NITROGEN AND PHOSPHORUS REMOVAL FROM DAIRYSHED EFFLUENT USING A SEQUENCING BATCH REACTOR

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Abstract

It is apparent that present dairyshed effluent treatment systems are not capable of complying with regulations generated by Regional Councils implementing the Resource Management Act 1991. This has created a need for research into dairyshed effluent treatment.

To develop an improved treatment system for dairyshed effluent, research was conducted with two main study objectives; to characterise effluent from the dairyshed holding yard and anaerobic pond, and to develop a sequencing batch reactor (SBR) for the removal of nitrogen and phosphorus.

The carbon characterisation showed that there was a large difference between dairyshed effluent and domestic effluent in the proportion of carbon in each fraction.

When treating dairyshed wastewater to reduce BOD, nitrogen and phosphorus concentrations it was not possible to treat either the yard effluent or the anaerobic effluent without addition of external materials.

The BOD reaction rate constant for the yard effluent at 0.2 d^{-1} was similar to a typical domestic wastewater value of 0.23 d^{-1}. The anaerobic pond effluent BOD reaction rate constant of 0.16 d^{-1} was lower than the yard effluent value indicating that the anaerobically treated effluent was hard to treat aerobically.

A pilot scale SBR treating dairyshed effluent was operated for 75 days. Startup procedure used a 50/50 mixture of anaerobic pond and aerobic pond effluents which was successful in establishing a biomass capable of nitrifying anaerobic pond effluent. The startup time to establish a nitrifying population was 17 days.

The sludge was found to settle well, with a maximum sludge volume index of 54 ml/g measured during the SBR operation. Sludge bulking was not seen as a problem.

Nitrification performance a large proportion of the bacteria were lost took only 5 days to recover. With the addition of alkalinity nitrification reliably reduced the effluent ammonia concentration to 5 mg/l.

From the cycle analysis the first order reaction rate constants for nitrification were; ammonia reduction 0.7 hr^{-1}, TKN reduction 0.4 hr^{-1} and nitrate formation 0.2 hr^{-1}. These constants could be used in future work to optimise stage times.

KEYWORDS: Sequencing Batch Reactor; Dairyshed effluent characterisation; readily available carbon; nitrogen and phosphorus removal; activated sludge; venturi aerator; Sludge Volume Index.

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1. LITERATURE REVIEW

1.1 Introduction and Background

1.1.1 Dairy effluent

The effluent produced in a dairyshed is made up of animal wastes and wastes produced from washing milk collection equipment, spilt milk, and animal health products. These substances can have considerable polluting potential if not handled correctly. The amount of waste produced is dependent on; herd size, milking time, herd management practices and shed washing practices. With large herds the volume of wastewater produced can be considerable. A design figure often used is 50 l/cow.day with a range from 20-90 l/cow.day (Vanderholm, 1984).

The number of cows in production is increasing each year. In the 1989/90 season there were 2,313,822 cows milked in New Zealand. In the 94/95 season this number increased to 2,830,977. This increase has been due mostly to increasing herd size. The average herd in 1989/90 was 159 cows which increased to 193 in 1994/95 (Dairy Statistics, 1994/95). These increases often overload the farm effluent treatment system (Manawatu Wanganui Regional Council, 1995).

1.1.2 Why control effluent discharges

Effluent pollution degrades the natural environment reducing the quality of waterways by causing excessive weed and algae growth, reduced aquatic animal life and making waterways unsuitable for recreation activities (Heatley, 1996).

Degradation may be from physical, chemical, biological or social effects.

1.1.2.1 Adverse physical effects

Adverse physical effects alter the physical properties of the waterway. Sediment may alter the colour, turbidity or temperature of a waterway. This may disrupt the aquatic communities that rely on light penetration, adversely affecting plant and algae photosynthesis. Sediments can also smother invertebrates on the stream beds and clog fish gills.

1.1.2.2 Adverse biological effects

Organic material can cause excessive growth of bacterial and fugal slimes. These growths use oxygen in the breakdown of organic material therefore reducing the amount of oxygen available for other aquatic life such as fish.

Aquatic weeds and algae play an important role in nutrient cycles of waterways. The normal cycle of plant growth in large aquatic ecosystems include the development of spring algal blooms which deplete the nutrients from the surface layer of waters for the entire summer. Summer nutrient levels in natural surface water rarely exceed 1 mg-N/l as NO_3 or NH_4, and 0.03 mg-P/l as PO_4 (Mulligan, 1970).

Effluent addition to waterways causes increased nutrient levels, which results in increased activity by the organisms present to remove the additional nutrients. Different organisms are more efficient at metabolising the nutrients due to a larger surface area to volume ratio. Populations can bloom smothering the other organisms present. Eventually single species populations develop (Mulligan, 1970).

Often planktonic algal blooms develop when excess nutrients are present. Algal blooms cause discolouration of the water and reduced light penetration to plants lower in the water. If the nutrient supply to the water is continuous, algal blooms will be

sustained and bottom weeds will not survive. Summer algal blooms are sign of deteriorated water quality.

1.1.2.3 Adverse chemical effects

Chemicals such as pharmaceuticals and cleaning agents can act as poisons to aquatic plants and animals.

The toxicity of ammoniacal-N depends strongly on receiving water pH. pH controls the relative concentrations of NH_4^+ and NH_3. The toxic form is NH_3. Ammoniacal-N imposes an oxygen demand on receiving waters where nitrification occurs. The USEPA[1] guidelines gives a final acute value (FAV) for stream invertebrate species of 0.52 mg/l free ammonia-N (Davies-Colley, 1996). A New Zealand study found that native stream invertebrate species were more sensitive and had a FAV of 0.15 mg/l (Hickey & Vickers, 1994 quoted by Davies-Colley, 1996). A provisional guideline that takes into account New Zealand's more sensitive stream invertebrate species has been set at 0.22 mg/l free ammonia-N.

1.1.2.4 Adverse social effects

Physical effects of pollution on waterways also has social effects. People no longer enjoy using water resources which have been rendered unsuitable for stock watering, bathing and other sporting activities because of effluent addition. Pollution detracts from the clean image New Zealand uses to remain competitive in marketing produce and tourism.

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[1] USEPA = United States Environmental Protection Agency

1.2 Requirements of Local Authorities

The statutory framework for managing water quality and restricting discharges to water is the Resource Management Act 1991 (RMA). Regional Councils are the authorities which implement the RMA. Regional Council's functions as described by section 30 of the RMA are the controlling authorities for all activities involving discharges to water, land or air. Under section 15 of the RMA every person who discharges a contaminant into water is required to obtain a discharge permit unless the discharge is specifically allowed by a rule in Regional plan or by regulation from the Regional Council (Forsyth, 1996; RMA 1991)

Every person has the duty to avoid, remedy or mitigate adverse effects into the environment under section 17 of the RMA (RMA, 1991).

New Zealand has twelve Regional Councils and four Unitary Authorities to administer the resource consent process and monitor consent compliance. Almost all of these Councils are preparing or have prepared a regional plan which deals with water management.

Regional plans must be prepared in accordance with the statutory requirements of the RMA. These plans must not be inconsistent with policies or plans from the Ministry for the Environment (MfE) or the RMA. Councils are required to consult with communities during the development of plans, and resolve public submissions through public hearings.

Under section 70 of the RMA, rules about discharges, a Regional Council can allow the discharge of effluent to be a permitted activity if it is satisfied that the discharge of effluent will not produce any of the following effects after reasonable mixing (Forsyth, 1996, RMA, 1991):

(1)(c) *The production of any conspicuous oil or grease films, scums or foams, and floatable or suspended material*

(d) *Any conspicuous change in the colour or visual clarity:*

(e) *Any emission of objectionable odour:*

(f) *The rendering of fresh water unsuitable for consumption by farm animals:*

(g) *Any significant adverse effects on aquatic life.*

There are two ways that consent authorities can meet their obligations under the RMA. One is to apply minimum standards on the effluent quality at the point of discharge. The other is to apply minimum standards on receiving water quality and require the consent holder to take responsibility for meeting those standards (Forsyth, 1996). To calculate receiving water quality a mass balance approach is required. This means that effluent quality can vary during the year and depending on the in stream conditions still comply with consents.

Discharging effluent to land is seen by most Regional Councils as the most acceptable disposal system, the primary concerns are; nitrate contamination of ground water, wastewater runoff to surface water, and objectionable odour beyond the property boundary.

Land application of effluent is a permitted activity in some regions and controlled or discretionary activity in others (Manawatu Wanganui Regional Council, 1994; Environment Waikato, 1994). All regions have placed restrictions on the N loading of the soil. The Manawatu Wanganui Regional Council has set a loading rate of 150-200 kg/ha.yr depending on the farm location(Manawatu Wanganui Regional Council, 1994). Southland Regional Council and Environment Waikato has a loading rate of 150 kg/ha.yr for a grazed pasture system; (Southland Regional Council, 1996; Environment Waikato, 1994).

The general view that land irrigation is more environmentally friendly has led many dairy farms to change from discharging pond treated effluent into water to land application. In the Waikato region 1800 farms have changed their system in the last 2 years (Selvarajah, 1996). In Southland only 25 out of 400 dairyshed discharge consents were for discharge of dairy effluent to water (Southland Regional Council, 1996).

1.3 Current Farm Treatment Systems.

1.3.1 Ponds

Dairyshed wastewaters in New Zealand have commonly been treated by pond systems consisting of an anaerobic pond followed by facultative (or aerobic) pond. An estimate in 1985 indicated that approximately half of the 14,000 dairysheds in New Zealand used pond systems to treat their effluent (Hickey & Quinn, 1992). Due to pond overloading and the increased awareness of the possible adverse effects from nutrients have on aquatic environments, the level of treatment provided by pond systems is now inadequate to safeguard the quality of many of New Zealand streams and rivers. High dilutions are required for waterways to safely assimilate these discharges (Sukias, 1996).

Pond system design has been based on standard design guidelines originally produced by the Ministry of Works. These utilised loading rates of 200- 280 kg-BOD_5/ha.d for anaerobic and 84 kg-BOD_5/ha.d for aerobic or facultative ponds (MAF, 1994). The appropriateness of these loading rates can be questioned when their history is considered. The anaerobic loading rate is from North American experience and the aerobic loading is based on 70 % removal of BOD_5 from the effluent by the anaerobic pond (MAF, 1994). Pond systems were primarily designed to remove BOD and suspended solids (SS), nutrients were not considered.

Anaerobic dairy ponds generally work well at their primary function of removing suspended solids and BOD. In contrast, facultative dairy ponds appear to provide lower levels of treatment than originally anticipated. This is attributed to pond overloading caused by actual loading rates being higher than the designed loading, or the design loading rate being to high for the amount of natural oxygen supply (MAF, 1994). The supply of oxygen may be reduced due to inadequate wind mixing and reduced algal photosynthesis from poor light penetration (MAF, 1994; Sukias, 1996).

There is considerable variation in effluent quality discharged over time and between ponds. In a study of dairy farm effluent treatment ponds the BOD concentration, a combined median for 11 sites monitored for at least 1 year was 98 mg/l with a 3 fold range of individual pond medians. The suspended solids (SS) concentration median was 198 mg/l with a 9 fold range. Available nutrients levels were high with 12.2 mg/l phosphorus and 75 mg/l for ammonium (Hickey et al, 1989).

There is enhanced awareness of the environmental impacts from contaminants in dairy wastewaters other than BOD and SS, such as ammoniacal nitrogen, phosphorus, pathogenic microorganisms, light attenuating materials and malodorous substances. This awareness has created a need for improvements to the present treatment systems if pollution from these contaminants is to be minimised (Environment Waikato, 1994).

The concern about pond treated effluent nutrient levels has led Environment Waikato to make discharging dairyshed effluent to surface water a discretionary activity (Environmental Waikato, 1994). Discharging large quantities of nitrogen into waterways from oxidation ponds is considered ultimately unsustainable (Environment Waikato, 1994). Other Regional Councils are also restricting discharges to water and are promoting sustainable discharges to land (Manawatu Wanganui Regional Council, 1995; Southland Regional Council, 1996).

1.3.2 Land application

Environment Waikato considers there are significant potential benefits to the environment from land treatment for disposal of effluent. The key benefit of land treatment and disposal is that it permits the management and conservation of nitrogen on farms. Land application of effluent enables nitrogen to be recycled and reused. Good nitrogen management can reduce nitrogen fertiliser applications and improve pasture growth, while avoiding the problems of weed and algae growth in streams that oxidation pond discharges can cause (Environmental Waikato, 1994).

Many land treatment systems have had problems with low nitrogen removal, ponding or overloading. The main cause of problems included; using an inadequate irrigation area (including poor chemical budgeting), poor distribution of wastewater, and seasonally high water tables (Environment Waikato, 1994). These problems can be avoided through good planning and farm management.

The area of land required for irrigation of wastewater can be determined by either, the allowable chemical loading rate, which is usually the nitrogen loading rate, or the allowable hydraulic loading rate when soil infiltration rates are more restrictive than chemical loading requirements.

Environment Waikato, (1994) suggest that the nitrogen loading rate should be determined according to the following;

1. the requirement to have a nitrate concentration which meet the New Zealand drinking water standard of 50 g NO_3 per cubic metre in the site drainage water which is discharged to the receiving environment;
2. the nitrogen removal by crop uptake (either agricultural or forest);
3. the nitrogen removal by denitrification;
4. the dilution afforded by ground water before flows enter the receiving environment;

5. other factors such as the form of nitrogen in the applied wastewater and seasonal changes in crop uptake rates.

1.4 Alternative Treatment Options

There are many different process configurations used for treatment of domestic and high strength industrial effluents. Many of these technologies would be applicable to treating dairyshed effluent, but the complexity and capital costs associated with some options makes them prohibitive for use on a small scale.

Three options that have shown particular suitability because of the low level of complexity, application on small scale and low operator attention are:

1. Upgrade present pond systems by adding a constructed wetland.
2. Rotating Biological Contactors as alternatives to ponds, and
3. Sequencing Batch Reactor treating combined raw and anaerobic pond effluent.

1.4.1 Constructed wetlands

Constructed wetlands are simple, natural treatment systems, that have the potential to provide cost effective, low maintenance tertiary treatment of dairyshed effluent (Tanner & Sukias, 1996). By utilising bacteria, plants and soils, constructed wetlands can reduce the polluting effect of many effluents.

In wetlands SS and organic material are removed through settling, filtration and microbial degradation. The hydraulic residence times are usually in the order of days allowing sufficient time for solids to settle. Settling and filtration may cause clogging of the wetland if influent SS levels are high. Because wetland can become blocked by

solid material, wetlands are not suitable for the treatment of primary effluent high in suspended solids or BOD as this can cause excessive biomass growth.

Sustainable nitrogen removal is achieved primarily through biological nitrification (NH_4 to NO_2 to NO_3) and denitrification (NO_3 to N_2). Another mechanism for removal is storage in sediments, detritus, and living biomass (Cooke et al 1992). As the internal environment is predominantly anaerobic, nitrogen removal is limited predominantly to denitrification of nitrate to N_2 gas, this anoxic process occurs if there is sufficient carbon available.

In wetland systems, nitrification is limited by low oxygen levels. There is limited dissolved oxygen because decaying organic material has a large O_2 demand. The oxygen needed to support the aerobic processes is supplied directly from the atmosphere via diffusion through the water-air surface, by photosynthetic O_2 production and a limited amount is transferred through the plant roots acting as pumps (Freeman, 1993; Knight et al, 1993).

Factors that affect removal of nitrogen in wetlands;

1. proportion of nitrogen incorporated in sediments,
2. the type of nitrogen species,
3. the amount of oxygen supplied,
4. the amount of available carbon and
5. the hydraulic retention time.

Phosphorus removal in wetlands is primarily through incorporation of the phosphorus into the sediments through adsorption, a small amount is initially included in plant material but as the growth rate stabilises this amount decreases. Removal by adsorption can initially be significant, however removal rates generally decline as soil sorption sites become saturated. Tanner and Sukias, (1996) found that net annual phosphorus removal decreased from 40% to 15% over a five year study.

1.4.2 Rotating biological contactors

Rotating biological contactors (RBCs) are a type of fixed film reactor developed by Bach and Imhoff in the 1930 from an idea conceived in Germany by Weigand in 1900. RBC can produce a superior effluent quality than trickling filters because of a lower organic loading per mass of biological solids, longer detention times in the biological stage and better control of short circuiting.

RBC consist of a series of slowly rotating partly submerged disks, about 40% of the surface area is submerged. During operation biological growths become attached to the surface of the disks and eventually form a slime layer over the entire wetted surface. The biomass alternately contacts the organic material in the water and air in the atmosphere. The water supplies dissolved oxygen and organic material which percolates through the biomass.

Disc rotational speed affects the amount of oxygen transferred, and the amount of biomass removed due to shearing forces created by the rotation. The removal of biomass prevents clogging of the media surfaces and maintains a constant thickness of microorganisms (Antoine, 1976). The ability of RBC to withstand hydraulic and organic surges is better than conventional activated sludge processes as most of the biomass is attached to the media. The sloughed sludge is a high density material which can be easily removed using a small secondary clarifier (Antoine, 1976).

The disadvantages of RBCs is that; a clarifier is required to remove the excess biomass, removal of phosphorus is limited to what is accumulated in the biomass for normal growth and a separate stage is required for denitrification where the media shaft is totally submerged. These systems are prone to mechanical breakage, the shaft failure is most serious. Other problems are media breakage, bearing failure and odour problems (Metcalf and Eddy, 1991).

1.4.3 Sequencing batch reactor

Sequencing batch reactor (SBR) is an activated sludge process that is operated on a fill and draw sequence. In an SBR all of the treatment processes are carried out in the same tank including sedimentation, this is different to a continuous flow treatment system where effluent comes into contact with different treatment processes in different tanks. The SBR treatment sequence is commonly fill, react (aerobic and anoxic), settle and decant treated effluent.

The advantages of SBR type systems over a continuous treatment system are; the reactor conditions can be changed quickly, and easily by adjusting the time sequence that the reactor follows, hydraulic surges do not cause washout of bacteria, and sludge age can be easily controlled by selecting how often sludge is wasted. The clarifying action is superior to a continuous flow secondary clarifier because conditions are completely quiescent. Settling times can be easily altered by changing the time allocated in the cycle.

Because of the ease of operation, the limited operator skill and the ability to remove nutrients from raw and primary treated effluent, SBR technology for the treatment of dairyshed effluent will be investigated further in this thesis. Homoda (1995) suggested that an SBR would be suitable for treatment of dairy farm wastewater in United Arab Emirates.

1.5 Sequencing Batch Reactor Technology

The sequencing batch reactor process is a variant of the activated sludge process. In activated sludge systems, aerobic micro-organisms in the suspended growth mode remove carbonaceous organic matter. By appropriate manipulation of the process conditions, nitrification, denitrification and phosphorus uptake may be achieved.

The activated sludge process was developed in the UK (Ardern & Lockett, 1914). It was named activated sludge because it involved the production of an activated mass of micro organisms, capable of stabilising waste aerobically. The original Ardern and Lockett process operated on a fill and draw basis. Each basin in the system was filled during a discrete period of time and then operated in a batch treatment mode. After treatment the bacterial flocs were allowed to settle and the clarified supernatant was withdrawn from the basin. Some bacteria remained in the reactor for treatment of the next batch. Excess bacteria were wasted to control the concentration of bacteria in the reactor. Subsequently due to the lack of control equipment available in the early 1920's and 30's the activated sludge process was developed into a continuous flow system (Norcross, 1992; Randall et al, 1992). There are numerous large scale applications of activated sludge around the world (Randall et al, 1992).

The SBR concept is based on the fill and draw type reactor, received renewed attention from the early 1970's onward, largely as a result of studies by R.L. Irvine and co-workers in the USA (Irvine and Davis, 1971; Irvine & Richter, 1978; Irvine et al, 1985). There was renewed interest in SBR technology because control technology had improved sufficiently to allow minimum operator attendance. Subsequently the SBR concept has generated considerable interest worldwide, to the extent that a first specialised international conference was held in March of 1996 to discuss SBR technology (IAWQ, 1996). Applications discussed included nutrient removal, treatment of industrial wastes, treatment of landfill leachate and transformation of hydrocarbons.

In activated sludge systems, the bacterial culture carries out the conversion of organic material forming bacterial cells, CO_2 and water, with the release of energy. Nutrients are required for cell formation and may also be released into solution during the process as shown by equation 1:

organic material + O_2 + nutrients →

→ CO_2 + H_2O + new cells + nutrients + energy 1.

1

Other reactions include the conversion of NH_4 to NO_3 (nitrification), the conversion of NO_3 to N_2 gas (denitrification) and phosphorus hydrolysis, which supports bacterial metabolism during anaerobic periods.

1.5.1 Process description

1.5.1.1 Cycle operation

SBR systems have five stages in common. They are carried out in the following sequence; (1) fill, (2) react, (3) settle, (4) draw, and (5) idle (Figure 1);

1.5.1.2 Description of the stages

(1) **Fill:** The purpose of the fill stage is to add substrate to the reactor. The fill increases the volume from the minimum to 100%. The time it takes to fill the reactor can effect the nutrient removal efficiency and reduction in coliforms and coliphages. As the fill time increases there is less time in the cycle available for the react stage (Sheker et al, 1993).

The fill stage may be static, mixed or aerated, depending on the treatment objectives. Static fill results in minimum energy input and high substrate concentration at the end of fill. Mixed fill results in denitrification of nitrates and reduction in oxygen demand due to carbon being used for denitrification. After denitrification is complete anaerobic conditions required for biological phosphorus removal may develop. An aerated fill extends the time available in the treatment cycle for aerobic reactions, a reduction in the react time is also

possible, and starting aerobic reaction keeps substrate concentrations low, which may be important if toxic materials are to be treated (Ketchum, 1996). A combination of these fill strategies can be used to tailor effluent treatment.

(2) **React:** During the react stage aerobic and anoxic reactions occur. This is the longest stage in the SBR cycle and when most of the nutrient removal is achieved. Because most of the nutrients are removed during the react stage, the time limits the final effluent concentration of the reactor effluent. To achieve nitrification-denitrification intermittent aeration is used so that aerobic and anoxic conditions occur.

(3) **Settle:** The settle stage allows solids separation to occur providing a clarified supernatant to be discharged as effluent. In an SBR this process is normally much more efficient than in a continuous flow system because the reactor contents are completely quiescent. The settling time typically ranges between 0.5 and 1.5 hr. This generally ensures that the sludge blanket remains below the withdrawal mechanism during emptying and minimises the risk of gas formation due to anoxic/anaerobic activity and consequent rising sludge, before the draw stage is completed.

(4) **Draw:** Clarified treated effluent is removed from the reactor during the draw stage. The withdrawal mechanism may take several forms. It may be as simple as a pipe fixed at the desired minimum tank volume or it can be a system of floating booms which carry a submersible pump so that effluent is withdrawn from just below the surface. This stops floating material being discharged. The floating system allows more flexibility in the operation of the plant and reduces the possibility of sludge being removed with the treated effluent. The time for draw can range from 5 to 30% of the total cycle time, however an overly extended time could cause problems with rising sludge.

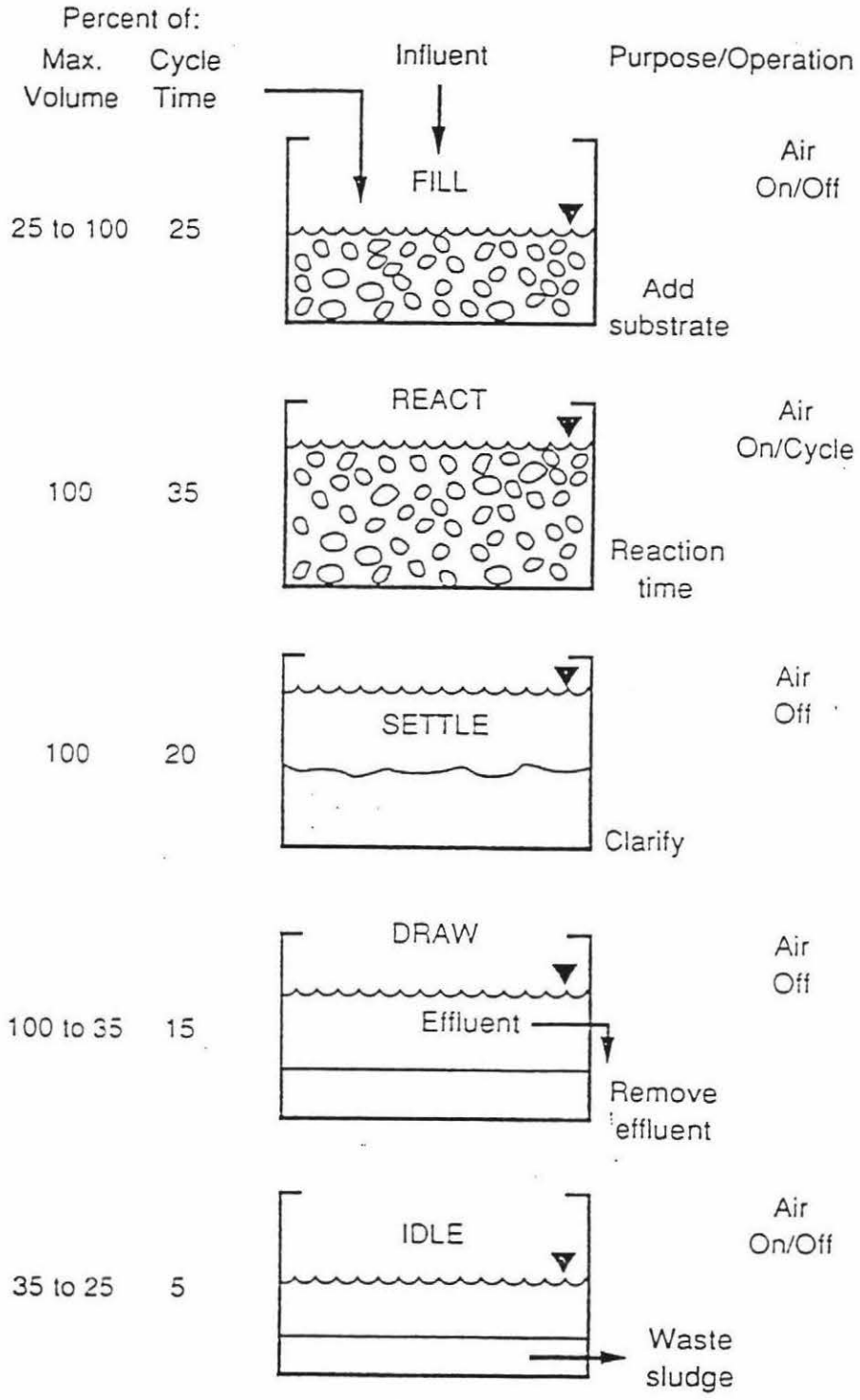

Figure 1 SBR operational sequence (Metcalf & Eddy, 1991)

(5) **Idle:** The idle stage in multitank systems is the time between draw and the next cycle fill. The purpose is to provide time for one reactor to complete its fill cycle before switching to the next unit. Idle is needed for flow equalisation and when there is uncertainty of flow rate fluctuations. Idle times can be near zero for peak design flows if the fluctuations are well defined (Ketchum, 1996). No idle period is included in the cycle for a single tank SBR.

Sludge wasting can occur during the react, settle, draw or idle stages. The sludge wasted during the react stage will be the least concentrated. The idle time is most often used because the sludge is at its most concentrated. Removal sludge during the idle stage requires separate equipment, whereas wasting sludge during the fully mixed react stage can be done using the decant pump. The frequency with which sludge is removed varies, depending on the required sludge retention time.

Table 1 illustrates common operating sequences to achieve different treatment objectives

Table 1 Common operating strategies (Ketchum, 1996)

Treatment objective	Fill Policies	React Policies
1. Organic carbon and suspended solid reduction. Minimum energy consumption or sludge production	Static, Mixed, then Aerated	Aerated
2. Organic carbon and suspended solids reduction, and nitrification	Static, Mixed, then Aerated	Aerated
3. Organic carbon and suspended solids reduction, and denitrification	Static, Mixed, then Aerated	Intermittent aerated, and mixing
4. Organic carbon and suspended solids reduction, and biological phosphorus reduction	Static, Mixed, then Aerated	Aerated
5. Industrial organic wastewater, toxic at high concentration	Mixed (short period) then aerated	Aerated (long period)

1.5.1.3 Advantages

The increasing popularity of SBR's may be attributed to the fact that they have several advantages over continuous flow systems. These include flow equalisation and attenuation of peak organic loads, both of which are inherent benefits of batch processes. Reactions that must be physically separate in continuous flow systems such as nitrification and denitrification can be carried out in the same tank by a single sludge.

SBR's are designed to operate under unsteady state conditions. Swings in the food to micro organism (F:M) ratio during the cycle, place selective pressures on the biomass. (The F:M ratio is defined as the ratio of BOD_5 concentration to mixed liquor volatile suspended solids (MLVSS) concentration in the reactor). During the fill period there is a high F:M ratio, leading to high BOD uptake rates. Conditions during this "feast"

period encourage the growth of floc forming bacteria. High F:M ratio conditions are less favourable for the growth of slow growing filaments which can cause sludge bulking in continuous flow systems. The selector concept employed to minimise bulking and poor settling characteristics in continuous flow systems operates on this principle. Sludge bulking is seldom a problem in SBR systems (Okada and Sudo, 1986; Randall et al, 1992).

1.5.2 SBR cycle requirements for nutrient removal

To achieve nutrient removal using SBR technology, conditions must be developed that selectively grow the necessary bacteria. The standard SBR sequence has an aerobic react phase, with no time in the cycle for anoxic or anaerobic reactions, apart from those which occur during the fill stage. By modifying the react stage so that the oxygen levels swing between aerobic and anoxic, nitrogen removal can be achieved. In order to create anoxic conditions during the react stage, mechanical mixing without aeration is used. This type of operation during the react stage can be described as intermittent aeration.

The time spent in aerobic conditions can be controlled by dissolved oxygen (DO) probes and timers. Anoxic conditions are controlled using oxidation reduction potential (ORP) probes. ORP will give information about when denitrification is complete and anaerobic conditions are starting to develop (Wouters-Wasiak et al, 1994). In an SBR system, denitrification may take place during idling, settling and fill phases, in addition to the phases with anoxic mixing.

Anaerobic conditions necessary for phosphorus removal are incorporated using mechanical mixing and should only occur once during any complete cycle. Anaerobic conditions usually occur during the fill stage or early in the react stage when readily available substrate concentration is the highest.

A suggested SBR sequence for nutrient removal is shown in Figure 2.

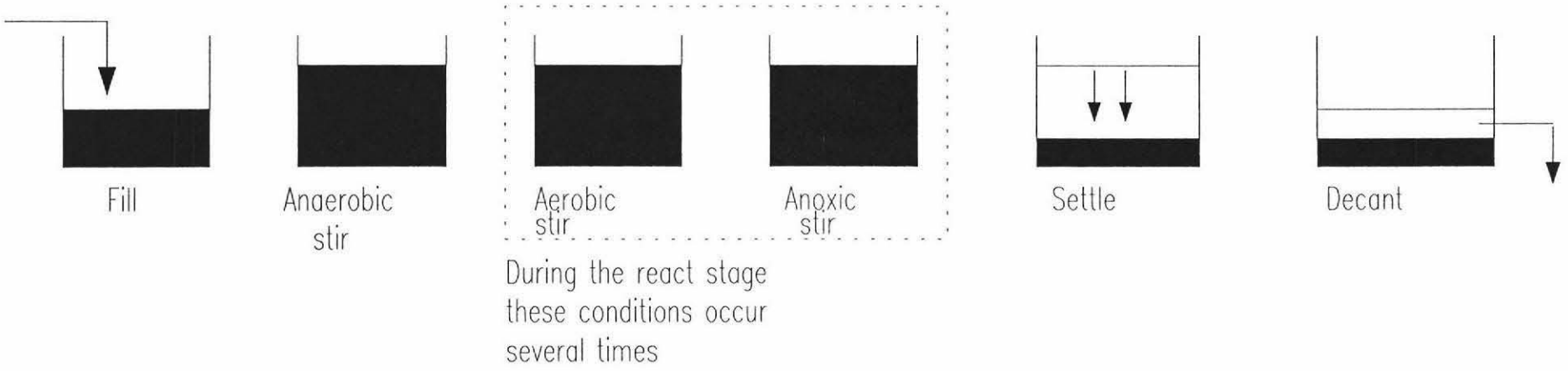

Figure 2 Suggested SBR sequence for carbon, nitrogen and phosphorus removal (Metcalf & Eddy, 1991)

1.6 Performance of SBR Technology

1.6.1 Domestic wastewater

Since Irvine's application of SBR technology in the 1970s and the improvements in control technology, the use of sequencing batch reactors for small and medium sized municipal treatment plants has gained widespread popularity (Rusten and Eliassen, 1993). Performance data shows that high levels of organic matter, nitrogen and phosphorus removals have been obtained under a range of conditions (Table 2).

From the early studies treating domestic wastewater and from subsequent work, R.L. Irvine and co-workers reported BOD removal from 160 g/m^3 to 10 g/m^3, ammonia removal from 25 g/m3 to 5 g/m^3 and phosphorus removal from 4.3 g/m^3 to 0.4 g/m^3 (Irvine & Davis, 1971; Irvine and Richter, 1978; Irvine & Ketchum, 1989). In a comparison of chemical and biological phosphorus removals, typical effluent, BOD, ammonia and phosphorus concentrations of 6 g/m^3, 15 g/m^3 and 0.4 g/m^3 respectively were reported, for the biological process (Ketchum, 1987).

SBR sewage treatment systems service lake shore settlements at two different locations in New Zealand. Median final effluent values for suspended solids of 11 g/m^3 and 13.5 g/m^3 and median total nitrogen levels of 7 $g\text{-}N/m^3$ and 17.5 g/m^3, were achieved. One facility had a higher effluent nitrogen level compared to the other, because of large fluctuations in population during holiday seasons, resulting in shock loads to the plant (Keys and Cooper, 1991). Phosphorus levels were not reported.

More recent reports have confirmed the ability of SBR systems to achieve high effluent quality when treating domestic wastewater (Imura et al, 1993; Rusten & Eliassen, 1993). SBR operation under cold climate conditions, temperatures ranging between 3°C and 8°C, was investigated by Marklund, (1993), mean effluent BOD_7 (seven day BOD) reductions from 124 g/m^3 to 23 g/m^3 and mean total phosphorus reductions from 6.0 g/m^3 to 1.6 g/m^3 were reported.

A review of treatment plants operating in Canada, United States of America and Australia revealed many different strategies for BOD and suspended solids removal (Arora et al, 1985). Of the eight facilities studied, seven produced effluent's with BOD_5 concentrations below 11 g/m^3 from daily flowrates up to 3,000 m^3. However, the operational strategy at each facility greatly effected power consumption. The power usage for the eight SBR plants ranged from 0.8 to 22.9 kWh/kg BOD applied. The plants using the least power achieved large removals of BOD during the anoxic stage.

Total and faecal coliforms may be removed during the react stage. Ng et al (1993) reported using a 2.5 hr react stage gave average removals from domestic sewage of 96% for total and faecal coliforms and 90% for coliphages.

A summary of SBR performance on domestic wastewater is shown in Table 2.

Table 2 Performance of SBR systems treating domestic wastewater

Parameter	Influent (g/m^3)	Effluent (g/m^3)	Removal (%)	Reference
COD	430	67	84	Rusten, 1993
BOD	-	3	-	Irvine, 1985[1]
	-	6	-	Irvine, 1985[2]
	162	6	96	Ketchum, 1987
	227	6.1	97	Imura, 1993
	151	8	95	Rusten, 1993[3]
	124	23	81	Marklund, 1993[3]
Nitrogen	160	10	94	Irvine, 1989
	150	8	95	Rusten, 1993
	34	6.3	81	Imura, 1993
Nitrogen (NH_3-N)	-	0.6	-	Irvine, 1985[1]
	-	1.1	-	Irvine, 1985[2]
	25	5	40	Irvine, 1989
Phosphorus	4.3	0.4	91	Irvine, 1989
	4.1	0.15	95	Imura, 1993
	6	1.6	73	Marklund, 1993
	6.8	< 1.0	> 85	Rusten, 1993

Note: 1 low loading

2 high loading

3 BOD_7 (seven day BOD)

1.6.2 Piggery wastewater

Treatment of piggery effluent has shown that SBR technology can successfully remove nutrients reduce nutrient concentrations in high strength wastewaters.

Laboratory scale studies on both raw and anaerobically pre-treated wastewater, using a 0.8 m^3 capacity aerobic SBR, were reported by Ng, (1987). Median total COD removals of 48% and 58% respectively and median soluble COD removals of 81% and 17% respectively were found for the two wastewaters. However, effluent COD figures were affected by relatively high suspended solids levels and by the accumulation of nitrite. Ng (1987) suggested that the difference in soluble COD reductions may also be due to the accumulation of difficult and non-biodegradable compounds resulting from the anaerobic process. Median removals of BOD_5 from 1167 g/m^3 to 182 g/m^3 for the raw wastewater and 456 g/m^3 to 80 g/m^3 for the anaerobic wastewater were reported, whilst levels of soluble effluent BOD_5 were 22 g/m^3 and 47 g/m^3. Typical reductions in ammonia from 350 g/m^3 to 55 g/m^3 and from 575 g/m^3 to 180 g/m^3 were achieved for treatment of the raw effluent and anaerobic effluent respectively.

In a pilot scale study, 1.4 m^3 of dilute swine wastewater was treated over 4 hours, with a reported reduction in BOD_5 from 640 g/m^3 to 56 g/m^3 (Lo et al, 1990). During the same period the COD was reduced from 3180 g/m^3 to 1335 g/m^3. Removal efficiencies for BOD_5 and COD were 91% and 42% respectively. Nitrification occurred but no denitrification was reported. Similar oxygen demand removals were reported for studies on piggery and dairy wastes, with a COD removal of 96%, from 3850 g/m^3 to 164 g/m^3 (Bortone & Piccinini, 1991). In this investigation nitrogen was reduced from 302 g/m^3 to 16 g/m^3 and phosphorus from 54 g/m^3 to 12 g/m^3.

High removal efficiencies for carbon, nitrogen and phosphorus were achieved treating swine wastewater in a bench scale study when using an intermittent aeration process (Osada et al, 1991). Total organic carbon (TOC) values were decreased from 552 g/m^3 to 26.5 g/m^3 and the corresponding BOD_5 concentrations calculated to be 1500 g/m^3

and 7.5 g/m^3 respectively. Nitrogen was reduced from 276 g/m^3 to 8.5 g/m^3 and phosphorus from 160 g/m^3 to 30.7 g/m^3. Corresponding removal efficiencies were as follows: BOD 99.5%, TOC 95.2%, TN 96.9% and TP 80.8%. The treatment efficiency decreased when the BOD to total nitrogen ratio decreased from 5.5:1 to 2.2:1. The largest drop was for TP which decreased to 19% removal. This study demonstrates the importance of carbon levels on the treatment efficiency.

High organic and nutrient removal rates have been reported for high strength piggery wastewaters, with reductions for; COD of 97%, nitrogen of 93% and ammonia of 97-99% (Fernandes et al, 1991; Fernandes, 1994). Nitrogen was decreased from 580 g/m^3 to 185 g/m^3 and a minimum effluent ammonia level of 2 g/m^3 was achieved.

Fernandes (1994) investigated the effect of temperature on the performance of an SBR treating piggery effluent. The removal efficiency for NH_4^+-N was 39.5%, 97% and 99.8% for reactors operating at 5°C, 10°C and 21°C, resulting in effluent ammonia nitrogen levels of 638 g/m^3, 34 g/m^3 and 2 g/m^3 respectively. The COD removal was 93-97% over the temperature range. Denitrification was achieved during the fill stage of the cycle, but as no anoxic stage was included in the sequence, nitrates accumulated in the effluent. To achieve greater denitrification during the react stage, intermittent aeration would be required.

Performance data for piggery wastewater are summarised in Table 3.

Table 3 Performance of SBR systems treating piggery wastewater

Parameter	Influent (g/m^3)	Effluent (g/m^3)	Removal (%)	Reference
COD	2,194	1,140	48	Ng, 1987[1]
	2,794	1,161	58	Ng, 1987[2]
	3,850	164	96	Bortone, 1991
	31,175	845	97	Fernandes,1991
	30,200	986	97	Fernandes,1994
COD (soluble)	2,028	391	81	Ng, 1987[1]
	982	817	17	Ng, 1987[2]
	1,500	7.5	99.5	Osada, 1991
BOD	1,167	182	84	Ng, 1987[1]
	456	80	82	Ng, 1987[2]
	640	56	91	Lo et al, 1990
BOD (soluble)	1,075	22	98	Ng, 1987[1]
	269	47	83	Ng, 1987[2]
Nitrogen	302	16	95	Bortone, 1991
	2,580	185	93	Fernandes,1991
	276	8.5	97	Osada, 1991
Nitrogen (NH_3-N)	350	55	86	Ng, 1987[1]
	575	180	69	Ng, 1987[2]
	1,126	34	97	Fernandes, 1994[3]
	1,265	2	99.8	Fernandes, 1994[4]
Phosphorus	54	12	78	Bortone, 1991
	160	31	81	Osada, 1991
	240	21.3	91	Osada, 1991

Note: 1 raw wastewater

2 anaerobically pre-treated wastewater

3 10 °C

4 21 °C

1.6.3 Dairyshed wastewater

Application of SBR technology for treatment of milking parlour effluent has been successfully demonstrated by Lo et al (1985; 1988), in both a bench scale experiment and a 200 litre pilot scale experiment, using cycle times from 3 to 6 hours. Oxygen demand removals of 78-85% for COD, concentrations of 633 g/m^3 reduced to 95 g/m^3 and, reductions of 86.5-95.5% for BOD concentrations of 260-270 g/m^3 reduced to 12-35 g/m^3, were reported. In the bench scale study (Lo et al, 1988) total nitrogen levels were reduced from 14.9 g/m^3 to 5.5 g/m^3, giving an efficiency of 63%. It is suggested that the reason for the poor nitrogen removals was that no separate anoxic stage was included in the treatment cycle. Denitrification may have been limited by the amount of available carbon. Subsequently, an effluent nitrogen level of 7.7 g/m^3 was obtained. Ammonia reductions to 1.6 g/m^3 and 2.9 g/m^3 were reported. In studies using the same type of effluent, COD was reduced from 919 g/m^3 to 155 g/m^3, BOD was reduced from 270 g/m^3 to 13 g/m^3 and ammonia nitrogen was reduced from 15 g/m^3 to 1 g/m^3 (Tam et al, 1986).

It should be noted that a conventional SBR process (Figure 1) was operated in these studies.

Available performance data for dairyshed wastewater are presented in Table 4.

Table 4 Performance of SBR systems treating dairyshed wastewater

Parameter	Influent (g/m^3)	Effluent (g/m^3)	Removal (%)	Reference
COD	633	136	78.5	Lo, 1985
	919	155	83	Tam, 1986
	975	144	85	Lo, 1988[1]
BOD	260	35	86.5	Lo, 1985
	270	13	95	Tam, 1986
	260	12	95.5	Lo, 1988[1]
Nitrogen	89	20	78	Lo, 1985[2]
	15	1	93	Tam, 1986
	21	7.7	63.5	Lo, 1988[1]
Nitrogen (NH_3-N)	16.6	2.9	82.5	Lo, 1985
	21	1.6	92.5	Lo, 1988[1]
Phosphorus	-	-	-	

Note: 1 21.8 °C
2 TKN

1.6.4 Anaerobically pretreated wastewater

Subramaniam et al (1994) investigated treatment of high strength wastewater from a abattoir using an SBR. Effluents had received two levels of prior anaerobic pretreatment. It was found that for the effluent with the highest level of anaerobic pretreatment, the performance of the SBR for nutrient removal was poor, although 70% of the COD was removed. There was considerable inhibition of nitrification, and no denitrification occurred during the aerobic stage, although simultaneous nitrification-denitrification it did take place in the SBR treating the stronger effluent. Phosphorus removal was poor and this was reported to be related to the limited availability of readily biodegradable COD (RBCOD) in the wastewater. The high NOx-N concentrations at the beginning of the cycle also meant that there was competition between the denitrifying bacteria and the phosphorus removal bacteria for the RBCOD.

The effluent which had received a lesser degree of anaerobic treatment had C:N ratios ranging from 2.3:1 to 8.3:1. This effluent was readily treatable, with COD, NH_3 and PO_4-P reductions of 94.9% 98% and 98% respectively. The carbon to nitrogen ratio of the more highly pre-treated wastewater ranged from 0.71:1 to 1.7:1, which is well below that recommended by Narkis et al (1979).

In a similar study using a dairy factory effluent which was rich in phosphorus (60 to 100 mg P/L) two different amounts of pretreatment occurred before the effluent was treated using a SBR. (Comeau et al, 1996).

The effluent which received only prefermentation had a COD concentration of 1550 mg/l, which contained 1230 mg/l VFAs measured as HAc. This effluent was treated to a higher level than an other effluent which had been treated using an upflow anaerobic sludge blanket reactor (UASB) to a COD concentration of 500 mg/l.

The effluent concentrations from prefermented effluent and UASB pretreated effluent for were phosphorus 6mg/l and 40 mg/l, for COD 40 mg/l and 150 mg/l respectively (Comeau et al, 1996).

These research results demonstrate that a readily biodegradable carbon source available in the prefermented effluent is required for biological nutrient removal.

The data for anaerobically treated effluent is shown in Table 5.

Table 5 Performance of an SBR system treating anaerobically treated wastewater

Parameter	Influent (g/m^3)	Effluent (g/m^3)	Removal (%)	Reference
COD	3500-6000	100-200	97	Subramaniam, 1994
	1500	150	90	Comeau et al, 1996
COD (soluble)	600-1500	-	-	Subramaniam, 1994
BOD	-	-	-	
Nitrogen	180-250	2-5	98	Subramaniam, 1994
	15	0.6	96	Comeau et al, 1996
Phosphorus	30-50	< 0.5	98	Subramaniam, 1994
	60	6	90	Comeau et al, 1996

1.7 Nutrient Removal Considerations

1.7.1 Phosphorus

To remove phosphorus from wastewater it is necessary for it to be accumulated within bacterial cells. By subjecting bacteria to alternating anaerobic and aerobic conditions, a bacterial population is selected that is capable of taking up phosphorus in levels beyond the stoichiometric requirements for growth and storing it as polyphosphates. The amount of accumulated phosphate in the sludge can be as high as 8% of the dry weight of cells (Comeau et al, 1996).

Biological phosphorus removal is based on the following two facts (Rodrigo et al, 1996):

1) In the anaerobic zone, phosphorus is released from the microorganisms. Accompanying the release, organic substrate (acetate and other short chain fatty acids) is taken up and stored intracellularly, most commonly as polyhydroxyalcanoates (PHA).

2) In the aerobic-anoxic zone storage products (PHA) are oxidised. At the same time, soluble orthophosphate provides for the resynthesis of intracellular polyphosphates.

Phosphorus removal requires small organic substrates which are stored in the cell. From the literature it is not clear what amount of COD is reduced during phosphorus release. Using the data presented by Rovatti et al (1995) for nine experiments it is possible to calculated a COD to phosphorus release ratio of 11.3.

The removal of phosphorus is linked to organic carbon uptake which is pH dependent. Acetate is not metabolism below pH 5.0. An optimum pH of 6.8±0.7 was proposed by Liu et al (1996) because relatively high acetate uptake rates with less energy consumption can be maintained by the bacteria.

To remove phosphorus it is necessary to waste a proportion of the sludge before the microorganisms present encounter anaerobic conditions again. Since phosphorus removal is achieved by removing excess sludge that has taken up soluble phosphorus from the wastewater, solids retention time (SRT) is one of the most important parameters in enhanced biological phosphorus removal system

Workers investigating phosphorus removal have found that SRTs of 10 days gave the highest removal (Choi et al, 1996; Rodrigo et al, 1996). Nitrogen removal efficiency decreases at aerobic SRTs below 12 days. Therefore in a combined system, nitrogen removal efficiency will need to be considered along with phosphorus removal when selecting a SRT.

For phosphorus removal, hydraulic retention times of 1.8-3 hr in the anaerobic zone and 1-4 hr in the aerobic zone are reported (Metcalf & Eddy, 1991). There is a critical time for the anoxic/anaerobic period in the phosphorus removal process. This was found for a synthetic wastewater to be 2 hr in work carried out by Okada & Sudo (1986).

It was also noted by these workers that true anaerobic conditions did not occur in the presence of high nitrate and nitrite (NOx-N) concentrations in the reactor, thus disturbing the release of phosphorus. Other workers found that as the ratio of total kjeldahl nitrogen (TKN) to total phosphorus increases there is a decrease in the amount of phosphorus removed (Choi, 1996). This may be due to increased nitrate concentration in the reactor, but if nitrogen species are fully removed before the anaerobic stage then this effect may not be apparent. It has also been found that if excess phosphorus is present the amount of phosphorus removed decreases considerably (Converti et al, 1995).

In recent experimental work (Vlekke, 1988; Kuba et al, 1996) dephosphatation was found to occur under anoxic conditions by denitrifying bacteria using nitrate as an electron acceptor. The anaerobic metabolism of denitrifying dephosphatation bacteria

(DPB) is identical to conventional anaerobic-aerobic phosphorus removing organisms. In the enriched DPB sludge, intracellular glycogen is utilised under anaerobic conditions for PHB (poly-β-Hydroxybutyrate) synthesis from HAc (acetic acid).

The main difference between anoxic metabolism by DPB and conventional aerobic metabolism is that electron transport phosphorylation uses nitrate oxygen and not molecular oxygen. The measured $P/NADH_2$ ratio in the electron transport phosphorylation with nitrate was approximately 1.0 mol- ATP/mol-$NADH_2$, which indicated that the anoxic energy production efficiency is approximately 40% less than the aerobic efficiency (Kuba et al, 1996).

In a five month period the effect of oxygen on the rate of dephosphatation was investigated. It was concluded that the presence of oxygen was not detrimental to the denitrifying dephosphatation activity (Kuba et al, 1996).

Kuba et al (1996) suggest that ideal conditions for phosphorus and nitrogen removal by DPB and nitrifiers in single sludge activated sludge systems should include the following;

Anaerobic Phase	- no COD carry over to the aerobic / anoxic phases;
	- no nitrate input from the anoxic phase;
Aerobic phase	- maximal nitrate production by nitrifiers;
	- minimal phosphorus uptake, which means minimal PHB oxidation inside DPB;
Anoxic phase	- complete phosphorus uptake and nitrate utilisation;
	- optimal NH_4^+ utilisation by growth of DPB.

These situations can promote maximum selection for DPB because the growth of other bacterial groups such as heterotrophic bacteria and nonpolyphosphate accumulating denitrifiers is prevented.

1.7.2 Nitrogen

Removal of total nitrogen is achieved partly by denitrification and partly by withdrawal of organic nitrogen contained within biomass and particulate material. Prior to denitrification, organic and ammonia nitrogen must be converted to nitrate by nitrifying organisms.

The nitrification is a two step reaction, Nitrosomonas bacteria oxidise NH_4^+ to NO_2^- and Nitrobacter bacteria convert NO_2^- to NO_3^-. The process is influenced by factors such as solids retention time (SRT), temperature, pH, dissolved oxygen and the carbon to nitrogen (C:N) ratio of the effluent. At a given temperature a minimum aerobic SRT is necessary for complete nitrification.

Relatively long SRTs or sludge ages are required for nitrification due to slow growth rates of nitrifying organisms. For domestic wastewater treatment systems using plug flow reactors SRT values of 8-20 days are reported for nitrification, whereas 1-5 days are indicated for denitrification (Metcalf & Eddy, 1991). However, a critical SRT range of 4-6 days for nitrification was found for an SBR reactor treating domestic wastewater. At a temperature of 10°C, a SRT of 12-13 days was necessary (Rusten and Eliassen, 1993). Typical hydraulic retention times for nitrification and denitrification range between 6-15 hr and 0.2-2 hr respectively (Metcalf & Eddy, 1991).

Oxygen requirements for nitrification can be significant in comparison to those for carbonaceous oxidation, with a stoichiometric demand of 4.57 g O_2/g NH_3-N, the actual demand is approximately 4.3 g O_2/g NH_3-N oxidised to nitrate (Metcalf and Eddy, 1991). Dissolved oxygen levels should be maintained above 1.0 g/m^3 for satisfactory nitrification performance. Anoxic conditions for denitrification may be sustained using intermittent aeration.

Denitrification rates decrease as oxygen concentration increases. At 1mg/l DO the rate of denitrification is zero. It has been found that if the reactor DO is about 0.5 mg/l, the

rate of nitrification equals the rate of denitrification. Simultaneous nitrification and denitrification would lead to shorter react times and simpler control strategies (Munch et al, 1996).

Temperature is an important factor as it influences the rate of enzymatically catalysed biological reactions and the rate of diffusion of substrate into the microbial cell. Nitrification and denitrification reactions are particularly temperature sensitive. A marked decrease in nitrification performance may occur below 10°C. The optimum temperature for nitrifying bacteria is in the range 28-36°C (Fernandes, 1994).

Sufficient alkalinity should be present to buffer the consumption occurring during nitrification, estimated at 7.14 g $CaCO_3$/g NH_3-N (Metcalf & Eddy, 1991). Alkalinity is however liberated in the denitrification process. From work using an aerated biofilter for nitrogen removal the average alkalinity requirement was estimated at 4.3 mg of alkalinity as $CaCO_3$ per mg N removed (Chui et al, 1996). This was higher than the theoretical value of 3.57 mg. A reason for the difference is that denitrification may not have been complete, reducing the amount of alkalinity released. Optimal pH ranges are reported for nitrification at 7.2-9.0 and denitrification at 7-8 (Metcalf & Eddy, 1991).

Biological denitrification has been shown to be a reliable and cost effective method of nitrogen removal from domestic wastewater (Abufayed & Schroeder, 1986).

Biological denitrification involves a respiratory mechanism in which nitrates replace molecular oxygen as the hydrogen acceptor in anaerobic systems. The nitrate is reduced to nitrite and subsequently to nitrous oxide and dinitrogen gas. These reactions are carried out under anoxic conditions by heterotrophic facultative bacteria in the presence of an energy and organic carbon source (Abufayed & Schroeder, 1986). The rate of denitrification is dependent on temperature, the NOx-N concentration and the type and concentration of the carbon source (Rusten and Eliassen, 1993).

Denitrification reaches a minimum at 0°C and a maximum at 40°C.

A lot of time has been spent investigating the type of carbon available for denitrification with the aim of reducing the amount of supplementary carbon added to the process (Alleman and Irvine, 1980; Demuynck et al, 1994; Surmacz-Gorska et al, 1995). Methanol, acetate, ethanol, and other readily biodegradable carbon sources have been used previously but due to increasing costs other carbon sources are required (Demuynck et al, 1994; Fang and Ling, 1995). Internal carbon sources are mainly raw or settled wastewater, storage induced carbon and endogenous respiration of activated sludge organisms are useful alternatives.

Use of internal carbon sources can limit the overall nitrogen removal efficiency, as raw wastewater introduces carbon, ammonium, and organic nitrogen. Using endogenous carbon results in very low denitrification rates and consequently larger reactor volume requirements. Workers investigating the kinetics of denitrification reactions found that the rate of denitrification using readily biodegradable carbon was 0.25 mg NO_3-N/mg VSS·d, while the rate decreased to 0.07 mg NO_3-N/mg VSS·d when endogenous carbon was the carbon source (Carucci et al, 1996). These problems have been overcome to some extent by exposing the sludge to alternating aerobic and anoxic environments. (Abufayed & Schroeder, 1986; Surmacz-Gorska et al, 1995).

Table 6 illustrates typical denitrification rates for various carbon sources.

Table 6 Denitrification rates for different carbon sources

Carbon Source	Denitrification rate Kg NO_3-N/Kg VSS·d	Temperature °C	Reference
Methanol	0.21-0.32	25	Metcalf & Eddy, (1991)
Methanol	0.12-0.90	20	Metcalf & Eddy, (1991)
Wastewater Readily biodegradable fraction	0.03-0.11 0.25	15-27 20	Metcalf & Eddy, (1991) Carucci et al, (1996)
Endogenous metabolism	0.017-0.048 0.07	12-20 20	Metcalf & Eddy, (1991) Carucci et al, (1996)

1.7.2.1 Production of N_20 during nitrification and denitrification

Products of denitrification may include N_2, NO and N_2O gases. It is important to minimise nitrous oxide (N_20) production since it is both a greenhouse gas and toxic to humans. N_2O as a green house gas is as important as carbon dioxide (CO_2). The concentration of N_2O in the atmosphere is currently much lower than CO_2. However, due to its long lifetime of 130 years, the 100 year spanned greenhouse effect of N_2O is 290 times greater than that of CO_2 on a weight basis (Zheng and Matsuo, 1996).

N_2O is a by product in nitrification and an intermediate product in denitrification. In denitrification, operational conditions such as low C:N ratios of wastewater and short SRT promote N_2O production (Zheng and Matsuo, 1996), as do low concentrations of DO and high nitrite concentrations (Schulthess et al, 1995). Wicht (1996) reported that lower COD concentrations lead to lower N_2O concentration and decreasing pH values lead to an increase in N_2O, not because of the pH dependency of the kinetics but of inhibition by nitrous acid.

Schulthess et al (1995) reported that during anoxic denitrification when the rate of nitrate reduction was approximately 50 gNO_3d^{-1} and nitrate was the first electron acceptor, little nitrite accumulated and almost no NO and N_2O was seen. When supplementary carbon was added and the rate of nitrate reduction increased some nitrite, NO and N_2O accumulated (Schulthess et al, 1995). In systems that have low amounts of carbon available during nitrification and denitrification N_2O is produced.

N_2O concentration rapidly decreases after carbon addition indicating N_2O can be utilised as a electron acceptor in denitrification (Zheng and Matsuo, 1996). These results suggest that in a combined nitrification-denitrification system, nitrifiers behave as producers of N_2O and denitrifiers play a role of both producer and consumer.

N_2O may also be removed from the system before conversion to N_2 gas by air stripping (Schulthess et al, 1995). Zheng and Matsuo (1996) observed that when the NH_4-N loading was doubled the percentage N_2O produced increased from 1% to 5%. But at the same time the fresh air flowrate was doubled so the observed increase may have been because of increased air stripping. This may also be the reason why in the continuously aerated reactor greater than 40% of the reduced nitrogen was converted to N_2O gas.

Other researchers found that N_2O emissions from continuous and SBR treatment plants were comparable. There was very little difference between the 2 systems, 9-13 % of the influent TKN was lost as N_2O (Willers et al, 1996). This was not the case for ammonia

emission. The average emission of ammonia from the aeration tank surface of the SBR was roughly three times that from the combined surface of the nitrification and denitrification stages in the continuous system. An explanation for this difference can be attributed to higher ammonium concentrations during the fill stage and initially in the react stage, whereas in the continuous system, peak concentrations are absent.

1.7.3 Carbon requirements for nutrient removal

In order for both nitrogen and phosphorus removal to occur, sufficient carbon must be present in the wastewater. The critical C:N ratio for denitrification has been reported as 2.3 g-BOD_5/g NOx-N (Narkis et al, 1979). This carbon must also be present in a form which is readily available to the bacteria. If the amount of carbon available is insufficient denitrification will not occur.

The rate of phosphorus release is dependent on the anaerobic rate of organic carbon uptake and storage as PHA (Liu, 1996; Rodrigo, 1996). It has been reported for one biological process, that in order to achieve effluent phosphorus levels less than 1 g/m^3, a BOD_5 to POx-P ratio greater than 10 is required (Metcalf & Eddy, 1991). The literature is not clear about the actual amount of soluble carbon reduced during phosphorus release, it has been reported at 1.2 g-BOD_5/g POx-P (Fang & Ling, 1995) and 13.8 g COD / g POx-P released (Rovatti et al, 1995). Researchers using acetate as the carbon source reported the ratio to fluctuate between 2:1 and 5:1 (Wentzel et al, 1985).

1.7.3.1 Carbon characterisation

As more complicated treatment process are introduced to meet stringent effluent criteria more information is required, so that the process can be optimised using mathematical models. Such models require a detailed knowledge of the carbon fractions present in the wastewater. The organic matter characterisation is the fractionation due to the rate of degradation. Specific compounds are not of primary interest, except VFAs and some

low molecular weight organics. These simple organic molecules can inmost cases be metabolised directly. Knowledge about the amount of carbon in each fraction can help predict the rate of phosphorus and nitrogen removal.

The different carbon fractions that are often used for modeling activated sludge include; inert soluble organics, readily biodegradable organic matter, rapidly hydrolysable organic matter, slowly hydrolysable organic matter and inert suspended organics (Henze, 1992).

Inert soluble organics are not transformed during the treatment process. They may also be produced from decaying bacteria.

Inert soluble organics can be estimated by determination of soluble total COD minus the soluble ultimate BOD (Henze, 1992). The soluble ultimate BOD is approximated by 1.5 time the soluble 5 day BOD.

$$\text{Soluble inert carbon} \approx S_{TCOD} - 1.5 * S_{BOD_5} \qquad 2.$$

Where S_{TCOD} = Total soluble COD

S_{BOD5} = Soluble five day BOD

It is useful to know the amount of inert material as it determines the minimum COD of an effluent after biological treatment.

When determining the inert fraction, BOD measurements must be done with suppressed nitrification. Nitrification is suppressed because bacteria can oxidised ammonia for growth. Ammonia oxidation will increase the value of BOD measured by approximately 4.57 mg O_2/ mg NH_3.

Readily biodegradable organic matter contributes 10-15% of the total COD of raw domestic wastewater (Henze, 1992). This fraction of directly metabolised organics is limited to small molecules of volatile fatty acids, carbohydrates, alcohols, peptide's and

amino acids. The volatile fatty acids, especially acetic acid, can be metabolised at a high rates under aerobic and anoxic conditions. Under anoxic conditions readily biodegradable carbon is used for denitrification (Carucci et al, 1996).

Readily biodegradable carbon is necessary in activated sludge plants which remove phosphorus biologically. Under anaerobic conditions this fraction is stored intracellularly as poly-β-hydroxybutyrate (PHB) and polyhydroxyvolerate (PHV) using energy from the release of poly phosphates (Henze, 1992; Rodrigo et al, 1996).

Measurement of readily biodegradable carbon can be achieved by direct measurement of specific compounds, physical-chemical procedure to separate solid material from soluble material and biologically using oxygen uptake rates.

Direct measurement involves testing for small specific compounds like VFAs, ethanol and glucose. These small compounds that can be transported directly in the cell, can account for 50-70 % of the readily biodegradable COD in raw municipal wastewater (Henze, 1992).

Physical-chemical method involves flocculation and precipitation of colloidal material before filtration through a 0.45 μm membrane filter. By flocculating small colloidal material, filtration is enhanced producing a filtrate containing truly soluble organic matter (Mamais et al, 1993).

One biological method uses a single oxygen uptake rate measurement. The oxygen consumption is related to readily biodegradable COD using a calibration curve. The calibration curve is obtained by measuring the oxygen consumption vs. acetic acid COD (Xu and Hasselblad, 1996). This method is affected by the bacterial population and experimental conditions used. Unless conditions are similar comparisons between different plants maybe limited.

The physical methods are not reliant on biological reactions and are therefore easily controlled and reproduced.

Rapidly hydrolysable organic matter can account for 15-25 % of total COD of raw municipal wastewater. Hydrolysis under aerobic conditions is rapid and will be finished within a few hours. Henze (1992) defines the rapidly hydrolysable organics as;

$$S_{hydrolysable} = S_{tCOD} - S_{inert} - S_{readily} \qquad 3.$$

It may also be possible to estimate the BOD of the readily biodegradable fraction and the rapidly hydrolysable organic matter using respirometric BOD measurements by assuming a reaction time and taking the BOD at that time. Eckenfelder (1989) suggests that most of the readily degradable substrates are consumed in the first 24-36 hours.

Slowly hydrolysable organic matter makes up 40-60% of the total COD in raw domestic wastewater Henze (1992). The rate of biological aerobic as well as anoxic degradation of slowly hydrolysable organic matter is limited by hydrolysis. This fraction can be determined from the respirometric BOD measurements after a selected time. A COD mass balance may also be used if all the other fractions are known.

Inert suspended organics are not degraded during the treatment process and contribute to the influent and effluent COD concentrations. The effluent concentration may be higher than influent as some inert suspended organics are produced from biomass decay.

1.8 Literature Review Summary

It is apparent that present dairyshed effluent treatment systems are not capable of complying with regulations generated by Regional Councils implementing the RMA. Many pond systems are over loaded from increased herd sizes using existing treatment facilities. Due to overloading many pond treatment systems do not reduce effluent nutrient levels enough, to avoid adverse environmental effects.

Under the statutory framework of the RMA Regional Councils have the role of protecting and improving the environment we live in. This is achieved by regulating waste producers. As this applies to dairy effluent , there is now a general view that land irrigation of dairy waste is the most environmentally sustainable way of disposing of the wastes produced on a dairy farm. This view has developed because of limited alternative treatment practices.

Present farm treatment systems can be classified into pond systems and land based treatment systems. The two pond systems were design primarily to remove BOD and SS. The effluent produced is often variable in quality in terms of BOD and SS and are high in nitrogen and phosphorus.

At present there are few alternatives to the use of ponds or land application for treatment of dairy effluent. An alternative that has been applied to a large number of different effluent types is the sequencing batch reactor.

Sequencing batch reactors are a type of activated sludge which is capable of reducing nutrient levels using a single tank. During a cycle different environmental conditions are achieved which promote the growth of specialised bacteria capable of nitrification, denitrification and luxury phosphorus uptake.

The requirements for nutrient removal using a sequencing batch reactor are; carbon species must be readily available for bacterial degradation, the ratio of carbon to nitrogen should be in excess of 2.3:1 mg BOD per mg NO_3-N. Available alkalinity exceeding 7.14 g $CaCO_3$ per gNH_3-N oxidised, pH in the range 7.2-9.0 for optimal growth of nitrifying bacteria and the sludge age greater than 12 days for growth of nitrifying bacteria.

Phosphorus removal is achieved by bacteria absorption during the aerobic period of the cycle. A proportion of the bacteria are then wasted from the system before anaerobic conditions are in counted again.

Sequencing batch reactors have been applied to the treatment of domestic effluent, results indicated good nutrient removal, 95% removal for nitrogen and phosphorus was achieved. Similar removal efficiencies were achieved treating piggery effluent and some high strength industrial effluents.

Application of SBR technology for the treatment of dairy effluent was limited to the use of a predominantly aerobic cycle for BOD reduction. Limited research has been under taken using SBR's to remove nutrients from dairyshed effluent. There is a need for research to determine the reactor volumes, and anaerobic, anoxic, and aerobic cycle times required for removal of nitrogen and phosphorus from this type of wastewater.

2. Wastewater Characterisation

2.1 Introduction

Dairyshed effluent contains considerable polluting potential if not treated properly.

Wastewater is generated daily when cows are brought to the dairyshed for milking. The high concentration of animals held in one place produces large volumes of waste. Treatment to reduce the polluting potential of the waste is required before the effluent can be safely discharged.

In order to develop effective treatment systems it is necessary to have an understanding of the nutrient levels in the waste and other wastewater characteristics such as pH, and available alkalinity. pH and alkalinity influence the rate of nitrification whilst carbon concentration effects the rates of phosphorus removal and denitrification.

Enhanced biological removal of phosphorus and nitrogen is a complicated treatment process because of the need for changes in the oxygen environment and readily biodegradable carbon. For such processes to operate reliably and meet stringent effluent criteria, information is required about the process variables, so that the process can be optimised. Process optimisation involves understanding the reactions taking place and substrates consumed. In order to optimise the nutrient removal process, reliable and accurate estimates of the parameters such as carbon fractions present in the wastewater, influent nutrient concentrations and biological growth rates are required.

Carbon characterisation involves fractionation based on the rate of bacterial degradation. Knowledge about the amount of carbon in each fraction can be used to help predict the reaction rates for phosphorus and nitrogen removal.

Carbon can be characterised into fractions which include; inert soluble organics, readily biodegradable organic matter, rapidly hydrolysable organic matter, slowly hydrolysable organic matter and inert suspended organics (Henze, 1992).

The objectives of this study were;

1). To gather information about the nutrient concentrations in the effluent from the dairyshed holding yard and the anaerobic treatment pond outlet.

2). To characterise the carbon material from the dairyshed holding yard and the anaerobic pond outlet into fractions based on availability for bacterial growth

Information was gathered to design a treatment system for dairyshed effluent.

2.2 Research Methods and Techniques.

2.2.1 Study farm

The dairy farm from which wastes were collected was located at Massey University, Palmerston North, New Zealand. The facilities were built for approximately 400 cows (farm data sheet 1982). The cows were milked in two herds and the yard was washed down at the end of milking.

The anaerobic pond received all the waste from the dairyshed and any run off water from the feed pad area. Solids from the feed pads were scraped into a stock pile and did not contribute to the pond loading.

2.2.2 Sample collection

Samples were collected fortnightly over an eighteen week period, from 7th August until 11th December 1996.

The raw yard effluent sample consisted of a two litre sample taken from a twenty litre composite collected during the morning yard wash down. Due to the constant wash down water flowrate and the relatively short time (twenty minutes) it took to wash the yard, effluent concentrations were expected to remain relatively constant and not require flow dependent sampling (Shilton 1996).

The yard effluent sample does not represent the total effluent load entering the anaerobic pond. The pond system receives effluent from the other activities such as, vat and milking machine washing, spilt milk and cow bail cleaning.

The reason for only selecting yard wash down water was that the yard washing practices are similar between farmers, composite samples were easily collected, and equipment for flow dependent sampling was not available. Effluent collection at the pond entrance would have required flow dependent sampling as water entered the pond during the whole time milking was in progress not just during the yard washing. The data gathered about effluent strength can be applied to other situations such as stock truck effluent disposal.

The anaerobic pond effluent sample consisted of a two litre grab sample collected from the pond outlet pipe after the morning washing of the yard was completed.

Samples were returned to the laboratory within 1 hour of collection. Analysis for Biological oxygen demand (BOD), Chemical oxygen demand (COD), alkalinity and pH occurred on return to the laboratory. Samples for volatile fatty acid (VFA), total nitrogen, total phosphorus, nitrate and ammonia were frozen before analysis.

2.2.3 Laboratory procedures:

2.2.3.1 Alkalinity

Alkalinity was measured by titration of a 25 ml sample with 0.02N HCl, as described in Standard Methods (1995).

Alkalinity was calculated using $A = \frac{a * N * 50000}{V}$ 4.

Where A = Alkalinity as mg $CaCO_3$/L

a = Volume of acid added (ml)

N = Normality of acid

V = Volume of sample (ml)

Alkalinity was measured to a pH endpoint of 4.5.

2.2.3.2 pH

pH was measured using a Hanna Hi 9025c Hand held pH meter. The meter was calibrated daily before use with 7.01 and 4.01 pH buffers. pH was corrected for temperature automatically by the meter.

2.2.3.3 Ammonia

Ammonia was measured using a version of the automated phenate method as described in Standard Methods (1995), the modification of this method developed by the Lime and Fertiliser Research Laboratory, Massey University, New Zealand. Nine standards were used over the range 0-12 ppm. The standard solutions were NH_4NO_3 and were stored at 4°C. Samples were analysed in duplicate at 1:10 and 1:15 dilutions. Samples were thawed on the day of analysis.

2.2.3.4 Nitrate Nitrogen

Nitrate nitrogen was measured using a version of the automated hydrazine reduction method as described in Standard Methods (1995), and modified by the Lime and Fertiliser Research Laboratory, Massey University, New Zealand. Standards and sample dilutions were the same as for ammonia determination.

2.2.3.5 Total Nitrogen

Total kjeldahl nitrogen analysis was carried out on a 10ml sample volume, to which 4ml of digestion acid was added. Digestion acid was 2.5l H_2SO_4 with 250g K_2SO_4 and 2.5g Selenium powder.

Samples and digestion acid were heated at 120°C for 10 hours, then 150°C until the sample volume was 4ml. Samples were then digested at 350°C for 4 hours allowed to cool and diluted to 50 ml. Samples were analysed using an autoanalyser colorimetric technique (Twine & Williams, 1971; Technicon, 1973).

2.2.3.6 Total Phosphorus

Phosphorus analyses were conducted on the total kjeldahl digestion by a automated colorimetric technique (Twine & Williams, 1971; Lime and Fertiliser Research Laboratory, Massey University, New Zealand.).

2.2.3.7 Carbon fractionation

The readily biodegradable carbon fraction was determined by VFA analysis and from CBOD data for the first 12 hour of incubation.

The rapidly hydrolysable carbon fraction was assumed equal to the CBOD value after 24-36 hours of incubation.

Slowly hydrolysable carbon fraction was determined from $CBOD_U$ less readily biodegradable and rapidly biodegradable fractions.

Inert carbon was determined from total COD less BOD_U. Inert suspended carbon was calculated from total inert carbon less soluble inert carbon. Soluble inert carbon was estimated from soluble COD less BOD_5. This was different to that proposed by Henze (1992), however the estimated error introduced by using BOD_5 instead of soluble BOD_5 will be small. This was confirmed using Mason's (1994) data. An average difference of 14 mg/l (6%) was calculated using data for the anaerobic pond effluent.

To determine the carbon fractions the following test were conducted on each effluent.

2.2.3.8 Total COD

Total COD was determined using the closed reflux colorimetric method as described by Standard Methods (1995). Samples were analysed in duplicate. Five standards were used to generate the standard curve for each batch. The calibration curve is shown in appendix 1.

2.2.3.9 Soluble COD

Soluble COD includes the dissolved and particulate material that stays in suspension after centrifugation.

Samples were centrifuged at 3000 rpm for 15 min and tested using the closed reflux colorimetric method as described by Standard Methods (1995). This procedure has

been used previously by (Mawson 1996). Duplicate samples were analysed and the COD was read from the calibration curve.

2.2.3.10 Carbonaceous BOD

Carbonaceous BOD was tested for using the proposed respirometric method (Standard Methods 1995). The instrument used was a Hach BODTrak (Hach Company, Loveland, Colorado. USA.). Sample BOD was measured in triplicate and each run lasted 10 days. Nitrification was inhibited by adding approximately 0.16 grams of Hach Nitrification Inhibitor (cat. No. 2533-35). The Ultimate BOD (BOD_U) and first order rate constant were calculated using the Fujimoto method (Metcalf and Eddy, 1991). Appendix 2 contains the average BOD curves used to determine BOD_U and K.

The Hach BODTrak was calibrated using a mixture of 150 mg/l each of glucose and glutamic acid as the BOD standard. A 10% by volume anaerobic pond effluent seed was used. Standards were run in duplicate, and the seed BOD was determined in triplicate. The corrected BOD of the standard solution should be 198 ± 30.5 mg/l(BODTrack operation manual. Hach Company, Loveland, Colorado. USA.). The corrected standards of 202 and 207 mg/l were within the limits, so the calibration of the unit was accepted.

2.2.3.11 Volatile Fatty Acid

20ml samples were treated with 4ml of a protein precipitant, centrifuged at 3000 rpm and frozen until analysis.

The protein precipitant consisted of 375g metaphosphoric acid dissolved in 900ml distilled water. To the metaphosphoric acid solution 500 ml of 100% formic acid was added along with distilled water to make the final volume 2 litres.

Standard VFA solutions were 1.0M acetic acid, propionic acid and n-butyric acid. 0.1M iso-butyric acid, iso-valeric acid and valeric acid.

Samples were analysed for VFA's with a Carlo Erba HRGC 5300 Mega series gas chromatograph using a external standard method. The adsorption column was a Stabilwax-Da column packed with Carbowax-PEG matrix. The column length was 15m long and 0.53mm internal diameter. A sample volume of 5µl was injected manually and carried by Hydrogen gas at a flowrate of 0.5 Kg/cm^2. VFAs were detected using a flame ionisation detector (FID).

Sample concentrations were calculated in meq/100ml and were converted to mg/l using the molecular weights in Table 7.

Table 7 VFA molecular weights

Acetic acid	60.05 g/mol
propionic acid	74.08 g/mol
Butyric acid	88.10 g/mol
Valeric acid	102.13 g/mol

Values calculated below 20 mg/l were reported as trace amounts and not included in total VFA calculations.

2.3 Results and Discussion

Appendix 3 contains tabulated results from the dairyshed effluent monitoring.

2.3.1 Alkalinity

The alkalinity contained in the yard effluent showed high variability. The average alkalinity was 1370 mg/l (sd 790). The anaerobic effluent had a more stable effluent alkalinity, the average being 1240 mg/l (sd 160). The anaerobic pond effluent alkalinity was more stable because the large pond volume buffered concentrations so daily variations were not seen, however seasonal trends maybe. Figure 3 illustrates the alkalinity concentration as $CaCO_3$, measured to pH 4.5.

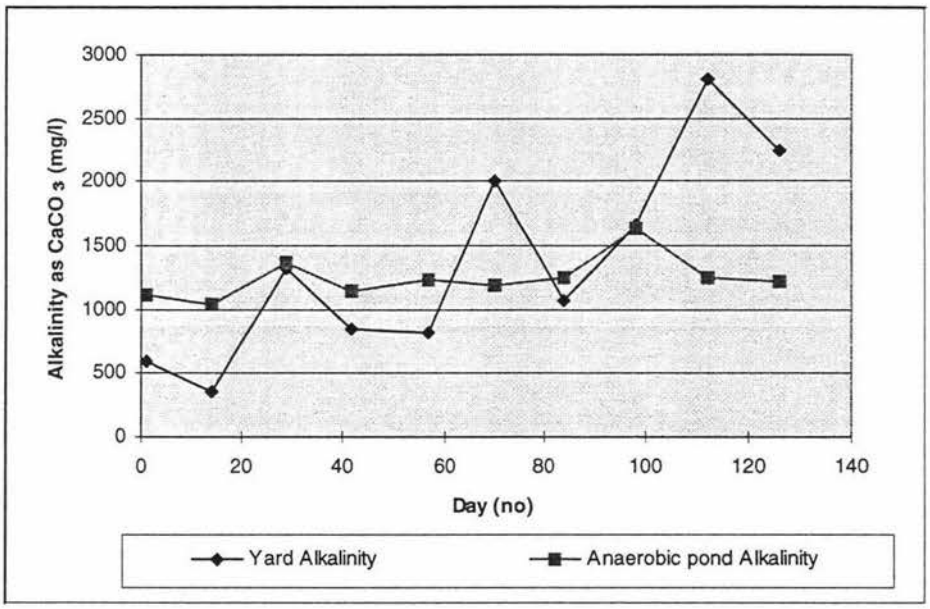

Figure 3 Alkalinity of the yard and anaerobic pond effluents.

2.3.2 pH

pH during the study period of both effluents was found to be stable (Figure 4). The average pH from the yard was 8.5 (sd 0.2), and 7.1 (sd 0.1) from the anaerobic pond. The yard effluent pH was within the optimum range for nitrifying bacteria and slightly

above the optimal for denitrifying bacteria. The anaerobic pond pH at 7.1 was near the optimum range of 7.2-9.0 for nitrification.

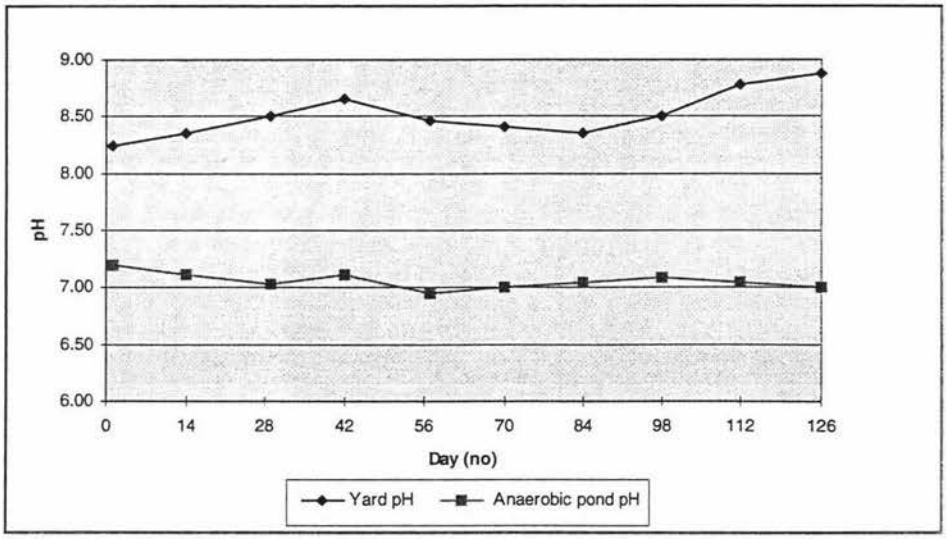

Figure 4 Yard and anaerobic pond effluent pH

2.3.3 Ammonia Nitrogen

Ammonia levels for anaerobic pond effluent during the study period were relatively stable with an average of 162 mg/l (sd 17.5). The average yard ammonia concentration of the first eight samples was 112 mg/l (sd 52) which was lower than the anaerobic pond average as can be seen in Figure 5. The yard effluent average when the final two samples were included increased to 209 mg/l, (sd 210), because of changing yard wash down practice which used a lower volume of water.

Ammoniacal nitrogen concentration in the anaerobic pond effluent are higher than the yard effluent concentrations if the last two the yard samples are not included because in the anaerobic pond organic nitrogen was converted to ammonia

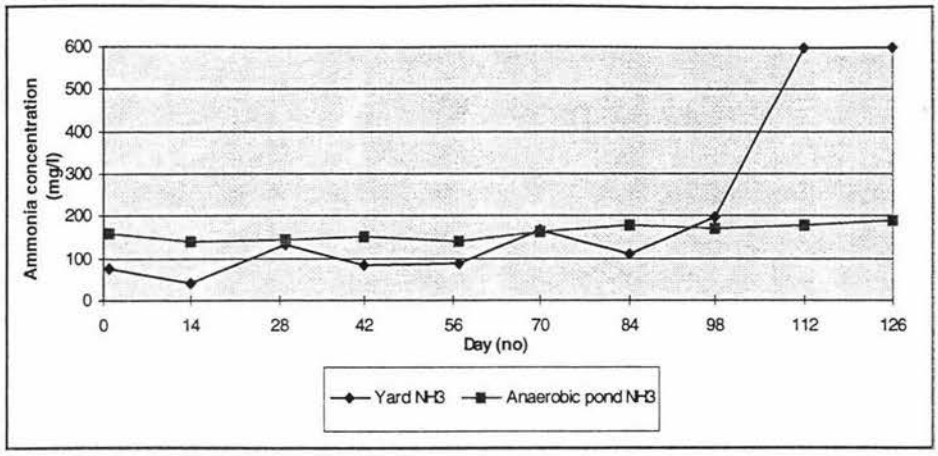

Figure 5 Ammonia concentrations for yard and anaerobic pond effluent.

2.3.4 Nitrate Nitrogen

Nitrate nitrogen levels were found to be low in both the yard effluent and the anaerobic pond effluent. (Figure 6). The yard effluent nitrate levels were higher than the anaerobic pond effluent because the yard effluent contained plant material that may not have been fully digested and denitrification may have occurred in the anaerobic pond. The final two yard concentrations were effected by the changing wash down procedure.

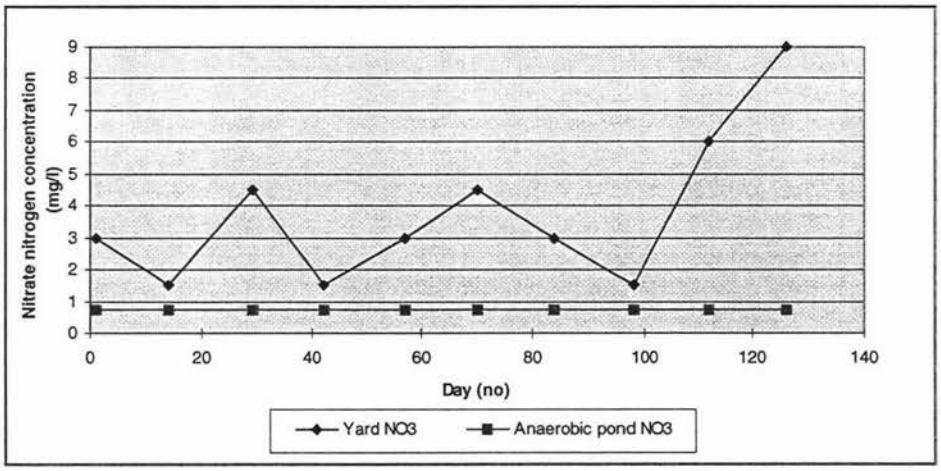

Figure 6 Nitrate nitrogen concentration for the yard and anaerobic pond effluents.

2.3.5 Total Kjeldahl Nitrogen

Figure 7 illustrates the total nitrogen concentration in the yard and anaerobic effluents. As with pervious analyses, the anaerobic pond effluent showed little change during the study period, the average was 240 mg/l (sd 23). The yard effluent total nitrogen was more variable, the average was 682 mg/l (sd 551). The average was affected by the last two samples which were about 3-6 times more concentrated than the other samples. The average decreased to 430 mg/l (sd 68) when the last two samples were excluded.

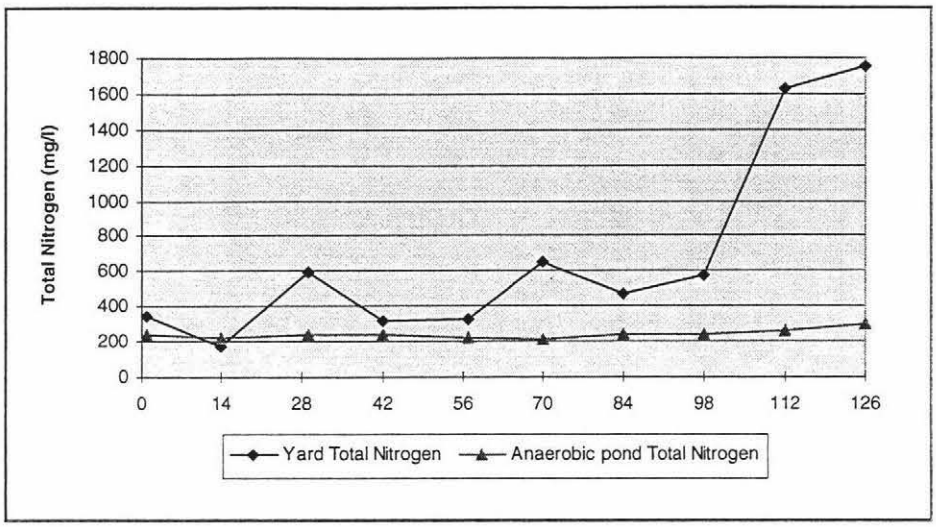

Figure 7 Total Nitrogen for the yard and anaerobic pond effluents.

Comparing the anaerobic pond effluent ammonia and total nitrogen concentrations shows that 32% (80 mg/l) of the nitrogen was organic. The proportion of total nitrogen associated with the organic fraction for the yard effluent was 70%. This was a significant amount of nitrogen that must be considered when attempting to treat the yard effluent. A longer time will be required for treatment of the yard effluent as organic nitrogen will need to be reduced to ammonia before nitrification can occur.

In order for the pH to remain above 7.0 excess alkalinity is required. The alkalinity to total nitrogen ratios (representing the worst case as not all of the nitrogen will be nitrified) for the yard and anaerobic effluent are 2:1 and 5:1 respectively. These ratios are lower than required for biological nitrification. If nitrification of the effluent was required, alkalinity addition would be needed. In a combined nitrification-

denitrification system approximately 3.57 g $CaCO_3$ is released per g NO_3-N denitrified (Chui et al, 1996). If 80% denitrification of the effluent was achieved then the alkalinity ratio would be 4.9:1 and 8.0:1 for yard and anaerobic pond effluent respectively. The alkalinity ratios for a combined nitrification-denitrification indicate that treatment of both effluents would require close monitoring, and the yard effluent would require additional alkalinity.

2.3.6 Total Phosphorus

Figure 8 Shows the total phosphorus concentration for the yard and anaerobic pond effluent. It can be seen that the yard effluent had a large amount of variability and contained high levels of phosphorus. The average for the yard effluent was 100 mg/l (sd 55) and for the anaerobic pond the average was 33 mg/l (sd 5). In the anaerobic pond a large amount of the phosphorus was removed by sedimentation. The increase of 17 mg/l seen in the anaerobic pond effluent concentration may be related to an increase in ambient temperature, causing settled sludge to rise, releasing phosphorus into solution.

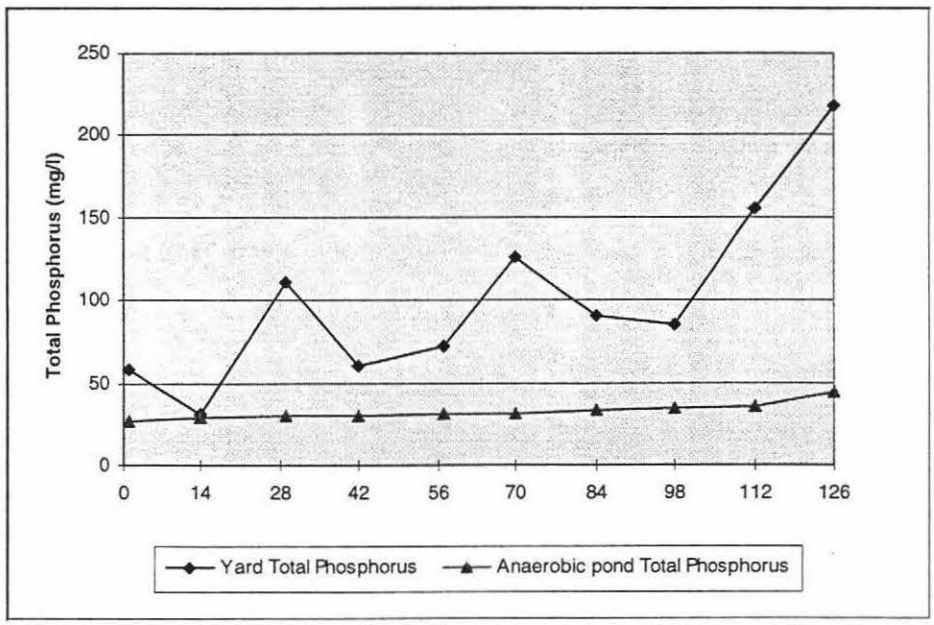

Figure 8 Total phosphorus concentration for the yard and anaerobic pond effluents

2.3.7 Yard effluent COD

Figure 9 shows COD levels of yard effluent. The yard effluent COD was variable during the monitoring period with the total COD average 10900 mg/l (sd 3830). The large variability was thought to be due to changing cow numbers, changes in the feed eaten by the cows due to natural variation in the pasture, changes to feed supplement and changes in yard washing practices during the last two collection periods. The yard was scraped to remove large solids then washed on days 112 and 126. The soluble effluent COD average was 4755 mg/l (sd 2170). The yard effluent total COD was higher than the soluble COD because it included the solid material. Removal of the solid material reduced the COD considerably, shown by the difference between total COD and centrifuged soluble COD (Figure 9).

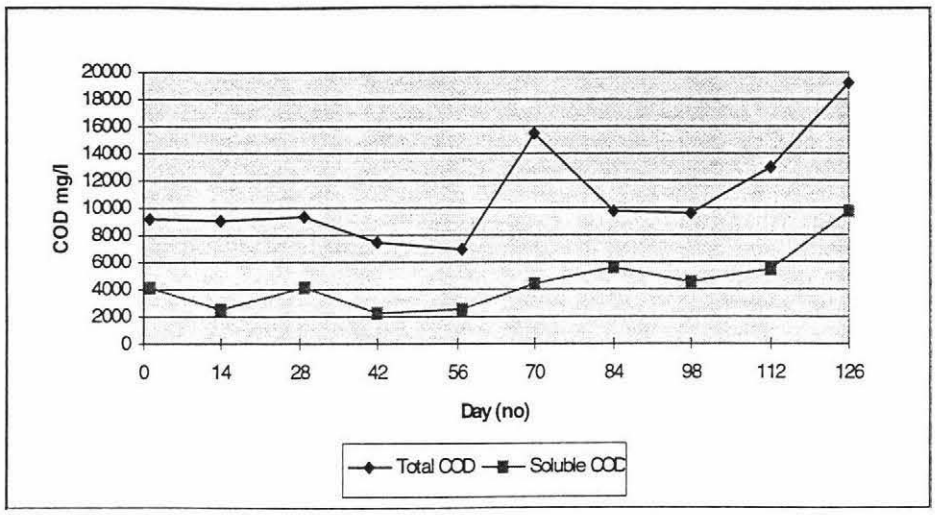

Figure 9 Yard effluent Total and Soluble COD

2.3.8 Anaerobic pond effluent COD

The anaerobic pond effluent COD is illustrated in Figure 10. There is a noticeable decreasing trend for the soluble COD, average 835 mg/l (sd 220), for the first seven samples. The decreasing trend is possibly due to increased biological activity in the anaerobic pond due to increasing ambient temperature. The increase after sample eight may be due to sludge rising to the surface due to gas formation. The total COD average was 1420 mg/l (sd 280). The total COD variation was probably due to changes in the solids content of the sample shown by the difference between the total and soluble COD curves (Figure 10).

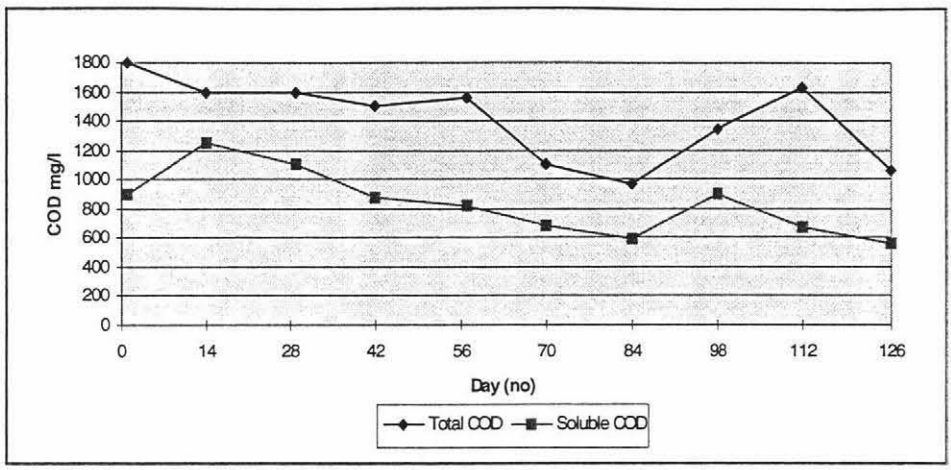

Figure 10 Anaerobic pond effluent Total and Soluble COD.

2.3.9 Carbonaceous BOD

2.3.9.1 Yard effluent

The yard effluent BOD follows the same trend as the soluble COD (Figure 11). Again, the increased BOD value of samples on day 112 and 126 may be associated with changes in the wash down practices used. The ultimate BOD average was 7213 mg/l (sd 3478). This average decreased to 5292 mg/l (sd 1070) when only the first six samples are considered. The first order reaction rate constant K at 20°C calculated using all nine samples was 0.20 d^{-1} (sd 0.06). This reaction rate constant is just below 0.23 d^{-1} which is a typical value for domestic wastewater (Metcalf and Eddy, 1991). As can be seen from the high BOD the polluting potential of raw dairy waste is considerable.

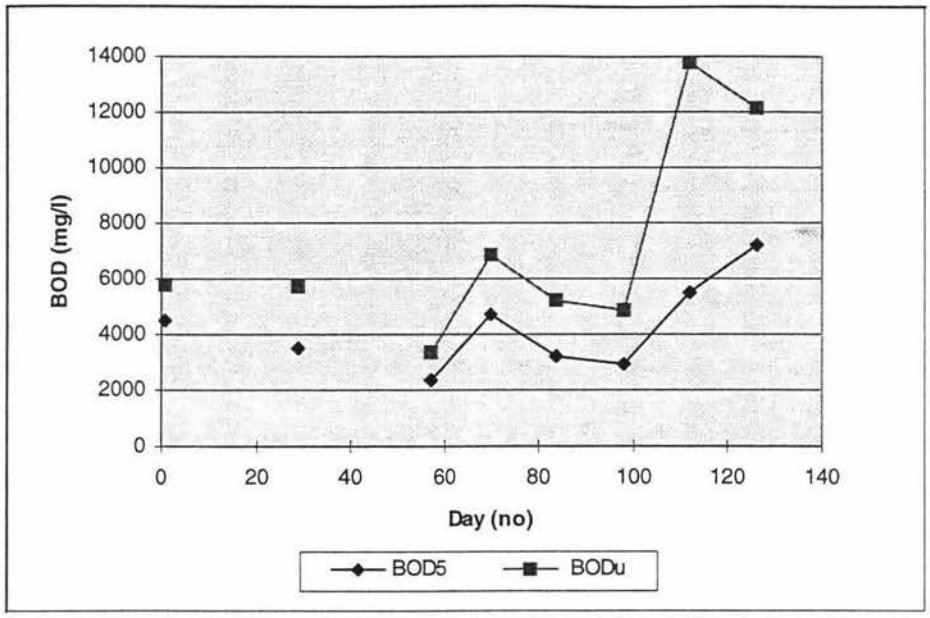

Figure 11 Yard effluent Carbonaceous BOD

2.3.9.2 Anaerobic pond effluent

Figure 12 below shows the anaerobic pond discharge. There was a similar decreasing trend as was seen with the COD. The decrease may be due to increased biological activity in the anaerobic pond with increasing ambient temperatures. The average ultimate BOD was 737 mg/l (sd 356). The reaction rate constant for the anaerobic pond effluent was 0.16 d^{-1} (sd 0.06). This was lower than the yard effluent rate constant of 0.20 d^{-1}. The lower rate constant indicates the complex nature of the waste and the reduced biodegradability after anaerobic treatment.

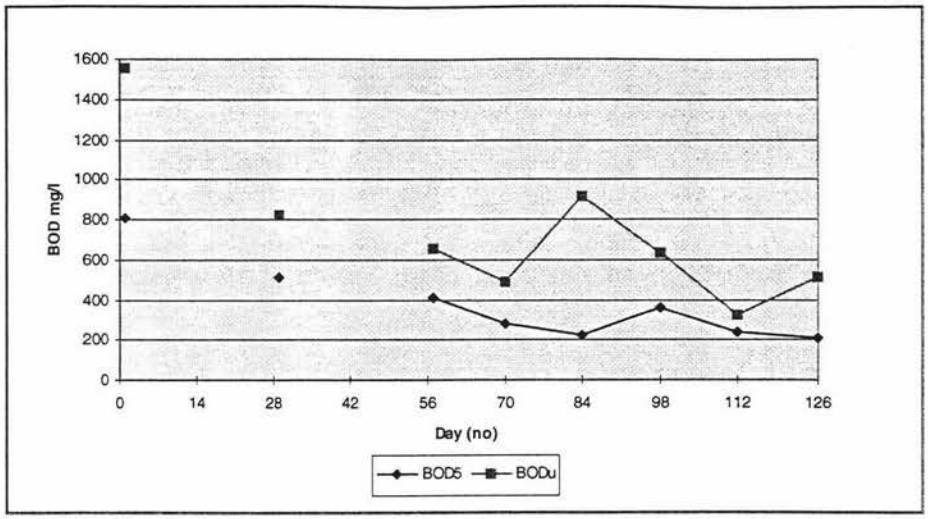

Figure 12 Anaerobic pond effluent carbonaceous BOD.

2.3.10 Volatile fatty acids

The trends shown by the volatile fatty acid concentration for both effluents were similar to the other parameters monitored (Figure 13). The average yard VFA concentration was 353 mg/l (sd 207), 80% was acetic acid, 12% was propionic acid and 4% was n-butyric. The remaining 4% consisted of trace amounts of iso-butyric, iso-valeric and n-valeric acids.

The anaerobic pond VFA concentration average was 92 mg/l (sd 38), 87% was acetic acid, 12 % was propionic acid and the remainder consisted of trace amounts of n-butyric and iso-valeric acid. Iso-butyric acid was found in trace amounts in only one sample, while n-valeric was not found. VFAs represent 9% of the yard effluent BOD_5 and 24% of the anaerobic pond effluent BOD_5.

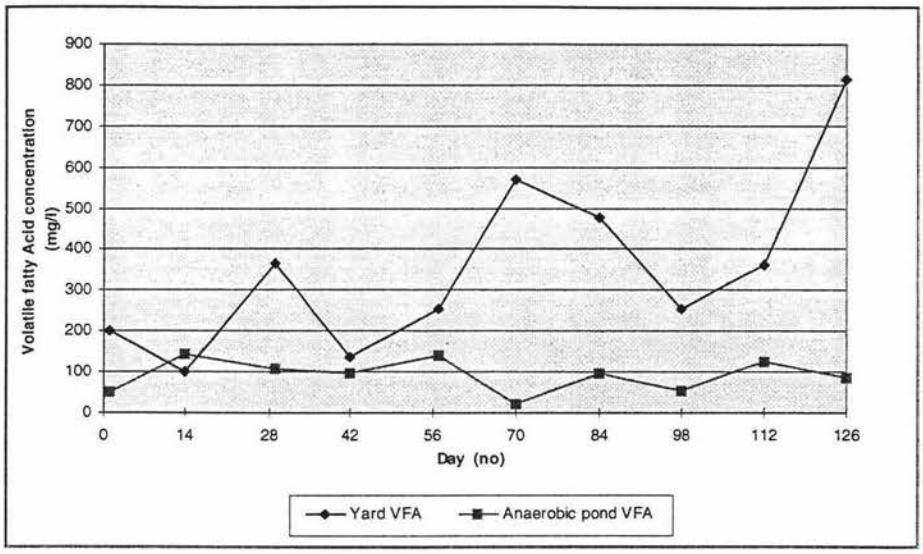

Figure 13 Yard and anaerobic pond effluent volatile fatty acid concentration.

2.3.11 Carbon fractions

Figure 14 shows the BOD contribution from VFAs, the amount of BOD exerted after 0.5, 1, 1.5, 5 days and the ultimate BOD of the yard effluent. The BOD after 0.5 days was due largely to VFAs assuming that VFAs are consumed early in the BOD test. The BOD after 0.5 day represented the readily biodegradable fraction of the BOD, the average was 957 mg/l, representing 14% of the total BOD.

Eckenfelder (1989) suggests that most of the rapidly hydrolysable substrate is consumed in 24 to 36 hours. Using the average BOD after 24 to 36 hours the rapidly hydrolysable fraction of the yard effluent BOD averaged 885 mg/l representing 13% of the total BOD. Slowly hydrolysable fraction is equivalent to the remaining BOD and averaged 4943 mg/l and represented 73% of the total BOD of the yard effluent. The 73% of the BOD represented by the slowly hydrolysable carbon fraction was higher than the 40-60% of COD predicted by Henze (1992) for raw domestic wastewater if the ratio for COD:BOD of 0.56:1 was taken into account.

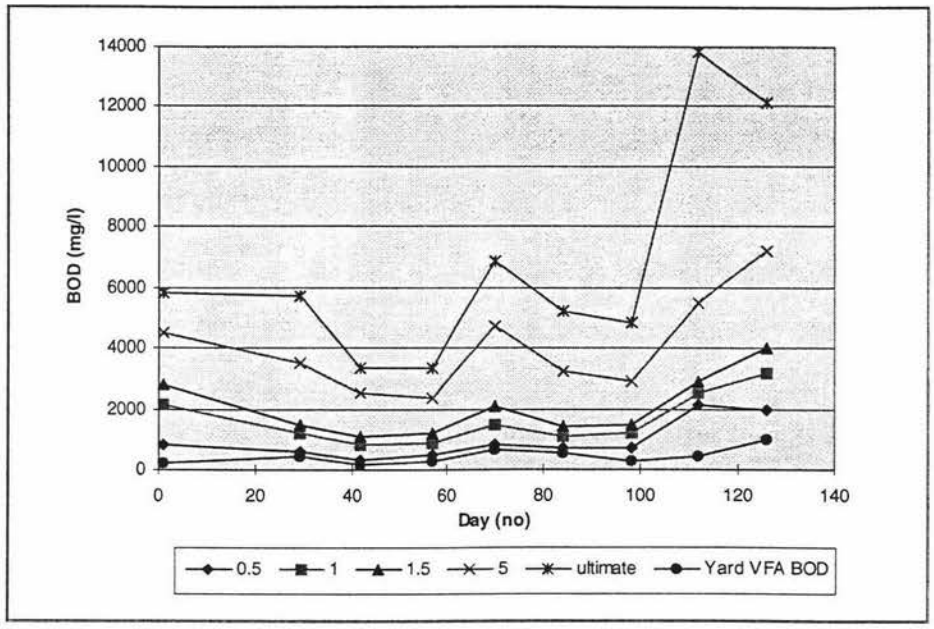

Figure 14 Yard effluent BOD profiles.

Figure 15 shows the BOD profiles with time for the anaerobic effluent. It can be seen that BOD contributed from VFAs made up a large proportion of the 0.5 day BOD and in some cases exceeded the 0.5 day BOD. Little difference was found between the 1.0 and 1.5 day BOD, 22 and 25% of the ultimate BOD had been exerted after each time. If the 0.5 day BOD represents readily biodegradable BOD then on average 121 mg/l BOD was readily biodegradable, representing 16% of the ultimate BOD. The rapidly hydrolysable fraction represented by the average value between 0.5 and 1.5 day BOD was 55 mg/l, 7% of the ultimate BOD. The remainder of the BOD was slowly hydrolysable and amounted to 575 mg/l, 76% of the ultimate BOD.

Soluble inert COD and suspended inert COD was calculated for each effluent. Due to a lack of soluble BOD data and the large variations in the yard BOD and COD, calculation of the soluble inert value gave negative results. The average total inert material calculated from the difference between total COD and ultimate BOD was 4936 mg/l (sd 1795). Anaerobic pond effluent inert soluble COD averaged 407 mg/l (sd 135) and suspended inert COD average was 240 mg/l (sd 204). When treating the anaerobic pond effluent there will be a large residual COD in the treated effluent due to the soluble inert material.

Comparing the soluble inert COD of 407 mg/l calculated using the anaerobic pond data collected in this study to the value calculated from Mason's (1994) data of 229 mg/l there was a difference of 178 mg/l, which may be attributed to increased loading on the anaerobic pond from the addition of the feed pad wastewater.

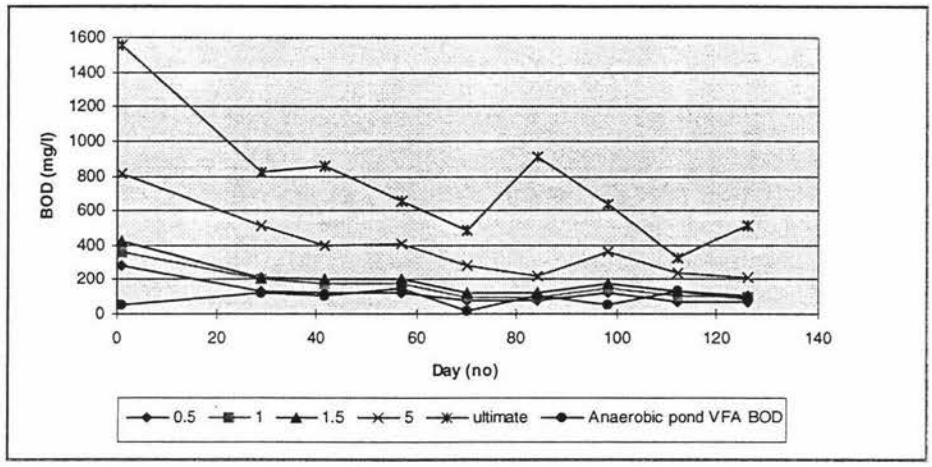

Figure 15 Anaerobic pond effluent BOD profiles

Using the information about carbon fractions it is possible to predict amounts of denitrification and phosphorus release. For total denitrification to occur the ratio of rapidly available BOD:NO_3 should exceed 2.3:1 (Narkis, 1979). The amount of nitrate formed from each effluent, assuming 80% of the total nitrogen is nitrified, is 570 and 190 mg/l for yard and anaerobic pond effluents respectively. The BOD used to calculate the ratio is the rapidly hydrolysable and readily available, this amounts to 1842 and 176 mg/l for yard and anaerobic pond effluents respectively. The ratios are 3.23:1 for the yard effluent and 0.93:1 for the anaerobic effluent. This means that the yard effluent could denitrify all the nitrate formed and still have carbon available for phosphorus release. For the anaerobic pond effluent only 40% of the nitrate would be denitrified making biological phosphorus removal impossible without supplementary carbon being added. This trend was also found in other research which attempted to aerobically treat an anaerobically pretreated effluent (Comeau et al, 1996; Subramaniam et al, 1994; Ng, 1987).

Table 8 Summary of carbon fractions in yard and anaerobic pond effluents

Parameter	Yard effluent	Anaerobic pond effluent	Domestic wastewater (Henze, 1992)
Readily biodegradable BOD	957 mg/l 14%	121 mg/l 16%	22% of total BOD
Rapidly hydrolysable BOD	885 mg/l 13%	55 mg/l 7%	37% of total BOD
Slowly hydrolysable BOD	4943 mg/l 73%	575 mg/l 76%	40% of total BOD
Total inert COD	4936 mg/l	647 mg/l	
Soluble inert COD	-	407 mg/l	
Amount of Denitrification possible	100%	40%	100%
Phosphorus release	YES	NO	YES

Table 8 shows the considerable difference between the relative amounts of carbon in each fraction found in dairyshed effluent and domestic wastewater.

2.4 Conclusions

Both yard and anaerobic pond effluent have high nutrient levels when compared to domestic sewage. The pH of both effluents was near the optimum required by bacteria for nitrification and denitrification.

Significant amount of nitrogen in both effluents was organic in nature. The anaerobic pond effluent contained 32% organic nitrogen and the yard effluent contained 70% organic nitrogen, therefore the yard effluent would take longer to treat.

The BOD reaction rate constant K for the yard effluent at 0.2 d^{-1} was similar to a typical domestic wastewater value of 0.23 d^{-1}. The anaerobic pond effluent reaction rate

constant K was 0.16 d^{-1} which is lower than the yard effluent value indicating that the anaerobically treated effluent was hard to aerobically degrade.

A significant amounts of the BOD in both effluents was from VFAs. 24% of the anaerobic pond BOD was from VFAs and 9% of the yard effluent BOD was from VFAs.

This research also showed that there was a large difference between dairyshed effluent and domestic effluent in the proportion of carbon in each fraction.

The yard effluent lacks alkalinity for nitrification, but contains large amounts of readily available carbon (1850 mg/l) which is needed for denitrification and phosphorus removal. Anaerobic pond effluent has much lower levels of readily available carbon (180 mg/l), but alkalinity levels would be adequate for nitrification, if denitrification was complete. Due to the low readily available BOD to nitrate ratio it is expected that less than 40% of the nitrate present would be denitrified. Phosphorus removal would not occur with out addition of carbon.

3. Oxygen Transfer

3.1 Introduction

Aeration is used to supply oxygen to microorganisms for enhanced aerobic biological activity. The natural rate of aeration in high rate activated sludge systems is insufficient to meet the oxygen demands of these bacteria. Mechanical aeration systems are required to supply oxygen at high rates.

The objectives of the work on oxygen transfer were to characterise the performance of the aeration unit supplied with the SBR and if the unit was unsuitable evaluate an alternative aerator.

3.2 Mechanisms of Oxygen Transfer

Aeration is a mass transfer process involving a gas/liquid mixture. Oxygen is transferred by diffusion along a concentration gradient, from high concentration, about 220 mg/g, in the gas phase to low concentrations in the liquid phase, about 9 mg/kg, at 12°C.

The diffusion process is defined by Ficks law where the rate of diffusion is dependent on the gas and liquid characteristics, the temperature, concentration gradient and the gas bubble surface area in contact with the liquid. Ficks law equation 5, states that the amount of gas transferred per unit time is proportional to the contact area and concentration gradient.

$$N = -D_L A \frac{dc}{dy} \qquad 5.$$

Where N = mass transfer per unit time

A = contact area over which diffusion occurs

D_L = diffusion coefficient

dc/dy = concentration gradient

Lewis and Whitman developed the 2 film theory in 1923 which assumes that the gas and water are at equilibrium at the interface. There are two resistances to mass transfer, one in the gas phase and the other in the liquid phase can be added to give an overall mass transfer coefficient.

Lewis and Whitman's equation 6,

$$N = K_L A(C_S - C_L) = K_G A(P_S - P) \qquad 6.$$

Where C_S = oxygen saturation concentration

C_L = concentration of oxygen in the liquid

K_L = liquid film coefficient

K_G = gas film coefficient

A = bubble surface area

Usually, to determine the amount of oxygen transferred into a system only the liquid situation is considered because for sparingly soluble gases, such as oxygen, the liquid film resistance controls the rate of mass transfer (Eckenfelder 1989).

The transfer of gaseous oxygen to the liquid phase is a physical process that follows a first order, decreasing rate kinetic pattern. Equation 7 shows the liquid film controlled expression of equation 6, in concentration units.

$$\frac{1}{V}N = \frac{dc}{dt} = K_L \frac{A}{V} (C_S - C_L) = K_L a (C_S - C) \qquad 7.$$

Where $K_L a$ = the overall film coefficient.

V = Reactor volume l

The liquid film coefficient and the area are considered as one coefficient because of the difficulty in measuring the bubble surface area accurately.

Integration of equation 7 yields

$$C_S - C_L = (C_S - C_O)e^{-K_L at} \qquad 8.$$

Where C_O = dissolved oxygen concentration at time zero

3.3 Methods of Measuring Oxygen Transfer.

There are several methods available for the measurement of $K_L a$. These can be divided into steady state, in which the dissolved oxygen rate of change is zero at any given point, and non steady state.

The steady state method requires the direct measurement of the oxygen uptake rate of the respiring biological system.

There are several non steady state methods:

1. Off gas analysis: The off gas method is a mass balance on oxygen that includes both the liquid and the gas streams. By comparing the composition of the gas leaving to the gas entering the tank, it is possible to calculate the oxygen transfer occurring.

2. Dynamic Method: This method is based on the measurement of the K_La of a deoxygenated liquid as a function of time, after the air flow is turned on. The deoxygenation can be obtained by passing nitrogen through the liquid.

3. Chemical sorption: These methods are based on the reaction of an absorbed gas (O_2) with a chemical added to the liquid phase. Na_2SO_3 is the most commonly used chemical and is often called the sulfite method. The O_2 converts the Na_2SO_3 to Na_2SO_4 in the presence of a Cu or Co catalyst.

   There are potential problems with this method when it is used for comparison of aeration equipment to theoretical or manufacturers data. The first issue is that there is a high salt concentration due to the Na_2SO_3 which reduces gas bubble size enhancing the oxygen transfer and secondly if comparison is made with another method it is necessary to determine the rate of chemical reaction. It can be difficult to determine the rate of the reaction as it varies with concentration. If the rate of reaction is not taken into account this method will give a higher K_La than methods 1 and 2. (Van't Riet, 1979)

The sulphite method was used to measure the K_La of the systems and comparisons were made as to which aerator was the most efficient. Because the conditions were constant for testing of each aerator it was not necessary to determine effects due to salt concentration or the order of the chemical reaction. It was considered valid to make the comparison.

3.4 Aeration Systems

There are many types of aerator designs including diffusers, surface aerators and jet aerators. This experimental work has focused on a type of jet aeration that uses a

venturi to entrain the air. Jet aeration equipment because the project was supplied locally built venturi aerator for which no performance data was available.

3.4.1 Jet aeration

The venturi aerator supplied works on a similar principle to that of the venturi meter. In a venturi meter the velocity is increased, and the pressure decreased, in the up stream cone. The pressure drop in the upstream cone is utilised to measure the flowrate of the fluid. The velocity is then decreased and original pressure is largely recovered, in the down stream cone (McCabe et al, 1985). To achieve a large pressure recovery the angle of the down stream cone is usually small between 5° to 15° (Ower & Pankhurst, 1977). A small angle stops boundary layer separation and the associated turbulence which decreases the pressure recovery.

The venturi aerator had no down stream cone, so as the fluid exits the upstream cone there is a sudden increase in cross sectional area. Boundary layer separation occurs because the rate of velocity decrease is slower than rate of cross sectional area expansion, creating a pressure drop which if open to the atmosphere can draw in air to fill the space between the fluid jet and the pipe wall. The air entrained mixes with the fluid and some dissolves. The mixture is then released into the tank, where the undissolved air bubbles to the surface, aerating the liquid.

The energy for this system is supplied by the pump recirculating the liquid from the tank to the venturi and back to the tank.

To calculate the theoretical pressure drop for a venturi meter equation 9 was used.

$$Q = Cd*E*A_2((2g*\Delta p/\gamma)^{1/2}) \qquad 9.$$

Where
- Q = Flowrate m^3/s
- Cd = Coefficient of drag
- E = Ratio of cone areas
- A_2 = Area of throat of meter
- Δp = Pressure drop across meter
- γ = Specific weight of water

By rearranging equation 9 the pressure drop can be calculated for different flowrates.

3.4.2 Submersible aerator

The submersible aerator design tested used an impeller to create a pressure drop which drew in air. The air is displaced outwards by rotational force where it combines with the water drawn in through the base impeller producing jets of air and water that radiate from the body of the aerator.

The submersible aerator can be tuned to increase the amount of O_2 transferred. Tuning is achieved by adjusting the air flowrate using a gate valve on the air inlet side. The pump is said to be tuned when the motor is drawing maximum current. When tuned the aerator was operating at it's peak water flowrate, this gave increased fluid turbulence which resulted in a decreased film thickness and hence increased K_L (Eckenfelder, 1989).

$$K_L = \sqrt{D_L r} \qquad 10.$$

Where D_L = liquid diffusion coefficient

r = rate of surface renewal frequency

Operating the aerator at high water flowrates replaces the water at the air-water interface at higher frequencies, therefore maintaining a larger concentration gradient.

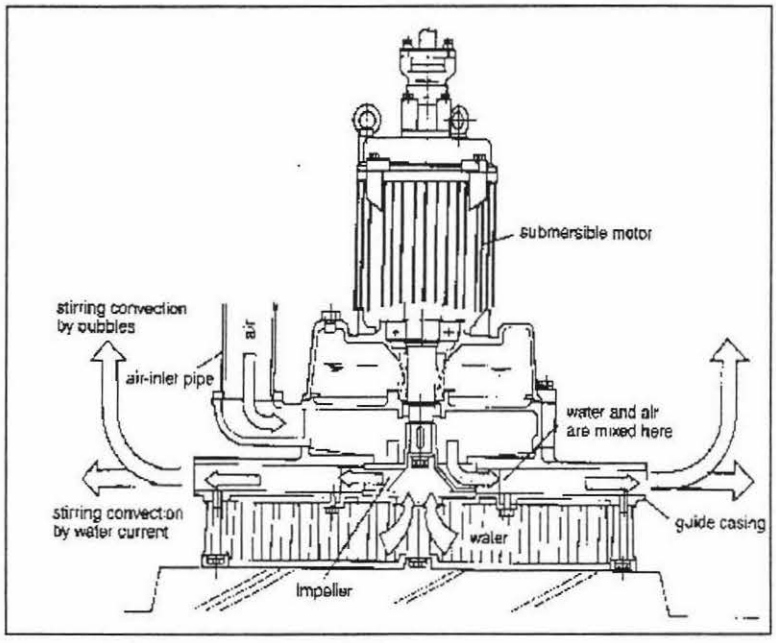

Figure 16 Tsurumi 15 TR 2 aerator

The experimental work with the venturi aerator supplied had the following specific objectives,

1. Quantify the amount of air entrained as a function of water flowrate.
2. Measure the pressure drop across the venturi as a function of flowrate.
3. Investigate the effect of water back pressure on venturi performance.
4. Investigate enhancing the performance of the aerator by using compressed air.

Objectives common to both aerators were

1. Measure the oxygen mass transfer coefficient, $K_La\ hr^{-1}$.
2. Calculate aerator efficiency, kg O_2/ kW.hr,
3. Select an aerator for use in the pilot scale SBR reactor.

3.5 Methods

The procedure of Eckenfelder (1989) for measuring oxygen transfer efficiency using the sulfite method was used. The procedure steps are;

1. Remove the dissolved oxygen in the aeration unit by adding sodium sulfite and cobalt chloride. A concentration of 0.05 mg/l of cobalt and 8 mg/l Na_2SO_3 per mg/l dissolved oxygen should be added.
2. Thoroughly mix the tank contents.
3. Start the aeration unit at the desired operating rate. Sample for dissolved oxygen at selected intervals until 90 percent of saturation is reached.
4. Record temperature and measure oxygen saturation
5. Compute the oxygen transfer rate, K_La by calculating the slope of a straight line from the plot of $\ln\frac{(C_S - C_L)}{(C_S - C_O)}$ vs. time.

Where: C_0 = dissolved oxygen concentration at time zero (mg/l)

t = time

The aerator efficiency was calculated using equation 11,

$$E = K_La * v * (C_s - C_O) / P \qquad 11.$$

Where E = Efficiency kg/kWh

v = Volume reaerated. l

P = Power of pump motor. kW

Pump motor power was determined by measuring the current and using equation 12

$$\text{power} = \text{amps} * \text{voltage} \qquad 12.$$

Air flowrates were measured using a electronic anemometer (Airflow Developments Ltd. High Wycombe, Bucks, England.). Initial venturi testing used mains pressure water so that the flowrate could be easily varied. The flowrate range was from 3.7 to 10.5 l/s. Water flowrates were measured using a "90° V notch" weir. At each flowrate the pressure drop across the venturi was measured using pressure tapping upstream and down stream of the venturi.

The back pressure effects were investigated by measuring the air flowrate into the venturi while varying the discharge depth. Forced aeration was tested by using compressed air feed of 25 m^3/hr into the venturi and measuring the reaeration constant with the sulphite method.

The aeration efficiency was measured at a depth of 1.6m using the sulphite method described earlier.

3.6 Equipment

3.6.1 Venturi aerator

This consisted of a fibre glass cone shape inserted into a PVC plastic tee bend. This aeration system was powered by a Darga 200 1.95 kW submersible pump. The water flowrate through the venturi was 7 l/s.

3.6.2 Submersible aerator

The submersible Tsurumi aerator tested was a 15 TR2 (Tsurumi Manufacturing CO., Ltd. Japan) this was a 1.5 kW unit capable of supplying between 1.0-1.4 kgO_2/Hr at a submersion depth of 3 meters (manufacturers data).

3.7 Results

3.7.1 Venturi aerator

As the water flowrate increased there was an increase in the pressure drop across the venturi which caused more air to be entrained. Figure 17 shows air flowrate and pressure drop as a function of flowrate.

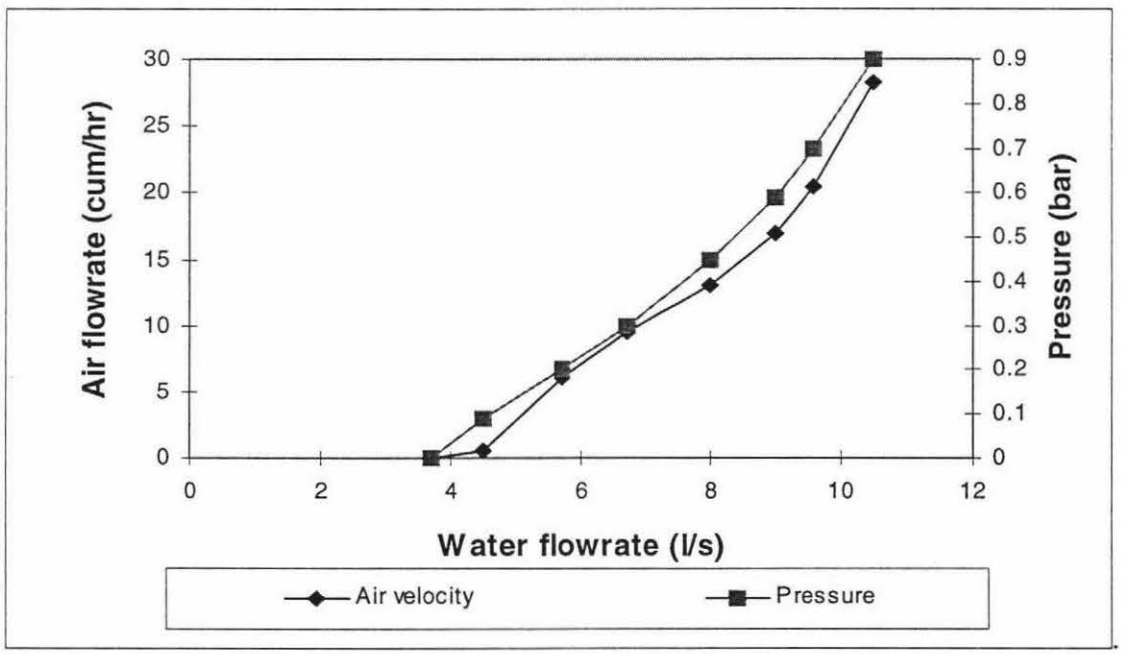

Figure 17 Air flow and pressure drop though the venturi as a function of water flowrate.

The effect of closing the air inlet to the venturi decreased the pressure drop. The amount the pressure decreased was effected by the flowrate. At a low flow of 5.7 l/s

the change in pressure was 0.2 bar, at the maximum flow tested 9 l/s the pressure drop was 0.28 bar. This is compared in Figure 18 with the theoretical pressure drop created by a venturi used for flow measurement

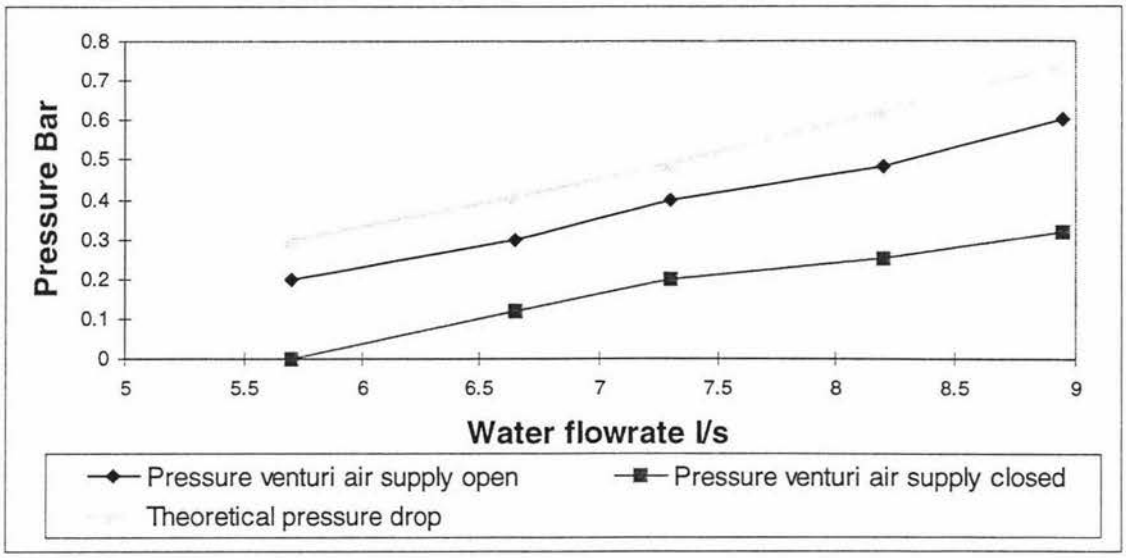

Figure 18 Theoretical pressure drop and actual pressure drop for a venturi aerator.

The effect of water back pressure on air flowrate was found to be significant. There was a large reduction in the air flowrate with increasing back pressure. From Figure 19 it can be seen that a minimum discharge head was required for the venturi to operate. The head was 60 to 65 cm of water at which an air velocity of 0.7 m/s was recorded. The velocity decreased to 3.5 m/s when the discharge head was 162 cm.

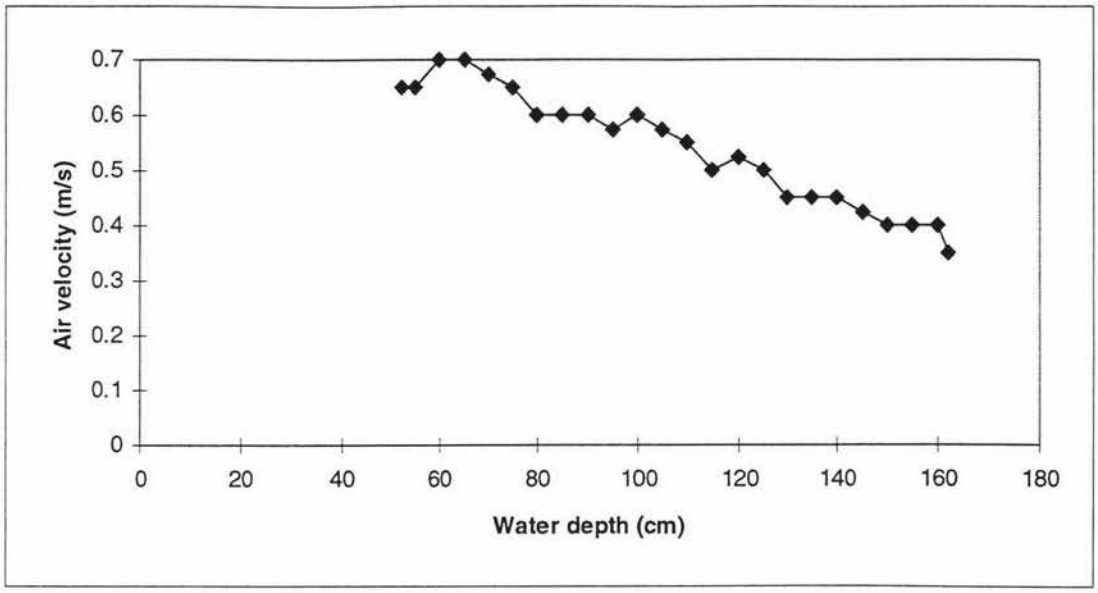

Figure 19 Air flowrate as a function of depth

It was found that there was a minimum down stream length required before the venturi would create a significant pressure drop. Placing a bend on the discharge side reduced the air flowrate from 0.7 m/s to 0.5 m/s.

Further work was carried out on tuning the venturi. The tuning involved changing the cone length, down stream pipe length and down stream flow patterns. It was found that a cone length of 14.5 cm with a discharge diameter of 32mm, followed by one meter of straight pipe gave the best performance.

Once the venturi was tuned the reaeration constant K_La was measured.

Average K_La_{20}	13.5 (sd 1.66)	hr^{-1}
Average efficiency kg/kWh	0.30 (sd 0.04)	kg/kWh
Average mass transfer	0.55 (sd 0.06)	kg/hr
compressed air K_La	17	hr^{-1}

3.7.2 Submersible aerator

Aeration testing was conducted with the aerator tuned and untuned. The power requirements were 1.5 kWh tuned and 0.8 kWh untuned, the mass of oxygen transferred was equivalent at 0.8 kg/h for both cases.

Average K_La_{20} tuned	19.45 (sd 0.225)	hr^{-1}
Average K_La_{20} untuned	19.23 (sd 1.51)	hr^{-1}
Average efficiency tuned	0.58 (sd 0.01)	kg/kWh
Average efficiency untuned	1.03 (sd 0.08)	kg/kWh
Average mass transfer tuned	0.8 (sd 0.01)	kg/hr
Average mass transfer untuned	0.78 (sd 0.02)	kg/hr

Attached in appendix 4 is a spreadsheet showing calculations of K_La and efficiency for both aerators.

3.8 Discussion

The venturi design created high air velocities when there was a high water flowrate with little or no back pressure. When the back pressure increased the overall pressure difference decreases and there is a significant reduction in the amount of air entrained.

From the comparison between the actual pressure drop and theoretical it can be seen that the venturi wasn't creating as much pressure drop as was theoretically possible, this means that less than the maximum amount of air for a certain flowrate is being drawn into the venturi. The reduction in air drawn in was because of restricted air flow patterns around the venturi cone.

The sulphite method was used to measure the K_La of the systems and comparisons were made as to which aerator was the most efficient. Because the conditions were constant

for testing of each aerator it was not necessary to determine effects due to salt concentration or the order of the chemical reaction.

The reaeration tests were performed once the maximum venturi performance had been achieved. The efficiency of the venturi aerator tested was poor with average efficiencies of 0.30 kgO_2/kWh. This is considerably lower than the efficiencies quoted by Eckenfelder (1989) for jet aerators 2.6-2.9 kgO_2/kWh, but only slightly lower than Cumby (1987) who found efficiencies for venturi aerators of 0.35-1.0 kgO_2/kWh. The performance of the submersible aerator at 1.03 kg/kWh when untuned was in the upper end of the range given by Cumby (1987).

The advantage of the submersible aerator over the venturi design is the dispersion of the air water mixture. The submersible aerator produces eight separate jets which rise up through the water column while the venturi has only a single jet. This single jet may cause the air bubbles to rise faster. The submersible aerator with eight jets has a higher frequency with which fluid at a oxygen concentration of C_L is replacing fluid from the interface with a oxygen concentration C_S.

The air water mixture leaving the venturi is in contact with less water that is at a low DO concentration. The submersible aerator's eight jets are in contact with more of the bulk liquid so there is a larger concentration gradient. Because of the larger gradient more of the oxygen is transferred. The mass transfer in clean water for the submersible aerator was 0.8 kg/hr and the venturi aerator was 0.55 kg/hr, the actual amount of air entrained was equivalent for each aerator.

The manufacturers of the submersible aerator suggest that the actual mass transfer in dirty water is 85% of the clean water value, the estimated dirty water mass transfer is 0.46 and 0.68 kg/hr for the venturi and submersible aerators respectively.

The next approach to improving oxygen transfer would be to increase the water depth as this would give a longer rise time for bubbles allowing more oxygen to dissolve in

the surround bulk liquid. If further improvements in oxygen transfer were required then a larger radius tank would be needed so that air-water jets were not forced to rise by hitting the external wall of the tank. At the wall air-water jets rise quickly reducing oxygen contact time with the bulk liquid.

3.9 Conclusions

From the reaeration tests it was determined that the venturi pump combination had poor oxygen transfer efficiency at 0.3 kgO_2/kWh. The venturi aerator's estimated dirty water mass transfer of oxygen, 0.46 kg/hr was less than the estimated 0.5 kg/hr oxygen requirement for treatment of dairyshed effluent to remove nitrogen and phosphorus. The performance did increase with the compressed air but the efficiency was still to low for it to be used in it present form. The submersible aerator had an efficiency of 1.03 kg/kWh. The submersible aerator's mass transfer of oxygen at 0.68 kg/hr exceeded the estimated requirements for biological treatment of dairyshed effluent.

The submersible aerator was chosen for use in the pilot scale reactor.

4. Sequencing batch reactor Design

4.1 Physical Design

4.1.1 Tank design

Initially it was proposed to use an Evolution treatment system, intermittent decant designed SBR (Fibre-form NZ Ltd, Palmerston North, New Zealand.) which was available from the business partner involved with this research.

The Evolution wastewater treatment SBR was designed for use in single dwelling or for small communities, where there is a requirement for a fully contained single tank system. The design consisted of a tubular tank with three compartments (Figure 20). The first compartment is used for gross solids size reduction as there is no primary sedimentation before the tank. Size reduction is achieved in the first compartment with a water-air mixture to break up solids retained by a screen between the first and second compartment. The second compartment is an aeration chamber, which uses the bacteria present to treat nutrients in the wastewater. The final compartment is also aerated during the react stage of the SBR cycle, but during the settle stage it is not agitated, and effluent settles before discharging.

For treatment of pretreated dairy effluent it was decided that no inlet chamber for solids removal was required as most gross solids would be removed during the pretreatment stage. One possible pretreatment is an anaerobic pond.

The use of a pretreatment stage also acts as a flow equalisation tank. This eliminates the need for a separate settle-decant chamber.

Whereas the Evolution SBR is a hybrid SBR system using continuous inflow and intermittent decant, the reactor designed to treat dairy effluent will be a discrete batch process.

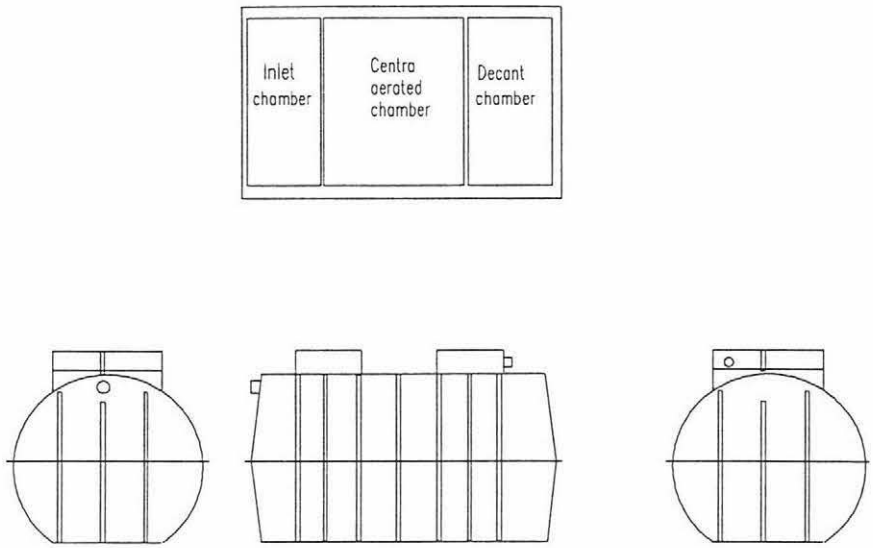

Figure 20 Evolution wastewater treatment SBR.

The size of the SBR was chosen from available moulds the business partner had. The tank volume was 4500 litres with a diameter of 1.8m at the centre and a height of 1.8 meters. Figure 21 shows the SBR dimensions in metres

The circular shape of the pilot scale SBR is easier to construct than the Evolution wastewater treatment SBR. The Evolution wastewater treatment SBR is designed for burial and are therefore considerably more expensive to manufacture than a circular tank due to the wall thickness required to withstand the pressures under ground. The cost difference between tank designs was $1400. The circular shape of the pilot scale SBR has good mixing characteristics, as there are no corners for solids to accumulate.

By using a single compartment unit instead of the Evolution treatment system, no internal transfer pump was required to pump the effluent from the react compartment to the settle-decant compartment.

The pilot scale SBR was installed on Massey University farm, Dairy N° 4 beside the

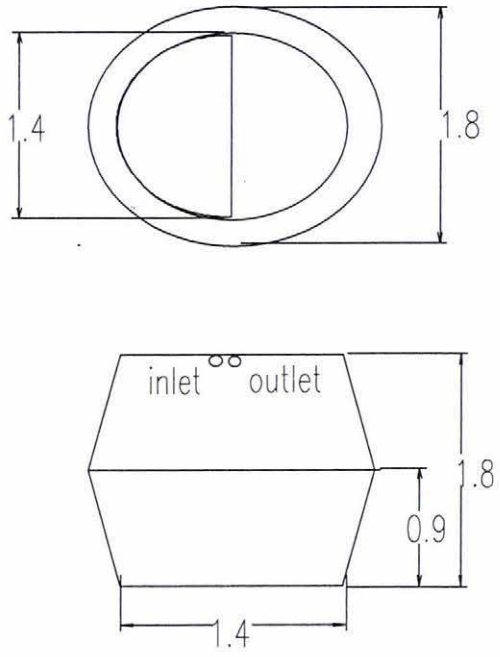

Figure 21 Pilot scale SBR diagram

Figure 22 Pilot scale SBR on site.

4.1.2 SBR operation

Influent enters through the tank wall near the top of the tank. From there it mixes with biomass and liquor present from the previous batch.

Aeration of the effluent during the react stage of the cycle was achieved using a submersible aerator. A submersible aerator was chosen after testing a venturi type aerator. Details about the aeration equipment and testing are given in section 3. A internal aerator was chosen for ease of installation, and the small amount of pipe work required. Locating all equipment inside the tank reduces possible damage from external sources. An external air compressor using dome diffusers were considered unsuitable because the diffusers are prone to blocking from sludge which can enter during the settling stage (Norcross, 1992).

Effluent was withdraw using a pump at a fixed height inside the tank. The tank contents were pumped out until a float switch turned the pump off. A fixed height decant pump was used because there was no minimum suspended solids concentration requirement for the effluent and to speed up installation on site. If suspended solids in the effluent become a problem a floating boom decant system maybe necessary.

To allow draining of the tank reactor a valve was placed near the base. In some situations this valve could be used to waste sludge if a more concentrated solution was required.

4.1.3 Electrical equipment

The pilot scale plant required three pumps and an aerator. The pumps were identical for ease of service if failure occurred. One pump was used for filling the tank, the second pump was located near the bottom of the tank for use as a mixer during anoxic

times, the third pump was used to decant the treated effluent. The pumps were Tsurumi HS3-4, 0.4 kW submersible pumps (Tsurumi Manufacturing Co., Ltd Japan), capable of pumping liquids slurries using an open type impeller.

Operation of the pumps and aerator was controlled by a Micro^3 programmable logic controller (Idec, Idec Izumi Corporation Japan) and two float switches. Float switches were used to limit maximum and minimum water levels during the fill and decant stages. Float switches also protected the aerator and internal pumps from operating below a safe water level. Programming of the controller using a hand held programmer was performed on site with assistance from the supplier. The program consisted of ladder logic control sequence operated from an internal real time clock. The controller allowed the SBR to operate automatically on a eight hour cycle treating approximately 1700 litres per cycle.

During the operation of the SBR aeration was controlled using on-off control. This was achieved by using a signal received from a dissolved oxygen probe in the tank as the input to the controller. To use the signal as an input to the controller, an amplifier and a analog to digital converter were required. The dissolved oxygen data for use in oxygen demand calculations was logged every 15 seconds using a 16 bit data logger (Pico Technology Ltd. Hardwick, United Kingdom.) and IBM 386 SX33 computer.

4.1.4 Inlet pump float design

To fill the SBR each cycle a pump was floated on the anaerobic pond near the outlet pipe.

The float was designed to support the weight of pump on the pond.

Float volume was calculated by equating the volume of water equal to the weight of the pump and float combination. Five kilograms was assumed for the weight of the float and the pump weight was 18 kg. The float volume displace 23 kg of water to float.

A circular float was built with an overall surface area of 0.075 m^2 and depth of 0.3 m. The pump was secured into the float by using 6mm stainless steel rods bolted to the base of the pump. The supporting rods were bent into shape so that the pump was submersed below the minimum safe working depth. To secure the float, lugs were fibre glassed onto the sides for ropes to be connected. The ropes were then tied to two posts on the pond edge. Shown in Figure 23 is the pump float combination installed on the pond.

Figure 23 Pump and float combination on pond

4.2 Process Design

Treatment of dairyshed effluent is a relatively new field of study with little data available about nutrient levels and reaction kinetics. To initially design the process side of the SBR it was necessary to use kinetics data for both dairy and swine wastewater (Lo et al, 1988; Fernandes, 1994). Due to the limited amount of data, nutrient concentrations for anaerobic pond effluent collected in 1993-94, (Mason, 1994) were used as for design time estimates. The anaerobic pond monitored is the same pond used in this study, though the loading entering the pond had increased due to the addition of feed pad wastewater.

To achieve phosphorus removal it is necessary to have a carbon to phosphorus released ratio between 1.2:1 and 11:1 (Fang and Ling, 1995; Rovatti et al, 1995) the 11:1 ratio will be used here. From the literature to achieve denitrification it was necessary to have a BOD:NO_3 ratio of greater than 2.3:1 (Narkis et al, 1979).

Table 9 Anaerobic pond effluent data

Parameter	Concentration average mg/l	Concentration range mg/l
Total BOD_5	159	94-275
Total COD	884	438-1220
Total Kjeldahl Nitrogen (TKN)	172	157-185
Ammonia (NH_3)	143	90-166
Total Phosphorus	25.3	17-29
Dissolved Reactive Phosphorus	16.8	6.4-25.2

From the anaerobic pond effluent data it was possible to calculate the following ratios;

BOD to Total phosphorus = 6.3:1

BOD to NH_3 = 1.11:1 assuming 100% nitrification.

To calculate the amount of phosphorus released, data from Rovatti et al (1995) was used. Rovatti et al (1995) used a synthetic mixture wastewater which had a COD of 550 mg/l and phosphorus content of 7 mg/l giving the COD/P ratio of 60. It was possible to calculate from the data presented a ratio of initial phosphorus to phosphorus released concentration. The average ratio calculated for 9 runs was 4.13:1.

When applying this ratio to dairy effluent, taking into account the initial dilution achieved in the SBR, the estimated phosphorus concentration at the end of the anaerobic phase would be 48 mg/l. During the release of phosphorus COD is consumed at a rate of 11.3 mg COD/mg P released. The amount of biodegradable COD required for P release is 416 mg/l. To determine this value in terms of BOD it was assumed that biodegradable COD is equivalent to BOD.

From the theoretical ratio for denitrification of 2.3:1 mg BOD:mg NO_3 it is possible to calculate the total BOD required for denitrification. Assuming 100% nitrification of ammonia to NO_3 the amount of readily biodegradable BOD required is 329 mg/l.

A first estimate of the total readily available BOD required for nutrient removal from dairyshed effluent is 750 mg/l. The majority of this BOD would be consumed during the anoxic-anaerobic stage at the beginning of the SBR cycle. During the react stage phosphorus would be reabsorbed, ammonia would be nitrified and slowly biodegradable BOD would be removed.

To calculate an appropriate cycle time and stage times first order rate kinetics based on dairyshed wastewater and piggery wastewater were used.

4.2.1 Assumptions;

- Kinetics for dairy waste and swine wastewater are valid
- MLVSS range 1000-10000 mg/l
- Nitrifier fraction depends on BOD/TKN ratio (Metcalf and Eddy, 1991)

Table 10 BOD:TKN ratio effect on nitrifying bacterial fraction

BOD:TKN ratio	Nitrifier fraction
0.5	0.35
1	0.21
2	0.12

4.2.2 Reaction rate kinetics

The first order kinetics used for process design:

Biological carbon oxidation (Lo et al, 1988)

$K_{BOD} = 0.017\ l\ mg^{-1}\ day^{-1}$

Nitrification (Lo et al, 1988)

$K_{NH4} = 1.28\ mg\ N\ g^{-1}\ VSS\ h^{-1}$

Denitrification (Fernandes, 1994)

$K_{DN} = 0.00285\ h^{-1}$

4.2.3 Equations

First order reaction model which includes biomass concentration equation 13,

$$\frac{S_e}{S_O} = e^{-kxt} \qquad 13.$$

Where So = Influent Concentration (mg/l)

S_e = Effluent concentration (mg/l)

x = Active Biomass concentration (mg/l)

k = Reaction rate constant (d^{-1})

t = Reaction time (d^{-1})

Substrate dilution during the fill stage can be calculated using equation 3 below, biological reactions during the fill time were assumed negligible.

$$So_a = So(1-f_T) + f_T S \qquad 14.$$

Daily oxygen requirements are calculated using equation 4

$$O_2\ kg/d = \frac{Q(S_O - S)}{f} - 1.42(P_X) + 4.57Q(N_O - N) \qquad 15.$$

Where: So_a = Actual effluent concentration at end of fill (mg/l)

f_T = Fraction of tank volume remaining after decanting

f = Factor to convert BOD_5 to BOD_U

Q = Volume of effluent treated per day

$N_{NH4\text{-}N}$ = Ammonia concentration mg/l

P_X = Sludge produced per day kg/d

$P_x = y(S_O - S)Q - k_d x$

y = Biomass growth yield

k_d = Biomass decay coefficient

From the rate calculations the following cycle time are required for nutrient removal of dairyshed effluent when adequate readily available carbon is present Table 11.

Calculations to determine cycle time for the SBR are attached in appendix 5.

Table 11 Estimated SBR stage times

Stage	Time (hour)
Fill	0.1
Anoxic (denitrification)	3.4
Anaerobic (Phosphorus release)	1.5
Aerobic (Nitrification, BOD removal)	3.0
Settle	1.0
Decant	0.2
Total cycle time	**9.3**

Phosphorus release time is an estimate from Randall et al (1992).

To determine the cycle time several initial assumptions were required, the validity of these assumptions will be confirmed using experimental data.

Assumptions were;

- The reactor volume decanted each cycle was 40%.
- MLSS concentration of 5000 mg/l.

A decant volume of 40% allowed for a sludge volume index of 110 ml/g sludge. A 40% decant volume retains 60% of the nitrate formed during the react stage to be denitrified at the beginning of the cycle.

MLSS concentration of 5000 mg/l was within the range Martin (1996) used in a pilot scale reactor investigating nitrification systems.

Sizing of aeration equipment was based the daily oxygen demand being supplied evenly during the aerobic react stage. It was recognised that the oxygen demand is not constant in an SBR but will change during the react cycle as nutrients are oxidised. The average oxygen demand required was calculated for an aerobically treated BOD of 750 mg/l was 0.5 kg/hr.

A total BOD of 1500 mg/l was used as an estimate for the oxygen demand because 40-60% of influent BOD is slowly hydrolysable which requires aerobic treatment (Henze, 1992). The other 750 mg/l is associated with the readily biodegradable BOD which is used during the anoxic and anaerobic periods for nutrient removal.

5. Sequencing Batch Reactor Operation

5.1 Introduction

The startup behaviour of a sequencing batch reactor treating dairyshed wastewater using only available pond effluents as a seeds is unknown at present. Very few researchers report startup procedures where no seed sludge is used. Startup without seed sludge was investigated to see if the cost of importing activated sludge could be eliminated. Reduced startup costs would make full scale application of SBR technology in the New Zealand dairy farming industry more feasible.

For bacteria growth it is necessary to have the correct biological conditions such as; supply of nutrients, a stable and a suitable oxygen environment which maybe anaerobic anoxic or aerobic, and pH of reactor contents near optimum levels.

The correct physical conditions are necessary to retain bacteria. It is important to have a good settling biomass or system of separating the solid biomass from the liquid effluent. Factors which effect solids retention are poor settling, foam formation, and the way effluent is removed from the reactor.

5.1.1 Biological conditions

5.1.1.1 Nutrient requirements

The principal nutrients for bacterial growth are carbon, nitrogen and phosphorus also required are substantial quantities of sodium, potassium calcium phosphate, chloride sulphate and bicarbonate (Metcalf and Eddy, 1991). Nutrient supply from the effluent is expected to be sufficient for bacteria growth. BOD:N:P ratio was 40:5:1 (Vanderholm, 1984).

5.1.1.2 Oxygen

Oxygen is consumed by bacteria for growth. During oxidation of carbon approximately 1.2 mg O_2 is consumed per mg BOD (Metcalf and Eddy, 1991). From equations 16 and 17 the approximate oxygen demand during the oxidation of ammonia is 4.3 mg O_2 consumed per mg of ammonia-nitrogen oxidised.

5.1.1.3 Alkalinity

Alkalinity is important in biological nitrogen removal because it consumed during nitrification. The alkalinity results from the presence of hydroxides, carbonates and bicarbonates. The approximate nitrification stoichiometry from Metcalf and Eddy (1991) are shown in equations 16 and 17.

Nitrification reaction for Nitrosomonas is;

$$55NH_4^+ + 76O_2 + 109HCO_3^- \rightarrow C_5H_7O_2N + 54NO_2^- + 57H_2O + 104H_2CO_3 \qquad 16.$$

for Nitrobacter the equation is;

$$400NO_2^- + NH_4^+ + 4H_2CO_3 + HCO_3^- + 195O_2 \rightarrow C_5H_7O_2N + 3H_2O + 400NO_3^- \qquad 17.$$

Alkalinity is consumed during the oxidation of ammonia to nitrate at a rate of 8.64 mg as HCO_3^- per mg of ammonia-nitrogen oxidised.

5.1.1.4 pH effects on reaction rates

Biological reactions are influenced by pH. The optimal pH range for nitrification is 7.2-9.0. The specific growth rate for combined sludge nitrification system treating

domestic wastewater can be calculated using equation 18 from Metcalf and Eddy (1991).

$$\mu = \frac{\mu_{mn}}{[1 - 0.833(7.2 - pH)]} \qquad 18.$$

Where μ = Specific growth rate, t^{-1}

μ_{mn} = Maximum specific growth rate, t^{-1}

5.1.2 Physical conditions

5.1.2.1 Sludge settling

To retain bacterial biomass for nutrient reductions, sedimentation is often used. Sedimentation can either occur in the same reactor, as in an SBR, or may occur separately in clarifier from which the sludge can be recycled back to the reactor.

An index used to measure sludge settlability is the sludge volume index (SVI). SVI is the volume in millilitres occupied by 1g of sludge (Standard Methods, 1995). To attain a low SVI it is necessary to discourage the growth of filamentous bacteria, which is achieved in SBR's due to a changing oxygen environment during the cycle (Albertson, 1987).

5.1.2.2 Foam production

Froth formation is caused by the presence of surfactants and is reduced when the MLSS concentration is high. Large quantities of foam can be produced during startup of a process (Metcalf & Eddy, 1991).

Foam contains sludge solids, grease, and large number of bacteria. This foam is both unsightly and potentially harmful due to it's slippery nature and the bacteria present. Methods for controlling the foam formation include using a water spray and or a antifoaming chemical additive. It was hoped that because the operation of the SBR includes a settle period any antifoam used would rise to the surface of the tank and not be discharged. This would reduce the amount of antifoam required.

5.1.2.3 Effluent withdrawal

Effluent is decanted from an SBR each cycle after a period of settling. The type of decanter mechanism used effects the solids concentrations in the effluent. A floating decanter which removes effluent from just below the surface produces the lowest effluent suspended solids concentrations. Fixed decanting mechanism are often used because of the ease of construction and reduced costs (Ketchum, 1996). The disadvantage of the fixed decanter is that solids can be removed if the sludge has not settled below the decanter. Floating decanters are more complex but allow effluent removal before settling is complete.

The objectives of monitoring the SBR startup process were to determine how easily a nitrifying activated sludge would develop and be retained, and determine the long term stability of the process.

5.2 Startup Procedure

At the time of startup facilities enabling the use of raw settled effluent were not available. In the interim it was deemed useful to startup the reactor using anaerobic pond effluent and establish a nitrifying biomass.

5.2.1 SBR cycle times

The reactor was initially operated on a 12 hour cycle treating approximately 1700 litres of anaerobic effluent per cycle. The amount decanted was equivalent to 40% of the reactor volume. The cycle was later changed to an 8 hour cycle when the three phase power was installed.

The 8 hour sequence started with the anoxic conditions in the reactor for half an hour this included the fill period and part of the react stage. Anoxic conditions were needed for denitrification. Following the anoxic stage were five and a half hour of aerobic react, then one and a half hours for settling, so that sludge produced would be retained.

The final half an hour of the cycle was for decanting the treated effluent and idle. The idle time was longer than required for pumping out because of the concern that pump blockage would slow the pump flowrate causing cycles to overlap. It was found that the pumps did not block and all pumps operated reliability during the study period. This meant that the idle time could be reduced to 5 minutes.

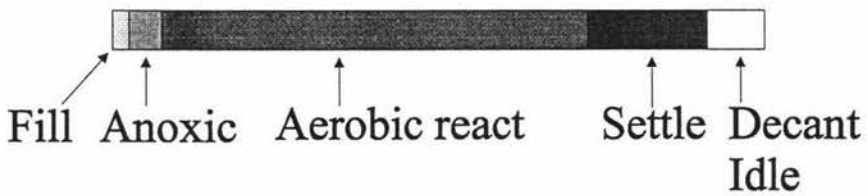

Figure 24 SBR cycle times

Table 12 Summary of startup cycle times used and estimated times.

Stage	Startup time used (hour)	Estimated Time section 4.2(hour)
Fill	0.1	0.1
Anoxic (denitrification)	0.4	3.4
Anaerobic (Phosphorus release)	0	1.5
Aerobic (Nitrification, BOD removal)	5.5	3.0
Settle	1.5	1.0
Decant	0.5	0.2
Total cycle time	**8**	**9.3**

The cycle was largely aerobic so that the environment encouraged the growth of nitrifying bacteria. The times used to treat the anaerobic effluent were different to those calculated in section 4.2 Process design, because of different nutrient concentrations.

The low carbon content of the anaerobic pond effluent meant that the BOD:NO_3 was less than 2.3:1. It was therefore decided to reduce the anoxic time in the cycle initially to 30 minutes, as little denitrification was expected. An anoxic period was still included to reduce the growth of filamentous bulking organisms. No time was allocated in the cycle for phosphorus release because high nitrate concentrations would suppress anaerobic conditions.

The aerobic react stage time is longer than was estimated for BOD reduction and nitrification, so that reaction kinetics could be determined. The extended aerobic react period was expected to reduce the slowly biodegradable BOD and give an indication of the inert COD remain after biological treatment.

5.2.2 Analytical methods

MLSS was determined by filtration through Whatman GFC filter paper as for suspended solids measurement (Standard Methods, 1995)

Sludge volume index (SVI) was determined as described in Standard Methods (1995). Settled sludge volume after 30 minutes was determined using a 1 litre graduated cylinder. To calculate the SVI the equation 19 was used,

$$\text{SVI (ml / g)} = \frac{\text{settled sludge volume (ml / l)} * 1000}{\text{suspended solids (mg / l)}} \qquad 19.$$

During the startup period alkalinity, pH, Ammonia, Nitrate nitrogen and Phosphorus were measured as described in section 2.2.3.

Soluble COD's of the reactor effluent were filtered through a filter with a nominal pore size of 1.2μm (Whatman GFC). COD was measured as described in section 2.2.3.

Dissolved oxygen was measured using a YSI model 58 -230V deluxe digital dissolved oxygen meter and H05521-00 dissolved oxygen field probe (Yellow Spring Instrument Co. INC. Yellow Springs, Ohio, USA.).

5.3 Results and Discussion

The reactor was seeded with a 50/50 mix of anaerobic pond effluent and aerobic pond effluent as the initial batch volume. A mixture of available effluents was chosen to maximise the diversity of bacteria which would reduce the time required to develop an active biomass that could cope with the environmental conditions present in the reactor.

Dissolved oxygen levels during the startup cycle reached a maximum of 8 mg/l which is considerably more than a typical activated sludge plant.

The startup procedure was successful, 17 days after startup nitrification was observed and 31 days after the reactor was started 95% of the inlet ammonia was nitrified reactor outlet ammonia concentration was below 5mg/l. The startup objective to develop and retain a nitrifying biomass had been achieved. The objective to develop stable nitrification was not achieved during the first startup period.

5.3.1 Mixed liquor suspended solids production

During startup MLSS was found to stabilise at around 6000 mg/l between days 24 and 40 of operation (Figure 24). Some biomass during this time was unintentionally wasted with the effluent. Because the MLSS concentration never rose above 6500 mg/l during the study no sludge was wasted intentionally. The MLSS concentration was at the top end of a range 3000-6000 mg/l (Martin (1996) used when investigating nitrification systems treating digested sludge filtrate with a pilot scale reactor.

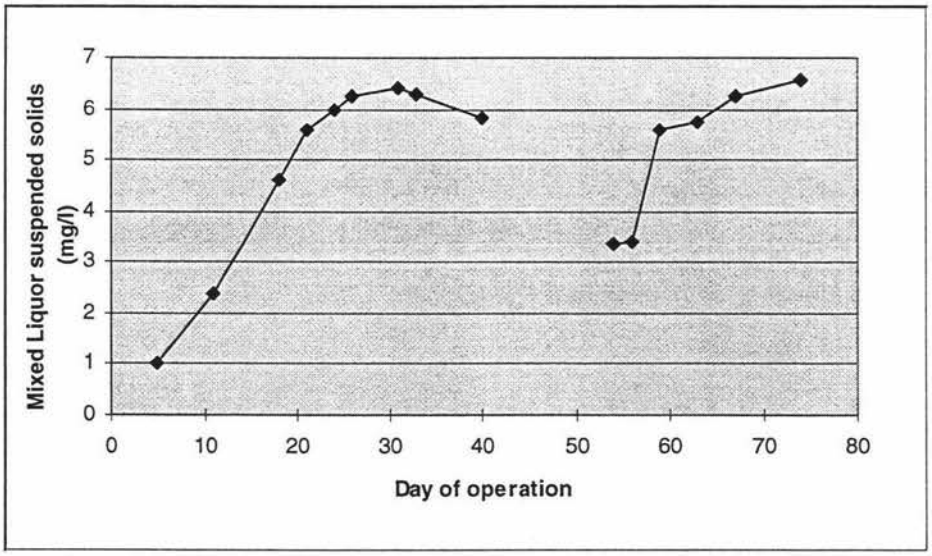

Figure 25 Mixed liquor suspended solids concentration

On day 50 a fault with the dissolved oxygen probe caused aeration to continue during the settle and decant stages for the cycle, resulting in the loss of a large proportion of the biomass from the system. The reactor was started again on day 51. Rapid biomass accumulation over a 5 day period was observed as shown in Figure 25.

On day 35, during the period of stable nitrification, the settled sludge volume was measured during the settling period (Figure 26). The settling time was later reduced from one and a half hours to one hour when it was found that most of the sludge had settled to a volume below 300 ml/l.

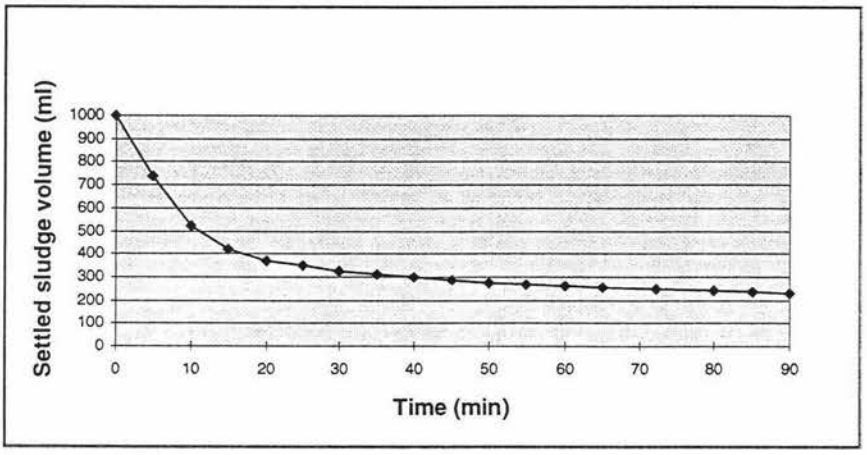

Figure 26 Sludge Settling rate on day 35 of operation

5.3.2 Sludge Volume Index

During the study period the maximum SVI measured was 54 ml/g. Though the sludge showed a low SVI indicating a very good settling sludge, significant amounts of solids were being discharged with the effluent. This was indicated by effluent SS concentration of 300-400 mg/l in the effluent.

The influent colour was a dark brown-green which after treatment had changed to a light brown. Effluent solids contributed to the colour though it was observed that after filtration (Whatman GFC) some colour still remained.

Sludge bulking can be a considerable problem for many activated sludge treatment plants it is therefore pleasing to have a sludge that settled well. If sludge bulking became a problem application of activated sludge technology would be limited as batch reactors are often used for controlling SVI by use of anoxic and aerobic times and swings in the food to microorganism ratio (Albertson, 1987).

5.3.3 Alkalinity and pH

As nitrification performance increased the amount of alkalinity consumed per cycle also increased. Alkalinity consumption caused a reduction in the pH buffering capacity of the effluent. The reactor contents decreased to pH 6.5 which is below the range 7.2-9.0 for optimum growth of nitrifying bacteria. The change in pH during the startup period is shown in Figure 27. From equation 20 at pH 6.5 the rate of nitrification is 40% lower than at pH 7.2 for domestic wastewater.

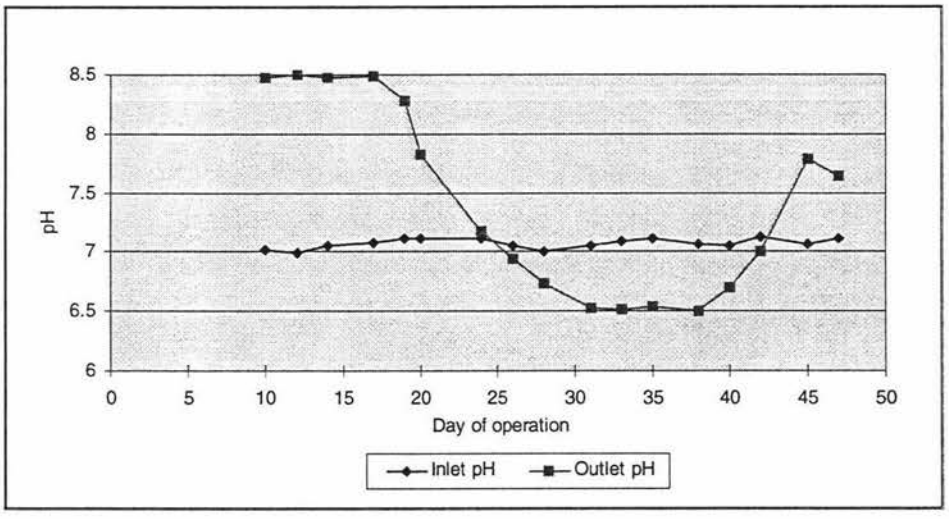

Figure 27 pH during SBR startup

Figure 27 shows that the pH reached a minimum of 6.5 between days 31 and 38. After day 38 pH began increasing, this maybe attributed to a decline in the nitrification performance and thus a increase in the available alkalinity. The changes in alkalinity

are shown in Figure 28. At the end of the startup period day 31 total alkalinity consumption was at a maximum.

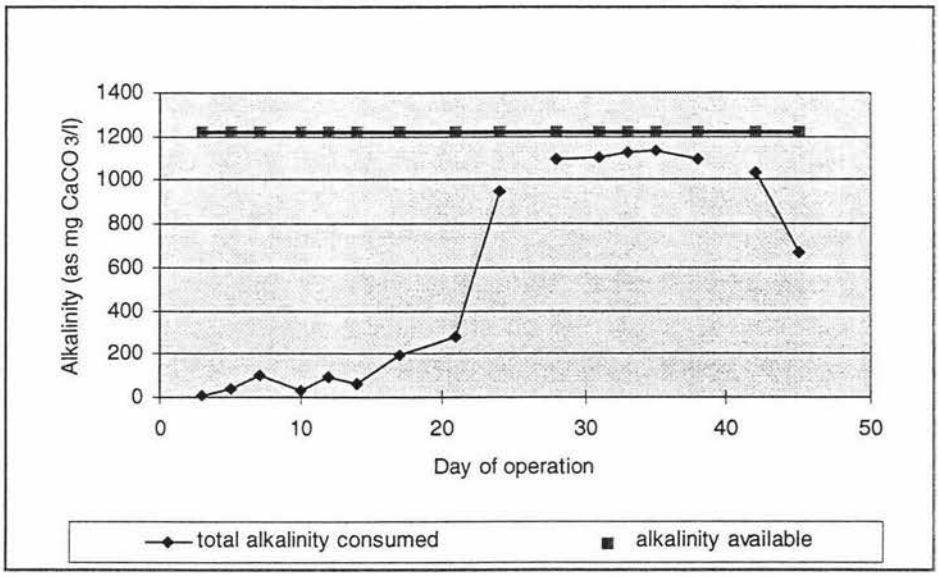

Figure 28 Alkalinity consumed during startup

5.3.4 Nitrification performance

The nitrifying capacity of the biomass was confirmed by the reduction in NH_4 and appearance of NO_3 after 14 days of reactor operation. After 28 days the NH_4-N was reduced from 196 mg/l to 7.5 mg/l. The nitrifying performance was related to pH and it can be seen that performance decreased after the startup period. The decrease in nitrification was due to low alkalinity concentrations in the reactor. The nitrification performance during the startup period is shown in Figure 29.

From a mass balance of nitrogen species it was concluded that denitrification did occur during the treatment cycle. Nitrogen deficit between inlet ammonia concentration, outlet ammonia and nitrate concentrations on day 31 amounted to 60 mg/l, 30% of the inlet nitrogen. The mass balance did not include nitrite nitrogen.

Nitrogen can be removed by bacterial growth, ammonia volatilisation and denitrification. The amount of nitrogen required for bacterial growth and lost via

volatilisation to the atmosphere was estimated at 25 mg/l NH_4 from the reduction in NH_4 on day 5 of the startup period when there was no nitrification. Thus the extra nitrogen removal between days 25 and 40 may have been due to bacterial denitrification. The amount of denitrification that could have occurred was limited by the amount of readily available carbon in the feed and the length of the anoxic stage in the treatment cycle. At times the pH was below the optimal for bacterial denitrification to occur.

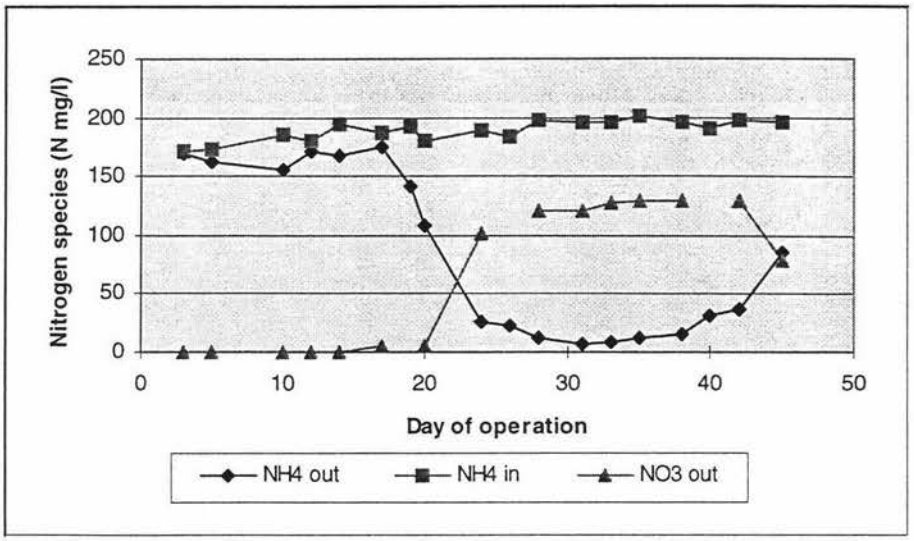

Figure 29 Nitrogen species during startup

5.3.5 Phosphorus reduction

Phosphorus removal during this study was limited to that required for bacterial growth. No enhanced biological phosphorus removal was observed because the reactor contents did not go anaerobic due to high nitrate concentrations. The low availability of carbon also limited the amount of phosphorus adsorbed. Figure 30 shows changes in phosphorus content of the effluent during the study period.

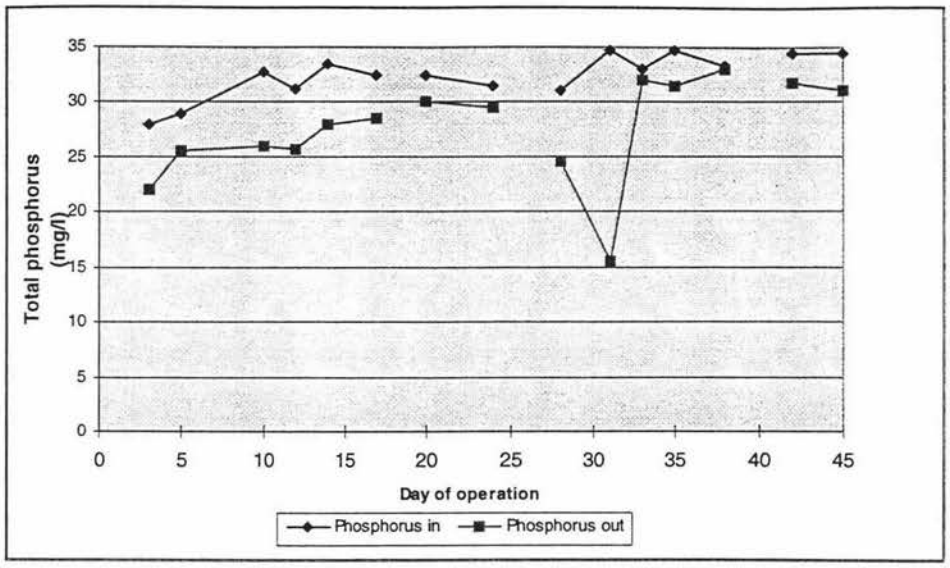

Figure 30 Phosphorus removal during the startup period

5.3.6 COD reduction

COD reduction during the time of operation of the SBR was limited. The highest percentage reduction was 80% and the average reduction was 60%. The aeration time of five and half hours gave an outlet COD of approximately 400 mg/l. Figure 31 shows the reduction between inlet COD and outlet COD.

From the work in section 2, characterising the anaerobic pond effluent, it was possible to estimate the inert soluble COD of the anaerobic pond effluent at 407 mg/l (sd 135). This was equivalent to the SBR outlet soluble COD, indicating that the SBR was removing all of the biodegradable material present.

It was suspected that because the wastewater was from animals fed lignin type material and was anaerobically pretreated, there would be a high residual COD, as found by Subramaniam et al (1994), Comeau et al (1996) and Ng (1987).

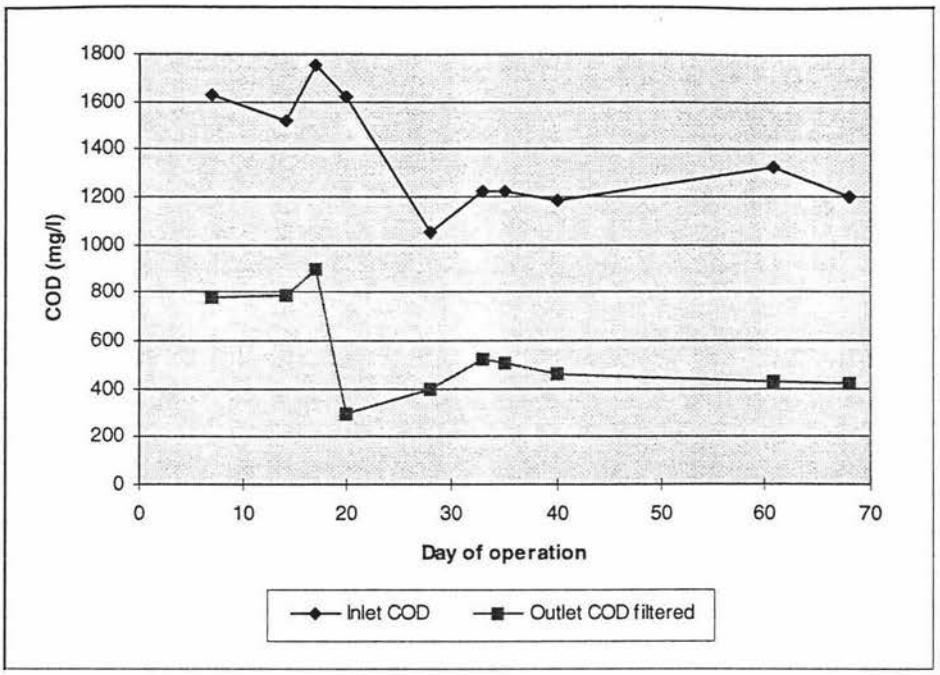

Figure 31 COD removal during SBR operation

5.3.7 Foam formation.

Foaming problems were observed when the reactor was aerated continuously during the react stage. Two antifoaming agents were tried Gensil 832 and Bevaloid 6618 (Rhone Poulenc Chemicals. Wingate, Lower Hutt, New Zealand.). Both products were successful in reducing foam production but it was necessary to dose chemicals each batch. It was also found that during the cycle monitoring when the aeration was controlled, foam production decreased to levels that were acceptable. This means that when large scale SBR's are developed for dairy effluent treatment a form of aeration control will be necessary. Aeration control could be achieved by using dissolved oxygen sensors or a simple timer sequence for both the aerator and the mixer.

5.4 Maintenance of Nitrification Activity

The reduction in nitrification activity after day 38 was due to a lack of alkalinity. From day 51, 500g of Na_2CO_3 was added daily to maintain the reactor alkalinity and pH. By buffering the alkalinity stable nitrification performance was achieved, illustrated in Figure 32 by the low outlet ammonia concentration. The recovery in nitrification activity took 5 days. The effluent ammonia concentration after 5 days was approximately 5 mg/l and effluent nitrate concentration was approximately 115 mg/l. From this work it was shown that addition of alkalinity was required for stable nitrification of dairyshed effluent. Addition of alkalinity allowed the objective of stable nitrification performance to be achieved.

A mass balance for nitrogen using the same assumptions as above indicated that denitrification was occurring on a limited scale. Again, denitrification may have been limited by several factors. The anoxic time which had been increased from 30 minutes to 1 hour was not long enough to allow complete denitrification, and simultaneous nitrification-denitrification could not occur because the dissolved oxygen concentrations were greater than 4 mg/l (Munch et al, 1995).

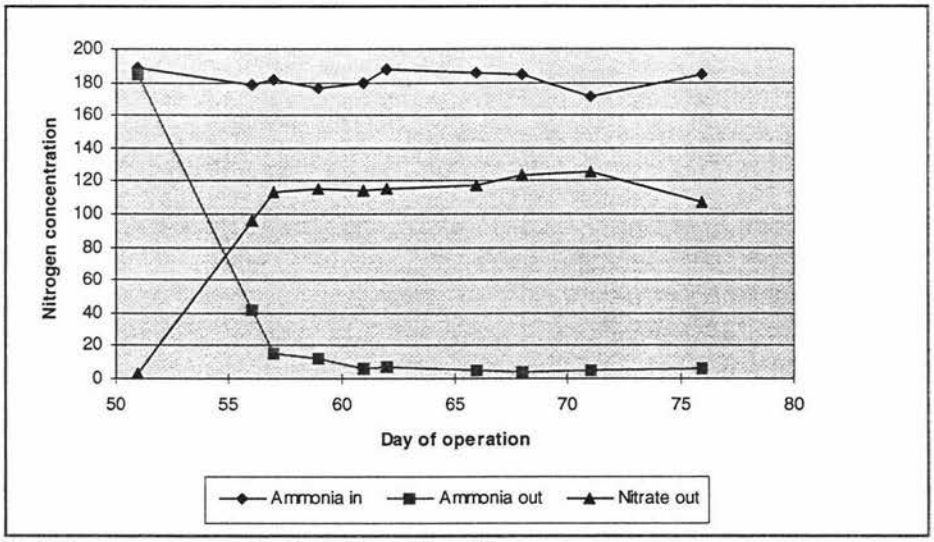

Figure 32 Nitrogen concentration from day 51.

5.5 Conclusions

Startup procedure which used a 50/50 mixture of anaerobic pond and aerobic pond effluents was used successfully as a seed in establishing a biomass capable of nitrifying anaerobic pond effluent. Startup time to establish a nitrifying population was 17 days, after 31 days maximum nitrification performance was achieved.

During startup the MLSS concentration reached 4600mg/l by day 20 and stabilised near 6000mg/l for 20 days. Depletion of biomass occurred on day 50 but within 5 days of the second startup the biomass had increased from 3200 mg/l to 5800 mg/l. The sludge settlability was good with a maximum SVI measured during both startup periods of 54 ml/g. Sludge bulking was not a problem.

After the second startup nitrification performance also returned in 5 days. Long term stable nitrification was not achieved during the first startup period. During the second startup period alkalinity was added to supplement the amount consumed during nitrification and stable nitrification was achieved.

Once alkalinity was added daily an effluent with 5 mg/l ammonia was produced reliably until the end of the study period.

From this work it can be seen that nitrification of dairy effluent to reduce the potential toxic effects of ammonia on aquatic organisms was possible provided alkalinity availability is maintained.

6. Cycle monitoring

6.1 Introduction

A sequencing batch reactor treats a discrete batch of effluent each cycle. This makes it possible to monitor the treatment cycle from the beginning fill stage through to effluent discharge for the same batch of effluent. By monitoring a single batch fluctuations in feed quality during the monitoring period are eliminated.

Prediction of oxygen transfer rates in wastewater is based on K_La values generated for aeration of clean water, which are then corrected for changes in mixing intensity, geometry of mixing chamber and constituents in the wastewater. The effects of mixing intensity and tank geometry are corrected for by an α factor

$$\alpha = \frac{K_L a(\text{wastewater})}{K_L a(\text{clean water})} \qquad 20.$$

The effects of wastewater characteristics are corrected for by a β factor.

$$\beta = \frac{C_s(\text{wastewater})}{C_s(\text{clean water})} \qquad 21.$$

Determination of correction factors allows actual oxygen transfer rate to be predicted for the wastewater.

The objectives of the cycle monitoring were to determine the reaction times, the amount of oxygen required and effluent concentrations of nutrient species, such as ammonia, nitrate, COD, phosphorus and alkalinity.

6.2 Methods

6.2.1 Sample collection.

Sample collection was every two minutes during the fill stage. After the fill stage had finished sampling frequency decreased to five minute intervals. Grab samples were collected from the reactor in 120 ml containers. Samples of the discharge were collected from the discharge pipe.

Immediately after collection sample pH and temperature were measured. Samples were than placed on ice to slow any biological reactions and routinely transferred from the field site to the laboratory where they were stored at 4°C before analysis.

6.2.2 Sample Analysis

COD analysis occurred immediately after sample collection was complete. A sample volume of 20 ml was centrifuged at 3000 rpm (1080 rcf) for 15 minutes to remove the bulk of the solid material. To facilitate rapid analysis centrifugation was chosen instead of filtration because the large biomass concentration meant that filtration was a slow process.

Analysis for COD, total nitrogen, and total phosphorus used the supernatant from the centrifuged samples to reduce the biomass interference. Total phosphorus in the supernatant was measured because phosphorus is transferred from the soluble phase to solid phase where it is accumulated in cell material. No reduction in total phosphorus would be seen if an analysis was conducted on the uncentrifuged sample.

All experimental techniques used are described in section 2.2.3.

Bacterial oxygen demand for the first cycle was calculated by measuring the rate of oxygen depletion when the aerator was turned off.

Dissolved oxygen concentration data for the second cycle was collected every 15 seconds during the cycle.

Bacterial oxygen demand was calculated over 15 second intervals for the react stage using the dissolved oxygen concentration data with the following relationship

$$\frac{dC}{dt} = K_L a(C_s - C) - r_M \qquad 22.$$

Where $\frac{dC}{dt}$ = Change in dissolved oxygen concentration with time

$K_L a$ = Reaeration coefficient in wastewater

C_S = Dissolved oxygen saturation concentration

C = Dissolved oxygen concentration

r_M = Rate of oxygen use by reactor biomass

It was assumed that the change in dissolved oxygen concentration over the interval was small and close to zero then $\frac{dC}{dt} \cong 0$ and equation 22 could be solved for r_M.

$$r_m = K_L a(C_S - C) \qquad 23.$$

From the oxygen data collected during cycle one $K_L a$ for the wastewater-biomass combination was determined by measuring the rate of reaeration near the end of the react stage when oxygen demand was low and conditions were assumed to be at steady state. It was assumed the demand was due to endogenous decay. The actual oxygen demand at this time was calculated by measuring the rate of oxygen consumption with time. The demand was then added to the $K_L a$ calculated from reaeration data.

The saturation concentration (C_S) of the system was estimated from the oxygen concentration at the end of the react stage which had been continuously aerated.

6.3 Results

A cycle analysis was conducted on two batches, day 47 and day 71.

6.3.1 Cycle One

The first cycle analysis was conducted on day 47. It was initially hoped to conduct the cycle analysis on day 31 when nitrification was occurring, but sampling was delayed because equipment for collection of oxygen concentration data was not available.

6.3.1.1 Oxygen demand

The oxygen demand data collected gave useful information about the oxygen demand when there was limited biological activity. The oxygen demand represents BOD removal and endogenous decay. No nitrification oxygen demand was exerted. Figure 33 shows the oxygen demand during cycle one.

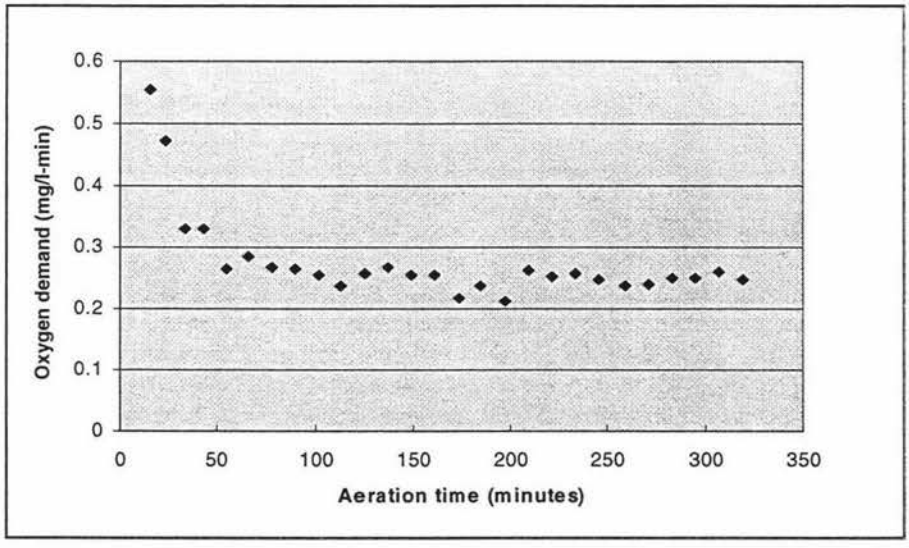

Figure 33 Oxygen demand for first cycle analysis

From the oxygen demand data the reaeration constant for the dirty water system was calculated. The average K_La_{20} for the dirty water was 11.10 ±1.27 hr^{-1}. This gave α

and β values of 0.58 and 0.88 respectively. The value for α was at the low end of the range 0.6-1.2 given by Metcalf and Eddy (1991) for mechanical aeration equipment. The β value was within the range 0.7-0.98 given by Metcalf and Eddy (1991).

It was thought that the low α was due to a increased resistance to oxygen transfer caused by the large amount of solids in the reactor that were not present in the clean water.

6.3.1.2 Nitrification performance

Because the performance of the reactor had decreased by the time analysis of the cycle was preformed very little nitrification occurred (Figure 34). The ammonia reduction between the inlet and the outlet concentrations is only 35 mg/l or 20%. This reduction in ammonia can be attributed to consumption during cell growth and losses to the atmosphere by volatilisation.

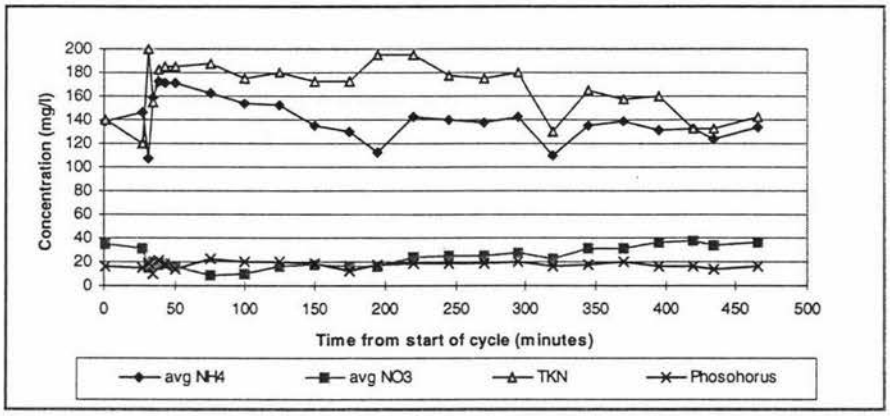

Figure 34 Nitrogen species and phosphorus during monitoring of cycle one.

Figure 34 also shows no significant change in phosphorus concentration during the cycle. This was expected because the high nitrate concentration prevented the reactor contents going anaerobic during the fill stage.

6.3.2 Cycle Two

By day 71 the reactor nitrification performance had increased and a second cycle analysis was conducted.

Sample collection and analysis details were the same as for cycle one.

6.3.2.1 pH

Figure 35 shows the pH of the reactor contents during the cycle analysis. The initially high pH of 8 was due to addition of 500g $NaCO_3$ to supplement the alkalinity consumed during nitrification. During the fill stage pH was reduced by the influent and as denitrification occurred releasing alkalinity, there was an increase in pH. pH reaches a maximum of 7.4 at 11:00 and then decreased as nitrification consumed alkalinity.

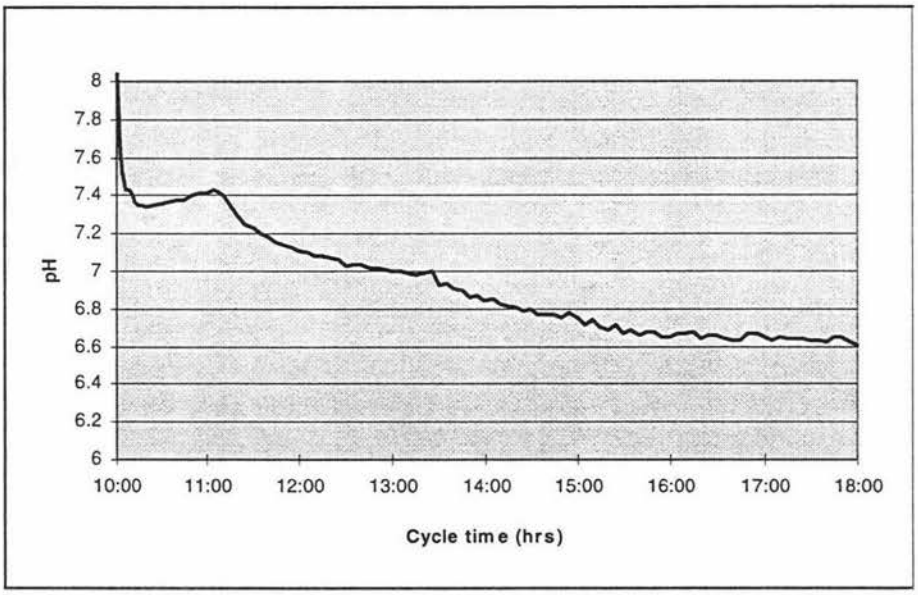

Figure 35 pH of reactor contents during cycle two.

6.3.2.2 Oxygen demand

The saturated dissolved oxygen concentration of 8 mg/l was estimated from Figure 36 for calculation of oxygen demand using in equation 23.

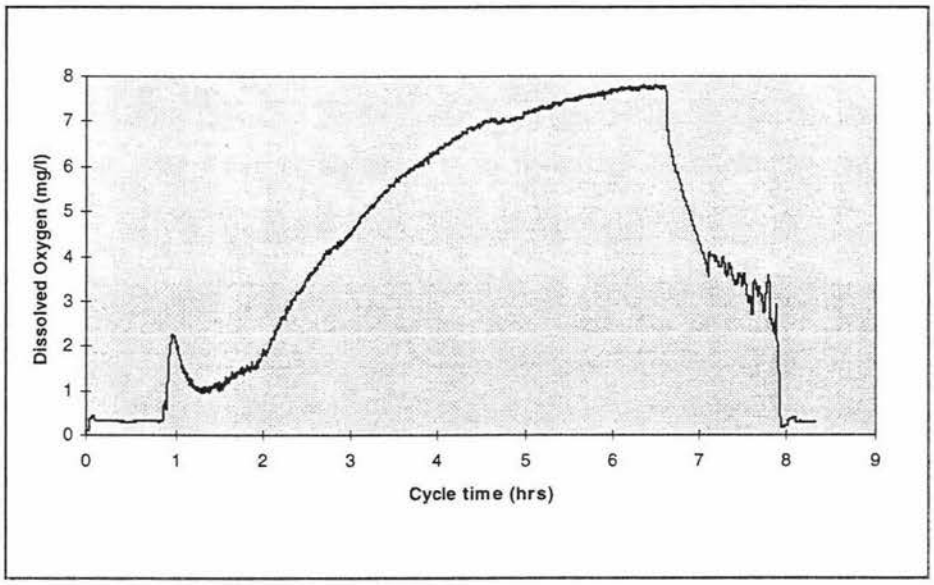

Figure 36 Reactor dissolved oxygen concentration without aerator set point control

Figure 37 shows the dissolved oxygen concentration during the react stage when the aerator was controlled between setpoints 0.5 mg/land 2 mg/l. The oxygen demand for the whole react stage could not be determined using Figure 37 because initial demand exceeded oxygen supplied, which meant that the aerator ran constantly and there was no oxygen depletion data. Dissolved oxygen data gathered near the end of the react stage was used to confirm the K_La value, 11.10 hr^{-1} calculated from data gathered during cycle one.

Figure 38 shows the biological oxygen demand calculated with equation 23. The initially high demand, due to the low oxygen concentration, decreased as reaeration began. After a short lag period oxygen demand increased due to bacterial activity oxidising carbonaceous material, ammonia and organic nitrogen. The rate of oxygen consumption exceeded the oxygen supply rate for approximately half an hour after aeration began. As carbonaceous demand decreased, the rate of aeration exceeded total

oxygen demand allowing the dissolved oxygen concentration in the reactor to increase, although there was still significant demand from nitrification for at least 2 hours.

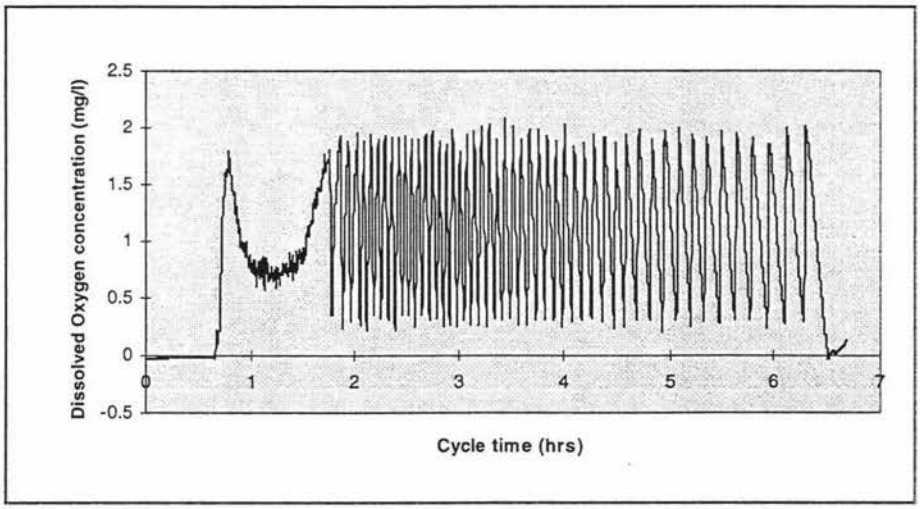

Figure 37 Reactor dissolved oxygen concentration with set point control of aerator

The oxygen demand from nitrification was confirmed by a linear decrease in ammonia nitrogen concentration for 2-3 hours after aeration began (Figure 40).

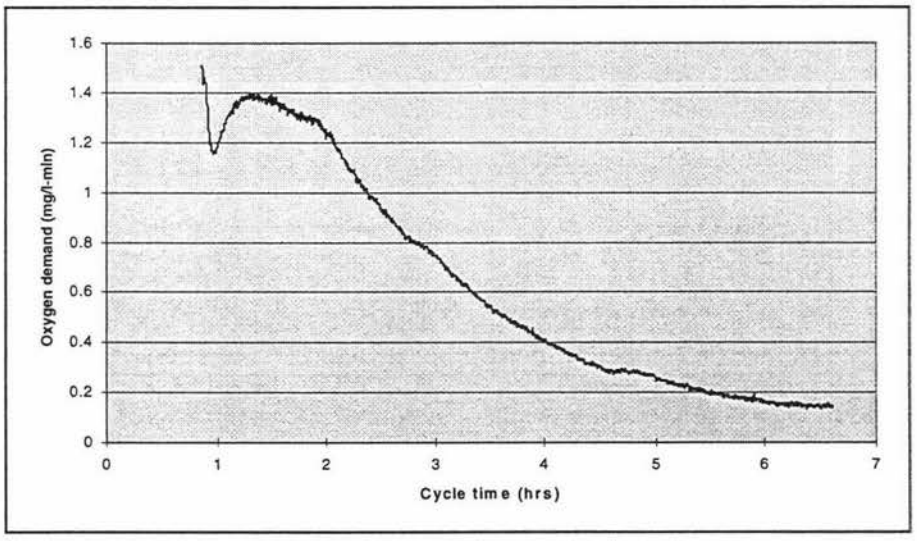

Figure 38 Oxygen demand during cycle two.

6.3.2.3 Nitrification

Figure 39 shows the total nitrogen concentration during the cycle. The first four data points represent the increase in total nitrogen concentration during the fill stage. At the end of the fill period the total nitrogen concentration was a maximum which subsequently decreased as nitrification occurred. Six hours after the cycle began, 75% of the total kjeldahl nitrogen had been removed.

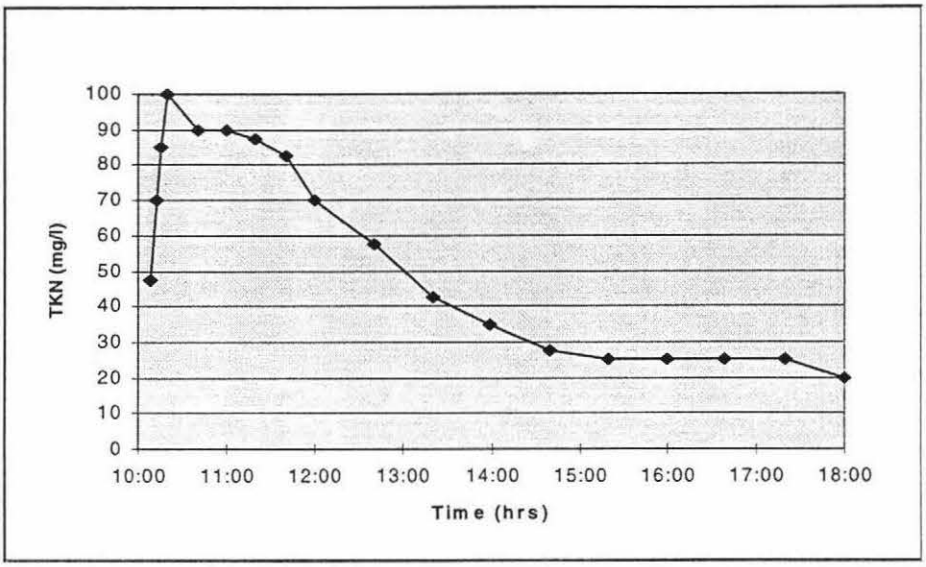

Figure 39 Total Kjeldahl Nitrogen during cycle two

Figure 40 illustrates the conversion of ammonia nitrogen to nitrate nitrogen during the cycle. The initial increase in ammonia and decrease in nitrate concentrations was due to influent being rich in ammonia and low in nitrate compared to the reactor contents retained from the previous cycle. The same graph shows that during the anoxic period, (10:00 to 11:00) little denitrification occurred as the nitrate curve during this time is relatively flat, after 11:00 aeration began and nitrification occurred. Nitrification was indicated by the decrease in the ammonia curve and an increase in the nitrate curve.

After three hours of aeration 88% of the ammonia was removed as estimated from the cycle time calculations in section 4.2. The first order rate constant for ammonia

reduction between 11:00 and 14:00 was 0.71 hr^{-1}. The rate constant for TKN reduction was 0.40 hr^{-1} and nitrate formation was 0.21 hr^{-1}.

First order reaction rate constants were calculated from the slopes of the straight line plots of -ln (C/C_s) vs. Time.

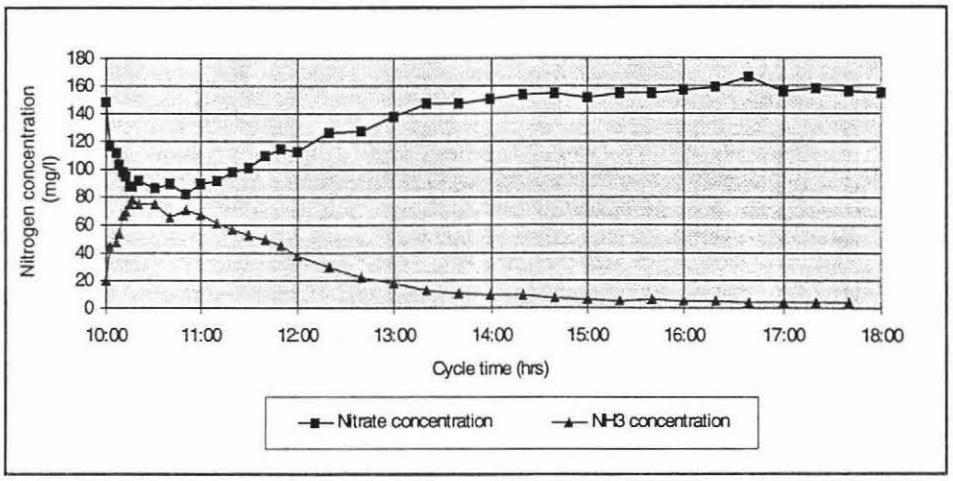

Figure 40 Ammonia and Nitrate concentrations during cycle two.

6.3.2.4 Phosphorus

Figure 41 shows the total phosphorus concentration during cycle two. There was very little change in phosphorus concentration during the cycle. Phosphorus removal was not expected because the reactor contents did not go anaerobic during the cycle due to high nitrate nitrogen concentrations.

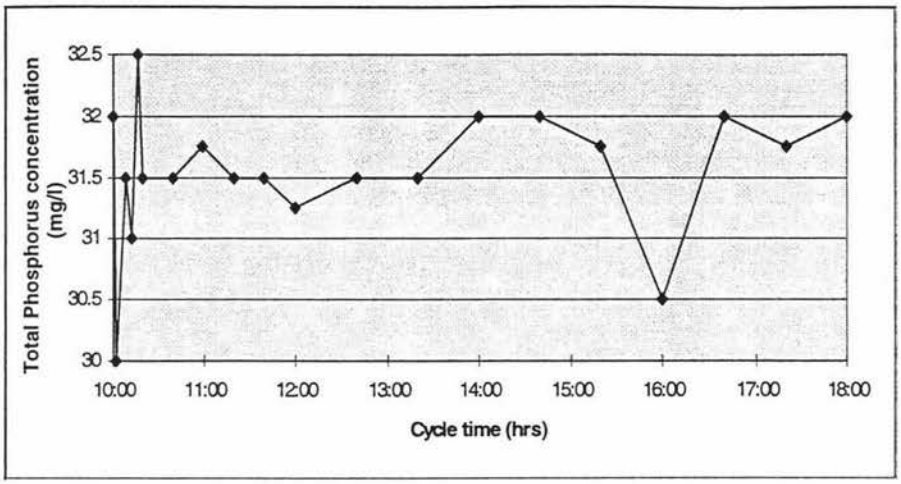

Figure 41 Total Phosphorus concentration during cycle two.

6.3.2.5 COD analysis

Figure 42 shows the COD concentration during cycle two. The inlet COD concentration was 1160 mg/l. Dilution during the fill stage decreased this to around 680 mg/l. There is no trend for COD removal illustrated in Figure 42 due to possible cell material remaining after centrifugation. Filtration or centrifugation at a high speed may be a better procedure for determining soluble COD, this was not confirmed.

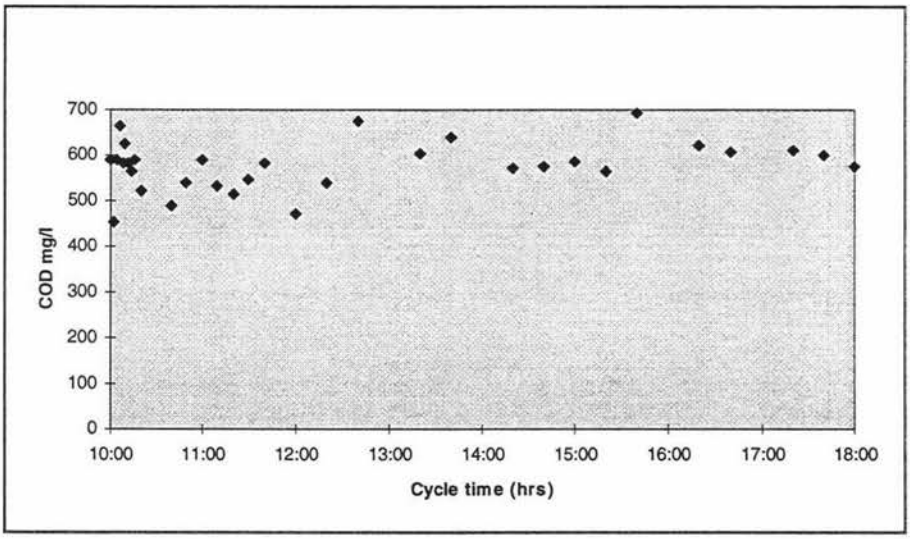

Figure 42 COD concentration during cycle two.

6.4 Conclusions

From the cycle analysis it can be seen that the size of the aeration equipment was adequate to meet the peak biological oxygen demand of 0.37 kg/hr for the wastewater system. The K_La_{20} for the submersible aerator was 11.1 hr^{-1}.

First order reaction rate constants for nitrification were; ammonia reduction 0.7 hr^{-1}, TKN reduction 0.4 hr^{-1} and nitrate formation 0.2 hr^{-1}. The nitrification time of 3 hours for an 88% reduction was the similar to the 3.4 hours estimated from Lo's (1988) data for piggery waste. This confirms that the assumption to use Lo's (1988) data was valid.

No significant phosphorus removal was observed during the cycle analysis because anaerobic conditions did not develop due to high nitrate concentration.

7. Recommendations

Future research is needed in the areas of process development work with the SBR, and identification of the chemical nature of residual COD and complete dairyshed effluent treatment systems.

7.1 Process development work

Process development work should focus on determining the optimum inlet concentration of waste so that there is enough readily biodegradable carbon available for both phosphorus release and denitrification.

Research is needed to optimise the cycle times to achieve target nutrient removal. Changing cycle stage times to include a longer anoxic time for denitrification will take advantage of the increased BOD. By enhancing denitrification, a true anaerobic period will be developed where phosphorus will be released ready for luxury uptake by bacteria during the aerobic react stage.

Investigation of the settling characteristics of the waste is needed to determine why there is such a high suspended solids concentration in the effluent leaving the reactor. Controlling the mixed liquor suspended solids concentration by actively wasting sludge and the development of a floating decanter may help reduce the suspended solids in the effluent.

7.2 Residual COD investigation

Investigation into the inlet wastewater COD chemical characteristics will give information about the probability of achieving target effluent nutrient concentrations. The effect of different effluent pretreatment on the final effluent nutrient concentration requires further investigation.

7.3 Total system for dairy effluent treatment

A total treatment system which requires investigation is the use of a short retention time anaerobic pond for effluent pretreatment before the SBR. The small pond would reduce nutrient concentrations by sedimentation and ferment slowly hydrolysable carbon material to form readily biodegradable carbon. Treatment of this effluent would require less energy in the form of aeration than the yard effluent and produce an effluent lower in nutrients than treatment of the anaerobic pond effluent.

The effluent from the SBR could be polished using wetland before discharge to water. The wetland would remove phosphorus and nitrates from the effluent and buffer the flow into the receiving environment.

8. Conclusions

This research had two main study objectives; to describe the effluent produced from the dairyshed holding yard and anaerobic pond, and to develop a sequencing batch reactor treatment system for the removal of nitrogen and phosphorus.

In order to accomplish the first objective regular grab samples were collected from the dairyshed holding yard and the anaerobic pond outlet. Both yard and anaerobic pond effluent have high nutrient levels when compared to domestic sewage.

The carbon characterisation showed that there was a large difference in the proportion of carbon in each fraction between dairyshed effluent and domestic effluent.

When treating dairyshed wastewater to reduce BOD, nitrogen and phosphorus concentrations it was not possible to treat either the yard effluent or the anaerobic effluent without addition of external materials. The yard effluent was lacking in alkalinity for nitrification, but contained large amounts of readily available carbon (1850 mg/l) which is needed for denitrification and phosphorus removal. Anaerobic pond effluent had much lower levels of readily available carbon (180 mg/l), but alkalinity levels would be adequate for nitrification, if denitrification was complete. Due to the low readily available BOD to nitrate ratio it was expected that less than 40% of the nitrate present would be denitrified. Phosphorus removal would not occur with out addition of carbon.

The BOD reaction rate constant K for the yard effluent at 0.2 d^{-1} was similar to a typical domestic wastewater value of 0.23 d^{-1}. The anaerobic pond effluent reaction rate constant K was 0.16 d^{-1} which is lower than the yard effluent value indicating that the anaerobically treated effluent was hard to aerobically degrade.

A significant amount of the BOD in both effluents was from VFAs. 24% of the anaerobic pond BOD and 9% of the yard effluent BOD was from VFAs.

The objective to develop a sequencing batch reactor for the removal of nitrogen and phosphorus was partly achieved with further work required to optimise the cycle times. The first part of the work was to test the aeration capacity of a locally built venturi aerator and an alternative submersible aerator. From the aeration testing it was determined that the venturi pump combination had poor oxygen transfer efficiency at 0.3 kgO_2/kWh. The venturi aerator's dirty water mass transfer of oxygen at 0.46 kg/hr was less than the estimated 0.5 kg/hr oxygen requirement for treatment of dairyshed effluent to remove nitrogen and phosphorus. The submersible aerator had an efficiency of 1.03 kg/kWh. The mass transfer of oxygen at 0.68 kg/hr exceeded the estimated requirements for biological treatment of dairyshed effluent. The submersible aerator was chosen for use in the pilot scale reactor.

An SBR design which consisted of a three chamber tank was compared theoretically to that of a single chamber circular tank. It was determined that a single chamber tank treating a pretreated effluent would be suitable for the treatment of dairyshed effluent. The circular tank for use in the pilot scale was estimated to be $1400 cheaper to construct than equivalent three chamber SBR.

The single tank SBR was operated for 75 days. The startup procedure used a 50/50 mixture of anaerobic pond and aerobic pond effluents and was successful in establishing a biomass capable of nitrifying anaerobic pond effluent. Startup time to establish a nitrifying population was 17 days, after 31 days maximum nitrification performance was achieved.

During startup the MLSS concentration reached 4600mg/l by day 20 and stabilised at 6000mg/l for 20 days. Depletion of biomass occurred on day 50, but within 5 days of the second startup the biomass increased from 3200 mg/l to 5800 mg/l. The sludge settlability was good with a maximum SVI measured during both startup periods of 54 ml/g. Sludge bulking was not found to be a problem.

After the second startup nitrification performance also returned in 5 days. Long term stable nitrification was not achieved during the first startup. During the second startup, alkalinity was added to supplement the alkalinity consumed during nitrification, and stable nitrification was achieved.

Once alkalinity was added daily an effluent containing 5 mg/l ammonia was produced reliably until the end of the study period.

From the cycle analysis it was found that the size of the aeration equipment was adequate to meet the peak biological oxygen demand of 0.37 kg/hr for the wastewater system. The K_La_{20} for the submersible aerator was 11.1 hr^{-1}.

From the cycle analysis the first order reaction rate constants for nitrification were; ammonia reduction 0.7 hr^{-1}, TKN reduction 0.4 hr^{-1} and nitrate formation 0.2 hr^{-1}. These constants could be used in future work to estimate stage times.

No significant phosphorus removal was observed during the cycle analysis because anaerobic conditions did not develop due to high nitrate concentration.

Future research work is required to optimise the treatment of dairyshed effluent using an SBR so that high nutrient removals can be achieved. This will give farmers an alternative to land application and pond systems for effluent disposal.

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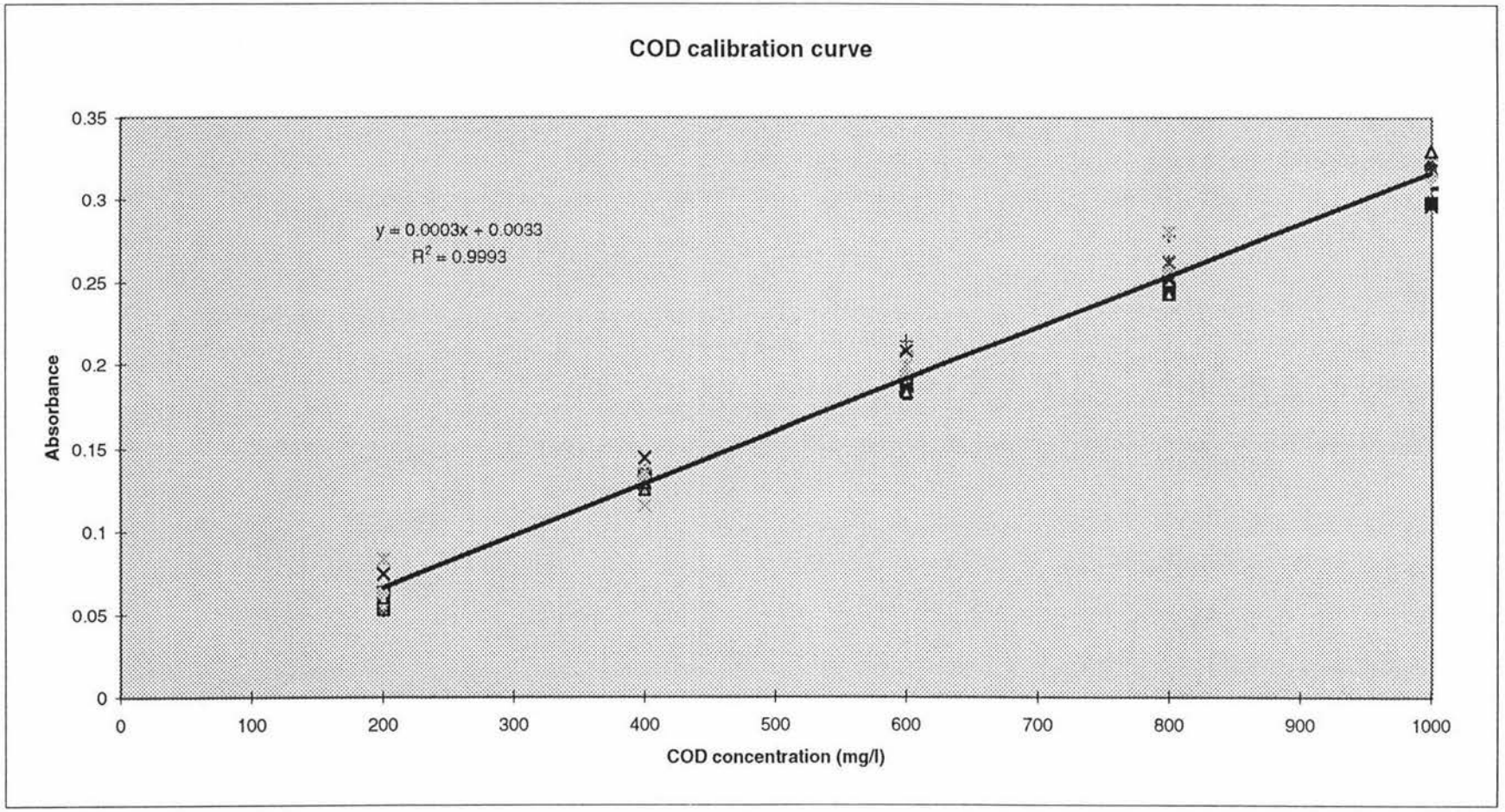
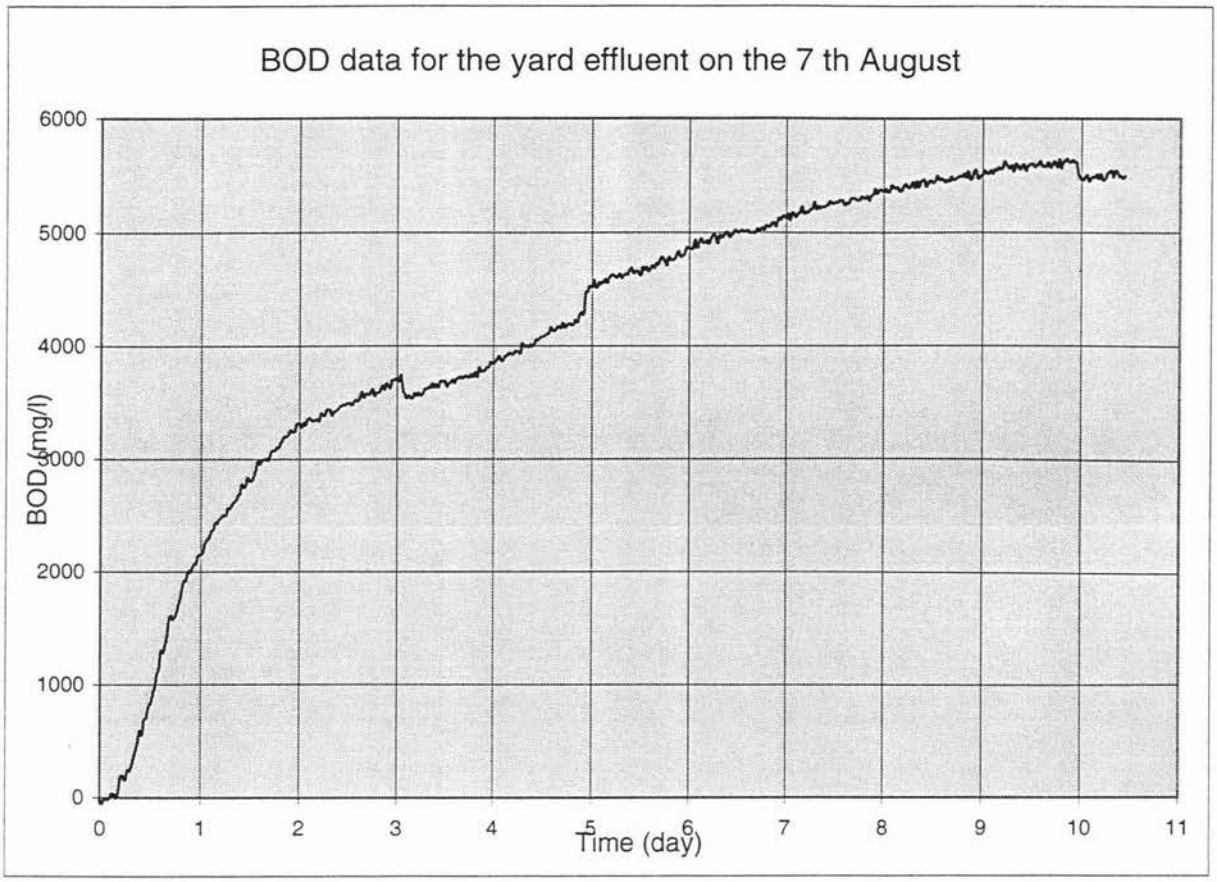
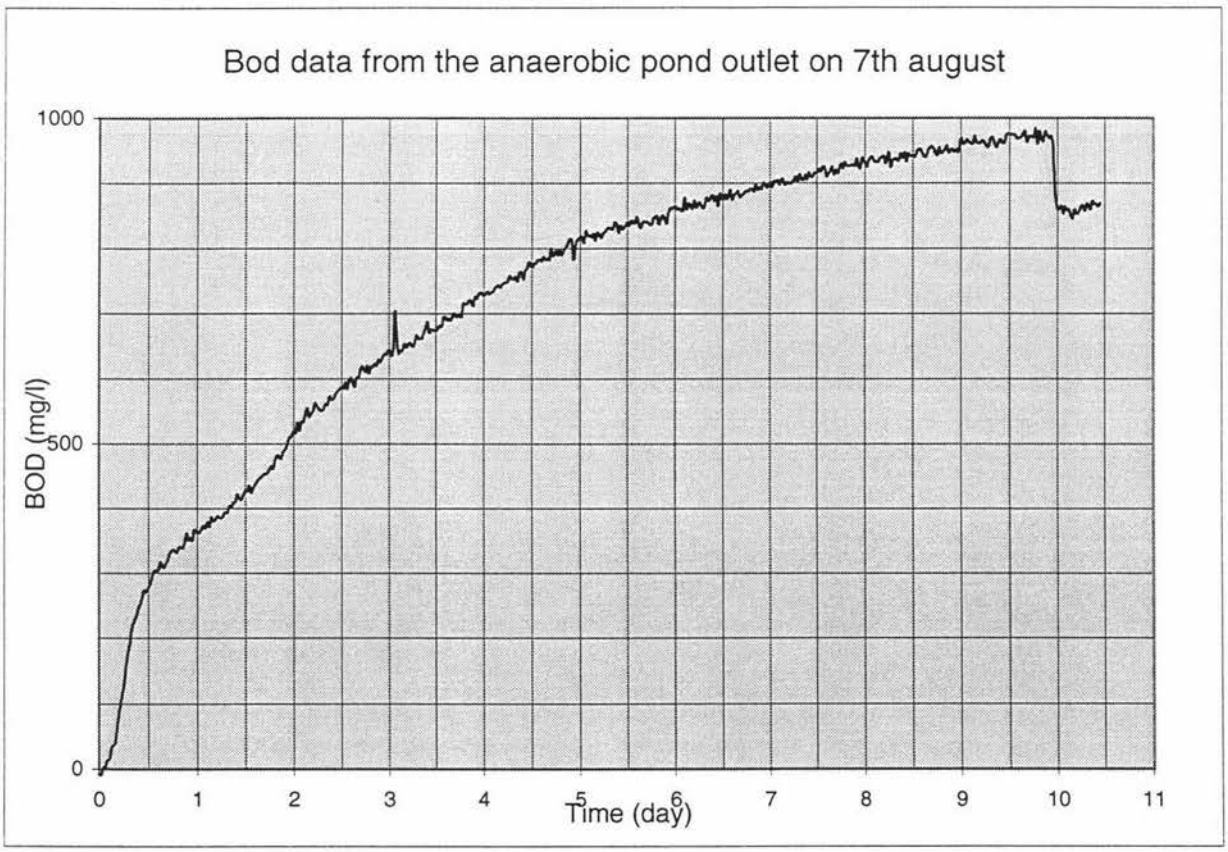
10. Appendices

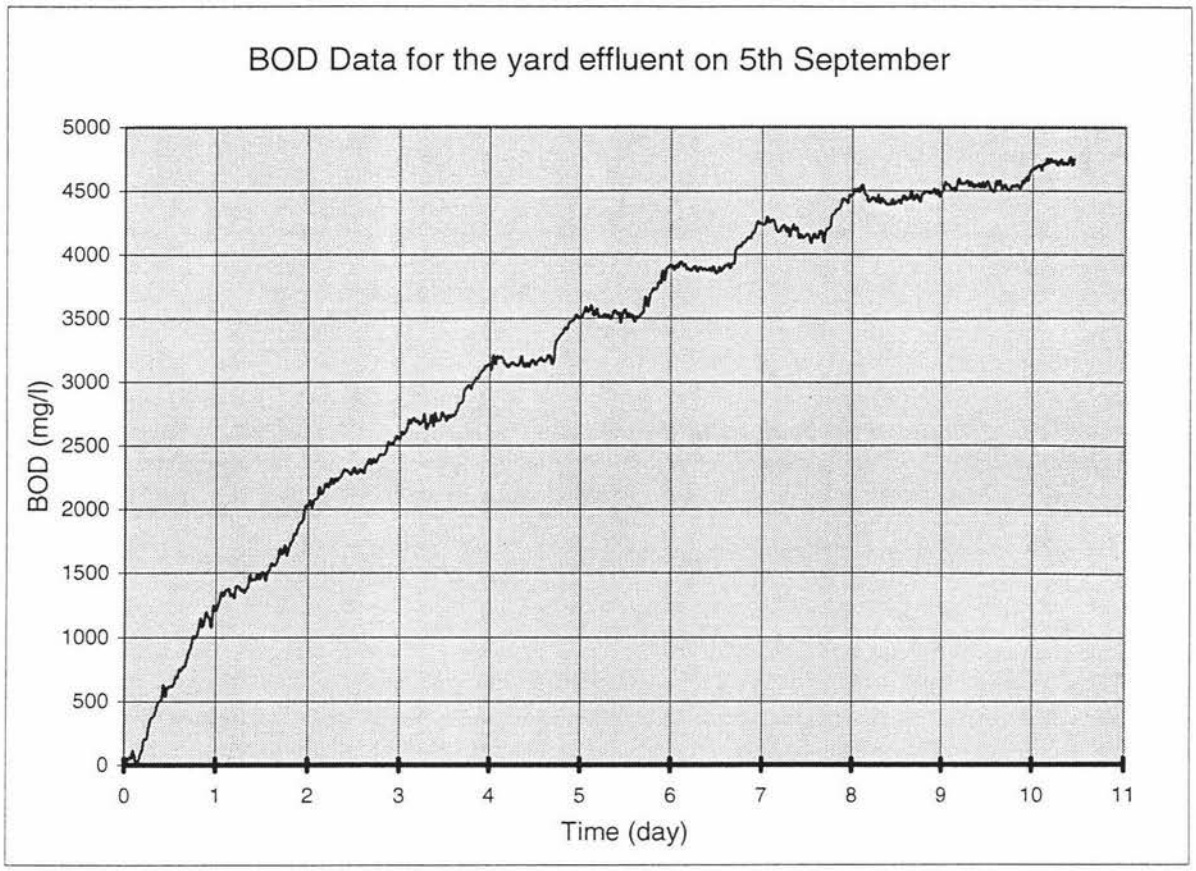
Appendix 1

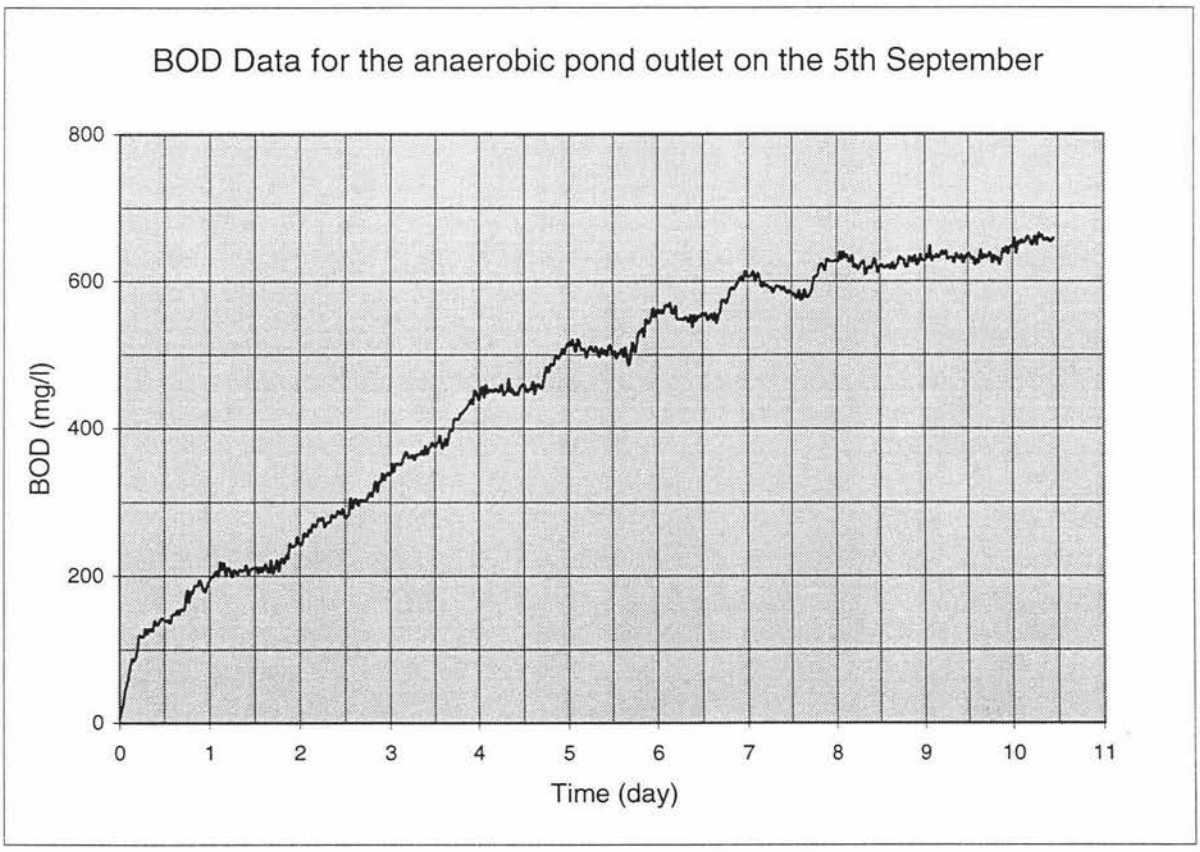

COD calibration curve
y = 0.0003x + 0.0033
R² = 0.9993
Absorbance
0.35
0.3
0.25
0.2
0.15
0.1
0.05
0
0
100
200
300
400
500
600
700
800
900
1000
COD concentration (mg/l)

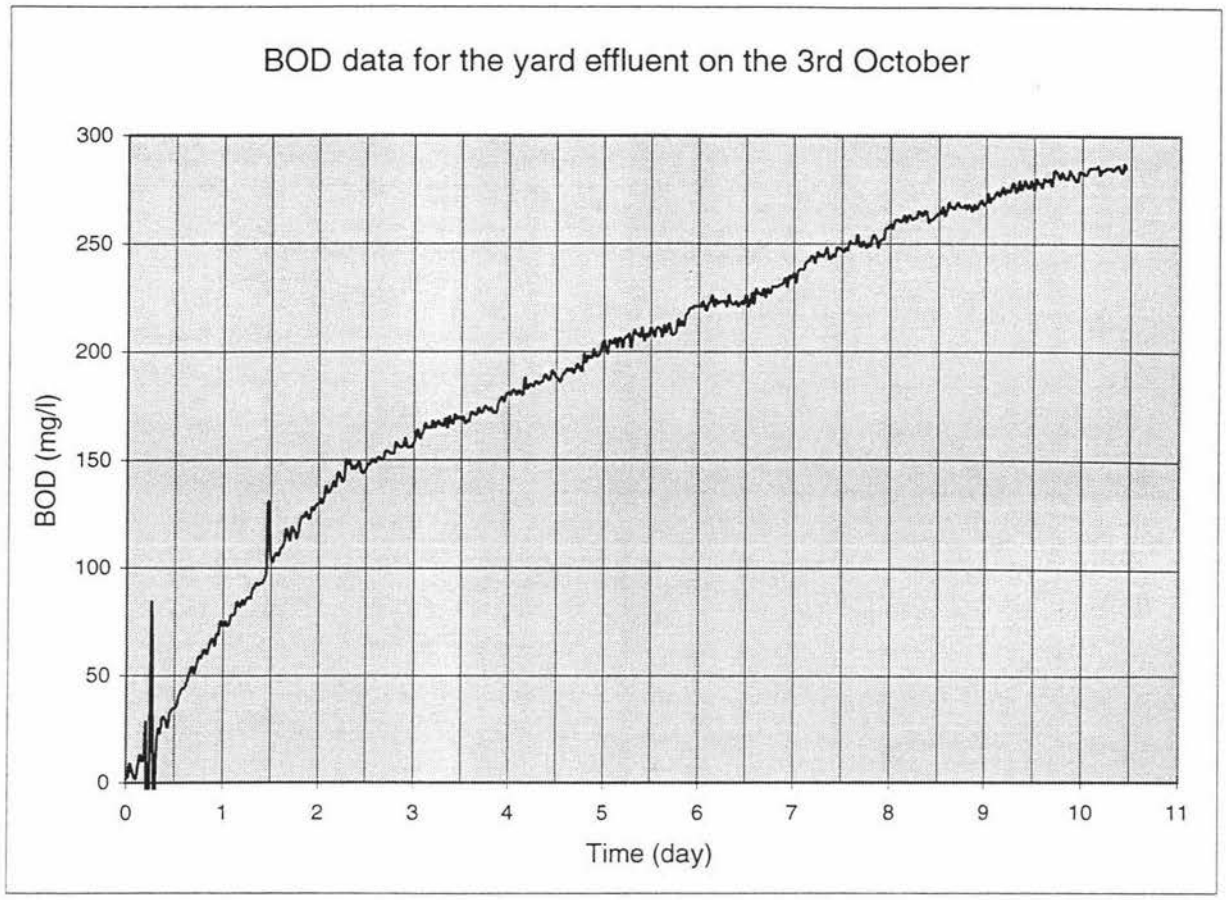
Appendix 2

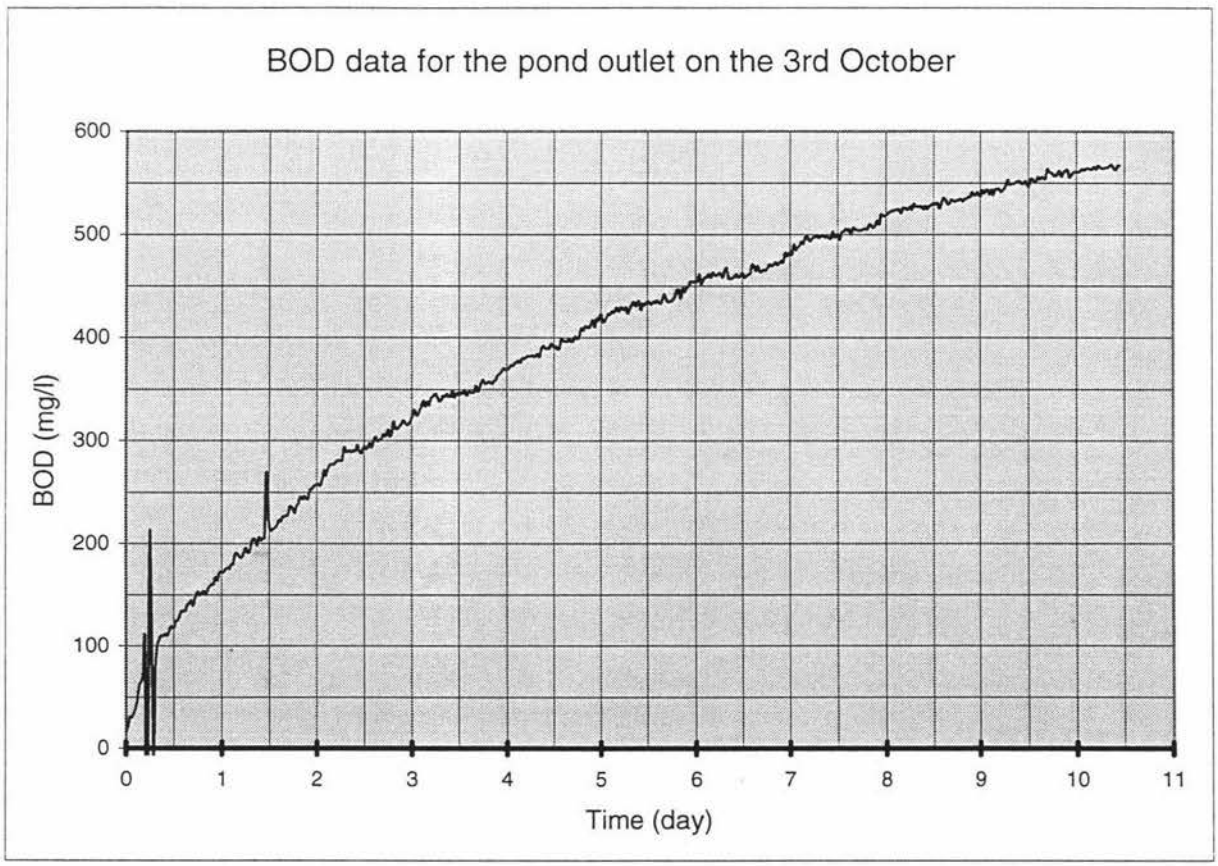

BOD data for the yard effluent on the 7 th August
BOD (mg/l)
6000
5000
4000
3000
2000
1000
0
0
1
2
3
4
5
6
7
8
9
10
11
Time (day)

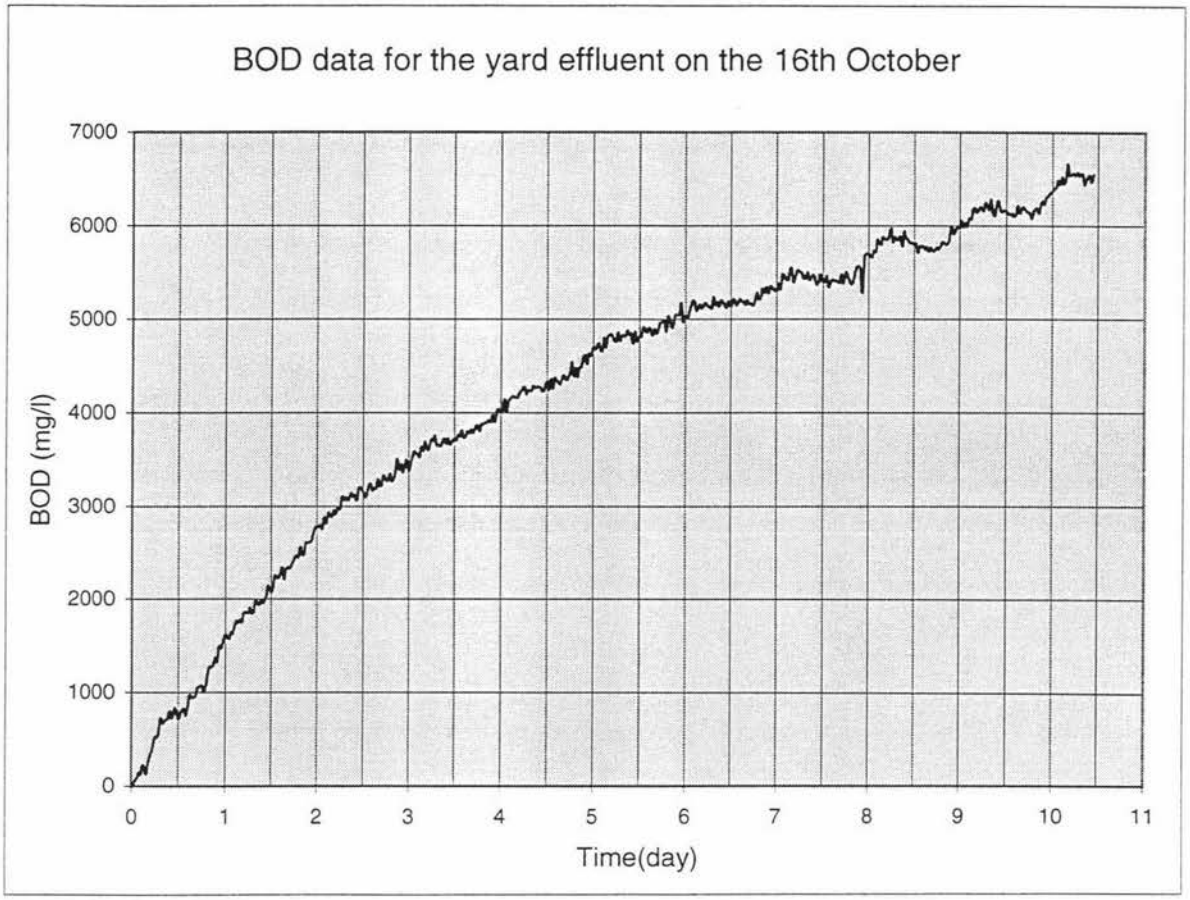

Bod data from the anaerobic pond outlet on 7th august
BOD (mg/l)
1000
500
0
0
1
2
3
4
5
6
7
8
9
10
11
Time (day)

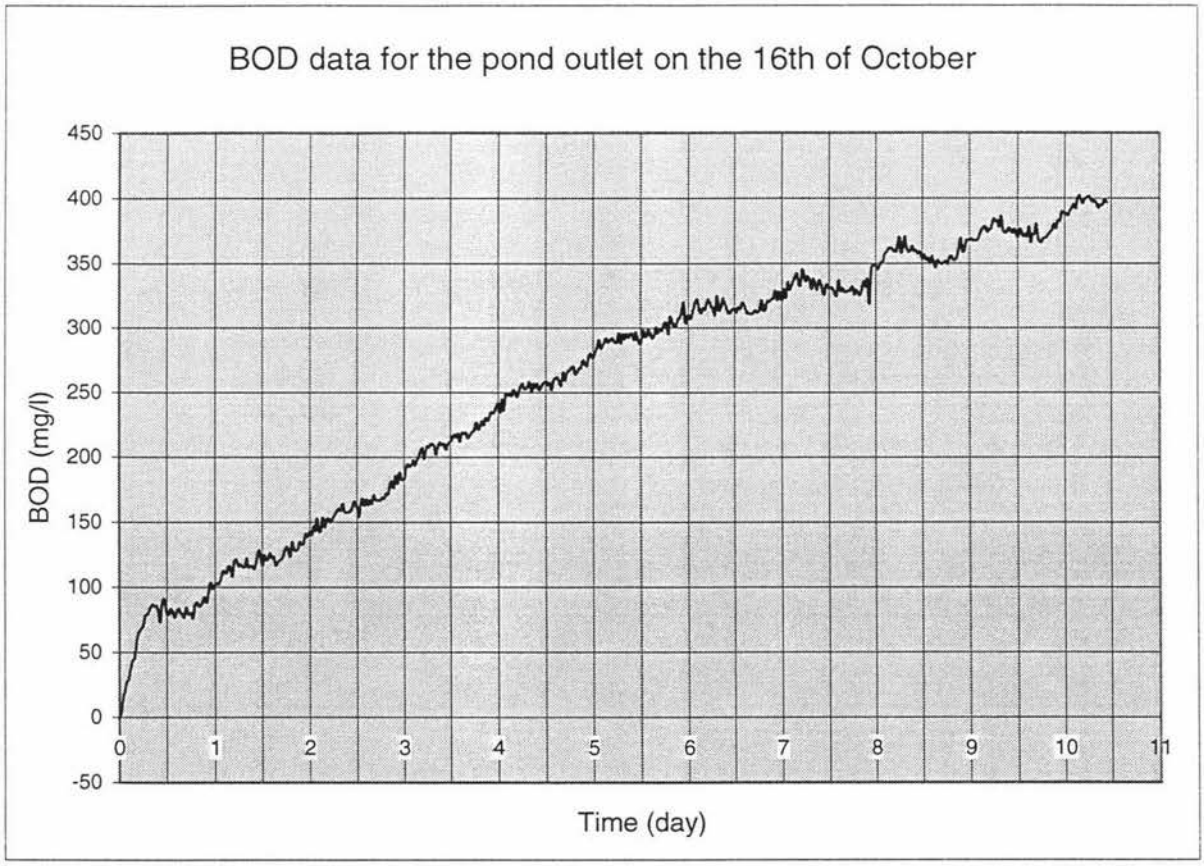

BOD Data for the yard effluent on 5th September
BOD (mg/l)
5000
4500
4000
3500
3000
2500
2000
1500
1000
500
0
0 1 2 3 4 5 6 7 8 9 10 11
Time (day)

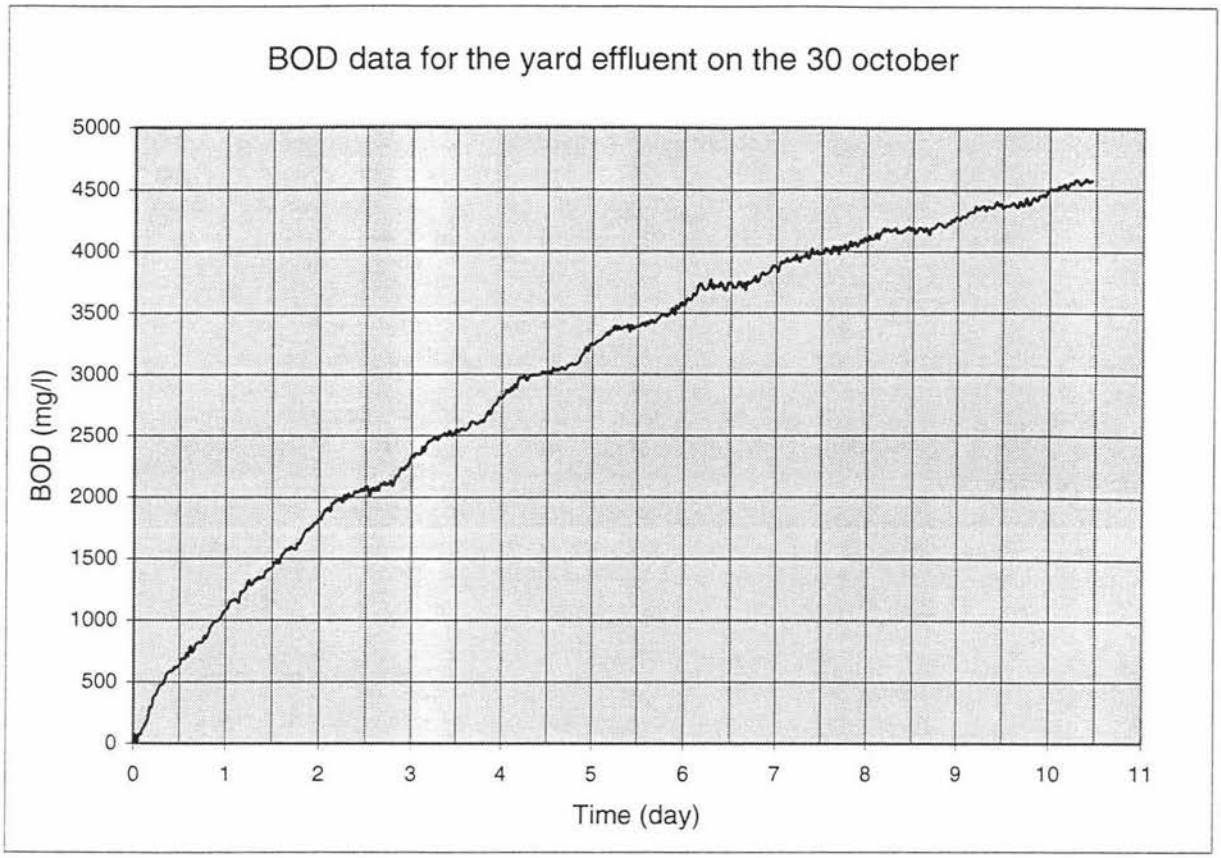

BOD Data for the anaerobic pond outlet on the 5th September
BOD (mg/l)
800
600
400
200
0
0 1 2 3 4 5 6 7 8 9 10 11
Time (day)

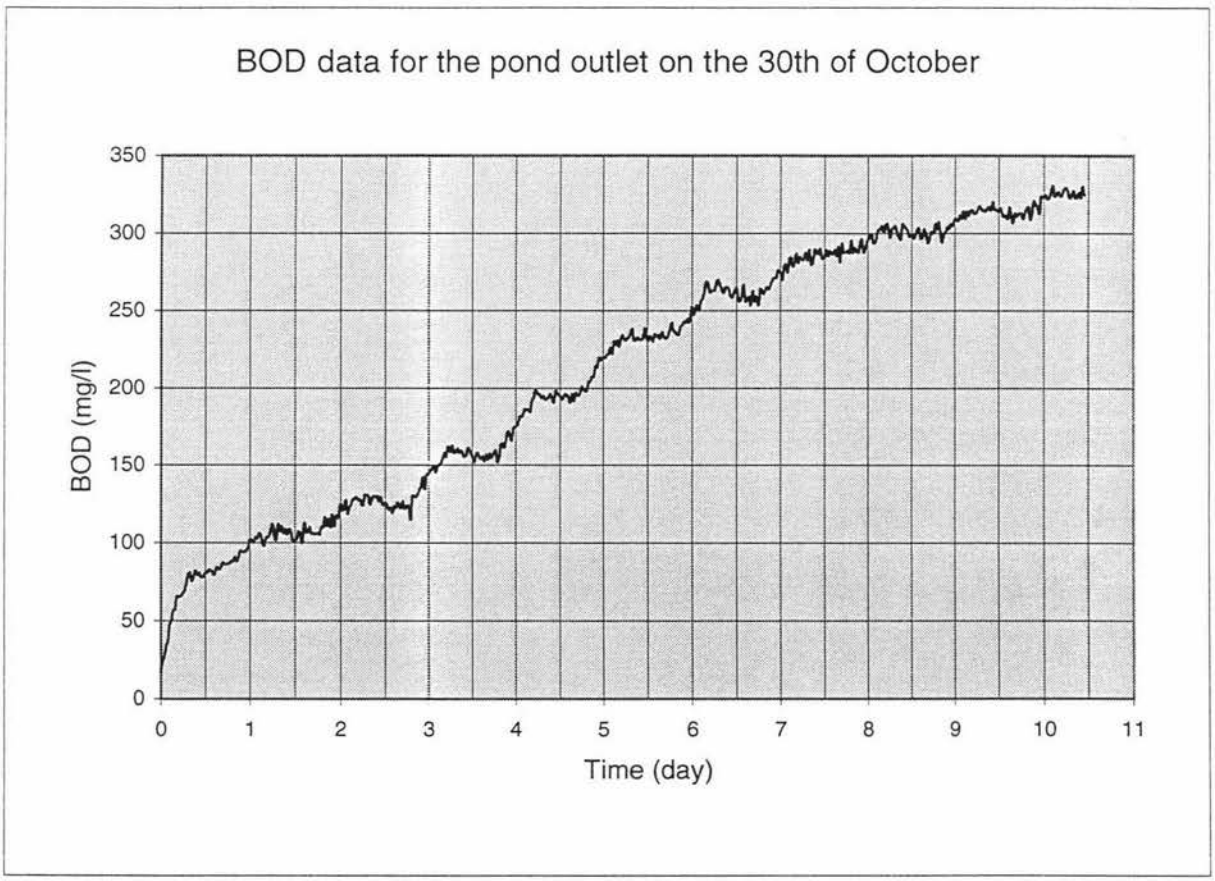

BOD data for the yard effluent on the 3rd October
BOD (mg/l)
300
250
200
150
100
50
0
0
1
2
3
4
5
6
7
8
9
10
11
Time (day)

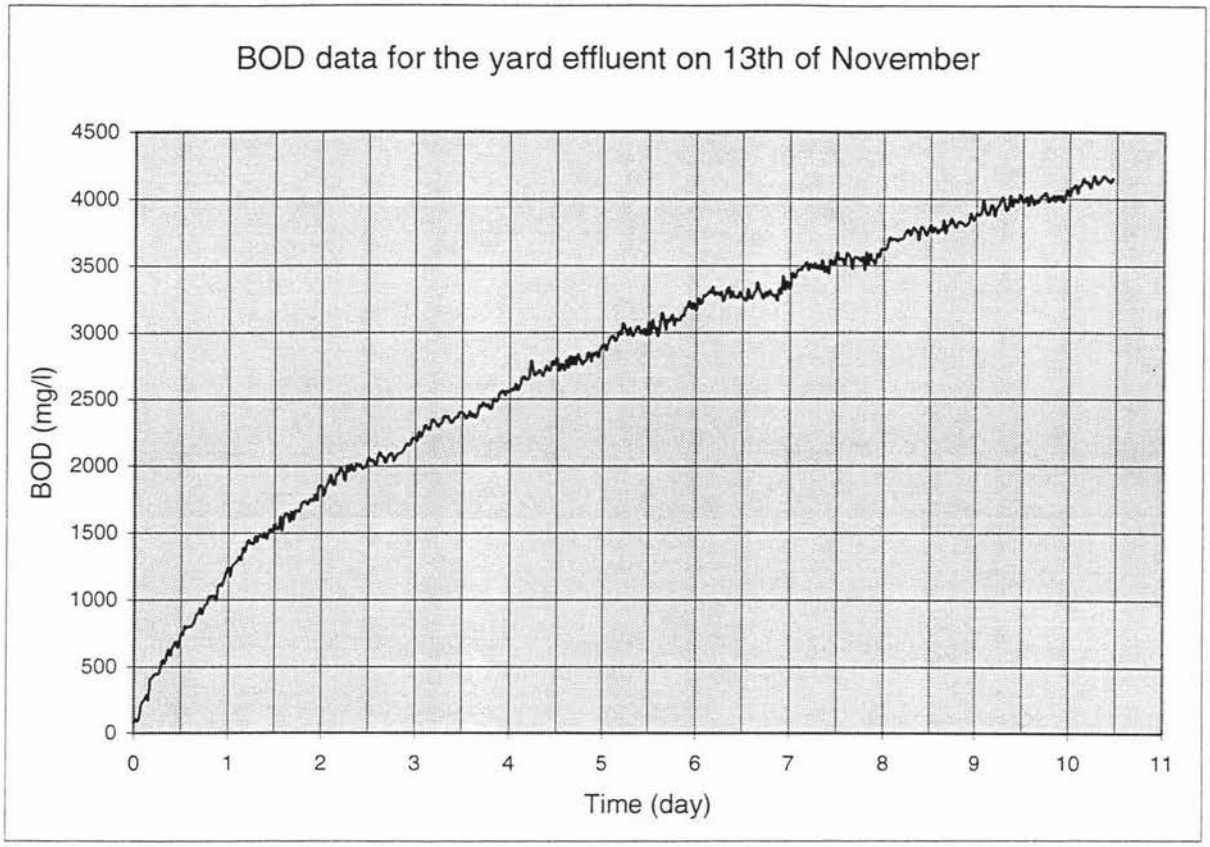

BOD data for the pond outlet on the 3rd October
BOD (mg/l)
600
500
400
300
200
100
0
0
1
2
3
4
5
6
7
8
9
10
11
Time (day)

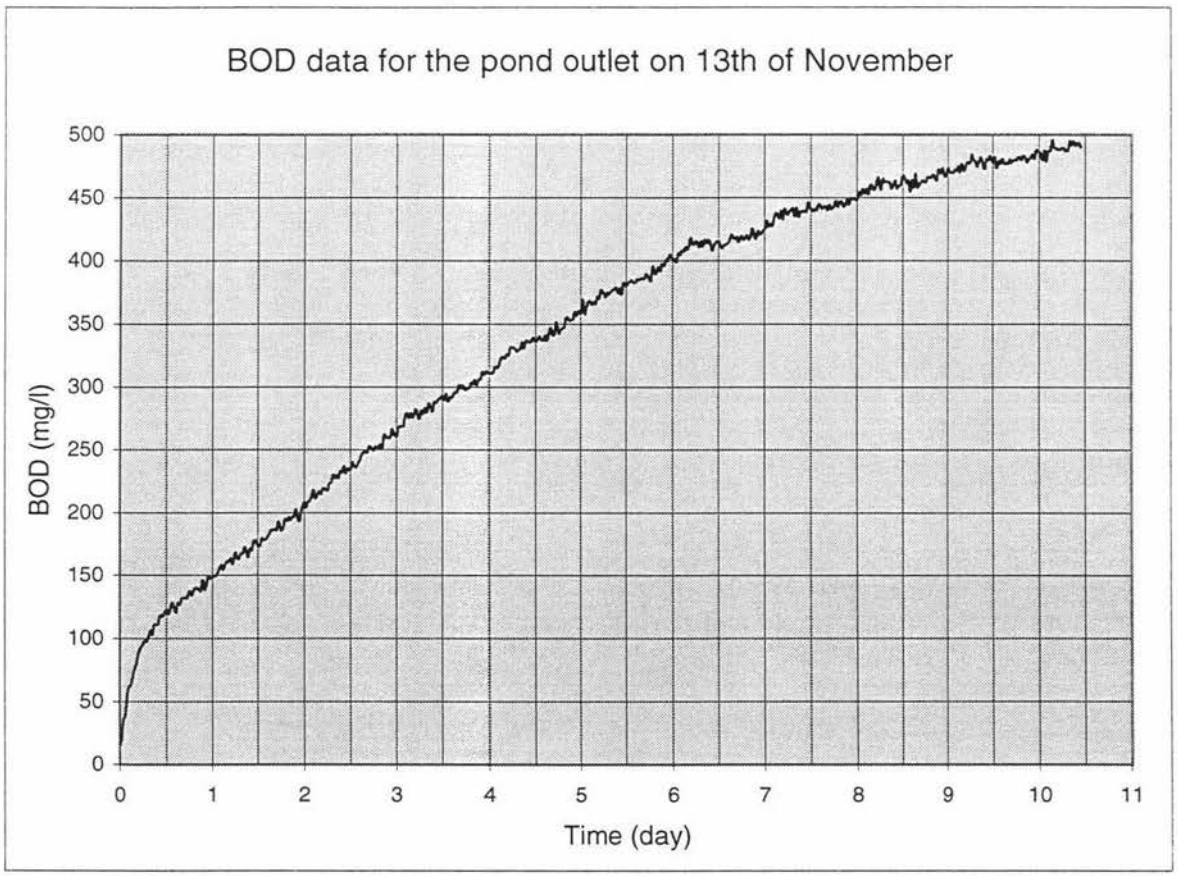

BOD data for the yard effluent on the 16th October
BOD (mg/l)
7000
6000
5000
4000
3000
2000
1000
0
0
1
2
3
4
5
6
7
8
9
10
11
Time(day)

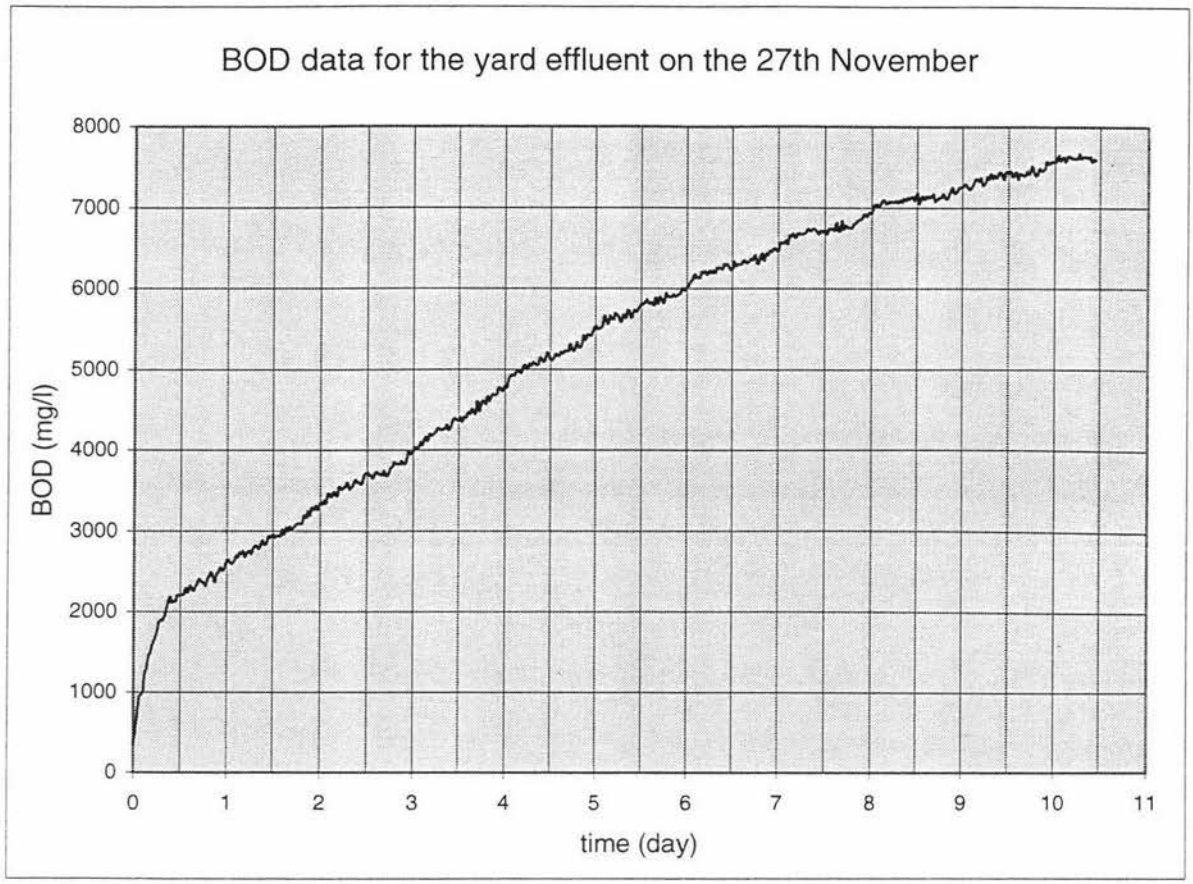

BOD data for the pond outlet on the 16th of October
BOD (mg/l)
450
400
350
300
250
200
150
100
50
0
-50
0
1
2
3
4
5
6
7
8
9
10
11
Time (day)

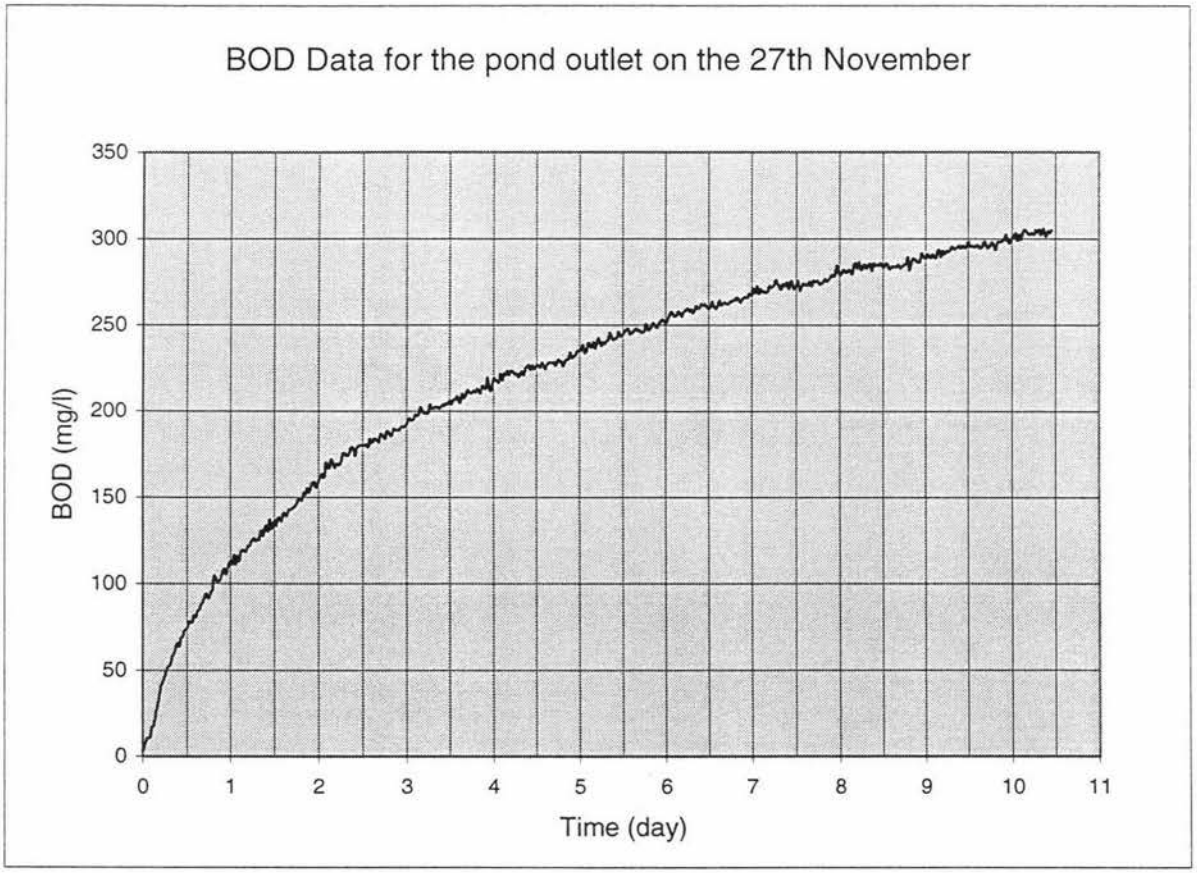

BOD data for the yard effluent on the 30 october
BOD (mg/l)
5000
4500
4000
3500
3000
2500
2000
1500
1000
500
0
0 1 2 3 4 5 6 7 8 9 10 11
Time (day)

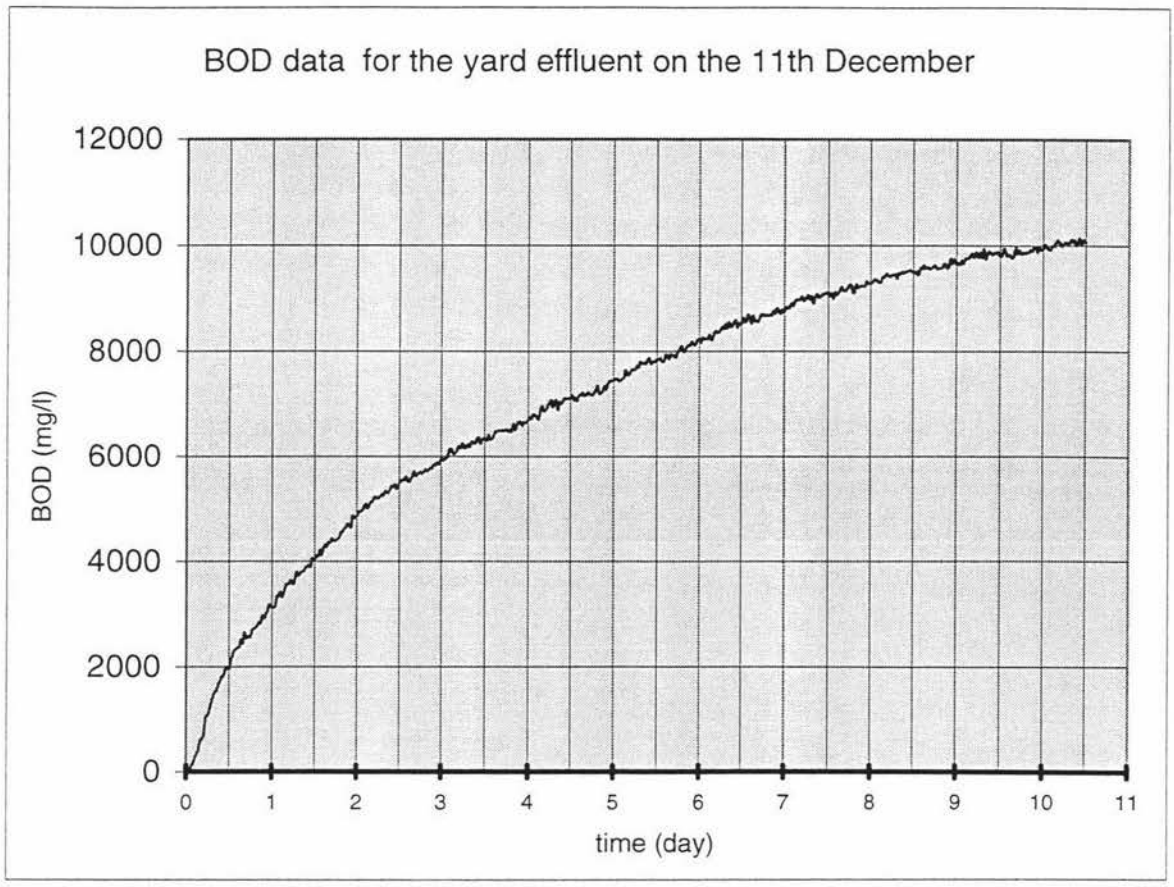

BOD data for the pond outlet on the 30th of October
BOD (mg/l)
350
300
250
200
150
100
50
0
0 1 2 3 4 5 6 7 8 9 10 11
Time (day)

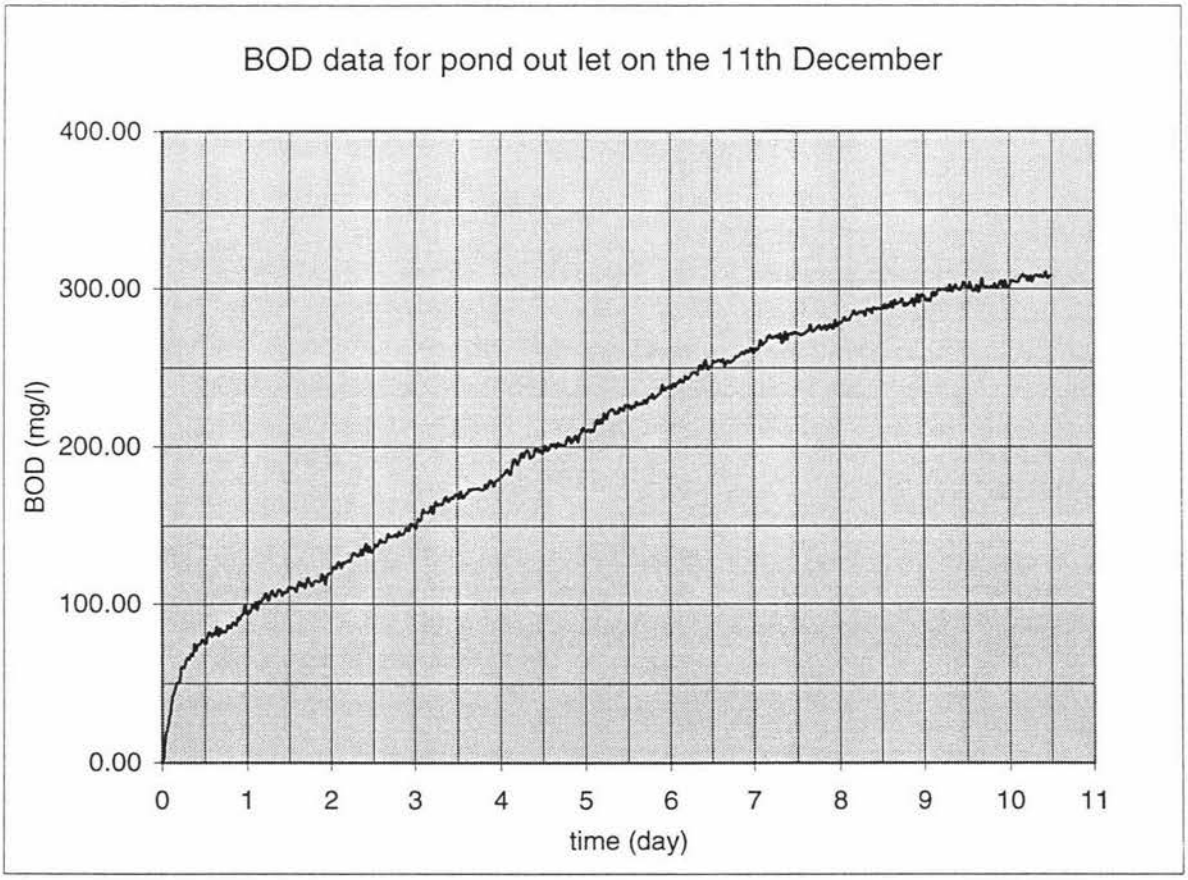

BOD data for the yard effluent on 13th of November
BOD (mg/l)
4500
4000
3500
3000
2500
2000
1500
1000
500
0
0 1 2 3 4 5 6 7 8 9 10 11
Time (day)


BOD data for the pond outlet on 13th of November
BOD (mg/l)
500
450
400
350
300
250
200
150
100
50
0
0 1 2 3 4 5 6 7 8 9 10 11
Time (day)


BOD data for the yard effluent on the 27th November
BOD (mg/l)
8000
7000
6000
5000
4000
3000
2000
1000
0
0 1 2 3 4 5 6 7 8 9 10 11
time (day)


BOD Data for the pond outlet on the 27th November
BOD (mg/l)
350
300
250
200
150
100
50
0
0 1 2 3 4 5 6 7 8 9 10 11
Time (day)


BOD data for the yard effluent on the 11th December
12000
10000
8000
6000
4000
2000
0
BOD (mg/l)
0
1
2
3
4
5
6
7
8
9
10
11
time (day)


BOD data for pond out let on the 11th December
400.00
300.00
200.00
100.00
0.00
BOD (mg/l)
0
1
2
3
4
5
6
7
8
9
10
11
time (day)

Appendix 3

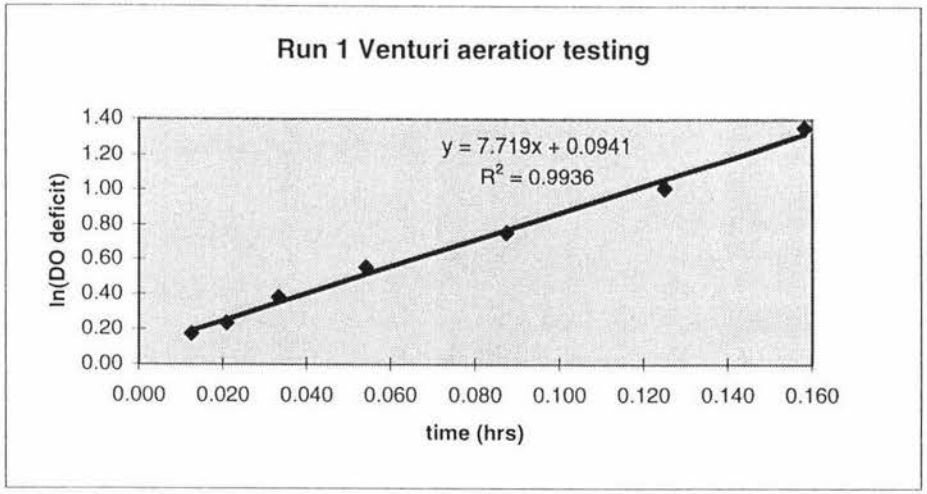
Dairy Shed holding yard data														
Day	Date	pH	BOD5	BODu	BOD K	VFA	%VFA in BOD5	Total COD	Soluble COD	Total Nitrogen	Total Phosphorus	NH_3	NO_3	Alkalinity
1	7-Aug	8.24	4500	5800	0.30	200	4.4	9150	4200	342.5	58	75	3	594
14	21-Aug	8.35				100		9063	2500	170	31.5	42	1.5	355.3
29	5-Sep	8.51	3500	5728	0.19	367	10.5	9340	4100	595	111.5	132	4.5	1325
42	18-Sep	8.66								320	60.75	84	1.5	840
57	3-Oct	8.47	2350	3343	0.24	255	10.8	6970	2480	325	72.5	87	3	810
70	16-Oct	8.41	4700	6840	0.23	568	12.1	15500	4415	655	126.5	168	4.5	2000
84	30-Oct	8.35	3250	5190	0.2	477	14.7	9765	5560	470	90	109.5	3	1075
98	13-Nov	8.51	2900	4855	0.18	256	8.8	9640	4517	570	85	198	1.5	1662
112	27-Nov	8.78	5500	13816	0.1	362	6.6	12890	5500	1625	155	600	6	2800
126	11-Dec	8.88	7200	12134	0.18	815	11.3	19150	9700	1750	217.5	600	9	2240
	average	8.52	4237.50	7213.25	0.20	377.88	9.91	11274.22	4774.67	682.25	100.83	209.55	3.75	1370.13
	Standard Deviation	0.19	1480.02	3478.26	0.05	204.56	3.02	3637.97	2022.72	522.64	51.71	199.94	2.25	748.29

Anaerobic Pond Data														
Day	Date	pH	BOD5	BODu	BOD K	VFA	%VFA in BOD5	Total COD	Soluble COD	Total Nitrogen	Total Phosphorus	NH_3	NO_3	Alkalinity
1	7-Aug	7.20	810	1559	0.15	50	6.2	1800	900	240	27	159	0.75	1108
14	21-Aug	7.12				145		1600	1250	220	29	141	0.75	1039
29	5-Sep	7.03	510	820	0.19	107	21.1	1600	1100	235	30.5	144	0.75	1359
42	18-Sep	7.11	400	857	0.12	98	24.6	1500	875	235	30.5	153	0.75	1137
57	3-Oct	6.95	410	655	0.2	140	34.1	1562	815	222.5	31.5	141	0.75	1229
70	16-Oct	7.00	280	487	0.17	20	7.1	1102	680	210	30.75	162	0.75	1190
84	30-Oct	7.05	220	912	0.06	97	44.2	970	598	240	33.5	177	0.75	1241
98	13-Nov	7.09	360	633	0.17	52	14.6	1350	905	235	35	172.5	0.75	1630
112	27-Nov	7.05	235	324	0.26	125	53.1	1625	675	260	35.5	180	0.75	1245
126	11-Dec	7.00	210	510	0.11	86	41.1	1065	554	292.5	44	190.5	0.75	1215
	average	7.06	381.67	750.78	0.16	92.12	27.35	1417.40	835.20	239.00	32.73	162.00	0.75	1239.30
	Standard Deviation	0.07	179.20	337.97	0.05	38.72	15.81	266.88	209.36	21.97	4.50	16.64	0.00	154.26

Appendix 4

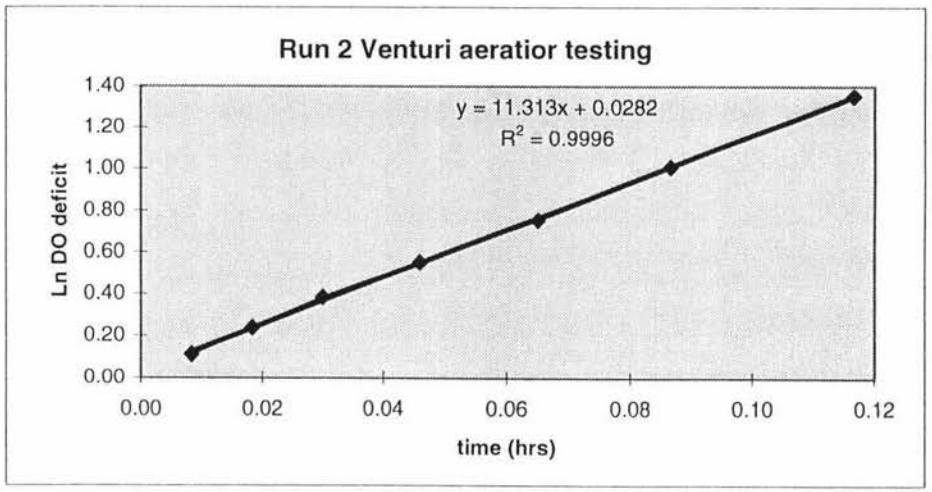
time (min)	time (hour)	DO	csm-do	-LN(csm-do)/DO sat
0.75	0.013	1.5	7.95	0.17
1.25	0.021	2	7.45	0.24
2	0.033	3	6.45	0.38
3.25	0.054	4	5.45	0.55
5.25	0.088	5	4.45	0.75
7.5	0.125	6	3.45	1.01
9.5	0.158	7	2.45	1.35

vol aerated	4450	l
DO sat	9.45	mg/l
Kla 20	8.03	mg/hr
power of pump	1.90	kw
kg of O2 transfered	0.33	kg/hr
Efficiency	0.17	kg/kw



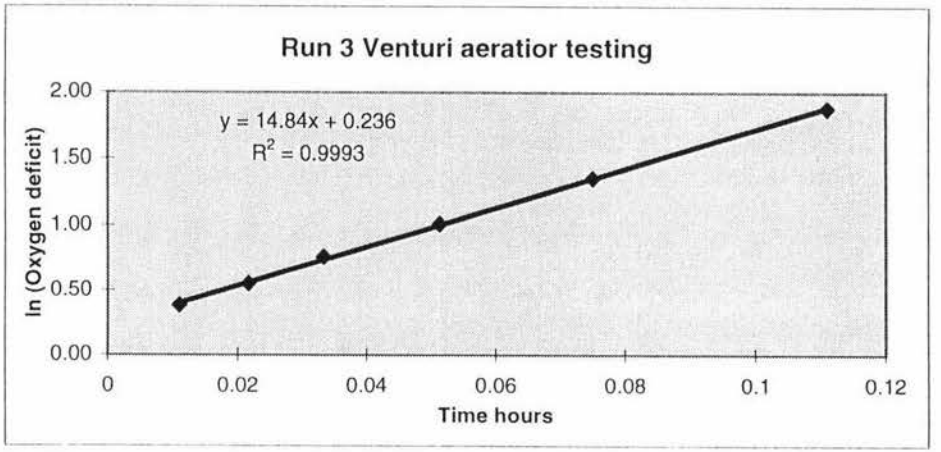
time (min)	time (hour)	DO	csm-do	-LN(csm-do)/DO sat
0.5	0.01	1.00	8.45	0.11
1.1	0.02	2.00	7.45	0.24
1.8	0.03	3.00	6.45	0.38
2.75	0.05	4.00	5.45	0.55
3.9	0.07	5.00	4.45	0.75
5.2	0.09	6.00	3.45	1.01
7	0.12	7.00	2.45	1.35

vol aerated	4450	l
DO sat	9.45	mg/l
Kla 20	11.77	mg/hr
power of pump	1.90	kw
kg of O2 transfered	0.48	kg/hr
Efficiency	0.25	kg/kw



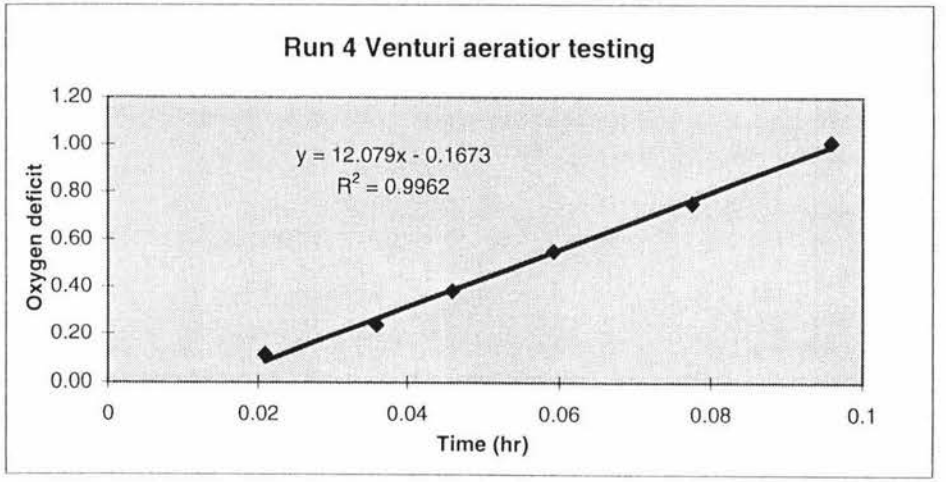
time (min)	time (hour)	DO	csm-do	-LN(csm-do)/DO sat
0.66	0.011	3	6.45	0.38
1.3	0.021667	4	5.45	0.55
2	0.033333	5	4.45	0.75
3.08	0.051333	6	3.45	1.01
4.5	0.075	7	2.45	1.35
6.66	0.111	8	1.45	1.87

vol aerated	4450	l
DO sat	9.45	mg/l
Kla 20	15.44	mg/hr
power of pump	1.90	kw
kg of O2 transfered	0.63	kg/hr
Efficiency	0.33	kg/kw



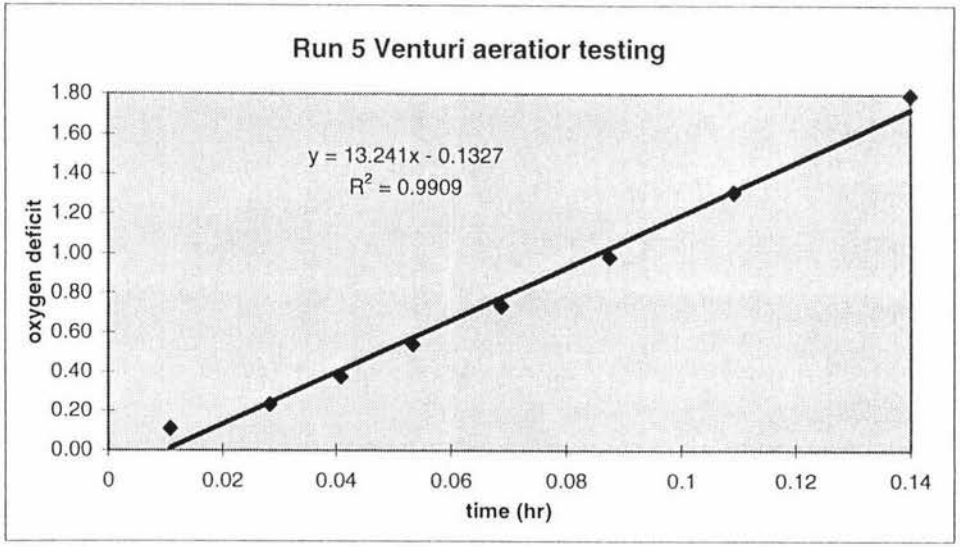
time (min)	time (hour)	DO	csm-do	-LN(csm-do)/DO sat
1.25	0.020833	1	8.45	0.11
2.15	0.035833	2	7.45	0.24
2.75	0.045833	3	6.45	0.38
3.55	0.059167	4	5.45	0.55
4.65	0.0775	5	4.45	0.75
5.75	0.095833	6	3.45	1.01

vol aerated	4450	l
DO sat	9.45	mg/l
Kla 20	12.57	mg/hr
power of pump	1.90	kw
kg of O2 transfered	0.51	kg/hr
Efficiency	0.27	kg/kw



time (min)	time (hour)	DO	csm-do	-LN(csm-do)/DO sat
0.65	0.010833	1	8.60	0.11
1.7	0.028333	2	7.60	0.23
2.45	0.040833	3	6.60	0.37
3.2	0.053333	4	5.60	0.54
4.125	0.06875	5	4.60	0.74
5.25	0.0875	6	3.60	0.98
6.55	0.109167	7	2.60	1.31
8.4	0.14	8	1.60	1.79

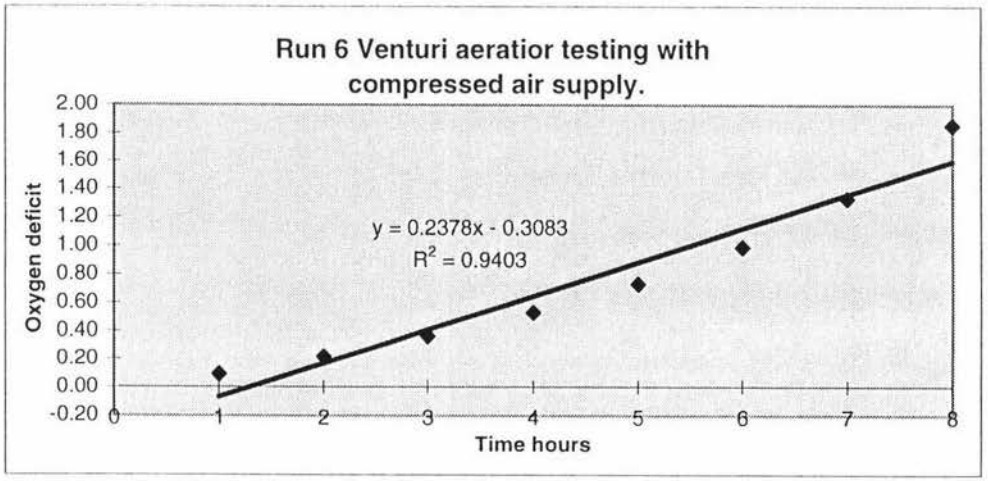
vol aerated	4450	l
DO sat	9.60	mg/l
Kla 20	14.33	mg/hr
power of pump	1.90	kw
kg of O2 transfered	0.59	kg/hr
Efficiency	0.31	kg/kw



run2	0.253614	average eff	0.29	kg/kw
run3	0.332681	standard d	0.04	kg/kw
run4	0.270786	percentage	0.13	
run5	0.322728			

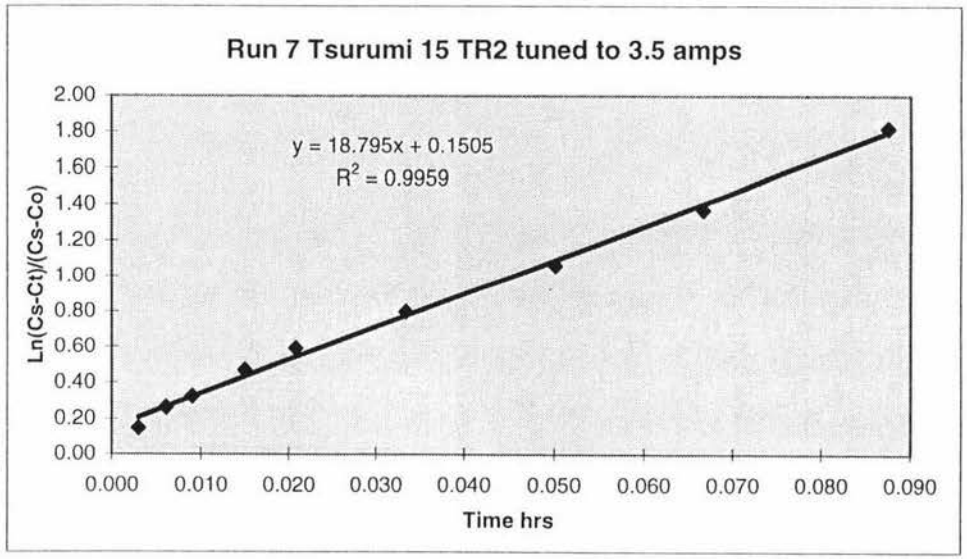
time (min)	time (hour)	DO	csm-do	-LN(csm-do)/DO sat
3.03	0.0505	1	8.45	0.09
3.58	0.059667	2	7.45	0.22
4.333	0.072217	3	6.45	0.36
5.033	0.083883	4	5.45	0.53
6	0.1	5	4.45	0.73
7.0666	0.117777	6	3.45	0.99
8.666	0.144433	7	2.45	1.33
10.783	0.179717	8	1.45	1.85

vol aerated	4450.00	l
DO sat	9.45	mg/l
Kla 20	18.76	mg/hr
power of pump	1.90	kw
kg of O2 transfered	0.79	kg/hr
Efficiency	0.42	kg/kw



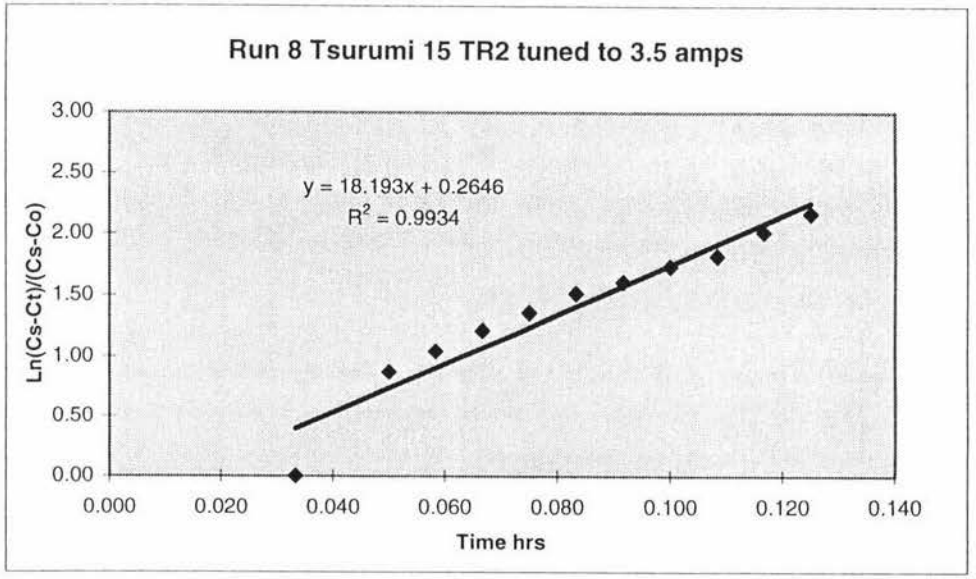
time (min)	time (hour)	DO	csm-do	-LN(csm-do)/DO sat
0.179	0.003	1.8	7.52	0.15
0.366	0.006	2.6	6.72	0.26
0.5416	0.009	3.0	6.32	0.32
0.9	0.015	3.9	5.47	0.47
1.25	0.021	4.5	4.82	0.59
2	0.033	5.4	3.92	0.80
3	0.050	6.3	3.02	1.06
4	0.067	7.1	2.22	1.37
5.25	0.088	7.9	1.42	1.81

vol aerated	4450	l
DO sat	9.32	mg/l
Kla 20	19.22	mg/hr
power of pump	1.40	kw
kg of O2 transfered	0.80	kg/hr
Efficiency	0.57	kg/kw



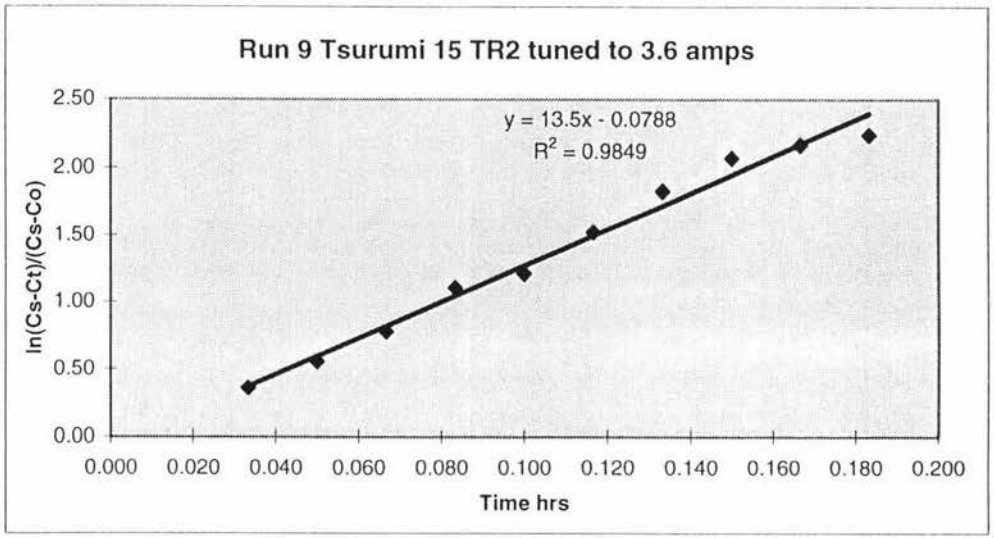
time (min)	time (hour)	DO	csm-do	-LN(csm-do)/DO sat
2	0.033	5.645278	3.654722	0.869599
2.5	0.042	6.22641	3.07359	1.042773
3	0.050	6.724523	2.575477	1.219584
3.5	0.058	7.056598	2.243402	1.357626
4	0.067	7.388673	1.911327	1.517822
4.5	0.075	7.554711	1.745289	1.608699
5	0.083	7.762258	1.537742	1.735304
5.5	0.092	7.886786	1.413214	1.819753
6	0.100	8.135842	1.164158	2.013622
6.5	0.108	8.30188	0.99812	2.167501
7	0.117	8.509427	0.790573	2.400617
7.5	0.125	8.716974	0.583026	2.705143

vol aerated	4450	l	l
DO sat	9.3	mg/l	mg/l
Kla 20	19.69269	mg/hr	mg/hr
power of pump	1.4	kw	kw
kg of O2 transfered	0.814982	kg/hr	kg/hr
Efficiency	0.58213	kg/kw	kg/kw



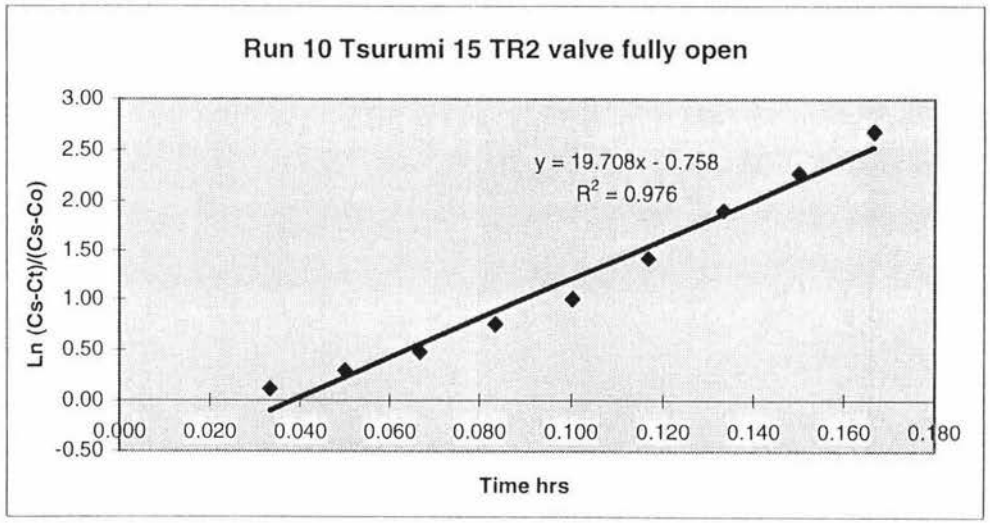
time (min)	time (hour)	DO	csm-do	-LN(csm-do)/DO sat
2	0.033	3.5	6.10	0.36
3	0.050	4.6	5.00	0.56
4	0.067	5.6	4.00	0.78
5	0.083	6.7	2.88	1.11
6	0.100	7.0	2.60	1.21
7	0.117	7.7	1.90	1.52
8	0.133	8.2	1.40	1.83
9	0.150	8.5	1.10	2.07
10	0.167	8.6	1.00	2.17
11	0.183	8.7	0.93	2.24

vol aerated	4450	l
DO sat	9.60	mg/l
Kla 20	14.61	mg/hr
power of pump	1.44	kw
kg of O2 transfered	0.62	kg/hr
Efficiency	0.43	kg/kw

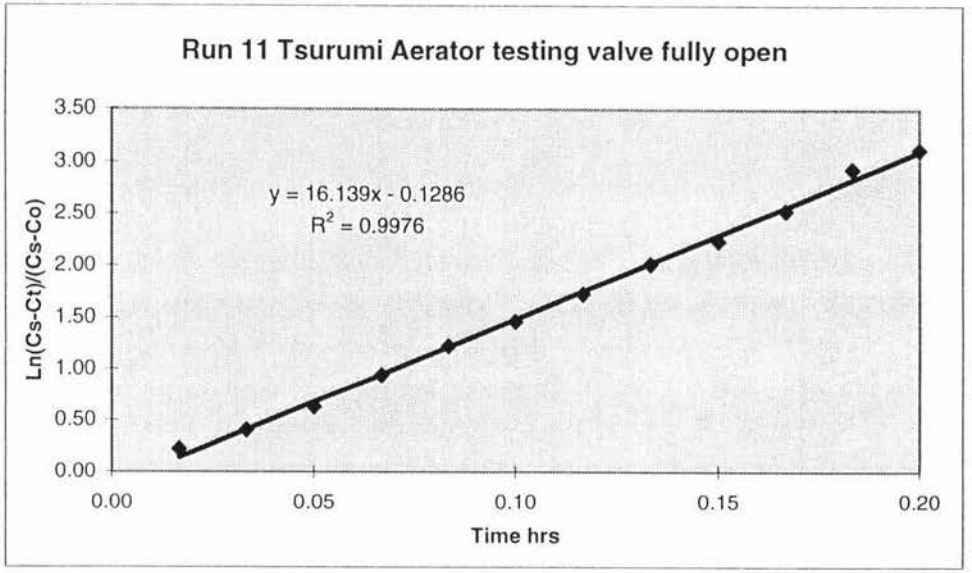


time (min)	time (hour)	DO	csm-do	-LN(csm-do)/DO sat
2	0.033	1.8	7.80	0.11
3	0.050	3.1	6.50	0.29
4	0.067	4.2	5.40	0.48
5	0.083	5.5	4.10	0.75
6	0.100	6.4	3.20	1.00
7	0.117	7.5	2.10	1.42
8	0.133	8.3	1.30	1.90
9	0.150	8.7	0.90	2.27
10	0.167	9.0	0.60	2.68

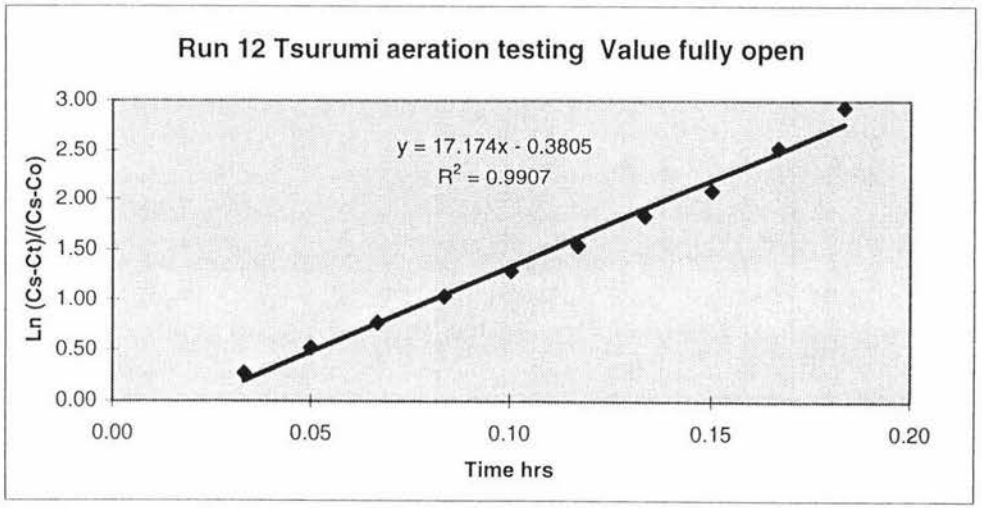
vol aerated	4450	l
DO sat	9.60	mg/l
Kla 20	21.33	mg/hr
power of pump	0.80	kw
kg of O2 transfered	0.91	kg/hr
Efficiency	1.14	kg/kw



time (min)	time (hour)	mm on cha	DO	csm-do	-LN(csm-do)/DO sat
1	0.02	30	2.4	7.20	0.22
2	0.03	45	3.6	6.00	0.41
3	0.05	60	4.8	4.80	0.63
4	0.07	76	6.08	3.52	0.94
5	0.08	87	6.96	2.64	1.23
6	0.10	94	7.52	2.08	1.46
7	0.12	100	8	1.60	1.73
8	0.13	105	8.4	1.20	2.01
9	0.15	108	8.64	0.96	2.24
10	0.17	111	8.88	0.72	2.53
11	0.18	114	9.12	0.48	2.93
12	0.20	115	9.2	0.40	3.11
	vol aerated			4450	l
	DO sat			9.60	mg/l
	Kla 20			17.82	mg/hr
	power of pump			0.80	kw
	kg of O2 transfered			0.76	kg/hr
	Efficiency			0.95	kg/kw



time min	hrs	mm on cha	do	csm-do	-LN(csm-do)/DO sat
2	0.03	34	2.72	6.88	0.27
3	0.05	53	4.24	5.36	0.52
4	0.07	68	5.44	4.16	0.77
5	0.08	80	6.4	3.20	1.03
6	0.10	89	7.12	2.48	1.29
7	0.12	96	7.68	1.92	1.54
8	0.13	102	8.16	1.44	1.83
9	0.15	106	8.48	1.12	2.08
10	0.17	111	8.88	0.72	2.53
11	0.18	114	9.12	0.48	2.93
	DO sat			9.6	
	vol			4450	
	Kla 20			18.59	
	O2 transfered			0.76	kg/h
	power			0.80	kwh
	efficiency			0.95	kg/kw



Appendix 5

Cycle time calculations using first order rate kinetics

BODi	0 mg/l	KBOD	0.0128	tank vol l	4450		
TKNi	185 mg/l	KNH4	0.59	decant pur	7250		
PO4i	25.3 mg/l	KNO3	0.00285	fill pump l/l	12750		
BODf	30 mg/l	%NO3	0.21	Aeration ef	0.85	kg/kw	
TKNf	10 mg/l	P release r	4.13				
NO3f	10 mg/l	BOD:Prele	11.3				
PO4f	2 mg/l						

bod	Excess BOD available	BODreact	TKN react	NO_3 react	PO_4 react	%tank emptied	Biomass	BOD time	TKN time	NO3 time	P release time	fill time	settle time
100	-603.093	-223.237	80	76	47	0.4	5000	0.0	3.4	3.4	1.5	0.1	1.0
200	-503.093	-183.237	80	76	47	0.4	5000	0.0	3.4	3.4	1.5	0.1	1.0
300	-403.093	-143.237	80	76	47	0.4	5000	0.0	3.4	3.4	1.5	0.1	1.0
400	-303.093	-103.237	80	76	47	0.4	5000	0.0	3.4	3.4	1.5	0.1	1.0
500	-203.093	-63.2372	80	76	47	0.4	5000	0.0	3.4	3.4	1.5	0.1	1.0
600	-103.093	-23.2372	80	76	47	0.4	5000	0.0	3.4	3.4	1.5	0.1	1.0
700	-3.09308	16.76277	80	76	47	0.4	5000	0.0	3.4	3.4	1.5	0.1	1.0
800	96.90692	56.76277	80	76	47	0.4	5000	0.2	3.4	3.4	1.5	0.1	1.0
900	196.9069	96.76277	80	76	47	0.4	5000	0.4	3.4	3.4	1.5	0.1	1.0
1000	296.9069	136.7628	80	76	47	0.4	5000	0.6	3.4	3.4	1.5	0.1	1.0
1100	396.9069	176.7628	80	76	47	0.4	5000	0.7	3.4	3.4	1.5	0.1	1.0
1200	496.9069	216.7628	80	76	47	0.4	5000	0.7	3.4	3.4	1.5	0.1	1.0
1300	596.9069	256.7628	80	76	47	0.4	5000	0.8	3.4	3.4	1.5	0.1	1.0
1400	696.9069	296.7628	80	76	47	0.4	5000	0.9	3.4	3.4	1.5	0.1	1.0
1500	796.9069	336.7628	80	76	47	0.4	5000	0.9	3.4	3.4	1.5	0.1	1.0
1600	896.9069	376.7628	80	76	47	0.4	5000	0.9	3.4	3.4	1.5	0.1	1.0
1700	996.9069	416.7628	80	76	47	0.4	5000	1.0	3.4	3.4	1.5	0.1	1.0

decant time	cycle time	cycles/d	vol/day	Biomas produced kg/d	O2kg/d	aerobic/hr	kg O2/hr	Aerator size
0.2	9.7	2.5	4423	0.07	1.4	8.3	0.2	0.2
0.2	9.7	2.5	4423	0.07	1.4	8.3	0.2	0.2
0.2	9.7	2.5	4423	0.07	1.4	8.3	0.2	0.2
0.2	9.7	2.5	4423	0.07	1.4	8.3	0.2	0.2
0.2	9.7	2.5	4423	0.07	1.4	8.3	0.2	0.2
0.2	9.7	2.5	4423	0.07	1.4	8.3	0.2	0.2
0.2	9.7	2.5	4423	0.07	1.4	8.3	0.2	0.2
0.2	9.7	2.5	4423	0.07	1.6	8.3	0.2	0.2
0.2	9.7	2.5	4423	0.07	1.8	8.3	0.2	0.3
0.2	9.7	2.5	4423	0.07	2.1	8.3	0.3	0.3
0.2	9.7	2.5	4423	0.07	2.4	8.3	0.3	0.3
0.2	9.7	2.5	4423	0.07	2.6	8.3	0.3	0.4
0.2	9.7	2.5	4423	0.07	2.9	8.3	0.3	0.4
0.2	9.7	2.5	4423	0.07	3.2	8.3	0.4	0.4
0.2	9.7	2.5	4423	0.07	3.4	8.3	0.4	0.5
0.2	9.7	2.5	4423	0.07	3.7	8.3	0.4	0.5
0.2	9.7	2.5	4423	0.07	3.9	8.3	0.5	0.6