ATKINS

Orange County Sanitation District

Anaerobic Baffled Reactor (ABR) Pilot Plant Demonstration

Feb 2002 – May 2004

Final Report (version 1.3)

Presented by Rupert Kruger
March 2005

EXECUTIVE SUMMARY

What is an ABR?

An Anaerobic Baffled Reactor (ABR) is a wastewater treatment technology that uses baffles to create multiple treatment zones in a primary clarifier. A sludge blanket is established in each baffled zone, and different microbiological populations establish themselves in each zone. The overall effect is to provide both primary treatment and some secondary treatment in a single basin. This compartmentalized design separates the solids retention time (SRT) from the hydraulic retention time (HRT), so wastewater can be treated anaerobically at short retention times (comparable to ordinary primary treatment retention times). ABRs can be viewed both as an alternative to primary clarification and as partial anaerobic secondary treatment providing *in situ* solids destruction. The particular ABR design being tested by OCSD is intended specifically to be a retrofit to existing primary basins.

OCSD and ABRs

ABRs traditionally have been used to treat high strength industrial wastewater. In 2001, Atkins completed a successful 2-year research project to develop and test ABRs for treating domestic wastewater. This work was funded by a group of UK water companies and OCSD. A full-scale ABR retrofit was then constructed and operated at a treatment plant in Northern Ireland.

During the original project the ABR benefits were demonstrated under conditions typical for UK treatment plants. These differed from typical Southern California conditions where the average temperature is warmer, primary clarifiers operate with shorter HRTs, the wastewater has lower TSS, and enhanced treatment with ferric salts and polymer may be used. To better evaluate ABRs for OCSD, a 5-compartment pilot plant was constructed treating up to 0.3 MGD. In a full-scale installation, the compartments would be formed by baffles inside a single primary basin.

Project Overview

The ABR was running for 15 months. The operation can be split into four distinct phases:

Phase 1: ***Feb – June 2003***
System startup, stabilization of the sludge blankets, and optimization.

Phase 2: ***July – Oct 2003***
Operation with reduced sludge blanket depths (a change from the previous UK operations).

Phase 3: ***Oct 2003 – Jan 2004***
Change from automated to manual desludging to allow more accurate mass balance calculations.

Phase 4: ***Jan – May 2004***
Operation at conditions comparable to conventional OCSD primary clarifiers with the addition of ferric and polymer to the primary influent.

Project Objectives

- To assess ABR performance under Southern Californian conditions
- To review the impact of polymer addition on ABR performance
- To assess ABR performance under different hydraulic retention times
- To compare ABR effluent quality to primary tank effluent quality
- To determine the solids destruction achieved by the ABR
- To quantify the risks associated with release of methane
- To estimate the costs and benefits of full-scale ABR installation

Summary of Results

HRT: 3.0 - 4.4 hours depending on the trial phase which was similar to the shorter primary tank retention times observed at OCSD

TSS: ABR removal rates average about 52% without polymer addition in Phase 3 and 60 - 65% with polymer addition in Phase 4. This was compared to A-side clarifier removal rates of approximately 70% during Phase 4.

BOD: ABR removal rates varied from 20% to 35% depending on the trial phase. This was compared to A-side clarifier removal rates of approximately 45 - 50% over similar periods. The ABR BOD removal rate might have been reduced due to partial digestion of some volatile solids, solubilizing them but not providing sufficient residence time to complete their conversion to methane.

VS (% of TS): ABR sludge VS content decreased through the ABR (62% in tank 1 decreasing to 53% in tank 5) confirming that solids destruction was occurring.

VS reduction: The volatile solids destruction ranged from 22 – 39% during Phase 3 and 4. Volatile solids destruction is by solubilization to chemical intermediates and subsequent conversion to methane.

Digestibility Laboratory assessment of conventional primary sludge and ABR sludge showed there was no difference in their digestibility

Dewaterability Laboratory assessment of conventional primary sludge and ABR sludge showed there was no difference in their dewaterability

Methane Risk Evaluation

A preliminary evaluation was conducted to identify and assess risks associated with the production of methane during solids digestion in an ABR. In normal day-to-day operation with the foul air collection systems operating as designed the methane concentration will not reach the minimum explosive threshold of 5% by volume. If the foul air collection system failed, the greatest potential for the methane concentration to reach an explosive concentration would be in the enclosed headspace above the primary basin overflow weirs on Plant 2.

ABR Costs and Benefits

The summary of costs presented in Table 1 was developed using the OCSD biosolids master plan (BMP) model. It should be noted that the costs do not take into account the treatment costs for primary tank effluent and only consider the solids handling costs. In every scenario the installation of ABR was found to generate net present savings which ranged from $35 million to $70 million.

Table 1 Summary of total present cost (saving) for ABR installation

	Total Present Cost (saving)	
	Plant 1 with ABR	**Plant 2 with ABR**
Baseline	($70 million)	($35 million)
Scenario 1	($35 million)	($36 million)
Scenario 2	($39 million)	($35 million)

The summary of net present costs (savings) in Table 2 takes into account the costs associated with treatment of the liquid effluent stream, but this analysis used a much less detailed model than the BMP model, so comparisons with Table 1 should be made very cautiously. It indicates that retrofit of ABR to Plant 1 has the potential to generate between $2.5 and $9 million savings based on a 20 year period discounted at 5%. If the capital cost of a retrofit is at the top end of estimated costs there will be no net savings for Plant 1 or Plant 2. Detailed design of the retrofit will enable more robust costs to be estimated for ABR retrofit.

Table 2 Summary of net present cost (saving) for Plant 1 and Plant 2

Scenario	**Net Present Cost (saving)**	
	Plant 1	**Plant 2**
Low cost retrofit with low operational savings	($2.5 million)	$5.2 million
Low cost retrofit with high operational savings	($9 million)	($4.6 million)
High cost retrofit with low operational savings	$11 million	$13.5 million
High cost retrofit with high operational savings	$4.4 million	$3.7 million

Summary and recommendations

The ABR pilot-plant has demonstrated that the ABR can convert volatile solids to methane and the associated reduction in the solids load onto downstream processes will generate net present savings. The net present savings are most likely to be realized at Plant 1 due to the reduced complexity and capital costs associated with a rectangular primary tank ABR retrofit.

The biochemical pathway to convert volatile solids to methane involves the production of intermediate compounds such as volatile fatty acids. Failure to convert all the intermediate compounds to methane can result in elevated concentrations of COD and soluble BOD in the ABR effluent. This would reduce the apparent BOD reduction in the ABR.

The potential impact of elevated concentrations of COD and BOD on future proposed down-stream secondary treatment processes will need to be assessed. It may have an impact on the generation of secondary sludge and the oxygen demand exerted during the treatment process.

A full-scale ABR retrofit has the potential to enhance the conversion of organic intermediates to methane because greater quantities of sludge can be retained in the ABR compartments given the increased depth of the full scale primary tanks and associated increase in sludge blanket depth.

In order to inform the decision about proceeding to a full-scale trial it is recommended that:

- Sensitivity analysis is conducted around a range of primary tank effluent BOD concentrations and downstream treatment aeration capacity and associated costs
- A potential site for a full-scale trial at Plant 1 is identified and a more detailed design for an ABR retrofit is produced in order to improve the estimate of the cost of retrofit

If the sensitivity analysis and additional design and costing work indicate that a full-scale ABR installation will still generate net present savings, it is recommended that a detailed study plan is produced for a full-scale trial. The plan should identify what the full-scale test is designed to accomplish in terms of better understanding of the process and obtaining essential design and operation data needed for detailed process and cost analysis. The full-scale trial would be designed to confirm the operational robustness of a full-scale ABR plant and to assess performance of a full-scale ABR against a full-scale control primary tank.

TABLE OF CONTENTS

1. AIMS AND OBJECTIVES

An Anaerobic Baffled Reactor (ABR) is a novel compartmentalized reactor design which enables solids retention time (SRT) to be separated from the hydraulic retention time (HRT). This makes it possible to anaerobically treat wastewaters at short retention times. ABRs can therefore be viewed both as an attractive alternative to primary settlement and / or anaerobic treatment.

The expected benefits of the ABR process include:

- Reduction in sludge production
- Enhanced solids capture in the ABR tanks
- Reduced solids loading downstream
- Low capital and operational costs

ABR operation at Orange County Sanitation District (OCSD) started in February 2003 and was completed in June 2004. Initially it formed part of the ongoing Micro-Filtration (MF) Demonstration Project at OCSD's Plant 2. The MF plant was shut down in January 2004. From that time onwards ABR operation continued as a stand alone project.

Raw wastewater flowed from the head works into the ABR (5 tanks / compartments). Performance was assessed in terms of effluent quality and *in situ* sludge reduction. The ABR was also evaluated as a treatment option to enhance the quality of the MF feed and to treat the backwash flows from the MF plant.

The main objective of the project was ***to demonstrate the economic, practical and technical feasibility of an ABR design for treating wastewater and reducing primary sludge production at OCSD.***

More specifically, the key objectives of the work were to:

- Compare the effluent quality from an ABR to conventional primary effluent
- Examine ABR performance under conditions specific to Southern California i.e. warmer temperatures, shorter HRTs and thinner wastewaters than the UK
- Evaluate the ABR performance with and without ferric dosing and polymer addition
- Assess the ABR performance at different hydraulic retention times and surface overflow rates
- Evaluate the ABR solids destruction and the impacts on downstream anaerobic digestion and dewatering
- Obtain data on ABR cost, sizing, and scale-up

2. BACKGROUND INFORMATION

2.1 Introduction

The treatment of domestic and industrial wastewater is often undertaken via biological means, as opposed to physical-chemical methods, primarily because of lower costs. However, due to a number of misconceptions, aerobic rather than anaerobic treatment is often the chosen treatment method. These misconceptions include the perception that anaerobic treatment is poor for treating low strength wastes (COD < 1000 mg/l), cannot tolerate inhibitory compounds, cannot operate at low temperatures (<35^0C) and has poor removal efficiencies.

However, aerobic treatment is plagued by the problem of bulking sludges (which can cause catastrophic process failure), it produces large amounts of waste activated sludge (which has to be treated before disposal) and it consumes substantial amounts of energy. In the last 15 years, considerable advances have been made in understanding the complex microbial processes that occur in anaerobic digestion and in designing reactors suitable for the process.

One of the major problems confronting engineers working in this field was due to the slow growth rate of mixed anaerobic cultures and the long residence times required for effective solids digestion. This resulted in very large (and costly) vessels. Furthermore, due to very low yield coefficients with dilute feeds (500 mg/l COD) the biomass that develops is very dilute without substantial cell recycle. Hence, the challenge has been to develop a reactor that enables the hydraulic retention time (HRT) to be separated from the cell or solids retention time (SRT).

The development of such a reactor began in 1969 with the anaerobic filter (AF), and then with the anaerobic attached film fluidized bed, and culminated with the up-flow anaerobic sludge blanket (UASB). The UASB has been used extensively in many countries for the treatment of strong industrial wastes. Although the UASB has many advantages, it sometimes takes many months for the granules to develop, on which its operation depends, and it is susceptible to hydraulic and organic shock loads.

The Anaerobic Baffled Reactor (ABR) is a development in the design of anaerobic digestion reactors. The ABR consists of a series of baffled compartments containing freely suspended biomass through which the wastewater is forced to flow. The ABR does not require the presence of granules to operate, and has been proven (in laboratory, pilot and full scale trials) to be very robust to most types of shocks.

2.2 What is an Anaerobic Baffled Reactor (ABR)

An Anaerobic Baffled Reactor (ABR) is a wastewater treatment technology that uses baffles to create multiple treatment zones in a primary clarifier. A sludge blanket is established in each baffled zone, and different microbiological populations establish themselves in each zone. The overall effect is to provide both primary treatment (solids settlement) and sludge destruction in a single basin. This compartmentalized design separates the solids retention time (SRT) from the hydraulic retention time (HRT), so wastewater can be treated anaerobically at short retention times (comparable to ordinary primary treatment retention times).

ABRs can be viewed both as an alternative to primary clarification and as partial anaerobic secondary treatment providing *in situ* solids destruction. The ABR design being tested by OCSD is intended specifically to be a retrofit to existing primary basins. Previous demonstrations have been designed to create multiple compartments using baffles such that the upflow velocity in each compartment does not cause substantial wash-out of the solids blanket. The number of compartments can be varied and for the purposes of the pilot demonstration at Orange County five compartments were used.

Methane fermentation is the consequence of a series of metabolic interactions among various groups of microorganisms. The first group of microorganisms secrete enzymes which hydrolyze polymeric materials to monomers such as glucose and amino acids which are subsequently converted to higher volatile acids and acetic acid. In the second stage, hydrogen-producing acetogenic bacteria convert the higher volatile fatty acids (propionic and butyric) to produce hydrogen, carbon dioxide and acetic acid. Finally, the third group, methanogenic bacteria convert hydrogen, carbon dioxide and acetate to methane and carbon dioxide.

2.3 The Expected Benefits of ABR Technology

ABR technology can be installed at full-scale, as previously demonstrated at Culmore, Northern Ireland (see section 3.4) with relative ease through simple modification / retrofitting of a conventional primary settlement tank. The costs associated with converting a primary tank to an ABR are an important element of the overall cost-benefit appraisal and vary depending on specific site conditions such as tank size, tank geometry and type of desludging mechanism. Conversion of a primary tank requires the installation of baffle walls to create separate compartments and modification to the desludging mechanism.

The expected benefits associated with ABR technology include:

- **Reduced sludge production:** A solids residence time that can exceed several days in the ABR results in significant *in situ* degradation of the primary sludge solids. It also offers potential for decreased loading on the secondary treatment plant so less secondary sludge could be expected. Previous work has demonstrated that a reduction in sludge production between 30 - 50% can be expected.

- **Short Hydraulic Retention Times (HRTs):** The retention times for the ABRs operated in the UK were longer (5 – 7 hours) compared with the US, where primary clarifiers typically operate between 2 – 4 hours. Since the hydraulic loadings on the ABR will be similar to those for the existing primary clarifiers, no new, costly or dedicated reactors would be required.

- **Treatment of low strength wastes in a primary tank:** The pilot trial at OCSD was designed to determine the precise levels of treatment that can be achieved with low strength wastewaters. Previous work had indicated that COD and BOD removal efficiencies in excess of 50% can be achieved.

- **Low operating cost:** The operating costs of an ABR are expected to be similar to those for a primary clarifier. The ABR can be regarded as a low cost treatment option, especially when compared to the high energy costs associated with conventional activated sludge secondary treatment.

- **Robust to shock loads:** As outlined below, previous work indicated that the ABR is a stable process which is robust to both organic and hydraulic shock loads.

2.4 OCSD and ABRs – desktop study

ABRs traditionally have been used to treat high strength industrial wastewater. In 2001, Atkins completed a successful 2 year research project to develop and test ABRs for treating domestic wastewater. This work was funded by a group of UK water companies and OCSD. A full-scale ABR retrofit was then constructed and operated at a treatment plant in Northern Ireland.

During the original project the ABR benefits were demonstrated under conditions typical for UK treatment plants. These differed from conditions specific to Southern California where the average temperature is warmer, primary clarifiers operate with shorter HRTs, the wastewater has lower TSS, and enhanced treatment with ferric salts and polymer may be used.

In January 2002 OCSD commissioned work to evaluate the costs of benefits associated with implementation of ABR at OCSD Plant 1 and Plant 2. The evaluation concluded that ABR implementation had the potential to greatly reduce or eliminate the need for future digester construction at both Plants 1 and 2, with commensurate savings in future capital cost outlays and a net present saving to OCSD.

The installation of ABR at plant 1 was estimated to have a net present value as high as $18 million over a twenty year period. This was based on a reduction in required digester volume and installation of full secondary treatment. The net present value was based on less extensive modifications being carried out to the primary basins. However, even with more costly modifications for primary tank modifications, the range of net present values for plant 1 was estimated at between $5.6 million and $8 million.

The potential savings from reduced operations costs and reduced future facility requirements identified in the report led to the decision that a further evaluation of ABR technology at OCSD should be conducted. To better evaluate ABRs specific to OCSD, a 5-compartment pilot plant was therefore constructed, treating up to 0.3 MGD. Note that for full-scale installation, the compartments would instead be formed by baffles inside a single primary basin.

3. PREVIOUS WORK BY ATKINS

In 2001 Atkins completed a successful 2 year research project to develop and test the application of Anaerobic Baffled Reactors (ABRs) for the treatment of domestic wastewaters. A consortium of water companies and authorities funded this work. These included Northern Ireland Water Services, Wessex Water, Thames Water, South West Water, North West Water and Orange County Sanitation District.

Following the success of the pilot plant operation at Ellesmere Port (United Utilities formerly North West Water) a full scale retrofit was planned for Northern Ireland Water Services at Culmore WwTP. The benefits of an ABR for minimizing sludge production could be easily realized at this site since the sludge from the primary clarifiers was being trucked more than 70 miles to Belfast for incineration.

3.1 Novelty of the Project

One of the most novel aspects of the original research work completed by Atkins was that it involved the treatment of relatively low strength, low temperature wastewaters in an ABR that could be retrofitted into an existing primary tank.

This is in contrast to previous research work concentrated on the application of anaerobic technology as an alternative process to conventional (aerobic) secondary treatment processes.

The expected benefits associated with retrofitting ABR technology into a primary tank include:

- Maximization of existing capital assets
- Reduced load on downstream conventional secondary treatment plant
- Reduced sludge production (both primary and secondary)
- Low operating cost (no moving parts)
- Robust to shock loads
- Relatively simple construction

3.2 Pilot plant operation at Ellesmere Port WwTW

A pilot scale demonstration plant was operated for 2 years at Ellesmere Port sewage treatment works in the North West of the UK. The plant was designed with 5 tanks, representing separate compartments in a conventional primary treatment basin (Figure 3.1).

Figure 3.1 ABR pilot plant at Ellesmere Port

The plant was operated with an HRT of between 5 – 7 hours (similar to a conventional primary tank in the UK). Towards the end of the experimental period the reactor was run under diurnal flow variations, representing the actual flow to full treatment experienced at a conventional wastewater treatment plant.

The plant was successful in separating the HRT and SRT (with the SRT being greater than 40 days on occasions). This long retention time maximized the potential of the reactor for solids degradation since there was a greater period of time over which the bacteria in the reactor could degrade the organic matter in the sludge.

The organic loading rates (OLR) ranged between 0.13 - 0.25 lb/ft^3/day (2 - 4 kg COD/m^3/day). The variations in OLR were due to changes in the influent wastewater characteristics and were beyond the control of the Atkins research team.

The pH of the supernatants remained at or above neutral and did not exceed 8. The percentage removals for total and volatile solids typically ranged between 70 - 90% (considerably higher than a conventional primary tank without chemical assistance). The lower removals tended to occur when the influent wastewater was dilute (as would be expected).

With regards BOD and COD removals these were between 30 - 70%, again with the lower percentage removals occurring when the influent wastewater was most dilute.

Biogas was not produced in significant quantity, mainly due to the low temperature of the effluents and the high solubility of methane and carbon dioxide at these temperatures. Warmer effluents would produce a more predictable quantity of biogas.

The volatile solids concentration of the sludges inside the ABR decreased significantly across the reactor, reducing from between 70 – 80% in the influent to below 50% in the final compartment.

A detailed solids mass balance was undertaken on the reactor. The volume, TS, VS and pH of sludge removed from each de-sludging port were monitored throughout the project. From this data, combined with the flow data and effluent data a detailed mass balance was completed. The mean solids destruction rate was 44% (i.e. the sludge production was 56% of the expected total).

To confirm the validity of the mass balance calculation a similar mass balance was undertaken on the inorganic fraction of the sludges. Clearly the inorganics in the sludge will not be degraded and thus it should be possible to account for most of the inorganics added to the reactor. The mean inorganics balance was 103% of the expected total, thus confirming the validity of the mass balance approach.

3.3 Scale Up Assessments/Costing Evaluation

A key part of the research project was to undertake a costing exercise to evaluate the benefits of retrofitting an ABR into an existing primary tank. A number of sites were visited across the UK and in the US (the project being funded by a consortium of UK and US partners).

As would be expected the primary tanks on each site tended to have very specific characteristics. However a detailed costing exercise was undertaken on two sites, a rectangular primary basin at OCSD and a radial primary basin at Culmore, Northern Ireland. However on both sites a design was proposed to ensure that the key characteristics of the ABR system (for example the separation of the solids retention time from the hydraulic retention time) were maintained.

An economic assessment of the ABR system indicated a payback period of between 1 – 1.5 years, depending on the population being served, the design of the primary tanks and, perhaps most importantly, the downstream process plant (i.e. whether or not the site had anaerobic digestion). The key saving related to the reduction in sludge production, although an additional saving in terms of reduced downstream aeration in the secondary treatment plant was also identified.

3.4 Full-scale retrofit at Culmore Wastewater Treatment Works

Culmore Wastewater Treatment works is in the Londonderry area of Northern Ireland and is operated by Water Services Northern Ireland. The works serves a population of between 80,000-100,000 with the intention to expand it in the future. The plant only had primary settlement (i.e. there is no secondary treatment). The primary sludge is thickened in a stirred tank and then trucked over 70 miles to Belfast for incineration, presenting a significant cost burden to Water Services.

The cost for sludge incineration in Northern Ireland is approximately $80/wet ton ($320 - $400/TDS). An alternative option is disposal to landfill but this costs $110 wet ton ($430 – $540/TDS). Recycling sludges to agriculture is considerably cheaper (at $16/wet ton or $80/TDS for transport and spreading) but this outlet is considered impractical in Northern Ireland due to the small area of land available for recycling organic material and competition from agricultural waste needing to be recycled to the same land.

The site was selected for the ABR retrofit for a variety of reasons, primarily to evaluate the benefits of ABR for minimizing sludge production and therefore offsetting some of the site disposal costs for Water Services. In addition, the site had the ability to pump sludges from a single converted primary tank into a dedicated sump. By fitting a fixed speed desludging pump to this line it was therefore possible to provide a mass balance on the sludge removed from the ABR retrofit system. Furthermore the flow into each of the primary basins was controlled by a weir (rather than a hydrostatic head). This gave considerable leeway for altering the flow into the retrofitted tank as part of the full-scale demonstration process. Also the tanks did not need to be covered as the site was not in any close proximity to local housing or sensitive properties.

The ABR was retrofitted into one of the unused primary basins (Figure 3.2). The retrofit work involved the installation of baffles to divide the primary basin into a 4 compartment ABR, and installation of a desludging system (Figures 3.3 and 3.4).

Figure 3.2 Primary Basin prior to ABR retrofit

Figure 3.3 Primary basin following ABR retrofit - note the three sets of baffles

Figure 3.4 Gantry showing the de-sludging lines

The Culmore ABR operated continuously for 230 days at target hydraulic retention times of approximately 5 – 6 hours. The performance data was in line with expectations and the benefits of the ABR technology were seen at full-scale.

In summary, the stability of the ABR plant was seen to improve as the project progressed, particularly during the last 6 - 8 weeks of the project:

- It operated more as a biological reactor than a conventional primary clarifier, despite the low winter temperatures
- Solids capture rates were comparable to the control primary basin (although the HRT for the ABR was in fact lower), with values in excess of 80%
- COD removal was also comparable to the control, performing in excess of 50%.

The solids retention time across the ABR was approximately 20 days for the first part of the project but decreased in the latter parts as a more frequent desludging regime was incorporated. Throughout the project, the SRT (in days) was significantly greater than the HRT (in hours) i.e. the ABR was successfully trapping the solids within the basin.

A mass balance was undertaken, indicating that the ABR was achieving a sludge reduction of approximately 20% for the last 6 – 8 weeks of operation. The data for the sludge production across the whole site (i.e. including the conventional primary tanks and imported sludges) was also examined. This too confirmed that the ABR was performing as expected.

The full-scale ABR application proved a number of important aspects of the ABR design in terms of operation and performance at full scale. It highlighted areas where the design compromised the performance to some extent. These will be useful for future ABR operation.

Other sites were also visited as part of the project in order to evaluate the potential for ABR installation and operation. Assessments were undertaken and typical ABR designs put forward for the future. Based on a number of assumptions, the costs and benefits for ABR retrofit were also evaluated. This exercise indicated that the conversion of the clarifiers with larger operating volumes presents a much more favourable option in terms of payback.

4. ABR FOR OSCD

As a result of the potential savings from reduced operations costs and reduced future facilities requirements (section 2.4) it was agreed that a further evaluation of ABR technology at OCSD should be conducted. Therefore to better evaluate ABRs specific to OCSD, a 5-compartment pilot plant was constructed, treating up to ~ 0.3 MGD (200 gpm). Initially this formed part of the ongoing micro-filtration project with Carollo Engineers.

The main drivers for ABR operation centered around increasing sludge processing and disposal costs and a move to full secondary treatment in the future.

Following confirmation of ABR performance through the pilot-scale testing, the MWH report recommended full-scale evaluations by converting two of the existing rectangular primary clarifiers into ABRs.

4.1 Site specific conditions

The objectives of the pilot trial were to confirm expected ABR performance and impacts under actual and site specific conditions, since these differed from previous ABR testing undertaken in the UK. The most significant operational differences were identified as:

- Warmer wastewater temperatures typical of Southern California
- Chemical addition (ferric and polymer)
- Lower influent wastewater concentration (lower TSS, BOD values)
- Shorter hydraulic retention times

A comparison of wastewater temperatures between the UK and southern California shows a difference of approximately 20 degrees F. Average influent wastewater temperatures in the UK are around 55 degrees F and in California approximately 75 degrees F. It was anticipated that this may enhance the *in situ* digestion capabilities. However, the increased gas production could also affect the settling characteristics of the sludge blankets.

Typically, in the UK chemicals are not added to the influent wastewater to enhance settlement in the primary clarifiers. OCSD adds ferric at the headworks and polymer to the distribution box. This allows the solids to settle more quickly. In addition, the HRTs in the primaries are much shorter in the US (3 hours compared with ~ 6 hours in the UK). The influent concentrations are also lower (in terms of TSS and BOD) and the particle size distribution found to be smaller. It was anticipated that these factors would also change the settling characteristics of the sludge.

For the reasons outlined above, it was important for OCSD to evaluate ABRs specific to their site conditions and mode of operation.

4.2 Design

The ABR plant was located at OCSD's Plant 2 and was initiated as part of the ongoing micro-filtration project. One of the objectives of the ABR was to enhance the quality of the feed to the MF plant and also to treat the backwash water from the MF plant. This complicated the design of the ABR because it had to be flexible in terms of flow rate. At full capacity, the MF plant would treat an influent flow of 200 gpm. However, the maximum flow for the back wash water was only 50 gpm. The design challenge meant it needed to treat flows varying between 50 – 200 gpm, but maintain an HRT of around 3 hours. A false baffle wall was therefore incorporated as part of the tank design (Figure 4.2) in order that the compartment size could be changed.

Design flexibility was maximized by representing each of the five compartments as a separate tank (Figure 4.1). Previous work highlighted that a 5 compartment ABR was effective. The tanks had interconnecting piping, feed and outlet piping manifolds, and individual drainage with dedicated isolating valves. Wastewater passed from one tank to the next via a series of connecting pipes and baffle plates (simulating the ABR weir). Each tank was sized at 14.8 ft (width) X 9.2 ft (length) X 7.2 ft (height) and designed to hold water to the full tank height. However, the operating volumes of the tanks were variable. This was achieved by moving the false baffle walls.

The moving baffle walls also enabled the up flow velocity to be optimized in each tank. Typically the up-flow velocity must be greater in the first compartment where the solids particles are larger. As the particle size distribution changes (decreases) down the length of the reactor, the up-flow velocities must be reduced to obtain the optimal settlement characteristics.

The raw influent wastewater flowed across the tank and the solids settled to form a sludge blanket, thus filtering the raw wastewater. Connector pipes and baffle plates directed the flow from the top of the first compartment to the base of the second tank. All tanks were connected in a similar manner.

The operation of the ABR in terms of sludge removal was automated. This could take place using either sludge blanket detectors to activate the desludge pumps when the level of the sludge blanket increased, or by running the desludge pumps from timers. Four fixed pipes extended into each ABR tank to remove the sludge.

Problems were experienced with the original method for desludging. Modifications were made and a manual desludge system was introduced after 6 months of operation so that a consistent sludge could be removed from each tank. This enabled a more robust mass balance to be calculated. When costing the ABR construction, it proved significantly cheaper to build in the UK and then ship to the US. Each tank was therefore designed to be transported by ship and then truck at either end. Figures 4.1 to 4.3 provide further detail on the ABR set-up.

Figure 4.1 Tank overview; 5 compartments

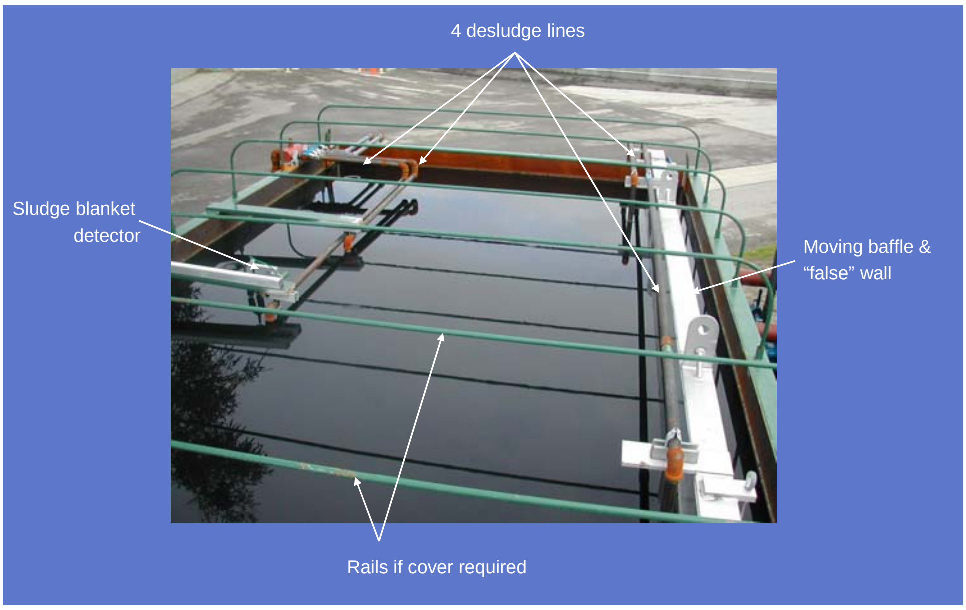


Figure 4.2 Varying compartment size

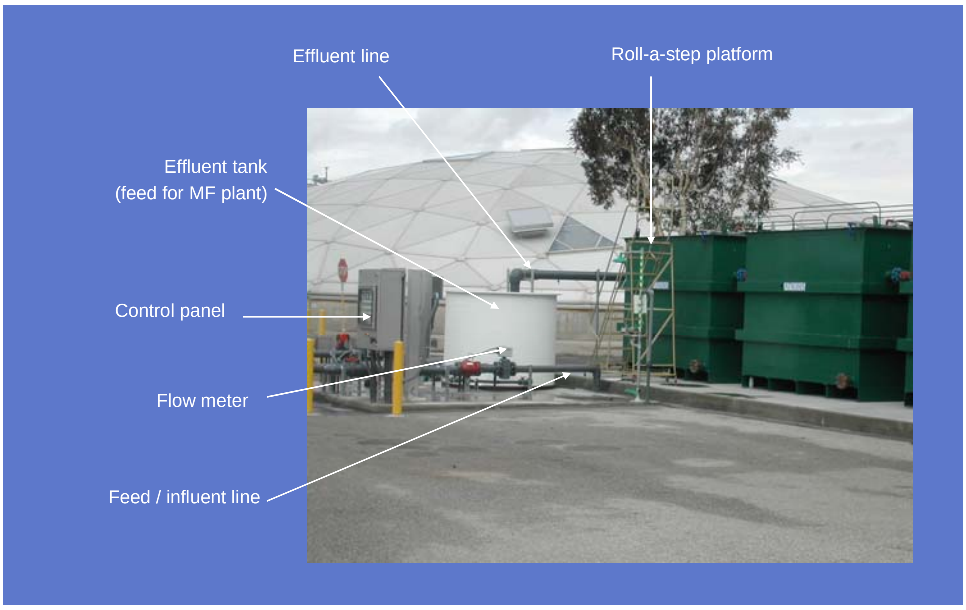


Figure 4.3 ABR controls

4.3 Shipping

The ABR was built in the UK and the tanks were shipped (unconnected) to the US where all the interconnecting pipe work, manifolds, pumps, control equipment and electrical instrumentation were installed. These all met with the required US standards. There was nearly an incident that would have severely delayed the project start-date. The original ship that was to transport the ABR tanks to Los Angeles never made it across and instead sank in the Atlantic (Figure 4.4).

Figure 4.4 What could have been!!

5. PROJECT OVERVIEW

The original scope of work for the ABR operation was from February to October 2003. The project was extended until May 2004 to assess performance using a manual de-sludging process and to generate performance data for a range of flows and polymer doses. During Phase 1 the sludge blankets were built up after seeding tanks 3 – 5 with primary sludge and the blanket depth was optimized. The ABR ran on a hydraulic retention time of 3.8 hours (130 gpm) with no chemical addition.

Table 5.1 Operational summary

Period of Operation February 2003 – May 2004	Period characteristics	Days run	Average flow gallons/day
Phase 1 (Feb 2003 – June 28)	Commissioning		
Phase 2 (June 29 – Oct 7)	Automatic de-sludging	100	189,500
Phase 3 (Oct 8 - Jan 11, 2004)	Manual de-sludging	95	184,100
Phase 4 a (Jan 12 - March 3)	Low flow/low polymer dose	51	161,900
Phase 4 b (March 4 - April 11)	High flow/low polymer dose	38	235,700
Phase 4 c (April 12 - April 30)	High flow/high polymer dose	18	230,200
Phase 4 d (May 1 - May 30)	High flow/low polymer dose	30	226,900

Table 5.2 Analytical measurements for sludge in tanks 1 thru 5.

Parameter	Units
Total volatile fatty acids	mg/l
Propanoic acid	mg/l
Acetic acid	mg/l
Butanoic acid	mg/l
Alkalinity	mg/l
Chemical oxygen demand	mg/l
Total solids	%
Volatile solids	%
Blanket depth	Feet

Table 5.3 Analytical measurements for influent and effluent

Parameter	Units
Turbidity	NTU
Total suspended solids	mg/l
Volatile suspended solids	mg/l
Volatile solids (as a % of total solids)	%
Chemical Oxygen Demand	mg/l
Total Biochemical Oxygen Demand (BOD-t)	mg/l
Soluble Biochemical Oxygen Demand (BOD-s)	mg/l
Alkalinity	mg/l
Ammonia	mg/l
Temperature	Degree F

5.1 Trial Extensions

During Phase 2 it proved to be difficult to calculate an accurate mass balance due to problems with the automated de-sludging mechanism and the inability to get a representative sample of the sludge removed from the tanks. The project was therefore given an initial 3 month extension and the plant was modified to a gravity-based manual de-sludging system (see section 5.2).

The first six months of testing had demonstrated that the ABR could achieve solids removal with no chemical addition. The visible evidence of gas production and the changing acid profile across the tanks indicated that the system was destroying solids *in situ*. Continuation of the trial provided the opportunity to develop more robust data with regards to:

- Solids destruction achieved by the ABR
- TSS removal with the addition of polymer
- Impact of chemical treatment on solids destruction and soluble BOD concentration

- Relationship between hydraulic retention time, chemical addition and TSS removal

Phase 4 provided data that could be used to generate a mass balance and assess the impact of polymer addition on ABR performance. Following an analysis and discussion of the results and potential for full scale ABR retrofit to primary clarifiers at OCSD a number of issues arose and an additional 3 month project extension approved. The following points were identified for investigation:

- TSS removal with the addition of polymer at a range of concentrations
- The impact of ABR operation on sludge digestibility
- The impact of ABR operation on sludge dewaterability both pre- and post-digestion
- The robustness of the methane solubility assumptions made in the solids mass-balance calculations for the ABR pilot plant
- The release of methane from solution downstream of the ABR and the associated risk of explosion

It was important to establish a better understanding of the critical parameters so that engineering options for primary basin retrofit could be explored and the original cost savings identified by MWH could be verified. The pilot-scale trial and associated laboratory and desk-top evaluations were designed to inform the decision whether or not to proceed to a full-scale retrofit demonstration.

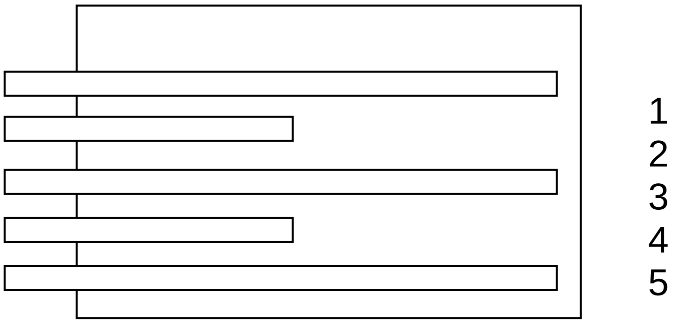
5.2 Desludging modifications

The desludging modifications were made to all ABR tanks in September 2003 to improve the desludging regime, enabling a more accurate measure of solids extraction thereby improving the accuracy of the mass balance. The main objectives of the changes were:

- To remove a known volume of sludge from each tank at regular intervals (de-sludging) in order to have a reasonably constant sludge blanket depth within the tanks
- To take samples during the desludging operation so that an accurate composite sample of the sludge could be obtained to establish the average solids concentration of the sludge removed from each tank
- To withdraw equal volumes of sludge from each pipe keeping the sludge as thick as possible during the process

Five pipes were connected to each tank. Three of the pipes were long (13 ft) and two were short (6 ft), to cover as much of the tank base area as possible. These could be pulled out and pushed back in. This enabled a more consistent sludge sample to be collected (Figure 5.1).

(a)



(b)

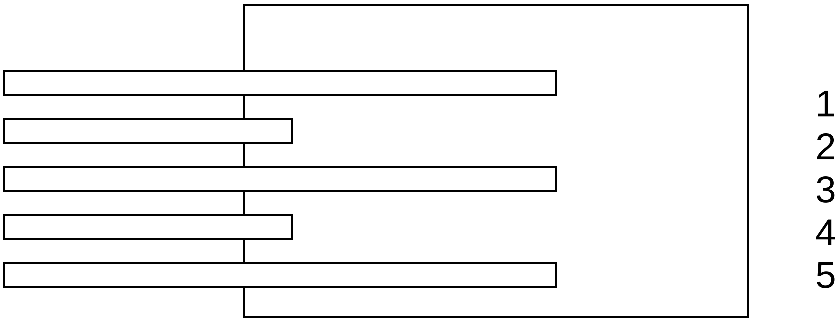


Figure 5.1 New desludge pipes in (a) the "in" position, (b) the "out" position

5.3 Chemical addition

The modified desludging process resulted in a more robust solids mass balance. The next step was to test the effect of chemical addition (using ferric and polymer) on solids removal. This was highlighted in the original MWH report as a significant parameter for further investigation. The chemical addition trials provided important data previously lacking from the project work, specifically direct comparison of primary basin with ABR performance given similar chemical dosing rates.

One month was the minimum time required to get a reasonable indication of the impact of chemical addition at a constant HRT. The intention was to compare performance of the pilot against the primary clarifiers. In addition, it provided useful data on ABR performance with and without chemical addition.

5.4 Methane generation

The digestion process in the ABR produces methane gas that can be dissolved in the liquid phase or be released as gas from the surface of the tanks. It was necessary to understand the cycle of methane absorption and release for 2 principal reasons:

1. The concentration of dissolved methane in the ABR influent and effluent formed part of the assumptions used in mass balance calculations.

2. The release of methane from solution within the ABR and in downstream processes presents a potentially explosive risk.

An increase in temperature could result in more dissolved methane coming out of solution. It was therefore important to assess the risk as part of the OCSD project rather than infer from data previously collected in the UK. Full details of the methane analyses performed and the results obtained are included in Section 6.

5.5 Digestibility/dewaterability

Since the ABR operates as a reaction vessel as well as a clarifier, the sludge produced can be considered as partially digested. In order to completely understand the potential costs and benefits associated with a full-scale retrofit of an ABR it was necessary to assess the inherent properties of ABR sludge.

A comparison of ABR sludge with conventional primary sludge was performed through digestibility and dewaterability tests at bench scale. It was anticipated the results would provide a qualitative assessment to the following points:

- The impact of the ABR on pre-digestion sludge thickening
- The impact of the ABR on post-digestion sludge thickening
- The impact of the ABR on gas generation during anaerobic digestion

Full details of the laboratory work performed are included in Appendix B.

6. RESULTS AND DISCUSSION

The data presented in the report covers the duration of the ABR pilot trial at OCSD from February 10, 2003 until May 30, 2004. This is best represented as four distinct phases of operation, as summarized in Table 6.1. This chapter will begin with a short overview of the operational conditions during each phase. The results and discussion will then focus on the objectives of the trial, as identified at the beginning of the report namely:

- To assess ABR performance under southern Californian conditions
- To review the impact of polymer addition on ABR performance
- To assess ABR performance under different hydraulic retention times
- To compare ABR effluent quality to primary tank effluent quality
- To determine the solids destruction achieved by the ABR
- To quantify the risks associated with release of methane
- To estimate the costs and benefits of full-scale ABR installation

The objectives listed above are divided into separate sections in the Results and Discussion chapter and will be presented in the same order. There is a degree of overlap between the sections but the report seeks to keep each section focused on the specific objective.

Table 6.1 Summary of Phases of Operation

Period of Operation Feb 2003 – May 2004	Key aspects of operation	Days run	Average flow gallons/day
Phase 1 (Feb 10 2003 – June 28)	Commissioning		
Phase 2 (June 29 – Oct 7)	Automatic de-sludging	140	189,500
Phase 3 (Oct 8 - Jan 11, 2004)	Manual de-sludging	95	184,100
Phase 4 a (Jan 12 - March 3)	Low flow/low polymer dose	51	161,900
Phase 4 b (March 4 - April 11)	High flow/low polymer dose	38	235,700
Phase 4 c (April 12 - April 30)	High flow/high polymer dose	18	230,200
Phase 4 d (May 1 – May 30)	High flow/low polymer dose	30	226,900

A short summary will conclude the Results and Discussion chapter, highlighting the main findings to emerge and areas that require more investigation before firm conclusions can be drawn.

6.1 Summary of operational conditions during each phase

Phase 1

During this time the ABR was under a period of commissioning, stabilization and optimization. The data from this phase is not presented in results as it is not representative of overall performance.

The influent to the ABR was ferric dosed during phase 1 to help the solids settle in the tanks and the sludge blankets to form. Tanks 3 thru 5 were seeded with primary sludge to reduce the time required for sludge blanket establishment and stabilization. Towards the end of phase 1 the sludge blankets in all tanks had built up to approximately 3 feet in depth. The desludging was automated using sludge blanket detectors to activate the desludge pumps. During phase 1 grab samples were collected. Based on the results during phase 1 it was observed that:

- The sludge blankets were too deep leading to excessive solubilization of COD and BOD. This in turn resulted in effluent COD and BOD concentrations exceeding the influent concentration.
- Solids removal across the ABR deteriorated as the depth of the sludge blanket increased.
- Production of volatile fatty acids increased during the period, resulting in a drop in the pH between the influent and effluent.

Phase 2

Phase 2 of operation ran from the end of June 2003 to beginning of October. No data was reported from May 10 to June 29 2003 because the feed (transfer) pump located at the headworks failed and had to be removed and repaired. During this time ABR operation was not representative because it was being fed with backwash water from the micro-filtration plant at a rate of 15 – 20 gpm. During the rest of the phase the flow averaged about 130 gpm to give an HRT of 3.8 hours.

By phase 2 the sludge blankets were properly established. The influent transfer from the headworks was therefore switched from ferric dosed to a non-ferric dosed feed. Based on the results from phase 1 it was apparent that the depth of the sludge blankets was leading to a deterioration of the effluent quality. Therefore the blanket depth was reduced and maintained at approximately 1.5 feet.

Operational problems with the automated desludging system led to its replacement with a manual system at the end of September 2003, as described in section 5.2.

Other operational changes included:

- Introduction of refrigerated composite samplers at the ABR influent and effluent, programmed to produce a 24 hour composite sample from a series of grab samples taken every 30 minutes
- Installation of gas collection cones on tanks 2 thru 5 to enable the gas volume to be calculated and to provide samples for analysis in the laboratory

Phase 3

Phase 3 ran from October 8, 2003 until January 11, 2004. The flow rate averaged 130 gpm to give an HRT of about 3.8 hours. The most significant change during phase 3 was the use of the manual desludging system.

The new manual system was much more labor intensive but it enabled thicker sludge to be withdrawn from each tank and a representative sample to be obtained. Five pipes were connected to each tank. These were designed to cover as much of the tank base area as possible and could be pulled out and pushed back in. During desludging composite samples was collected for analysis (total solids and volatile solids) and the volume of sludge removed was calculated based on the fall in the level dropped inside the tank.

Phase 4

Phase 4 incorporated a program of chemical dosing at low and high flow rates. The key operational parameters that were adjusted during phase 4 included influent flow and polymer dose. The polymer dose was based on the dose used for the A-side primary basins at plant 2 to facilitate comparison between ABR performance and full-scale performance. The ABR received a ferric-dosed feed from January 12, 2004 and polymer dosing commenced on January 19 2004 (days 337 and 344 respectively). Ferric was dosed at the headworks and anionic polymer was dosed via a small polymer dosing pump upstream of the ABR influent. Polymer additions were calibrated to mirror the dosing concentrations used on site. Polymer additions and flow rates were varied according to the periods detailed in table 6.2.

Table 6.2: Polymer addition and influent flow rates during phase 4

Dates (2004)	Average flow (gpm)	Polymer dose (mg/l)
Phase 4a **Jan 12 – March 3**	Low (113)	Low (0.2)
Phase 4b **March 3 – April 12**	High (162)	Low (0.2)
Phase 4c **April 12 – April 30**	High (162)	High (0.4)
Phase 4d **April 30 – May 30**	High (158)	Low (0.2)

6.2 Data analysis and interpretation

This section presents a detailed summary of the performance data. The data from Phase 1 is not considered because the plant was being commissioned and the sludge blankets needed to stabilize. The short duration of phase 4b thru d resulted in a small data set being produced for these respective phases, particularly with regard to COD and BOD data which were not taken on a daily basis. Phase 4a ran for a longer period from Jan 12 to March 3 2004 and therefore provided a more robust and statistically significant dataset. Table 6.3 records the number of data-points obtained in each time-period of phase 4. Caution needs to be applied to interpretation of the results for sub-phases because of the limited dataset for COD and BOD concentrations.

Table 6.3 Number of data-points for specific parameters during Phase 4

	TSS	VSS	COD	BOD
Phase 4a	31	31	13	20
Phase 4b	21	21	6	12
Phase 4c	15	15	6	9
Phase 4d	18	18	7	11

6.2.1 ABR Performance under southern Californian conditions

It was noted in section 4.1 that climatic and operational conditions in southern Californian differ significantly compared to Northern Ireland and the UK where previous full-scale and pilot demonstrations for the ABR have been conducted. Figure 10 presents the hydraulic retention time for the entire ABR and for individual tanks. The data presented is based on a running seven day mean used as standard for presentation of data. A seven day running mean takes into account weekly trends and provides smoothing of data to aid with the interpretation of results. Tanks 2 thru 5 were the same size and therefore have identical HRTs and superimpose on the graph. The compartment size for tank 1 was slightly smaller. The up-flow velocity in tank 1 can be greater because the particle size of solids that settle in tank 1 are typically larger and therefore the solids settle faster.

The HRT averaged between 3.5 and 4 hours during phase 2, with a maximum of 4.5 hours HRT in phase 3 and 4 and a minimum of 3 hours in phase 4. Previously, the ABR demonstrations had operated with an HRT of 5 – 6 hours and therefore the operation of the ABR at OCSD successfully tested performance at significantly shorter HRTs.

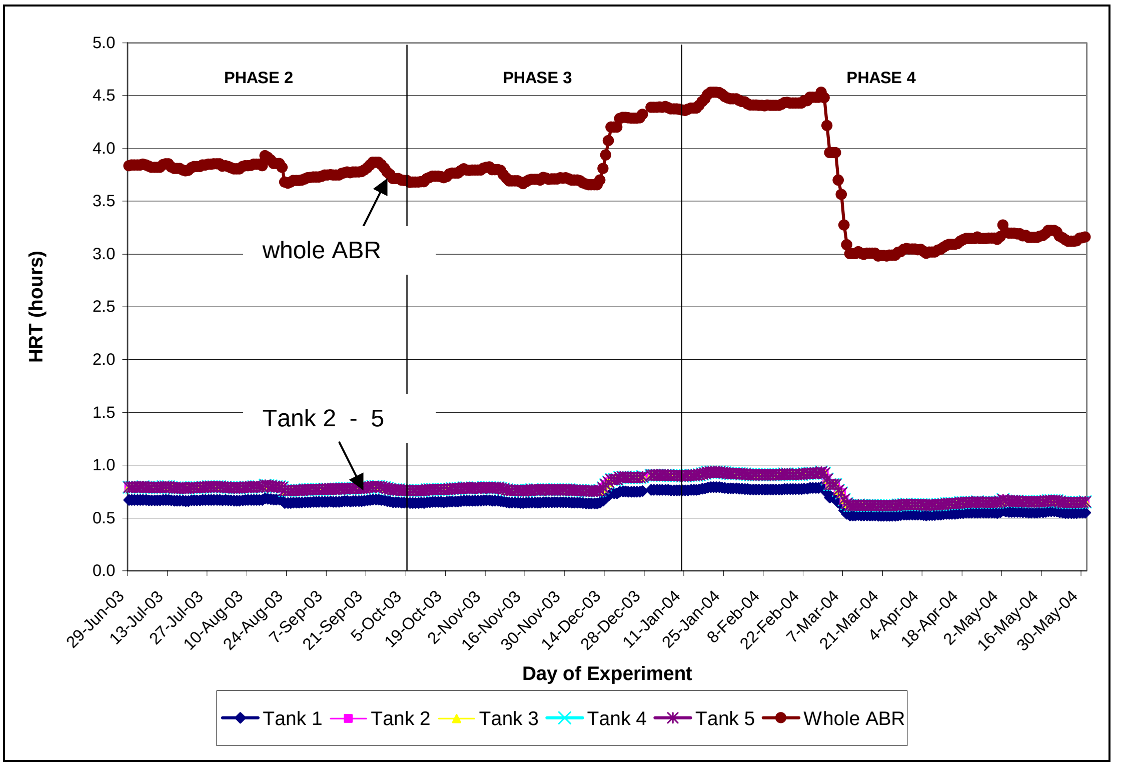


Figure 6.1 Mean HRT (whole ABR and for each tank – 7 day running mean)

One of the initial objectives of the work for OCSD was to evaluate the effect of warmer temperatures, compared with the UK, on ABR operation and performance. Figure 6.2 presents the seasonal variation in temperature during the trial and it was found that California wastewater is approximately 20 degrees warmer than wastewater in the UK.

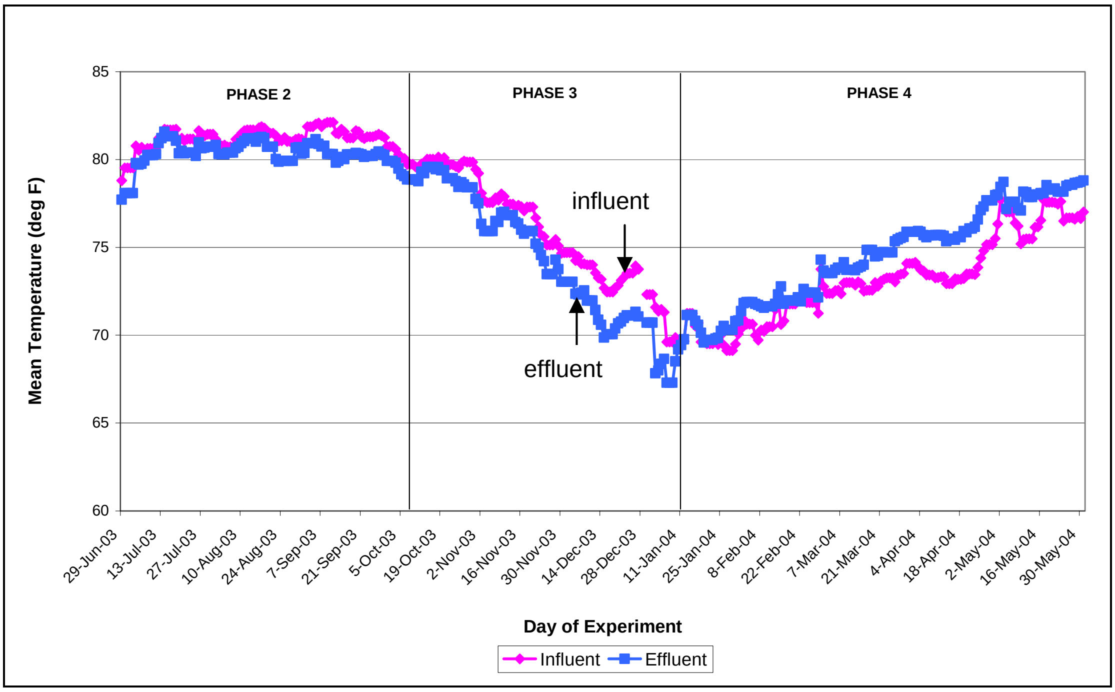


Figure 6.2 Influent and effluent temperatures (daily)

The elevated temperature was expected to have two impacts. Firstly the warmer temperatures could enhance the *in situ* sludge digestion and secondly it may result in more methane release as the solubility will decrease with increasing temperature. In addition, increased gas production may prevent some of the finer solids settling. The small fluctuation in wastewater temperature throughout the year (about 15 degrees) indicates that there will be limited seasonal impact of temperature on ABR performance.

Brief summary of selected performance parameters

Three aspects of performance are included in this section, specifically:

- pH across the ABR tanks
- volatile solids within the tanks
- variation in turbidity

More performance data will be presented later in the chapter in the sections that are most appropriate. The solids removal across the ABR will be discussed in section 6.2.4. Section 6.2.6 explores the release of methane gas from the ABR, although in the absence of gas sampling for the full-scale Northern Ireland installation or the pilot demonstration at Ellesmere Port, a relationship between methane release and temperature cannot be deduced.

One of the principal benefits of ABR operation concerns the *in situ* solids destruction. The design of the ABR separates the hydraulic retention time from the solids retention time and the different compartments help to isolate the various stages of digestion. Microbes acclimatize to the conditions specific to each tank and digestion environments are established with hydrolysis followed by acidogenesis and finally methanogenesis in separate tanks.

The trend in the sludge pH as seen in figure 6.3 is characteristic of an ABR, with increasing sludge pH down the length of the reactor. Hydrolysis and acidogenesis take place at the front end of the reactor (hence the more acidic environment) and methanogenesis further down. This increase in pH is one indicator to demonstrate that the ABR is operating as anticipated. Note that the pH seen in tank 1 is higher than tank 2. This may not be expected but can be attributed to the removal of sludge from tank 1 on a regular basis.

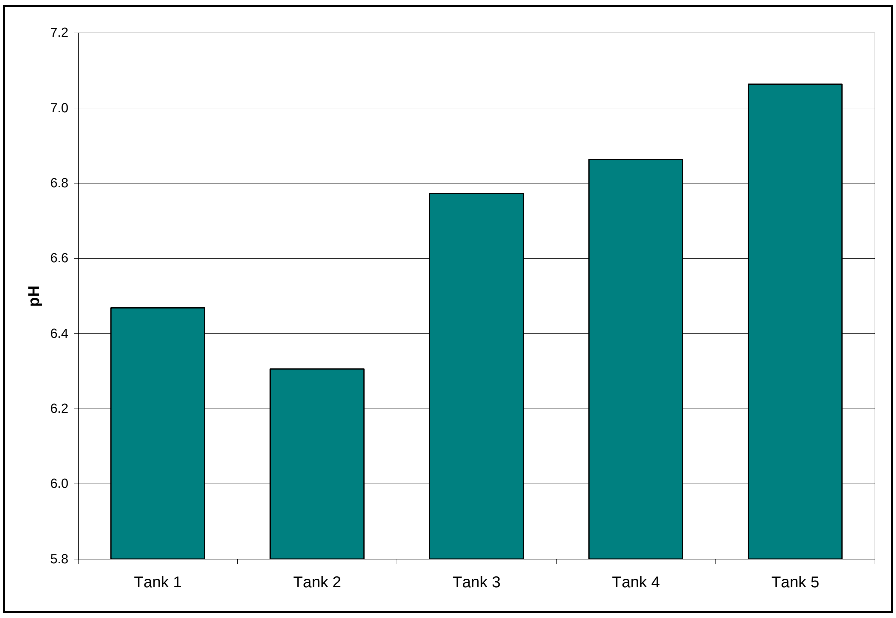


Figure 6.3 Mean sludge pH in each tank for phases 2 thru 4

The data in figure 6.4 also demonstrates *in situ* sludge digestion. The volatile solids concentration (as a percentage of the total solids) was highest in tanks 1 and 2 at nearly 62% and decreases down the length of the reactor to less than 53% in tank 5.

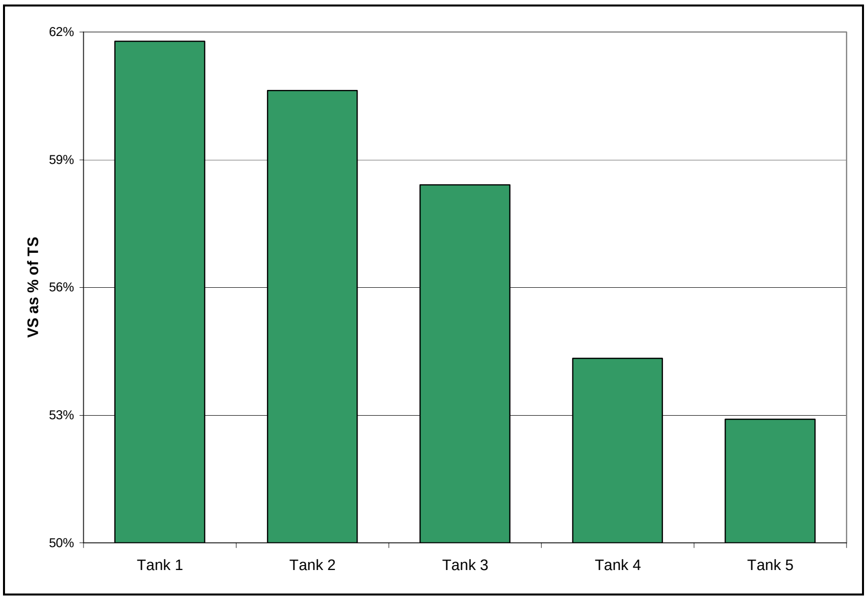


Figure 6.4 Mean sludge VS (as a percentage of TS) for each tank

The turbidity values (figure 6.5) demonstrate an interesting trend in terms of diurnal variation. The influent turbidity readings are considerably higher in the afternoon when compared to the morning. This demonstrates the importance of taking a composite sample, therefore representing the whole day of operation. The effluent samples were taken at the same time as the influent. Consistently the levels are very similar for the morning and the afternoon indicating that the ABR acted as a buffer and was robust to diurnal variations in turbidity.

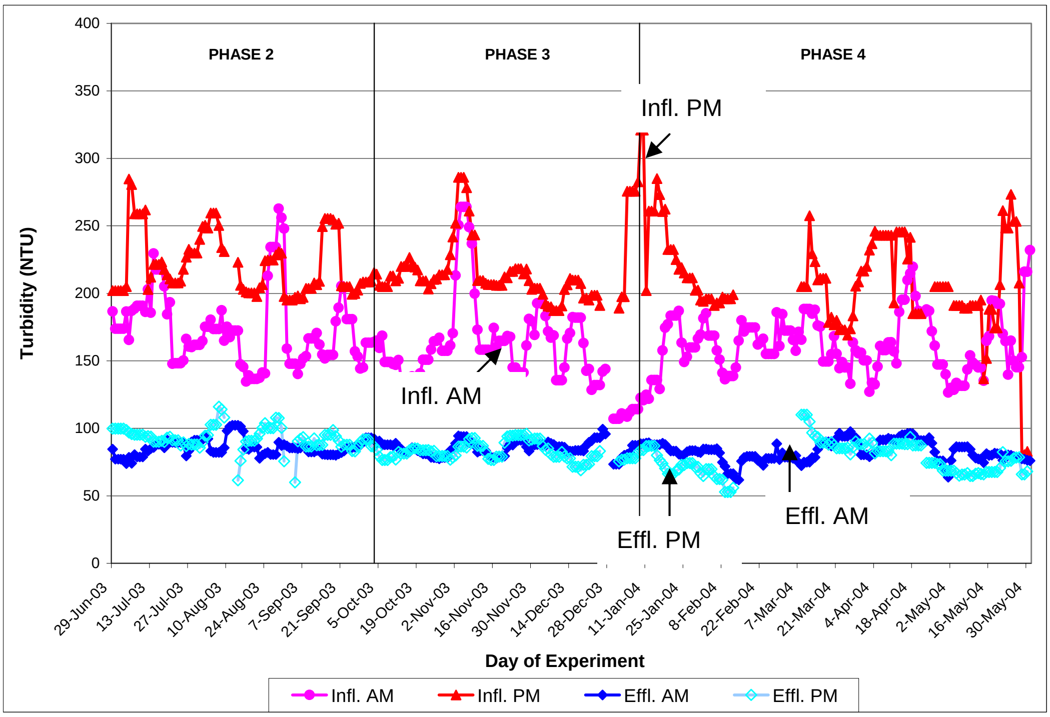


Figure 6.5 Mean influent and effluent turbidities

6.2.2 Impact of polymer addition on ABR performance

Anionic polymer was added to the ABR influent during phase 4 as previously outlined in table 6.2. The overall performance of the ABR is presented in table 6.3. The polymer used to dose the ABR was taken from the full-scale polymer dosing plant. It was intended to dose polymer at concentrations used in the primary tanks. However, due to a malfunction of the meter used to measure the polymer makeup water on the full-scale plant, the polymer solution was more concentrated than assumed.

Therefore, the ABR and full-scale primary tanks received a polymer concentration of 2 mg/l and above during the period of the trial, higher than the target of 1 mg/l. This may explain in part the lack of improvement in performance when the polymer dose was increased from 2 mg/l to 4 mg/l between phase 4b and 4c. If the polymer was already being dosed at the maximum effective concentration (identified at about 2 mg/l from empirical full-scale evidence), any increase in dose was unlikely to improve performance and may even have had a detrimental effect by causing bulking and floating of sludge. Such impacts have been observed at full scale and floating sludge was also recorded on tank 1 of the ABR during periods of high polymer dosing.

The comparison between phases provides evidence that VSS and TSS removal was enhanced by the addition of polymer. It is difficult to draw any firm conclusions about the impact on COD and BOD removal due to the small size of the dataset, but the average COD and BOD removal during phase 4 exceeds phase 3.

Table 6.3 Summary of ABR performance with and without polymer dosing

Phase	HRT (hours)	Polymer dose (mg/l)	% TSS removal	% VSS removal	% COD removal	% BOD removal
2	3.8	0	50	53	23	20
3	3.9	0	52	55	17	20
4a	4.4	2	65	66	29	24
4b	3.1	2	58	57	23	30
4c	3.1	4	58	59	25	35
4d	3.2	2	62	64	25	25

TSS removal – no polymer dosing

Figure 6.6 provides an overview of the influent and effluent TSS concentration. It was highly variable on a day-to-day basis but the effluent TSS concentration was more stable indicating that the ABR acts as a buffer to variations in the influent load. During phases 2 and 3 the influent TSS concentration averaged 230 mg/l and the effluent TSS concentration averaged 108 mg/l. The average TSS removal was 52% during phase 2 and 53% during phase 3 without the addition of any chemicals.

TSS removal - with the addition of polymer and change in HRT

During phase 4, the changes made to chemical addition and flow rate had a significant impact on TSS removal achieved by the ABR. Addition of ferric and polymer initially led to a drop in effluent TSS concentrations to a minimum around 70 mg/l and a corresponding increase in TSS removal to a maximum of 70% (Fig 6.6 and 6.7). The flow rate was then increased and effluent TSS concentrations rose to around 110 mg/l with a corresponding TSS removal of about 60%.

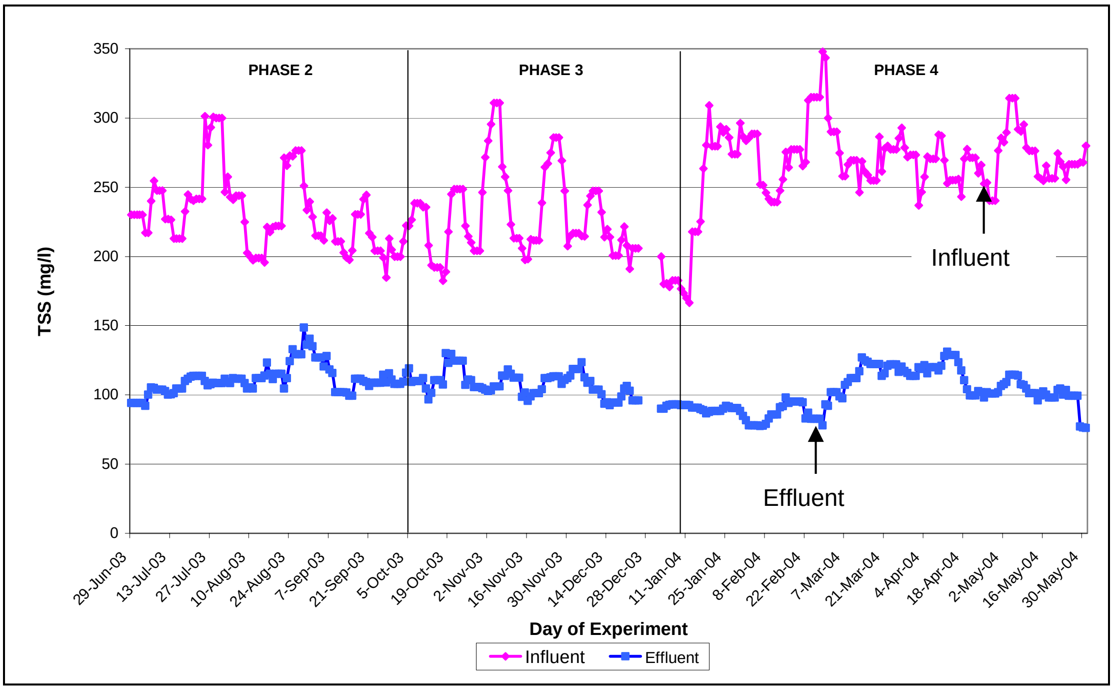


Figure 6.6 Mean influent and effluent TSS – 7 day rolling average

Note that, despite phase 4b thru 4d having a shorter HRT (3.1 hours) than phase 2 and 3, the ABR performed better. Figure 6.7 provides evidence of the enhanced %TSS removal following the addition of polymer in phase 4. The impact of HRT on overall performance will be discussed in the next section in more detail. In summary, it is reasonable to state that:

- Addition of polymer to the ABR improved performance in terms of %TSS and %VSS removal
- Due to a failure of the meter for the polymer makeup water, polymer was dosed at higher concentrations than intended. It is likely that the optimal polymer dose was below 2mg/l as evidenced by the lack of improvement in performance when the dose was increased to 4 mg/l.
- The polymer mixing system for the ABR constituted the short length of pipe between the dosing point and the influent to the ABR. This was unlikely to have been optimal.
- The average COD and BOD removal (as a % of influent) increased during phase 4 compared to phase 2 and 3 when chemicals were not dosed.

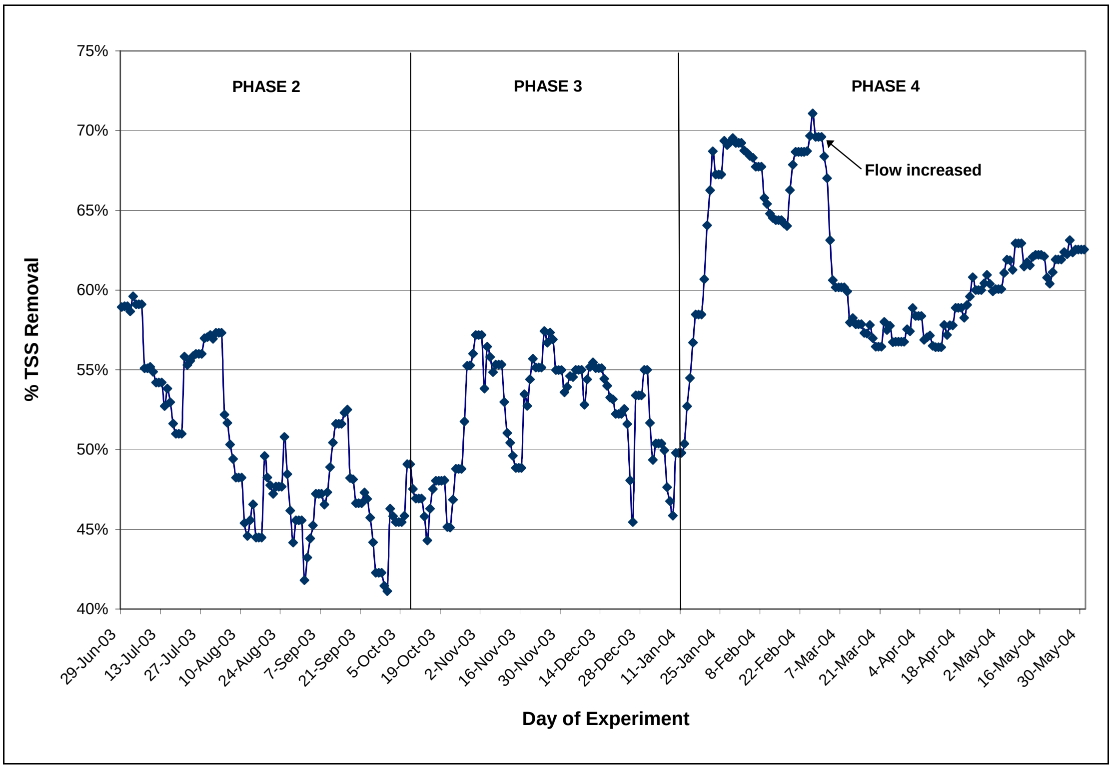


Figure 6.7 Mean TSS removal rates across the ABR (7 –day mean)

Overall chemical addition enhanced the performance of the ABR but it was not possible to optimize the dosing regime in order to obtain the best quality effluent.

6.2.3 ABR performance under different hydraulic retention times

The hydraulic retention time for the ABR ranged from 3.0 to 4.5 hours. There were two significant step changes in HRT. The first occurred in phase 3 when the retention time for the tanks was increased from about 3.7 hours to about 4.3 hours as presented earlier in fig. 6.1. The second step change occurred during phase 4 when the HRT was reduced from 4.4 hours to 3 hours. The reduction in HRT during phase 4 was during the period of ferric and polymer dosing.

A summary is presented in table 6.4. There was no discernable impact on performance in terms of % TSS and VSS removal in phase 3 when the HRT increased. The only parameter that shows a significant change was total BOD removal. This decreased as the HRT increased which at first appears counter-intuitive.

The change in performance when the HRT was decreased during Phase 4 produces a similar result for BOD removal. It is seen to decrease from 30% to 24% when the HRT increases from 3 hours to 4.4 hours. Further examination of the results is required to explain why solids removal is enhanced as the HRT increases (as would be expected in a primary settlement tank) yet BOD removal decreases.

Table 6.4 Performance of ABR at different hydraulic retention times

	TSS removal (%)	VSS removal (%)	COD removal (%)	tBOD removal (%)
Phase 3 (3.7 hours HRT)	53	55	17	21
Phase 3 (4.3 hours HRT)	51	54	18	14
Phase 4a (HRT 4.4 hours)	65	66	29	24
Phase 4 b,c,d (HRT 3 hours)	60	60	24	30

In order to understand better the inverse relationship between HRT and BOD removal, the ratios of COD to BOD were examined for the influent and effluent (see table 6.5 and 6.6). The influent and effluent ratios of COD:tBOD were found to be similar suggesting COD and tBOD were removed in similar proportions across the ABR (see figures 6.8 and 6.9).

The total BOD in the ABR effluent had a greater fraction in the soluble phase compared to the influent total BOD. Similarly the effluent COD:sBOD ratio was lower than the influent because there was a greater proportion of soluble BOD. One explanation for the decline in performance of the ABR in terms of BOD removal is that BOD is being solubilized as the HRT increases, without subsequently being converted to methane. This has important implications for the solids mass balance as will be discussed in section 6.2.5.

One consequence of the decrease in ratio of tBOD:sBOD (increase in proportion of soluble BOD) will be a potential impact on the downstream secondary biological treatment process. Substrate (BOD) that is available in a soluble form may be easier to biodegrade than substrate bound up with solid matter. This provides for a possible benefit; effluent from the ABR may be more amenable to biological treatment compared to effluent from normal primary tanks because more of the BOD is in solution. This will be discussed in more detail in section 6.2.4

Table 6.5 Ratio of influent COD and BOD concentrations

	COD:tBOD	tBOD:sBOD	COD:sBOD
Phase 2	2.39	2.64	6.32
Phase 3	2.25	2.40	5.39
Phase 4	2.21	3.22	7.12

Table 6.6 Ratio of effluent COD and BOD concentrations

	COD:tBOD	tBOD:sBOD	COD:sBOD
Phase 2	2.26	1.86	4.20
Phase 3	2.40	1.52	3.66
Phase 4	2.38	1.66	3.95

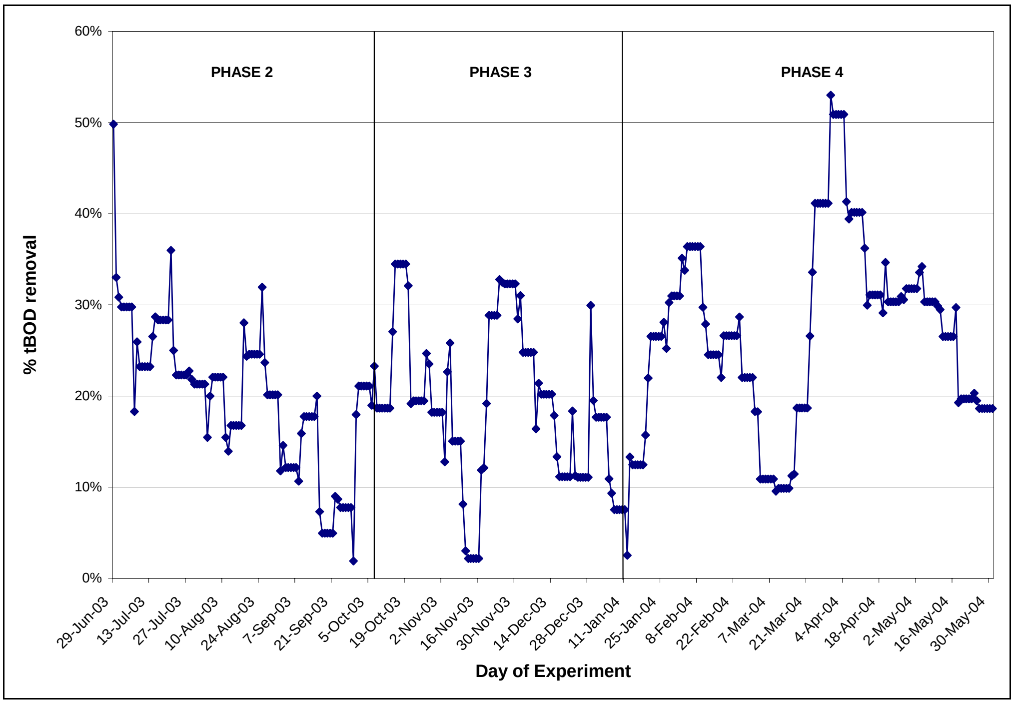


Figure 6.8 Mean total BOD removal – 7 day running mean

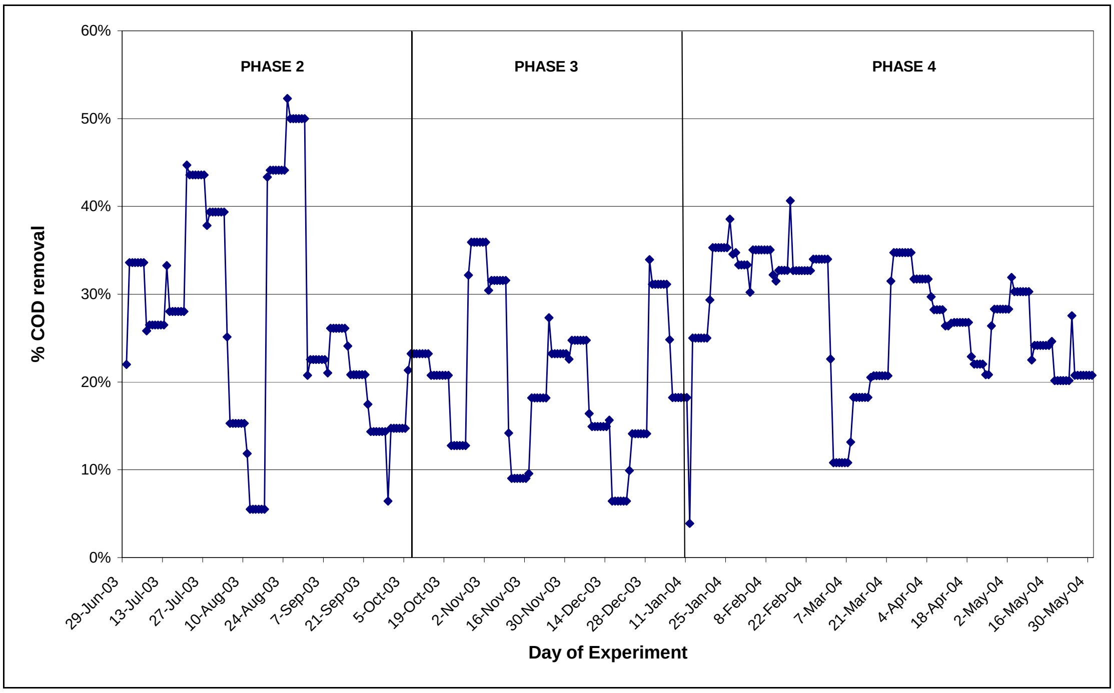


Fig 6.9 Mean COD removal across ABR – 7 day running mean

6.2.4 Comparison of ABR and primary tank effluent quality

The addition of polymer and ferric to the ABR influent in order to compare performance against full-scale primary tanks provided a number of challenges. The effluent quality and removal rates achieved by the ABR and OCSD A-side primaries are shown in Tables 6.7 and 6.8 below. There are a number of points to note:

- BOD, COD and TSS concentrations were higher for the ABR effluent than for the full-scale A-side primary effluent
- Effluent ammonia concentrations were the same and the ABR had no impact on ammonia concentration (average influent concentration was 26 mg/l)
- Polymer dosing was not optimized due to problems with the on-site polymer mixing flow meter

Table 6.7 Hydraulic Retention Times and effluent quality from the ABR and A-side primary basins

Phase	ABR					A-side primaries				
	HRT hours	tBOD mg/l	COD mg/l	TSS mg/l	NH_3 mg/l	HRT hours	tBOD mg/l	COD mg/l	TSS mg/l	NH_3 mg/l
2	3.8	158	356	112	24.7	2.9	100	257*	73*	23.9
3	3.9	155	372	107	24.4	2.7	102	257*	72*	25.7
4a	4.4	145	357	88	25.6	3.0	112	284	76	27.0
4b	3.1	164	377	117	25.2	2.8	111	272	77	27.0
4c	3.1	152	380	110	26.3	3.0	102	268	67	25.0
4d	3.2	160	350	105	26.3	3.0	114	268	66	27.0

[* data appear to be in error]

The removal rates in table 6.8 show that BOD removal and TSS removal was lower for the ABR than for primary basins. The full-scale primary data was collected from monthly MSO reports and shows some inconsistencies between 2003 and 2004 results. There is a discontinuity between the data for phase 3 and phase 4a where a considerable drop in primary BOD and TSS removal occurs. An initial review of the data suggested there was a fault with 2003 removal calculations but further analysis will be required to confirm this fact.

Table 6.8 Hydraulic Retention Times and removal rates by the ABR and A-side primary basins

Phase	ABR			A-side primaries		
	HRT hours	BOD removal %	TSS removal %	HRT hours	BOD removal %	TSS removal %
2	3.8	20	52	2.9	72*	86*
3	3.9	20	53	2.7	70*	84*
4a	4.4	29	68	3.0	48	69
4b	3.1	23	56	2.8	47	70
4c	3.1	25	57	3.0	50	72
4d	3.2	25	64	3.0	45	72

[* data appear to be in error]

The difference in performance between the A-side primary tanks and the ABR may appear to be a cause for concern but a number of factors need to be taken into consideration:

- It is not possible to compare directly the performance of a pilot-scale ABR with a full-scale primary tank. Only a full-scale retrofit with a control, or a pilot ABR with a pilot primary tank would provide a definitive comparison
- The polymer dosing and polymer mixing for the ABR during phase 4 was not optimized and is unlikely to represent the improvements that could be delivered at full-scale.
- The sludge blanket depth in the pilot plant was substantially less than would be used in a full-scale plant. This results in a reduced depth of sludge to capture fine solids and to generate methane gas.

However, it was noted in section 6.2.3 that BOD had been solubilized. This in turn would explain the relatively poor BOD removal despite TSS removal rates exceeding 60% and a lack of correlation between BOD and TSS removal. The following section that presents the results of the solids mass balance will explore this issue in more detail.

It is worth noting that the reduced depth of sludge will in turn reduce the mass of methanogens available to convert VFAs into methane. Operating sludge blankets at full-scale should enable a greater depth of sludge to be maintained with an increased quantity of methanogenic micro-organisms and therefore the potential exists to convert a greater quantity of VFAs to methane. The potential for methane gas to become entrained with the sludge blanket and cause solids to be washed out of the ABR is a parameter that can be monitored through analysis of solids removal. However, the results for TSS removal during the different phases of the trial indicate that the upflow velocity and the addition of polymer will have a greater impact on the performance of the ABR.

6.2.5 Solids digestion achieved by ABR pilot plant

Overview of section

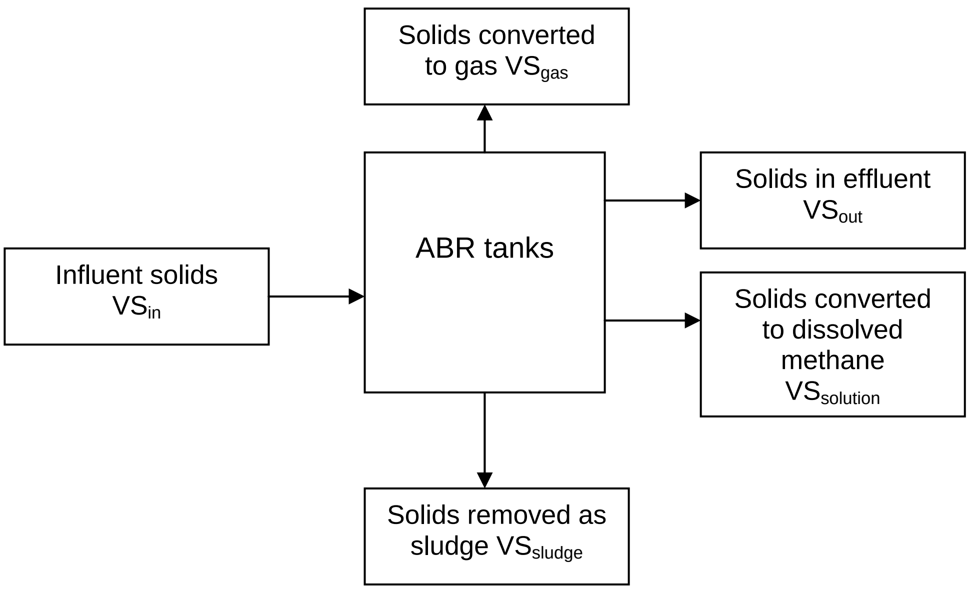
This section presents data for the volatile solids digestion across the ABR and will be divided into a number of components. First there will be a brief overview of the initial mass balance model used to determine the solids destruction achieved by the ABR. Two equations will be presented to compute the level of volatile solids destruction. The input and outputs components for the model will be discussed highlighting some of the challenges with acquiring robust data and methods of estimation in the absence of good data.

Sensitivity analysis around the concentration of methane in the final effluent forms an important part of the solids mass balance and detailed results are presented. In light of the initial mass balance results, a modified model is presented to take into account the possible conversion of volatile solids to soluble BOD. The section concludes with a summary of the findings and areas requiring further exploration.

Methods to determine volatile solids mass balance across ABR

There was a single input of volatile solids to the ABR from the influent flow, and a number of possible outputs for the volatile solids, or the products of volatile solid digestion. If the five ABR tanks are treated as a single system for the purposes of the analysis, a simple overview of the volatile inputs and outputs represented in figure 6.10 .

Fig 6.10 Proposed input and output of solids across ABR



Notation

VS_{in}	Mass of influent volatile solids
VS_{out}	Mass of volatile solids in wastewater effluent
VS_{sludge}	Mass of volatile solids removed during desludging
VS_{gas}	Mass of volatile solids converted to gas
$VS_{solution}$	Mass of volatile solids converted to methane in solution

The influent solids mass (VS_{in}) and effluent mass (VS_{out}) is the product of the flow and concentration of volatile solids. A refrigerated composite sampler at the influent and effluent combined with a flow meter and data-logger provided a reasonably accurate measurement of VS_{in} and VS_{out} .

The sludge removed (VS_{sludge}) is the sum of the mass of sludge removed from each tank during de-sludging. As discussed in the outline of Phase 3 operation, the tanks were modified to improve desludging and enable a more accurate sludge sample be taken during the desludging regime.

The measurement of VS_{gas} required the capture of gas generated by the digesting sludge and released from the top of the tanks. The gas was captured in an inverted calibrated cone filled with water. The gas displaced the water in the cone and the volume of gas could be measured over time. Samples were analysed to determine the ratio of specific gases.

The final component of the mass balance is represented by $VS_{solution}$, methane that has been dissolved in solution. A number of assumptions can be made to determine the mass of volatile solids converted to dissolved methane. A direct and reliable measure of the concentration of dissolved methane in the ABR effluent to determine $VS_{solution}$ was difficult to achieve as will be discussed later.

Two methods can be used to calculate the quantity of volatile solids that are destroyed by the ABR. The simple method does not need a measure of VS_{gas} or $VS_{solution}$ and assumes the following relationship, where $VS_{destroyed}$ is the mass of volatile solids destroyed by the digestion process:

$$VS_{destroyed} = (VS_{in} – (VS_{out} + VS_{sludge}) \quad \text{[Equation 1]}$$

This simplified mass balance assumes that the mass of volatile solids not contained in the effluent or in the sludge removed during desludging must have been destroyed. A more complicated mass balance attempts to account for the mass of solids being converted to gas and to dissolved methane, whereby:

$$VS_{destroyed} = (VS_{gas} + VS_{solution}) \quad \text{[Equation 2]}$$

$$VS_{gas} + VS_{solution} = (VS_{in} – (VS_{out} + VS_{sludge}) \quad \text{[Equation 3]}$$

$$(VS_{in} – (VS_{out} + VS_{sludge} + VS_{gas} + VS_{solution}) = 0 \quad \text{[Equation 4]}$$

Measuring VS_{gas}

The estimation of VS_{gas} was based on the quantity of gas captured by inverted cones, using the following assumptions and data:

Average flow to ABR	= 24,240 ft^3/day
Area of cone capturing gas	= 1.226 ft^2
Area of tank	= 140 ft^2
Ratio of cone to tank surface area	= 1:114
Conversion factor ft^3 :litres	= 1:28.3
Volume of methane per lb VS destroyed	= 9 ft^3 methane gas
Solubility of methane in clean water	= 3.3 ml/100ml

The volume of methane gas produced by the tanks during phase 3 and phase 4 is shown in Table 6.9. The gas volume was recorded in liters/hour and converted to ft^3 / day. One significant observation with phase 4 was the production of gas in tank 1. During phase 2 and 3 no gas bubbles were observed in tank 1 and therefore no measurements were taken for gas production. The addition of polymer caused more solids to be retained in tank 1 which in turn may have resulted in the digestion of volatile solids and gas production.

Table 6.9 Volume of gas released by ABR tanks

Phase	**Gas volume (ft^3/day)**				
	Tank 1	**Tank 2**	**Tank 3**	**Tank 4**	**Tank 5**
3	0	39	59	54	62
4	16	47	43	32	22

The methane gas collected in the cones was measured and found to be variable with a maximum near 80% by volume (table 6.10). There were a number of potential causes for the variation including:

- Different solubility of gases in wastewater
- Residual air in the cone prior to collection of gas
- Air leaking into the cone during gas collection
- Air leaking into the gas sampling bag during transfer from cone

It should be noted that the solubility of carbon dioxide in water is 90ml/100ml which may explain maximum readings of methane that exceeded 80% volume. Usually in anaerobic digestion methane is produced at 60 to 65% by volume and this would be the anticipated level in gas exiting the ABR.

Table 6.10 Volume of specific gases produced by ABR tanks

Gas (% volume in cone)	Tank							
	2		3		4		5	
	Avg	Max	Avg	Max	Avg	Max	Avg	Max
CH_4	47	78	57	81	55	84	50	82
CO_2	9	41	5	7	5	8	4	7
O_2	11	20	9	18	11	20	13	44
N_2	31	44	29	48	30	53	34	60

It has been assumed that the destruction of volatile solids in the ABR follows the same biochemical pathway as that experienced in anaerobic digestion resulting in about 60% methane by volume and for the purposes of the mass balance calculations this has been used as the default. The mass of volatile solids that would need to be converted to methane gas in order to generate the volumes produced by the ABR is presented in table 6.11. A methane gas production of 9 ft^3 for each lb of VS has been assumed. Methane gas that is dissolved in solution has been accounted for separately as described below.

Table 6.11 Estimated mass of volatile solids converted to produce methane gas

	Estimate mass of VS (lb) converted to methane and released as gas					
Phase	tank 1	tank 2	tank 3	tank 4	tank 5	Total
3	0	245	374	344	393	1356
4	104	301	270	203	141	1019

Challenges with measuring $VS_{solution}$

Review of literature indicates that methane has a solubility of 3 ml/100ml in clean water which equates to about 23 mg/l. Data for solubility of methane in primary effluent is not readily available and therefore it was decided to measure the dissolved methane concentration in samples taken from the ABR tanks.

The samples were sent to an external laboratory for analysis and the results are presented in table 6.12. An attempt was made to reach a saturation concentration of methane in ABR effluent by bubbling laboratory gas through a diffuse filter placed in a sample from tank 5.

Table 6.12 Measured concentration of methane in ABR wastewater

Date (2004)	Methane concentration (mg/l)			
	Influent	Tank 5 Effluent	Tank 4 Effluent	Saturated Tank 5 effluent
April 21st	0.08	1.3		
April 28th		0.89		
April 29th		0.35		12
May 5th		0.31		
May 11th			0.82	
May 13th			0.84	

The single data-point for the ABR influent indicates that there was a negligible quantity of dissolved methane, which would be as expected assuming that there was limited generation of methane in the sewer collection system. Based on results from Ellesmere Port it was anticipated that most of the methane would end up in solution, with very little released as gas. However methane concentrations in the effluent were lower than expected, and indicate the ABR effluent was not methane saturated. There are, however, a number of reasons to question the laboratory data for the concentration of methane in the ABR effluent:

- The sampling process could lead to methane being lost from solution
- There is an unexplained three-fold range in methane concentrations results when the flow to the ABR remained constant
- The effluent methane concentration from tank 4 is not consistent with the tank 5 effluent concentration (which should be higher)
- Analytically it is difficult to get accurate measurements of methane in solution

The lack of robust data for the concentration of methane in the ABR effluent provides a challenge when using equation 2 to determine a volatile solids mass balance. This will be discussed in more detail in the analysis of data.

Relationship between $VS_{solution}$ and $VS_{destruction}$

Given the uncertainty with the concentration of methane in solution, it was decided to produce a mass balance using a range of methane concentrations. A worked example for the calculation of the values in table 6.13 has been presented below and the values in table 6.13 will be used later in the mass balance analyses.

Table 6.13 Mass (lb) of VS needing conversion to achieve given concentration of dissolved methane in ABR effluent

	Assumed dissolved methane concentration (mg/l)				
	5	10	15	20	25
	Mass of VS (lb) requiring conversion				
Phase 3	1840	3680	5520	7360	9200
Phase 4	2068	4136	6204	8272	10340

Worked example to show how the values for Table 6.13 were calculated:

Phase 3 average flow = 8.07 liters/sec
Days run = 95
Daily flow = 697,200 liters

At 5 mg/l dissolved methane;
Mass of methane in effluent = (697,200 x 95 x 5)/1,000,000
= 331 kg

Conversion of volatile solids to dissolved methane

Assume 1 lb of VS converts to 9 ft^3 methane
1 mole methane = 16 grams and occupies 22.4 liters at STP
1 ft^3 gas = 28.3 liters
Therefore 16 grams methane = 22.4/28.3
= 0.79 ft^3

1 lb volatile solids = (16)(9)/0.79
= 180 g methane

331 kg methane = 331/0.180
= 1839 lb volatile solids

Therefore to obtain 331 kg methane in the ABR effluent (equating to an average of 5 mg/l methane during the whole of phase 3) would in theory require the destruction of 1839 lbs of volatile solids assuming all the methane produced ended up in solution.

Calculating mass balances based on the model in figure 6.10

So far the discussion has proposed two methods by which the volatile solids destruction can be calculated (equations 1 and 2). Data for measuring VS_{in} and VS_{out} was more robust and improvements were made to desludging in order to overcome shortfalls with the measurement of $VS_{sludge.}$ The challenges associated with measuring $VS_{solution}$ and VS_{gas} have been discussed.

The estimated destruction of solids for the ABR using equation 1 is presented in Table 6.14. It is not possible to estimate the %VS destruction for phase 2 because of the lack of good data for the mass of sludge removed during the automated desludging process. Addition of ferric and polymer in phase 4 was the most significant operational difference and may account in part for the elevated solids destruction. More solids are retained in the tanks and can therefore be exposed to the digestion process.

Table 6.14 Volatile solids destruction using simplified mass balance

Phase	VS_{in} (lb)	VS_{out} (lb)	VS_{sludge} (lb)	$\%VS_{destroyed}$
2	29700	13400	No data	No data
3	27200	12100	8900	22
4	53400	20000	12400	39

Table 6.15 presents the results of the volatile solids mass balance with the inclusion of the additional components for conversion of VS to dissolved methane ($VS_{solution}$) and conversion to methane released as gas (VS_{gas}). From the experimental work it was apparent that a number of potential values could be used for the concentration of methane in the ABR effluent. Table 6.15 assumes an effluent concentration of 10 mg/l methane. The mass of volatile solids destroyed is the sum of VS_{gas} and $VS_{solution}$ (see equation 2).

Table 6.15 Volatile solids destruction assuming 10 mg/l methane in effluent

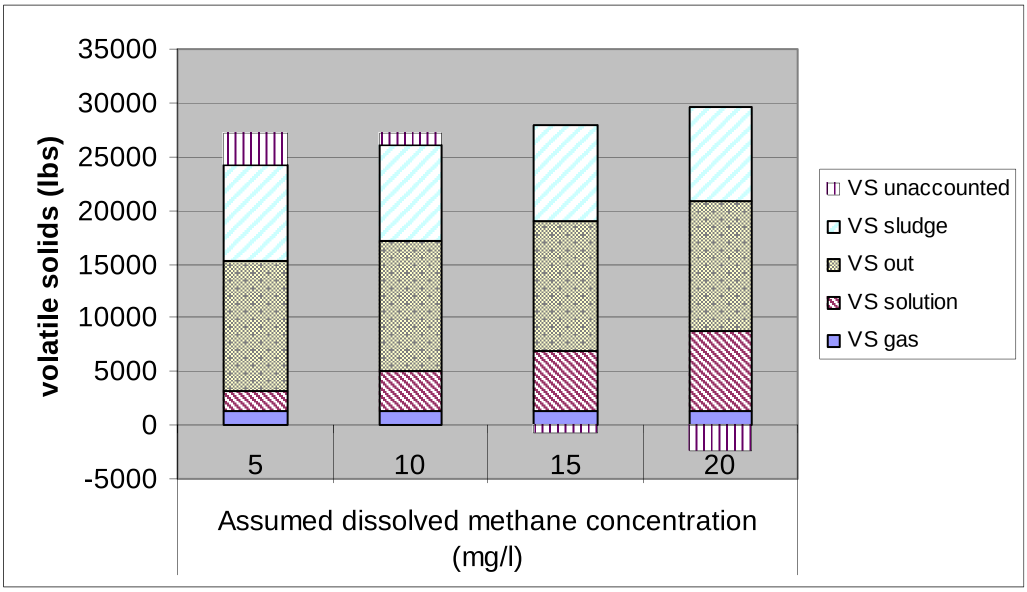
Phase	VS_{in} (lb)	VS_{out} (lb)	VS_{sludge} (lb)	VS_{gas} (lb)	$VS_{solution}$ (lb)	% VS destruction
2	29700	13400	No data			
3	27200	12100	8900	1320	3680	18
4	53400	20000	12400	1020	4136	10

The difference in the values for table 6.14 and table 6.15 requires further investigation. The first observation is that the volatile solids do not balance for table 6.15. Given the lack of certainty around the concentration of methane in the effluent, a mass balance sensitivity analysis was conducted for a range of methane concentrations.

Sensitivity Analysis for dissolved methane concentration

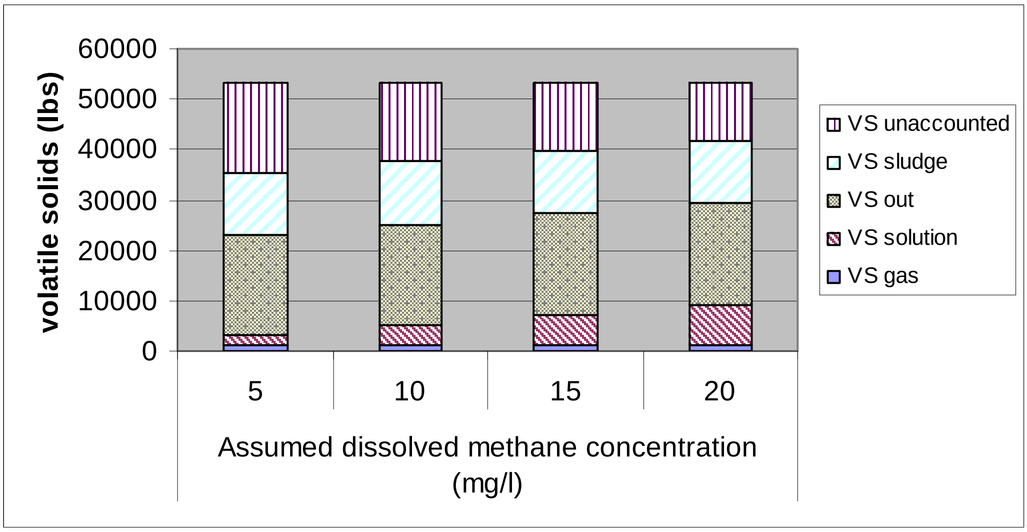
The results for phase 3 and phase 4 are presented in Fig 6.11 and Fig 6.12. The mass of VS exiting the ABR (in the various states) needs to balance the influent mass of VS. It is interesting to note that for phase 3 there is a very small mass of volatile solids that cannot be accounted for when the assumed $VS_{solution}$ is set at 10mg/l. At a methane concentration above 10 mg/l there is a negative mass balance which means the quantity of VS exiting is greater than that entering the ABR. This is a situation that cannot exist and therefore implies an upper limit of about 10 mg/l methane in solution.

Figure 6.11 Mass balance for volatile solids in phase 3



For phase 4 there was a significant mass of VS for which the model did not account even assuming a concentration of 20 mg/l of methane in solution. A number of factors may have led to the poor mass balance. The quantity of gas measured exiting the tanks may not have been accurate. This is unlikely as repeated samples were taken and the cones were moved to various points around the tank. The influent and effluent solids mass may have been inaccurate. However, composite samples were taken so the data is reasonably robust. It was recognized that the desludging system did not allow for good composite sampling leading to the modifications outlined earlier in the report and it is possible that the quantity of sludge removed was under-estimated even after modifications.

Figure 6.12 Mass balance for volatile solids in phase 4



Revised mass balance model to consider partial digestion of Volatile Solids

Two key findings highlight the need to revise the assumptions used to calculate the mass balance as presented thus far. A proportion of influent VS cannot be accounted for by the outputs presented in the mass balance model in figure 6.10. In addition, it has been noted that the soluble BOD in the effluent is greater than that in the influent. In light of these findings a revised mass balance model is proposed as presented in figure 6.13. It introduces one additional output component namely VS_{sBOD} to represent the mass of volatile solids that are solubilized but are not subsequently converted to methane. The chemical constituents that would produce an increase in soluble BOD concentration are likely to be complex but will include volatile fatty acids. VFA analysis across the ABR tanks is presented in figure 6.14. The effluent contains VFAs at around 100 mg/l which will contribute to the soluble BOD in the effluent.

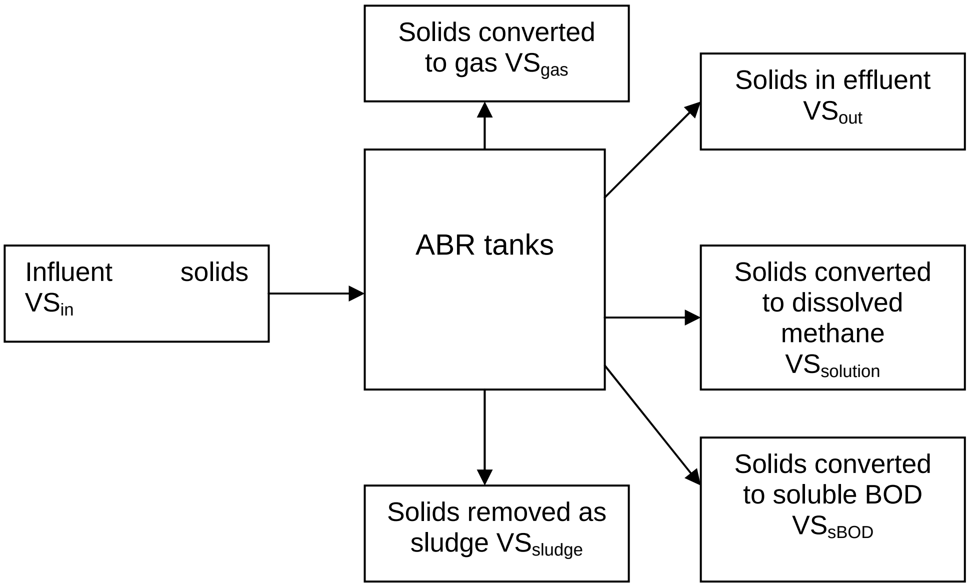


Figure 6.13 Revised mass balance model to account for soluble BOD

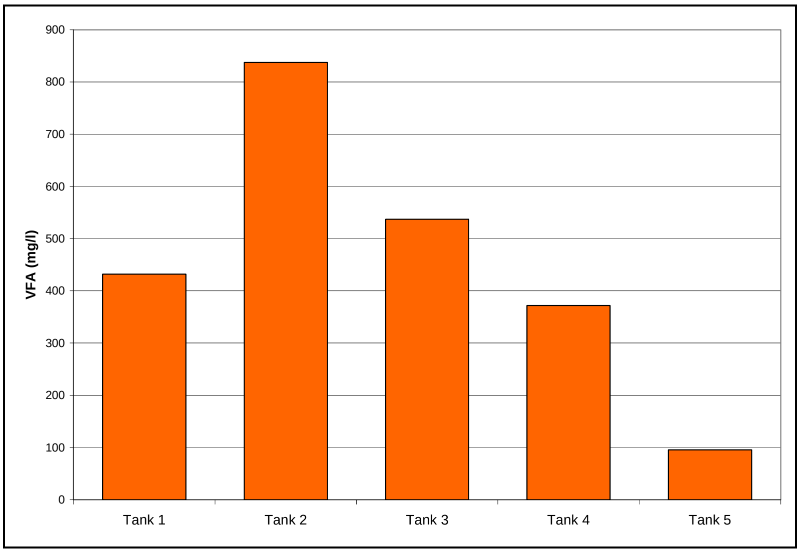


Figure 6.14 Mean sludge VFA values in each tank for phases 2 thru 4

Figure 6.15 and figure 6.16 present the concentration of soluble BOD required to create a mass balance for phase 3 and phase 4. The left hand y-axis is the mass of unaccounted for volatile solids ($VS_{unaccounted}$) for a range of dissolved methane concentrations (from Fig 6.11 and 6.12). The right hand y-axis is the concentration of soluble BOD in the effluent that would equate to the mass of 'unaccounted' volatile solids, thereby producing a mass balance. It assumes that 1 lb of VS would be converted to 1 lb of soluble BOD. Further work would be required to confirm the validity of this assumption.

Figure 6.15 Phase 3: Concentration of sBOD that equates to unaccounted VS

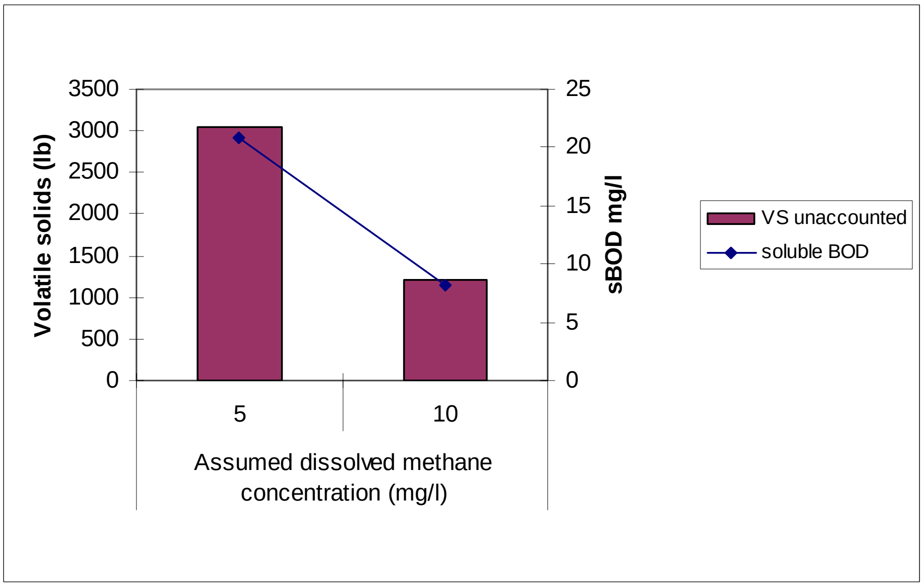
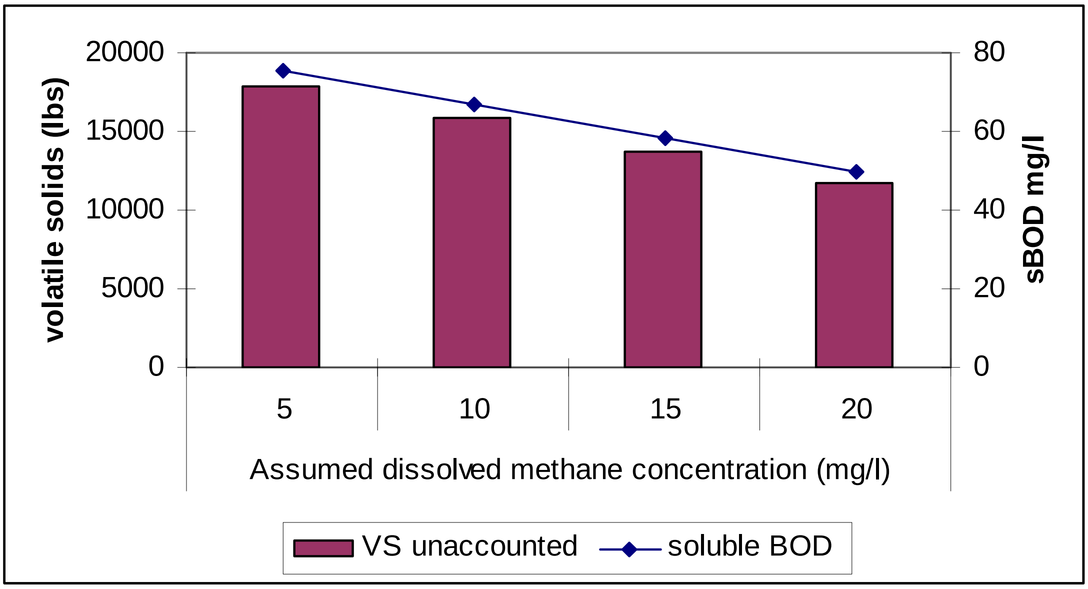


Figure 6.16 Phase 4: Concentration of sBOD that equates to unaccounted VS



The conversion of $VS_{unaccounted}$ to soluble BOD in phase 3 is consistent with the data for effluent soluble BOD concentrations (Fig 6.17). During phase 3 the average influent sBOD was 88 mg/l and effluent sBOD was 104 mg/l representing an increase of 10 mg/l In phase 4 the influent sBOD averaged 71 mg/l and the effluent averaged 92 mg/l, an increase of 21 mg/l.

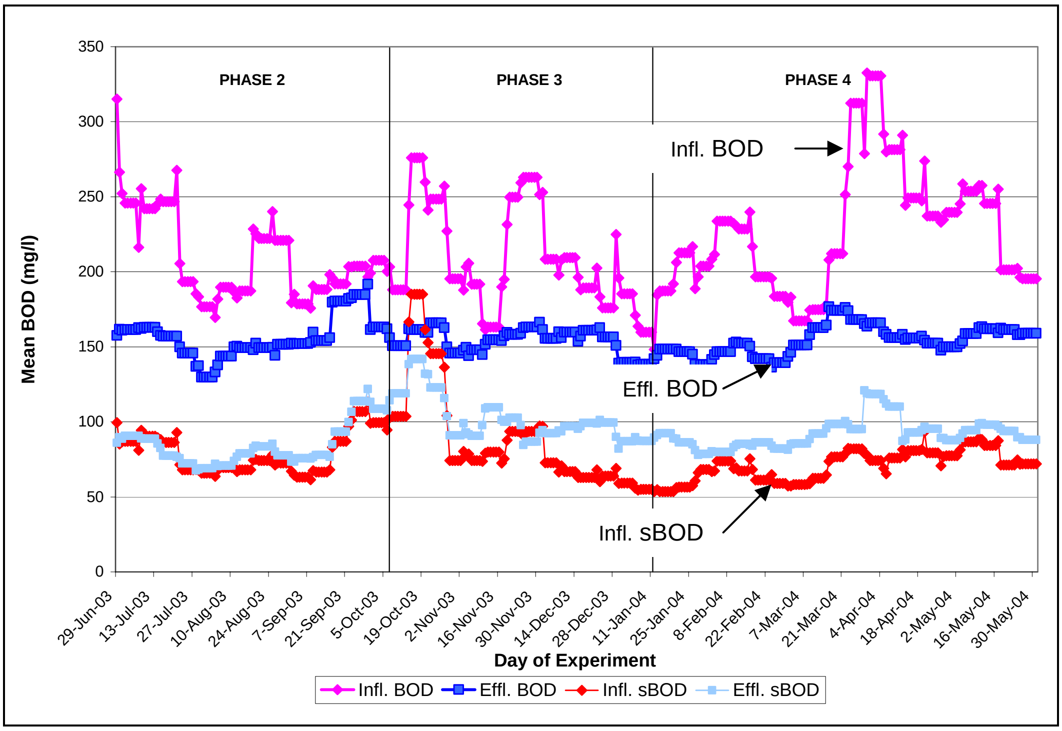
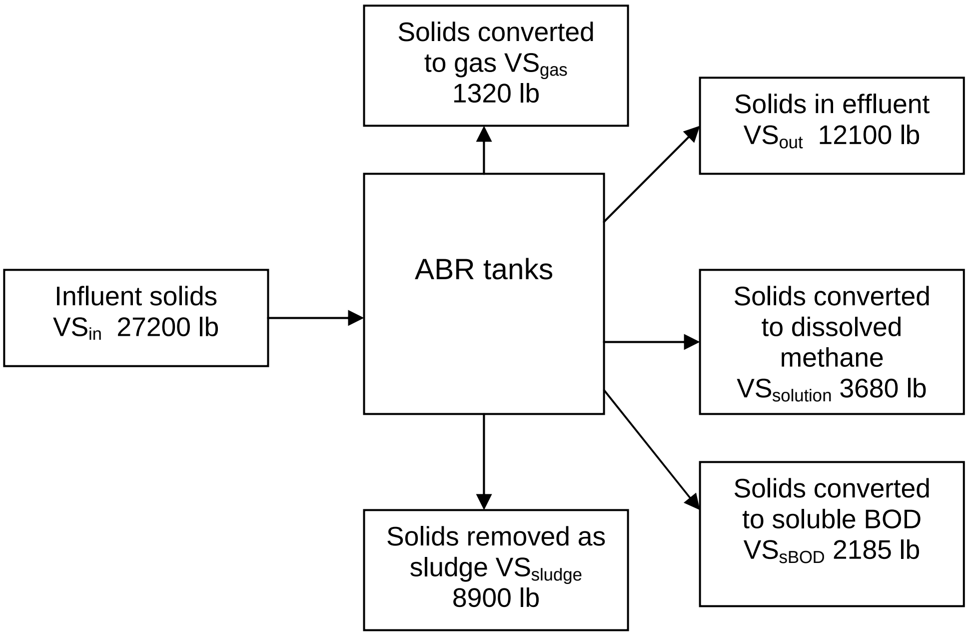


Figure 6.17 Mean influent and effluent BOD – 7 day running mean

Based on the data discussed, figure 6.18 presents a mass balance for phase 3 assuming the methane concentration in the effluent was 10 mg/l and the soluble BOD increased by 15 mg/l. Using this data the inputs and outputs to the system balance to within about 1000 lb, which is less than 4%.

Figure 6.18 Phase 3 mass balance



It can be seen from figure 6.16 that the required increase in soluble BOD in phase 4 to produce a mass balance will be over 60 mg/l. The increase recorded was only 21 mg/l. It is possible that the earlier assumption about 1 lb of VS converting to 1 lb of BOD is incorrect and solubilization of more than 1 lb of VS is required to generate the equivalent of 1 lb BOD.

Summary of mass balance results

Table 6.19 presents a summary of the mass balance results. There are a number of key points to note that would require further exploration. The mass balance is reasonably accurate for phase 3 but there is still a significant component of volatile solids in phase 4 that cannot be accounted for. The 16% of solids which do not balance may have been solubilized, but not in the ratio of 1 lb VS: 1 lb BOD as has been assumed.

Table 6.16 VS destruction assuming 10 mg/l dissolved methane (upper limit)

Phase	VS_{in} (lb)	VS_{out} (lb)	VS_{sludge} (lb)	VS_{gas} (lb)	$VS_{solution}$ (lb)	VS_{sBOD} (lb)	% VS destruction	%VS unaccounted
3	27200	12100	8900	1320	3680	2185	26	4%
4	53400	20000	12400	1020	4136	7085	23	16%

Phase 4 also saw a shift in methane production as more solids settled out in tank 1 during polymer addition. The changing dynamics of the ABR tanks in terms of the impact on VFA production was not captured for phase 4.

In summary the lower limit for VS conversion is 23% in phase 4 with 16% $VS_{unaccounted}$. The upper limit as calculated using the simplified mass balance model is 39% which assumes conversion of all $VS_{unaccounted}$. The ability to convert chemical intermediates to methane and thereby reduce BOD load to the secondary aeration tanks will be an important consideration in an ABR retrofit. This will be of particular interest if there are design limits on the BOD load into the aeration tanks.

The pilot system is not optimized to deliver the methanogenic phase of digestion due to the reduced sludge depth required to retain solids in the shallow pilot-scale tanks. At full scale it is possible to operate with deeper blankets to retain a greater mass of methanogens within the system and enhance conversion of VFAs to methane.

6.2.6 Quantification of risks associated with release of methane

The digestion process in the ABR produces methane gas that can be dissolved in the liquid phase or be released as gas from the surface of the tanks. It was necessary to understand where the methane gas may be released in the full scale treatment process in order to assess associated explosive risks. Methane can escape from solution for a number of reasons:

- The liquid phase is saturated and therefore any additional production of methane will not be dissolved
- Agitation or turbulence causes the release of gas when the liquid passes over a weir
- Aeration in the activated sludge plants displaces methane from solution

There are therefore two mechanisms of methane release from the ABR that need to be considered at full-scale:

1. Methane release directly from the ABR compartments as a gas.
2. Methane release from solution during turbulent conditions, specifically at the overflow weir of the primary tanks and in the activated sludge plants.

An increase in temperature could also result in dissolved methane coming out of solution but this has not been considered as there is not likely to be a significant temperature change across the treatment processes. This section summarizes the quantification of the risk associated with the release of methane, based on the potential solubility of methane in the ABR effluent and the quantity of methane released from the ABR pilot tanks (section 6.25).

Estimate of methane production from the ABR pilot plant

The volumetric production of methane by the ABR was derived from the mass balance and direct capture of gas. The ABR tanks had a surface area of about 700ft^3 and produced about 220 ft^3 of gas. The average flow to the ABR was ~ 686,000 litres/day or about 24,240 ft^3.

Based on the data gathered and the assumptions used for calculating the mass balance, a maximum methane concentration of 10 mg/l in the effluent from a full-scale retrofit ABR plant has been assumed. In order to assess the explosive risk associated with methane production, an upper limit of 40% volatile solids destruction was assumed.

Estimate of methane release from solution at the ABR pilot plant overflow

The pilot ABR overflowed into an effluent holding tank (see Fig xx). The water drop was between 1 and 2 inches, creating turbulent conditions. Details of the experiment to assess methane release from the ABR effluent tank are included in Appendix B. It was found that it took approximately 30 minutes for methane levels to build up to a steady state of 12% by volume in a headspace of 18.7 ft^3.

At full scale the overflow arrangement would differ considerably from the pilot plant. Therefore the data obtained in this investigation can only be used to give an indication of the potential release of methane from solution at full scale as described below.

Full scale scenarios investigated

Analyses were performed looking at a range of scenarios for the primary tanks and activated sludge process in view of retrofitting ABR to the existing primary basins on both plants 1 and 2. Data was extracted from engineering drawings where available. The scenarios studied are identified in Table 6.20.

Table 6.20 Range of scenarios

	Primary tanks	**Activated sludge process**
Plant 1	Circular and rectangular clarifiers	Diffuse aeration
Plant 2	Circular clarifiers	Pure oxygen aeration

The results of the analysis are detailed in Appendix B and summaries for plant 1 and plant 2 are presented in Table 6.21 and 6.22. There are two potential areas of risk. Firstly, the headspace above the overflow weirs of the circular clarifiers on plant 2 is separated from the main headspace. Therefore methane released from solution as the primary effluent passes over the weir could collect in the small headspace volume. However the air extraction system for the circular tanks removes air from the headspace above the weir as well as air from the main headspace above the primary tanks. It was not possible to assess the separate extraction rates from the weir headspace and the main headspace, and the precise dynamics of methane release from a full scale overflow weir was also unknown.

Table 6.21: Plant 1 summary of potential methane build-up

	Plant 1 primary tanks	
	Circular tanks	**Rectangular**
Extraction rate (ft^3/min)	800	6000
Headspace volume (ft^3)	210,000	16000
Time to reach 5% methane (minutes)*	440	1800
Time to extract headspace volume	26 minutes	2 minutes

*without ventilation

Table 6.22: Plant 2 summary of potential methane build-up

	Plant 2 circular primary tanks	
	Weir headspace	**Total headspace**
Extraction rate (ft^3/min)	Unknown	8000
Headspace volume (ft^3)	2700	210000
Time to reach 5% methane (minutes)*	7	500
Time to extract headspace volume	Unknown	26 minutes

*without ventilation

Rectangular clarifiers have a considerably smaller overflow weir with less surface area for methane release. This is offset by an increase in the flow speeds at the weir enhancing the turbulence and therefore possibly increasing methane release.

The second area of risk relates to the activated sludge plant and a summary is presented in Table 6.23. As with the overflow weirs it was not possible to quantify the rate of methane stripping associated with the different aeration systems therefore a worse case scenario was used. It was assumed that the primary effluent had a methane concentration of 10 mg/l when it reached the activated sludge plant and all the methane was removed from solution by the aeration system. The time taken to reach an explosive limit assumes no removal of air from the headspace, a scenario that is unlikely to exist because the air and oxygen used for aeration will have to displace headspace gas otherwise there will be a pressure build-up.

Table 6.23 Potential for methane build-up in activated sludge plant

	Plant 1	**Plant 2**
Flow (MGD)	70	65
Volume methane released (ft^3/day)	140,000	130,000
Headspace volume (ft^3)	640,000	285,000
Time to reach 5% methane (min)	330	160

Summary of methane build-up risk

In normal day-to-day operation with the abstraction system operational for the primary tanks, there will be no significant risk of the methane concentrations reaching an explosive level of 5% by volume in the headspace. The only scenario that could result in an explosive concentration would be the release of all dissolved methane (at 10 mg/l) into the confined headspace above the overflow weirs of the circular primary tanks. The total air extraction rate for the circular clarifiers is known, but not the split between extraction from the headspace above the primary tank and the headspace above the overflow weir. This information would be required to better quantity the possible rate of build of methane.

In the event that methane with a concentration of 10mg/l reached the activated sludge plant and was completely stripped from solution, the time taken to reach an explosive limit in plant 1 would be 330 minutes. However, the volume of air introduced by the aeration system will replace the headspace volume in about 20 minutes. Data for the volume of oxygen used in plant 2 was not obtained therefore the rate at which headspace volume could be replaced has not been calculated.

6.2.7 Costs and benefits of ABR installation

Before the ABR pilot-plant project was commissioned, an earlier desk-top study had been conducted by Atkins Water and MWH. The desk study investigated four scenarios based on the cost to retrofit ABRs to existing primary tanks and the potential operational savings associated with an ABR installation. The four scenarios were:

- Lower cost estimate for retrofit of ABR to plant 1 with minimum savings
- Upper cost estimate for retrofit of ABR to plant 1 with maximum savings
- Lower cost estimate for retrofit of ABR to plant 2 with minimum savings
- Upper cost estimate for retrofit of ABR to plant 2 with maximum savings

The minimum and maximum operational savings were based on the estimated difference between operational costs with a retrofitted ABR and operational costs without ABR. Operational costs for the following components were included:

- Aeration in activated sludge plant
- Sludge thickening
- Digestion
- Post digestion dewatering

The original desk-study for plant 1 and plant 2 was revised to take into account standard estimating assumptions used at Orange County Sanitation District for projects still at the conceptual design phase. The results are presented in Table 6.24. It should be noted that the figures in parentheses represent a net present saving. The net present cost was based on a 20 year period with a discount rate of 5%. A more detailed summary is provided in Appendix A.

Table 6.24 Summary of estimate net present costs for Plant 1 and Plant 2

Scenario	Net Present Cost (saving)	
	Plant 1	Plant 2
Low cost retrofit with low operational savings	($2.5 million)	$5.2 million
Low cost retrofit with high operational savings	($9 million)	($4.6 million)
High cost retrofit with low operational savings	$11 million	$13.5 million
High cost retrofit with high operational savings	$4.4 million	$3.7 million

A separate exercise was conducted by CH2MHill to estimate the total present worth for solids handling with and without ABR, using the biosolids model developed for capital planning purposes. The model explored 4 scenarios:

- Capital and O&M costs at Plant 1 for solids handling (without ABR)
- Capital and O&M costs at Plant 1 for solids handling (with ABR)
- Capital and O&M costs at Plant 2 for solids handling (without ABR)
- Capital and O&M costs at Plant 2 for solids handling (with ABR)

The costs for solids handling relate to all processes downstream of the primary tanks. The assumptions for the quality of the effluent from the primary tanks are shown in table 6.25 assuming ferric and polymer dosing of the primary influent:

Table 6.25 Assumed primary tank effluent quality for biosolids model

	Total BOD (mg/l)	TSS (mg/l)
Plant 1 primary effluent	130	65
Plant 2 primary effluent	145	80

The processes that were considered in the capital investment scenarios are set out in table 6.26. In each scenario it is necessary to build additional process capacity or add new processes

Table 6.26 Type of new process capacity required for biosolids handling

	Plant 1 New Process capacity	**Plant 2 New Process capacity**
Baseline scenario	Dissolved air flotation thickeners (DAFT) Mesophilic digestion Belt filter press (BFP)	Dissolved air flotation thickeners (DAFT) Mesophilic digestion Belt filter press (BFP)
Planning Option 1	Primary settlement tank centrifuges Gravity belt thickeners Dewatering centrifuges Ultrasound	Dewatering centrifuges Ultrasound
Planning Option 2	Primary settlement tank centrifuges Gravity belt thickeners Dewatering centrifuges	Dewatering centrifuges

A summary of the total present worth for the three planning scenarios, with and without the addition of anaerobic baffled reactors is presented in table 6.27 for plant 1 and table 6.28 for plant 2. More detail relating to the capital and O&M costs can be found in Appendix A.

Table 6.27 Total Present Worth with and without ABR installation at Plant 1

	Total Present Worth	
	Without ABR	**With ABR**
Baseline	$497,000,000	$427,000,000
Scenario 1	$336,000,000	$301,000,000
Scenario 2	$347,000,000	$308,000,000

Table 6.28 Total Present Worth with and without ABR installation at Plant 2

	Total Present Worth	
	Without ABR	**With ABR**
Baseline	$276,000,000	$241,000,000
Scenario 1	$241,000,000	$205,000,000
Scenario 2	$242,000,000	$207,000,000

The difference in the total present worth with and without ABR represent the whole life cost savings that can be achieved by the installation of ABRs in the primary tanks. It should be noted that the capital planning model does not take into account the impact of ABRs on the liquid stream and the associated O&M costs to process the liquid stream.

Summary of costs and benefits

Desk-top exercises have been performed to help understand the costs and potential savings associated with retrofitting ABRs to primary settlement tanks. Table 6.29 summarizes the outputs from the biosolids masterplan model.

Table 6.29 Summary of Total Present Worth for ABR installation

	Total Present Cost (saving)	
	Plant 1 with ABR	**Plant 2 with ABR**
Baseline	($70 million)	($35 million)
Scenario 1	($35 million)	($36 million)
Scenario 2	($39 million)	($35 million)

When considering the biosolids processing costs, installation of ABR has been estimated to produce whole-life cost savings of between $35 million and $70 million. Further work is needed to quantify the impact that the ABR will have on the costs associated with treating the liquid stream.

6.2.8 Assessment of ABR sludge properties

Following analysis and discussion of the potential for full scale ABR retrofit to primary clarifiers at OCSD questions were raised about the properties of the ABR sludge and requirements for further work were identified as follows:

- Assessment of the impact of ABR operation on gas generation during anaerobic digestion and therefore the potential impact on co-gen revenue.
- Assessment of the potential impact of the ABR on sludge thickening costs pre- and post-digestion

The ABR sludge properties were investigated through a series of bench-scale digestibility and dewaterability tests. Appendix B details the methodologies used, results obtained, and conclusions drawn from the testing.

Following the comparison of ABR sludge with primary sludge the following conclusions have been made:

- There was no significant difference in the volume of gas produced per lb of volatile solids fed to a lab-scale batch digester for primary and ABR sludge.
- More gas was produced for each lb of volatile solids destroyed as one progresses from tank 1 to tank 5.
- Overall there was no significant difference in the digestibility of primary sludge and ABR sludge and no reduction in gas yield.
- There was no significant difference in the pre-digestion dewaterability of primary sludge and ABR sludge (for a composite sample of ABR sludge).
- There was no significant difference in the post-digestion dewaterability of primary sludge and ABR sludge (for a composite sample of ABR sludge).

7. CONCLUSIONS AND RECOMMENDATIONS

The information evaluated as part of this report accounts for 15 months ABR operation at Orange County Sanitation District (Feb 2003 to May 2004). The main objective of the work was to demonstrate the economic, practical and technical feasibility of an ABR for treating wastewater and reducing primary sludge production.

Conditions for ABR operation at OCSD in Southern California were different from previous plants operating in England and Northern Ireland. More specifically, the wastewater was warmer and more dilute, and the primary clarifiers ran on shorter HRTs.

Initial evaluations carried out in 2002 by MWH, with assistance from Atkins, indicated that ABR implementation could offset future capital construction costs and instead provide a net present saving to OCSD.

A 5 compartment pilot plant was therefore constructed and operated at OCSD's Plant 2, treating flows up to 0.3 MGD. There were four distinct phases of operation. During phase 1, the ABR was under a period of commissioning, stabilization and optimization. The data collected in phase 2 indicated that the sludge blanket depth was too high and optimal operation instead required a lower blanket depth. Modifications were made to the desludging system during phase 3 so that a robust mass balance could be obtained. Phase 4 evaluated ABR operation with polymer addition so that a more direct comparison could be made between ABR and conventional primary treatment. There was also a period of laboratory work which compared the digestibility and dewaterability properties of ABR sludge with that of conventional primary sludge.

The ABR pilot-plant has demonstrated that the ABR can convert volatile solids to methane and the associated reduction in the solids load onto downstream processes could generate net present savings. The net present savings are most likely to be realized at Plant 1 due to the reduced complexity and capital costs associated with a rectangular primary tank ABR retrofit.

The biochemical pathway to convert volatile solids to methane involves the production of intermediate compounds such as volatile fatty acids. Failure to convert all the intermediate compounds to methane can result in elevated concentrations of COD and soluble BOD in the ABR effluent. This would reduce the apparent BOD reduction in the ABR.

The potential impact of elevated concentrations of COD and BOD on future proposed down-stream secondary treatment processes will need to be assessed. It may have an impact on the generation of secondary sludge and the oxygen demand exerted during the treatment process.

A full-scale ABR retrofit has the potential to enhance the conversion of organic intermediates to methane because greater quantities of sludge can be retained in the ABR compartments given the increased depth of the full scale primary tanks and associated increase in sludge blanket depth.

In order to inform the decision about proceeding to a full-scale trial it is recommended that:

- A sensitivity analysis is conducted to determine the impact of primary tank BOD concentrations on downstream treatment costs. This is necessary because the ABR has the potential to increase primary tank effluent BOD concentrations if the intermediate chemical compounds are not converted to methane.
- A potential site for a full-scale ABR retrofit with a control at Plant 1 is identified and more detailed design is developed for an ABR retrofit. The detailed design will enable robust cost-estimates to be constructed.
- If the cost of retrofit and the sensitivity analysis indicate that there is still a net present saving to be made it would be recommended to proceed with a full scale demonstration.

The full-scale demonstration would be designed to:

- Confirm that an ABR retrofit can be operated satisfactorily and does not cause impairment of associated downstream process performance or operation.
- Provide a direct comparison with a control primary and thereby quantify any potential change in effluent quality.
- Confirm the volatile solids destruction that can be achieved and the extent of full conversion of volatile solids into methane gas.

APPENDIX A

THE POTENTIAL COSTS AND BENEFITS OF ABR RETROFIT

Net Present Costs for installation of ABR retrofits at Plant 1 and Plant 2

The assessment of costs and benefits (Table A1 and A2) are expanded versions of the summary tables presented in section 6.2.7. An initial report produced by MWH and Atkins Water (Anaerobic Baffled Reactor Evaluation, August 2002) was revised by Atkins Water to generate the figures in table A1.1, A1.2, A2.1 and A2.2. The lower capital investment requirements for Plant 1 and Plant 2 assumed that there would be minimal retrofit requirements for the ABR. The upper costs for capital investment assumed more significant investment requirements for the ABR and are therefore more conservative.

Lower and upper estimates for operational cost savings take into account the potential savings in digestion operational costs given the reduced solids load to the digesters. Operational savings are also assumed for sludge thickening, post-digestion dewatering and sludge disposal, given that there will be a reduced solids load to the digesters. The operational cost savings assume full secondary treatment and therefore may need to be reviewed depending on future projected secondary treatment capacity.

Assumptions for Table A1.1 and A2.1 – minimum operational savings

ABR volatile solids reduction	20%	
Imported energy cost	14	cents/kWh
On-site energy cost	6	cents/kWh
VS destruction in digesters	55%	
TWAS solids concentration	6.00%	

Assumptions for Table A1.2 and A2.2 – maximum operational savings

ABR volatile solids reduction	45%	
VS destruction in digesters	55%	
TWAS solids concentration	6.00%	
Imported energy cost	14	cents/kWh
On-site energy cost	6	cents/kWh

Table A1.1 Summary for net present cost (saving) for ABR conversion at Plant 1 with minimum operational savings

Item	Modifications Lower estimate	Upper estimate
Potential project cost	$2,750,000	$16,250,000
Annual differential operations costs (savings) minimum	($421,000)	($421,000)
Annual differential operations costs (savings) present worth (5% 20 years)	($5,250,000)	($5,250,000)
Net Present cost	($2,500,000)	$11,000,000

Table A1.2 Summary of net present cost (saving) for ABR conversion at Plant 1 with maximum operational savings

Item	Modifications Lower estimate	Upper estimate
Potential project cost	$2,747,697	$16,249,309
Annual differential operations costs (savings) maximum	($948,000)	($948,000)
Annual differential operations costs (savings) present worth (5% 20 years)	$11,814,175	$11,814,175
Net Present cost	($9,066,479)	($4,435,134)

Table A2.1 Summary for net present cost (saving) for ABR conversion at Plant 2 with minimum operational savings

Item	Modifications Lower estimate	Upper estimate
Potential project cost	$13,000,000	$21,250,000
Annual differential operations costs (savings) minimum	($625,000)	($625,000)
Annual differential operations costs (savings) present worth (5% 20 years)	($7,800,000)	($7,800,000)
Net Present cost	$5,200,000	$13,450,000

Table A2.2 Summary for net present cost (saving) for ABR conversion at Plant 2 with maximum operational savings

Item	Modifications Lower estimate	Upper estimate
Potential project cost	$13,000,000	$21,250,000
Annual differential operations costs (savings) maximum	($1,410,000)	($1,410,000)
Annual differential operations costs (savings) present worth (5% 20 years)	($17,600,000)	($17,600,000)
Net Present cost	($4,600,000)	$3,650,000

From the net present costs it is apparent that there is a wide range in estimates for the costs of retrofit to Plant 1 and Plant 2 and the potential savings. The retrofits to Plant 1 are less expensive because the tanks are rectangular in shape and easier to retrofit than the circular tanks at Plant 2. Further analysis around the potential impact of an ABR retrofit on the operational costs of the proposed secondary treatment system is recommended. Sensitivity analysis based on a range of effluent BOD concentrations would help establish a cross-over point at which the project would be deemed economically viable and worth pursuing.

APPENDIX B

Assessment of ABR sludge properties, dissolved methane concentrations and methane release from the ABR overflow

6. INTRODUCTION

Following analysis and discussion of the potential for full scale ABR retrofit to primary clarifiers at OCSD a number of issues arose and the requirements for further work were identified as follows:

- Assessment of the impact of ABR operation on gas generation during anaerobic digestion and therefore the potential impact on co-gen revenue.
- Assessment of the potential impact of the ABR on sludge thickening costs pre- and post-digestion
- Assessment of the robustness of the methane solubility assumptions made in the solids mass-balance calculations for the ABR pilot plant.
- Assessment of methane release from solution downstream of the ABR and the level of risk that needs to be managed.

The issues identified were investigated through a series of bench-scale digestibility, dewaterability and dissolved methane tests. This report details the methodologies used, results obtained, and conclusions drawn from the testing.

7. METHODOLOGIES

During each trial the performance of the ABR was compared to that of the A-side primary basins only. This is because ABR operation best represents current operation of the A-side basins in terms of ferric and polymer addition.

7.1 ABR Sludge Digestibility

The digestibility of ABR sludges was compared to that of plant 2 primary sludge through the use of bench scale batch-digesters. Sludge from the ABR Tanks 1-5 and plant 2 clarifiers was incubated and the gas produced by each sample collected. The trial was repeated twice in order to test the repeatability of the results and observe potential changes in ABR sludge digestibility under different ABR operating conditions. Trial A ran from 12th April – 26th April with ABR sludge from a 3 hr HRT and 0.2mg/l anionic polymer dose. Trial B ran from 26th April – 10th May with ABR sludge from a 3 hr HRT and 0.4 mg/l anionic polymer dose.

7.1.1 Sludge Sampling

Composite samples were taken from all the ABR tanks during de-sludging. Plant 2 primary sludge was taken from basin F which is fed from the A-side. Digested sludge was taken from digester F which is fed by primary basin F. Both locations were sampled as a composite over a period of 30 mins. Following collection, all sludges were stored in refrigerated conditions.

7.1.2 Laboratory Procedure

Two water baths were prepared with 6 x 2 litre glass flasks in each. 6 x 2 litre graduated flasks were arranged alongside each bath ready for gas collection. Both baths were set to 35 °C (95 °F) and the shaking platforms set at 60 RPM.

1000 ml of sludge from ABR tanks 1 – 5 and primary basin F was measured into each of the incubated flasks. For Trial A sludges were tested without any adjustment. This resulted in the primary sludge solids load to the digesters being about four times higher than that for the ABR sludge because of the higher concentration of solids in the primary sludge.

Therefore, for Trial B primary sludge was diluted so that the volatile solids load to the digesters was approximately the same for primary and ABR sludges. The primary digesters were run in duplicate for trial B.

The digesters were seeded with 250 ml of digested inoculum which was added to each incubated flask. 50 ml of the mixed solution was removed from each flask for pH, TS and VS analysis. Each flask was connected to the gas collectors and the time noted. The gas production was monitored at the start and end of each day and the gas collection flasks refilled with water once a day, or more often when required.

7.2 ABR Sludge Dewaterability

7.2.1 Sludge Sampling

All sludges tested were taken from the same composite samples used during Trial A digestibility testing.

7.2.2 Laboratory Procedure

Pre-digestion dewaterability

A composite sample of pre-digestion ABR sludges was produced according to the ratios of sludges removed from each tank during routine de-sludging over the previous 4 weeks operation. This constituted a ratio of 29:1:1:1:1 by volume for T1:T2:T3:T4:T5.

Preliminary investigations were performed on this solution in order to investigate the belt pressure and polymer dose needed to produce TS levels similar to those proposed for pre-digestion dewatering at plant 2. It was found to be difficult to produce a homogenous sample with TS levels < 10%. A belt pressure of 3 lbs and polymer dose of 0.4ml/500ml-sludge sample (equivalent to 0.5lbs/ton) was chosen to be the most suitable to produce homogenous sludge with low TS levels. Prior to testing, primary sludge was diluted to the same consistency as the ABR sludge to ensure that polymer addition was comparable. For each test the TS was taken before and after dewatering.

12 tests were performed on primary sludge and a further 12 tests on the ABR sludge solution.

Post-digestion dewaterability

Following the completion of gas production in Trial A, matching sludge was taken from each bath and mixed to produce a composite. ABR sludges from T1 – 5 were then mixed to the same ratio as in pre-digestion dewatering. Primary sludge was diluted to approximately the same consistency as the ABR solution, as before.

The number of tests, and the procedures used, were the same as for pre-digestion dewaterability, apart from the following points:

1. Due to a limitation on the volume of digested sludge available for testing the volume of sludge used in each test was reduced to 150 ml.
2. In order to simulate the post-digestion dewatering operations used at plant 2 the polymer dose was adjusted to the equivalent of 8lbs/ton which constituted an increase to 7ml/500ml sample (2ml/150ml sample).
3. In order to produce dewatered sludge with high TS levels the belt pressure was increased to 50lbs.

7.3 Methane Analysis

7.3.1 Methane solubility

Samples were taken from the ABR influent and effluent and analysed for methane concentration. As a comparison, a saturated sample was prepared in the laboratory by bubbling natural gas through ABR effluent for 6 hours. The sampling system at the exit of tank 5 caused the effluent to spray which in turn may have led to some loss of dissolved methane as the effluent came into contact with air. Additional samples were also taken from the Tank 4 effluent where it was thought there was less opportunity for methane loss through turbulence. All analyses were performed by Del Mar laboratories.

7.3.2 Methane Release

The effluent tank was covered with plywood and a sealing plastic sheet. An in-situ gas analyser was used to log methane concentrations. Evacuation of the headspace gas was performed daily during routine desludging when the tank was drained of all effluent.
A photograph of the tank covering is shown in Figure B1, below.

Figure B1 ABR pilot plant at effluent tank covered to assess build-up of methane gas

8. RESULTS AND DISCUSSION

8.1 ABR Sludge Digestibility

Average gas productions during Trial A, expressed as cuft/lbVS feed and cuft/lbVS destroyed, are shown in Figures B2 and B3.

Average gas productions during Trial B, expressed as cuft/lbVS feed and cuft/lbVS destroyed, are shown in Figures B4 and B5.

Figure B2 Average gas production in cuft/lbVS feed during Trial A digestibility testing

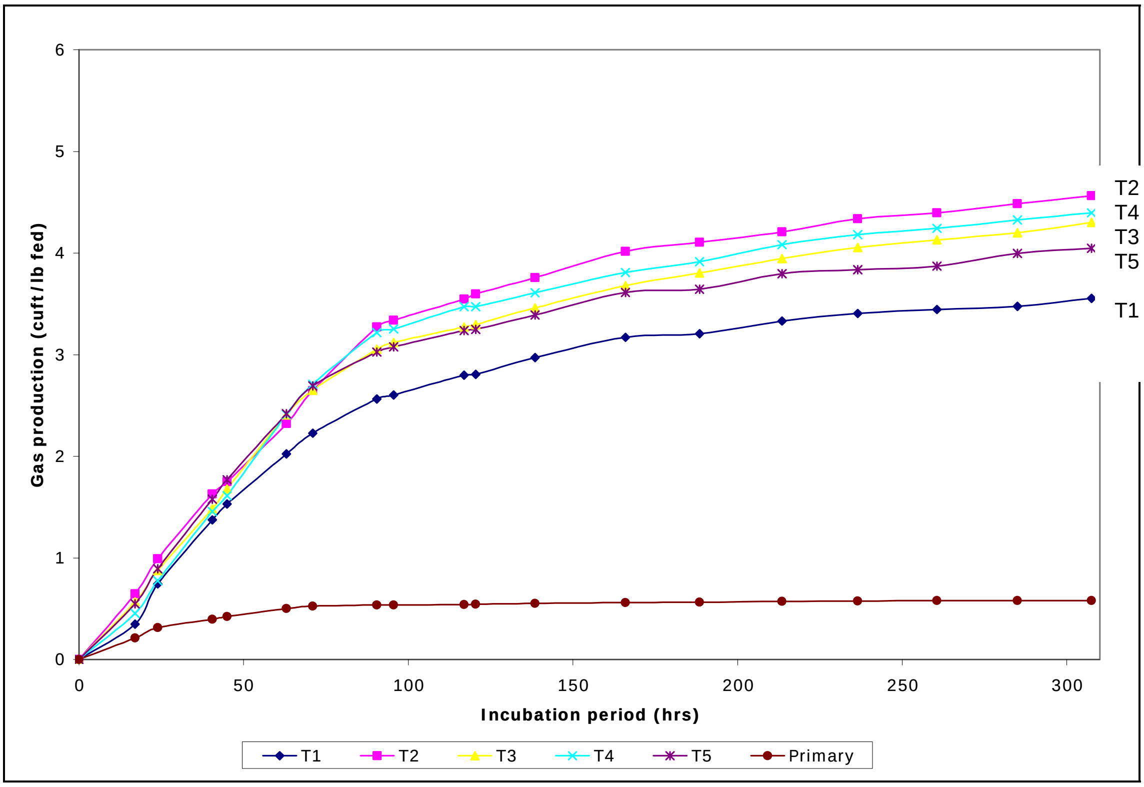


Figure B3 Average gas production in cuft/lbVS destroyed during Trial A digestibility testing

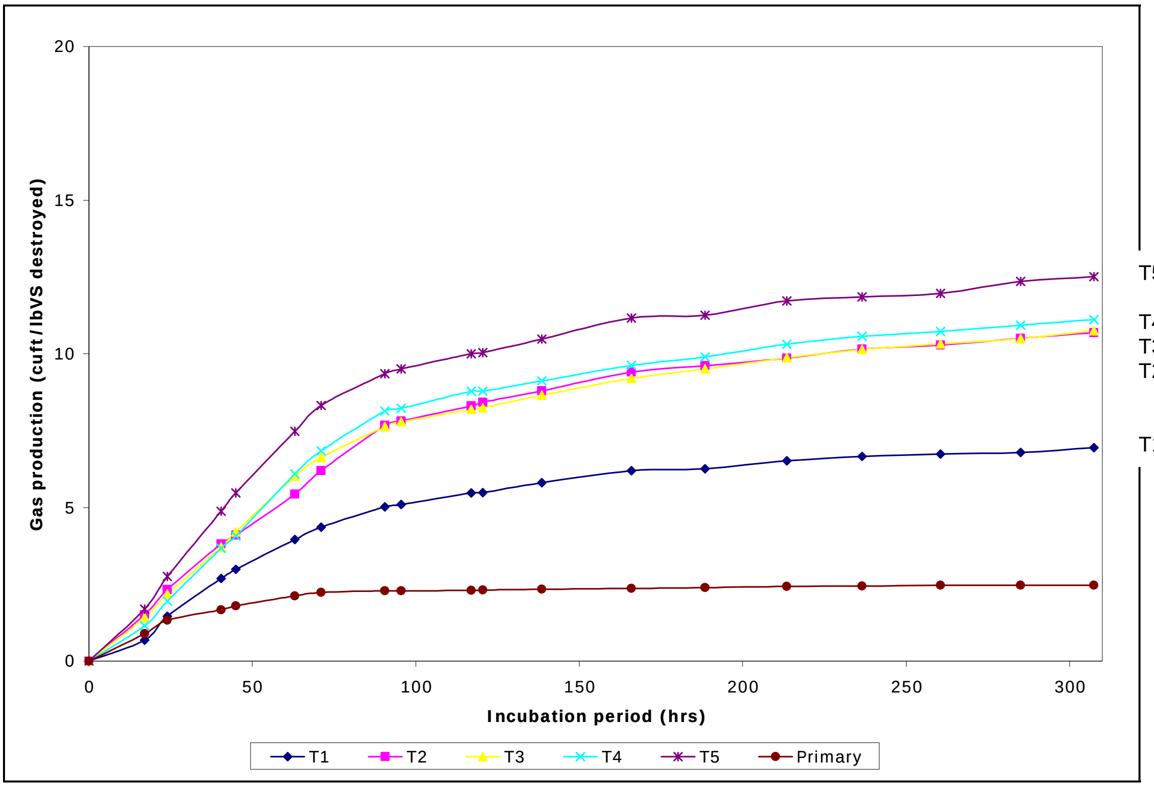


Figure B4 Average gas production in cuft/lbVS feed during Trial B digestibility testing

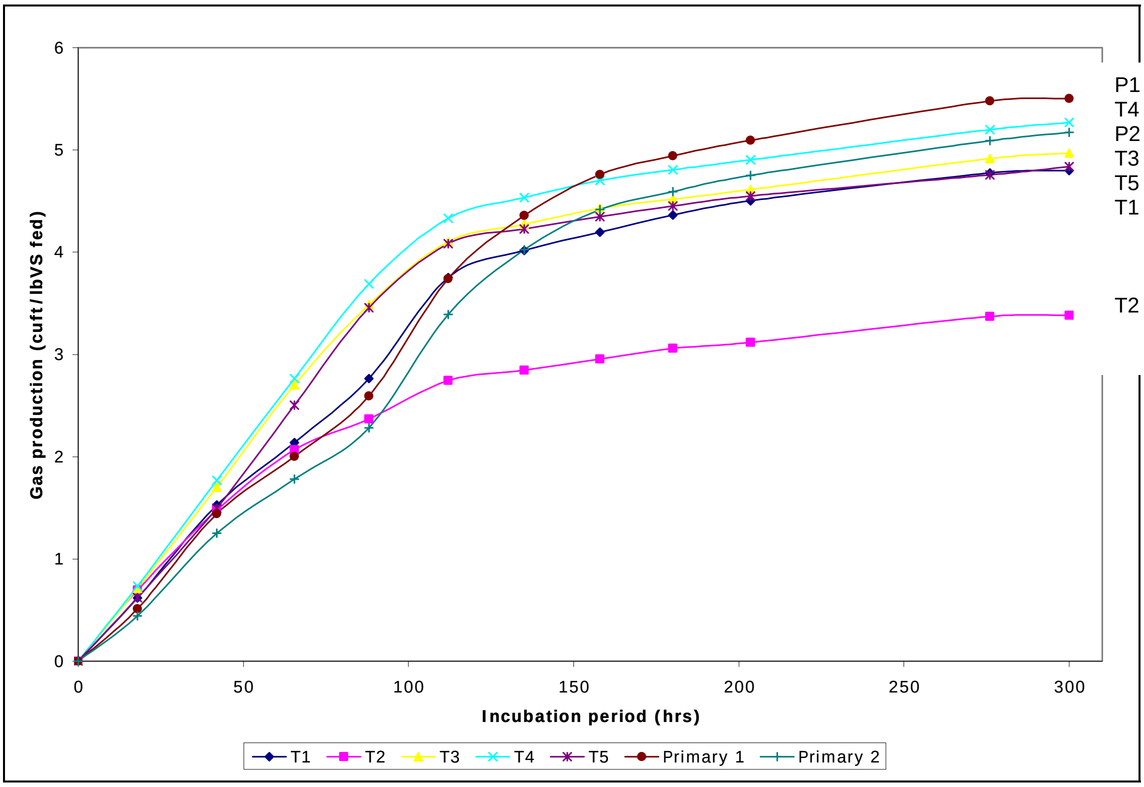
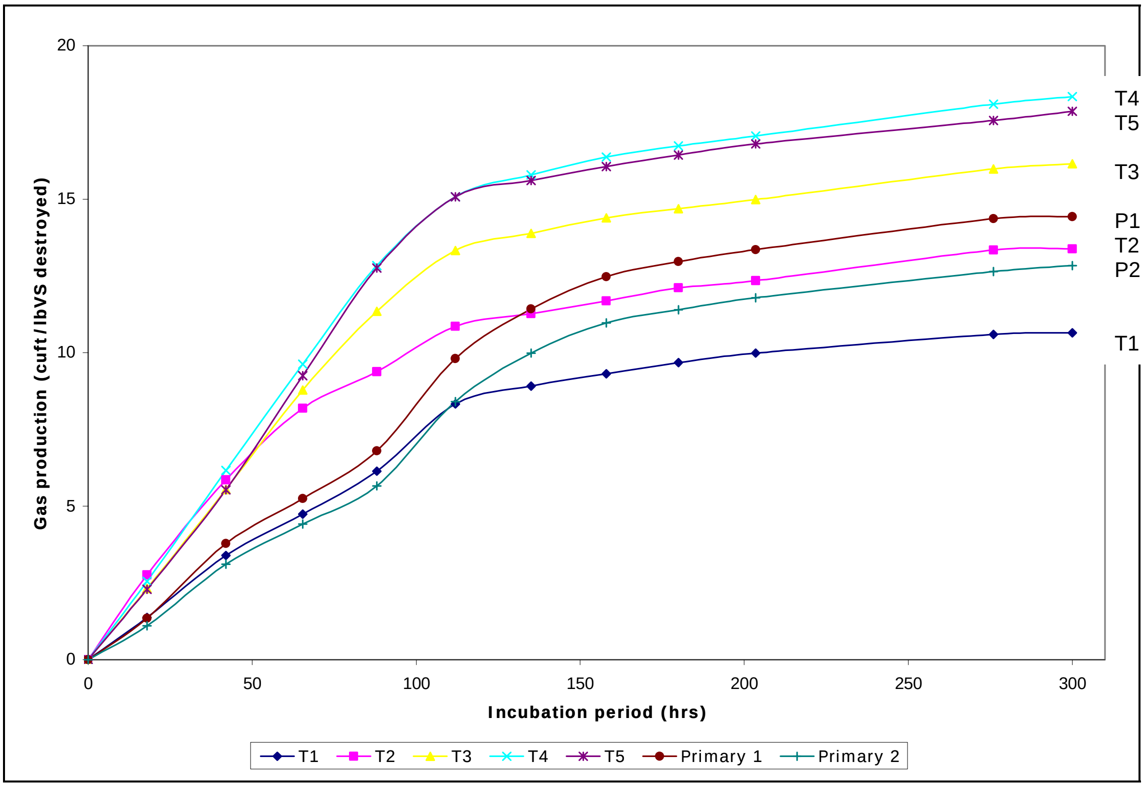


Figure B5 Average gas production in cuft/lbVS destroyed during Trial B digestibility testing



In Trial A the performance of the digester with primary sludge was significantly below that of the digesters with sludge from ABR tanks 1 to 5 (figure B2 and B3). The reason for the poor performance was believed to be the high concentration of solids in the primary sludge resulting in the digester being overloaded. The solids concentration for the primary sludge was about 4% compared to about 1% for the sludge from Tanks 1 to 5. The pH of the primary digester fell to 6.6, further evidence that the digester had failed. For digesters with sludge from tanks 1 to 5 the pH at the end of the incubation period was between pH7.5 and pH7.7.

The gas production for tanks 2 thru 5 based on volatile solids destroyed was very similar (figure B3). It is interesting to note that gas production over the period of incubation for sludge from tank 1 is approximately half that of tank 5 in terms of ft^3/lb VS destroyed. One possible explanation is that the partial digestion of solids in tanks 2 thru 5 facilitates the additional digestion during the incubation period.

This finding appears to be supported by Trial B (Fig B5). The total gas production per lb volatile solids destruction is lowest for tank 1 and increases steadily for sludge from tank 2 thru tank 5. The total volume of gas produced for each pound of volatile solids destroyed is also in line with that expected for full-scale anaerobic digestion, at around 15 ft^3. This is supporting evidence that the batch-type incubation provides a reasonable surrogate measure for the digestion process. However, it should be noted that the experimental protocol was not intended to replicate the actual operation of full-scale digesters which would undergo regular feeding and withdrawal of sludge over such a time-period. From the results it is reasonable to conclude that sludge from an ABR would be expected to digest in a similar manner to primary sludge, generating similar quantities of gas.

8.2 ABR Sludge Dewaterability

Total solids levels achieved during the dewaterability testing of ABR and primary sludges are shown in Table B1.

Table B1 TS concentrations in dewatered primary and ABR sludge

Type of test		Primary Sludge (% TS)	ABR Sludge (% TS)
Pre digestion *3 lbs pressure*	Pre DW	0.61 ± 0.02	0.68 ± 0.04
	Post DW	11.9 ± 0.5	12.7 ± 0.77
Post digestion 1 *50 lbs pressure*	Pre DW	0.77 ± 0.01	0.77 ± 0.02
	Post DW	13.9 ± 0.49	13.5 ± 0.85
Post digestion 2 *250 lbs pressure*	Pre DW	0.84 ± 0.01	0.83 ± 0.01
	Post DW	15.8 ± 0.61	16.0 ± 0.59

There was no significant difference in the *pre-digestion* dewaterability of primary sludge and ABR sludge (for a composite sample of ABR sludge). There was also no significant difference in the *post-digestion* dewaterability of primary sludge and ABR sludge (for a composite sample of ABR sludge).

8.3 Methane Analysis

8.3.1 Methane Solubility

Results from the analysis of ABR influent and effluent samples for dissolved methane are shown in Table B2

Table B2 Dissolved methane concentrations in ABR influent and effluent

Date 2004	Methane concentration (mg/l)			
	Influent	T5 Effluent	T4 Effluent	Saturated T5 effluent
21st April	0.08	1.3		
28th April		0.89		
29th April		0.35		12
5th May		0.31		
11th May			0.82	
13th May			0.84	

The results show that the ABR influent contained only negligible quantities of methane. Results for the effluent show unexpectedly low methane concentrations and indicate the ABR effluent was not methane saturated. There are, however, a number of reasons to question the laboratory data for the concentration of methane in the ABR effluent:

- The sampling process could lead to methane being lost from solution
- There is an unexplained three-fold range in measured methane concentrations when the flow to the ABR remained constant
- The effluent methane concentration from tank 4 is not consistent with the tank 5 effluent concentration (which would be higher given methane production in tank 4)
- The laboratory results are not consistent with the mass balance as detailed in section 6.2.5
- The laboratory results are not consistent with the observed increase in methane gas in the headspace above the ABR overflow tank (see section 3.3.2).

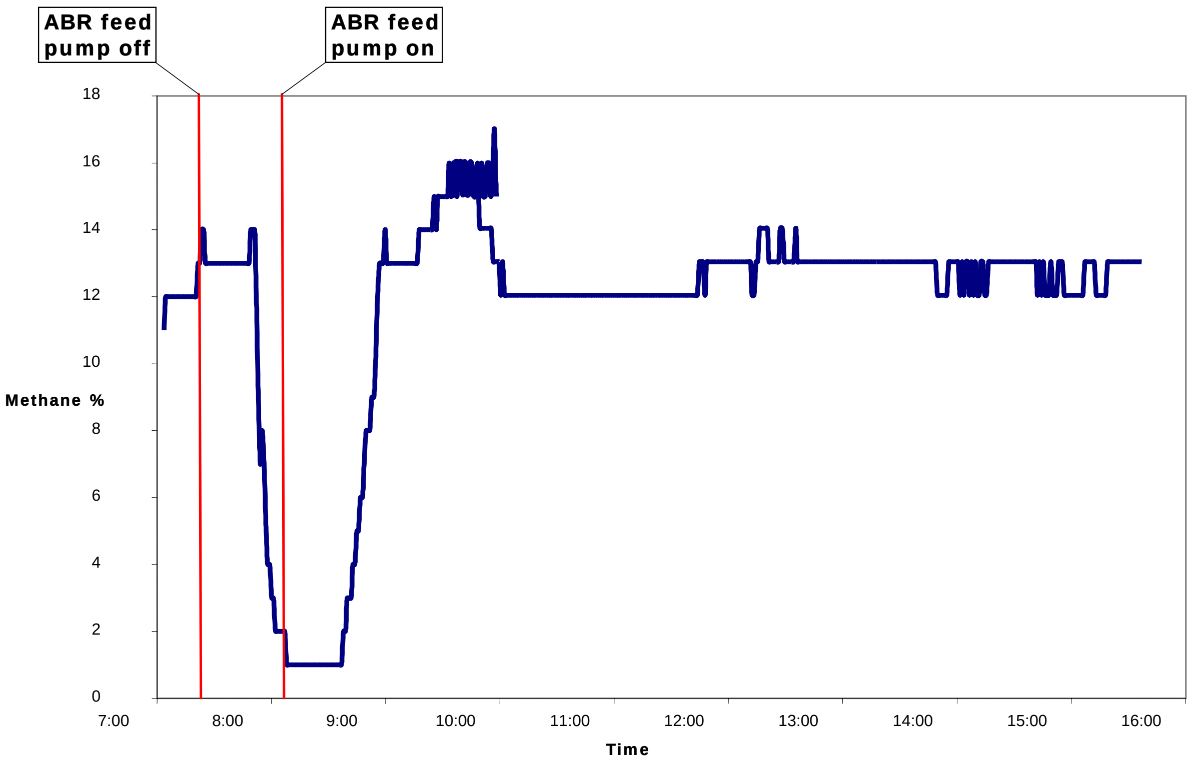
8.3.2 Methane Release in turbulent conditions

In the absence of full-scale data relating to the release of methane from solution at an overflow weir it was decided to use the ABR effluent tank to generate experimental results on the rate of build of methane in the headspace. The results were informative for two reasons:

- It provided information about the potential rate of release of methane from solution
- Based on the rate of release of methane from solution, the results provided a possible range for dissolved methane concentration
- It provided a comparator with the laboratory measurements of methane concentration in solution to see if the results were consistent
- The results helped confirm that the assumptions used in the risk assessment study (section 6.2.6) were valid and tended towards a worst-case scenario

The ABR pilot-plant effluent tank was covered (figure B1) so that a confined headspace of know volume was created. The flow into the ABR was then turned off to empty the effluent holding tank. The methane concentration in the headspace fell to around 1%. When the flow was turned back on (figure B5) it took approximately 30 minutes for methane to reach a steady-state concentration of 12-13% in the confined headspace. The only source of methane that could cause the increase in headspace gaseous methane was that dissolved in the ABR effluent.

Figure B5 Methane build-up in ABR effluent tank headspace following headspace evacuation



By considering the volume of the effluent tank headspace, the concentration of methane reached in the headspace and the period of time required for that concentration to be attained, it is possible to calculate the necessary concentration of methane in solution as shown below:

~ Volume of the headspace at maximum flow = 18.71 cuft
  Volume of the headspace when tank is empty = 129.8 cuft

~ Methane concentration reached a maximum of 15% in 19 cuft, therefore the volume of methane released from effluent = 0.15 x 19 = 2.85 cuft

~ Given it took approximately 30 mins to reach a maximum concentration, the rate of release = 5.7 cuft/hour (160 liters/hour)

~ Knowing the effluent flow = 10 liters/sec = 36,000 l/hour
  At a concentration of 1 mg/l methane there will be 36 g methane/hour in the ABR effluent

~ If 1 mole gas = 16 grams methane = 22.4 liters volume at STP
  Then 36g methane = 36/16 x 22.4
  = 50.4 liters/hour
  = 1.8 cuft

Table B3 below shows the relationship between the methane concentration in solution and the percent of methane that needs to be released from solution in order to produce observed methane gas concentrations in the headspace.

Table B3 Relationship between methane concentration and release rate

% dissolved methane released from solution at overflow point	Methane concentration required in effluent to generate 15% headspace concentration (mg/l)
25	12.8
50	6.4
100	3.2

In order to reach 15% by volume methane in the headspace of the effluent tank in about 30 minutes, it would require 100% of dissolved methane to be released from effluent containing 3.2 mg/l. If it is assumed that the measurement of headspace methane concentration is accurate, there is an inconsistency with the recorded levels of dissolved methane (table B2).

9. CONCLUSIONS

Following the comparison of ABR sludge with primary sludge and the investigation of dissolved methane concentrations and its release from ABR effluent, the following conclusions have been made:

- There was no significant difference in the volume of gas produced per lb of volatile solids fed to the batch digester for primary and ABR sludge.
- More gas was produced for each lb of volatile solids destroyed as one progresses from tank 1 to tank 5.
- Overall there was no significant difference in the digestibility of primary sludge and ABR sludge and no reduction in gas yield.
- There was no significant difference in the pre-digestion dewaterability of primary sludge and ABR sludge (for a composite sample of ABR sludge).
- There was no significant difference in the post-digestion dewaterability of primary sludge and ABR sludge (for a composite sample of ABR sludge).
- Methane build-up in the effluent tank headspace reached a steady state 30 minutes after tank evacuation and methane concentration remained around 12-13% by volume.
- It would not be possible to reach 12% methane in the headspace with a concentration of 1 mg/l methane in solution during a 30 minute period, given the effluent flow rates.
- A methane concentration in the ABR effluent of up to 10mg/l is consistent with the rate of build-up of methane in the effluent tank headspace and the saturation concentration achieved in the laboratory.