The overall turnover has increased as shown in figure 2 since 1988. Again the increased revenue is due to increases in price as the connected population has not increased greatly in the last five years (table 1). After a period of under investment in the water industry during the early 1980s the industry has had to invest heavily in new treatment plants in order to meet the tighter environmental standards now set by the EC.

With the increasing environmental awareness of recent years anti-pollution legislation has been getting increasingly stricter. Many of the environmental standards are now being set by the European Commission. It is then up to the UK government to put the standards into law, after which the water companies must comply with them. The control on discharges is only likely to get stricter with the water companies having to improve treatment facilities to meet ever tightening consents. The required process improvements will need a great deal of investment, approximately £30 billion at 1992 prices (Ernst, 1994; Hassan *et al.*, 1996) in the UK to meet the requirements of the EC Urban Waste Water Treatment Directive (UWWTD) (EC, 1991).

6.2.3 Trade Effluent

Industrial water users can choose to treat their own effluents to the required standard for discharge to the environment or discharge the effluents either untreated or partially treated to the sewers of their regional sewerage service provider. The WSC owning the sewer will charge the industrial discharger according to a published formula, the Mogden Formula. The charges take into account the strength of the effluent relative to the average sewage strength in the region and the level of treatment provided. Trade effluent charges provided a total revenue of £189.45 million to the WSCs in 1997/8 (Water UK, 1998).

6.2.4 Non Core Business

The water companies, or their parent companies, can carry out business that is not regulated . However non-regulated activities cannot be used to subsidise the regulated business and vice versa. The majority of non-regulated activity is carried out by the holding companies.

6.2.5 Water and Sewerage Service Providers in England and Wales

There are ten providers of sewerage services in England and Wales and a total of 27 water suppliers. The total revenue of the regulated industry has increased continuously since 1987, reaching £6 460.93 million in the year to 1998 (figure 2).

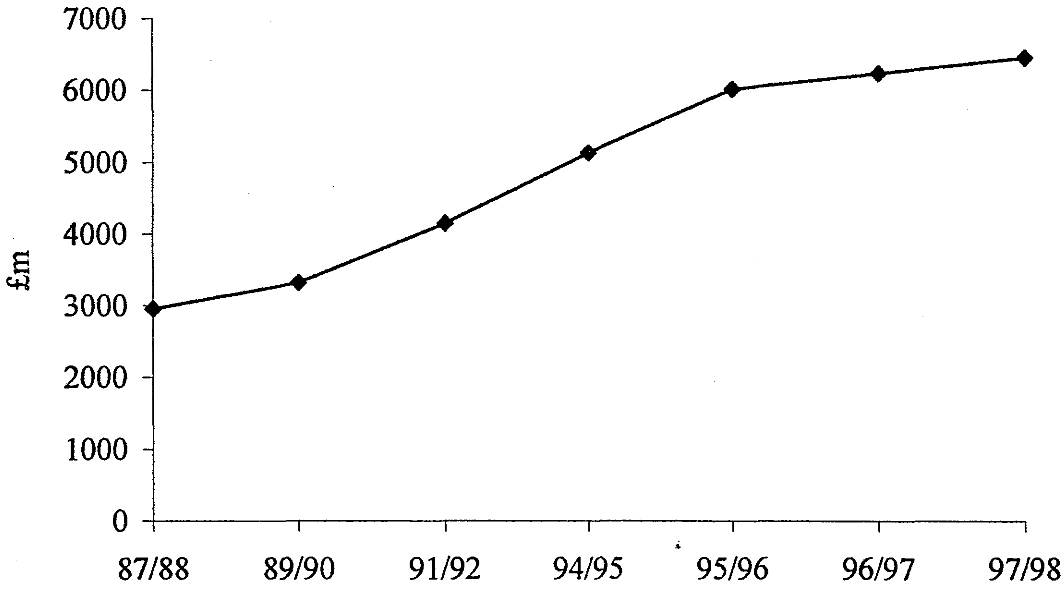

Figure 2. Total turnover of water companies regulated business in England and Wales.

During this period all of the companies in the industry have experienced increasing turnover. However, although the business for each of the companies has grown, it can be seen from table 2 that the individual market share of each of the companies has remained static. This is due to the regional monopoly conditions prevailing in the industry which restricts the opportunities for companies to expand geographically.

Table 2. Percentage share of the total market turnover of the ten water and sewerage companies in England and Wales.

	Market share (of turnover) of the water and sewerage companies						
	87/88	89/90	91/92	94/95	95/96[a]	96/97[a]	97/98[a]
Anglian	12.2%	12.1%	12.4%	12.3%	10.9%	11.0%	11.0%
Dwr Cymru	7.6%	7.7%	8.0%	8.1%	7.0%	6.8%	6.8%
Northumbrian	4.2%	4.5%	4.2%	4.3%	4.9%	4.9%	5.0%
North West	14.9%	15.4%	15.4%	15.3%	13.5%	13.4%	13.7%
Severn Trent	15.8%	16.4%	16.7%	16.5%	14.9%	14.7%	14.4%
Southern	7.2%	6.8%	6.6%	6.8%	6.0%	6.2%	6.4%
South West	3.5%	3.6%	3.8%	4.6%	3.9%	3.9%	3.8%
Thames	20.2%	18.4%	18.2%	17.9%	15.9%	15.6%	15.7%
Wessex	4.3%	4.4%	4.3%	4.5%	3.8%	3.9%	3.9%
Yorkshire	10.1%	10.7%	10.4%	9.8%	8.7%	8.7%	8.5%

[a] Figures for last three years include total turnover of WOCs.

(Data from WAA, 1988; WSA 1990, 1992, 1995, 1996 and 1997; Water UK, 1998)

Since the existing WSCs and WOCs are limited to a fixed share of the market they have to look outside of the regulated business in order to grow. This can be done through the WSCs and WOCs themselves or separately by the holding company of the WSCs or the parent company of the WOCs that are owned by other companies.

On a global scale there is considerable work to be done to provide drinking water and sewerage treatment systems. It has been estimated that 20% of the world population has no access to safe drinking water and around 50% has inadequate sanitation (WHO, 1999). World wide investment in water and sanitation projects has been estimated at £200 billion a year and use of private capital could become the largest source of funding for such projects (Smith, 1997). However, to compete for projects around the world will mean competing with very large international companies.

6.2.6 Environmental Factors

To predict how an industry is likely to change in the future it is important to consider what environmental factors are affecting the organisation and which of these are the most important at the present time and in the next few years (Johnson and Scholes, 1993). The environmental factors likely to affect the water industry in England and Wales have been considered below.

Political / legal

When the water industry was privatised the 10 water utilities were given a private monopoly within their region for drinking water supply and sewerage services. The Collins dictionary defines a monopoly as

> "**1.** exclusive control of the market supply of a product or service. **2.a.** an enterprise exercising this control. **b.** the product or service so controlled......"

In practice it is not often that a single firm has 100 % of the market share and for the purpose of investigation by the Monopolies and Mergers Commission a monopoly is defined as " a single firm or interrelated group of firms controlling 25% or more of the market" (Jeffreys, 1985).

"Neo-classical economics" is based on all decisions being motivated by a desire to maximise something. This includes individuals maximising the benefit obtained from a level of expenditure and firms maximising output from given levels of input and also profits. In competitive markets this leads to an optimum situation for society, i.e. maximum welfare from existing resources.

Given that a firm is assumed to seek maximum profits, where Profit = Revenue - Costs or

$$\Pi = R - C$$

where both R and C are related to the quantity produced per unit time, Q.

In order to decide the level of output at which profits are maximised two conditions are required.

1 $\quad \frac{d\Pi}{dQ} = 0 \quad$ and $\quad$ 2 $\quad \frac{d^2\Pi}{dQ^2} < 0$

$\therefore \quad \frac{dR}{dQ} - \frac{dC}{dQ} = 0 \qquad \frac{d^2R}{dQ^2} - \frac{d^2C}{dQ^2} < 0$

$\therefore \quad \frac{dR}{dQ} = \frac{dC}{dQ} \qquad \frac{d^2R}{dQ^2} < \frac{d^2C}{dQ^2}$

i.e. marginal revenue = marginal cost

Hence the firm would prefer to produce at the output, Q, where marginal revenue = marginal cost, i.e. at Q_1 in figure 3.

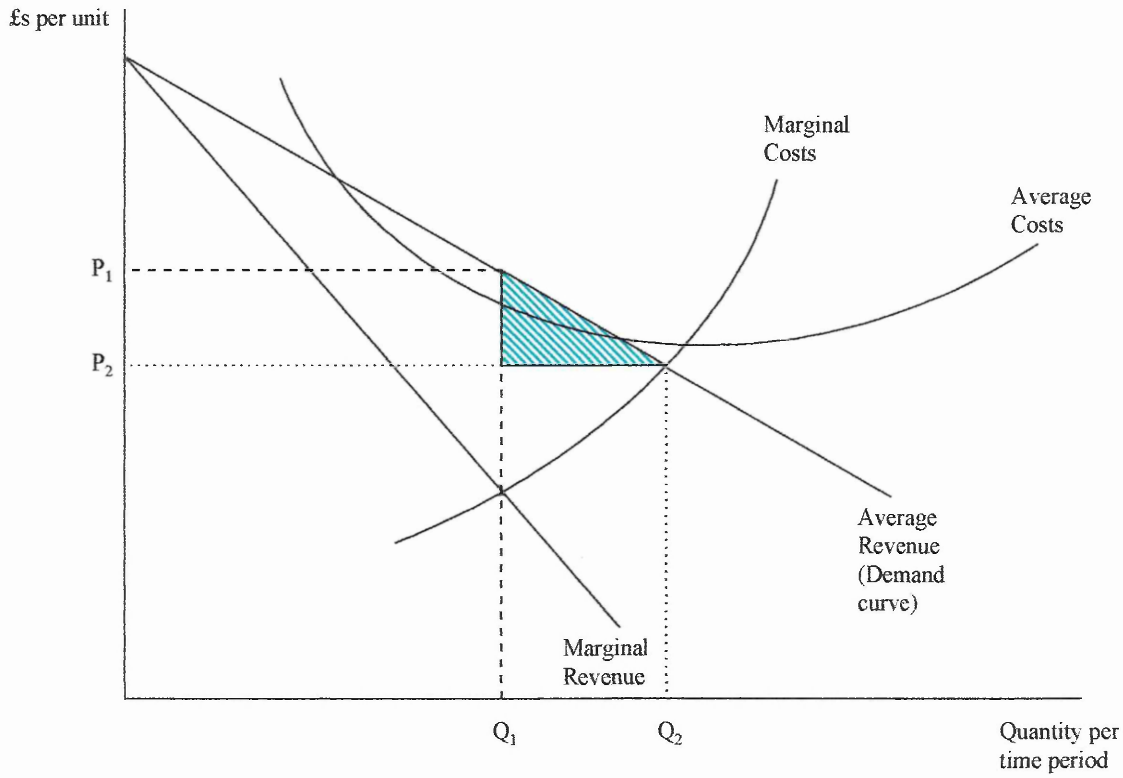

Figure 3. Output and price under monopoly conditions.

The average revenue curve represents the demand curve and the marginal cost curve can be considered as the value the consumer puts on goods foregone as a result of producing this good (Jeffreys, 1985). The output that would maximise society's welfare is the one where the marginal benefit to the consumer (given by the demand curve) equates to the value placed on the additional unit. Therefore the consumer would prefer the firm to produce at an output of Q_2 and a price of P_2 , i.e. at the point where the marginal cost curve intersects the demand curve (figure 3). However, in the absence of competition the monopolist would produce at the point where the marginal revenue equates to the marginal cost, i.e. at an output of Q_1 and price P_1 (assuming the costs remain the same). The consumer surplus would be reduced. Some of this reduction would a gain for the producer, however there is also a loss to society as a result of the monopoly which is represented by the shaded triangle. It is very difficult to quantify this loss due to the number of factors involved. In order to limit the welfare loss monopolies are investigated to ensure that they are operating in the public's interests. In some cases it might be in the consumers interest for a product or service to be supplied under monopoly conditions. This situation is termed a natural monopoly which has been defined by Sharkey (1982) as

> "...there is a natural monopoly in a particular market if and only if a single firm can produce the desired output at a lower cost than any combination of two or more firms. Natural monopoly is defined in terms of a single firm's efficiency relative to the efficiency of other combinations of firms in the industry."

Historically industries such as water and wastewater services, electricity and gas supply and the rail industry have been considered to be natural monopolies due mainly to their use of networks which it would be inefficient to duplicate. Given that these services have been provided under monopoly conditions even following privatisation, regulation is required to prevent abuse of the monopoly conditions by the firms providing the services. The nature of these services, especially for energy and water, provides a further need for regulation. For most commodities the consumer has a choice as to whether or not they purchase them, however,

> "Universal access to clean water and reliable supplies of energy is generally recognised as one of the fundamental quality-of-life benchmarks in late twentieth century society"

and as such regulation is required

> "...to offset limitations of applying conventional market principles and market processes to the utility sector and in order to provide protection for ordinary customers." (Ernst, 1994)

Given that water is supplied under regional natural monopoly conditions and that the customer has no choice over the supplier, regulation is required to protect the consumer interests. The water and sewerage industry differs from the other utility industries in that it is also regulated in terms of drinking water quality and discharges to the environment in order to protect public health and the environment.

Financial regulation

The Office of Water Services (Ofwat) regulates the financial dealings of the utilities, ensuring that:

- water and sewerage functions are properly carried out throughout England and Wales
- undertakers are able to finance their functions (by securing a reasonable rate of return
- no undue preference or discrimination is shown in the way companies fix and recover charges, i.e. the customer's bill should reflect the cost of service provision
- rural customers are protected
- customer interests, including cost and quality of service, are protected
- economy and efficiency on the part of the undertakers are promoted

The director general of Ofwat also has a role to 'facilitate' effective competition between suppliers and potential suppliers and to settle disputes between companies and customers.

The Monopolies and Mergers Commission (MMC) acts as a court of appeal for the director general and the companies. Any proposed mergers within the industry where each party has assets valued at more than £30 million are automatically referred to the MMC.

Environmental regulation

Prior to privatisation, environmental protection was policed by the Regional Water Authorities. However, the role of the Water Authorities as both potential polluters and enforcers of pollution control led to some conflict of interest in that they were unlikely to take action against themselves (Summerton, 1998). Furthermore there was the possibility of consents being adapted to suit the requirements of the industry rather than the environment (Parker and Sewell, 1988; Hassan *et al.*, 1996), reducing the need for the government to invest heavily in modernisation within the industry. This was in line with government policy at the time of reducing public spending (Helm and Rajah, 1994). The results of this situation were that by the mid 80s much of the infrastructure was in a poor state of repair and the water quality in many rivers and coastal areas was declining (Parker and Sewell, 1988; Hassan *et al.*, 1996).

As part of the privatisation process the National Rivers Authority (NRA) was created as an independent regulator to enforce environmental protection. The NRA has since been combined with Her Majesty's Inspectorate of Pollution (HMIP) and local authority Waste Regulatory Authorities to create the Environment Agency (EA) which is responsible for control of discharges to air, land and water. The Environment Agency (EA) regulates environmental water quality, discharges, abstractions, flood defences, etc. Regulation of drinking water quality is carried out by the Drinking Water Inspectorate (DWI).

Having different bodies applying financial and environmental regulation can lead to a conflict of interests in that Ofwat may want a company to restrict capital spending so as to reduce the price for the consumer. On the other hand the EA may want increased capital expenditure in order to achieve further environmental improvements (Helm and Rajah, 1994).

Financial and environmental regulation are features of the water industry in England and Wales that will continue for the foreseeable future due to the necessity of water and ever increasing concern for the environment.

Changes of government could have an impact on the water industry as different political parties have different ideas on privatisation and competition in the utility industries. As an example, it is felt by some that although the previous government was actively pushing for more competition in the industry (DoE, 1996), it does not appear to be so high on the agenda of the current government (Stedman, 1998a).

Economic

Due to the necessity of water and energy and the absence of feasible substitutes (especially for water with its importance for personal hygiene and public health issues) the consumer cannot exercise the power of exit. Hence the demand for these services does not show the same price and income elasticity as that for other commodities (Ernst, 1994). Although little data is available for water demand, the data represented in figure 4 shows that demand for fuel is income inelastic when compared to other products (Central Statistic Office, 1991).

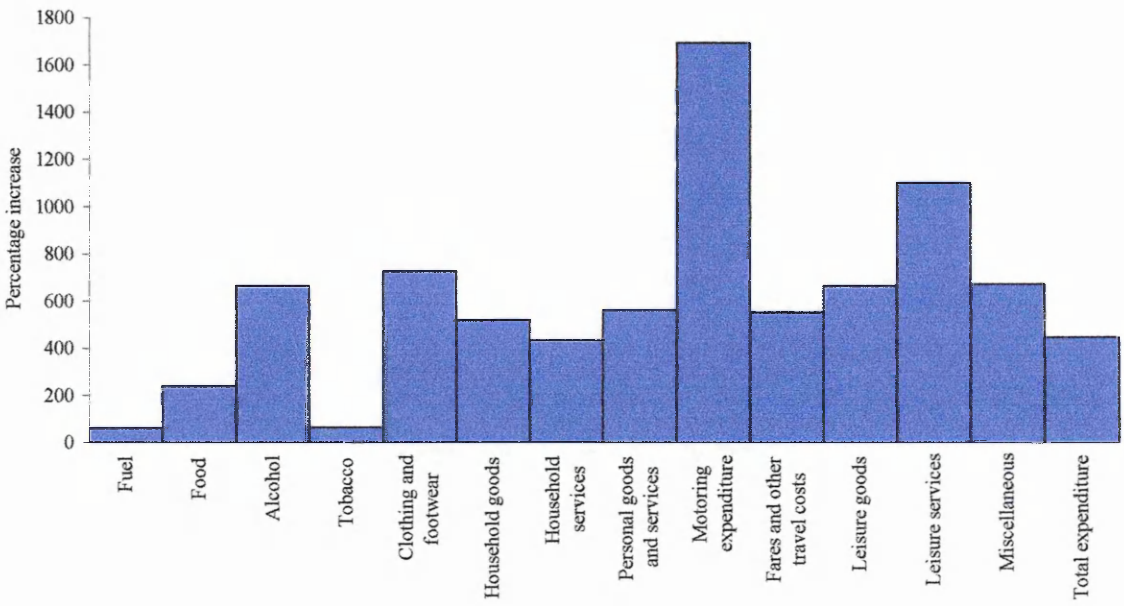

Figure 4. Percentage increase in average weekly spending on major commodities between the 1st and 5th income quintile groups.

Since demand for water and sewerage services is relatively price and income inelastic it is not therefore affected to the same extent as other industries by the usual economic cycles. Hence the income to the core business is virtually guaranteed and generally companies know in advance what the demand and hence their turnover will be.

However, profits will be affected by the usual cycles as the costs paid to suppliers can fluctuate. For example, the recession in the early 1990s led to a big fall in construction costs (table 3) which meant that the costs of construction projects were lower than planned when the K factor was set (Helm and Rajah, 1994). Any substantial benefits would be limited as the regulator could make the company pass on any cost savings to the customers at the time of the next price review along with any other efficiency savings made (Summerton, 1998).

Table 3. DoE Construction Price Index (COPI).

Year	COPI
1988-1989	89.50
1989-1990	99.25
1990-1991	99.25
1991-1992	91.25
1992-1993	83.75
1993-1994	82.25
1994-1995	87.00
1995-1996	95.25
1996-1997	98.30

(Data from WSA 1997)

The stock market can be nervous about how the regulator might intervene in the industry which can lead to fluctuations in the share prices of the companies for no real reason (White, 1992).

Socio-cultural

The population growth will lead to increased demand for water supply and wastewater treatment but this growth is very slow. The total amount of water supplied in England and Wales increased by only 13 % between 1973 and 1997 (figure 1). The increased demand will result in increased revenue for the company but the increase in profit will depend on the marginal costs of meeting it.

Water is often taken for granted by customers, an attitude that results in more water being used than is really required. Better education of consumers could result in less wastage of water, reduce water consumption and help to avoid water shortages as experienced in 1995 and 1996.

Since privatisation the water companies have had to look at how they are perceived by their customers. As customers now get separate bills for water supply and sewerage provision they expect to get a good quality service and good value for money, especially since they have little or no choice in who supplies them. It has been recognised by the privatised water companies that they have to give customer satisfaction (Harper, 1992; Noone, 1993; Smith *et al.*, 1995). To do this they need to know what levels of service the customer expects and what they are willing to pay for. This will become increasingly important if competition is forced onto the water industry by the regulator.

Technological

New processes are continually being developed which should enable companies to meet the environmental restrictions imposed by legislation with improved efficiency and reduced costs.

Development of the technology to treat water and / or sewerage at a much smaller scale could remove one of the main barriers to entry, water company ownership of the distribution / collection networks, would be removed, opening up the market for increased competition. Such processes are already under development, e.g. packaged plant for treating domestic wastewater, in house recycling and re-use of greywater.

6.2.7 Competitive Environment

A tool which can be used to determine how competitive an industry is the five forces model put forward by Porter (1980), the five forces being rivalry amongst existing firms, barriers to new entrants, bargaining power of buyers, bargaining power of suppliers and threat from substitute products or services.

Although the conditions of slow growth and high exit barriers present in the water industry would normally result in intense **rivalry amongst existing firms** it is reduced by the limited opportunities for new entrants and geographical expansion.

When the regional water authorities in England and Wales were privatised in 1989, it was felt that, despite the natural monopoly conditions which existed, there were at least 5 elements of competition that would apply (Water Bulletin, 1988).

Comparative competition is where companies are encouraged to improve their performance in order to be perceived to be as good or better than the others. There are difficulties in applying comparative competition, such as on what basis asset values and costs should be compared between companies given that they can be affected by factors beyond the companies' control, e.g. topography (Cubbin, 1993). Having considered such problems, the director general of Ofwat will apply efficiency targets to the regulated companies at the next periodic review in 1999. These targets, taking into account costs and quality of service, will be set with reference to the most efficient companies (Summerton, 1998).

Industrial water users might choose to treat their own wastewater or abstract and treat water from **alternative supplies**.

Failure to perform well could lead to one company being the subject of a **take-over** bid from another. Since privatisation there have been a number of mergers and take-overs involving the water only companies such that there were only 19 remaining in 1997

(Smith, 1997) of which many are owned wholly or partly by other companies. Some of the WSCs have been taken over by companies from outside of the UK water industry. Northumbrian Water was taken over by the French company Lyonnais des Eaux SA in 1995 and Southern Water was taken over by ScottishPower PLC in 1996 (Smith, 1997). More recently Wessex Water has been taken over by Enron.

Two bids by existing WSCs (Severn Trent and Wessex Water) to take over a third (South West Water) were prevented from proceeding, after referral to the MMC, on the grounds that the resulting savings promised for customers would not be sufficient to compensate for the loss of a comparator company (Haddon, 1996).

The share prices of the companies give an indication of how well they are perceived to be operating. Hence the **capital market** can lead to increased competition.

The only opportunity for competitive entry to the water and sewerage industry was through **inset appointments**. This approach enabled new entrants to supply water and sewerage services to sites which were not connected to the water mains or sewerage network of another company and was 30 metres or more from the mains or sewer network. Existing water and sewerage companies could also bid to supply new customers in the region of their competitors.

In subsequent attempts to increase competition through inset agreements the 30 metre rule was removed and companies were allowed to bid to supply to the large customers (those using 250 megalitres or more of water a year) of the existing regional supplier. Inset applicants were also permitted to obtain a bulk supply of water from the existing water supply company and sell it at a profit to the large industrial customers (The Competition and Service (Utilities) Act of 1992). In 1996 the government proposed that customers eligible for inset appointments should be expanded to premises in 'common ownership but which are separated by highways, railways and / or watercourses'. The 250 megalitres a year rule would still apply but to the combined premises. A second proposal to make inset appointments more attractive to potential entrants was that they should be allowed for a limited, defined time period, e.g. 5 or 10 years, rather than being perpetual appointments (DoE, 1996).

Despite efforts to make inset appointments more attractive to new entrants, there have been few successful applications to date. Up until recently there had been only three inset applications accepted, for one of which the customer (Shotton Paper) had yet to sign a contract with the appointee (Enviro-Logic) (Stedman, 1998a). The other two accepted appointments were made by an existing WSC, Anglian Water at Buxted Chickens and RAF Finningley (Byatt, 1998). As the pressure to increase competition increases, more WSCs are starting to apply for inset appointments (Thames Water has won a contract to supply water and sewerage to an army base in Hampshire) but Ofwat have heard from only two potential new entrants (Haddon, 1998).

To increase competition further Ofwat supports reducing the minimum supply volume for inset appointments from 250 megalitres a year to 100 megalitres a year. This would significantly increase the number of customers who could potentially benefit from inset appointments.

A significant effect of the threat of losing customers through inset appointments is that the majority of companies supplying water and sewerage services have introduced large user tariffs (Cowan, 1997). This will reduce the possibility of new entrants 'cherry-picking' the most profitable customers. It is up to the regulator to ensure that the large user tariffs are not offered at the expense of the remaining customers being charged unfairly high prices. The logical result of this is a move towards cost reflective charging structures, e.g. rural customers could face higher charges due to higher distribution charges (Cowan, 1997; Stedman, 1998b).

Other methods of competition have been considered since privatisation. The Competition and Service (Utilities) Act of 1992 introduced legislation allowing **Cross-border competition**, i.e. enabling domestic customers to connect to the water mains or sewerage network of a neighbouring regional provider rather than the one in whose region they are located. Connection costs would be the responsibility of the customer.

In 1996 the government proposed to expand the scope of cross-border competition to include smaller agricultural, industrial and commercial customers not eligible for inset appointments (DoE, 1996).

Further proposals made in 1996 to increase competition in the water industry included allowing the customer (through a third party contractor) to make the connection to the mains - provided it is up to the standard stipulated by the company owning the mains. This would allow contractors to compete with the existing companies for **mains connection**. There is currently no motivation for the existing water and water and sewerage companies to reduce the costs for connections. The most controversial proposal was to allow **common carriage**, i.e. allowing 'any person with a suitable source of supply and prospective customers' access to the water mains of the existing companies. This would apply to the existing WOCs and WSCs and prospective new entrants deemed fit to supply potable water. Anybody supplying water through another company's network would have to pay an agreed fee to that company. If no agreement could be reached the fee would be determined by the Director General of Ofwat (DoE, 1996).

The introduction of common carriage would require a government act and whilst this was proposed by the last government (DoE, 1996) and supported by Ofwat (Byatt, 1998; Haddon, 1998) it may not be on the agenda of the current government (Stedman, 1998a).

There are issues surrounding common carriage which need to be addressed before it could be implemented. These include practical issues such as the effects of mixing water from different sources, increased flows through some sections of pipework and water quality issues. There are also legal issues such as who would be responsible in the event of a breach of water quality (Stedman, 1998b). The practical issues were the subject of an investigation by the Water Research Centre (DoE, 1996).

Despite the mechanisms in place to encourage competition in the water industry interest from potential new entrants has been limited (Haddon, 1998). This could be due to the existence of further **barriers to entry**. The capital costs of entry are very high for treatment plants and supply networks. Furthermore duplication of the networks is not really economically or practically feasible (Summerton, 1998). The close regulation of the industry and the considerable experience of the existing companies could be further factors which deter potential entrants.

The **bargaining power of buyers** is not significant currently as there is a very high number of customers, most of them being very small users, e.g. households, and have little or no choice over their water and sewerage supplier. In the few cases where the consumer could switch suppliers the associated switching costs, e.g. providing new pipelines, could be prohibitive. Some very large industrial users might be able to negotiate different prices and some water companies now offer a large user tariff (Cowan, 1997).

The major supplier to the water industry is the construction industry. As such the **bargaining power of suppliers** is likely to be low due to the competitive nature of construction with several companies bidding for major capital schemes. The power of suppliers is further reduced as most water and sewerage companies source up to 13% of capital expenditure from sister companies (Smith, 1997).

Since water is a necessity for life there is no **threat from substitutes**. Although many people use bottled water for drinking, the transportation costs would prevent bulk supply of bottled water for other purposes. There could be a possibility of alternatives to certain aspects of the industry. It has been suggested, for example, that although no national grid exits for water, bulk water movements could be carried out using the existing canal networks rather than pipelines (Heslop, 1997). Reduced or zero emission technology such as in house recycling or a move away from water as the medium for domestic waste removal could also be seen as a possible substitute (Summerton, 1988).

6.2.8 Future Developments

Examination of the environmental factors and the competitive forces that impact on the water industry in England and Wales has led to three possible scenarios for the future development of the industry. These scenarios are outlined below together with the likely consequences for the existing companies in the market.

Existing companies retain their regional monopolies.

Although mechanisms are in place for the existing water suppliers to provide water to some of the customers of their competitors, it is considered unlikely, by some observers, that firms will compete on this basis (Cowan, 1997). However the threat of competition could lead to a change in the charging structure to reduce the risk of competitors attracting business away from the existing suppliers.

If the existing companies continued with their existing markets - they would need to reduce costs and improve efficiency. They are required to do this by Ofwat but it would also help them to retain a strong position in order to avoid take-over and to continue to make large profits in the face of price cuts imposed by the regulator.

Should the market continue along similar lines for the foreseeable future then it would remain very difficult for firms to increase their market share in an overall market that is growing only slowly. In this event firms will have to apply their core competencies outside of the core business, either through extending the product range or finding new markets, in order to achieve growth. Failure to do this could leave the firms open to being taken over.

Full competition introduced in water industry.

Despite the doubts raised by some observers over the introduction of full competition in the water industry in England and Wales, the Director General of Ofwat remains determined to carry it out. He recently expressed his desire for the companies to agree on codes of access for the introduction of common carriage (WET News, 1999).

Full competition would give existing firms and potential new entrants the opportunity to increase their market share. Conversely it would also leave the existing companies open to losing some of their existing market share to their competitors.

In order to compete companies would have to offer consumers the product and service they want at a price they are willing to pay. Hence firms would have to increase the perceived use value of the services they provide and at the same time reduce costs, allowing them to compete on price if necessary. In order to sustain competitive advantage in the long run firms must aim to be the lowest cost supplier. It is felt by some that levels of service rather than price will be the main focus of competition.

Should this scenario occur firms will still have to improve efficiency and reduce costs - both to satisfy the regulator and to remain competitive.

Ofwat and MMC allow mergers/take-overs of water and sewerage companies.
Should this situation arise it would open up another route for WSCs to expand their market share. In order not to be taken over companies would have to be in a strong position, i.e. efficient companies with a strong financial position. Such firms would then be in a position to take over other firms themselves or to merge with other firms. This would allow the newly formed larger firm to take advantage of scale economies and to make efficiency savings through less duplication of effort, e.g. in research and development effort. Such take-overs and mergers will only be permitted to take place if there are considerable benefits passed on to the consumer.

6.3 POTENTIAL WATER INDUSTRY STRATEGIES

All three of the possible scenarios outlined in the previous section will require the companies in the water industry to improve their efficiency in order to stay competitive. This is equally important for competition in the core market and to enable firms to compete globally. It is also probable that firms will have to start increasing their activities in areas outside of the core business in order to achieve growth. Some of the strategies adopted by firms operating in the water industry in England and Wales are discussed below.

6.3.1 Consolidation

Given that they cannot easily increase their market share and are under constant pressure from the regulator to reduce charges to customers, all of the water companies have to consolidate their position within the industry. This includes improving efficiency and reducing costs.

The companies that did this without looking to expand business elsewhere have been vulnerable to take-over, e.g. Southern Water and Wessex Water. These companies now have no interests outside of their core activities but are part of large groups with numerous outside interests.

Companies looking to expand outside of the regulated business still need to consolidate and streamline their utility operations as this is the base from which they are expanding. One way to improve efficiency is to split the utility business up into a number of functions and focus on those functions in which the company has the most competence and where it can add the most value, e.g. conceptual or process design, plant operation, network provision and maintenance. Partnerships or alliances can then be formed with the outside partners providing competence and adding value in the other areas, e.g. detailed design, purchasing, construction.

Many companies have restructured in order to reduce overheads, often involving cutting the workforce and outsourcing some activities, e.g. billing and customer services being provided by a service company within the group but outside of regulatory control.

6.3.2 Market Development

Since it is difficult for the water companies in England and Wales to increase their market share in the regulated water and sewerage sector they have to look to new markets in order to exploit core skills to expand the business.

Several of the WSCs have looked to overseas markets with varying degrees of success. Although the potential market is massive, an estimated £200 billion a year, there are also large risks for companies with little or no experience in the overseas markets. There are high costs involved in setting up international operations which probably contributed to the losses made by companies, including Thames Water and Anglian Water, during the first years of overseas operations.

Improvements have been made though and in the year to March 1998 Thames Water made £6 million profit on its overseas activities. In the year to March 1999 United Utilities and Severn Trent also made an operating profit on international activities but in the same year Northumbrian Water Group made a loss. Despite increasing international success, these operations still only account for a small percentage of total group income.

Some companies have split off their services business, e.g. IT, billing, customer call centres, etc. into companies separate to the regulated business. The regulated business then pays the service company for services provided. Where this has happened the service company can make use of its experience to expand externally within the UK. United Utilities and Hyder have both increased the external work carried out by the service subsidiaries.

Similar opportunities exist for companies to make use of their existing experience of running environmental laboratories to develop more external work an opening Yorkshire Water is exploiting through its subsidiary Alcontrol.

6.3.3 Product Development

Some of the WSCs have developed business outside of the core activity but in related areas. The waste management industry is one in which several of the WSCs are prominent, notably Northumbrian Water Group, Pennon Group (parent of South West Water), Severn Trent and Yorkshire Water. Waste management activity enables these groups to offer a complete package of waste treatment and disposal, solid and liquid, to industrial and commercial clients. For example, Northumbrian Water handles all of the waste from Nissan's car plant through its waste management subsidiary. White Rose Environmental, owned by Yorkshire Water, is the market leader in clinical waste disposal - a growing market due to stricter legislation.

Products and services related to water and wastewater treatment, e.g. process plant and instrumentation, present an opportunity for companies to increase their income.

Pennon, for example, has focused on the environmental instrumentation market and has up to 50% market share in some key areas.

There are now a number of multi-utility companies operating in the UK, the result of mergers and take-overs involving water and electricity utility companies. Hyder (Welsh Water), United Utilities (North West Water) and Scottish Power (Southern Water) are all involved in the sale of water and sewerage services, electricity and gas. Scottish Power, having taken over Southern Water, have sold off all of the non-core businesses of the former WSC. However, Hyder and United Utilities have continued non-core activities in water and sewerage in their home markets and abroad.

The combined utility operation enables these companies to benefit from cost reductions through having a single service company providing billing, customer services, IT, etc for all the utilities, i.e. economies of scope. Both Hyder and United Utilities have cited the success of their service companies as contributing to the improved levels of customer service that they have achieved. This could become even more important if competition develops in the domestic water and sewerage market and customer service becomes a key competitive factor.

The multi-utility companies can offer packages to customers which include gas, electricity, water and telecommunications, e.g. United Utilities contact to provide all of these services to the Trafford Centre in Manchester. With domestic gas and electricity opened up to full competition in 1998 and 1999 respectively, provision of all utilities provides these companies to increase their market share in domestic markets.

Most of the WSCs have property subsidiaries which are used to make the most out of land and property deemed surplus to requirement by the parent company.

6.4 POTENTIAL BENEFITS OF PROCESS MODELS AND FIXED FILM / ACTIVATED SLUDGE HYBRID PLANTS

The development of mathematical models of wastewater treatment processes has led to a better understanding of the processes being used. The improved understanding can be applied, using the models, to the benefit of companies treating wastewater. As discussed above, it is important for companies in the water industry to reduce costs and improve efficiency in order to sustain competitive advantage. There are a number of applications of models which can help to achieve this.

Environmental regulation has led to more stringent discharge consents for wastewater treatment plants, normally requiring substantial capital investment. However, it is possible in some cases to improve the performance of existing plants without major construction work. Process models have been used to identify situations where this was possible (Coen *et al.*, 1996; Kurata *et al.*,1996). Dupont and Sinkjaer (1994) and Leeuw *et al.* (1996) found that by optimising existing plants, effluent quality could improved by up to 40% in terms of total nitrogen discharged.

Should new plant be required, it is in the interests of the water companies to keep the required expenditure to a minimum. Use of modelling during the design of treatment plants saved a UK water utility construction costs of approximately £70 000 by enabling sizing of the plant 23% smaller than it would have been with manual calculations (Monro, 1996). Similar capacity reductions were identified by Leseouf *et al.* (1992) when considering different upgrading options for a wastewater treatment plant.

A major operating cost of a wastewater treatment plant is the energy cost. Improved design of processes and control strategies using mathematical models has enabled energy savings of up to 21% possible in activated sludge plants (Pederson, 1992; Brouwer *et al.*, 1994; Thornberg and Thomsen, 1994; Horan and Chen, 1998). Energy savings having also been achieved for other processes such as sequencing batch reactors (Demuynck *et al.*, 1994) and oxidation ditches (Parker *et al.*, 1992).

Improved operator training could result in improved performance of existing wastewater treatment processes. Models have been used in training simulators for operators which enable them to see the effects on the processes of altering various control parameters. This should help to improve decision making for example if there was a process disturbance (Schreiber *et al.*, 1992; Metzger, 1994; Barnett *et al.*, 1995).

Improved planning of sewerage services by considering the interactions between the sewer systems, wastewater treatment plants and river quality has been shown to be feasible through the use of process models of theses three components. Crabtree *et al.* (1996) demonstrated that this approach led to selection of an upgrading option with a lower cost than options arrived at through traditional methods.

As discussed, in numerous cases, the use of mathematical models has led to significant savings of both capital and operational expenditure. However, because of the high costs of model calibration, any works that are to be modelled have to be selected carefully to ensure that there will be a cost benefit to the company.

The cost of model calibration can be as high as £30 000, which is made up of 2 main components. The most significant cost is the analysis required for wastewater calibration which depends on the complexity of the works and how comprehensive the survey is. The cost ranges from £6 000 for a 5 day sampling programme with 7 sample points (Shafibeik, 2000) to upwards of £20 000 for a more comprehensive 4 week survey with daily sampling for 3 weeks and 2 hourly sampling for a week. The other main cost is the time of the modeller which could be up to £10 000 depending on the complexity of the works. Other costs include overheads such as computer hardware and software licence costs.

In order to be of benefit to a water utility, modelling has to be targeted at works where the potential savings will exceed the costs of the modelling exercise. For example, it is unlikely that significant savings would be made at a small works which doesn't have a strict effluent discharge. However for a larger works which might have a new or stricter ammonia consent to be applied in the future, or where substantial growth has been identified, there is a high possibility that modelling could lead to cost savings in excess of the calibration costs.

For a capital scheme with a value of £2 million, a saving of 2.5% would result in savings of £30 000 after the costs of calibration assuming £20 000 costs of model calibration. For more complex, larger schemes, a 2.5% Capex saving would give savings £95 000 after the costs of a more intensive calibration process costing £30 000. With large numbers of schemes coming up in the next AMP period, careful targeting of sites to be modelled could yield significant savings to the water utilities.

Recently the EA has required the water companies to meet increasingly strict discharge consents from their wastewater treatment plants. At the same time Ofwat has demanded that the companies increase efficiency and has reduced the price that can be charged to customers. This has meant that innovative solutions are required to meet the new treatment standards whilst minimising capital and operational expenditure. Any solution where the performance of the plant can be enhanced using the existing reactor volumes will help to meet the new performance targets.

By incorporating fixed film biomass into conventional activated sludge tanks, the process can be made more intensive and hence meet tighter effluent quality standards. Examples include carbon only removal plants that have been upgraded to nitrogen removal within the existing reactor volumes. The MBBR is a high rate biofilm process that has been combined with a conventional ASP in a hybrid plant at Colchester STW. Full scale trial and modelling results have demonstrated that retrofitting the whole plant with the hybrid process would enable the plant to meet its new discharge consent without having to construct any new tanks. The estimated saving of this solution over other options is approximately £1.75M.

6.5 CONCLUSIONS

Due to regulatory constraints the 10 water and sewerage companies in England and Wales are having to continually improve the performance of their treatment plants to meet new standards whilst having their income from customers restricted.

Options for expanding the customer base to increase income are limited within the regulated business and so to maintain profits the companies are having to become more efficient. Companies have to find ways of achieving competitive advantage.

Most of the companies that have not been taken over by overseas companies have themselves developed business outside of the regulated utility business. However the utility business still remains the main source of income and so efficiency still has to be improved within the core business if the firms are to avoid take-over.

In order to meet the new performance targets, both quality and financial, the companies have to look for low CAPEX and low OPEX solutions. Process modelling is a tool that enables opportunities for reducing operational expenses and avoiding major capital expense to be identified.

Hybrid fixed film / activated sludge systems can enable existing wastewater treatment plants to be upgraded to meet new discharge consents for a relatively low capital outlay.

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CHAPTER 7 - DISCUSSION

7.1 DISCUSSION

Analysis of the competitive environment within the regulated water industry in the UK shows the industry to be growing in terms of turnover, from £3bn to £6.5bn in the 10 tears to 1998. However due to tight financial regulation the opportunities for an individual company to increase its market share are limited within the regulated business. More competition in varying formats is likely to be enforced in the industry over the next few years which offers both an opportunity and a threat to the existing companies in the market.

The financial regulator, Ofwat, is committed to reducing the cost of water in real terms for the customer over the next 5 years. At the same time conflicting pressures are being exerted on the water companies by the EA. In the most recent AMP determinations even stricter discharge consents have been applied to the wastewater treatment works of the water and sewerage companies - many works will have to be upgraded for nitrogen and/or phosphorus removal.

These pressures are forcing the water companies to become more efficient and to look for innovative solutions for asset investment when trying to meet the new environmental obligations. This is required in order for firms to maintain competitive advantage and failure to do so would lead to reduced profits . This would have a knock on effect on the share price and could lead to take-over. Even as firms look increasingly to non-regulated business for income, they still need to maintain the core business which continues to be the main source of revenue and profits.

Wastewater process modelling is a tool which can be used to identify opportunities for improving the performance of wastewater treatment plants, potentially to meet stricter discharge consents without new build. Modelling can help on several different levels.

Developing, calibrating and application of process models increases the level of understanding of that process. This allows a plant to be run under optimised conditions and closer to the limits of performance.

Modelling can lead to increased confidence in a process which then leads to the safety factors typically applied during design procedures to be reduced. It could be argued that this happens anyway with increasing experience. However, there might be no inclination in practice to find the real limitations of a works but if a modelling exercise highlighted potential savings it could be the catalyst for decreasing safety factors.

In order to reliably calibrate a process model the modeller has to consider in detail the physical and operational factors affecting plant performance. Major differences between simulated and actual performance are as likely to be due to operational discrepancies as to process kinetics.

For example, during calibration of the model of the Colchester STW hybrid plant the MLSS could not be calibrated without adjusting the surplussing rate of the ASP. When the mixed liquor transfer pumps were beaten by the influent flow, mixed liquor flowed over a stop log, resulting in a higher surplus rate than intended. During the trial this was thought to be insignificant as nitrification was maintained, however the modelling work indicates that the sludge age was 4 days rather than 5 days during the latter stages of the trial.

Combinations of fixed film processes have been demonstrated to increase the efficiency of wastewater treatment processes, in terms of treatment per unit volume. Such hybrid processes have been used to increase treatment capacity in carbonaceous treatment plants, (Morper and Wildmoser, 1990; Gebara, 1999) and to upgrade existing plants for ammonia removal (Lessel, 1994; Randall and Sen, 1996). Improved performance is due to a much increased total biomass within the same reactor volumes which leads to higher overall sludge ages and lower F:M ratios. A further benefit is that since a large proportion of the biomass is retained in the aeration tank, additional final sedimentation tanks are not required.

The use of such processes can lead to large savings in capital expenditure when upgrading plants with Randall and Sen (1996) reporting potential savings of \$14.8m. After successful pilot scale trials of various configurations of the MBBR/AS process (Jaouen, 1996; Lievre, 1997; Michel, 1998) the process was identified as an option for upgrading Colchester STW to ammonia removal within the existing reactor volumes. Use of this option would realise savings of up to £1.75m over the other options identified to meet the new ammonia consent.

Full scale trials were carried out by retrofitting the hybrid process into one of the aeration lanes at the works. During the first period of operation the hybrid plant was not isolated hydraulically from the rest of the plant with common final sedimentation tanks being used. Although during the summer some nitrification was achieved, full nitrification could not be maintained in the plant. The sludge age for the whole works during this period was 4 days and while this was considered sufficient for nitrification in the hybrid plant it did not enable nitrification to occur in the other aeration lanes. The design F:M ratio for the hybrid (based on the total influent BOD_5 and the MLSS) was 0.43 kg BOD kg^{-1} MLSS d^{-1}. At this loading rate nitrification would not be expected although Lessel (1994) reported an effluent ammonia of 2 mg l^{-1} for an F:M of 0.4 kg BOD kg^{-1} MLSS d^{-1} based on the suspended biomass.

During the second phase of the trial the hybrid was isolated from the rest of the works and the sludge age was controlled independently. For a sludge age initially thought to be 5 days but later calculated as 4 days full nitrification was achieved and maintained for a month. The load during this period was only an average of 0.8 kg BOD_5 m^{-3} d^{-1} on the MBBR during this period. This equated to an F:M for the influent BOD_5 and MLSS only of 0.09 kg BOD_5 kg^{-1} MLSS d^{-1}. For this loading rate nitrification would be expected in an activated sludge plant. The rate of ammonia removal achieved in the AS zone during this period was 0.08 kg m^{-3} d^{-1} which is comparable to the rate of 0.12 - 0.15 kg m^{-3} d^{-1} reported by Bonhomme *et al.* (1990) for an ASP with fixed biomass. The nitrification rate was within the range of those reported in pilot trials for a similar plant configuration, 1.75 gNH_4-N kg^{-1} MLSS h^{-1} for this work compared to 1.02 and 3.96 gNH_4-N kg^{-1} MLSS h^{-1} (Lievre, 1997).

Although a number of authors have reported improved settleability in plants that combine fixed and suspended biomass (Wanner, 1988; Lessel, 1994; Jones *et al.*, 1998), this was not observed at Colchester during the trials. The $SSVI_{3.5}$ for the hybrid plant was not significantly lower (13 %) for the hybrid plant than for the other aeration lanes at this site.

Overall the performance of the MBBR/AS plant compared favourably to other fixed film activated sludge systems and is summarised in table 1.

Table 1. Comparison of performance of different hybrid systems.

	Removal (%)				
Determinand	MBBR/AS	Ringlace[1]	FM-AS[2]	Linpor[3]	TF-SC[4]
TSS	86	...	88	...	96
BOD_5	85	94	95	...	98
COD	84	87	88	84	...
Ammonia	88	95	95	96	...

([1] Lessel, 1994; [2] Han-Chang and Thomas, 1993; [3] Lessel, 1994; [4] Parker *et al.*, 1993)

Results from the second stage of the trial gave promising indications that upgrading to the MBBR/AS hybrid process would enable Colchester STW to meet its new ammonia consent. A longer period of testing with the hybrid isolated from the other lanes would have been preferable. This would have enabled the hybrid to be operated at the higher loading rates expected under normal operation and under more stable operating conditions.

In order to obtain further information to support the hybrid option for Colchester, a process model of the MBBR, being used for the first time, was calibrated and used to predict the performance of the full scale hybrid plant under future loading conditions.

The model was found to predict the average plant performance to within 20 % of the measured effluent quality. When the dynamic performance of the plant was simulated it was found that the model predicted the trends in the measured data, although some of the absolute values were not predicted. This was attributed to reactor hydraulics and anomalies in the sample analysis. The average effluent quality predicted for the full scale upgrade was 22 mg l^{-1} TSS, 12 mg l^{-1} BOD_5, 0.2 mg l^{-1} NH_4-N and 26 mg l^{-1} TON. This performance is within the new consent obligation for the plant, indicating that the hybrid would be suitable for upgrading Colchester STW. With the anoxic zones in use in the full scale plant it would be expected that some denitrification occurred, however the model did not predict this.

Recently the newly upgraded plant has been commissioned and the model prediction is comparable to the results obtained during early commissioning work. Observed performance during the first few months of commissioning was 8.3 mg l^{-1} TSS, 4.5 mg l^{-1} BOD_5 and 1.6 mg l^{-1} NH_4-N and 29 mg l^{-1} TON. (Wright, 2000). The actual plant data shows a lower observed TSS and BOD_5 in the effluent although it is interesting to note that denitrification did not occur in the full scale plant, in line with the modelled results.

Despite the model predicting performance similar to that which was observed in practice, the confidence in the model is limited. The confidence in any model corresponds to the quality of input data. The influent characteristics for the Colchester model were obtained by using the BOD based influent model in GPS-X. This model relates the influent TSS, BOD_5 and TKN to the state variables used in the model using a number of stoichiometric ratios. The ratios used include BOD_5:BOD_{20}, particulate COD:VSS and ammonia:TKN. When the sampling was carried out a number of problems were encountered with the analysis of some of the determinands, particularly BOD_{20} and TKN results. Similar problems have been reported elsewhere (Smith and

Dudley, 1997). The problems with the analysis led to COD fractions being calculated which were outside of the typical ranges reported by other authors. The most notable difference was for the particulate inert COD which was overestimated by the BOD based model.

Sensitivity analysis on the parameters Y_H, K_S and μ_{max} showed that only changes to μ_{max} had a significant effect on the effluent quality predicted, although variations in μ_{max} of ±50% from the default value only resulted in the predicted effluent quality changing from -20% to +30% of the value obtained using the default value.

In order to get over the problems encountered with influent characterisation and gain more confidence in the MBBR model, a second modelling exercise was carried out on a large pilot scale plant. Influent characterisation using methods described by Mamais *et al.* (1993) and Leseouf *et al.* (1992) led to estimates of COD fractions within the typical ranges reported in literature. These COD fractions were then used to set up one of the other influent models in GPS-X, the ASM2 model influent, to give the correct COD fractions for the COD and TSS under average conditions. Although under average conditions the correct fractions were predicted, as the ratio of influent COD:TSS varies so will the various COD fractions. As the ratio moves outside of a given value unrealistic values for the COD fractions can be predicted, in this case, if COD:TSS < 2.02, the readily biodegradable substrate became <0 which cannot happen.

Despite the problems with the stoichiometry under certain conditions the model prediction was within 5% of the measured data for all measured determinands. To achieve this μ_{max} for the heterotrophs was adjusted to 8 d^{-1} and the molecular diffusion coefficients were set to 80% of those in water. Under dynamic conditions the model prediction showed less variation than the measured data although it was generally within 20% of the measured data.

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CHAPTER 8 - CONCLUSIONS AND RECOMMENDATIONS

8.1 CONCLUSIONS

1. Combining fixed and suspended growth processes, in a single stage or 2 stage process, has been found to improve process performance. Process efficiency in conventional carbonaceous treatment ASPs has been improved by the addition of fixed biomass and carbonaceous plants have been upgraded to nutrient removal within the same reactor volumes through the use of fixed and suspended biomass. Further benefits include improved process stability and better settling of the activated sludge.

2. Calibrated process models have been found to benefit companies by enabling more efficient process design, identifying cost effective upgrading options, giving a better process understanding and enabling more effective operator training. Although significant savings can be made through the use of modelling, these savings have to exceed the cost of calibrating the model which can be significant due to the high cost the analysis required for influent characterisation. Care has to be taken in characterising the influent as the quality of the input data affects the levels of confidence in the model.

3. The full scale hybrid trial plant was found to achieve an average effluent quality of 25/10/4 mg l^{-1} for TSS, BOD_5 and ammonia respectively at a sludge age of 5 days for the activated sludge zone. This effluent quality was within the new effluent consent for Colchester STW but was achieved at very low loading rates, 0.8 kg BOD_5 m^{-3} d^{-1} on the MBBR and an F:M ratio of 0.09 kg BOD_5 kg^{-1} MLSS d^{-1}.

4. Calibration of the MBBR process model for full scale and pilot plants showed the model to be able to predict actual performance to within 20% under steady state and dynamic conditions. However, caution should be applied when applying the model calibrated for these plants due to problems encountered during influent characterisation.

5. Simulation of a full scale hybrid upgrade at Colchester STW predicted an average effluent quality of 22/12/0.2 mg l^{-1} for TSS, BOD_5 and ammonia respectively and performance under typical diurnal influent variations was also within the future consent for the works. The modelling supports the findings of the trial that the hybrid plant would enable nitrification to be achieved within the existing tank volumes at Colchester STW. However, as discussed some caution should be taken in interpretation of these results.

6. Water companies in the UK are facing conflicting pressures in having to meet increasingly stringent discharge consents from wastewater treatment plants whilst having their income restricted and reduced in real terms. In order to maintain profits companies have to become more efficient in terms of capital and operational expenditure. Process modelling can be used to identify potential savings in both capital and operational costs when upgrading existing plants. Any process such as the MBBR/AS hybrid which can be used to upgrade a plant without having to build new tanks can lead to significant operational savings.

8.2 RECOMMENDATIONS

1. The load during the full scale trial of the hybrid plant was significantly lower than the design loads. Further trials at higher loading rates would give more confidence in the process.

2. Modelling of the hybrid process was limited by the quality of the data obtained prior to the trial being ended. For application of the model in process design of hybrid plants, further modelling development is required, with great care being taken over the influent characterisation.

3. Once greater confidence has been obtained in the model, further simulation should be carried out to determine the operational limitations of the hybrid plant in comparison to a conventional ASP.