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THESIS TITLE:

THE APPRAISAL OF ANAEROBIC DIGESTION IN IRELAND TO DEVELOP IMPROVED DESIGNS AND OPERATIONAL PRACTICE

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IN MEMORY OF

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ABSTRACT

Mesophilic Anaerobic Digestion treating sewage sludge was investigated at five full-scale sewage treatment plants in Ireland. The anaerobic digestion plants are compared and evaluated in terms of design, equipment, operation, monitoring and management. All digesters are cylindrical, gas mixed and heated Continuously Stirred Tank Reactors (CSTR), varying in size from 130m^3 to 800m^3. Heat exchanger systems heat all digesters. Three plants reported difficulties with the heating systems ranging from blockages to insufficient insulation and design. Exchangers were modified and replaced within one year of operation at two plants. All but one plant had Combined Heat and Power (CHP) systems installed. Parameter monitoring is a problem at all plants mainly due to a lack of staff and knowledge. The plant operators consider pH and temperature the most important parameters to be measured in terms of successful monitoring of an anaerobic digester. The short time taken and the ease at which pH and temperature can be measured may favour these parameters.

Three laboratory scale pilot anaerobic digesters were operated using a variety of feeds over a 144-day period. Two of the pilots were unmixed and the third was mechanically mixed. As expected the unmixed reactors removed more COD by retention of solids in the digesters but also produced greater quantities of biogas than the mixed digester, especially when low solids feed such as whey was used. The mixed digester broke down more solids due to the superior contact between the substrate and the biomass. All three reactors showed good performance results for whey and sewage solids. Scum formation occurred giving operational problems for mixed and unmixed reactors when cattle slurry was used as the main feed source.

The pilot test was also used to investigate which parameters were the best indicators of process instability. These trials clearly indicated that total Volatile Fatty Acid (VFA) concentrations was the best parameter to show signs of early process imbalance, while methane composition in the biogas was good to indicate possible nutrient deficiencies in the feed and oxygen shocks. pH was found to be a good process parameter only if the wastewater being treated produced low bicarbonate alkalinities during treatment.

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CHAPTER 1

INTRODUCTION TO ANAEROBIC DIGESTION

1.1 SLUDGE MANAGEMENT

1.1.1 GLOBAL SLUDGE MANAGEMENT

Sludge management is becoming a more complex problem due to increased difficulties in locating disposal sites and meeting more stringent environmental quality requirements. The ideal solution to the sewage sludge disposal problem would be to develop sludgeless sewage treatment systems. Assuming that this is unlikely to occur anytime soon, the first goal should be to develop cost-effective sewage treatment processes, which produce less sludge, or processes that produce sludge which has beneficial use (Spinosa *et al.*, 1994).

Sludge stabilisation and dewatering methods are regarded as imperative prior to any disposal option. Recently there has been a resurgence of interest in incineration due to the increasing pressure on disposal options, such as agricultural use and landfilling. New processes and alternative uses of sludge such as deep-hole wet oxidation in The Netherlands and low pyrolysis to produce oil in Canada have been proposed and experienced, but most of them still require further investigations to evaluate effectiveness and reliability (Spinosa *et al.*, 1994).

In any case, it is not realistic to search for a global solution to the problem, but the best solutions must derive from local and site-specific considerations. Each Country and Community must decide what is the most suitable way of solving their sludge management problems in terms of cost, local factors and circumstances, including environmental considerations and benefits obtained from sludge disposal.

An example of a well-established process that reduces sludge production, while stabilising the sludge, is the anaerobic digestion process. This involves the anaerobic degradation of organic matter and as a direct result of this action, there is a reduction in solids of up to 50%. There are many more processes, which can reduce the solids of sludge, however the anaerobic digestion process has the added benefit of producing as a useful by-product, the fuel methane gas. As with most treatment processes, every system still has problems, even though significant developments have occurred during the past few years. The present project investigates and discusses the developments of the anaerobic digestion process, with particular emphasis on the future of anaerobic digestion of sewage sludge in Ireland.

1.1.2 SLUDGE MANAGEMENT IN IRELAND

Municipal wastewater (sewage) sludge might only comprise of 1% of the total non-hazardous sludge in Ireland, however it is potentially the most dangerous generated. The range of potential contaminants in sewage sludge is greater than in most other sludge types (agriculture and industrial) because of the diffuse nature of the discharges into the municipal wastewater stream. Sewage sludge may contain disease-causing organisms, heavy metals and organic micropollutants. Anaerobic digestion is a process, which will reduce these undesirable compounds before the sewage sludge is safely disposed of.

Options, for disposal of sewage sludge until recently include *Landfill, Agricultural Land,* and *Sea.* Virtually all sludge arising from municipal wastewater treatment is disposed of to landfill however the proposed Landfill Directive is likely to have implications for this disposal option. In particular, gate fees for sludge are likely to increase significantly, primarily due to the unsuitability of sludge for landfill disposal. Sludge consumes landfill

capacity, the high water content of sludge makes it unstable and leads to the creation of odour nuisances and gas generation when biodegrading (D'Alton, 2000).

The use of wastewater sludge in agriculture is regarded as the most sustainable method of sludge management (Department of the Environment, 1999). However, only limited volumes of sewage sludge (10%) is returned to land. The direct use of sewage sludge as a fertiliser on land is receiving increased scrutiny (D'Alton, 2000). This application to land, can lead to several environmental problems, which include surface runoff, leaching of nitrate into groundwater, possible dissemination of pathogens, persistence of toxic organic contaminants, heavy metal leaching and uptake by plants, and threat to human health due to pathogenic microorganisms (Spinosa *et al.*, 1994). In Ireland, sewage sludge is the only sludge used in agriculture, which is governed directly by legislation (Directive 86/278/EEC adopted by S.I. No. 183 of 1991 and by S.I. No. 148 of 1998).

In the past, most of the sewage sludge produced in Ireland was disposed of to the sea. This practice is no longer permitted under the 1991 Urban Wastewater Treatment Directive. All major sludge deposition to sea around the Irish coast had now ceased. As well as placing a ban on sea disposal this Urban Wastewater Treatment Directive (91/271/EEC) as adopted by S.I No. 419 of 1994 also requires treatment of waste discharging from major population centres by 31 December 2005. Consequently, while the volume of sewage sludge is set to increase substantially, traditional disposal routes will become unavailable.

The Department of the Environment & Local Government (DoELG) in Ireland, aware of the effect which the Urban Wastewater Treatment Directive would have on municipal wastewater sludge volumes and disposal methodologies, commissioned a *Strategy Study on*

Options for the Treatment/Disposal of Sewage Sludge in Ireland in 1992. The resulting report, prepared by Consultants Fehily Timoney Weston was completed the following year. This report is also known as the "Weston Report". It subdivided the country into regions for municipal wastewater sludge treatment. At least one hub-centre for the treatment of wastewater sludge was designated within each region and recommendations appropriate to that region were made for the use of the treated sludge arising at each hub-centre. Anaerobic digestion was the most recommended process for the stabilising of sewage sludge. Other options included aerobic digestion, composting, lime treatment, N-Viro/Agri-soil process, thermal drying, bio-drying, incineration and long term storage. Landspreading and forestry were the most popular disposal options recommended in the report for the disposal of sludge. Other options included land reclamation, restoration of peat bogs and land co-disposal.

Subsequently, Fehily Timoney & Co. completed a number of reports on behalf of the DoELG:

- *Inventory of Non-Hazardous Sludges in Ireland (1997)*

  The conclusions of this *Inventory* were that 93% of all non-hazardous sludges in Ireland arise from the agricultural sector. A further 6% are generated by industry, while 1% comprises municipal wastewater and water treatment sludges.
- *A Study of International Practices on the Use of Biosolids in Agriculture (1998)*
- *Code of Good Practice for the Use of Biosolids in Agriculture (1999)*
- *Model Sludge Management Plan for Tipperary South Riding (1999).*

For the most part, these documents relate to the sustainable management of municipal wastewater sludge. However, the principles as outlined are relevant to management of all non-hazardous sludges.

The *Code of Good Practice for the Use of Biosolids in Agriculture* has defined Biosolids as being:

> *"The organic by-product of urban wastewater treatment which, by being treated to an approved standard, can be beneficially used in agriculture".*

Directive 86/278/EE permits the use of untreated municipal wastewater sludge in agriculture provided that it is incorporated into the solids after application. In contrast, under the *Code of Good Practice for the Use of Biosolids in Agriculture,* untreated municipal wastewater sludge may not be used for agricultural purposes. Only municipal wastewater sludge, which has been so treated to achieve the following defined standards of pasteurisation, may be used as a fertiliser:

- *Faecal coliform* < 1,000 MPN*. g^{-1} dry solids
- *Salmonella spp.* < 3 MPN*.4g^{-1} dry solids

Note: * Most Probable Number (MPN) method for measuring coliform indicator bacteria in a laboratory (No. 9-75, Standard Methods, 1995).

This Code advised six treatment processes which, when operated to specified conditions of time and temperature, may treat municipal wastewater sludge to achieve high quality Biosolids. These include:

- Mesophilic anaerobic digestion with pre- or post-pasteurisation
- Thermophilic anaerobic digestion
- Thermophilic aerobic digestion
- Composting - windrowing
  - static pile or in-vessel
- Alkaline stabilisation
- Thermal drying

However, these processes are recommended only and their specification in the Code of Good Practice does not preclude developments of existing processes or the introduction of new processes, provided that pasteurisation standards are achieved. The Code of Good Practice also encourages beneficial use and quality control of biosolids (D'Alton, 2000).

Efforts to overcome the sludge challenges began in Ireland within the last 15 years. Two mesophilic anaerobic digestion plants already existed and were in operation before the Weston report was commissioned in 1992. Following the recommendation in this report three more anaerobic digestion plants have been commissioned and three are under construction. It is thought that over 10 more mesophilic anaerobic digestion plants are at design stage. However, due to the subsequent studies, Local Authorities may review these anaerobic digestion designs to include pre- or post pasteurisation treatment with mesophilic plants or alternatively use thermophilic digestion to achieve effective pathogen kills. Anaerobic digestion seems to be the most favourable option in treating sewage sludge in Ireland at the moment. This present research study investigates the anaerobic digestion option for sewage sludge stabilisation in Ireland. Investigations were conducted on both full-scale operational anaerobic digestion sites and pilot-scale anaerobic digesters.

1.2 INTRODUCTION TO ANAEROBIC DIGESTION

1.2.1 DEFINITION

Anaerobic digestion is the microbial degradation of organic compounds in the absence of oxygen with the production of a gaseous mixture called Biogas. This biogas usually consists of 50-70% methane (CH_4), 30-40% carbon dioxide (CO_2) and trace amounts of hydrogen (H_2), ammonia (NH_3) and hydrogen sulphide (H_2S).

The primary objectives of anaerobic digestion treating sewage sludge, emphasised today, are to produce a low volume, harmless, odourless sludge that can be disposed of easily. Therefore anaerobic digestion is regarded as a pollution-control system with the added advantage of producing a fuel.

1.2.2 BRIEF DESCRIPTION AND APPLICATIONS OF THE ANAEROBIC DIGESTION PROCESS

Anaerobic digestion involves the decomposition of organic matter into biogas, in the absence of molecular oxygen. This process is carried out in an airtight container, known as a digester or reactor. The contents of the digester are heated and mixed depending on the reactor configuration. Sludge, introduced continuously or intermittently, is retained in the rector for varying periods of time (Retention Time). The stabilised sludge is withdrawn continuously or intermittently from the reactor with a reduced organic and pathogen content and is nonputrescible. The biogas produced during this process is normally collected in a gas collection tank and can be used to heat and mix the contents of the digester or other uses outside the digestion process, e.g. heating buildings, electricity generation etc. Recognition of

these advantages has resulted in a broadening application and use of anaerobic treatment process throughout the world. Nowadays, the anaerobic digestion process has been adapted to convert intractable plant residues, agricultural wastes, manures, effluents from the food and beverage industry, paper and pulping, landfill leachate, municipal solids use and some chemical industry wastes into useful by-products.

1.3 PRINCIPLES OF THE ANAEROBIC DIGESTION PROCESS

The anaerobic digestion of organic compounds to methane and carbon dioxide is a multistep process involving different physiological groups of microorganisms (Pavlostathis & Giraldo-Gomez, 1991). As shown in Figure 1.1, anaerobic treatment processes require the presence of a diverse and closely dependent group of bacteria to bring about the complete conversion of complex mixtures of substrates to methane gas. A good understanding of the microbiological parameters and operating principles necessary for successful methanogenesis is essential for the utilisation of anaerobic treatment. All anaerobic biological treatment involves a consortium of bacteria and is based on a series of reactions, the slowest of which will determine the overall safety factor for that system (Speece, 1996). There are two possible rate-limiting steps in the anaerobic digestion of sludges. The first is the time taken for hydrolytic bacteria to break down complex long chain macromolecules such as cellulose. To achieve this reactors must have sufficiently long retention times to ensure the breakdown of such molecules (10 - 20 days). The second rate-limiting step involves the sensitive and slow growing methanogens (producing methane from a small variety of substances) (van Haandel & Lettinga, 1994). Under favourable environmental conditions the multitude of other microbes involved in the anaerobic consortium would also function well.

A controlling factor generally accepted at the present is that about two thirds of the methane produced in an anaerobic reactor is derived from acetate and the remainder from H_2 and CO_2 (Speece, 1996). This was proven by Smith and Mah (1966) when 73% of the methane produced in a domestic sewage digester was deduced from acetate (Hobson *et al.*, 1992).

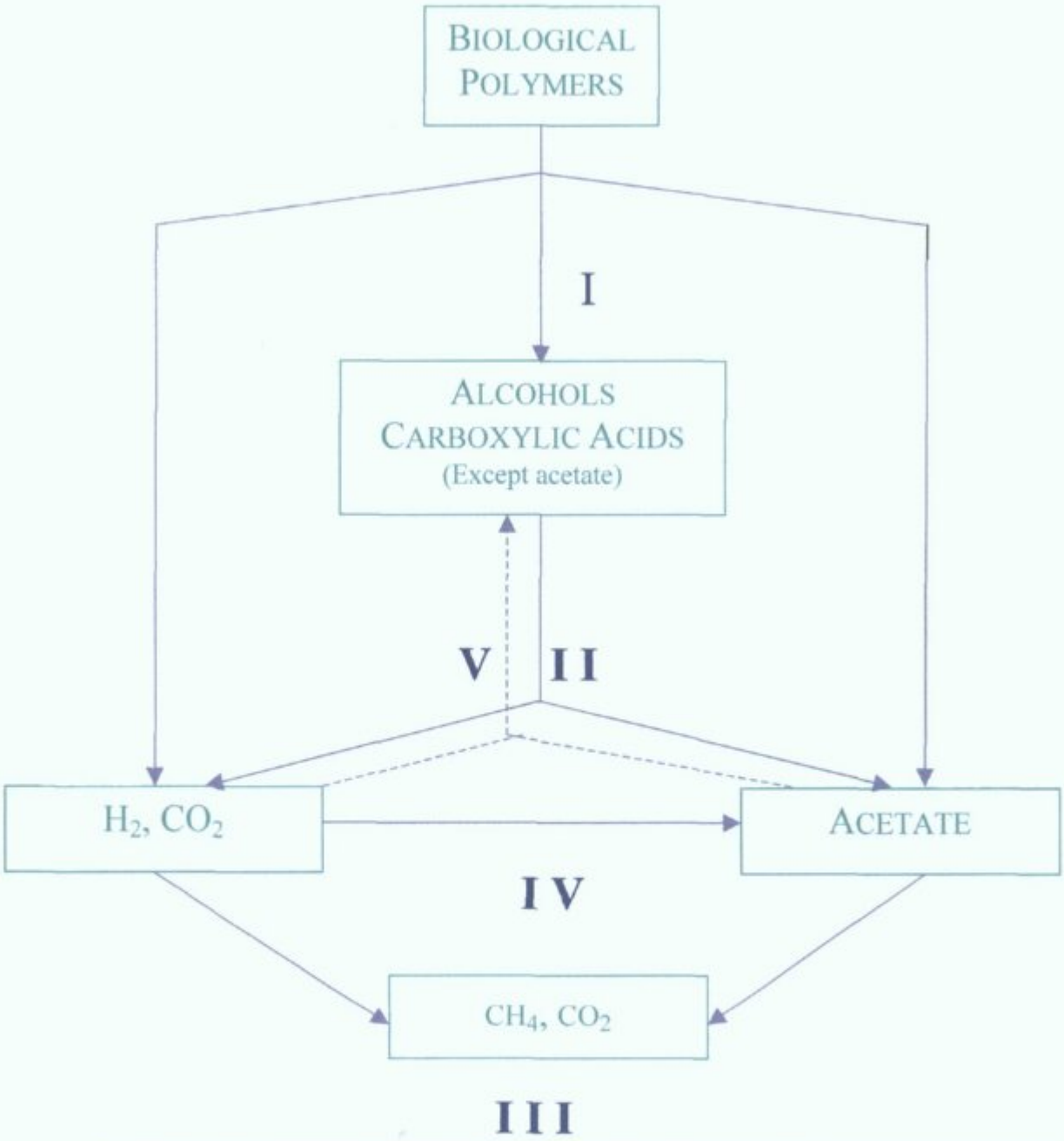
1.3.1 THE MICROBIOLOGY PROCESS

The microbial consortia active in anaerobic treatment execute a complex process involving many classes of microorganisms, mostly bacteria, in several steps (Figure 1.1). Four different stages can be distinguished in the overall conversion process.

1.3.1.1 Hydrolysis (*Hydrolytic Bacteria*)

Consortia of anaerobic bacteria break down complex organic molecules (proteins, cellulose, lignin, and lipids) into soluble monomer molecules such as amino acids, glucose, fatty acids, and glycerol. The monomers are directly available to the next group of bacteria. Hydrolysis of the complex molecules is catalysed by extracellular enzymes such as cellulases, proteases and lipases. However, the hydrolytic phase is relatively slow and can be limited in anaerobic digestion of wastes such as raw celluloltic wastes, which contain lignin (Polpraseert, 1989; Speece, 1983; Parkin & Owen 1986; Bitton, 1994).

FIGURE 1.1 FLOW DIAGRAM OF THE ANAEROBIC DIGESTION PROCESS IN ANAEROBIC DIGESTERS



I = Hydrolytic and fermentative bacteria
II = Acetogenic bacteria
III = Methanogenic bacteria (a) hydrogenophilic, (b) acteophilic
IV = Homoacetogenic bacteria
V = Fatty acid-synthesizing bacteria

(Bryant *et al.*, 1967, Zehnder *et al.*, 1982)

1.3.1.2 Acidogenesis (*Fermentative Acidogenic Bacteria*)

Acidogenic (i.e., acid-forming) bacteria convert sugars, amino acids, and fatty acids to organic acids (e.g., acetic, propionic, formic, lactic, butyric, or succinic acids), alcohols and ketones (e.g., ethanol, methanol, glycerol, acetone), acetate, CO_2, and H_2. Acetate is the main product of carbohydrate fermentation. The products formed vary with the type of bacteria as well as with culture conditions (temperature, pH and redox potential) (Bitton, 1994). Hydrogen is inhibitory to many of the acid-forming bacteria and must be removed from the system if acid production is to continue. Fortunately, hydrogen is an energy source for some methanogenic bacteria and is rapidly consumed in the reduction of carbon dioxide to methane (Parkin & Owen 1986).

1.3.1.3 Acetogenesis (*Acetogenic Bacteria*)

Acetogenic bacteria (acetate and H_2-producing bacteria) convert fatty acids (e.g., propionic acid, butyric acid) and alcohols into acetate, hydrogen, and carbon dioxide, which are used by the methanogens. This group requires low hydrogen tensions for fatty acid conversion. Under relatively high H_2 partial pressure, acetate formation is reduced and the substrate is converted to propionic acid, butyric acid, and ethanol rather than methane. There is a symbiotic relationship between acetogenic bacteria and methanogenic bacteria. Methanogens help achieve the low hydrogen tension required by acetogenic bacteria. Ethanol, propionic and butyric acid are converted to acetic acid by acetogenic bacteria for use as substrate by the methanogens. Acetogenic bacteria grow much faster than methanogenic bacteria. The former group has a μ_{max} (maximum growth rate) of approximately 1/hr, whereas the μ_{max} of the latter is around 0.04/hr (Hammer, 1986, Bitton, 1994).

1.3.1.4 Methanogenesis (*Methanogenic Bacteria or Methanogens*)

Methanogens are subdivided into two subcategories:

- *Hydrogenotrophic methanogens* (i.e., hydrogen-utilising chemolithotrophs) convert hydrogen and carbon dioxide into methane. The hydrogen-utilising methanogens help maintain the very low-level partial pressures necessary for the conversion of volatile acids and alcohols to acetate (Speece, 1983).
- *Acetotrophic methanogens*, also called acetoclastic bacteria or acetate-splitting bacteria, convert acetate into methane and CO_2. The carbon dioxide produced either escapes as gas or is converted to bicarbonate alkalinity (Parkin & Owen, 1986). The acetoclastic bacteria grow much more slowly (generation time = a few days) than the acid-forming bacteria (generation time = a few hours). About two thirds of methane is derived from acetate conversion by acetotrophic methanogens. The other one third is the result of carbon dioxide reduction by hydrogen (Mackie & Bryant, 1981; Bitton, 1994).

1.3.2 Series Metabolism

As explained by Speece (1996), series metabolism is similar to a bucket brigade of water where the buckets are progressively passed down a line of people to douse a fire located away from the water source. Both types of series, the bucket brigade and the methanogenic series, are rate-controlled by the slowest member involved in the process. In the case of anaerobic treatment the slowest step is characterised by an accumulation of substrate build-up found just prior to the rate-controlling step. If the form of substrate is a non-acid organic (e.g. alcohol) there may be no adverse impact on the overall consortia. The slowest members of the consortia often are the propionic- or acetic acid-utilisers, so that an accumulation of these organic acids can overwhelm the reserve bicarbonate alkalinity. Such a malfunction may

cause a drop in pH, which can have a drastically adverse impact upon the entire microbial consortia. Unfortunately the greatest inhibition of a low pH may also be directed at the propionic- and acetic acid-utilisers themselves, compounding the problem.

The anaerobic process works well as long as each subsequent class of organisms processes the organic intermediaries at least as fast as they are produced. Since microbial processes function at a rate proportional to their substrate concentration, an accumulation of substrate may result before they are able to process it as fast as it is passed on to them (Speece, 1996).

1.3.3 COD EQUIVALENCE OF METHANE

It is relatively easy to determine a mass balance on an anaerobic treatment process. The COD equivalent of methane is as follows:

$$CH_4 + 2\,O_2 \rightarrow CO_2 + 2\,H_2O$$

From this equation it can be determined that for each mole of methane consumed (22.4L @ 0°C), two moles of oxygen equivalent are destroyed (64g). Thus 0.35L (22.4 L/64g) of CH_4 at 0°C and 760mm Hg pressure (STP) is equivalent to 1 g COD destruction. To compensate for a higher temperature, the CH_4 equivalence is 0.395 L at 35°C and one atmosphere (Parkin and Owen 1986; Speece, 1996). Wheatley (1990), stated there should be a constant production of methane from the same amount of organic matter metabolised, i.e. about 0.5m^3 methane/kg of (volatile solids) VS or 0.35 m^3 CH_4/kg COD removed.

1.3.4 CONCEPT OF WASHOUT AND SAFETY FACTOR

Predicting *bacterial generation time* and/or *bacterial washout time* can assist rational design of anaerobic systems. The size of an anaerobic digester should be adequate to ensure that the solids retention time in the system never falls below a critical minimum solids retention time. This critical value indicates the SRT at which washout of microorganisms begins, i.e. the point where the rate of organisms leaving the system exceeds the rate of generation of these organisms inside the system.

Parkins & Owen (1986) gives the bacterial generation time as $\theta_c^{lim} = (yk - b)^{-1}$

and the bacterial washout time as $\theta_c^{min} = \left(\frac{ykS^o}{K_c + S^o} - b \right)^{-1}$

where

y	=	Bacterial yield g VSS/g COD,
k	=	Maximum specific substrate utilisation rate g COD/g VSS/d,
b	=	Bacterial decay rate/day,
S^o	=	Influent biodegradable substrate concentration g/l,
K_c	=	Fatty acids + propionic acid + acetic acid g/l.

The design solids retention time (SRT_d) should be calculated by applying a safety factor to the bacterial generation time or the bacterial washout time. Typical safety factors inherent to the design of conventional high-rate sludge digesters range from 2.5 - 10. The safety factor applied to conventional full-scale digesters usually includes compensation for inefficient mixing, varying quantities and composition of the feed sludge over time and sludge temperature fluctuations. Once the SRT_d is selected, process efficiency and methane production rates can be estimated (Parkin & Owen, 1986, US EPA, 1979). Process success,

depends on the selection of appropriate safety factors built into systems to ensure that the process meets target effluent standards while maintaining the correct conditions for the satisfactory metabolism of the biomass (Speece, 1996).

1.3.5 MINIMUM GENERATION TIME

The biomass responsible for conversion of carbohydrates, proteins, lipids, acetate, propionate, and H_2 may be considered rate controlling in certain cases for soluble substrates. Pavlostathis and Giraldo-Gomez (1991) calculated the minimum values of SRT for anaerobic treatment of various substrates (Table 1.1). Note that fatty acids, fats and H_2 utilising microbes have very long generation times compared with their aerobic counterparts of less than approximately 0.2 days.

TABLE 1.1 LIMITING ANAEROBIC GENERATION TIME (AT 35 °C)

Substrate	*Days*
Carbohydrates	0.18
Proteins	0.43
Acetate	3.9
Propionate	3.3
Butyrate	2.0
Lipids	3.2
H_2	1.2

(Pavlostathis & Giraldo-Gomez, 1991)

1.4 REQUIREMENTS FOR OPTIMAL ANAEROBIC TREATMENT

Important environmental factors affecting anaerobic sewage digestion are temperature, pH, the presence of essential nutrients and the absence of excessive concentrations of toxic compounds in the influent. In the case of sewage, the latter three factors normally do not need consideration unless the sewers receive untreated wastes from chemical or other industries. An adequate and stable pH is set by the presence of the carbonic system and no chemicals are needed to correct the pH. Nutrients (macronutrients, nitrogen and phosphorus, and micronutrients) are abundantly available in sewage (Parkin & Owen, 1986; van Haandel *et al.*, 1994).

1.4.1 TEMPERATURE

Anaerobic digestion, like other biological processes, strongly depends on temperature. With respect to the conversion rate of digestion processes, there are maxima between 35 and 40°C for the mesophilic range and at approximately 55°C for the thermophilic range. For sewage treatment, mesophilic digestion is more common (van Haandel *et al.*, 1994). However, because thermophilic digestion removes more pathogens than mesophilic, there may be an increase in thermophilic digestion for the treatment of sewage sludge to reduce the amount of potentially dangerous pathogens.

The influence of temperature on the rate and extent of anaerobic digestion has been the subject of many investigations. From Figure 1.2 Henze and Harremoes (1983) came to the following conclusions:

(1.) the optimum temperature range is between 30 and 40°C and

(2.) for temperatures below the optimum range the digestion rate decreases by 11 per cent for each °C temperature decrease, or according to the Arrhenius expression:

$$r_t = r_{30}\,(1.11)^{(t-30)}$$

where

t = temperature in °C

r_t, = digestion rate at temperature t,

r_{30} = digestion rate at temperature 30°C.

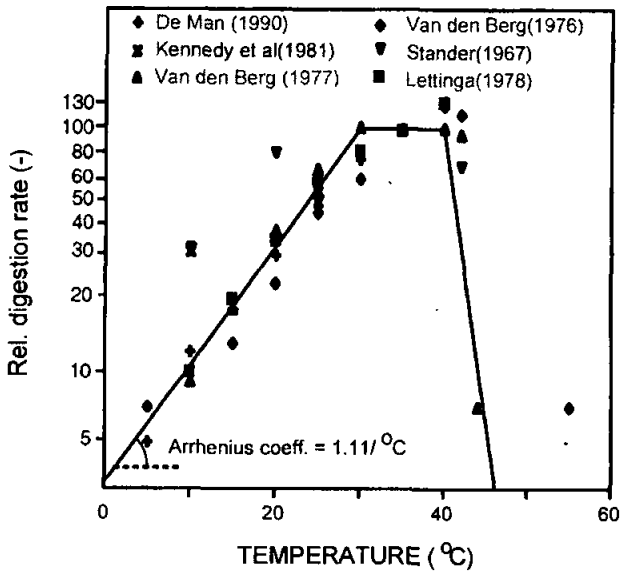


FIGURE 1.2 INFLUENCE OF TEMPERATURE ON THE RATE OF ANAEROBIC DIGESTION IN THE MESOPHILIC RANGE, AFTER HENZE AND HARREMOES.

(van Haandel *et al.*, 1994)

The influence of temperature on anaerobic digestion is not limited to the rate of the process. The extent of anaerobic digestion is also affected, as found by O'Rouke (1968) and van der Last (1991). Figure 1.3 shows the achieved extent of digestion for settled sewage solids (primary sludge) in relation to digestion time at different temperatures, according to the results of O'Rouke (1969). This diagram clearly reveals the strong dependence of solids digestion on temperature. The decrease in the fraction of organic matter degraded can be attributed to a low rate of hydrolysis (van Haandel *et al.*, 1994).

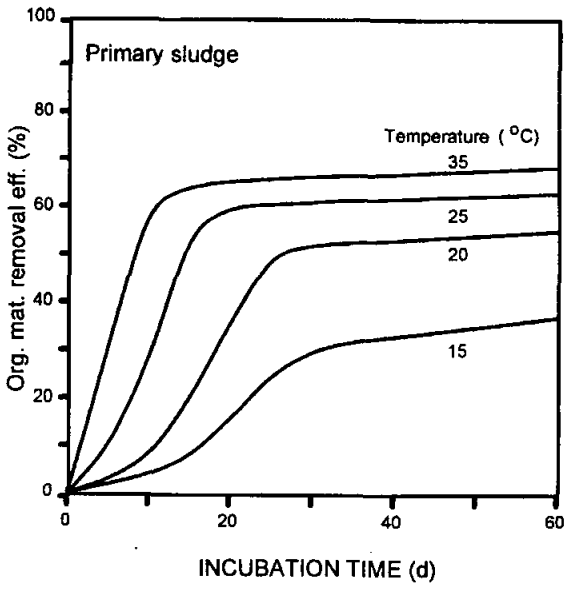


FIGURE 1.3 INFLUENCE OF TEMPERATURE ON THE EXTENT AND RATE OF ANAEROBIC DIGESTION OF PRIMARY SEWAGE SLUDGE, AFTER O'ROUKE (1968)

(van Haandel *et al.*, 1994)

1.4.2 pH

The pH of the environment within a reactor must range between approximately 6.5 to 8.2 to make possible successful anaerobic processing (Speece, 1996). Methanogenesis only proceeds at a high rate when the pH is maintained in the neutral range. The literature generally supports this range with slight variations e.g. van Haandel and Lettinga (1994) reports that at pH values lower than 6.3 or higher than 7.8 the rate of methanogenesis decreases. Parkins and Owen (1986) reports an acceptable range of 6.5 - 7.8 for good process efficiency. Methanogenesis will continue at pH 6.0 and even lower at reduced rates but the bicarbonate alkalinity does not buffer well under such conditions, and this environment tends to result in considerable instability (Speece, 1996). Acidogenic population are significantly less sensitive to low or high values, and hence acid fermentation will prevail over methanogenic fermentation, which may result in 'souring' of the reactor contents (van Haandel *et al.*, 1994).

Unless the system contains sufficient buffer capacity, the pH will drop to unacceptable low levels and methane production will decrease and may eventually cease if the pH is of sufficient magnitude or duration (souring). The main buffering system in anaerobic digestion is the bicarbonate system, where buffering capacity is generally measured as alkalinity. Maintenance of sufficient alkalinity guards against failure from pH drop (Parkins & Owen, 1986).

Some microbial action can significantly alter the pH of the feedstock. It is wasteful and possibly even detrimental to attempt to neutralise the wastewater itself. Acetic acid that has been completely neutralised with sodium, for example, presents potential excessive pH problems for anaerobic treatment because the gas production would be composed of 100%

methane with no CO_2 remaining to counter the high alkalinity. Such conditions would cause the pH potentially to rise so high that further methanogenesis would be precluded should the sodium acetate concentration in the feed also be sufficiently high. Components such as CO_2 and volatile short chain fatty acids tend to lower pH, while alkalinity-generating cations like ammonium ions from protein degradation and sodium from soap degradation, increase the alkalinity and the pH (Speece, 1996).

1.4.3 ADEQUATE MACRONUTRIENTS AND MICRONUTRIENTS

Nutrients must be present in sufficient quantities to ensure efficient digestion. Nitrogen and phosphorus are the macronutrients required in the highest concentrations. The carbon to nitrogen ratio (C:N) is frequently utilised to describe this micronutrient requirement. The carbon can be measured as chemical oxygen demand (COD). Typical COD:N ratios of about 400:7 and 1000:7 have been estimated as required at high and low substrate loadings, respectively (Henze *et al.*, 1983, van den Berg *et al.*, 1978).

The empirical formula for bacteria, is $C_5H_7O_2N$, in which nitrogen comprises approximately 12% of the bacterial cell mass. The phosphorus requirement for bacterial growth is about one seventh to one fifth that of the nitrogen requirement (Parkins & Owen, 1986; Speece, 1996). Nitrogen concentrations within the reactor ranging from 40 to 70 mg/l, must be provided to prevent nitrogen limitation (Speece, 1996). Both N and P are necessary to synthesise different compounds contained in the sludge. Domestic sludge usually contains enough quantities of nitrogen (in the form of protein, urea and ammonia) and phosphorus for efficient digestion (Parkins & Owen, 1986).

In addition to these two elements (N and P) required by the anaerobic microbial system some sulphide precursor must be available, commonly in the sulphate form. The methanogens manifest an obligate requirement for sulphide even though, this need may be satisfied by maintaining very low concentrations of both ions in the reactor. Requirements for macronutrients other than nitrogen are proportionally lessened when compared with aerobic processing due to the much-reduced amount of biomass synthesis characteristics of anaerobic processes (Speece, 1996).

The crucial role of trace metals in methanogenesis, if properly recognised, can make possible creative applications to a wide range of effluents. Iron appears to be required in the highest concentrations relative to all of the other trace metals, while cobalt, nickel, and zinc are most often reported as stimulatory (Speece, 1996).

Macronutrients and micronutrients are generally abundant in sewage sludge therefore deficiency of these compounds poses no problem in the anaerobic digestion of sewage sludge.

1.4.4 Absence of Excessive Toxicants

The nature of toxicity in biological waste treatment is often misunderstood, especially with regard to anaerobic digestion. Methane-producing bacteria are no more sensitive to toxicants than aerobic and/or facultative organisms. Whether a substance is toxic to a biological system is a matter of the nature of the substance, concentration and acclimation. Many substances will stimulate the reactions in low concentrations; however as concentration increases the effect becomes inhibitory (Parkins & Owen, 1986).

It is commonly known that failure in the anaerobic digestion process has often been explained by the sensitivity of the methanogenic bacteria to potential inhibitors, such as heavy metals (Kugelman & Chin, 1971; Graef & Andrews, 1974; Kroeker *et al.*, 1979; Ahring & Wastermann, 1985; Wong & Cheung, 1996). However, the presence of these compounds at inhibitory concentrations is unlikely in sewage. Potentially toxic compounds that might be present are oxygen and sulphide. Some oxygen may be introduced in the influent distribution system, but it will be used for oxidative metabolism in the acidogenesis process. Thus, no dissolved oxygen will be present in the anaerobic reactor, unless air is entrained together with the influent, so that its introduction will be of no consequence for the performance of the reactor. Sulphide can be formed in the process due to the reduction of sulphate. However, according to results of Rinzema (1989) the sulphide concentration to be expressed in anaerobic sewage treatment systems (up to 50mg/l) is far lower than the minimum concentration for noticeable toxicity. Therefore, toxicity will normally not be a problem in anaerobic sewage systems (van Haandel *et al.*, 1994).

1.5 OTHER REQUIREMENTS FOR OPTIMAL ANAEROBIC TREATMENT

In order for a microbial process to function effectively a number of conditions must be satisfied. The previous conditions are the most obvious, however Parkin (1986) has itemised ten requirements which he calls a checklist. The required items are proper pH, adequate macronutrients and micronutrients, temperature, toxicity (all of which are discussed in the pervious section), mass transfer, metabolism time, carbon source, electron donor and electron acceptor. The latter conditions will be discussed in this section.

1.5.1 MASS TRANSFER OF POLLUTANT INTO THE MICROBIAL CONSORTIA

It is important when designing to ensure adequate mass transfer between organic pollutants in the wastewater and the biomass responsible for its bioconversion by making possible both intimate contact and adequate exposure time. Intimate contact is much related to the technique used for mixing and cell immobilisation e.g. suspended growth systems or attached growth systems. (Parkins & Owen, 1986, Speece, 1996). For sewage anaerobic digesters attached growth systems, such as UASB, are usually not used and the adequate exposure time is directly related to the hydraulic retention time.

1.5.2 ADEQUATE METABOLISM TIME

Two measures of time are involved; Hydraulic Retention Time (HRT) and Solids Retention time (SRT) and can be defined as follows:

$$SRT = \frac{\text{Mass of solids in tank, kg}}{\text{Rate of solids removed, kg/day}}$$

$$HRT = \frac{\text{Working volume, L}}{\text{Rate of sludge removed, L/d}}$$

(Kiely, 1997)

HRT defines the window of opportunity afforded the microbes to accomplish their task and SRT determines which organism can replicate and predominate within the system as well as what biomass inventory (biological safety factor) can be maintained. These two types of time

intervals affect the system in different ways. Microbial treatment necessitates adequate contact time (HRT) for metabolism to occur and is very much related to the nature of the substrate, which is reflected in the degree of difficulty encountered by the biomass in metabolising it. Simple low molecular weight VFA, sugars and alcohols can be metabolised within minutes but large, complex, or chlorinated molecules may require hours or even days. Adequate SRT must be maintained to allow the biomass to regenerate and accumulate to a satisfactory inventory concentration (Speece, 1996). To date in sewage treatment systems SRT = HRT, however in many industrial effluent treatment systems such as anaerobic filters the microbial consortia are immobilised inside the reactor providing a long SRT and short HRT. Such systems are only appropriate for wastewater with high soluble organic wastes with low solids concentrations (Iza *et al.*, 1991; Kiely, 1997; Malina *et al.*, 1992; Killilea, 1992).

1.5.3 CARBON SOURCE FOR SYNTHESIS

Except for autotrophic methanogens converting H_2 to methane, anaerobic systems are mainly heterotrophic. The organics in the substrate provide the carbon source for synthesis of heterotrophic biomass. For the autotrophic hydrogen utilisers, the carbon source can be the dissolved CO_2 in the reactor (Speece, 1996).

1.5.4 ELECTRON DONOR AND ACCEPTOR

The electron donor that provides energy for the biomass activity is the organic pollutant in the feedstock - the biodegradable COD. Anaerobic systems operate in the absence of oxygen and utilise CO_2 or sulphate as their electron acceptors. CO_2 reduction results in CH_4

production sulphate reduction results in H_2S production. The acid producing organisms are fermentative, oxidising a fraction of the substrate and reducing the remainder. Aerobic systems reduce oxygen (as the electron acceptor) to water, while anoxic systems reduce nitrate/nitrite to nitrogen gas (Speece, 1996).

1.6 OPERATIONAL CONSIDERATIONS, MONITORING AND CONTROL

A working knowledge of the basic principles of anaerobic digestion process, including potential operational problems, is necessary for the successful application of the technology. Proper monitoring and control can prevent a minor operation problem from turning into a major plant failure.

1.6.1 NORMAL OPERATION

If digestion is proceeding satisfactorily, the pH value of the digesting sludge normally ranges from 6.6 to 7.4, alkalinity (as $CaCO_3$) from 1000 to 5000 mg/l and concentrations of volatile acids (as acetic acid) less than 250mg/l (Metcalf & Eddy, 1991). In general biogas production rates of between 0.3 - 0.5 m^3/kg VS might be expected with between 1 - 5 m^3/m^3 digester volume, although large variations have been found in operational digesters depending on process configuration (Combs, 1990; Kiely, 1997). The expected percentage methane of a process operating well would be approximately 70% with carbon dioxide 30%.

1.6.2 PROCESS INSTABILITY

Process instability of the anaerobic treatment can be triggered by many variations in the process such as changes in organic or hydraulic loading, changes in organic feed characteristics, changes in temperature, or introduction of toxic substances. Typically during process imbalance the volatile acid concentrations will increase (an increase of 200 - 300mg/l would be a cause for concern), while bicarbonate alkalinity, pH, gas production, percent methane, and the destruction of organic matter all decrease. Therefore careful monitoring of these parameters should allow digestion plant operators to observe the onset of stress and take appropriate remedial measures to prevent system failure.

The literature is varied and sometimes conflicting as to which parameter will give the quickest and best description of process performance. Much work has and still is being carried out in this area of parameter monitoring to try and find the best indicator parameter of digester performance. As part of this present research study, a pilot trial was set-up to investigate this topic and will be discussed in chapter 4 as well as a literature review on the subject.

Generally, measurement of only one parameter will not yield an adequate description of process performance. For reliable process monitoring, there is no doubt the most informative and safest way to investigate the performance of a reactor, if possible, is to measure all parameters frequently.

1.6.3 PROCESS CONTROL

If process instability occurs the general practice to recover the process is to cut the feed and add a chemical to correct the pH, which in time will reduce the VFA and the process will begin to recover. The most common chemicals used for pH correction in digesters are lime, sodium bicarbonate, sodium and potassium hydroxide and ammonia. The preferred chemical for pH control is sodium bicarbonate, although lime has been shown to yield rapid recovery from organic overload.

1.7 ADVANTAGES OF ANAEROBIC TREATMENT

The oil crisis of 1973 stimulates renewed interest in anaerobic digestion as a wastewater treatment system, as it did not require fossil fuel expenditure, but in fact generates net quantities of methane for use as a fuel in the treatment works (Schink, 1988). The primary objective of anaerobic digestion, emphasised today, is to produce a low volume sludge, which can be disposed of easily, regarding the process as a pollution-control system with the added advantage of producing a fuel.

Over the past decades many installations, embracing a variety of wastewaters, have demonstrated conclusively the positive features of the anaerobic digestion process in biotransforming organic pollutants to methane. Table 1.2 highlights some of these positive features.

TABLE 1.2 ADVANTAGES OF ANAEROBIC DIGESTION

Positive features of anaerobic digestion
▪ Reduction of waste biomass & installation space requirements
▪ Reduction in the number of pathogens
▪ Production of methane gas with ecological and economical benefits
▪ Retains nutrient values & reduction in odour
▪ Elimination of off-gas air pollution

1.7.1 REDUCTION OF WASTE BIOMASS

Due to the conversion of organic matter in the volatile solids to methane, carbon dioxide etc., the mass and volume of the sludge is reduced by approximately 25-35% (Malina, 1992). Anaerobic digestion negates the need for aerobic oxygen transfer with its associated high microbial synthesis characteristics, thus significantly lessening the disposal costs involved with excess biomass synthesis. Considerable reduction in space requirements further increases financial savings (Speece, 1996).

1.7.2 REDUCTION IN THE NUMBER OF PATHOGENS

The warm temperatures and the anaerobic conditions are key factors in the reduction of pathogens. Thermophilic digestion (~55°C) obviously remove more pathogens than the mesophilic process (~ 37°C) due to its higher temperatures. Thus anaerobically digested sludge is more hygienic and makes disposal to land less hazardous.

1.7.3 PRODUCTION OF METHANE GAS WITH ECOLOGICAL AND ECONOMICAL BENEFITS

Methane gas is a useable source of energy and the quantity produced, at a sludge treatment facility, is usually in excess of that required to maintain the temperature of the digesting sludge and to meet energy requirements for mixing (Malina, 1992). The surplus energy has been known to heat buildings, to drive the engines for other energy requiring processes, or to generate electricity, which can be used to drive pumps etc.

Anaerobic treatment produces 12 x 10^6 BTU as CH_4 per 1000 kg of COD. Because no oxygen transfer is required, the need for the 500-2,000 kw hrs of energy per 1,000 kg of oxygen transfer normally required for aerobic treatment is negated, making energy conservation possible with its accompanying ecological and economical benefits (Speece, 1996).

1.7.4 NUTRIENT VALUES RETAINED AND REDUCTION IN ODOUR

During the anaerobic digestion process the nitrogen, phosphorus and other organic material values of the sludge is not altered. The forms in which they occur in the sludge are altered and when they are applied to soil they are in a more available form. Therefore the digested sludge has a considerable fertiliser value that can improve the fertility and texture of the soil. The odour associated with raw sludge is markedly reduced to a musty odour by the anaerobic digestion process.

1.7.5 Elimination of Off-Gas Air Pollution

Many organic contaminants are volatile and tend to be air stripped from the wastewater during aerobic treatment before they are biodegraded, thus contributing to air pollution. This significant drawback cannot be omitted from the design process of aerobic systems, but eliminated when anaerobic treatment is utilised (Speece, 1996).

1.8 Disadvantages Of Anaerobic Treatment

In weighing the merits of anaerobic treatment for a given wastewater, certain disadvantages also need to be kept in mind. Sometimes it would not be practical to use anaerobic treatment as might be the case in processing low temperature, or dilute wastewater, insufficient alkalinity wastewater, or effluent requiring exceptionally low BOD for final discharge regulations. Careful examination of each situation in light of these and other disadvantages listed below may sometimes dictate aerobic biotechnology as the better choice. Possible disadvantages of the anaerobic digestion process are shown in Table 1.3.

Table 1.3 Disadvantages of Anaerobic Digestion

Possible disadvantages of anaerobic treatment
▪ Long start-up requirements for development of biomass inventory
▪ Capital costs are high
▪ Long hydraulic retention times are required to develop and maintain a population of methane producing bacteria.
▪ Insufficient methane generation from dilute wastewater to provide heating for the process

1.9 COMPARISON OF ANAEROBIC AND AEROBIC TREATMENTS

Table 1.4 sets forth a concise comparison of the operating features for treating a readily degradable wastewater by either anaerobic of aerobic treatment.

TABLE 1.4 COMPARISON OF ANAEROBIC/AEROBIC PROCESSES

Comparison of anaerobic/aerobic biotechnology
▪ Volumetric organic loading rates 5 - 10 times higher than for aerobic processes
▪ Biomass synthesis rates of only 5 - 20% of those for aerobic processes
▪ Nutrient requirement of only 5 - 20% of those for aerobic processes
▪ Anaerobic biomass preserved for months or years without serious deterioration in activity
▪ No aeration energy requirements for anaerobic processes vs. 500 - 2000 kwhr/1000 kg COD for aerobic processes
▪ Methane production of 12,000,000 BTU/1000 kg COD destroyed

A combination of tighter restrictions on air pollution, hazardous waste disposal, odour control, groundwater contamination, and sludge disposal site location, in addition to other factors in recent years have had a substantial impact on the viability of aerobic treatment of wastewaters. In future for a growing number of wastewaters the best choice must be the most environmentally desired choice. With the amount of available applied research now published anaerobic biotechnology may well become the solution.

Operational problems often plaguing a sludge process (such as chronic bulking with attendant secondary clarification and thickening facility) coupled with greater familiarity with anaerobic processes, should influence many future decision makers to explore the possibility

of anaerobic treatment. A thorough knowledge of anaerobic principles and applied research will make feasible intelligent replacements for any aerobic processes, which continue to present ongoing operational and disposal problems (Speece, 1996).

1.10 ANAEROBIC PROCESS DESIGNS

Anaerobic technology can be divided up into two broad categories: Flow-through systems and Retained Biomass systems.

1.10.1 FLOW-THROUGH SYSTEMS

The septic tank and Imhoff tank are examples of early flow-through systems (Figure 1.4). In both these systems the sewage flows through the system in the upper part, while the anaerobic sludge rests at the bottom of the tank. The settleable solids present in the sewage will sediment and are degraded by the anaerobic sludge. In later developments of the Imhoff tanks the accumulated solids are conveyed to a heated digester, thus increasing the rate of anaerobic digestion. A later version of this system is the 'Standard-rate' anaerobic digester, which is still commonly used today (Figure 1.5 A.). In the standard-rate digestion process, the contents of the digester is usually unheated and unmixed. Retention times vary from 30-60 days.

It was discovered in the 1950's that mixing overcame problems such as the formation of a thick scum layer. Not only did it reduce the scum layer but also enhanced the rate of digestion by bringing the bacteria into closer contact with the organic waste (McCarty, 1982, Killilea, 1992). These mixed digesters were called Continuously Stirred Tank Reactors (CSTR's) and

the process is termed 'High- rate' digestion (Figure 1.5 B). In a high rate digestion process the contents of the digester are heated and mixed completely (Metcalf *et al.*, 1991).

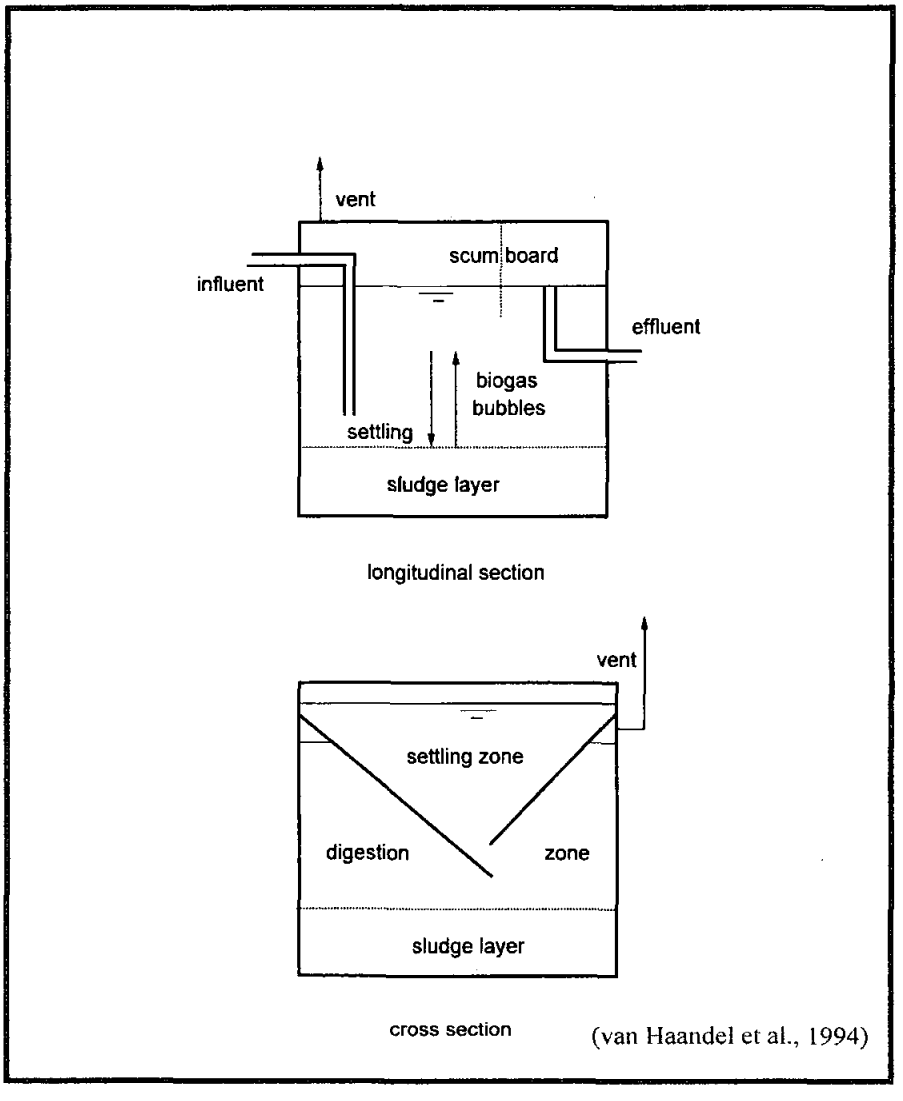


FIGURE 1.4 **EARLY ANAEROBIC TREATMENT PROCESSES: SEPTIC TANK (UPPER) AND IMHOFF TANK (LOWER)**

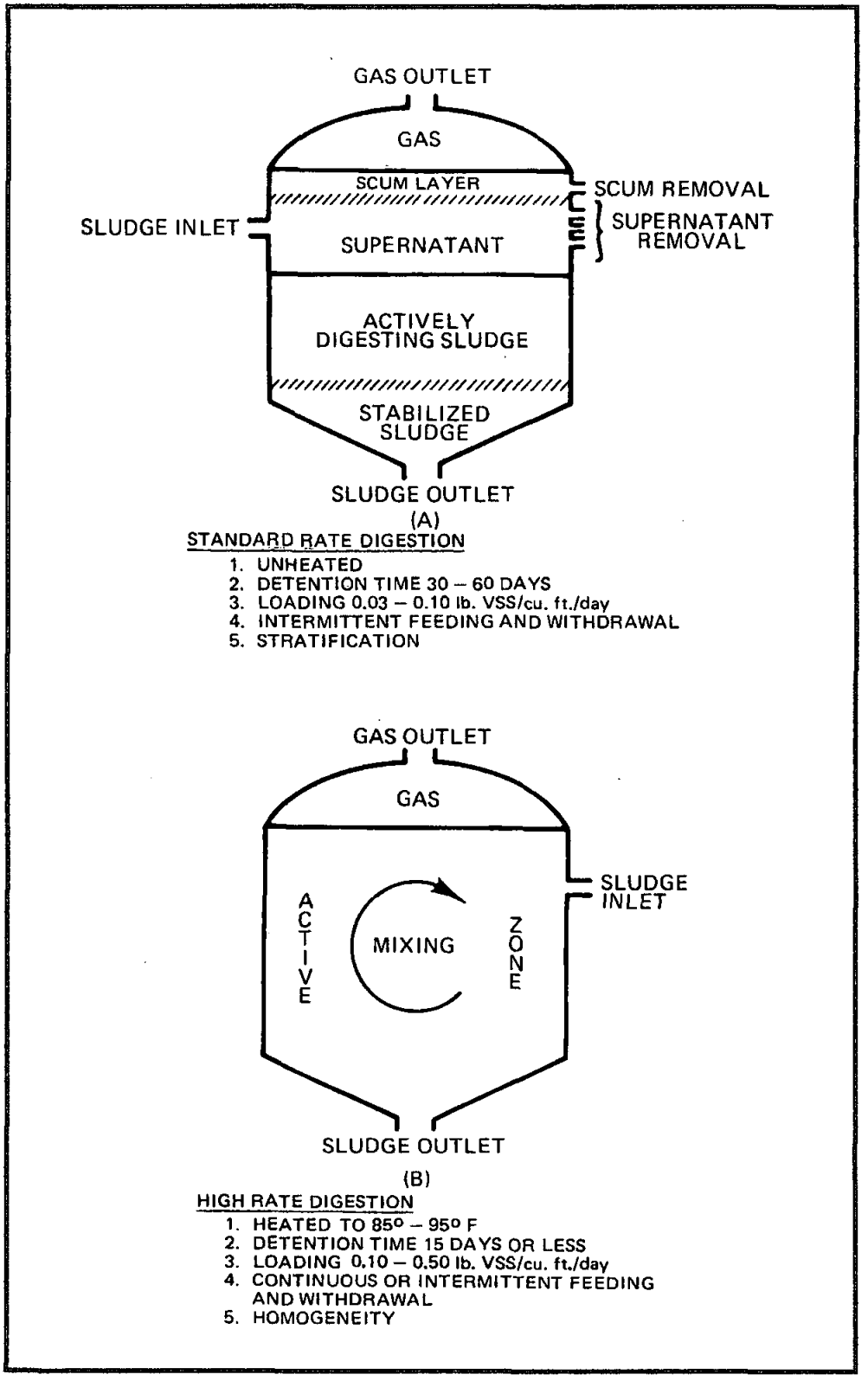


FIGURE 1.5 (A) STANDARD-RATE AND (B) HIGH-RATE DIGESTION (SINGLE-STAGE)

Both single stage reactors and two-stage processes exist, the latter being more predominant in Britain. The two-stage process is a combination of standard rate and high rate processes (Figure 1.6). The primary function of the second stage is to separate the digested solids from the supernatant liquor, however, additional digestion and gas production may occur (Metcalf & Eddy, 1991).

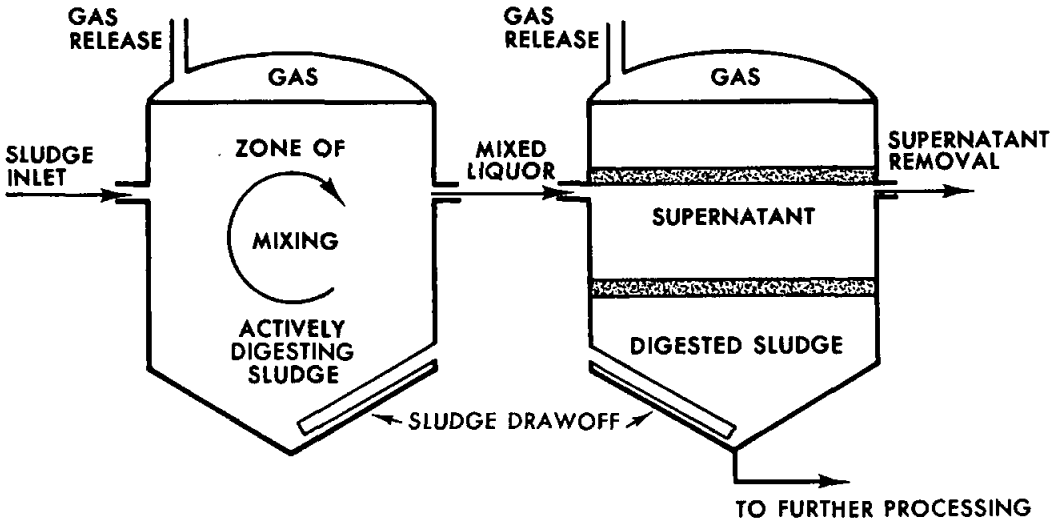


FIGURE 1.6 TWO-STAGE ANAEROBIC DIGESTION

CSTRs are used to treat sludge produced in municipal sewage treatment plants throughout Europe. At Mogden sewage treatment works, west London, the domestic effluent from 1.4 million inhabitants is treated. The sludge produced is digested in 20 CSTRs having a total installed capacity of 70,000 m^3 and a HRT of 20 days (Buvet, 1986, Killilea, 1992). However, in Ireland unheated digestion processes such as the Imhoff tank were the only anaerobic systems in operation in municipal sewage treatment works until 1987 (Finnegan, 1990,

Killilea, 1992). In this year a heated, gas mixed digester at Tullamore Co. Offaly was constructed. Since then four more of this type have been constructed around the Country and many more are planned. These plants will be discussed later.

Another of the flow-through reactor designs is the Plug Flow digester. A major advantage of this system is that it allows simple tank configuration, resulting in significant savings in capital cost. This type of digester has been in operation successfully since 1986 at Bethlehem Cistercian Abbey, Portlenone, in Northern Ireland (Killilea, 1992).

1.10.2 RETAINED BIOMASS SYSTEMS

In CSTRs the solids retention time (STR) is equal to the hydraulic retention time (HRT). As breakdown of long chain polymers in the sludge can take several days the HRT is usually in the order of 15 days. In addition, due to the slow growth of the methanogenic and syntrophic bacteria, reductions of the HRT in CSTRs risks causing washout of the active biomass and the consequent process failure. (Colleran, 1991). However, use of such lengthy retention times for the treatment of high volume, low strength industrial and agri-industrial wastewater is clearly impractical (Switzenbaum, 1983; Killilea, 1992). Retained biomass reactors overcome this problem by maintaining the microbial population within the reactor independent of waste flow. This has resulted in the successful operation of reactions at low retention times.

Early examples of such systems would be:

1. The Contact Digester (Figure 1.7)
2. The Anaerobic Filter (Figure 1.8 A)

And later developments include

3. The Upflow Anaerobic Sludge Bed Reactor (UASB) (Figure 1.8 B)
4. The Fluidised/Expanded Bed Reactor (Figure 1.8 C), and
5. Hybrid/Anhybrid Reactor (Figure 1.8 D).

A number of these systems successfully operate in Ireland treating industrial wastewater. An example of this would be an Anaerobic Filter operating at ADM, Co. Cork. However, these types of systems are much more popular throughout Europe. Retained biomass systems would not be appropriate for sewage solid treatment as long retention times are required for the breakdown of long chain organic molecules that are present in sewage sludge solids. Retained biomass reactors are a very viable alternative for short chain high strength organic wastewater such as brewery wastewater etc.

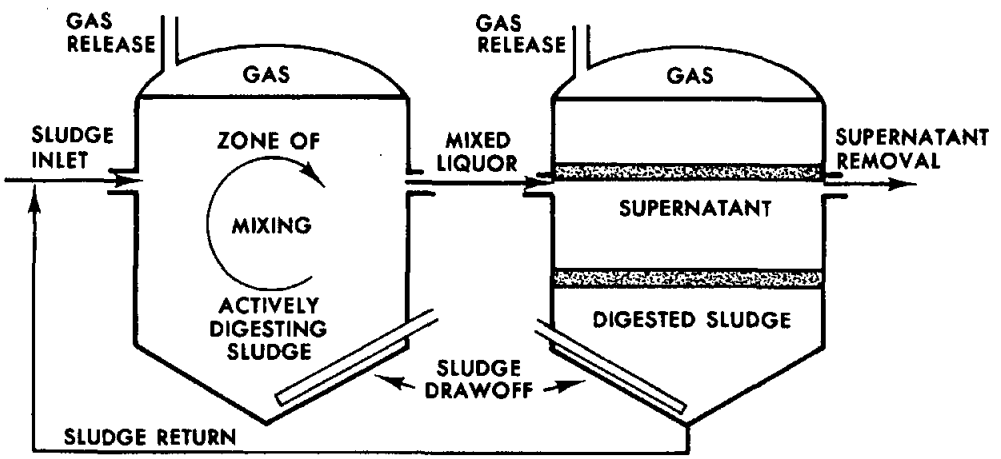


FIGURE 1.7 ANAEROBIC CONTACT DIGESTION

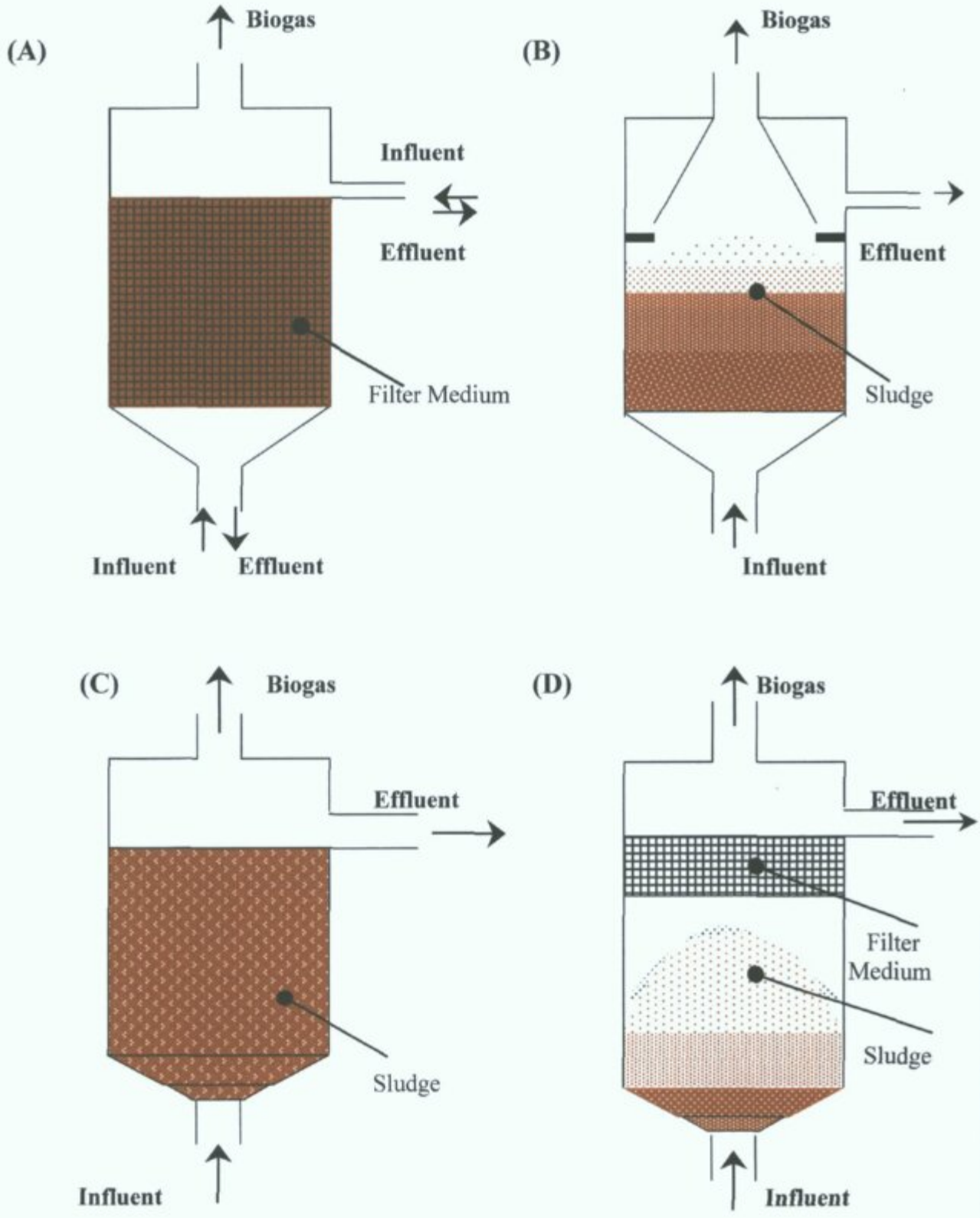


FIGURE 1.8 (A) ANAEROBIC FILTER; (B) UASB REACTOR; (C) FLUIDIZED/EXPANDED BED REACTOR AND (D) HYBRID/ANHYBRID REACTOR

1.10.3 New and Broader Applications of Anaerobic Digestion

From Table 1.5 it can be seen that the range of waste types that can or is being treated via anaerobic digestion processes has, in the recent past, been expanding at a rapid pace due to new reactor designs (Verstraete & Vandevivere, 1999). For example, the low strength wastewater can now be treated, even under psychrophilic conditions, by using specific hydraulic conditions in the expanded granular sludge bed (EGSB) reactor (Kato, 1994; Rebac *et al.*, 1994; Reynolds *et al.*, 1995; van Lier *et al.*, 1996). Solids wastes are treated anaerobically with the thermophilic 'high-solids' fermentation technology (Baeten & Verstraete, 1993). New reactor designs permit S° recovery from SO_2- rich waste gases (Buisman, 1996). Another significant trend of anaerobic digestion technologies is higher treatment efficiency. Higher efficiency is made possible by adequate pre- or post treatments and by various types of additives or co-substrates that improve sludge retention, composition, metabolic diversity, or resistance toward toxicants (Thaveesri *et al.*, 1994; Field *et al.*, 1995; Wirtz & Daque, 1996; Kupferle *et al.*, 1995).

In 1997 Lettinga *et al.*, introduced the Staged Multi - Phase Anaerobic (SMPA) reactor. This system will be feasible for all temperature conditions, viz. From very low (<10°C) to very high (>55°C) and many very different types of wastewater, including those contained quite inhibitory compounds of a variety of heavily polluting chemical industrials. Full-scale applications of this system have not yet been reported.

TABLE 1.5 NEW AND BROADER APPLICATIONS OF THE ANAEROBIC TREATMENT OF WASTES

Current Status	Recent Developments	Challenges
Sewage		
UASB reactors in warm climates	Sequential anaerobic-aerobic	Nutrient removal
Limitations are	(e.g., USB - trickling filter	Odour Removal
VSS/COD > 0.1	Psychrophilic conditions	Pathogen removal
High K_m, T , 18°C	(e.g., EGSB reactor)	Profit opportunities
Little N, P removal	Nigh pulse of ANS	
Industrial Wastewater		
UASB reactors	Xenobiotics removal	Marketable by-products
Limitations are	Granulation-enhancing additives	Engineered sludge
Toxicity and lag phase	Toxicity control (GAC)	Nitrate-driven
Floatation	Desulfurization	desulfurization
Rapidly acidified COD/COD		
High-sulphate effluents		
Slurries		
Fully mixed reactors	Thermophilic digesters	Nutrient removal
Limitations are	ADUF reactor	
Long retention times		
Pathogens removal		
Remobilization of N, P		
Municipal Solid Waste		
'High-solids' anaerobic digesters	Co-digestion	Biochemical 'rerouting'
Limitations are	Adaptation to developing	Inter-regional cycling
Low-value end-product	countries	High-value end-products
Non-biocompatible wastes		
Excess water		

(Verstraete *et al.*, 1999)

1.11 ANAEROBIC DIGESTION IN EUROPE

1.11.1 HISTORY

The first application of anaerobic digestion for sewage sludge treatment is presumably the airtight chamber developed by the end of the 19th century in France by M. Mouras. In this so-called 'Mouras automatic scavenger' settleable solids from sewage are 'liquefied' (McCarty, 1982). By the early twentieth century, several new anaerobic treatment systems were developed, e.g. the septic tank (Figure 1.4) by Cameron in England and the Imhoff tank (Figure 1.4) by Imhoff in Germany. Primary treatment of sewage by anaerobic digestion found widespread application in the years between the World Wars. More than 12 million people in Germany were served by anaerobic treatment systems, mostly versions of the Imhoff tank. In many cases the biogas produced from sewage digestion was purified and compressed and used as a fuel (van Haandel *et al.*, 1994).

1.11.2 CURRENT SITUATION IN ANAEROBIC DIGESTION

1.11.2.1 Denmark

For over 70 years anaerobic digestion has been used for the stabilisation of sewage sludge in Denmark. In 1980, the concept of farm-scale biogas plants (anaerobic digestion plants) was introduced. Most of them co-digest animal manure and organic waste. The concept of centralised biogas plants has been developed in Denmark since 1987. At present 20 plants are operating, with capacities ranging from 50 to 599 tonnes biomass feedstock per day. Approximately 80% manure, mainly slurries, is co-digested with 20% organic wastes from

abattoirs, other food industries and municipalities. A few plants also co-digest sewage sludge or the organic fraction of source-separated household waste. Table 1.6 shows the total number of anaerobic digestion plants in operation in Denmark treating the different wastes generated.

TABLE 1.6 ANAEROBIC DIGESTION PLANTS IN 1996

Type of Biogas Plant	Denmark	Austria	Sweden
	No. of Plants	No. of Plants	No. of Plants
Wastewater treatment plants	64	88	134
Landfill plants	10	31	59
Industrial waste treatment plants	5	20	8
Centralised biogas plants, co-digestion	20	0	0
MSW treatment plants	-	3	4
Farm scale plants	20	50	6
Total	**119**	**192**	**211**

(Braun, *et al.*, 1999).

1.11.2.2 Austria

Anaerobic digestion has a long tradition in Austria with 88 municipal sewage sludge digesters installed. Since 1978, 50 farm-scale are in operation. Additionally, anaerobic waste pre-treatment plants are implemented in more than 20 agro-, food-, pharma-, and paper- industry factories. Recently anaerobic digestion was also introduced for treatment of biowaste from source-separated collection of municipal solids waste. Biogas is also collected from 31 major landfill sites (Braun, *et al.*, 1999).

1.11.2.3 Switzerland

Biogas production in Switzerland has a relatively long tradition in wastewater treatment. The first anaerobic digesters were built in the thirties for the stabilisation of sewage sludge. Currently there are approximately 100 plants in operation treating sewage sludge, 140 plants treating agricultural manure (farm-scale) and 3 centralised anaerobic digestion plants.

1.11.2.4 United Kingdom

Around 70% of the UK sewage sludge is treated by heated anaerobic digestion. The first scheme was initiated in Birmingham in 1923. Now anaerobic digestion is carried out at around 240 sewage treatment works in the UK, treating some 760,000 tds annually (Meadows, 1993). About 45 farm-scale digesters have been installed in the UK since 1975. Of the 45 units installed only about 25 are currently operating. These farm-scale digesters have suffered several problems due to inadequate designs and lack of operator training. Very few digesters have been installed in the last few years since the removal of grant funding. Recent interest has focused on larger centralised schemes due to support available from the Non-Fossil Fuel Obligation (NFFO). To date seven centralised anaerobic digesters have received NFFO contracts. All of these contracts are in development and none have yet proceeded to construction (Higham, 1999). The UK also has a small proportion of successful industrial anaerobic digestion plants (approximately 15 in 1990) treating industrial wastes (Wheatley, 1990).

For over a half a century anaerobic digestion of different wastes, in Europe, has been well established, tried and tested in Countries discussed above and in others such as Finland, Italy, The Netherlands, Portugal, Spain, Germany and France. Sewage sludge was the pioneer waste to be successfully treated by anaerobic digestion.

1.12 ANAEROBIC DIGESTION IN IRELAND

1.12.1 SEWAGE SLUDGE DIGESTION

Anaerobic digestion has existed for many years in Ireland in the form of Septic tanks and Imhoff Tanks for the treatment of sewage. However, high-rate, heated and mixed mesophilic anaerobic digestion is a relatively new arrival to Ireland for the stabilisation of sewage sludge. In 1987, Tullamore, Co. Offaly, was the Countries' first pioneer in this area when anaerobic digestion was introduced in the form of a Continuously Stirred Tank Reactor (CSTR) facility, to handle the Town's sewage sludge. Tullamore was shortly followed by Buncrana, Co. Donegal in 1991.

Then in 1992 the Department of the Environment in Ireland commissioned a National Sludge Strategy Study for 25 counties of the Irish republic (excluding Dublin). This report was to guide the department on policy and practice in relation to the treatment and re-use of sewage sludge (see section 1.1.2). The resulting study recommended a further 23 viable sites for the anaerobic digestion of sludge (Figure 1.9). Local authorities were requested to implement the first recommendation of the Strategy Study which required them "to prepare a detailed sludge management plan for each region within the framework of the Strategy Report but based on a more comprehensive database than that on which the Strategy Study was based".

Due to the recommendation of the above report and further strengthened by the requirements of the Urban Wastewater Directive 1991, local authorities begun preparing sludge management plans for their regions. As a subsequent reaction of the above pressures, anaerobic digestion of sewage sludge was introduced in Greystones, Co. Wicklow (1996) and

more recently digestion plants were constructed in Clonmel, Co. Tipperary and Tralee, Co. Kerry (Figure 1.10). Both of these plants were commissioned in 1999. These anaerobic digestion plants will be discussed in detail in Chapter 3. Currently (2000), there are four anaerobic digestion sites under construction in Dundalk, Drogheda, Roscrea and Galway (Figure 1.10). Many more anaerobic digestion sites are at design stage and it is estimated that approximately twenty anaerobic digestion plants treating sewage solids will be in operation by the year 2010.

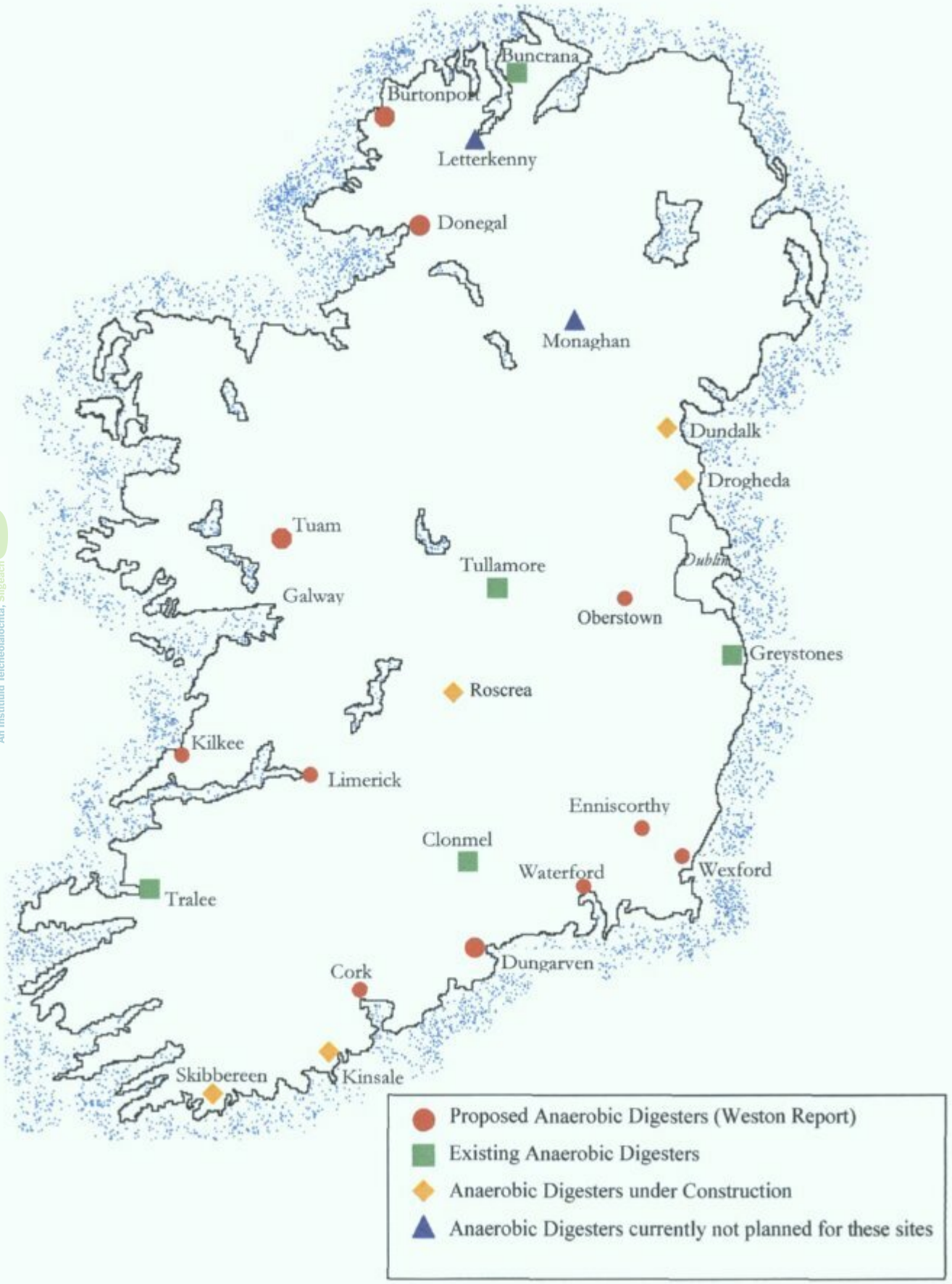
However, not all regions incorporated the recommendations of the Weston Strategy Study. After the preparation of the sludge management plans in Monaghan and Letterkenny both towns decided not to included anaerobic digestion for the treatment of sewage sludge. Monaghan's decision was purely based on economics and Letterkenny decisions may have been influenced by the fact that the digester in Buncrana (which is in the same county as Letterkenny) has not been very successful to-date (Chapter 3).

1.12.2 INDUSTRIAL AND FARM-SCALE DIGESTERS

There are currently three on-farm digesters located in Ireland (Co. Waterford, Co. Wexford and Co. Kilkenny). All of these plants were built in the last 10 years and operate at mesophilic temperatures with digester sizes ranging from $144m^3$ to $490m^3$ (Guest, 1999). There are also several industrial anaerobic treatment processes in Ireland treating agri-food waste. Examples include ADM, Co. Cork and Kerry Foods, Co. Kerry.

FIGURE 1.9 PROPOSED AND EXISTING ANAEROBIC DIGESTION SITES IN IRELAND (1993)

FIGURE 1.10 PROPOSED AND EXISTING ANAEROBIC DIGESTION SITES IN IRELAND (2000)



1.13 AIMS AND OBJECTIVES OF THIS STUDY

The aims and objectives of this current study are as follows:

- To describe in detail the current state-of-the-art anaerobic digestion technology for sewage treatment in Ireland;

- To access the success of this technology in terms of process stability, solids removal and biogas production;

- To ascertain the most appropriate methods for monitoring process performance and highlighting process instability by both plant operators at full-scale plants and by laboratory trials;

- To investigate the effects of environmental conditions such as substrate type, oxygen shock, temperature and mixing on process performance using pilot trials.

The above was achieved by extensive on-site investigations of five full-scale plants throughout Ireland and a 144-day pilot trial using three laboratory scale anaerobic digesters.

CHAPTER 2

MATERIALS AND METHODS

2.1 INTRODUCTION

The methods of research employed in this project involved:

- Field investigations of existing full scale anaerobic digestion plants
- Laboratory scale testing of the anaerobic digestion process

2.1.1 FIELD INVESTIGATIONS OF EXISTING ANAEROBIC DIGESTION PLANTS

All existing anaerobic digestion sites were visited on at least three separate occasions. The first visit was of an introductory nature in order to become acquainted with plant personnel and a detailed description of each plant was drawn up. A database of process, mechanical and electrical installations was recorded. Results on plant performance were reviewed (solids concentrations, gas production etc.). Data collection forms/questionnaire were then compiled based on Comprehensive Performance Evaluation data forms given in the US EPA Handbook on "Retrofitting Publicly Owned Treatment Works (POTWs)" (1989). These forms were modified to yield more information on the design and performance monitoring of anaerobic digesters. For example, a ranking system was devised to highlight the plant operators' perception of what are the most importance parameters requiring attention for optimum process performance (with "1" meaning of "little or no importance" and "5" meaning "essential").

2.1.2 LABORATORY TESTING OF THE ANAEROBIC DIGESTION PROCESS

The laboratory scale study involved the use of three 5-litre anaerobic digestion plants and an activated sludge (AS) plant. The AS plant was commissioned 36 days prior to the anaerobic digesters. This was used in order to generate enough sludge to feed the anaerobic digesters once commissioned. The main reason for using an AS plant to generate sludge was to mimic the operation at four out of the five sewage treatment plants visited where waste activated sludge system was a feed for the anaerobic digestion process. The pilot study was to be modelled as near as possible to a full-scale sewage treatment works that had incorporated the anaerobic digestion process.

The whole trial was carried out in a laboratory at the Institute of Technology, Sligo. Both the anaerobic digesters and the activated sludge plant were on site and had previously been used in other studies.

2.2 DESCRIPTION OF ACTIVATED SLUDGE (AS) PLANT

The equipment used to imitate a full-scale AS process was the "Biocontrol Mark 2" system. The main components making up the Biocontrol process are:

- A control unit (providing control of air supply and feed/sludge pumping)
- An aeration tank (where the mixed liquor suspended solids digested a whey and nitrogen feed)
- A settlement tank (where suspended solids were separated from supernatant and returned to the aeration tank).

Figure 2.1 shows a schematic layout of the principle components used in the AS process and Plate 2.1 shows a photograph of two 'Biocontrol Mark-2' AS plants in operation (only one plant was used for this study).

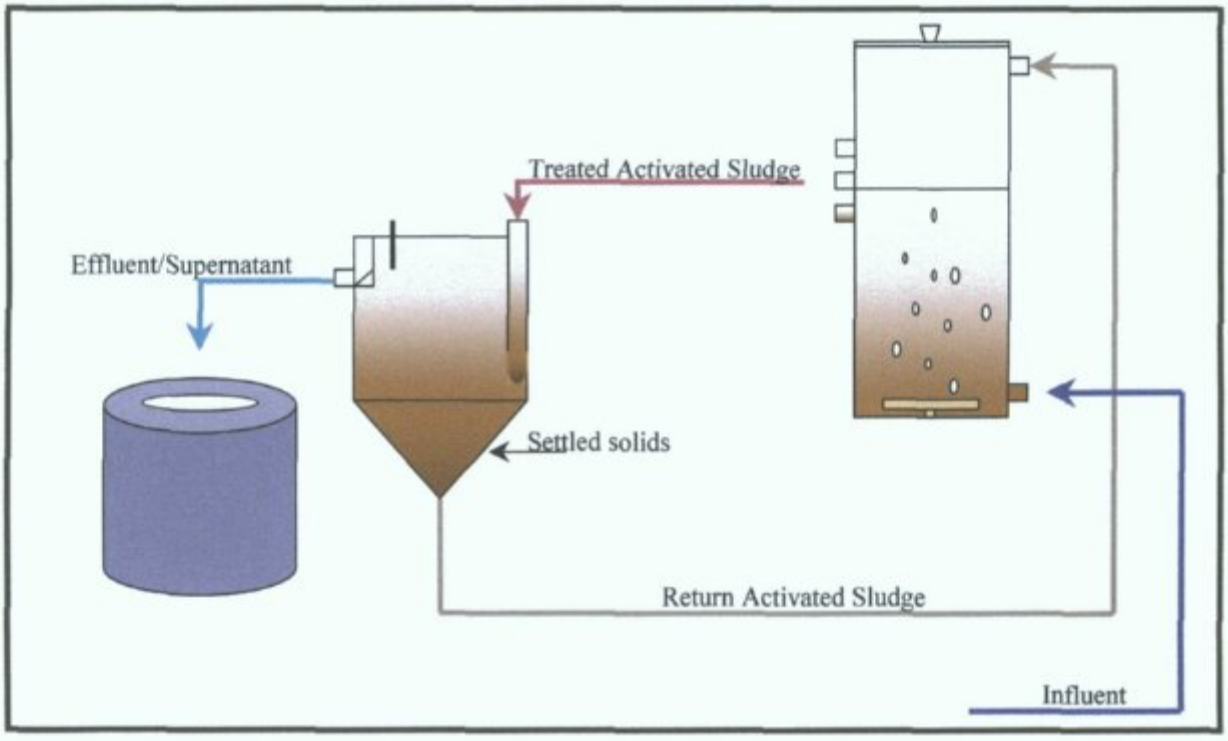


FIGURE 2.1 SCHEMATIC OF THE PRINCIPLE COMPONENTS USED IN THE AS PROCESS

PLATE 2.1 TWO 'BIOCONTROL - MARK 2' AS PLANTS IN OPERATION

2.3 SEEDING THE ACTIVATED SLUDGE PLANT

The pilot scale activated sludge plant was seeded with activated sludge from a local sewage treatment plant at Grange, County Sligo. This treatment plant treats domestic sewage for a population equivalent of 750, approximately. Ten litres of activated sludge was collected from Grange sewage treatment plant. The container was allowed to settle and then the supernatant was decanted into a sewer. Three litres of the settled activated sludge was then poured into the aeration tank.

2.3.1 SUBSTRATES TO ACTIVATED SLUDGE PLANT

The activated sludge plant was fed with the substrate, spray dried cheese whey. North Kerry Milk Products under the brand name "Actopro" manufactured this product. The nutritional characteristics of the whey power are given in Table 2.1. The initial feed rate was calculated by using the food to microorganisms (f/m) ratio of 0.05-0.15. The quantity of microorganisms present was obtained by calculating the Mixed Liquor Suspended Solids (MLSS) content of the seed activated sludge. Based on this MLSS and the ratio above, the concentration of feed (gCOD/d) was calculated for any one day. Once the correct biomass quantity had been established an f/m ratio of up to 1 was used.

If a biological system is to function properly, nutrients must be available in adequate amounts (Metcalf & Eddy, 1991). Examination of the whey's composition (Table 2.1) indicated a deficiency in nutrients necessary for the AS process. The principle nutrients are nitrogen and phosphorus. Based on an average composition of cell tissue, approximately 12.4 % by weight

of nitrogen will be required and phosphorus requirement is usually assumed to be one-fifth of this value (Metcalf & Eddy, 1991). However, the COD:N ratio used here was 100:10. No additional phosphorus was added because it was estimated that sufficient quantities were available in the whey feed (Table 2.1). Laboratory grade ammonia solution was used as the nitrogen source.

When the required quantities of the whey and ammonia were calculated and measured out, they were brought up to the required volume with tap water. The substrate dissolved in water was placed in a large bucket and fed to the aeration basin intermittently throughout the day.

TABLE 2.1 CHARACTERISTICS OF "ACTOPRO" SPRAY DRIED CHEESE WHEY

Parameters	Composition (%)
Moisture	4.5
Total acidity	0.2
Chemical Composition	
Protein	11.5
Fat	1.2
Mineral substances	
(including trace amounts of Ca, Mg, K, P , Na)	8.4
Lactose	74.2

2.4 ANALYTICAL METHODS FOR THE ACTIVATED SLUDGE PLANT

2.4.1 DISSOLVED OXYGEN DETERMINATION

Aerobic biological activity is independent of dissolved oxygen above a minimum critical value. Below this concentration, the metabolism of microorganisms is limited by reduced oxygen supply. Critical concentrations reported for various systems range from 0.2 to 2.0 mg/l. (Hammer *et al.*, 1996).

The DO of the activated sludge was determined by using an industrial DO_2 meter, Model 9090. The Model 9090 Dissolved Oxygen Meter is a main operated bench mounted unit. The unit is housed in a waterproof casing. Dissolved oxygen readings can be displayed in mg/l or percentage (%), with automatic temperature compensation over the range 0 to 40%.

Depending on this result the air pump on the Biocontrol unit would be adjusted to keep the DO value at around 2 mg/l, the optimum for growth of microorganisms.

2.4.2 MEASUREMENT OF OTHER PARAMETERS

Mixed liquor suspended solids and pH were measured using Standard Methods for the Examination of Water and Wastewater (1995) as outlined for anaerobic digestion analyses.

Suspended solids are the residual left on filter paper after sample filtration. The determination is expressed as mg/l in accordance with section 2540-D of the Standard Methods. The method used for pH measurement was Standard Method No. 4500-H+B.

2.5 DESCRIPTION OF THE ANAEROBIC DIGESTION (AD) PLANT

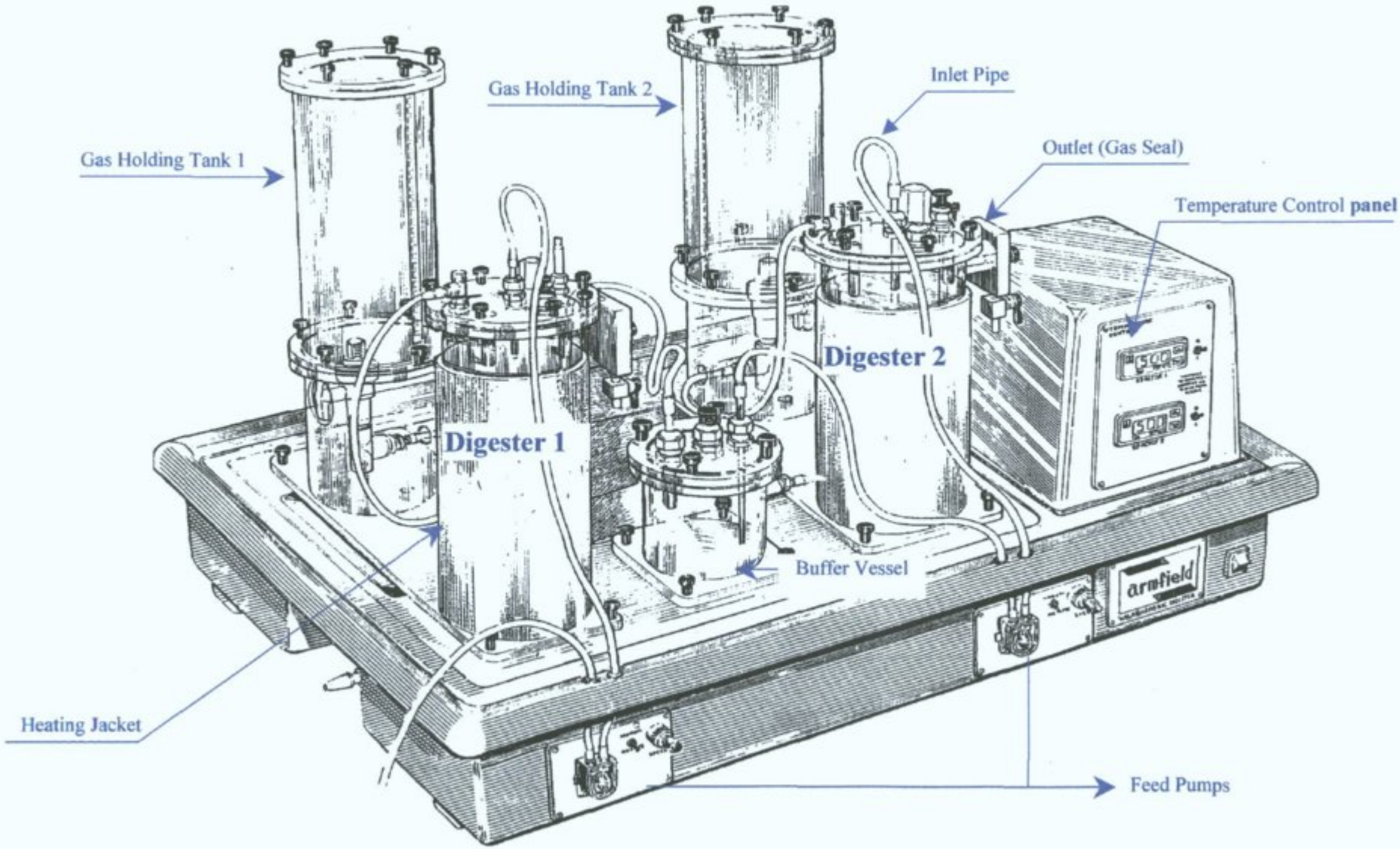
Two bench-mounted pilot anaerobic digesters were purchased from Armfield Ltd, England. These Armfield Ltd Anaerobic Digesters (W8) are small scale demonstration units which have been designed to allow thorough investigation into this biological process.

Each unit contains two anaerobic digesters. Each digester has a total liquid volume of 5.0 litres with accompanying gas collection tanks. Both vessels were constructed from 150mm diameter acrylic tubing and mounted on a fabricated base which housed both overflow and control facilities. Originally, the reactor design by Armfield Ltd was constructed to function as a two-phase anaerobic process, with the first and second reactors representing the hydrolytic and methanogenic phase of the process, respectively (Figure 2.2). Operation, with significant differences in HRT and buffering requirements, was made possible by the incorporation of a two-litre buffering vessel between reactors. However during this study the reactors were modified to suit the mesophilic single-phase anaerobic digestion process.

2.5.1 DESCRIPTION OF ORIGINAL TWO-STAGE ANAEROBIC DIGESTION PROCESS

Figure 2.2 shows the layout of the original two-stage anaerobic digestion process by Armfield Ltd. Each reactor (1 and 2) has a total liquid volume of 5-litres. The feed or influent is pumped by variable peristaltic pumps through a central pipe having an output near the base of the reactor. The liquid outlet from the reactor is through a gas seal which ensures any gas produced cannot escape and also creates a liquid take-off which is lower than the operating level of the reactor to prevent any "scum" formed on the liquid surface being taken off.

FIGURE 2.2 ORIGINAL TWO-STAGE ARMFIELD ANAEROBIC DIGESTION PROCESS



Liquor leaving the first reactor enters a buffer vessel. This vessel allows the first reactor to be operated at a higher throughput than the second reactor, the excess being taken off through overflow. Feed to the second reactor is by variable speed pump in an exactly similar way to reactor 1 through a central pipe and leaving through a gas seal.

Any gas produced by the reaction taking place in reactor 1 or 2 is collected in 5 litre calibrated vessels. The gas collection is by water displacement through a constant head device, which also creates a liquid seal between the gas tank and reactor. There is a water overflow chamber situated in the middle of the base of the unit, which collects the displaced water as the gas is being produced.

Each reactor is heated to its operating temperature by electrical heating mats (jackets). Temperature sensors transmit the reactor contents temperature to a 3-term controller, which automatically adjust the electrical power to the mats to maintain the desired operating temperature at a constant level. The maximum operating temperature of both reactors is 55°C.

The reactors were suitable for filter configuration with the addition of plastic media. This reactor configuration was used in a previous study (up-flow anaerobic filters).

This two-stage process described above was modified during this study in order to run three single stage anaerobic digesters. This process and the modifications are described next.

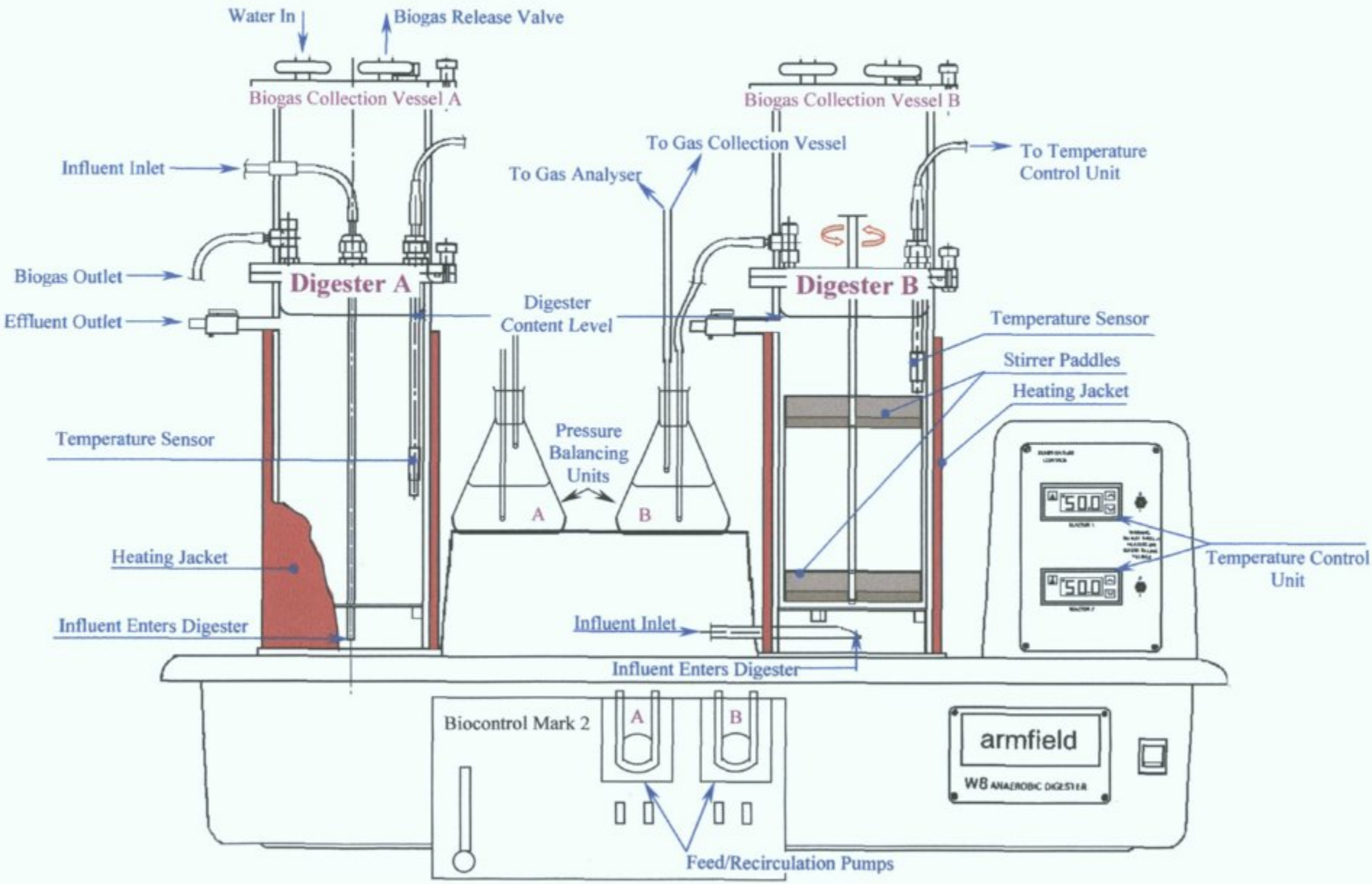
2.5.2 DESCRIPTION OF THE MODIFIED ANAEROBIC DIGESTERS

The two Armfield units (4 reactors) were used, however only 3 of the digesters were set-up, the fourth was used for spare parts, which proved to be vital in the maintenance of the other digesters. Figure 2.3 shows the modifications carried out to the reactors. Reactor (A) in the figure represents the alterations carried out to reactors 1 and 2 and reactor (B) represented alternations carried out to reactor 3.

The Armfield units were used in a pilot trial previously in another research project. Modifications were carried out on the pilot digesters to cater for a different reactor configuration. Several design problems were experienced while operating these trials units. To learn from these problems encountered and improve the design, modifications were made to the reactors. Modifications were also made to suit the single-stage anaerobic digestion process. The description of the three single-stage anaerobic digesters is the same as above with the following modifications:

I. U tube outlet: These U tubes were meant to act as a gas seal to ensure that no gas escaped when the effluent was exiting the digester. However, this system never worked, the U tube was constantly getting blocked with sludge. It was decided to remove these tubes in all three of the digesters and replace them with simple valves. To ensure gas did not escape through these valves while the effluent was leaving the digester, the sludge level was to be kept above the outlet at all times.

FIGURE 2.3 MODIFIED ANAEROBIC DIGESTERS



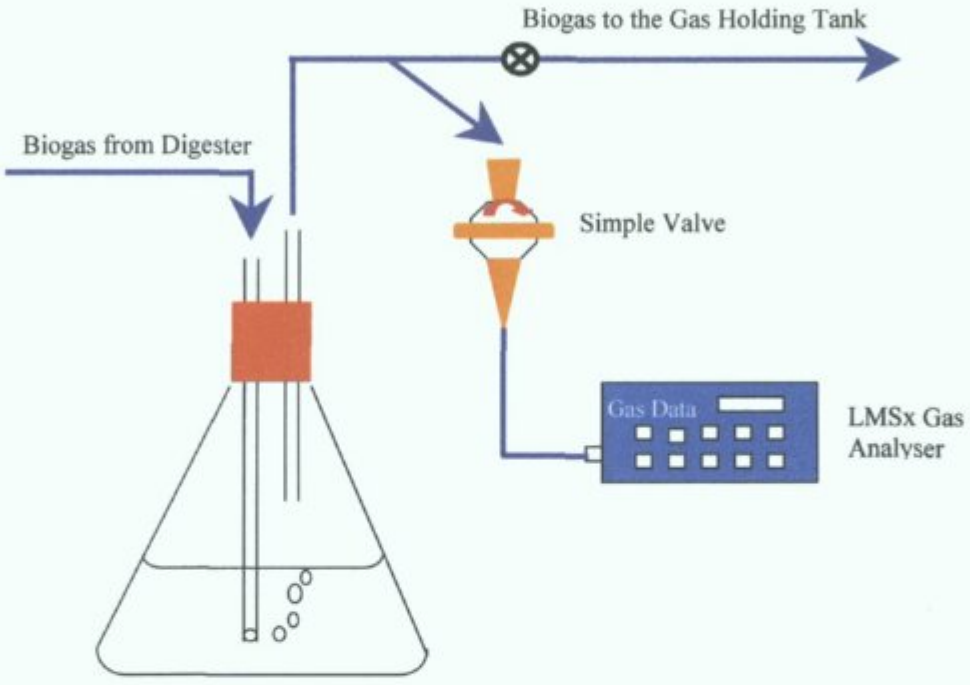
II. The gas measurement system: A gas analyser was purchased to enable the measurement of the biogas composition. As the reactors were not adapted for this type of analyser it was necessary to tap into the biogas system in order to measure the biogas composition. A tap was inserted along the biogas line from the digester to the gas collection vessel (Figure 2.4). However, this idea proved to be unsuccessful because when the gas analyser was in use, it sucked biogas out of the digester, thus causing negative pressure inside the digester. This negative pressure was counteracted by sucked water from the gas holding tank into the digester. This was an undesirable condition for the digestion process, as this water would eventually fill the digester and it would also water down the digester contents.

In order to solve this problem a pressure-balancing device was designed, constructed and installed. It consisted of a conical flask filled half way with water and a rubber bung sealing the flask. The rubber bung held in place two lengths of glass tubing (Figure 2.4). The gas pipe from the digester to the gas holding tank was cut in two. The pipe coming from the digester was connected up to the glass tube, which was immersed in the water contained in the flask. The gas pipe to the gas holding tank was connected to the other glass tube in the bung. This tubing was not immersed in the water. The tap to the gas analyser was on this pipeline, from the flask to the gas holding tank. This layout can be seen in Figure 2.4.

Using this apparatus when the gas analyser was switched on the biogas would be sucked out of the space in the flask and subsequently the digester, depending on the amount of biogas present in the system. In this way the negative pressure occurred in the flask and the water would displace back into the flask not the digester, unless there

was a very large quantity of biogas drawn from the system. This proved to be a successful way of taking the biogas from the gas system without disturbing the solids to biogas equilibrium in the digester.

FIGURE 2.4 APPARATUS FOR GAS MEASUREMENT



III. Mechanical mixing of Reactor 3: Variable-speed-stirring device was installed in reactor 3. This was placed down the centre of the digester (Digester B, Fig. 2.3). The stirrer had an upper and lower paddle, which were both coated in varnish to prevent corrosion.

IV. The feeding system: The feed tubing and the feed pumps were only suitable for solids less than 2 %. As the solids composition of the sludge is considerably larger than this (2-10%) the existing feeding system was deemed unsuitable. All new tubing

was purchased with an internal diameter of 8 mm. The plastic central inlet tube entering at the top and going down through the length of the digester was replaced with bigger steel tubes coated with varnish to prevent corrosion. This was carried out for reactor 1 and 2. In reactor 3 the inlet pipe was positioned at the bottom of the reactor because a mechanical stirrer was situated down the middle of this reactor.

V. Recirculation: This was achieved by connecting the inlet and outlet piping to the pumps used for feeding. These pumps were from the Biocontrol Mark 2 apparatus (Section 2.2) from the activated sludge system. The pumps were set to come on for 15 minutes every 30 minutes. The recirculation was not introduced to Reactor 3 until the second Biocontrol unit was available i.e. when the on site AS plant was shutdown. The same pumps carried out the recirculation and feeding. When feed was to take place the recirculation would be switched off and the feed was administered through the recirculation line.

VI. Removal of the buffer vessel: The buffer vessel, which was the all-important link in the two-stage digestion process, was omitted in the current operation. In this single-stage process each of the digesters were feed separately and both the hydrolytic and the methanogenic phases took place in the same vessel for each digester.

VII. The gas holding tanks were modified with the addition of gas valves and tubing on the top of the tank. This allowed controlled addition of water to the gas measuring vessels and purging of the collected biogas outside of the laboratory.

These adjustments were carried out either before or during the start-up period. The process was operated at 35°C +/- 1°C (mesophilic temperatures). Thermostats placed in the reactors regulated this temperature. The influent was fed manually to each of the three reactors in through the recirculation lines. As the influent was being feed the same quantity of effluent was displaced out through the outlet. Gas generation within the digester accumulated in the space above the digester contents (approximately 0.3 litres), until sufficient pressure forced its release to the gas collection tank, which operated by water displacement.

Plate 2.2 shows the set-up of the bench mounted modified Armfield digesters with the 'Biocontrol Mark-2' pumps being utilised as the feed/recirculation pumps.

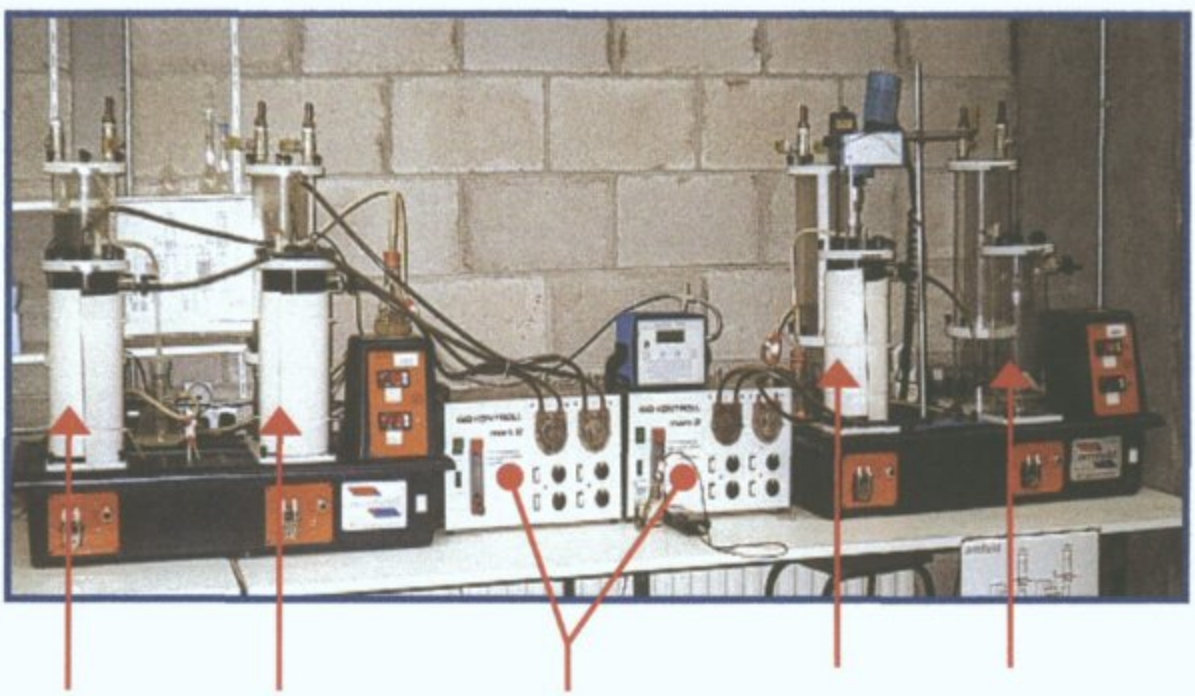


PLATE 2.2 SET-UP OF THE PILOT SCALE ANAEROBIC DIGESTION PLANT

2.6 SUBSTRATES TO ANAEROBIC DIGESTION PROCESS

The main bacterial population and the reactions of the digester will be determined by the nature of the feedstock (substrate). The feedstock of anaerobic digesters can vary in physical form as well as in chemical composition. The three substrates used during this Study were Sewage Solids, Whey and Animal (cattle) Slurry.

2.6.1 SEWAGE SOLIDS

Three types of sewage solids were added to the digesters at different times; a) settled activated sludge generated on site, b) settled activated sludge collected from sewage treatment plants around the Sligo area and c) primary sludge collected from Strandhill sewage treatment plant in Co. Sligo. Both types of activated sludge (a) and (b) had a TS concentration of between 2-6% and the TS in the primary solids were 7%. The COD concentration of the activated sludge and primary sludge ranged from 7 - 17 g/l and 14 g/l, respectively. Table 2.2 shows the approximate composition of an average sanitary wastewater.

2.6.2 WHEY

The whey used was spray-dried cheese manufactured by Kerry Foods, Co. Kerry. Composition of this whey is already discussed in see Section 2.3.1, as it was the same whey that was added to the AS plant. Whey was added to the sewage solids to supplement the organic load entering the digesters. It was also added on its own for certain periods. The composition of the whey power is described in Table 2.1.

TABLE 2.2 APPROXIMATE COMPOSITION OF AN AVERAGE SANITARY WASTEWATER BASED ON 450 L/PERSON/D

Parameters	Raw (mg/l)	After Settling (mg/l)	Biologically Treated (mg/l)
Total Solids (TS)	800	680	530
Total volatile solids (VS)	440	340	220
Suspended Solids (SS)	240	120	30
Volatile Suspended solids (VSS)	180	100	20
Biochemical Oxygen Demand (BOD)	200	130	30
Inorganic Nitrogen as N	22	22	24
Total nitrogen as N	35	30	26
Soluble phosphorus as P	4	4	4
Total phosphorus as P	7	6	5

(Hammer *et al.*, 1996)

2.6.3 ANIMAL SLURRY

The animal slurry was obtained from a slatted house, which housed dairy cows for the winter months. Their main foodstuff for the winter was silage and concentrates (cattle nuts). The animal slurry was watered down and sieved, to remove solids such as grass, which could clog pumps. The TS content entering the digester was approximately 10% and the COD value of this content was approximately 73 g/l. Tables 2.3 and 2.4 show the physical and nutrient characteristics of animal slurry in general. The volatile solids (organic matter) content varies but 70% VS of the TS is about average for the wastes.

TABLE 2.3 EXCRETA PRODUCTION FROM DAIRY COWS AND THE BIOGAS PRODUCTION FROM THE EXCRETA IN MESOPHILIC DIGESTION

Animal	Excreta/animal Day (kg)	TS in excreta (%)	Biogas (litres/kg VS)	Methane in biogas (%)
Beef heifer	21	13	220-300	55-60
Dairy Cows	45	12	200-400	55-60
Pigs	5	6-10	300-400	68-70

(Wheatley, 1990)

TABLE 2.4 AGRICULTURAL WASTE NUTRIENT VALUES

Nutrients	Kg./tonne (units/1000 gals).	
	Cattle Slurry	Pig Slurry
N	5.0 (45)	4.3 (39)
P	0.7 (6)	1.4 (13)
K	5.0 (45)	2.2 (20)

(Fertilisers Association of Ireland)

2.6.4 GENERAL CHEMICAL COMPOSITION OF ALL SUBSTRATES UTILISED

Wastes composed mainly of human and animal excreta contain all the necessary bacterial nutrients in the form of intestinal bacteria, intestinal secretions and food residues undigested in the gut. In addition domestic sewage contains kitchen wastes and material from drainage water (Hobson & Wheatley, 1993). These substrates contain compounds of carbon, nitrogen and other elements, which supply the nutrients needed by the bacteria. If in excess some compounds can be toxic and if deficient can cause slow microbial growth. Therefore, it is important that these nutrients are present in correct quantities in order for the anaerobic digestion process to operate at optimum. A typical nutrient ratio would be 1000:30:5 as COD:N: P (Ghose & Das, 1982). From the nutrient tables above it can be seen that the nutrient was present in near enough quantities to this ratio, therefore nutrients were not

supplemented to the substrates. The only cause for concern was nutrient deficiency when Whey was being added on it own. On one occasion different quantities of NH_3 was added to each reactor, however this shows no change (either negative or positive) in the digestion process, so it was decided there was no need to add it again. Whey was only added on its own for a short period.

2.6.5 INITIAL FEED TO THE DIGESTERS

Throughout the duration of the study there were three different types of sewage solids added to the digester. The first type was from the activated sludge plant on site, which was seeded with activated sludge from a sewage treatment plant. This plant was fed with whey as a source of food in order to generating solids to be feed to the anaerobic digesters (Section 2.3). The average percentage total solids of this feed entering the digesters were approximately 6%. The initial feed rate was based on a retention time of 14.1 days, which was equal to a volumetric loading rate (VLR) of 0.333 l/d. Analysis of the sludge feed revealed the COD effluent to be 7.54 g/l. This was equal to an organic loading rate (OLR) of 2.5 gCOD/d. The f/m ratio of 0.05 was used for start-up. The volatile solids of the seed anaerobic sludge were 71.8g/reactor. Based on the above f/m ratio and the VS of the seed sludge the minimum OLR should be 3.6 gCOD/l. Therefore, 1.1g of whey was added to the feed sludge to bring the 2.5 gCOD/d up to 3.6 gCOD/d. The reactor was feed manually through feed pumps once every two-day for the first stage of the trials, and once a day for the second stage when the animal slurry was introduced.

2.7 ANALYTICAL METHODS FOR THE ANAEROBIC DIGESTION PLANT

Analytical techniques used to monitor the anaerobic reactors during the study periods were taken from the 19th edition of the Standard Methods for Examination of Water and Wastewater, (APHA, AWWA, WEF., 1995).

2.7.1 BIOMASS - SOLIDS ANALYSIS

The most important physical characteristic of wastewater is its total solids content, which is composed of floating matter, settleable matter, colloidal matter, and matter in solution (Metcalf & Eddy, 1991). Operation efficiency of various treatment units is defined by solids removal, for example, sludge digestion. Analytically, the total solids content of a wastewater is defined as all matter that remains as residue upon evaporation at 103 to 105°C. Matter that has a significant vapour pressure at this temperature is lost during evaporation and is not defined as solid.

Solids can be classified further on the basis of their volatility at 550 +/- 50°C. The organic fraction will oxidise and will be driven off as gas at this temperature, and the inorganic fraction remains behind as ash (fixed solids). The volatile solids analysis is applied most commonly to wastewater sludges to measure their biological stability (Metcalf & Eddy, 1991).

Total Solids Determination

Method: Standard Methods (19th edition, 1995), No. 2540 B (Total Solids Dried at 103-105°C)

Procedure: A measured volume of sample was placed in a porcelain evaporation dish. Water was evaporated from the porcelain dish in a drying oven at approximately 2°C below boiling to prevent splattering. The evaporated sample was dried for at least 1 hour in an oven at 103°C to 105°C and cooled in a desiccator to a constant weight. The milligrams of total residue was calculated as the difference between the cooled weight of the dish and the original weight of the empty dish. Concentration of the total residue was calculated using the following equation:

$$\text{Total solids (mg/l)} = \frac{(A - B) * 1000}{\text{Sample volume (ml)}}$$

Where A = Weight of evaporation dish + Dried residual (mg)

B = Weight of evaporation dish

The determination was complete when successive weight differences over a period of hours was negligible.

Fixed and Volatile Solids Determination

Method: Standard Methods (19th edition, 1995) No. 2540 E, (Fixed and Volatile Solids Ignited at 550°C)

Procedure: The volatile fraction was determined by igniting the residue on evaporation, at 500°C +/- 50°C in a muffle furnace supplied by Carbolite Ltd, England. Dried solids are burned for 15 to 20 minutes then cooled in a dessicator. Volatile solids were calculated using the following equation:

$$\text{Volatile solids (mg/l)} = \frac{(C - B) * 1000}{\text{Sample volume (ml)}}$$

Where C = Weight of evaporation dish + Ash residual (mg)

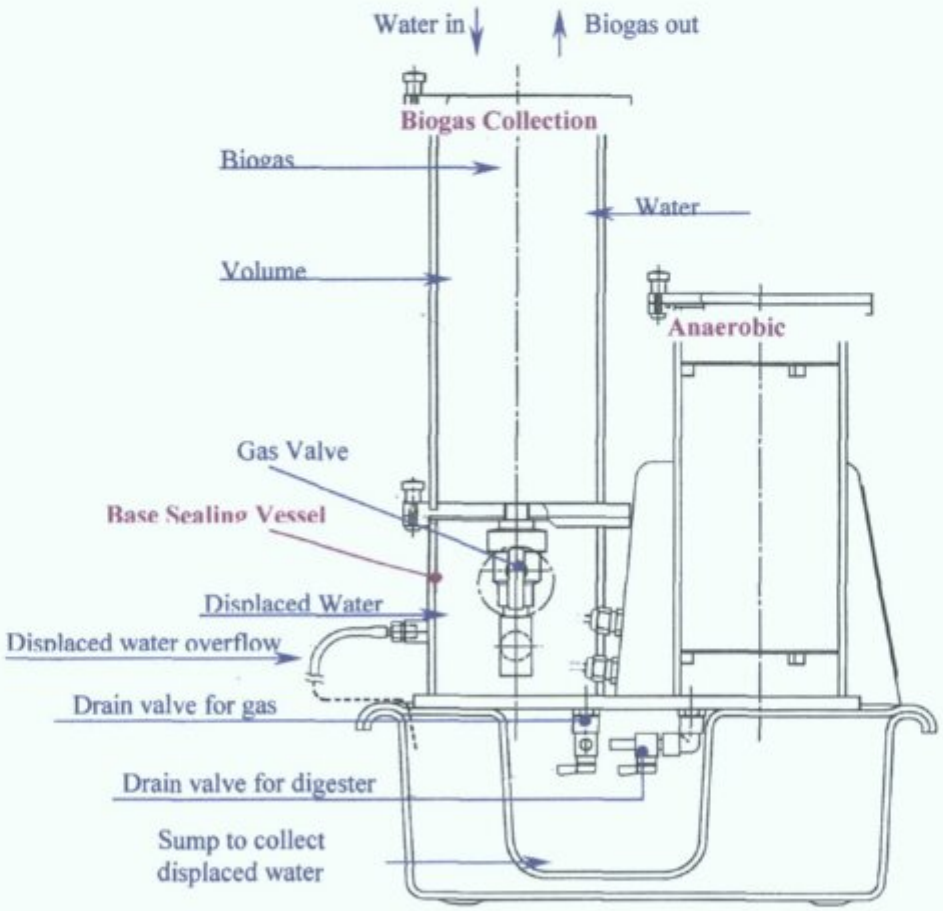
B = Weight of evaporation dish

The total and volatile solids were established on the digester contents, at the start-up and the shutdown of digestion process. At start-up the volatile solids contents were initially used to calculate the starting feed to each reactor. At the end of the trials the solids were obtained from each reactor content and compared with the solids at the onset of the study. At this time the solids content of the scum layer was also analysed and included in the shut down figure so that the correct solid removal could be calculated.

2.7.2 BIOGAS - VOLUME MEASUREMENT

The quantity of biogas produced during the bacterial digestion process was collected in the gas collection tanks using a water displacement system, which was designed to impart a small and constant back pressure to the reactor vessels. Water displaced by the gas, simply overflowed from the base of the vessel into the to the sump (Figure 2.5). This sump was large enough to cater for the 2 reactors displacing water at the same time (10-litre capacity).

FIGURE 2.5 CROSS-SECTION OF THE ARMFIELD UNIT SHOWING THE GAS COLLECTION VESSEL AND THE ANAEROBIC DIGESTER



Biogas generation was measured once a day and expressed as litres per gram of COD applied to the reactor, or alternatively as litre of gas produced per litre of reactor volume.

2.7.3 BIOGAS - COMPOSITION ANALYSIS

The quantity of Methane (CH_4) produced in the biogas, is one of the monitoring parameters, which reveals relative health or malfunction in the anaerobic digestion process. If there is no methane production, there is no energy yield and no biomass growth of the crucial *methanogens*, nor any BOD/COD reduction of the organic pollutant, because a reduced end product must be withdrawn from the feedstock under anaerobic conditions to reduce BOD/COD (Speece, 1996). In a digestion process operating well the percentage of CH_4 would be approximately 65-70% and the CO_2 between 25-30%.

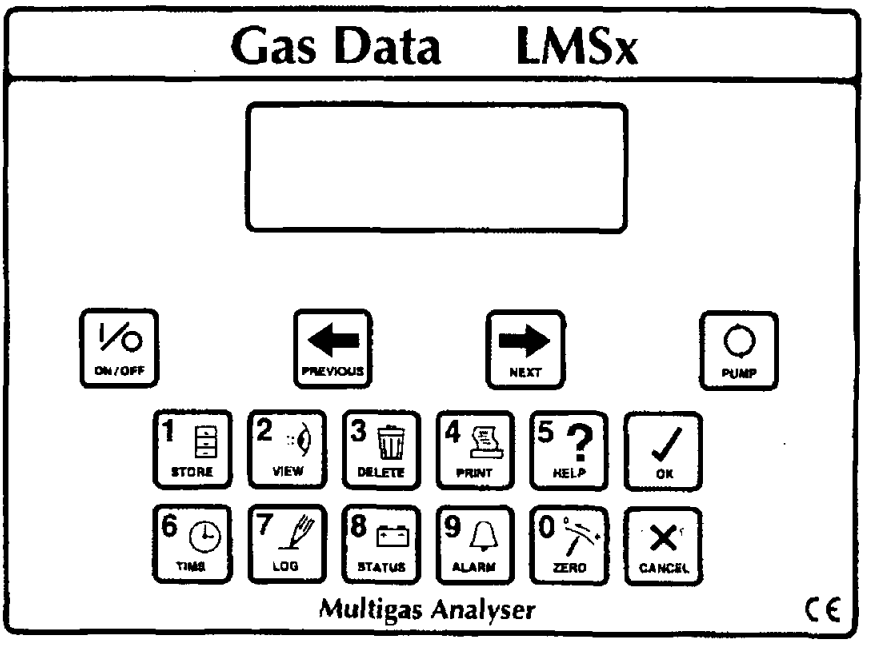
Gas composition was monitored using the Gas Data model LMSx, Multigas Analyser. This analyser was bought from Mason Technology, Dublin and was manufactured by Gas Data, Coventry, UK.

This analyser could measure, simultaneously, methane, carbon dioxide and oxygen concentrations, plus atmospheric pressure, borehole flow rate, and site temperature. Methane and carbon dioxide concentrations are measured using an in built non-dispersive infra red analyser. These parameters could then be stored with the current time, date and site name. Readings may be viewed on-site or downloaded to a host computer for report complication and printing. The LMSx had an integral gas-sampling pump, and came complete with a hands free carry case, battery charger and gas filter. Menu-driver PC software and a communications cable are also available.

The percentage CH_4, CO_2 and O_2 generated in the biogas from the three anaerobic digesters in these trials was measured and stored by the LMSx gas analyser. The biogas composition was measured between one and three times a day. The gas analyser was connected up to a closed valve, especially modified for the analyser, on the gas lines from the digester to the gas collection vessel (Figure 2.4). The pipe continuing to the gas collection vessel was sealed while biogas analysis was being carried out. The simple gas valve to the gas analyser was opened and the analyser was turned on. The internal pump in the analyser then was turned on. This would pump a small quantity of biogas from the gas system. The display would then read % CH_4, % CO_2 and %O_2 in the biogas. This would then be stored under the digester site name. Each digester, 1, 2 and 3 had a site name, 1, 2 and 3, respectively in the gas analyser.

The menu-driven software and communication cable was purchased and the information stored in the analyser was downloaded on a computer. However this was not that easy, as the software program was very old (DOS mode) and very difficult to use. After down loading the information still had to be converted to a Windows format so the information could be used in Excel. This process proved cumbersome and the software was antiquated by modern standards. In addition the software did give problems associated with Y2K non-compliant systems. Figure 2.6 shows the basic keyboard layout of the gas analyser.

FIGURE 2.6 BASIC KEYBOARD LAYOUT OF GAS DATA LMSX



2.7.4 EFFLUENT MONITORING - pH

The value and stability of the pH in an anaerobic reactor is extremely important because methanogenesis only proceeds at a high rate when the pH is maintained in the neutral range. At pH values lower than 6.3 or higher than 7.8 the rate of methanogenesis decreases. (Haandel & Lettinga, 1994).

The pH of the influent and effluent was determined each time the digesters were feed. The pH was taken of the influent just before entering the digester and of the effluent just after exiting the digester. When measuring the effluent it was important to measure the pH within minutes. If the sample is allowed to stand exposed to the air for a few minutes CO_2 would escape, causing the pH to rise. Also if the sample is stirred or agitated it will lose CO_2 even

faster and increase the difference between the pH measured in the sample and within the reactor even more (Speece, 1996).

Analysis was conducted using a hand held probe purchased from Lennox Ltd, Dublin. Standardisation against buffer solutions of pH and 10 was made as appropriate.

Method: Standard Method No. 4500-H+ B, (19^{th} edition, 1995).

Apparatus: HACH model pH meter

Buffer Solution; e.g. buffer solution at pH 4, 7 and 10

Procedure: The electrodes were rinsed with distilled water and wiped gently with tissue paper. The pH meter was calibrated using a buffer solution of known pH. The pH of the buffer solution was as close to the pH of the solution to be measured as possible. The range selector of the pH meter was turned to the pH buffer solution and the pH was measured of the water or wastewater sample, the temperature selector on the pH meter was adjusted to the temperature of the sample being tested.

2.7.5 EFFLUENT MONITORING - ALKALINITY

The alkalinity of a water/wastewater is a measure of its capacity to neutralise acids and is due primarily to the presence of the hydroxides, carbonates and bicarbonates. Low pH conditions may be caused by two sources of acidity, H_2CO_3 and volatile fatty acids (VFA), which are generated in microbial reactions. These acids require alkalinity for neutralisation so that the

microbial activity is not hindered by the pH depression. However, the major requirement for alkalinity in well operating anaerobic processes is neutralisation of the high H_2CO_3 which results from high partial pressure of CO_2 in the reactor (volatile acids concentrations commonly are low). If the acids concentrations (H_2CO_3 and VFA) exceeds the available alkalinity, the reactor will "sour" (a drop in pH), severely inhibiting microbial activity, especially the methanogens. When methane production becomes "struck" (ceases) the VFA may continue to accumulate, exacerbating the situation further. (Speece, 1996). Typical values of alkalinity during a well-operated digestion process range between 2000-4000 mg/l $CaCO_3$ (Killilea, 1992).

The alkalinity during these trials was not measured until the VFA began to increase. Then the alkalinity of the effluent was obtained approximately three times a week. The determination was reported in units of $CaCO_3$ as specified in section 2320-B (Titration Method) of the Standard Methods (19th edition, 1995). Dilutions were made and alkalinity was determined using a sample size of 20 ml in accordance with the following formula.

$$\text{Alkalinity (mg CaCO3/l)} = \frac{0.1\text{N } H_2SO_4 \text{ (ml)} * 1000}{\text{Correction factor} * \text{sample (ml)}} * \text{Dilution factor}$$

2.7.6 EFFLUENT MONITORING - VOLATILE FATTY ACIDS

Volatile Fatty Acids (VFA) are formed as intermediates during the anaerobic digestion of carbohydrates, proteins and fats. As described in the previous section if these VFA concentrations are allowed to build up and exceed the available alkalinity, the reactor will

"sour" severely inhibiting methanogenic activity. Normal operational VFA values would be between 100-300mg/l.

Method: HACH Method, Volatile Acids (0 to 2800 as mg/l acetic acid) Method 8196 for digester sludge Esterification Method.

Apparatus: Spectrophotometer (HACH model DR 2000); water bath; centrifuge;

Reagents: Ethylene glycol; 19.2 N Sulfuric Acid; Hydroxylamine Hydrochloride; 4.5 N Sodium Hydroxide Standard; Ferric Chloride Sulfuric.

Procedure: The stored program number for volatile acids as acetic acids was entered into the spectrophotometer (770). The wavelength dial was then rotated until the small display showed 495nm. Using a pipette, 0.5 ml of deionised water was put into a dry 25-mL sample cell (the blank). The sample was centrifuged and 0.5 ml of this sample was placed into another dry 25-mL sample cell (the prepared sample. Then 1.5 ml of ethylene glycol was pipetted into each sample cell and swirled to mix. 0.2 ml of 19.2 N Sulfuric Acid Standard Solution was pipetted into each cell and swirled to mix. Both cells were placed into a boiling water bath for 3 minutes. When the timer beeped, the solutions were cooled to 25°C (until cell are cold) with running tap water. Then 0.5mL of Hydroxylamine Hydrochloride Solution was pipetted into each cell and swirled to mix. 2.0 ml of 4.5 H Sodium Hydroxide Standard Solution was pipetted into each cells and swirled to mix. Then 10 ml of Ferric Chloride Sulfuric Acid solution and 10 ml of deionised water was added to each and

again swirled to mix. A 3-minute reaction period was allowed to begin. When the timer beeped, the bank was immediately placed into the cell holder and zeroed. Then the prepared sample was placed into the cell holder and measured. The display showed the result in mg/l acetic acid.

Summary of Method

The volatile acids test is designed specifically for the determination of volatile acids in digesting sludges. The method is based on esterification of carboxylic acids present and determination of the esters by the ferric hydroxamate reaction. All volatile organic acids present are reported as their equivalent mg/L acetic acid.

2.7.7 EFFLUENT MONITORING - CHEMICAL OXYGEN DEMAND (COD)

The Chemical Oxygen Demand (COD) is the oxygen equivalent of the organic matter contained within a liquid. The test measures the amount of oxygen required for chemical oxidation of organic matter in the sample to carbon dioxide and water. It differs from the standard water and wastewater test, Biochemical Oxygen Demand (BOD) by virtue of the inclusion of some inorganic material in its determination.

Method: Standard Method No.5220 (19th edition, 1995)

Reactor Digestion Method

Apparatus: COD reactor, Spectrophotometer (HACH model DR2000)

Reagents: COD vials, manufactured by HACH

Procedure: 2 mls of sample to be tested was pipetted into the appropriate vial type (low range, high range or high range plus). The vial was inverted several times before placing into the preheated COD reactor. This procedure was repeated to prepare the blank with deionised water. The vials were left in the COD reactor for 2 hours at 150°C. After two hours, the COD reactor was turned off and the COD vials were allowed to cool to 120°C. Each vial was then inverted while still warm before being placed in a rack to cool to room temperature.

The *Colorimetric Determination* is as follows:

The stored number for COD for the required range was entered in to the spectrophotometer: High range & high range plus = 435 Enter

Then the wavelength dial was rotated until the small display showed 420 nm for low range and 620nm for high range & high range plus. The blank vial was then placed into the vial adapter in the spectrophotometer and zeroed. Finally the sample vial was placed into the adapter and measured. The result was displayed in mg/l COD.

It was necessary to establish the organic load entering the digesters to ensure that the digesters were not being over or under loaded. Therefore the COD of the influent was determined. The effluent COD was measured in order to assess the ability of the reactors to remove organic matter and to compare actual biogas generation with theoretical gas production based on COD removed. The difference in the two values was the organic matter removal rate. Dilutions were required of each sample to ensure that the COD determination remained in the 0 to 1500 mg/l vial range. Multiplication by the subsequent dilution factor gave the true COD reading.

CHAPTER 3

INVESTIGATION OF FULL SCALE APPLICATIONS OF ANAEROBIC DIGESTION OF SEWAGE SOLIDS IN IRELAND

3.1 INTRODUCTION

In order to obtain a comprehensive knowledge of the fundamentals behind the design, operation and maintenance of full-scale anaerobic digestion sites in Ireland, site visits were conducted. Each of the five anaerobic digestion sites in Ireland were investigated and studied. To make the most out of the site visits a questionnaire was invented prior to visiting the plants. This questionnaire was modelled on the layout and structure of the data collection forms used by the US EPA to evaluated sewage treatment plants (Section 2.1.1, Chapter 2).

This procedure was applied to existing and newly commissioned AD plants in Ireland in order to deduce process performance. The purpose of this chapter is to describe in detail the current state-of-the-art of anaerobic digestion of sewage sludge in Ireland.

3.1.1 CONDUCTING THE SURVEY AND GATHERING INFORMATION

All existing anaerobic digestion sites were visited on at least three separate occasions. The first visit was of an introductory nature in order to become acquainted with plant personnel and a brief description of each plant was drawn up. On subsequent visits, personnel were questioned in detail with the help of data collection forms (questionnaires) filled out on site. These forms were then summarised in tabular form and returned for comments and corrections by on-site personnel at each plant. A database of process, mechanical and electrical installations was recorded. Results on plant performance were reviewed (solids concentrations, gas production etc.). The plant details obtained during each site visits are described in the following sections.

3.2 TULLAMORE SEWAGE TREATMENT PLANT

HISTORY Tullamore is a medium sized town in the central plain of Ireland and is the Capital town of County Offaly. It had a population of approximately 9,000 at the time the present sewage treatment works was designed and constructed in the late 1980's. Today, the population of the town has increased to approximately 11,000 persons with a further 3,000 persons population equivalent consisting of industry, students, shoppers, office, commercial and industrial employees. Over the last ten years the population has increased by 15% and continues to grow. The present treatment works is the third to be located in the same area. The present treatment works was constructed in 1986 and was very successful, however to improve it efficiency even more a slight upgrade to the plant took place in 1997. Figure 3.1 shows the flow diagram of the wastewater treatment stream at Tullamore.

PLANT LOADING AND SIZE The plant is designed to serve an equivalent population of 16,000. The current P.E. entering the plant is 14,000 therefore the plant will reach full capacity within the next few years. However there have been provisions made in the plant design for future expansion.

PRELIMINARY TREATMENT The first compartment of the inlet works contains two chambers, the influent reception area and the storm overflow chamber. A screen in the storm overflow chamber eliminates the problem with plastics and rags entering the river through a water overflow pipe during periods of heavy rain. Flows in excess of 3 DWF are allowed overflow to the river from this chamber.

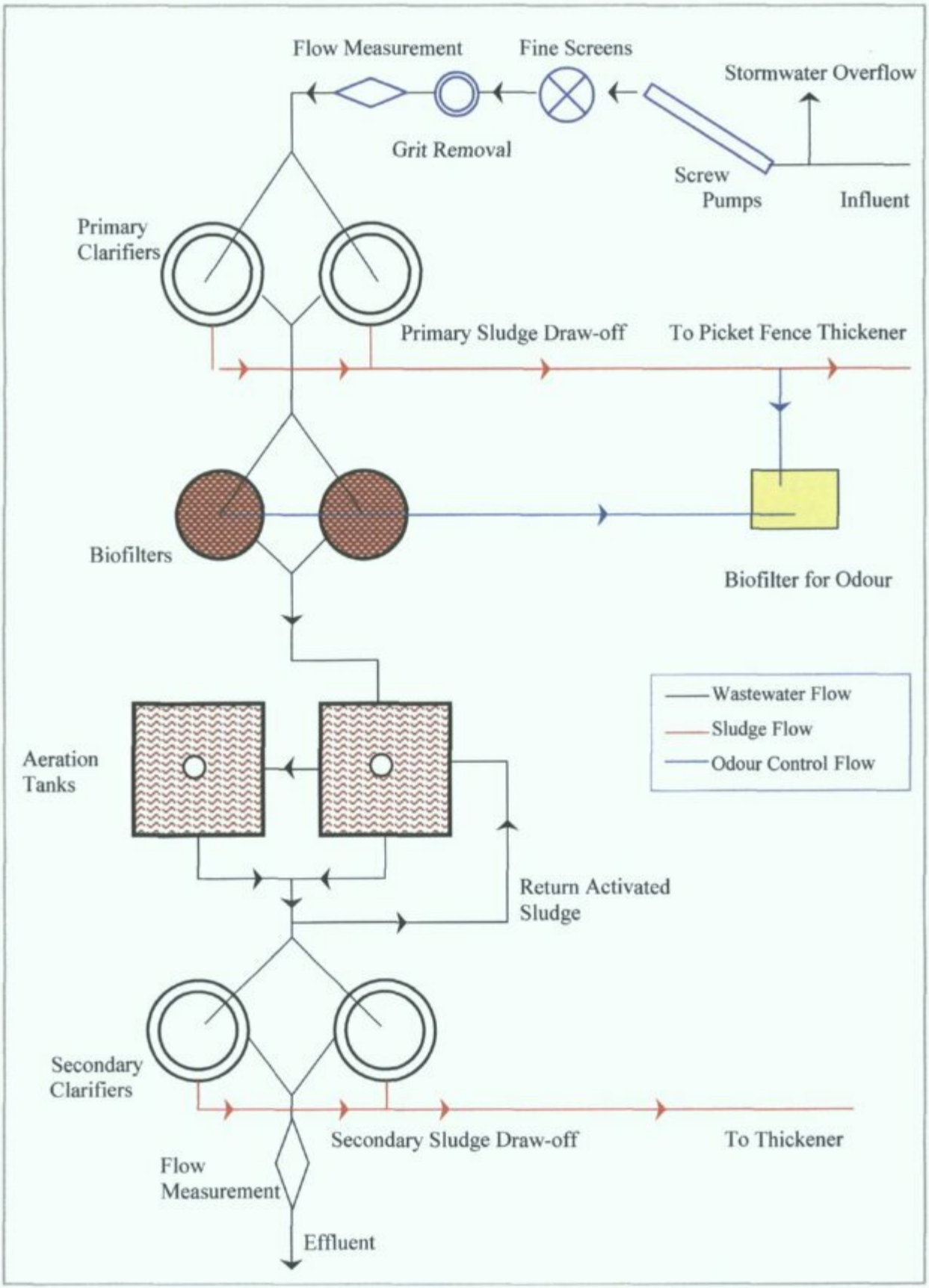


FIGURE 3.1 TULLAMORE SEWAGE TREATMENT PLANT WASTEWATER FLOW DIAGRAM

The influent then enters two screw pumps (Figure 3.1 & Plate 3.2.1) where it is lifted up to a height, so as it can flow by gravity throughout the rest of the plant. There are provisions for a third screw pump in the case of future duplication of the works.

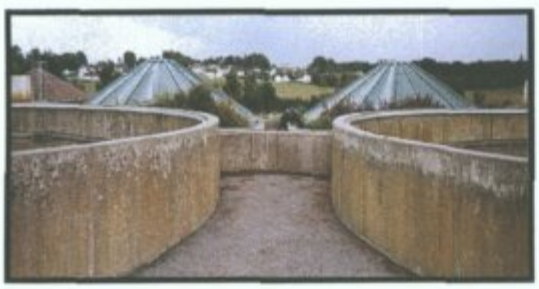
The wastewater then passes through two fine screens (Figure 3.1 & Plate 3.2.1), that are part of the recent upgrade. These screens separate any rags, plastics etc. from the wastewater. From the screens the wastewater flows into the grit trap (Plate 3.2.1) where any heavy material such as, sand, gravel, road chipping's etc. contained in the flow are removed. There are provisions for a second grit trap in the case of future duplication of the works.

PLATE 3.2.1 SCREW PUMPS, FINE SCREENS & GRIT CHAMBER

PRIMARY TREATMENT There are two primary settling tanks (clarifiers) on site (Figure 3.1 & Plate 3.2.2). Both are 14 metres diameter having a side wall depth of 2 metres and a floor slope of 10°. As automatic desludging is in use, sludge storage is not a critical parameter and $6m^3$ of storage is provided in each sludge hopper.

From the primary settlement tanks the wastewater flows into the two primary trickling filters (Figure 3.1 & Plate 3.2.2). Each are 9.8 metre diameters and 1.8 metres deep having a total volume of media of 272 m^3. The liquid is sprayed over the surface by water driven rotary distributors.

PLATE 3.2.2 THE PRIMARY CLARIFIERS & THE TRICKLING FILTERS



SECONDARY TREATMENT The effluent from the trickling filters flows directly to the aeration tanks where it mixes with the recycled activated sludge at the inlet point. The aeration tanks (Plate 3.2.3) consist of two 12m x 12m x 2.6m deep tanks with a further 1100 mm freeboard providing a liquid capacity of 750m^3. The tanks are designed to operate in series. Either tank can be used on its own in case of mechanical breakdown. Aeration is carried out by a centrally located surface aerator in each tank and are controlled depending on the oxygen demand as indicated by dissolved oxygen probes installed in the liquid. Activated sludge is recycled by means of duplicate screw pumps driven by two speed motors and each capable of recycling the sludge at rates up to 100% DWF.

Chemical Phosphorus removal in the form of ferric chloride addition takes place in the aeration tank. Ferric chloride is pumped slowly at the inlet with the return sludge going back into the aeration tank.

The two final settling tanks (Plate 3.2.3) are scraped tanks with the same diameter as the primary tanks but having 3m deep sidewalls and a 15% floor slope. Desludging is carried out on a continuous basis by hydrostatic head to the sludge return screw pump inlet chamber.

PLATE 3.2.3 THE AERATION TANK & THE TWO SECONDARY CLARIFIERS

SLUDGE PREPARATION Figure 3.2 shows the flow diagram of the sludge treatment stream. Primary sludge, from the primary clarifiers is pumped down to a sludge reception chamber system, before being pumped to the picket fence thickener (Figure 3.2 & Plate 3.4). Some waste activated sludge (secondary sludge) was previously mixed with the primary sludge. However, because this sludge contained only 0.7% solids it made the primary sludge too thin entering the picket fence thickeners reducing the efficiency of the thickener. Therefore as part as the 1997 upgrade a new thickening centre was built containing two new thickening presses (Plate 3.2.4 & Figure 3.2). This receives all the waste activated sludge for thickening before being feed to the anaerobic digester.

The primary sludge is the only sludge now being fed to the picket fence thickener. The thickener is a covered concrete tank, 5.8m in diameter and 4.5m deep having rotary floor scrapers and full depth picket frame. A continuous supply of primary sludge is pumped to the picket fence thickener throughout the day via the automatic desludging equipment.

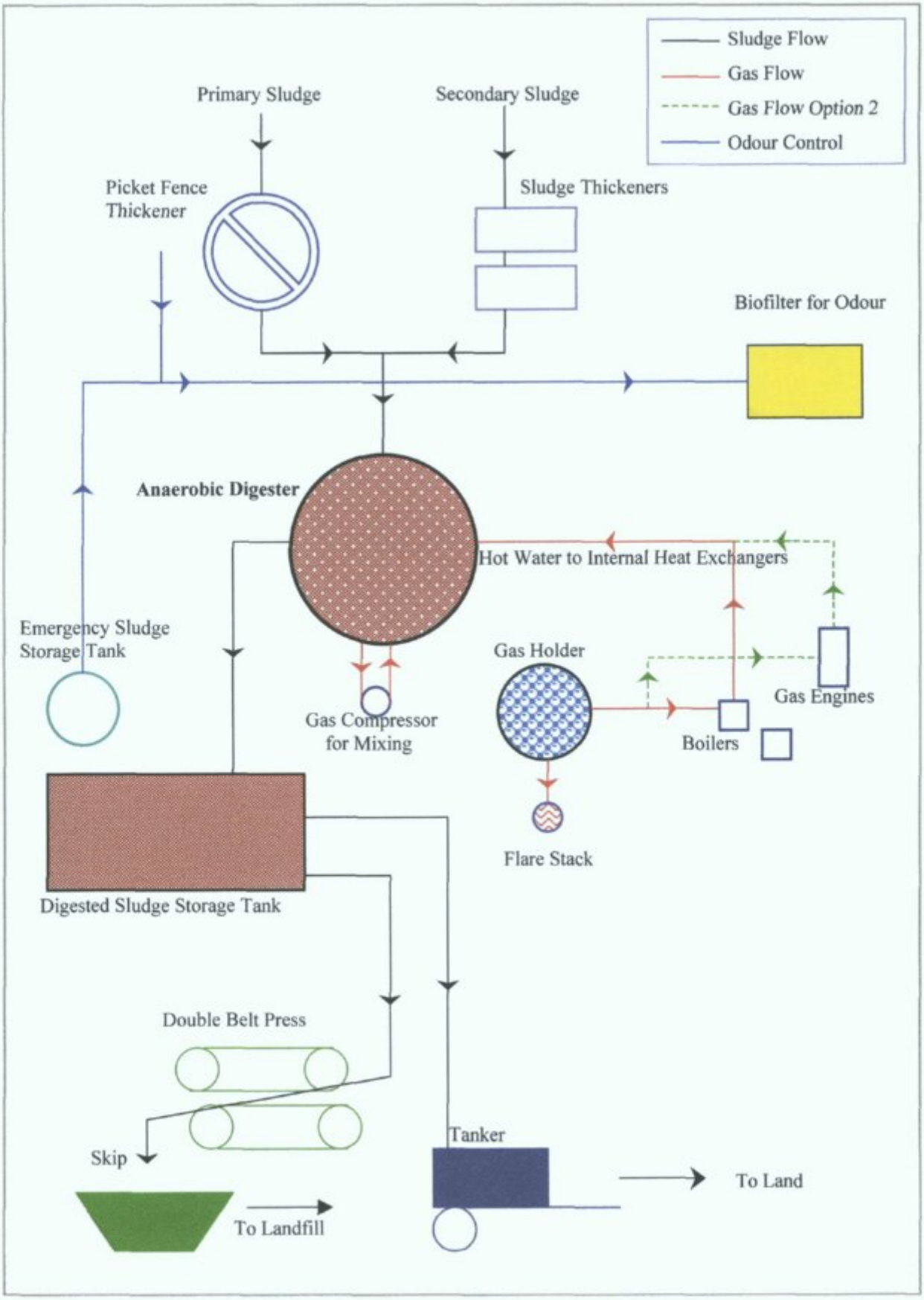


FIGURE 3.2 TULLAMORE SEWAGE TREATMENT PLANT SLUDGE FLOW DIAGRAM

PLATE 3.2.4 PICKET FENCE THICKENER & SLUDGE THICKENERS

SLUDGE TREATMENT (THE ANAEROBIC DIGESTER)The anaerobic digester (Plate 3.2.5) is an above ground cylindrical tank constructed from glass-coated steel with an *in-situ* concrete base. The tank dimensions are approximately 7.6m diameter and 7.3m high, giving a sludge volume of 330m^3. The digester wall is insulated externally with twin-skin glass reinforced plastic (g.r.p) and polyurethane foam panels, each panel corresponding to the sheet of the tank beneath it. The tank roof is fixed, using insulated g.r.p construction.

PLATE 3.2.5 TULLAMORE ANAEROBIC DIGESTER

DIGESTER FEEDING Thickened sludge from the picket fence thickener and the waste activated sludge thickeners are pumped to the anaerobic digester, by two feed pumps (duty/standby), in small batches at different times throughout the day. Both sludges enter the digester at the inlet port on the digester roof. The contents of the digester are heated to mesophilic temperature and mixed by gas recirculation. The sludge inlet port consists of a wide diameter tube projecting down into the sludge a distance of approximately 1000mm. This provides a seal for the gas in the crown of the digester roof. The inlet is fitted with level probes to detect high sludge level in the digester. The sludge overflow weirs consist of a g.r.p assembly bolted to the side of the tank. Sludge passes to the weir through a 200mm square opening in the tank, passing over the weir to the downpipe.

DIGESTER HEATING The contents of the digester are maintained at a constant temperature of 35°C by circulation of hot water through internal heat exchangers fitted in the base of the digester. There are three separate processes for water heating:

1. One gas boiler designed to burn methane gas from the digestion process (normally used if the CHP unit is out of commission).

2. One gas boiler designed to burn commercial propane gas (used for start up or in case of failure of methane gas supply).

3. Combined Heat and Power (CHP) Unit, which uses methane gas to generate electrical power while the heat, recovered from the exhaust, cooling and oil system satisfies the heating requirements of the digester.

The digester is fitted with eight internal annular heat exchangers, fabricated in epoxy coated steel, fixed to the digester base in a circular array. The heat exchanger dimensions are 610mm diameter and 1200mm high. Each heat exchanger is located over one of the stirring pipes to ensure a good flow of sludge around the surface. The exchangers are fed from a 2" supply and return water main. The digester is fitted with one water pump on the heating water circuit, which is designed to operate at all the times, irrespective of whether the generator or boilers are operating. The boilers are separate propane and methane gas boilers.

DIGESTER MIXING Digester mixing is by gas recirculation through a series of 12 stirring nozzles around the base of the tank. Gas is drawn from the gas holder and pumped by the gas pump to the rotary valve. The rotary valve is an L port valve driven by reduction gears and a ratchet drive off the gas pump motor. The valve is rotated by the ratchet, which moves one port position every minute, to give one minute of stirring for each stirring pipe while the gas pump is running. The operation of the gas pump is automatic, controlled by the gas pump timer on the main digester control panel. Manual operation is also provided.

GAS STORAGE AND TREATMENT The sludge gas produced in the process is taken off at the roof of the digester and flows to the gas holder (Plate 3.2.6). The gas holder is a bell-over-water type, of a twin -skinned g.r.p construction, located on a concrete slab at the back of the plant. This is a low capacity type ($10m^3$), designed less for bulk gas storage than for pressure regulation, gas buffering and gas filtering. Three independent outlets are taken from it; for the gas main to the digester, for the gas supply to the boiler in the control room, and for the supply to the gas pump for digester mixing.

Gas collecting at the surface of the digesters is generally dirty and saturated with water. To reduce the quantity of dirt and water entering the gas lines and the gas holder, gas is collected in a large chamber at the top of the digester, called the splash trap, from which the gas outlet is taken. The splash trap operates by drawing gas from the digester slowly, allowing solids carried by the gas to fall back into the digester. The trap is not insulated in order to provide some cooling of the gas and condense out some of the vapour in the gas. The outlet pipe turns up inside the trap to draw gas from its highest point.

The low point of the low-pressure gas pipes (gas lines to the digester, boiler and gas pump) is fitted with a condensation trap located next to the gas holder. This is an A.B.S. water-filled trap that collects condensation contained in the biogas, before entering the boilers, pumps and engines.

A gas/sludge pressure relief valve protects the digester tank from either overpressure or vacuum conditions. A gas vent is mounted at the top of the splash trap in the centre of the digester roof. On power failure or if the panel isolator is turned off, this gas vent allows gas to be vented from the system.

ENERGY GENERATION The CHP unit (Plate 3.2.6) consists of a gas engine specially designed for use on "sour" gas with high hydrogen sulphide content from oil well heads and it has an output of 21 kW. The generator is an asynchronous type and consequently cannot be operated independently of the Electricity Supply Board (ESB) supply. However, the electrical power generated is fed back to the Main Control Panel thus reducing the demand from the ESB grid, (the ESB did not agree to meter this or to pay for it).

As a precautionary measure a flare stack (Plate 3.2.6) was installed, as part of the 1997 upgrade, to cater for the burn off of excess biogas when required. The flare is rarely operated as maximum usage is derived from gas in the CHP unit.

PLATE 3.2.6 THE GAS HOLDER, CHP UNIT & FLARE STACK



DIGESTED SLUDGE TREATMENT/DISPOSAL The digester sludge overflows at the digester overflow, spilling over into the adjacent concrete holding tank (Figure 3.2 & Plate 3.2.7). The sludge emerges at a fairly constant stream, due to the regular loading cycle. This tank is a rectangular opened concrete tank 24m x 7.5m x 2.9m deep with 300mm freeboard. The capacity is 529m^3. Its purpose is to store digested sludge pending removal for dewatering or land disposal, allowing any remaining gas escape from the digested sludge or allow removal of further supernatant liquid. It also acts as an emergency holding tank (Plate 3.2.7) for raw sludge if the digester is out of commission for any reason. A smaller sludge holding tank was installed beside this tank, for emergencies (Figure 3.2).

The digester is provided with a bottom drain fitted with two valves where the drain passes out from the digester foundations. The bottom drain discharges to the adjacent digester sludge holding tank.

For approximately 3 months of the year when the sludge cannot be spread on the land, digested sludge from the storage tank is dewatered in a double belt press (Figure 3.2). It is a 1.5 metre wide double belt filter press and designed to increase the solids content of the sludge from 5% to 17%. Because of the storage provided in the sludge holding tank it was not considered necessary to provide a standby machine but sufficient space has been provided for its future installation if necessary. This dewatered sludge is then disposed of, to landfill.

There is a slurry spreader designed to spread slurry on farmland on site. When conditions allow this is the preferred method of dealing with digested sludge because of the potential savings in energy and polyelectrolyte costs if dewatering can be omitted.

PLATE 3.2.7 DIGESTED SLUDGE HOLDING TANK & EMERGENCY HOLDING TANK

ODOUR CONTROL The biggest and most obvious upgrade in the plant is the addition of the new odour control system (Plate 3.2.8). The primary tank desludging units, the trickling filters, the picket fence thickener and the emergency sludge holding tank are all covered sealed units (Figure 3.1 and 3.2). These units are connected up to a series of pipes, which in turn are connected to the odour control biofilter where the odours are neutralised so as no dangerous, odourous gases escape in to the atmosphere.

PLATE 3.2.8 ODOUR CONTROL SHELL FILTER

DIGESTER MONITORING There is a full-equipped laboratory on site in which one full-time technician is involved with carrying out daily testing of the plant treatment processes. Monitoring of the digester involves approximately half of the testing procedures carried out in the plant. The pH and temperature is measured each day and the total solids are measured 2 to 3 times weekly for the undigested sludge. On the digested sludge the pH and temperature are measured each day and the total solids, volatile fatty acids and the alkalinity are measured approximately 2 to 3 times weekly depending on the performance of the digester at the time of analysis. The methane composition of the biogas is also measured every day. The digester has no online monitoring facility.

PLANT MANAGEMENT There are four full-time staff employed at the plant: the plant operator, who is in charge of the entire plant operation and maintenance and makes all the crucial decisions in the daily running of the plant with the help of one laboratory technician and two maintenance personal. The plant is thoroughly cleaned, landscaped and is well maintained.

3.3 BUNCRANA SEWAGE TREATMENT PLANT

HISTORY Buncrana is the second largest town in Co. Donegal, which lies on the northwest Coast of Ireland. Buncrana itself is located along the coastline. Prior to 1990, Buncrana Town had no sewage treatment and raw sewage flowed directly in to the sea. In 1990 the treatment work was constructed.

PLANT LOADING AND SIZE The plant was designed for a population equivalent of 13,000 persons, however at the moment only a load equivalent to 7,000 PE, is been treated. Buncrana is the smallest sewage treatment plant that was surveyed. The plant consists of primary treatment only followed by anaerobic digestion of sludge. The principle units for the treatment of wastewater in the works are described in the following sections. Figure 3.3 shows the flow diagram of the wastewater and sludge treatment stream.

PRELIMINARY TREATMENT The influent wastewater flows into the inlet works through a channel. A wall separates the inflow from the overflow. Inserted in this wall are bar rack screens to stop large particles getting into the final pumping station before been disposed of to the sea. Also in the overflow chamber there is a rectangular horizontal flow grit chamber to remove grit from the overflow stream. The inflow then enters a building where preliminary treatment takes place. The wastewater first flows through two comminutions, where course solids are grinded up without being removed from the flow (Figure 3.3). The flow then leaves the comminution chamber and enters two aerated grit chambers where the grit is removed. The wastewater then goes out to a splitter chamber and where influent flow is measured by ultrasonic gauging.

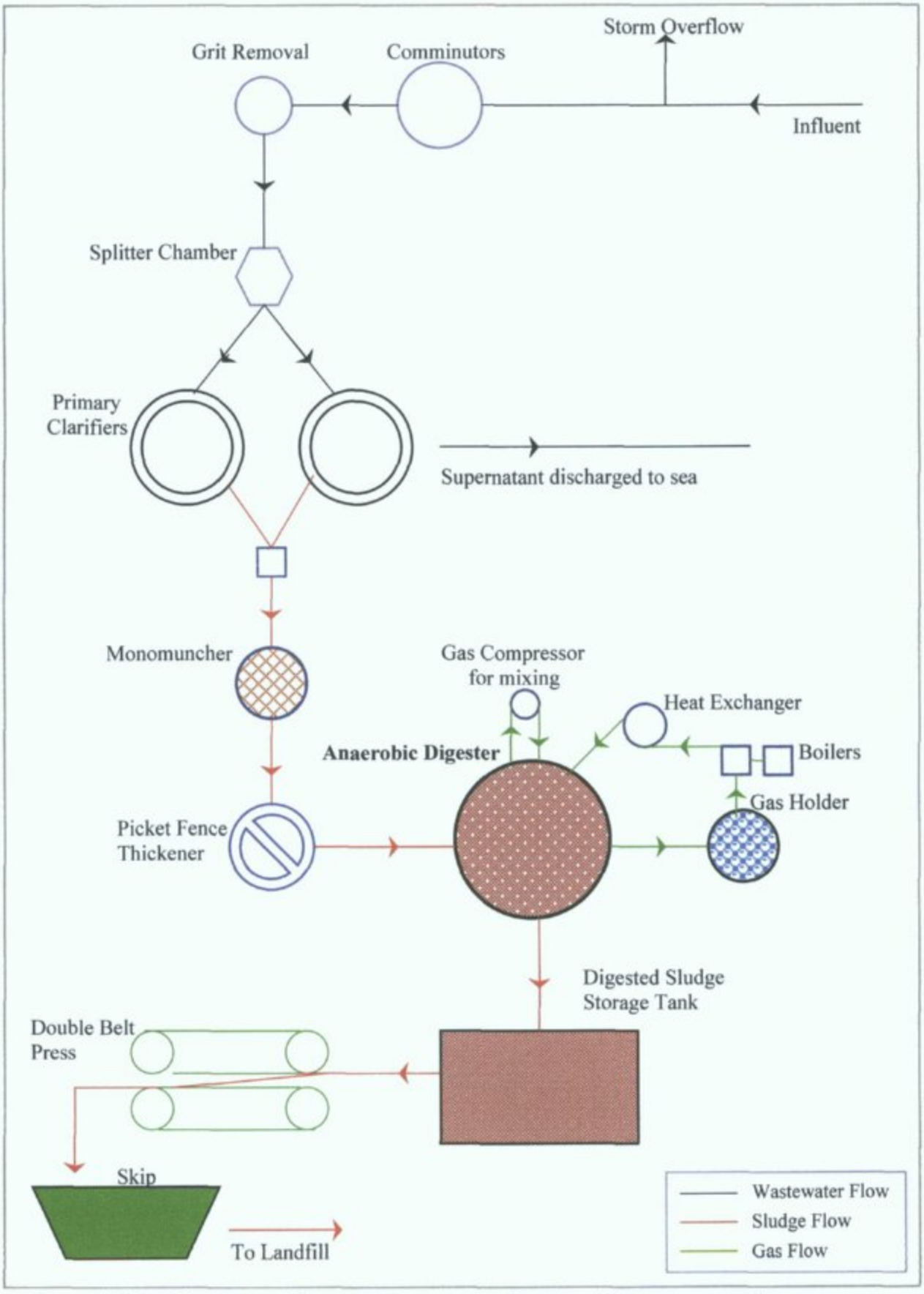
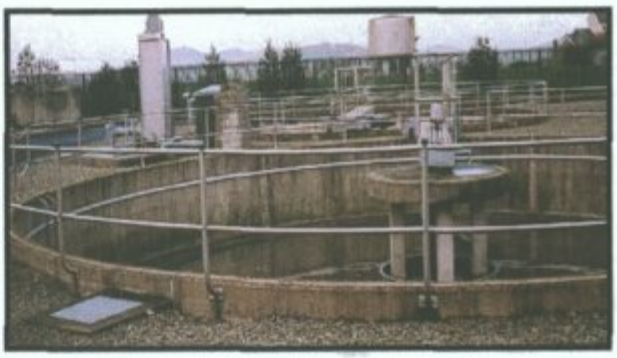


FIGURE 3.3 BUNCRANA SEWAGE TREATMENT PLANT: WASTEWATER, SLUDGE AND GAS FLOW DIAGRAM

PRIMARY TREATMENT The flow is split equally between two primary settlement tanks (clarifiers) (Plate 3.3.1). These clarifiers are 12.3m in diameter and 3.5m deep. From the primary settlement tank is the weir overflow (supernatant), with no further treatment, flows to the main pumping station where it is then discharged, via an 800m discharge pipe, to the sea. This practice could cause significant pollution if the primary settlement tank was hydraulically overloaded leading to loss of primary solids over the weir. In the current operation it is not possible to remove significant quantities of pathogenic organisms because no secondary treatment or disinfection is exercised at the plant.

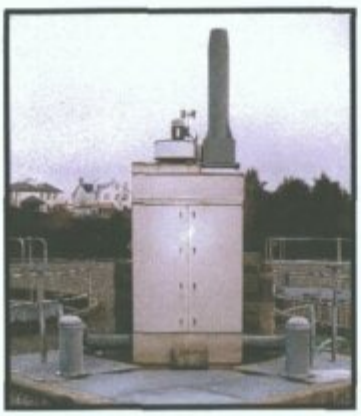
PLATE 3.3.1 SPLITTER CHAMBER & ONE OF THE TWO PRIMARY CLARIFIERS



SECONDARY TREATMENT There is no secondary treatment at the plant. The practice of discharging a wastewater, which has only been primary treated, to sea, is no longer permitted under the urban wastewater directive 1991 and the bathing directive 1976. Buncrana like many other seaside towns will have to upgrade their wastewater treatment to include secondary treatment by the year 2005.

SLUDGE PREPARATION The sludge, which settles to the bottom of the primary settlement tank (clarifier) goes into a sump and after this by the use of air blowers, the sludge is blown up to the monomuncher. The monmuncher (Plate 3.3.2) further chops up the gross solids in the sludge before entering the picket fence thickener. Centrifugal type pumps (2 duty and 1 standby), pump the chopped sludge to the picket fence thickener. In the picket fence thickener (Plate 3.3.2), the solids concentration of the sludge is increased by removing the liquid fraction. After this the sludge is then pumped in to the anaerobic digester.

PLATE 3.3.2 MONOMUNCHER & PICKET FENCE THICKENERS



SLUDGE TREATMENT (ANAEROBIC DIGESTION) By the year 1991 the anaerobic digester at the Buncrana Plant was constructed and commissioned. It was the second heated, gas mixing digester in Ireland. The anaerobic digester (Plate 3.3.3) is cylindrical in shape with a diameter of 5.1m and a height of 6.88m. The total volume is 133m^3 with a sludge volume of 120m^3. The tank is steel glass lined with a fixed roof. The walls are insulated with 4" cladding and the roof is not insulated.

PLATE 3.3.3 BUNCRANA ANAEROBIC DIGESTER

DIGESTER FEEDING The sludge is manually fed intermittently twice daily through the top of the digester. As the digester is being fed, equal quantities of sludge is displaced out through a bell mouth outlet at the top of the digester. Two alternating pumps are used to pump the thickened sludge to the digester. These pumps are situated on the top of the picket fence thickener. The current pumps are relatively new pumps as the previous pumps could not cope with the thick sludge coming from the picket fence thickener. The old pumps were continuously blocking.

The feed sludge to the digester comes directly from the picket fence thickener. There is no storage tank or sump to store the sludge after thickening. This has proven to cause problems as the sludge in the picket fence thickener gets extremely thick once the retention time in the thickener goes over the required retention time. This can happen if the feed to the digester is reduced and the sludge from the primary clarifier is still entering the picket fence thickener. There is no room for this sludge in the thickener as the feed to the digester has been reduced.

This thick sludge in turn causes problems for the pumps and the digester and also for the thickener itself. The sludge may also go septic outside the digester causing serious odour problems. When this build up of sludge occurred at the plant the untreated sludge had to by pass the digester and into the digested sludge holding tank. This contaminated the digested sludge.

DIGESTER HEATING The digestion process is operated at mesophilic temperatures and is heated by a heat-exchanger unit (Plate 3.3.4). This unit is incorporated into the sludge circulation system. The exchanger is a hot water heater, consisting of two co-axial tubes with the sludge flowing through the inner tube and hot water through the outer tube in opposite directions. The tubes are in the shape of a coil like a snail shell. The unit is located inside and there is only one unit on site, there is no back-up heat exchanger. This heat exchanger has been very successful and no blockages have been recorded since its installation.

Hot water heating is provided from two boilers (Plate 3.3.4); one burns methane gas from the digestion process and the other burns imported propane gas. Two pumps (Plate 3.3.4) keep the hot water circulating from the boilers to the heat exchanger. A compressor (Plate 3.3.4) keeps an even pressure of methane to the boiler and similarly for the propane boiler. They are on a cut-in and cut out system, obviously when there is ample supply of methane gas being produced by the digestion process the methane boiler is in use. However, if a problem should occur with the process that would cause insufficient methane production, then the propane boiler will cut in to heat the water. This should only be used as a back-up system as it is not economical to be burning propane gas when methane is available. Both boilers are also housed inside with the heat exchanger.

PLATE 3.3.4 HEAT EXCHANGER, BOILERS, GAS COMPRESSOR & HOT WATER PUMPS



A temperature probe transmits the temperature into the control room and depending on the reading the boilers are adjusted manually to correct the temperature. There is no cooling radiator on site; this would cool the digester contents if they rose too high due to high weather temperatures or boiler misfunction.

There are 2 centrifugal type sludge circulation pumps, (1 duty and 1 standby), that pump the sludge from the digester through the heat exchanger and back to the digester again. One pump is new (duty) as the older mono pumps were very expensive to run due to high maintenance and electricity costs. The new pump has been running successful since its installation and is running for a fraction of the price of the older pump. There is still one old pump (standby) installed in case a breakdown occurs with either of the centrifugal pumps. The pumps are installed indoors with the heat exchanger and the boilers. The pipes running to and from the digester carrying the recirculated heating sludge have within the last few years been insulated to reduce heat loss.

DIGESTER MIXING Mixing in the digester is by means of gas injection. There are 8 gas pipes located near the bottom of the digester and spread evenly around the diameter of the

digester. There is a gas compressor (Plate 3.3.5) situated outside the digester, which compresses the gas to the 8-gas pipes in the digester. The gas entering the digester is controlled by pneumatic valves, which are located on each of the gas pipes at the entrance to the digester. The air compressor controlling the opening and closing of the valves is located in the sludge building. The gas system works as follows: Two gas valves are opened and the gas is compressed into the digester for a certain period, mixing that part of the tank. Just before these two valves are closed the neighboring two valves are opened and gas is compressed and expressed to these two valves. Then the two other valves cutout. Therefore, four valves are on together just for a short period. This pattern continues round the 8 valves and continues for 24 hours a day to keep the sludge contents well mixed.

GAS STORAGE AND TREATMENT The biogas used for the methane boiler and mixing of the digester contents is stored in a bell-over-water type gas holding tank located alongside the digester (Plate 3.3.5). Observation of the gas bell rising up and down was the only way of guessing the volume of gas being produced. There is no flare stack or combined heat and power unit on site. The excess gas produced is vented to the atmosphere. This is not a recommended practice, for safety and environmental purposes. There should be at least a flare stack on site so the gases could be burnt before reaching the atmosphere.

PLATE 3.3.5 GAS HOLDER & GAS COMPRESSOR FOR MIXING

ENERGY GENERATION As discussed above all methane is used for reactor heating or vented to atmosphere. There is no electricity generation on site.

DIGESTED SLUDGE TREATMENT/DISPOSAL The digested sludge is stored in a sludge holding tank before being dewatered by two double belt sludge belt presses (Figure 3.3). The solids concentration leaving the presses is approximately 30%. These solids are then taken away to Letterkenny sewage treatment plant where they are spread on a specific allotment of land and burnt.

ODOUR CONTROL None

DIGESTER MONITORING There is no laboratory staff or technicians on site. Samples for testing are carried out at a laboratory in Letterkenny. These tests are normally solids, volatile fatty acids and alkalinity and samples are monitored twice weekly. There is a portable pH unit on site and the pH is measured before and after digestion every day. The methane composition of the biogas is also measured once a week. The only online monitoring device used in this digester is for temperature. There is a temperature probe situated mid way up the wall of the digester.

PLANT MANAGEMENT There is only one full-time staff member on-site at the Buncrana plant and this is the plant operator. The urban district council staff carries out any maintenance that is necessary on a part-time basis. The plant is clean and well maintained.

In 1993 the Weston Report recommended that the existing anaerobic digestion treatment at Buncrana should be expanded to cater for the sludge generated in the area by the year 2013.

If this recommendation is to be carried out, the Buncrana site will not be large enough and a new site will have to be located. However no decision has been taken on this recommendation yet. This decision may depend on the operation of the Buncrana digester. Since the construction and commissioning of the plant, the digester at Buncrana has not being reaching design performance level at any stage. Operational difficulties include low pH values, little or no gas production and odourous sludge and presently (2000) it is decommissioned for remedial works, including the repairs of holes in the roof due to corrosion. The future of the Buncrana plant is thus in doubt.

3.4 GREYSTONES SEWAGE TREATMENT PLANT

HISTORY Greystones is a town in Co. Wicklow, situated on the East Coast of Ireland, lying just south of Dublin City. Like Buncrana, Greystones lies on the coastline and its potential growth in tourism is based in no small part on the quality of seaside amenities available in the town. The importance of sewage treatment for such a town has come into greater focus in recent years with the enactment into Irish Law of Environmental Legislation. Specifically, the Bathing Water Directive and the Urban Wastewater Directive have special relevance in this context.

Historically, Greystones was served by two treatment works - one in the north and the other to the south of the town. Due to a large population increase (Greystones is now a satellite town for Dublin) and the development of Greystones as an important coastal town the sewage treatment system needed to be upgraded. It was decided to replace the old system with a totally new treatment works to the south of the town at Killincarrig. The collection system was radically modified to provide a new configuration of pipelines, leading to the site of the

new treatment works in Killincarrig. The first few months of 1996 saw the completion of the sewerage scheme and the treatment plant with the final tie-ins leading to the de-commissioning of the two existing treatment plants later that same year.

PLANT LOADING AND SIZE The new sewage treatment works was designed for a 30,000 P.E with provisions made for a further 10,000 P.E. Special attention has been given to the aspects of noise, odour and visual impact in the design of the works. Where practicable all equipment was housed in buildings with suitable soundproofing. All processes leading to odour were enclosed and gas was collected and scrubbed before emission to the atmosphere.

The plant is just treating Greystones, which has a population of approximately 13,000, which is all-domestic with no significant industry. The town is known as a sleeper town on the outskirts of Dublin. People come to sleep in Greystones and work in Dublin. There is one hotel and two public houses in Greystones. This leads to large fluctuations in flow on a daily basis and this causes problems for the treatment works.

The existing plant is designed for 30,000 P.E and when the additions are made in the future, it will be able to take 40,000P.E. However, at the moment there is only approximately 13,000 P.E entering the plant. This low flow to the plant has caused some operational difficulties at the plant and some processes have yet to reach optimum performance conditions. Figure 3.4 shows the flow diagram of the wastewater treatment stream.

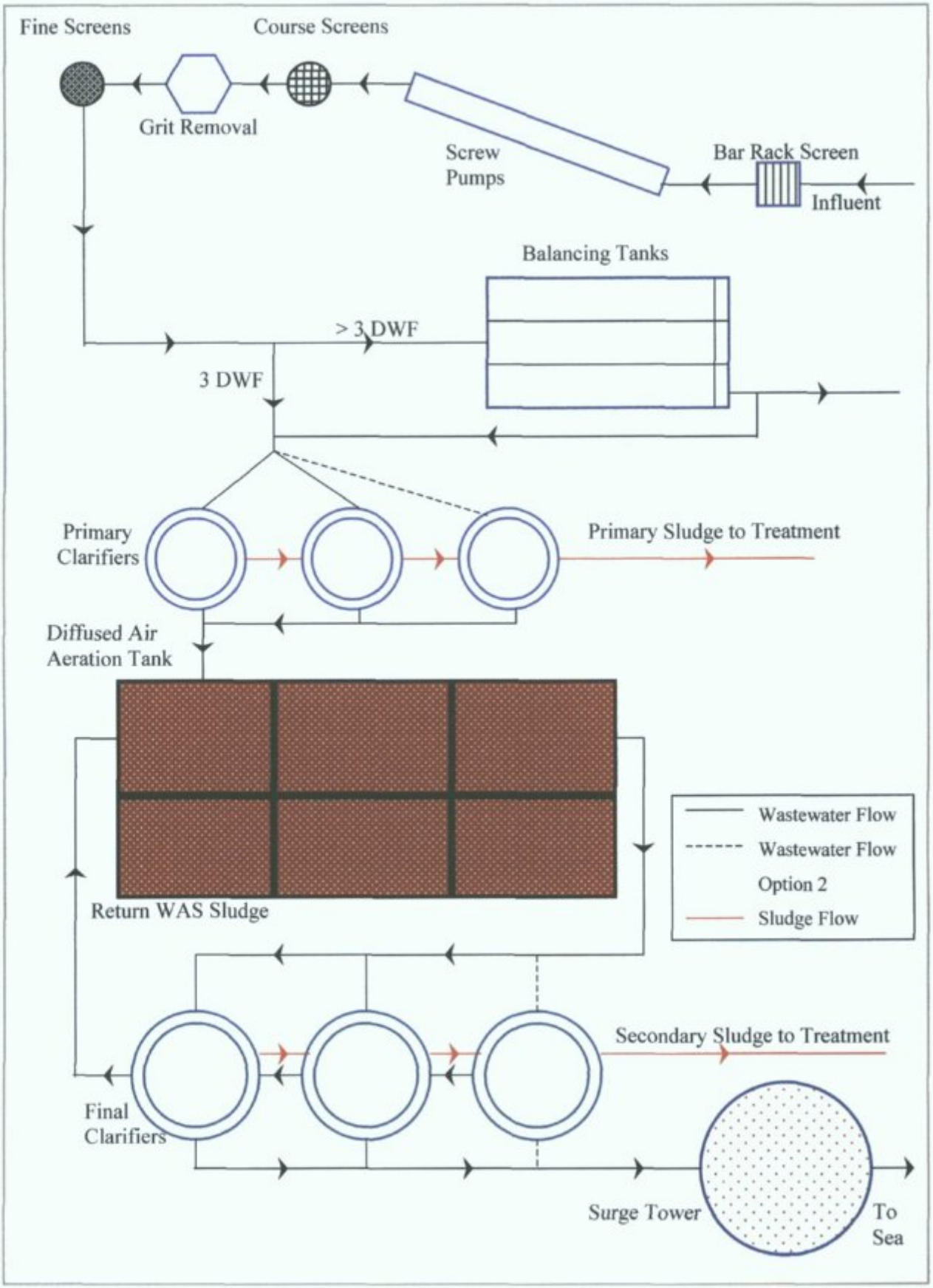


FIGURE 3.4 GREYSTONES WASTEWATER FLOW DIAGRAM

PRELIMINARY TREATMENT The entire preliminary works is housed in Building 1 (Plate 3.4.1). The influent flowing in to the works is screened using rotating bar interceptors followed by main lift pumping (3 No. archimedeaen screw pumps), similar to those found at Tullamore sewage treatment plant. The influent then flows through a flume measurement device and is measured before going on to the course bar screens and the grit trap (Figure 3.4). The course screens are curved bar screens with 12mm spacing. The grit falls into a sump, where air blowers remove the grit.

After this there are two fine screens (Figure 3.4). They are large drum screens that are constantly rotating. The flow enters the screens and is meant to drop through the screens, however because they are 5mm screens they tend to block easily.

The entire preliminary works is designed to treat 12 DWF. If 12 DWF enters the treatment work all is coursed and fine screened, and the grit is removed. Then after this 9 DWF of the 12 DWF will go back to the storm tanks and the remaining 3 DWF will go into the plant for secondary treatment. So, in extremely heavy flow, most of it will go through the preliminary treatment and then only 3 DWF will undergo primary and secondary treatment further on in the plant. The rest will be stored in the storm tanks (balancing tanks) (Plate 3.4.1) until such time as the flow coming into the plant will decrease. When the flow decreases the pumps in the storm tank will pump the stored effluent back into the works for further treatment.

PRIMARY TREATMENT There are three 17.3m diameter primary settlement tanks (clarifers) (Figure 3.4.1) with provisions made for a fourth. The flow comes from the inlet works and into a splitter chamber which divides the flow evenly to the 3 clarifiers. There are probes in these tanks to detect the sludge blanket level. When the sludge blanket level is

detected the bellmonth drops and the hydrostatic pressure pushes the sludge out. This primary sludge then goes into a sump.

PLATE 3.4.1 ADMINISTRATION AND PRELIMINARY TREATMENT BUILDING & BALANCING TANKS

SECONDARY TREATMENT The effluent that overflows from the clarifiers goes on into the aeration tanks. There are 6 cells in the aeration tanks but only 3 are in use. The size of the entire aeration tank is 12m long, 12m wide and 6m deep. Aeration method is by fine bubble diffused air pods in the bottom of each cell. The cells are used in sequence. There are DO probes in each of the cells and a MLSS probe at the outlet. In building 2, air blowers are located to supply the aeration tank requirements.

There are three 20.0m diameter secondary settlement tanks with provisions made for a fourth. The flow comes from the aeration tank and into a splitter chambers where it is distributed equally between the tanks. There are probes in these tanks to detect the sludge blanket level. When the sludge blanket level is detected the bellmonth drops and the hydrostatic pressure pushes the sludge out.

Final effluent is pumped to the surge tower and from there flows by gravity to a marine outfall of 482mm diameter, approx. 900m long. A storm outfall of 1194mm diameter lies alongside this marine outfall and terminates in a specially designed multi-port diffuser component.

SLUDGE PREPARATION Figure 3.5 shows the flow diagram of the sludge treatment stream. The sludge from the primary sludge sump is pumped to two-covered picket fence thickeners (Figure 3.5 & Plate 3.4.2). These thickeners are glass lines tanks, 4.2m in diameter and 4.2m high. Only one thickener is in use presently. The thickener has a diameter of 4.2m and is 4m high and the sludge is thickened to approximately to 4-5% dry solids.

The waste activated sludge from the secondary clarifers is pumped to Building 3, where it is thickened. There are 2 rotary drum thickeners used. The solids are thickened to approximately 5-6% DS.

The thickened primary sludge and the thickened waste activated sludge are pumped to the same covered mixing tank. Both sludges are blended together in this mixing tank before entering the anaerobic digester. There are 2 mixing tanks but only one is commissioned. The tanks are glass lined with a 7.68m diameter and are 2.8m high.

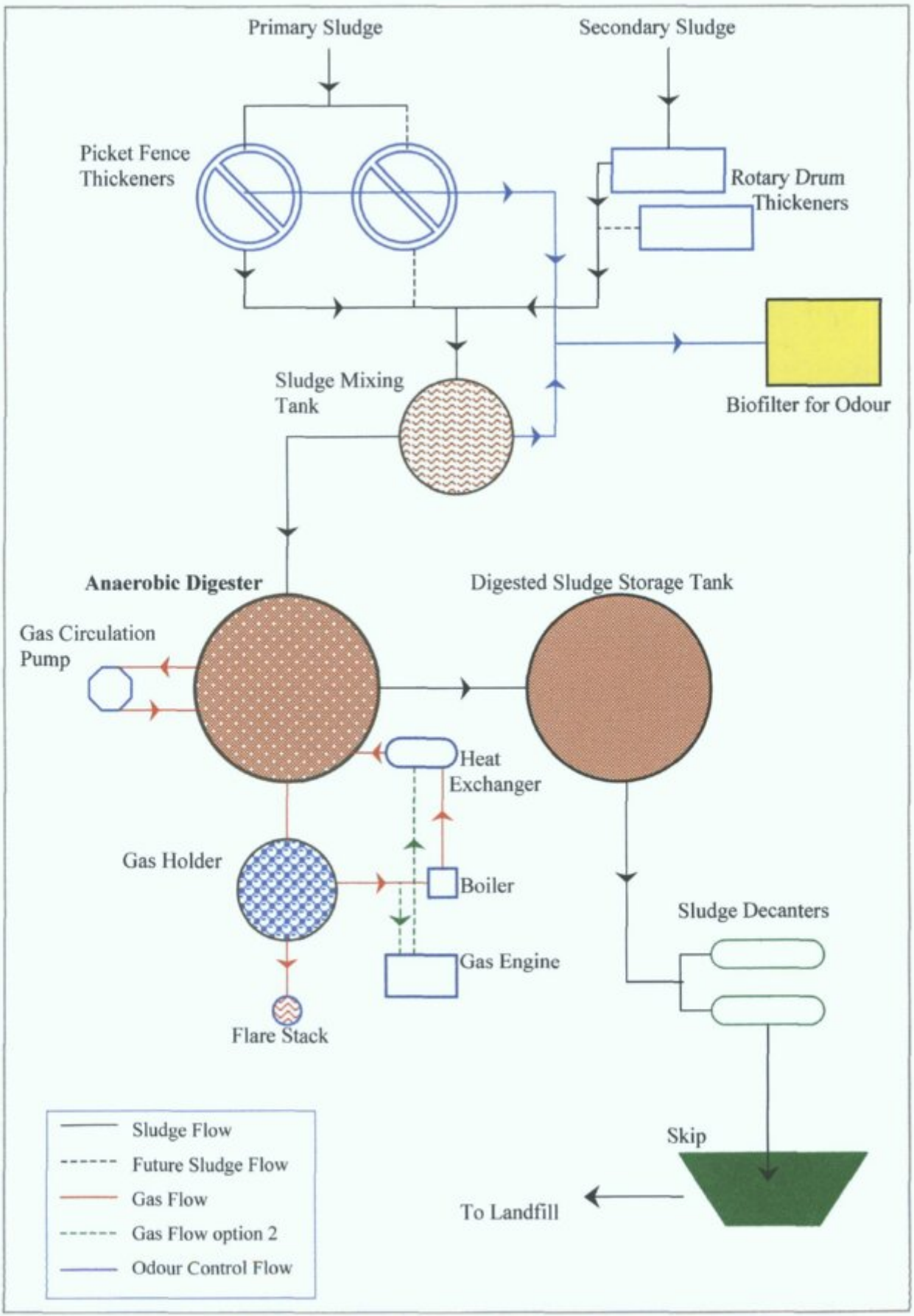


FIGURE 3.5 GREYSTONES SLUDGE TREATMENT FLOW DIAGRAM

PLATE 3.4.2 THE TWO PICKET FENCE THICKENERS

SLUDGE TREATMENT (ANAEROBIC DIGESTION) The anaerobic digester is a gas mixed mesophilic anaerobic digester (Plate 3.4.3). The digester has a 4.2m diameter and is 8.4m high with a fixed roof and a sludge capacity of 570m^3. The digester is constructed from reinforced steel glass and is insulated with 200mm rockwool on the side-walls and roof.

PLATE 3.4.3 GREYSTONES ANAEROBIC DIGESTER

DIGESTER FEEDING The digester is automatically fed intermittently once every hour. Mono pumps with a capacity of 15 m^3/hr pump the mixed sludge to the anaerobic digester. These digester pumps are of the positive displacement type (1 duty and 1 standby). The feed sludge enters at the recirculation line before the external heat exchanger. Therefore the sludge is heated fully before entering the digester at the bottom.

DIGESTER HEATING One external concentric-tube hot water heat exchanger unit (Plate 3.4.4) heats the contents of the digester. It consists of 4-heat exchangers, 2 on the bottom and 2 on the top. Each exchanger is approximately 1.5m long with a 225mm diameter pipe. The sludge travels in sequence through each exchanger starting at the bottom and up to the top. The exchangers consist of two co-axial tubes with sludge flowing through the inner tube and hot water flowing through the outer tube in opposite directions. Experience has shown that the inner tube must be at least 100mm diameter, which reduces the likelihood of blockages but gives less efficient heat exchange. (Manual of British Practice, 1979).

Problems occurred with this heat exchanger unit and modifications had to be carried out. During cold weather periods the exchanger was struggling to keep the contents of the digester heated to the optimum temperature. It was assumed that the depth of the sludge in the exchanger was not allowing the inner part of the sludge to be heated, therefore the inner tube as designed was too large. To correct this a plastic bore tube was placed in the centre of each heat exchanger so as to reduce the size of the inner tube were the sludge passed. This solved the problem of insufficient heating, however the modification caused a slight problem. The solid bore is causing blockages to occur at the corners of the heat exchanger when the sludge is leaving the bottom exchanger and entering the top. Normally the exchanger has to be cleaned out once every six months to prevent a serious blockage occurring. When this

cleaning occurs the digester contents has to go without heat for up to 4-5 hours as there is no standby heat exchanger unit. If the temperature decreases too much this would have severe repercussions for the digestion process. A standby heat exchanger is always recommended in case of exchanger service or breakdown. The sludge is circulated constantly to the heat exchanger from the digester by two mono positive displacement pumps (1duty and 1 standby) with a capacity of $37m^3/hr$.

One water boiler (Plate 3.4.4) heats the hot water fed to the heat exchanger unit. This boiler was dual-fired i.e. was run on either methane or propane fuel. Currently there is no standby boiler in the case of existing boiler breakdown. The hot water is then circulated to the heat exchanger unit by centrifugal type hot water feed pumps (1 duty and 1 standby).

Two temperature probes in the digester automatically control the boiler. These two probes also give a good indication of mixing within the digester. The reading from both probes should be constant for good adequate mixing.

PLATE 3.4.4 THE HEAT EXCHANGER, BOILER AND CHP UNIT

DIGESTER MIXING The digester is mixed by gas injection and is similar to that of Buncrana. The only differences being: the pipes enter the digester at the top and go down the side walls of the digester, eventually turning out towards the centre of the digester and the gas entering the digester is controlled by solenoid valves.

GAS STORAGE AND TREATMENT There is one gas holding tank (Plate 3.4.5) situated beside the digester. It has a bell over water roof and had a gas capacity of 16.44m^3. The level in the gas holder is controlled by a level switch. When the gas holder gets to 95% of its capacity the flare stack cuts in, flaring excess gas until the level in the gas holder is back down to 75% of its capacity. It also has a breathing safety valve. Before the gas reaches the gas holder it goes through condensate pots (Plate 3.4.5), which removes moisture from the gas. Gas from the gas holder is used in the boiler to heat hot water from the heating unit. Before the gas enters the boiler it again goes through another condensate pot. These pots are cleaned once a week as they can become blocked trapping the gas in the digester. This can be a very dangerous situation. There is no gas scrubbing or filtering. This is causing problems for the gas compressor, as the compressor had to be cleaned recently because there was scaling building up.

There is a pressure release valve and a vacuum valve (Plate 3.4.5) located on the top of the digester in case a blockage may occur in the gas pipes or in case a vacuum may occur in the digester. These are vital safety devices associated with anaerobic digesters.

PLATE 3.4.5 GAS HOLDER, CONDENSATE POTS & VACUUM AND RELEASE VALVE

ENERGY GENERATION There is a CHP unit on site however it was still not commissioned at the time of writing. The reason for this is that there has never been enough gas produced.

DIGESTER SLUDGE TREATMENT/DISPOSAL There is a sludge holding tank beside the digester (cold digestion or secondary digester) to provide approximately 15 days storage. This digester is not mixed or heated. It is a sealed tank with carbon active units at the top to eliminate any smells that may be coming from this tank.

The digested sludge is then sent to 2 decanters. The sludge is dosed with polymers just before going into the decanters. The DS content of the decanted sludge is 22%, which goes to the sludge skips and then on to landfill. Laboratory experiments undertaken by the material research agency, Teagasc, have shown that this sludge is suitable for spreading on land.

However, as currently $60m^3$ of sludge is produced per month the volume is considered insufficient to warrant a landspreading option.

ODOUR CONTROL The gas from the picket fence thickener and the mixing tank is abated, by passing it through a shell biofilter before being discharge to the atmosphere (Figure 3.4.2).

DIGESTER MONITORING There is a fully-equipped laboratory on site where all the testing is carried out by the plant operator. The plant operator has found it difficult to carry out both the daily operation of the plant as well as the daily testing due to the time demands involved. For example if a problem occurred in the plant the daily testing would not be a priority and would sometimes be neglected.

There is a pH probe in the digester, however there is no online adjustment if the pH drops. The pH was also monitored with a portable probe every day in the digester influent and effluent. There are two temperature probes, one in the upper and one in the lower section of the digester. There is a third temperature probe in the recirculation line. Other parameters measured included solids (influent and effluent), volatile fatty acids and alkalinity in the digester effluent. These parameters are measured between two and three times a week. The biogas volume and the percentage methane were measured every day.

PLANT MANAGEMENT During normal operation of the plant there are four full-time staff working: one plant operator, one mechanical/electrical engineer and two maintenance personnel. The plant operator did express the need to have a laboratory technician on site so, that the monitoring of the plant would not be neglected when other problems arouse. On a visual level the plant was very clean was well maintained.

Due to persistent problems throughout the plant mainly caused by the low and fluctuating flows, the local authority have currently sub-contracted the management and operation of the plant to outside contractors to try and re-commission the plant. However the above personnel still remain in their respective employment on-site.

3.5 CLONMEL SEWAGE TREATMENT PLANT

HISTORY Clonmel town is situated on the River Suir, with the Comeragh Mountains to the south and Slievenamon to the east, in the south of Ireland. The Suir has been a major influence on the town's development. Clonmel is an expanding industrial town and a thriving commercial and business centre with many fine hotels, pubs and guesthouses.

The sewage treatment plant was constructed between 1996 and 1998, and is located to the south west of the town. Prior to the existing plant there was no sewage treatment in the town and raw sewage was discharged to the river Suir. Due to Urban Wastewater Directives this kind of discharge is no longer permitted. Clonmel town was required to install a wastewater treatment plant before the River Suir became seriously polluted. Figure 3.6 shows the flow diagram of the wastewater treatment stream.

PLANT LOADING AND SIZE The treatment plant as outlined in Figure 3.6 was designed for a P.E. of 60,000 – 70,000. The maximum that can be received by the plant 100 l/s. When the flow reaches 300 l/s, storm sump weir opens up (just before the primary clarifers) and takes the flow to the balancing tanks, similar to those in Greystones. After the high flows have abated the wastewater in the balancing tanks can re-enter the treatment works.

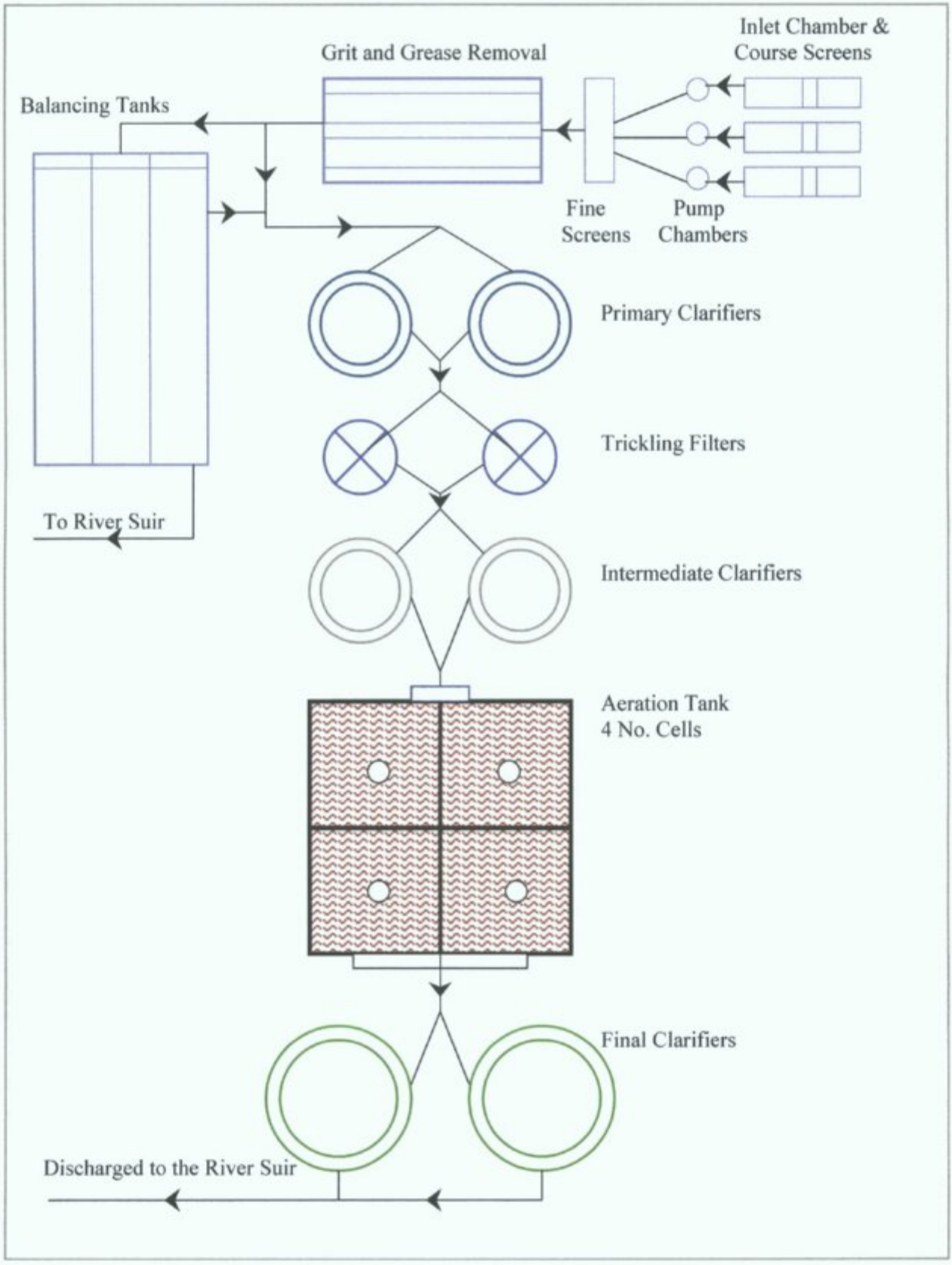


FIGURE 3.6 CLONMEL WASTEWATER TREATMENT FLOW DIAGRAM

PRELIMINARY TREATMENT There are three separate inlet chambers (Figure 3.6 & Plate 3.5.1). The first chamber takes the flow from the town, south of the river. The second chamber, the main sewer (inceptor chamber) takes the flow from the town centre and surrounds, and the third chamber (northern and eastern chamber) takes the inner relief road. The flow in the eastern sewer is mainly industrial. On entering the inlet works the wastewater flows through four course screens. As the flow passes through these screens debris is collected on the surface of the screens. The course screened wastewater then falls into a sump in the pumping chambers (Plate 3.5.1). The north-east chamber contains 3 duty submersible pumps and 1 standby, as does the interceptor chamber. The south chamber contains 2 duty submersible pumps and one standby. These pumping chambers are very deep and sensor probes, which indicate to the pumps when to cut-in and out control the wastewater level.

There are 3 fine screens situated at the top of the inlet chambers. These screens remove fine particles contained in the wastewater flow. From here the wastewater flows by gravity through the remaining processes.

Grit and grease is then removed (Figure 3.6). Both of these processes take place in a big rectangular tank called the Hartman Tank (Plate 3.5.1). There is a bridge on top of the Hartman tank, which travels over and back along the tank. The movement and the shape of the tank helps to settle the grit to the bottom and allows grease to float to the top. The grit is collected, by two submersible pumps running along the bottom of the tank, and pumped out to a grit classifier. The grease floating on the top of the tank is sent to a side channel. Here it is scrapped into a hopper and then into a grease holding tank. The grease in the sludge holding tank in used in the digestion process. There is a tanker on site, which transports the grease to the digester. The grease is conveyed using grease-loading pump to the digesters.

PLATE 3.5.1 INLET CHAMBERS, TOP OF SUBMERSIBLE PUMPS & HARTMAN TANK

PRIMARY TREATMENT The wastewater flows into the primary influent channel. Located along this channel is the storm dump weir. The storm dump weir discharges the water into the balancing tanks. If neither stream can take the flow or the pumps break down there is another emergency channel, which flow straight to the river. There is mostly a combined sewer in town so there is a storm overflow chamber along the sewer as well, used in the case of emergencies.

From the primary flow channel the wastewater flows into 2 large primary clarifiers (10.6m) (Figure 3.6). Solids settle out and the supernatant discharges over the weirs and onto 2 trickling filters (biotowers). The wastewater is lifted here again so as it will flow by gravity through the remaining wastewater treatment units. After the biotowers the wastewater flows into two intermediate clarifers where settlement of the solids occurs. The supernatant overflows from these clarifiers and into the aeration tanks.

SECONDARY TREATMENT There is one square aeration tank containing 4 aeration cells with centrally located mechanical aerators. Following aeration the effluent flows into two

final clarifiers. Here the remaining solids settle out of suspension before the effluent leaves the clarifiers and is discharged to the River Suir. Along this final effluent channel before the effluent leaves the plant, there is a measurement flume, measuring the effluent flow, a pH probe, and a temperature probe. These parameters are connected up to an alarm system, in the main control room.

SLUDGE PREPARATION Figure 3.7 shows the flow diagram of the sludge treatment stream. The sludge from the primary, intermediate and secondary clarifiers is stabilised by two mesophilic primary anaerobic digesters and one secondary digester (storage). Prior to digestion the primary and intermediate sludge is thickened in a picket fence thickener. The picket fence thickener (Plate 3.5.2) is a cover concrete tank, with a sludge capacity of 185m^3.

The sludge from the final clarifier is thickened in one sludge thickener (Plate 3.5.2) in the sludge treatment building. After thickening this sludge is pumped, to a blending tank (Plate 3.5.2) where it meets the thickened sludge from the picket fence thickener. The tank is fitted with vertical axis agitator to blend the incoming sludge. The tank is constructed from concrete and is covered. It has a sludge capacity of 35m^3.

PLATE 3.5.2 PICKET FENCE THICKENER, SLUDGE THICKENER & BLENDING TANK



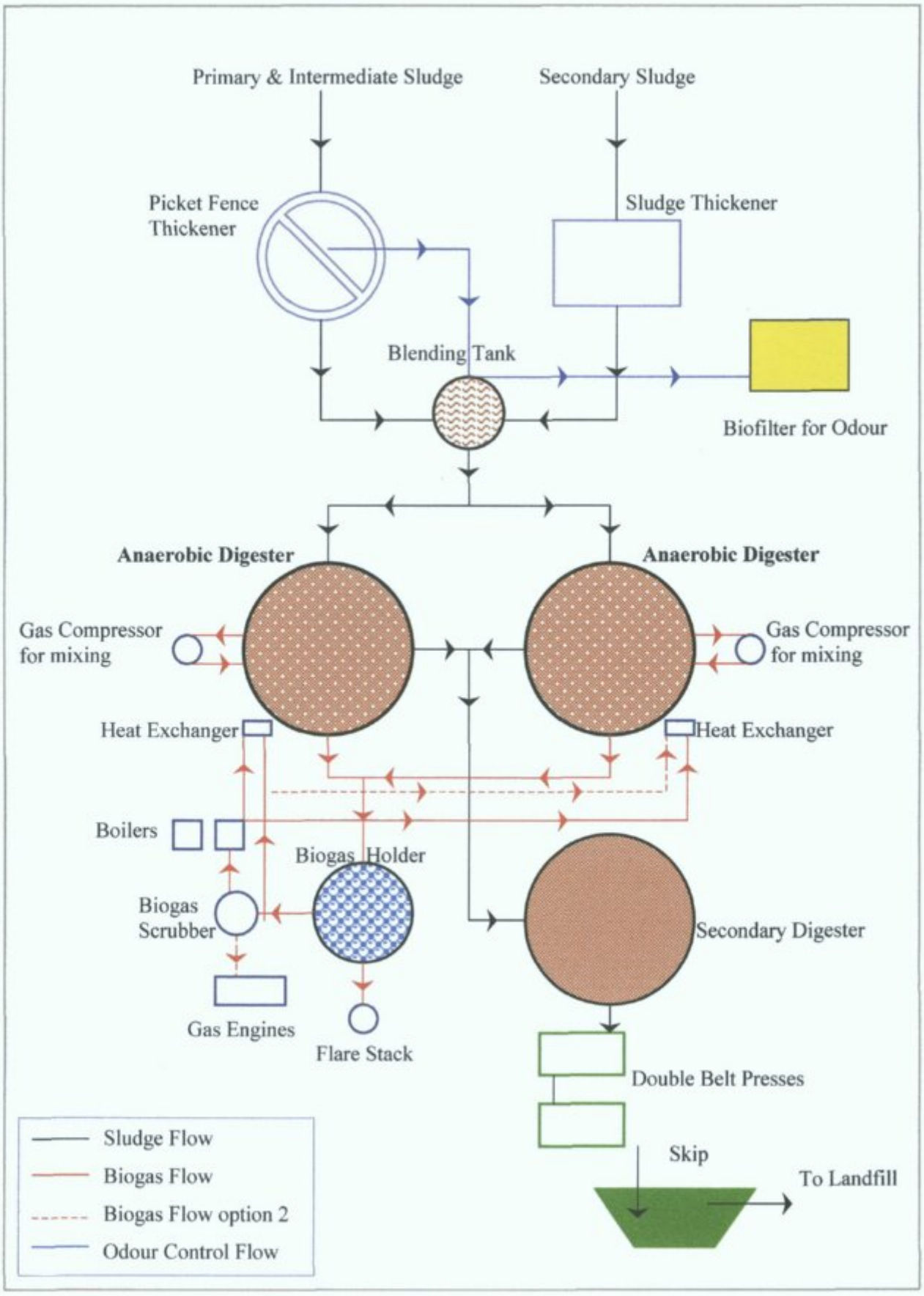


FIGURE 3.7 CLONMEL SLUDGE TREATMENT FLOW DIAGRAM

SLUDGE TREATMENT (ANAEROBIC DIGESTION) There are two anaerobic digesters in Clonmel (Plate 3.5.3). They are the only digesters to be constructed from concrete (side-walls and roof) and are the largest digesters treating sewage sludge in Ireland, with a sludge capacity of $800m^3$ each. They have a height of 10m, of which 4m is underground and an internal diameter of 10m. The walls are 400mm thick with a roof thickness of 200mm. The digesters are insulated with 100mm of mineral wool fibre.

PLATE 3.5.3 CLONMEL ANAEROBIC DIGESTERS

Currently, one anaerobic digester operates at full capacity and the other is operating at approximately half capacity. The digesters are working in parallel. The digester operating at full capacity is fed for 30 minutes every two hours. The operation of the pumps is automatically controlled by the PLC installed in the digestion control panel. The pumps are designed and controlled to cease if there is a low level in the blending tank and if there is a high level in the digester-feeding pit.

DIGESTER FEEDING The blended raw sludge is pumped from the sludge-blending tank on a semi-continous basis by two groups of duty/standby pumps to the anaerobic digesters. Each digester has 1 duty and 1 standby pump. The pumps provided are progressive cavity type, driven by a speed variator to allow a flow rate range from 2.4 to 12m^3/hr. Also grease from the inlet works can be added to the digesters through separate grease loading pumps. This process only takes place 4 times, as only a small quantity of grease is contained in the wastewater.

The inlet sludge is mixed with the recirculated heated digested sludge in the feeding pit located at the top digester and then conveyed by gravity to the sludge mass.

DIGESTER HEATING The sludge digestion temperature is kept constant by external heat exchangers, one for each digester (Plate 3.5.4) through which the sludge recycled from the digesters bottom is recirculated and heated with hot water. The heat exchangers are jacketed pipe type, in which the hot water is pumped counter-current to the sludge flow through a concentric pipe surrounding the sludge pipe. This system is similar to that of Greystones. Two horizontal centrifugal pumps (Plate 3.5.4), one for each digester provide recirculation of the sludge, from the digester through to the heat exchanger and back to the digester inlet pit.

PLATE 3.5.4 HEAT EXCHANGERS WITH SLUDGE RECIRCULATION PUMPS

The hot water used to heat the sludge circulating through the heat exchanger is generated by the energy recovery system from the two gas engines (CHP unit) and/or in a boiler fed by methane gas or propane gas. There are two boilers on site working in parallel and both are double feed (methane or propane gas). The hot water is circulated to the heat exchanger by means of two horizontal centrifugal pumps, one for each digester.

DIGESTER MIXING Biogas produced is taken from the top of the digester, compressed and injected through a series of 4 drop pipes (Plate 3.5.5), located on the roof, descend vertically to the bottom of the digester; in this way big gas bubbles go through the sludge mass, mixing the contents of the digester. Drop pipes are uniformly located inside the digester along one circumference. One rotary vane compressor (per digester) and one on standby (Plate 3.5.5) recirculate the biogas. The biogas passes through a gravel filter (Plate 3.5.5) to remove all condensate and impurities in the biogas to protect the rotary vane compressors. Solenoid valves (one for each pipe), for sequential distribution of the gas to the digester gas mixing distribution pipes, are provided. The mixing system operates on a continuous basis, 24hr/d.

PLATE 3.5.5 GAS DROP PIPES ENTERING AT THE TOP OF THE DIGESTER, GAS COMPRESSORS & GRAVEL FILTERS

GAS STORAGE AND TREATMENT The biogas produced during sludge digestion, recovered from the dome at the top of each digester is partly sent to storage digestion and partly recycled to mix the sludge in the digestion tank.

Biogas produced in the digestion phase of both digesters passes through gravel filters and from here is fed to one common gas-holder with a floating cover (Plate 3.5.6). The gas-holder is composed of a concrete tank (8m diameter and 4m height) with a monolithic constructed mild dome with vertical guides. It has a total capacity of 200m^3, which is equal to approximately 6 hours storage capacity. The floating cover is equipped with a breathing (double function) safety valve, operating both for high or low pressure. Also the floating cover is equipped with high, very high, low and very low level switches.

In parallel to the gas holder, a safety flare (Plate 3.5.6) is installed to burn excess biogas in case of high gas production or lack of consumption in the gas engines. The flare system automatically lights up when the very high level is reached on the gas holder floating cover, which also opens the automatic valves of biogas feeding to the flare.

Before reaching the gas run engines the biogas is purified of the sulphurous deviatives (H_2S) by passing through one gas scrubber. The biogas scrubber is composed of:

- Washing tower (Plate 3.5.6)
- Water storage tank and washing pump
- Sodium hydroxide (NaOH) storage tank and relevant dosing pump (Plate 3.5.6)
- Heating system installed in the NaOH storage tank.

This purification will lengthen the life of the CHP unit and the boilers.

PLATE 3.5.6 GAS HOLDER AND FLARE STACK, WASHING TOWER & Na_2CO_3 TANK

ENERGY GENERATION In addition to methane boilers a CHP unit is on site. This unit consists of a gas engine coupled to a generator. There are two such units on site one duty and one standby. Thermal energy is recovered from the flue gas discharge and cooling water/oil system. On the water pipe coming from the heat exchanger an emergency cooling system (radiator) has been installed. The emergency cooling system is necessary to ensure the gas

engine inlet water temperature does not get to high especially during the summer months to avoid overheating of the digester contents. This system was commissioned shortly after the digesters were commissioned, when enough gas was being produced.

DIGESTED SLUDGE TREATMENT/DISPOSAL Digested sludge is withdrawn from the two digesters and sent by gravity to a holding tank, which is acting as intermediate storage (Figure 3.7). The digested sludge holding tank may be converted to a digester at a later date without civil/structural alternations, as it is identical in size and construction to the other two digesters. Since the function of this tank is sludge storage only, no mixing or heating is provided. However, biogas can still build up in this tank and such gas can be released from the tank by extraction with a line connected to the primary digesters gas. A safety overpressure valve is installed, with solenoid valves provided to vent biogas in case the safety overpressure valves fault.

The sludge flows by gravity by means of telescopic valves from the storage tank to the dewatering building. The digested sludge is dewatered by 2 double belt presses (Figure 3.7) and disposed of to landfill. Plans for lime stabilisation of the digested sludge are being looked into before sludge will be spread on land.

ODOUR CONTROL The gases from the picket fence thickener and the blending tank are trapped in the tanks and pipes to a shell biofilter (similar to that of Tullamore) where the gases are neutralised.

DIGESTER MONITORING There is a laboratory on site where some tests are carried out. An external contractor commissioned the digestion plant and they carried out the necessary

start-up procedure. The only online monitoring and control for the digesters are temperature probes. The digester influent and effluent pH is measured each day. The solids in the digester influent and effluent are measured 3 times a week and the volatile acids are measured once a week. The percentage methane in the biogas is sampled twice a week.

PLANT MANAGEMENT Currently the plant is only fully commissioned for approximately one year. The plant is being managed and operated by an external contractor who also designed the plant. Clonmel Corporation will eventually get the plant handed over when they are satisfied that the plant in running correctly. The plant was very clean, well landscaped and maintained well.

3.6 TRALEE SEWAGE TREATMENT PLANT

HISTORY Tralee Co. Kerry is located in the south west of Ireland. Like Buncrana and Greystones the town is situated on the coastline. The population of Tralee is approximately 18,000 - 20,000 and growing. The town has a number of large industries and has a triving tourism trade. There was no sewage treatment in the town before the existing plant was commissioned in 1998. The raw wastewater from the town was being discharged to sea.

PLANT LOADING AND SIZE The treatment plant was designed for a P.E. of 42,000 and is currently treating approximately 25,000. The plant consists of primary and secondary treatment with anaerobic digestion treating the sludge. Figure 3.8 shows the flow diagram of the wastewater treatment stream.

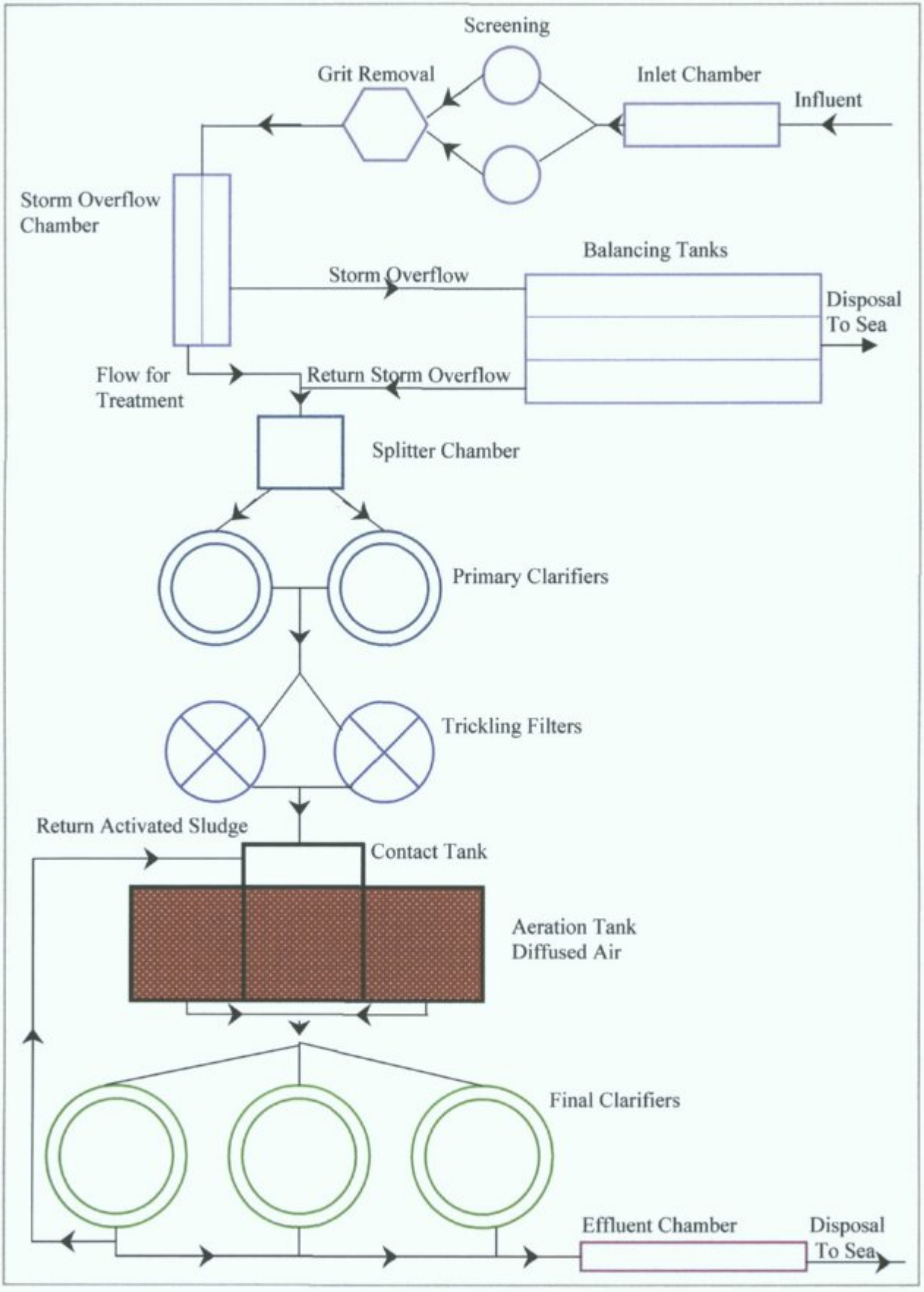


FIGURE 3.8 TRALEE WASTEWATER TREATMENT FLOW DIAGRAM

PRELIMINARY TREATMENT The wastewater enters the treatment works through the inlet channel. The total flow entering the treatment works goes through the preliminary works. It enters the two vertical drain fine screens and a grit trap before entering the storm overflow chamber (Figure 3.8). Here the flow if greater than 3 DWF overflows to the balancing tanks. These storm tanks (balancing tanks), like Greystones and Clonmel, have three compartments with three tipping buckets for cleaning purposes when emptying. It also has a sea out fall for extreme flows. The treated effluent is used for the tipping buckets for cleaning the tanks. So in extremely heavy flow all wastewater will go through the preliminary treatment and then only 3 DWF will undergo primary and secondary treatment further on in the plant. The rest will be stored in the storm tanks until such time as the flow coming into the plant will decrease.

PRIMARY TREATMENT The wastewater then flows into a splitter chamber where the wastewater is split equally between two large primary clarifiers (22m). Here the primary solids settle out and the supernatant overflows onto two trickling filters (Plate 3.6.1) . These are uncovered 23.6m diameter filters in which the wastewater flows down through filter media.

SECONDARY TREATMENT After filtration, the wastewater flows into a large aeration tank (Figure 3.8 & Plate 3.6.1). The aeration tank contains 3 aeration cells with a total volume of 840m^3. The tanks operate in parallel, with a contact tank before the aeration process. Aeration method is by fine bubble diffused air pods in the bottom of each cell.

After aeration the wastewater enters three secondary clarifiers where final settlement takes place before the supernatant flows out to sea. After leaving the final clarifiers, the effluent

flow is measured by a measurement flume and a composite sample takes samples at regular intervals throughout the day. Before the effluent leaves the plant it flows into a tidal holding tank and is discharged regularly depending on the tides.

PLATE 3.6.1 TRICKLING FILTER, DISSOLVED AIR AERATION TANK & TIDAL TANK

SLUDGE PREPARATION Figure 3.9 shows the flow diagram of the sludge treatment stream. Prior to anaerobic digestion the primary and secondary sludges from their respective clarifiers are thickened in a picket fence thickener (Figure 3.9 & Plate 3.6.2). This thickener is an uncovered concrete tank with an internal diameter of 8.35m and a capacity of approximately 200m^3.

PLATE 3.7.2 PICKET FENCE THICKENER

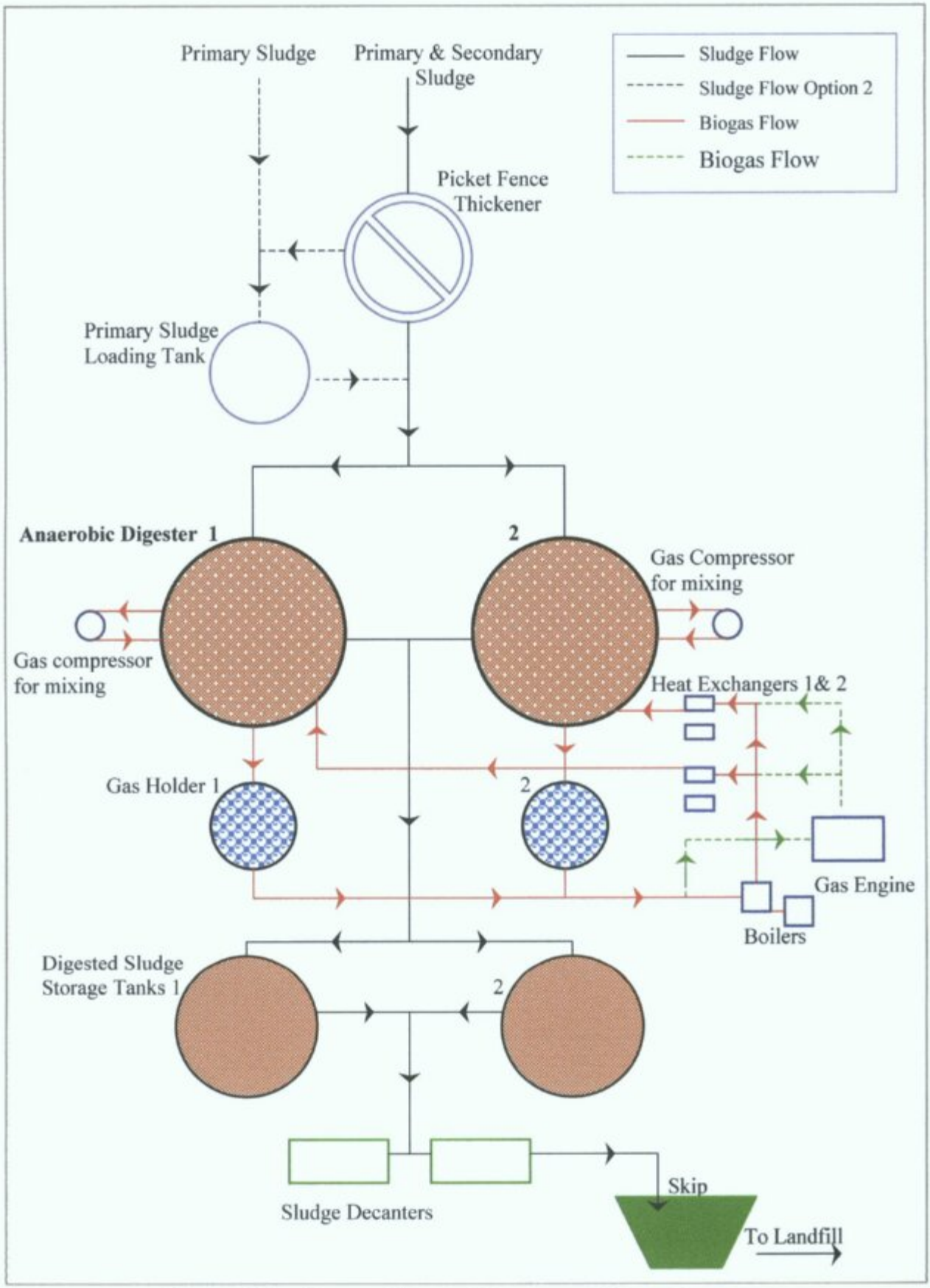
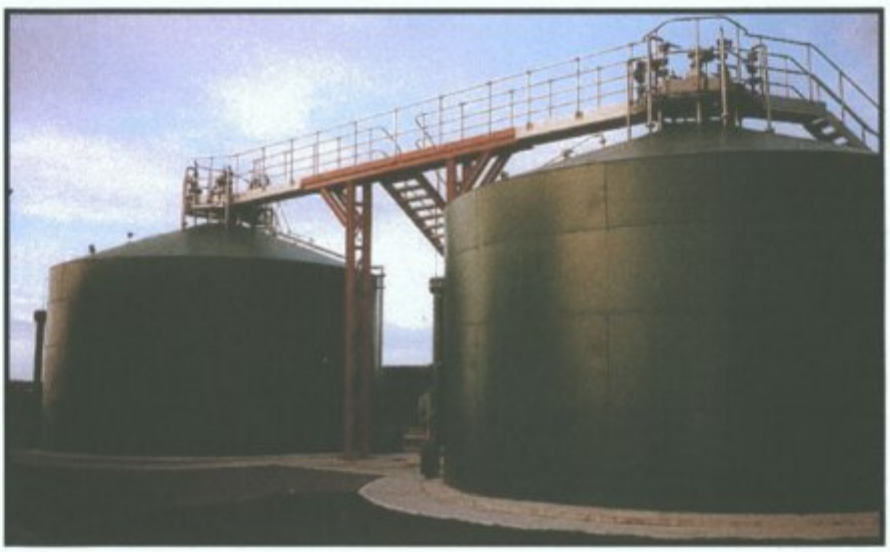


FIGURE 3.9 TRALEE SLUDGE TREATMENT FLOW DIAGRAM

SLUDGE TREATMENT (ANAEROBIC DIGESTION) There are two mesophilic anaerobic digesters (Plate 3.6.3) constructed at Tralee sewage treatment plant. Each digester has a capacity of 400m^3, with a diameter of 8.54 and a height of 8m, of which 4m is underground. The section of the digesters over ground is constructed of glass lined section steel, with the walls and roof insulated with 50mm thickness polyurethane foam, encapsulated in rigid g.r.p so as to be completely waterproof. The underground substructure is constructed of concrete, on which the upper part of the digesters is supported.

PLATE 3.6.3 TRALEE ANAEROBIC DIGESTERS



Presently, only one digester is commissioned. The digesters are designed to work in parallel. The digesters and associated equipment are sized for a fully developed indigenous wastewater treatment plant, which will cater for the domestic and industrial loads emanating from Tralee Town and its environs. A second digestion stream may be provided in future to cater for imported sludge from outside the Tralee area

DIGESTER FEEDING The thickened sludge is pumped from the picket fence thickener and/or primary sludge loading tank by means of variable speed positive displacement sludge pumps, one duty and one standby for each digester. A primary loading tank is available to take sludge in the case of emergencies where the picket fence thickener or the digesters are not working properly.

Sludge is fed to the digester on a time-controlled basis with flow measurement on each of the digester feed lines monitoring instantaneous and total volumes of sludge fed to the digester. The digesters are designed to operate on the displacement method with digested sludge displaced by the incoming sludge and gravitating to the secondary digestion tanks (storage cold digestion). The feed sludge enters the digester in the recirculation line after the heat exchanger. Therefore the cold sludge combines with the heated sludge before entering the digester.

DIGESTER HEATING Heating the contents of the digesters is provided by external heat exchangers, one unit for each digester. Problems occurred with the type of exchangers commissioned in the plant and within a year they were replaced. These exchangers were spiral-tube sludge hot water heat exchangers. These have narrow spiral passages, the casing being circular with flat sides. An advantage of this type of heater is that, with the sludge flowing in a curved path, turbulence is increased which facilitates mixing and results in an increased rate of heat transfer. However, the pathway in these exchangers was too narrow and the sludge got completely blocked and cleaning of this exchanger unit became too regular and time consuming. They were replaced with simple concentric-tube sludge hot water heat exchangers (Plate 3.7.4), consisting of two co-axial tubes with the sludge flowing through the inner tube and hot water through he outer tube in opposite directions. Experience has shown

that the inner tube must be at least 100mm diameter, which reduced the likelihood of blockages but gives less efficient heat exchange. This proved to be a much more efficient system than the former.

A temperature probe in the digester controls digester heating. Sludge will be continually fed through the heat exchanger by sludge recycle pumps (positive displacement type).

The hot water is supplied to the heat exchangers by means of a boiler. There are two dual fired boilers (Plate 3.6.4). The biogas is taken from the gas-holder and compressed by two rotary vane compressors to the boilers. If there is not enough biogas produced the boiler can run on propane gas stored on site. Provisions are also made for the water to be heated by a gas engine (CHP unit) (Plate 3.6.4) to run on biogas from the digester, however this unit is not commissioned presently. When in operation the maximum possible heat will be extracted from the unit to heat water for sludge heating. The waste heat from cooling water, lubricating oil and exhaust gases will be collected and passed through a heat exchanger of which the secondary circuit is connected to the hot water circuit. If the CHP unit is producing more heat than is required for the hot water circuit, the excess heat will be routed to the dump radiator.

PLATE 3.6.4 HEAT EXCHANGER UNIT, BOILERS & CHP UNIT



DIGESTER MIXING Gas mixing is provided by the even distribution of 6 separate unconfined gas mixing pipes around one circumference of the digester. These distribution pipes enter at the top of the digester and drop vertically down to the bottom of the tank. The biogas is taken from the top of the digester and compressed by gas compressors through pneumatic valves on a sequential basis, similar to that of the mixing in the Buncrana digester. A vacuum and pressure release valve is fitted on the top of the digester (Plate 3.6.5).

GAS STORAGE AND TREATMENT There are two gas-holders on site, one for each digester, to store the biogas produced in it the digestion process. The capacity of both gas-holders is $15m^3$. Both of these tanks have a breathing safety valve and level switches that are connected to the flare stack, which cuts in automatically when the level in the gas-holder is high and flare the excess biogas. The flare automatically cuts out when sufficient quantities of excess biogas are burned off.

There are three condensate pots for each gas holder. One on the gas line coming from the digester to the gas holder. One for the gas leaving the gas holder before entering the boilers or gas engine and one for gas leaving the gas holder before entering the other gas holder (in case of an emergency).

PLATE 3.6.5 VACUUM AND RELEASE VALVE, GAS MIXING PIPES & GAS HOLDERS

ENERGY GENERATION In addition to methane boilers there is a CHP unit (Plate 3.6.4) installed on site, however at the time of writing it had not been commissioned.

DIGESTED SLUDGE TREATMENT/DISPOSAL The digested sludge is sent via gravity to two sludge holding tanks. These are uncovered concrete tanks with a sludge storage capacity of approximately $300m^3$ each. From here the sludge is sent to 2 alternating centrifugal decanters where the sludge is dewatered and then disposed of to landfill (Figure 3.9).

ODOUR CONTROL None

DIGESTER MONITORING There is a fully-equipped laboratory on site in which all the daily tests are carried out by the part-time laboratory technician. There is pH and temperature monitoring in the digester, however the pH probes in the digester never worked properly. Each day the pH and solids are measured in the digester influent and the pH, solids, volatile fatty acids and COD measurements are taken for digester effluent. Alkalinity is measured twice a week in the digester effluent. The biogas volume and percentage methane in the gas is also measured every day.

PLANT MANAGEMENT There are three full-time and one part-time personnel employed at the plant. There is one plant operator and two maintenance personnel, and one part-time laboratory-technician. The plant operator emphasised the importance of the lab-technician and would prefer if it was a full-time position. As with all plants surveyed this treatment plant is very clean, landscaped and well maintained

CHAPTER 4

COMPARISON OF FULL SCALE ANAEROBIC DIGESTER PERFORMANCE IN IRELAND

4.1 INTRODUCTION

In 1993 the Irish Government commissioned a strategy study to address the difficulties encountered with sludge management and disposal in this Country. Many Local Authorities have complied with the recommendations of this strategy study by incorporating Anaerobic Digestion of sewage solids in the design of new and retrofitted sewage treatment plants. Irish Local Authorities have little experience in anaerobic technologies compared to the more traditional mechanical and aerobic processes. This in turn led to very little backup available to those persons involved with the design and operation of these processes. However, over the past few years, there is an enhanced confidence in the process as several anaerobic digesters are operating successfully in Ireland.

Several International Consultancies were commissioned to contribute to the design of Anaerobic Digesters in Ireland. Thus the sizing, heating, insulation and mixing systems are similar to those found throughout Europe. Information on process monitoring and operation of anaerobic digesters was found to be a greater problem for Local Authorities in Ireland. Conflicting views concerning the parameters to be measured and the methods employed for measurement emerged during this survey.

The main aim of this chapter is to compare and evaluate different designs, equipment, operation etc., of the anaerobic digestion technology in relation to sewage sludge in Ireland based on the information presented in chapter 3. From the results of this research, recommendations for the design, operation and maintenance of AD plants is presented in Chapter 6.

4.2 COMPARISON OF DIGESTER DESIGNS

The following section compares the different digester designs at the five anaerobic digestion plants treating sewage sludge in Ireland. The design and operation of the five anaerobic digestion plants are summarised in Table 4.1. All designs are gas mixed CSTRs operated at mesophilic conditions.

DIGESTION PLANT SIZE As seen from Table 4.1, the more recent designs (Clonmel and Tralee) are of larger capacity and have two digestion tanks as opposed to one, which allows for greater flexibility in terms of plant maintenance and feeding. Buncrana is the smallest with a digester tank volume of $120m^3$.

DIGESTER CONSTRUCTIONAL MATERIAL AND INSULATION In four out of the five digestion plants the digesters are constructed from reinforced steel glass with insitu concrete bases. The digesters at the fifth plant, Clonmel, are constructed entirely from concrete. Capital costs of concrete digesters are high and restrict the applications of such tanks generally to larger treatment plants (Hobson & Wheatley, 1992). Clonmel is the largest of the treatment plants and this may be a reason why concrete digesters were selected. Another property of concrete digester is that they are self-insulated (Hobson, 1990), however digesters in Clonmel included 100mm of mineral wool fibre as supplementary insulation to the concrete. Smaller sized digesters made from glass-steel are generally 70% cheaper then concrete tanks. These tanks require insulation and this can be internal but is generally external (Wheatley & Hobson, 1992). Of the four digestion plants constructed with glass-steel, all are insulated with external insulation material. Concrete digesters have a higher guaranteed plant life of 40 years compared with 20 years for glass-steel tanks (Wheatley & Hobson, 1992).

TABLE 4.1 COMPARISON OF THE MESOPHILIC GAS MIXED ANAEROBIC DIGESTION PLANTS IN IRELAND

	Tullamore	Buncrana	Greystones	Clonmel	Tralee
Primary Sludge Thickening	Picket Fence Thickener	Picket Fence Thickener	Picket Fence Thickener	Picket Fence Thickener	Picket Fence Thickener
Secondary Sludge Thickening	Mechanical Sludge Thickeners	No Secondary Sludge	Mechanical Sludge Thickeners	Mechanical Sludge Thickeners	Picket Fence Thickener
Volume of Digester(s) m^3	1 No. 330	1 No. 120	1 No. 570	2 No. 800 each 2 commissioned	2 No. 400 each 1 commissioned
Design PE (Current PE)	~ 16,000 (~ 14,000)	~ 11,000 (~ 7,000)	~ 30,000 (~ 12,000)	~ 80,000 (~ 60,000)	~ 42,000 (~ 25,000)
Current Feed Rate m^3/d	17	-	18	40 (digester A) 12 (digester B)	23 (digester 1) 0 (digester 2)
Design Gas Yield m^3/d	~408	-	606	1162 per digester	622 - 809 per digester
Actual Gas Yield m^3/d	400 - 600	-	~ 400	~ 910 cumulative	200 - 300
Gas Holder Volume m^3	1 No. 10	1 No.	1 No. 16.4	1 No. 251.2	2 No. 15
Present Gas Usage	Heating & mixing the digester contents Heating buildings Generates electricity	Heating & mixing the digester contents	Heating & mixing the digester contents Heating buildings	Heating & mixing the digester contents Generates electricity	Heating & mixing the digester contents
Operating pH Range	6.8 - 7.3	-	6.8 - 7.2	7.0 - 7.3	6.95 - 7.1
pH Control if necessary	Lime addition before the PFT	Lime addition	Sodium hydroxide at recirculation pumps	Lime addition at blending tank	Caustic soda addition at recirculation pumps

TABLE 4.1 CONTINUED COMPARISON OF THE MESOPHILIC GAS MIXED ANAEROBIC DIGESTION PLANTS IN IRELAND

	Tullamore	Buncrana	Greystones	Clonmel	Tralee
Hot Water Boilers	2 No. boilers 1 propane & 1 methane	2 No. boilers 1 propane & 1 methane	1 No. Dual fired	2 No. Both dual fired	2 No. Both dual fired
Heating System	8 internal heat exchangers	1 external heat exchanger	1 external heat exchanger	1 external heat exchanger per digester	1 external heat exchanger per digester
CHP unit	2 Gas engines 1 duty/ 1 standby	No CHP unit	1 Gas Engine (not commissioned)	2 Gas engines	1 Gas engine (not commissioned)
Mixing System	Gas injection 12 No.gas distribution pipes	Gas injection 8 No. gas distribution pipes	Gas injection 8 No. gas distribution pipes	Gas injection 4 No. gas distribution pipes	Gas injection 6 No. gas distribution pipes
Gas for mixing is taken from	Gas holder	Gas holder	Gas holder	Top of the digester	Top of the digester
Control valves on gas distribution pipes	Rotary Valves	Pneumatic Valves	Solenoid Valves	Solenoid Valves	Pneumatic Valves
Tank Material	Glass coated steel plates bolted together	Steel Glass Lined	Steel Glass Lined	Concrete Tanks	Steel Glass Lined
Insulation	External twin-skin g.r.p. and polyurethane foam panels	100mm of cladding	Rockwool 200mm	Minerial & Fibre 100mm	Polyurethane 50mm
Scum Removal Facility	No	-	Yes	No	Yes
Grit Removal Facility	Yes	-	No	No	No

DIGESTER HEATING All of the digesters are heated with hot-water heat exchangers in which the hot water is supplied by biogas/natural boilers or a CHP unit. Tullamore has 8 internal heat exchangers (as described in Chapter 3, section 3.2) whereas all of the other plants have external exchangers. Generally, externally heat exchangers are preferred due to easy assess for cleaning and maintenance compared with heat exchangers in the digester. However the internal heat exchangers in the digester at Tullamore have being running successfully for 13 years with little trouble. The plant operator puts their success down to the fact that they are enclosed in an anaerobic atmosphere and are therefore not susceptible to corrosion by air. Over the years the plant operator at Tullamore has had problems with keeping the temperature up during the winter months. His associates this problem with the central location of Tullamore as during the winter months the temperature can get considerably lower than the rest of the Country. In his opinion there is inadequate insulation for this drop in temperature.

There are many different configurations of external heat exchangers. Greystones, Clonmel and Tralee have concentric straight pipes in which the hot water and sludge flows though in opposite directions and the pipes in the Buncrana exchanger is tube-in-shell shaped. All of these are described in detail in Chapter 3. Initially, Tralee had spiral tubed heat exchangers installed, however after one month of digester operation, these had to be replaced because they were constantly blocking. The exchanger at Greystones had to be modified (Chapter 3, Section 3.4) as it was unable to heat the digester contents to its optimum during the winter months. The heat exchanger was either under-designed or the insulation in the digester is inadequate. However, after the modification the exchanger unit worked successfully providing sufficient heat to maintain the digester contents at 35°C during winter months.

External heat exchangers are more expensive (Brade *et al.*, 1982). They need sludge pumps to circulate the sludge to and from the digester. If internal heat exchangers are considered they are preferred in small reactor designs.

DIGESTER MIXING Each digester is gas mixed and mixing is supplemented in all of the reactors, apart from Tullamore, by sludge recirculation through the external heat exchangers. No mixing system is the same and differences include; the number and position of distribution pipes entering the digester, different non-return valves and whether the gas is extracted from the digester roof or the gas holder. Table 4.1 summaries this information. Even though no mixing system is physically the same, the principle of injecting gas under pressure sequentially through the distribution pipes is the same for all reactors. Generally all plant operators were satisfied with the mixing systems, however they expressed the difficulty of actually knowing whether or not the mixing is operating successfully. One operator monitored the mixing by watching the difference in temperature between the top and bottom of the reactor. Another operator expressed the need to keep a close watch on the non-return valves as they can break down leading to no mixing in a section of the digester. It was clear in Buncrana when the digester was decommissioned and empty that the mixing system was totally inadequate as serious stratification had occurred within the digester, actually engulfing the mixing lances, however there was no obvious sign of this problem while the digester was operating.

To reduce clogging problems in gas distribution pipes provisions should be made for flushing the gas lines with high pressure water (U.S., EPA, 1979).

There is no mechanically mixed digester in Ireland, however in Countries such as Denmark mechanical mixing with propellers is preferred to gas mixing especially in large digesters (Lyhne, P, personnel communication, 1999). Mechanical mixers are the most efficient type of mixer when new (Brade & Noone, 1981), but rapidly become choked with rags in sewage sludge and are eventually damaged by corrosion (Hobson & Wheatley, 1992).

GAS STORAGE AND TREATMENT All of the gas holders are bell-over-water type and are constructed of the same material as the accompanying digester. Only one of the plants (Clonmel) treats the biogas (as described in Chapter 3, Section 3.5) by removing the hydrogen sulphide, before entering the boilers and gas engines. Hydrogen sulphide can cause problems of wear in engines and valves. There are various economic arguments for treating the biogas to get longer engine life, however in practice this is feasible and economical only on a large scale (Hobson & Wheatley, 1992). All the digesters have condensate pots or splash traps to remove different vapours and dirt trapped in the biogas.

ENERGY GENERATION Buncrana is the only digestion plant that has no combined heat and power (CHP) unit installed on site. Tullamore and Clonmel have two gas engines coupled to a generator (1duty/standby) and when possible the biogas produced goes straight to the gas engine in which the primary task is to heat water for the heat exchangers with excess hot water used to heat the buildings. Then the biogas is used to generate electricity which is generally used for lighting etc. in the administration buildings. Flare stacks are on site in both plants, however they are rarely used.

The CHP system in Tullamore and Clonmel works well because both plants are producing sufficient quantities of biogas because they are operated near their design capacities, in

particular Tullamore. In Greystones and Tralee similar CHP units are installed, however to date they have not been commissioned. This would relate to the fact that the plants are not generating enough biogas to supply a constant flow to these engines. One reason for this is that the plants are not running at design capacity, both plants are not treating enough sludge to generate sufficient quantities of biogas. The biogas produced is used in the methane boilers with excess being burned through the flare stacks.

4.3 THE FULL SCALE OPERATION OF THE TULLAMORE ANAEROBIC DIGESTER

Process performance over a four-month period is summarised in Table 4.2. The pH entering the digester was on average just less than 6.6. However, alkalinity remained high in the influent with greater alkalinity released inside the digester therefore there was no need for buffer addition. Effluent pH continued to remain above 7.

TABLE 4.2 TULLAMORE ANAEROBIC DIGESTER SLUDGE ANALYSIS (AVERAGE) 1997-1998

Period	VFA mg/l	Alkalinity mg/l $CaCO_3$	pH influent	pH effluent	Solids in % DS	Solids out % DS
Jan '98-Apr'98 (Range)	36 (30-41)	5600 (5000-7000)	6.38 (5.7 - 6.79)	7.1 (7.01 - 7.16)	4.5 (3.9 - 5.9)	2.4 (2.2 - 2.9)

These results accurately reflect normal operation at Tullamore. The high alkalinity and gas yield (400 - 600m^3/d) suggest excellent process performance. This gives an average gas yield of 29m^3/m^3 inflow, which compares favourably with findings of Li *et al.*, (1996) operating an egg-shape digester yielding 21m^3 gas @ 61% CH4 per m^3 sludge feed. In fact the actual gas yield (400 - 600m^3/d) is consistently above the design figure of 408 m^3/d.

The survey showed that the success of the plant was due to a combination of factors including:

- Successful operation of the plant for a period of 12 years. Therefore a good database of information on different flow types and troubleshooting was available,

- One member of the supervising team has been on site since commissioning in 1986. Thus personnel are available with specialised experience for plant operation and maintenance,

- Emphasis was placed on the success of the internal heat exchangers employed in the digester to heat it to the design temperature of 36°C. These heat exchangers are open-ended cylindrical water jackets which stand on 3 legs on to the floor of the digester. A series of pipes from the boilers convey hot water to these 8 cylindrical water jackets which heat the sludge as it is mixed around in the digester. As the heat exchangers are inside the digester, there is no need for sludge recirculation therefore the energy cost and pumping costs are automatically kept to a minimum. It is interesting to note that none of the other designs use internal heat exchangers, but rather the more traditional external heat exchangers. Indeed difficulties have been reported where clogging and insufficient heat exchange has occurred in external exchangers.

- Waste activated sludge (WAS) is mechanically pre-thickened before feeding to the digester. This improves gas yields and is used in 72 out of 164 AD plants surveyed in Germany. (Buer *et al.,* 1996). All of the Irish designs surveyed also use mechanical pre-thickened WAS before digester feeding.

4.4 COMPARISON OF TULLAMORE WITH THE OTHER DIGESTION PLANTS

The design and operation of the anaerobic digester at Tullamore is compared with the four more recent designs in Table 4.1. Buncrana process performance details are not compared with the others plants because it is currently decommissioned for remedial work. As seen from this Table 4.1, the more recent designs are all of larger capacity, except Buncrana. Clonmel and Tralee have two digestion tanks as opposed to one, which allows for greater flexibility in terms of plant maintenance and feeding. All designs are gas mixed continuously stirred tank reactors operated at mesophilic conditions.

In terms of m^3 gas production per m^3 digester volume and energy efficiency, the digester at Tullamore could be considered to be the most successful ($29m^3/m^3$). However, Greystones operating for over two years is treating comparable amounts of feed and producing approximately equal quantities of biogas ($22m^3/m^3$ feed). Some of this gas is used in the boiler, however to-date the CHP unit has not been commissioned. Greystones treats all of the sludge generated on site ($18m^3/d$), however, the digester is designed to treat greater quantities of sludge than is generated at the moment. Clonmel and Tralee have only been operating two years and are producing significant quantities of biogas and no significant difficulties have been reported for the start-up operation.

Comparison of actual gas production versus design gas yields show that Clonmel and Tralee are still in start-up mode. The CHP unit and one of the digestion tanks at Clonmel are operating at design capacity and the second tank is currently operating at half design load. Similar progress is reported for Tralee. A slow start-up time was reported for some of the digesters. Part of the explanation could be that it takes a long time for an active consortium of

different groups of bacteria to develop in the digestion tanks. Until a microbial equilibrium is established, reactor performance can be unstable where toxic effects due to excessive VFA production can lead to process souring. Many researchers have reported that seeding with active biomass from other digesters at start up can reduce the risk of this occurrence. This seeding operation was performed at Tralee and few start-up difficulties were reported at this site.

Greystones reported a foaming problem in the digester after a bulking problem was encountered in the activated sludge tank. The bulking/foaming problem was attributed to the presence of filamentous bacteria in the activated sludge. Chemical treatment of the activated sludge eliminated the bulking problem and additional chemical treatment was required to eliminate the foaming problem in the digester. Pagilla and co-workers (1997) investigated the cause and effects of foaming in anaerobic sludge digesters. They observed that presence of excessive levels of *Nocardia* filaments (> 10^6 number/g VSS) in the activated sludge caused greater foam problems in gas-mixed than in mechanically mixed digesters. However, these authors reported greater solids removal for the gas mixed designs. All digesters in Ireland are of the gas mixed type.

4.5 DIGESTER MONITORING AND SIGNIFICANCE OF PARAMETERS MEASURED

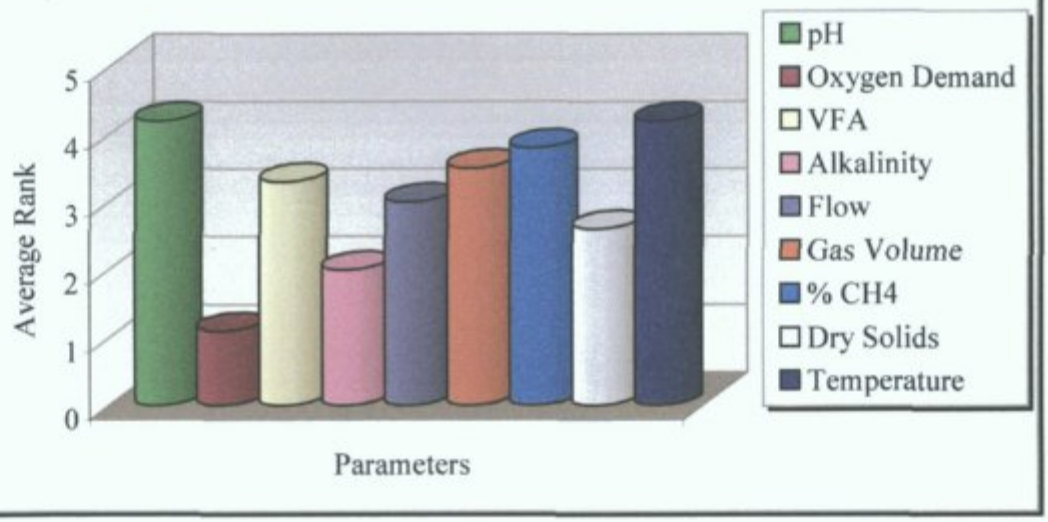
In order to monitor process performance and ensure stable operation of anaerobic digestion several measurement criteria have to be selected and checked on a regular basis. All plant operators agreed that pH was the most critical parameter to use as a yardstick for process stability (Figure 4.1). One operator noted that when a problem did occur due to insufficient mixing in the digester that significant pH changes were observed long before VFA

concentrations increased. The least successful of the digesters studied in terms of solids reduction and gas production showed pH levels consistently below 7. Figure 4.1 gives the average ranking of the parameters in terms of importance according to all plant operators working on existing plants in Ireland.

IWPC (1979) stated that if digestion is proceeding satisfactorily, pH control is unnecessary since the natural buffering capacity of digested sludge (based on bicarbonate and ammonia ions) usually maintains the pH close to the optimum of 7. The survey showed that pH measurements are taken every day at all plants usually from sample ports situated in the feed and effluent lines or from the recirculation line. All operators now try and maintain a pH inside the digester of greater than 7. Manual pH correction is usually not necessary except at start-up or after a shock feed to the digester.

Temperature is ranked at equal importance with pH in terms of process control (Figure 4.1). Temperature is also recorded every day. Some plants have on-line temperature probes in the digester to help control operation temperature and give a warning if there is a temperature drop due to problems with boilers, heat exchangers etc. Two of the digesters surveyed do not have on-line temperature probes. A decrease in temperature of the digesting sludge especially during winter months may be a consequence of inadequate heating capacity. Causes of persistent low temperatures are scaling or blocking of heat exchangers or increased requirement by sludge having a low solids content (IWPC, 1979). One operator observed a drop in temperature and traced the problem to a blockage in the heat exchangers. Another observed heat exchangers having difficulties providing sufficient temperature during the winter months. Inadequate tank insulation was also considered as a factor for lower tank temperatures in winter.

Figure 4.1 Averaged Ranked Importance of Parameters Measured



Biogas volume and percentage methane are the next most important parameters according to plant operators. Biogas volumes are usually measured or estimated on a daily basis with percentage CH_4 measured between 1 and 5 times a week.

There was a large variation in attitudes to the importance of Volatile Fatty Acids (VFA) measurement, alkalinity and flow measurement. There was also variation in the procedures used for the measurement of VFA and Alkalinity. Many plant operators have found VFA measurement to be time consuming and less informative than pH measurement. However, the IWPC (1979) stated that an increase in VFA concentrations of 100-200 mg/l was a sign of process stress.

Flow measurement is still recommended on a daily basis as it is necessary to check pump operation and estimate expected gas yields. Regular dry solids measurements (usually once or twice a week) are required for the same purpose, i.e. to see if the desired solids destruction of greater than 50% is achieved and to check the operation of the picket fence thickener.

Optimum digester feed DS concentrations appear to be around 5%, as lower concentrations result in less efficient solid removal in the digesters, and higher concentrations tend to cause mechanical problems with pumps, heat exchangers and mixing units. This leads to increased maintenance workload due to wear and cleaning of equipment (Buer *et al.,* 1996).

The use of toxicity and activity tests was almost totally absent from laboratory procedures for all of the plants surveyed. Only one operator used laboratory assays to check for toxicity. The test used was the measurement of ammonia concentrations. Indeed, ammonia is a well-known inhibitor of digestion, although inhibitory concentrations are a little indefinite and digester bacteria will adapt to otherwise toxic concentrations (Hobson, 1988). The utilisation of a "one-off" methanogenic activity test (as described by Colleran *et al.*, 1992), to determine toxicity thresholds for ammonia might save valuable time and give more useful information.

There were no checks for pathogens on-site in any of the plants. However, in future when thermophilic treatment is used for sludge sterilisation, checks for indicator microorganisms may be required.

Only two of the five plants surveyed measured organic content in any form. One measured total volatile solids and the other measured COD only. This is not surprising, as the main function of anaerobic digestion of sewage sludge is to reduce solids mass. The percentage of organic matter will be reflected in gas production.

No attempt at sludge characterisation was made at any of the plants. Sludge is a particularly difficult material to characterise in a quantitative manner that is both fundamentally based and useful on an engineering scale (Dentel, 1997).

CHAPTER 5

INVESTIGATION OF MIXING AND PROCESS MONITORING ON ANAEROBIC DIGESTER PERFORMANCE AT LABORATORY SCALE

5.1 INTRODUCTION

Having visited and investigated the five existing full-scale sewage treatment plants which have incorporated anaerobic digestion into their design, it was evident that there was some confusion and discrepancy in the operation and monitoring of the digestion process (Chapter 4). Temperature, pH, volatile fatty acids (VFA), biogas production and percentage methane (CH_4) composition of the biogas are all significant parameters in terms of process performance, stability and control (Switzenbaum *et al.*, 1990; Ahring & Angeledaki, 1997; Nordberg *et al.*, 1999). However, plant operators were not consistent as to which parameter(s) was the most important to ensure reliable process control. Many authors who have completed extensive research in this area also show conflicting results. Two examples of this are: Rozzi & Labellarte, 1984, Wheatley *et al.*, 1987 found that measurement of alkalinity in conjunction with pH to be a good process indicator and Ahring *et al.*, 1995 found that VFA and in particular individual VFA measurement are good parameters for predicting process instability.

Process instability and consequently process failure was the biggest fear for plant operators. Therefore, it is very important that the parameters indicating the first sign of process failure be obtained. To complete this Study in a full and comprehensive way it was necessary to try and alleviate some of these conflicting views by first hand experience in operating a pilot plant. In order to do this, pilot-scale anaerobic digestion plants, similar to those visited and studied, were set-up and operated in a laboratory at IT, Sligo. The main objective of this Study was to confirm which parameters indicates process distress first and to observe what happens when using different feeding and mixing systems.

5.2 DETAILS OF THE PILOT STUDY

There were three pilot anaerobic digesters commissioned called reactor 1, 2 and 3. Reactor 1 and 2 were recirculated, unmixed reactors and are identical in all aspects including size and feeding characteristics. Reactor 3 was recirculated and mixed with a mechanical stirring device as described in Figure 2.3, Chapter 2. Also operated along side the digestion plant was an Activated Sludge plant. This plant was to provide the secondary sludge solids feed for the digesters in the first stage of the study. The Study was carried out over 144 days.

Accurate monitoring of process parameters is essential for effective control and efficient operation of anaerobic digestion plants. Table 5.1 shows the parameters tested on the influent and effluent and frequency of measurement, during this study.

TABLE 5.1 PARAMETERS MEASURED DURING MONITORING OF PILOT DIGESTERS

Parameter	Influent	Effluent	Frequency
COD	Yes	Yes	Every day
Solids TS, VS & FS	Yes	Yes	Every day
pH	Yes	Yes	Every day
VFA	No	Yes	Every day
Alkalinity	No	Yes	Every day

Biogas volume and biogas composition (CH_4, CO_2 and O_2) were also measured every day.

A combination or all of the above parameters were used to determine the effect the following had on each of the reactors:

- organic loading rate,
- organic composition,
- stirring.

5.3 COMMISSIONING AND SEEDING OF THE DIGESTERS

All three pilot scale digesters were seeded the same and commissioning began on day 1. The seed for the digesters was obtained from anaerobic digesters at 3 sewage treatment plants around the country (Tralee, Greystones and Clonmel). Two sludge samples came from Clonmel because there are two digesters commissioned at this plant. Before sludge was mixed and put into the individual reactors the solid content of each sample was calculated (Table 5.2).

TABLE 5.2 SOLIDS CONTENT OF SLUDGE SAMPLES FROM TREATMENT PLANTS

Plant	Average Volatile Solids g/l	Average Fixed Solids g/l	Average Total Solids g/l
Clonmel A	15.7	8.65	24.381
Clonmel B	20.9	15.187	35.513
Tralee	28.015	21.035	49.05
Greystones	14.755	10.780	25.535

The four sludge samples (seen above) were mixed together to form the seed sludge for the pilot digesters. Each reactor received 3.5 litres of sludge with a total solids content of 121.9g/reactor. The VS content measured was 71.8 g/reactor. The reactors were then topped

up to 4.5 litres with water. Sodium bicarbonate buffer solution (Na_2CO_3) was also added in varying quantities to maintain a neutral pH.

Once the 3 reactors were filled and buffered the reactors were sealed. The heating jackets were then switched on and the contents of the digesters were heated to 35°C, which is generally considered the optimum mesophilic temperature (Hammer *et al.*, 1996).

The three reactors were then left to acclimatise to their new conditions over the weekend. The stirring device in reactor 3 was commissioned on day 4 and was set to operate for 15 minutes every 4 hours.

On day 7 the heating jacket of reactor 3 failed and the reactor contents fell below the optimum temperature required for stable operation. The jacket was replaced immediately with a spare jacket on site and the reactor was heated again.

Recirculation was introduced to reactor 1 and 2 on day 7. Recirculating lines and pumps were set up and started operating at a rate of 15 minutes every hour. Also, on day 7 monitoring of gas composition commenced.

This starting-up period (days 1-12) provided an opportunity to become familiar with the operation and general maintenance of the anaerobic digestion equipment. Problems were solved both with the biogas and heating systems and awareness was raised as to the possible problems that may be encountered with the equipment over the rest of the pilot study.

On day 13 feeding of the reactors commenced.

5.4 FEED COMPOSITION ENTERING THE DIGESTERS

All 3 digesters were subjected to the same feed on the same days. The 144-day period was divided into separate periods depending on the feed composition as shown in Table 5.3.

TABLE 5.3 FEED COMPOSITION

Period	Days	Feed Composition	Stage of operation
		None	Set-up/start-up
I	1-12		commissioning
II	13-34	Secondary Sludge & Whey	Full operation
III	35-61	Mainly Whey	Full operation
IV	62-76	None	No feed (reactors shut down)
V	77-94	Primary Sludge, Secondary Sludge & Whey	Full operation
VI	95-108	Animal Slurry and Whey	Full operation
VII	109-140	Animal Slurry	Full operation
VIII	140-144	None	Down time decommissioning

The feed was manually batch fed into the digesters on a daily basis using peristaltic pumps, pumping approximately 333ml (+/- 80 mls) over approximately 3 minutes (+/- 1 minute). In reactor 1 and 2 the feed entered the digester through the top and down the middle (Section 2.5.2 III, Chapter 2). In reactor 3, because the stirring device was placed down through the middle of the digester, the feed had to enter the digester under the stirrer at the bottom (Section 2.5.2 III, Chapter 2).

Throughout this study different feed compositions were used. There were three main feeds used: sewage solids, whey and animal slurry. The sewage solids were either generated on site

from the laboratory pilot scale activated sludge plant (secondary sludge), or obtained from sewage treatment plants (primary & secondary sludge) around the Sligo area. This pilot scale activated sludge plant was commissioned 36 days before the commissioning of the pilot anaerobic digesters, in order to generate enough solids for feed to the digesters.

In order to produce a comprehensive pilot study, it was necessary to model the pilot plant as closely as possible to a full-scale sewage treatment plant. Four out of the five anaerobic digestion plants studied in this project were fed using both primary and secondary sludge (settled waste activated sludge). The pilot activated sludge plant was operating to simulate this situation. The plant was seeded with activated sludge (Chapter 2) from sewage treatment plants around the Sligo area and was continuously fed with whey as substrate.

5.4.1 Calculating the Initial Feed Quantity for the Digesters

The first feed was based on the volatile solids (VS) content of the seed sludge in each reactor, and the food to microorganism (f/m) ratio of 0.05 - 0.15. The VS content of the seed sludge was calculated to be 71.8 g/reactor. For the 0.05 f/m ratio the organic load required was calculated to be 10.77 gCOD/d and for the 0.15 f/m ratio the organic load was calculated to be 3.6 g/COD/d. The average of these two figures was calculated and the first feed was based on this figure (3.6 gCOD/d).

5.4.2 FEEDING

On days 13-61, the digesters were only fed once every two days i.e. Mondays, Wednesdays and Fridays. On Fridays and Mondays greater volumes were fed to the digesters (400mls) to make up for the absence of feed during the weekend.

On day 13 the anaerobic digesters were fed for the first time with secondary sludge, from the activated sludge (AS) plant, and whey. However after three weeks the AS plant was not able to generate sludge fast enough to continue feeding the 3 anaerobic digesters. Therefore days 35-47 consisted of feeding the reactors with just whey. The AS plant was given a few days to recover and to start generating significant amount of sludge again. Then on day 48 and 50 the digesters were fed with secondary sludge from the pilot AS plant again. However this was short-lived as the AS plant could not generate enough sludge to feed the digesters, so it was decided to decommission the AS plant and find an alternative feed. For the remainder of this period days (51- 61), the reactors were feed with whey.

From days 81 - 140 feeding took place once a day, with greater volumes on Mondays and Fridays to make up for no feed at the weekend. Initially after the non-feeding period (days 77-81) the reactors were allowed to heat up. On day 81, whey was added for two days and then primary sludge mixed with whey was introduced for the following 3 days. This primary sludge was obtained from Strandhill treatment plant, Co. Sligo. The use of primary sludge was discontinued after 3 days due to the difficulties with handling, and the risk of pathogenic infection. On days 88 to 94 secondary sludge was obtained from treatment plants throughout the Sligo region. Again this sludge was mixed with whey and administered to the digesters.

On days 95- 108 animal slurry was introduced as feed to the digesters. First, the animal slurry was mixed with whey and then on days 109 to 140 the slurry was added on its own. The animal slurry was obtained from the tanks of a slatted house, which stored dairy cows for the winter months. This was thought to be a good alternative feed for the digesters as it has a good organic content.

The reactors were fed for the last time on Day 140.

Due to the different feeds being introduced to the reactors, the process had to be monitored very closely to observe the effects each feed type had on the process. The microorganisms would have to acclimatise to the new substrate each time. The parameters used to monitor the process, would pick up any changes that could occur.

For ease of explanation, from day 13 onwards the operation, maintenance and monitoring of reactors 1, 2 and 3 is discussed separately in the following sections. The outcomes and the results of each digester are then compared in a later section.

5.5 Operation of the first Unmixed, Recirculated Digester (Reactor 1)

5.5.1 Period I, Start-up, days 1-12

The reactor was seeded and commissioned as in section 5.3. A summarised version of the results taken for Reactor 1 is contained in Table 5.5.1. The biogas volume was recorded from day 5. Figure 5.5.1 shows the biogas production for the first stage of the study starting on day 5. The average produced for this period was 0.17 l/l reactor/d (l/l/d) (Table 5.5.1). The methane composition of the biogas was recorded from day 6 and the mean for this period was 61% (Table 5.5.1) varying from 37-71%. On day 12 the percentage CH_4 acutely dropped to 37%, however by the next morning it had recovered to 59%. Figure 5.5.2 shows the CH_4 production for the first stage of the study starting on day 6.

On day 12 the recirculation pump was turned on and began recirculating the sludge at a rate of 1.8 l/hr, which was equal to 9.2 times the reactor volume in 24 hours. The sudden drop in methane on day 12 may be related to the recirculation being turned on, that same day.

Figure 5.5.1 R1 Biogas Production

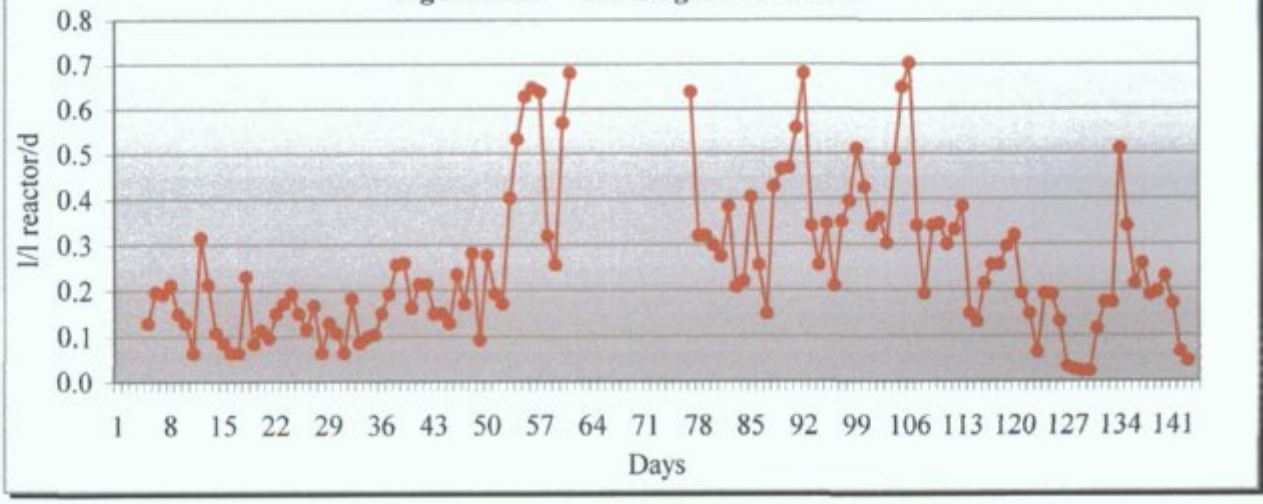
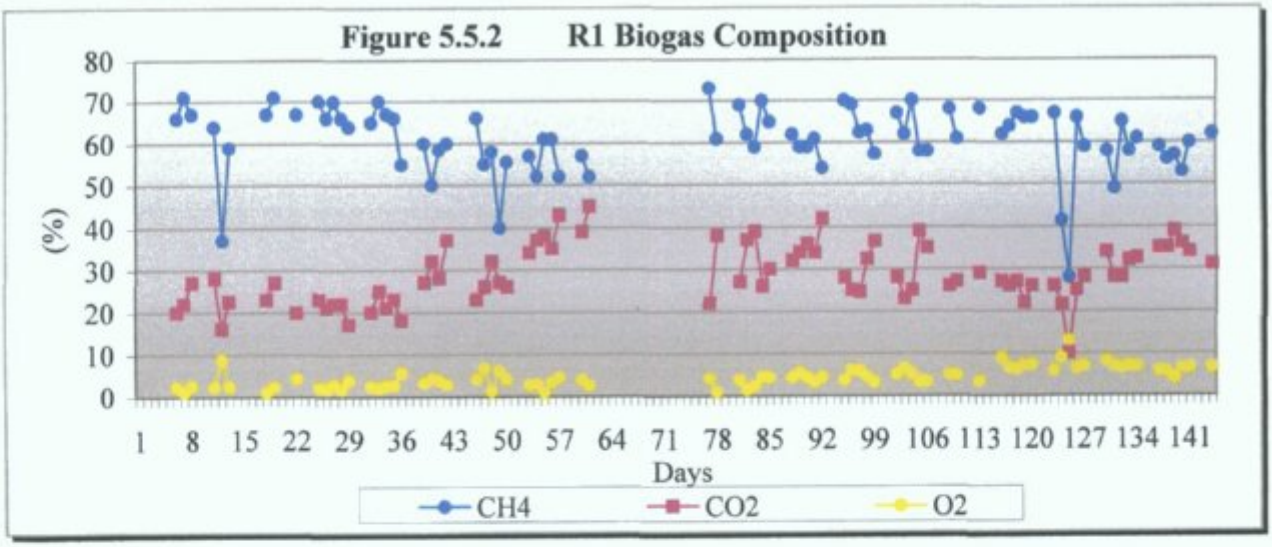


TABLE 5.5.1 REACTOR 1 RESULTS SUMMARY TABLE

Period (Days)	Organic Loading Rate (OLR) gCOD/d		COD (effluent) g		Retention Time (RT) days	Volumetric Loading Rate (VLR) gCOD/l/d		Biogas Volume l/l/d		Methane Composition %		PH (effluent)		VFA (effluent) mg/l		Total Solids Removal %		COD Removal %	
	M	*S.D.*	*M*	*S.D.*	*M*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*
1-12								0.17	0.07	61	13.6								
13-34	3.8	0.57	0.35	0.11	15.5	0.81	0.12	0.12	0.05	66.8	3.3	7.54	0.15			89.4	8.2	90.7	3.15
35-62	5.13	1.5	1.43	1.41	56.2	1.09	0.32	0.3	0.19	56.4	6.1	7.47	0.13	155.5	57.8	65.5	49.2	72	34.6
63-75																			
77-94	5.4	1.83	1.82	1.53	26	1.15	0.39	0.37	0.15	62.8	5.45	7.37	0.16	151	149	41.1	31.8	66.3	47.9
95-108	14.9	4.06	3.44	1.05	24.9	3.19	0.86	0.4	0.15	63.6	5.0	7.52	0.12	1082	620	67.5	4.49	77	4
109-140	18.2	4.75	7.70	3.18	19.9	3.89	1.01	0.21	0.12	59.1	9.3	7.49	0.16	3074	1387	54.2	9.45	57.7	8.37
140-144								0.32	0.23										

M = Mean S.D. = Standard Deviation

Figure 5.5.2 R1 Biogas Composition



5.5.2 PERIOD II, DAYS 13-34

On day 13, reactor 1 was fed for the first time with secondary sludge and whey, at an average organic loading rate of 3.8 gCOD/d with a mean retention time of 15.4 days (Table 5.5.1). Figure 5.5.3 show the organic loading rate and the feed composition on the days the reactor was fed. This is basically the same for all three reactors. The mean volumetric loading rate was 0.81 gCOD/l/d, based on a reactor volume of 4.7 litres.

The mean biogas volume produced during this period was slightly below the former period at 0.12 l/l/d (Table 5.5.1). This drop in production can be attributed to the introduction of feed during this period. The microorganisms were acclimatising to the substrate. The mean methane content for this period was 66% (Figure 5.5.2).

The mean percentage COD removal was calculated to be 90.7%. From this percentage it was calculated that approximately 1.2 l CH_4/d should have been produced. However only 0.12 l/l/d biogas was produced, which was equal to 0.564 l/d. Of this biogas 66.8% was methane, which was equal to 0.376 l CH_4/d production. These contradicting figures would suggest that either, only 28.2% COD was converted to gas with the remainder retained as solids in the digester or else a substantial quantity of biogas escaped from the gas collection system. The following is a sample calculation of how the above figures were arrived at.

Example calculation for R1, days 13-34:

From Table 5.5.1, the average COD in the influent and effluent was 3.8g and 0.35g respectively,

therefore, COD removed by reactor was 3.8g - 0.35g = 3.45g

therefore, % removed was 3.45/0.038 = 90.7%

From Parkins & Owen (1986), 1 g COD removed = ~ 0.35 l CH_4

Therefore 3.45g*0.35 l CH_4 = 1.2 l CH_4 should have been produced

From Table 5.5.1, the average biogas production was 0.12 l/l/d, which was equal to 0.564 l/d, (based on a reactor volume of 4.7 l). Of this 0.564 l/d produced, 66.8% was CH_4.

This was equivalent to only 0.376 l CH_4 indicating that only 28.2% COD was removed by digestion [(0.376/0.35)/0.038)].

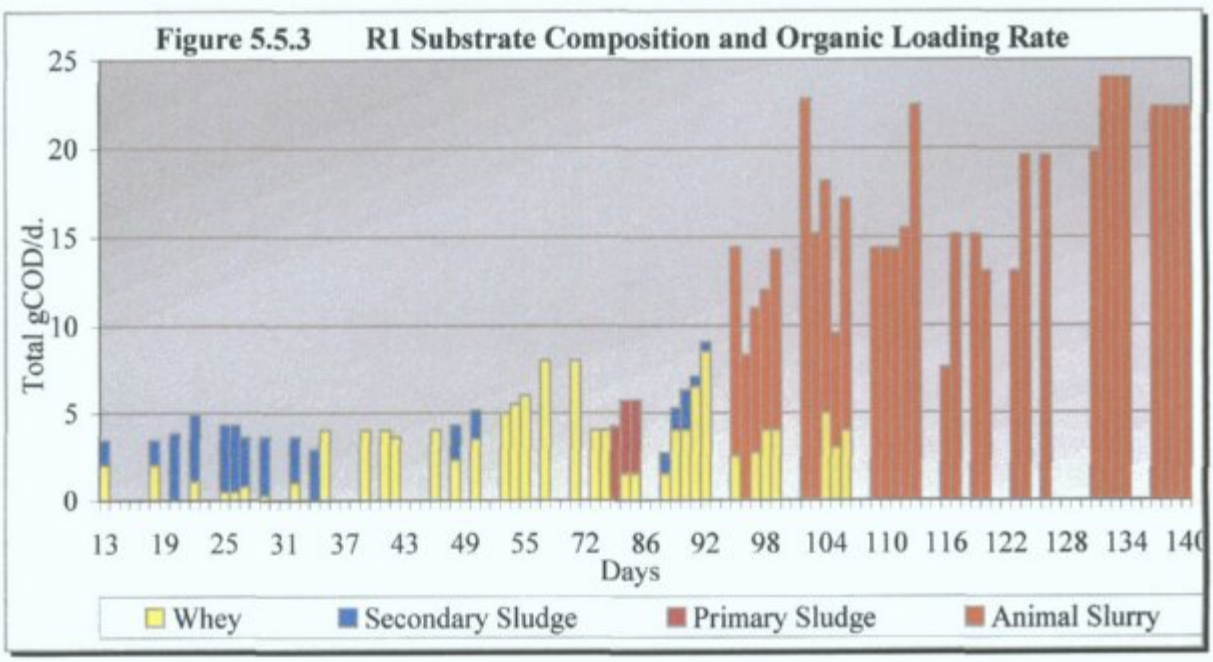
The difference between observed and theoretical COD conversion to biogas may be due to:

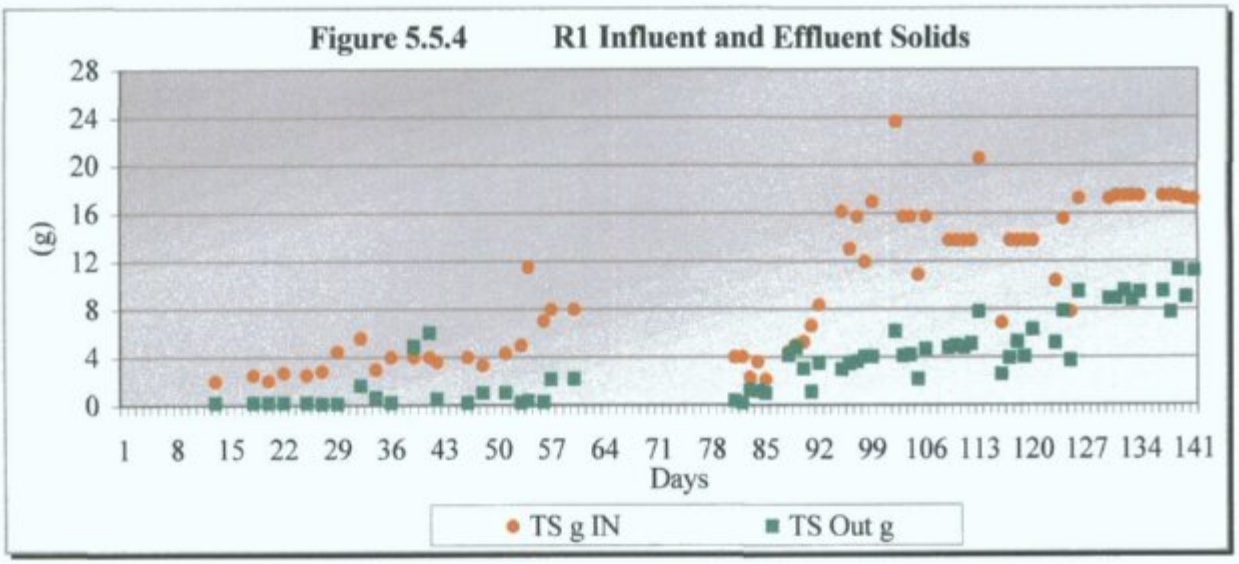
1. Biogas leaking from the gas measurement device,
2. Solids retained by settlement in reactor but not digested,
3. A percentage of the organic matter entering the digester may have been converted to new bacterial cells (microbial growth).

As discussed in a later section the second possibility did occur.

The average solids removal for this period was 89.3%. This would not be a representative figure because it is highly unlikely that a digester would remove that many solids by digestion alone. The reason was obviously due to settlement in the digester due to lack of mixing with the supernatant leaving in the effluent stream. The recirculation rate was doubled in an effort to reduce settlement. This did not solve the problem immediately, however it was realised that it would take some time for the reactor solids to thicken up and this would depend on the percentage solids entering the digester at any one time and the retention time. Figure 5.5.4 shows the total solids in grams entering the reactor in the influent and exiting the reactor in the effluent.

The monitoring of pH started on day 20. The average pH for this period was 7.5, which was within the optimum range for good process performance.





5.5.3 PERIOD III, DAYS 35-62

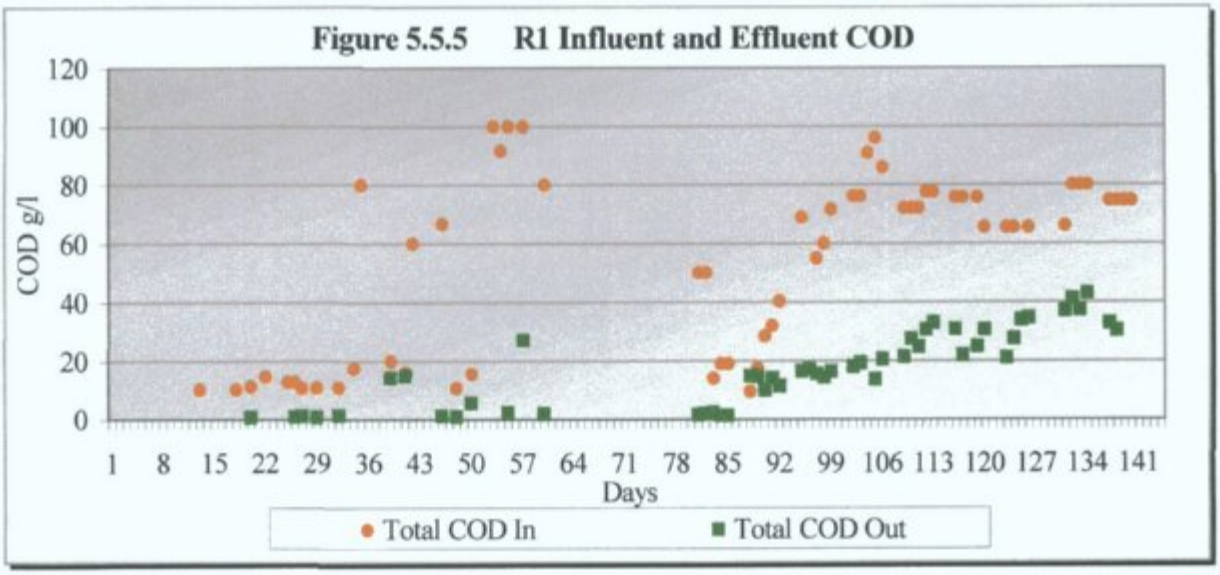
During this period, the digester was mainly fed with whey, apart from two days (day 48 & 50) when secondary sludge was mixed with the whey (Figure 5.5.3). The average organic loading rate was of 5.13 gCOD/d (Figure 5.5.3), and the mean retention time was 56.2 days. Whey was being added with water and in order to avoid washout of solids and microorganisms with large quantities of water entering the digester, it was decided to cut the volume entering the digester.

The biogas production increased to an average 0.3 l/l/d. This increase may be due to the fact that whey is more easily digested than the secondary sludge, as it consists of shorter chain molecules than those present in the sludge. On day 48 and 50, when the secondary sludge was re-introduced, biogas volume increased once again to 0.28 l/l/d. The next day and again on day 50 after the secondary sludge was added, the biogas production was up to 0.4 l/l/d. For the remainder of this period (days 53-61) when whey was fed to the digester on its own the biogas volume increased to it highest so far in the study (Figure 5.5.1).

The mean percentage methane in the biogas decreased, to a mean of 56.4 % (Table 5.5.1). It can be seen from Figure 5.5.2, that on day 36 immediately after the whey was added, the methane content dropped to 55% with an increase to 60% following the weekend, and again on day 40, it decreased by 10% to 50%. This would suggest that the methane-producing microorganisms reacted adversely to the organic content of the whey at the beginning (maybe perhaps due to a deficiency in nutrients that were present in the secondary sludge). The consortium of microorganisms that was building up would have been based on the contents of the secondary sludge. However, on day 41, the percentage methane content began to rise and continued this trend until day 46 where it reached 66%. This would suggest that the microbial consortium were changing and beginning to adjust (acclimatising) to the new substrate, whey.

On day 49 when the secondary sludge was re-introduced to the reactor, the percentage methane dropped. This can be attributed to acclimatisation of the methanogens to the new substrate and organic loading. The following day the methane content rose to 55.5%. When whey was fed on its own for the remainder of this period the percentage methane showed an average decrease, varying between 52 and 61%. This decline may have been due to a lack of nutrients in the whey and to detect this on day 56, 0.2 mls of NH_3 was added with the whey to see if this would enhance the performance of the digester. This did not have any significant effect, positive or negative, on digester performance.

The COD removal during this period was 72% (Table 5.5.1). This would be equal to 1.3 l CH_4/d, however only 0.8 l CH_4/d was produced (see Sample Calculation, Section 5.5.2). Again indicating that only 44.3% COD was removed by digestion or that biogas was escaping from the system. Figure 5.5.5 shows the COD in the influent and effluent throughout the duration of the study.



The mean solids removal was 65.5%. The digester contents were seen to be thickening and actually for two days the solids coming out of the digester were greater than what was going in, resulting in negative figures for solids removal. This was because the solids entering the digester were of a low solids concentration compared to what was in the effluent (Figure 5.5.4).

The mean pH at this time was 7.47 and the VFA was measured for the first time and found to be quite low at a mean of 155.5 mg/l (Table 5.5.1).

5.5.4 PERIOD IV, DAYS 63-75

Feeding was discontinued on day 62 and the reactor was shutdown for 15 days. From a previous pilot scale study, this shutting-down of the process over a period of time did not pose a problem during start-up again. The process was still viable (Shannon, 2000).

5.5.5 PERIOD V, DAYS 77-94

In stage 2 of the study feeding re-commenced on day 81. During this stage the reactor was fed every day once a day, except for weekends. The initial feed for the first two days (days 81 & 82) was whey and then primary sludge was mixed with whey and was fed to reactor 1 for days 83, 84 and 85. Following the weekend, the substrate was changed to secondary sludge mixed with whey for one week (days 88 to 92) (Figure 5.5.3). The mean organic loading rate for this period was 5.4 gCOD/l/d and the retention time was 19.2 days. This was equal to a mean volumetric loading rate of 1.15 gCOD/l/d, based on a reactor volume of 4.7 litres.

The biogas volume decreased after the whey was added but recovered quickly. Again on day 84 there was a significant decrease in biogas production. This may be attributed to the introduction of primary sludge at a higher loading rate than the previous days. The microorganisms would have to adjust to the new substrate and conditions. The following two days showed a steady increase in the biogas volume. With the introduction of secondary sludge and whey on day 88, biogas production increased and continued this trend for the remaining of the period (Figure 5.5.1). The mean biogas production was 0.32 l/l/d (Table 5.5.1).

The mean methane production was 62% (Table 5.5.1). Thus, the 0.32 l/l/d biogas produced corresponds to 0.2 l/l/d as methane gas. On days 81-82 the percentage CH_4 dropped from 69 to 59%, as the microbial consortium would be acclimatising to the new feed. On day 83, primary sludge was added without whey and there was a significant increase in the methane production equal to 70%, suggesting the methanogens thrived on the contents of the primary sludge. The following day the primary sludge was added with whey and this caused a decrease in methane to 65%. This would suggest that whey was slightly inhibiting the methane-producing

microorganisms. On day 62, after secondary sludge and whey was added the CH_4 decreased to 59%. The following day it stabilised at 59% and increased to 61% (Fig. 5.5.2) on day 91. This increase may have been caused by the increase in secondary sludge content entering the digester. On day 91, the whey content was increased and the secondary sludge content decreased. Within one day percentage CH_4 decreased by 7% to 54%, again suggesting slight inhibition of the whey to the methanogenesis.

The percentage COD removal during this period was 66.3%. This would suggest that 1.25 l CH_4/d should have been produced. However only 1.1 l CH_4/d was produced. This would indicate that 57.7 % COD was removed by digestion or else some biogas was escaping out of the gas collection system (see Sample Calculation, Section 5.5.2). Figure 5.5.5 shows the COD entering and exiting the digester in g/l. The percentage solids removal was 55.3 %. Solids entrapment was obviously still occurring inside the digester. Figure 5.5.4 shows the solids entering and exiting the digester.

The value for average pH was 7.36. On day 92 the pH dropped to 7, this also coincided with the drop in methane to 54% on the same day, suggesting that something may have upset the digester contents. The VFA on that day was 135 mg/l, which is not considered as a high value. If the VFA were high it would have been thought that the process was at the beginning stages of souring. The mean VFA for this period was 151 mg/l.

5.5.6 PERIOD VI, DAYS 95-108

On day 95 animal slurry was introduced as feed to reactor 1. It was mixed with whey and was fed at a mean organic loading rate of 14.9 gCOD/d and a retention time of 24.9 days. The mean

volumetric loading rate was 3.19 gCOD/l/d. This was a significant increase in loading compared with the pervious periods (Figure 5.5.3).

The initial volumetric loading rate was approximately 3.07 gCOD/l/d, after this was added, by the next day the biogas volume decreased. The loading rate was lowered to 2.34 gCOD/l/d and the biogas volume still dropped. However, this may be due to the microbial consortium acclimatising to the new conditions. An increase in biogas production was observed on days 98 & 99. The following week the loading rate was increased and the biogas volume dropped but increased dramatically to 0.53 l/l/d on day 104. The loading rate was decreased slightly on day 104 and this caused a slight decrease by the next day however by day 106, 0.65 l/l/d of biogas was produced (Figure 5.5.1), this was the most biogas produced in any one day by reactor 1, since start-up. The average percentage methane production for the period was 63.6% (Figure 5.5.2).

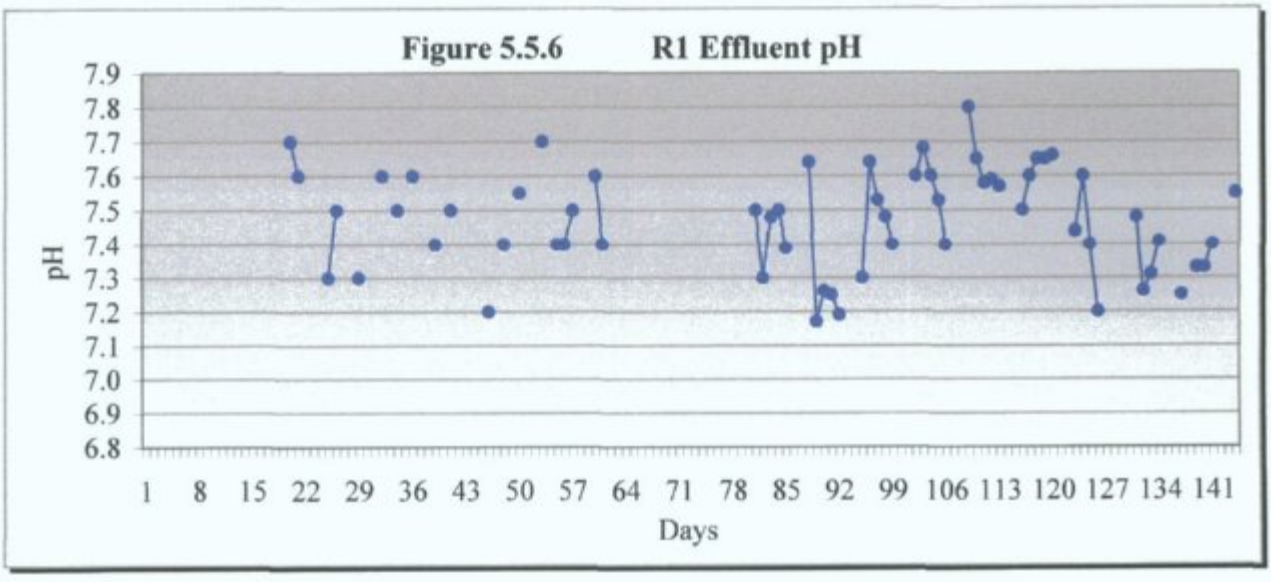
The COD removal rate for this period was 77%. This would be equivalent to 4 l CH_4/d but it can be seen from the actual volume of biogas produced and the methane composition of that biogas that only 1.2 l CH_4/d was produced. This would suggest that only 23% of COD was actually removed by digestion or else some biogas was escaping from the gas system.

On day 103, the recirculation pump became blocked overnight leading to a burst recirculation pipe. The pump then sucked in air and pumped out sludge until the sludge level had dropped below the outlet level. It was necessary to add old sludge (approx. 500mls) to make up the digester contents to the correct level. This problem did not seem to cause the digester performance any undue distress. Within this period a scum layer was observed to be forming on the top near the outlet of the reactor.

The solids removal was not representative again here. This was due to the large increase in the percentage solids entering the digester (Figure 5.5.4). It would take a while for the solids in the reactor to thicken up as this reactor was not mechanically mixed.

The pH remained stable (Figure 5.5.6), however the mean VFA increased rapidly during this period. This would be a concern as a build up of VFA would lead to souring of the digester contents, eventually leading to reactor failure. On day 34, VFA levels rose by 400 mg/l and the next day they increased by ~1000 mg/l before decreasing by 900 mg/l on the following day (Figure 5.5.7). However, this build up did not seem to have any effect on the pH so the digester contents was sufficient to be self-buffering. The definite increase in the VFA concentrations suggests digester distress. The fact that levels began to fall after a few days suggests acclimatisation of bacterial population to the newer more complex substances present in the slurry feed.

Figure 5.5.6 R1 Effluent pH



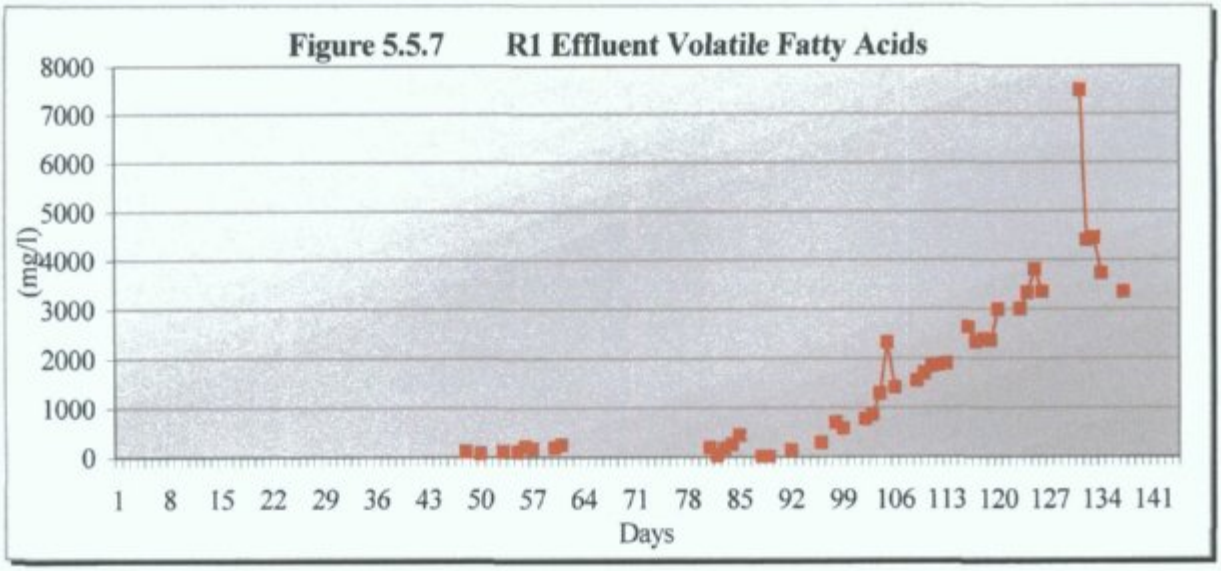


Figure 5.5.7 R1 Effluent Volatile Fatty Acids

5.5.7 PERIOD VII, DAYS 109-139

On day 109, the animal slurry was fed to the reactor by itself at a mean organic loading rate and retention time of 18.2 gCOD/d and 19.9 days, respectively. For the first two weeks (days 109-122) of the digester running on animal slurry there was no obvious problems. The mean volumetric loading rate (days 109-122) was 3.12 gCOD/l reactor/d and the retention time was 23.7 days. During the first week the biogas production showed a decrease each day, however the following week it begin to show an increase each day, therefore the microorganisms were adjusting to the new substrate. The average biogas production for these two weeks was 0.27 l/l reactor/d. The percentage methane remained at an average of 65.6%.

The COD removal for these two weeks was 57.7% (Table 5.5.1). This would suggest that there should have been 3.6 l CH_4/d produced however there was only 0.58 l CH_4/d produced. This suggests that only 9.2 % COD removal (see Sample Calculation, Section 5.5.2).

The average solids removal for these two weeks was 64%. This could be considered as high. However, it would relate to the high solids content entering the digester as it would take a while for the new removal rate to settle out (~ 1 retention time).

Unfortunately pipe blockage started to become a regular problem since the addition of the animal slurry with its grassy composition. By this time a thick scum layer was forming at the top of the digester, which was also attributed to the composition of the animal slurry. The week after these blockages were first observed, the digester exploded under extreme pressure on day 124.

5.5.7.1 CAUSE OF REACTOR EXPLOSION

The gas pipe and the inlet pipe had both blocked, therefore as the gas was being produced there was no way for it to escape, so eventually, it burst out the inlet pipe causing an explosion of sludge and gas from the top of the reactor. This action also sucked in a lot of air into the reactor and was stirred into the remaining reactor contents. The main cause of the pipes blocking was the formation of a scum layer. This layer filled the space between the digester contents and the lid thereby blocking the entrance to the gas storage tank. The inlet pipe was blocked due to grassy material building up inside the pipe, until it eventually clogged. Both of these pipes blocking was caused by the build up of the scum layer, which was directly linked to the composition of the animal slurry. Plate 5.1 and 5.2 shows the scum in the digester.

The reactor was topped up with 250 mls of animal slurry and some old sludge (approx. 700mls). This action would have caused the air to be agitated throughout the whole contents of the digester. The reactor was cleaned up as quickly as possible and the parameters were monitored.

PLATE 5.1 SCUM LAYER FORMATION IN THE TOP OF THE DIGESTER

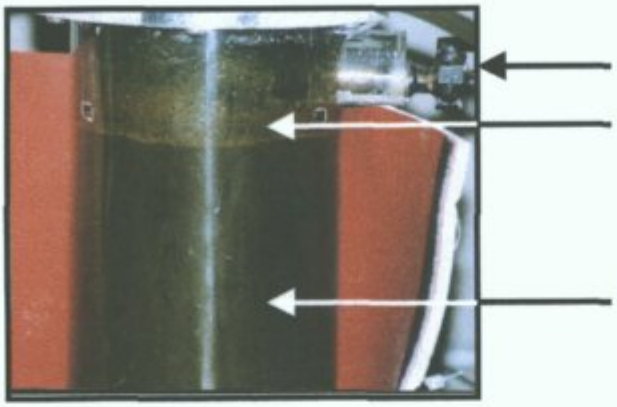
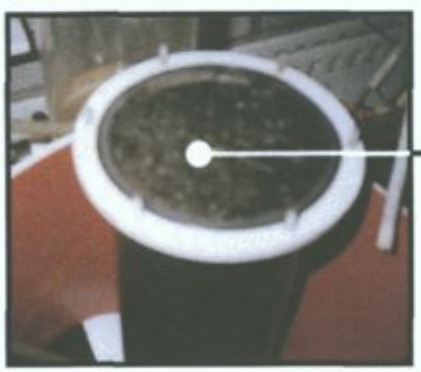


PLATE 5.2 SCUM LAYER BEFORE AND AFTER REMOVAL



Top of digester after the scum layer was removed

5.5.7.2 DIGESTER PERFORMANCE AFTER EXPLOSION

The shock caused the gas production to drop to 0.076 l/l/d that day. However, it increased to 0.19 l/l/d (Figure 5.5.1) the following day (day 125). For the remainder of the week it decreased and after the weekend it only had produced 0.08 l/l/d, which was an average of 0.04 l//d over the weekend period. This low biogas production could also be attributed to the fact that reactor 1 was not fed for the remainder of the week after the explosion, giving it times to recover. It was also possible that this digester had completely failed at this stage. That day the percentage methane decreased to 41.5% (Figure 5.5.2) and on day 125 it decreased even further to 28%. This would suggest that the methanogens bacteria were inhibited or maybe even killed off by the air intrusion, as they are very sensitive to O_2. By day 126, the percentage methane had recovered to 66% even though it was producing very little gas. The VFA concentration was measured the day of the explosion and they were 3350 mg/l (Figure 5.5.7). They were not measured again until day 131 (because the digester was not fed there was no effluent sludge) and they showed a significant increase to 7485 mg/l. This increase in VFA was also reflected in the percentage methane for that day which dropped to 49%. However by the next day the acids had subsided to 4405 mg/l. Also the percentage CH_4 increased to 65% corresponding to the drop in VFA. On day 126, the pH dropped slightly to 7.2 (Figure 5.5.6). It would correspond to the increase in VFA and the decrease in biogas volume and percentage methane for that day. The alkalinity on that day was high at 7900 mg/l which was significant to keep the pH above 7.

The digester was opened on day 130 and 500mls of scum was removed with as little disturbance as possible to the reactor contents (Plate 5.2). Again the digester was topped up with old anaerobic sludge (the effluent from the digester was in a storage container). This same day feeding commenced again. For the remainder of the study the mean volumetric loading rate was 4.81 gCOD/l/d and the retention time was 15.6 days. The biogas production started to increase

gradually day by day, however the percentage methane began to decrease day by day (Figure 5.5.1 & 5.5.2).

5.5.8 DECOMMISSIONING OF REACTOR 1

Feeding was discontinued to R1 on day 141. The recirculation and heating was switched off. The heating jacket was removed from around the outside of the reactor. It was now possible to see into the digester contents and the suspected layering was obvious. Again between day 128 and day 140 the scum layer had formed to a depth of 300mm on the top. From the bottom to about half way up the digester it was apparent that the solids were thicker. Therefore confirming that the settlement had taken place. The reactor was allowed to cool down for the weekend before further tests were taken.

On day 144, the digester was opened and the 250 mls of a scum layer was again scooped out. The solids content of this was measured. Then the remaining contents in the reactor were stirred well and a sample was taken so that the solids could be measured. The results of the final solids taken are given in Table 5.5.2.

TABLE 5.5.2 SOLIDS IN R1 AT THE START AND FINISH OF THE PILOT STUDY

Row	Sample	Total solids g
1	Day 144 sludge sample TS	285.14
2	Day 1 sludge sample (seed) TS	121.93
3	Cumulative TS entered R1	657.52
4	Cumulative TS exited R1	251.38

Row 1 represents the total solids contained in the digester at the end of the study including the solids in the 750mls of scum layer. Row 2 represents the total solids that were in the reactor on the first day of the Study (the seed sludge). Row 3 represents the total solids that entered the digester throughout the entire study and row 4 represents the total solids exited the reactor during the entire study. The total solids removed throughout the entire pilot trial in reactor 1 were 36.9% solids removal. The normal solids removal in a reactor should be between 30 to 50%, therefore reactor 1 removed almost 13% less than expected in terms of solids removal. This can be partially explained by the constant changes in feed composition resulting in insufficient time for microbial populations to develop and degrade the complex macromolecules in the solid fraction of the feed. Also the solids retained in the reactor were equally to 163.2g. A reason for this could be that the solids were settling to the bottom of the reactor due to a lack of mixing.

From the Table 5.5.3 it is obvious that there was quite a large percentage of organic matter (volatile solids) and solids in the scum layer.

TABLE 5.5.3 R1 COMPOSITION OF SLUDGE AT THE START AND FINISH OF THE PILOT STUDY INCLUDING THE SCUM LAYER

	Total Solids g (%)	**Volatile Solids g (%)**	**Fixed Solids g (%)**
Scum layer	60.3 (100)	47.5 (78.7)	12.66 (20.9)
Day 144 reactor contents	224.8 (100)	152.7 (67.9)	72.14 (32.0)
Day 1 reactor contents	121.93 (100)	71.68 (58.8)	50.78 (41.6)

5.6 Operation of the Second Unmixed, Recirculated Digester (Reactor 2)

Reactor 2 was identical to reactor 1 in terms of size, recirculation and feed composition. The results for this reactor are also similar as can be seen from Table 5.6.1 and Figures 5.6.1 to 5.6.6.

Figure 5.6.1 R2 Biogas Production

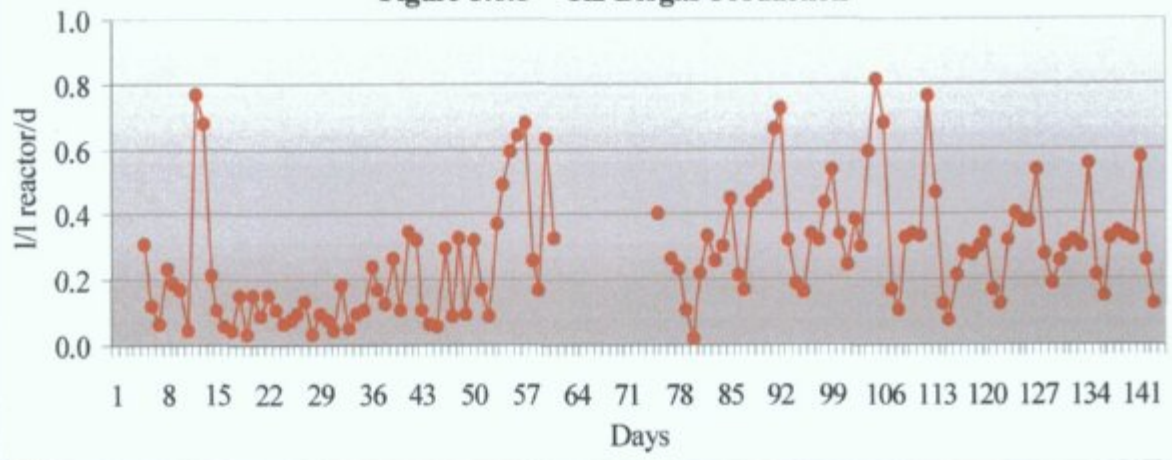


Figure 5.6.2 R2 Biogas Composition

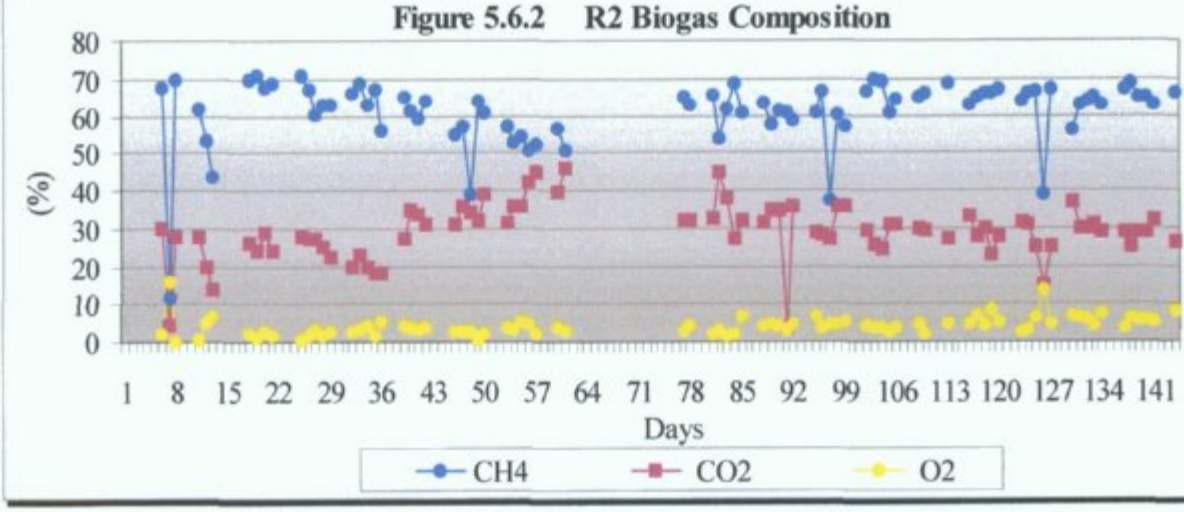


TABLE 5.6.1 REACTOR 2 RESULTS SUMMARY TABLE

Period (Days)	Organic Loading Rate (OLR) gCOD/d		COD (effluent) g		Retention Time (RT) days	Volumetric Loading Rate (VLR) gCOD/l/d		Biogas Volume l/l/d		Methane Composition %		PH (effluent)		VFA (effluent) mg/l		Total Solids Removal %		COD Removal %	
	M	*S.D.*	*M*	*S.D.*	*M*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*
1-12								0.24	0.23	53	24								
13-34	3.8	0.57	0.33	0.09	15.5	0.81	0.12	0.12	0.13	64.9	7	7.44	0.13			82.1	12.3	91.3	2.46
35-62	5.13	1.5	1.02	1.08	56.2	1.09	0.32	0.28	0.19	56.8	6.6	7.39	0.12	225.3	113	82.4	12.5	80	14
63-75																			
77-94	5.4	1.83	1.5	1.45	26	1.15	0.39	0.31	0.19	61.8	3.96	7.27	0.11	126	194	34	25.4	72.2	44.5
95-108	14.9	4.06	2.95	1.43	24.9	3.19	0.86	0.39	0.20	61.3	9.3	7.44	0.13	826	505	72.2	5.5	80.2	4.15
109-140	18.3	4.64	9.26	3.68	19.9	3.9	0.99	0.31	0.14	63.9	6.19	7.45	0.1	3450	2038	48.7	11.7	49.4	10.4
140-144								0.32	0.23										

M = Mean S.D. = Standard Deviation

Figure 5.6.3 R2 Influent and Effluent Solids

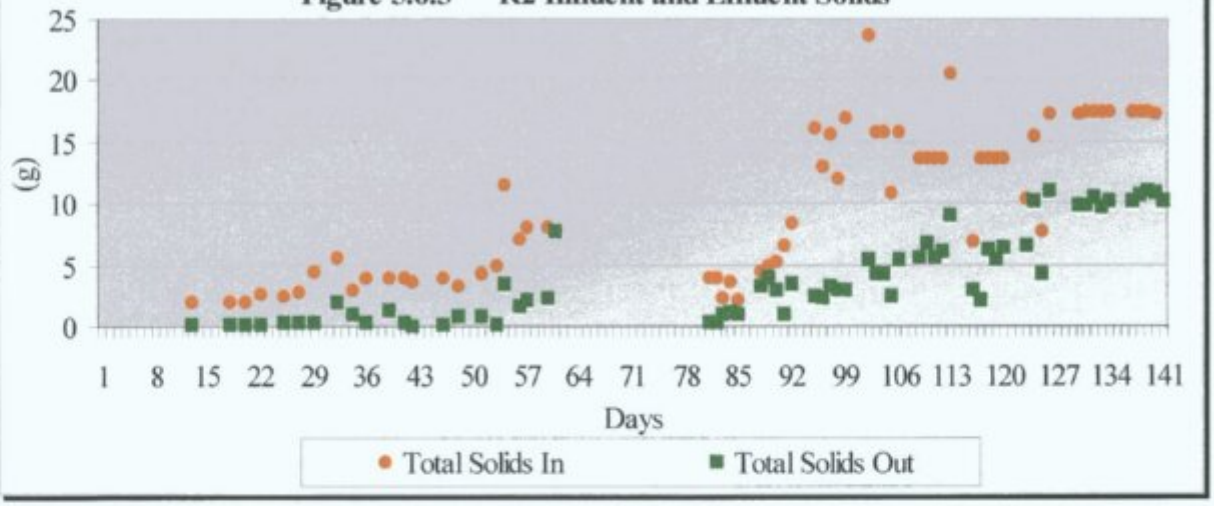


Figure 5.6.4 R2 Influent and Effluent COD

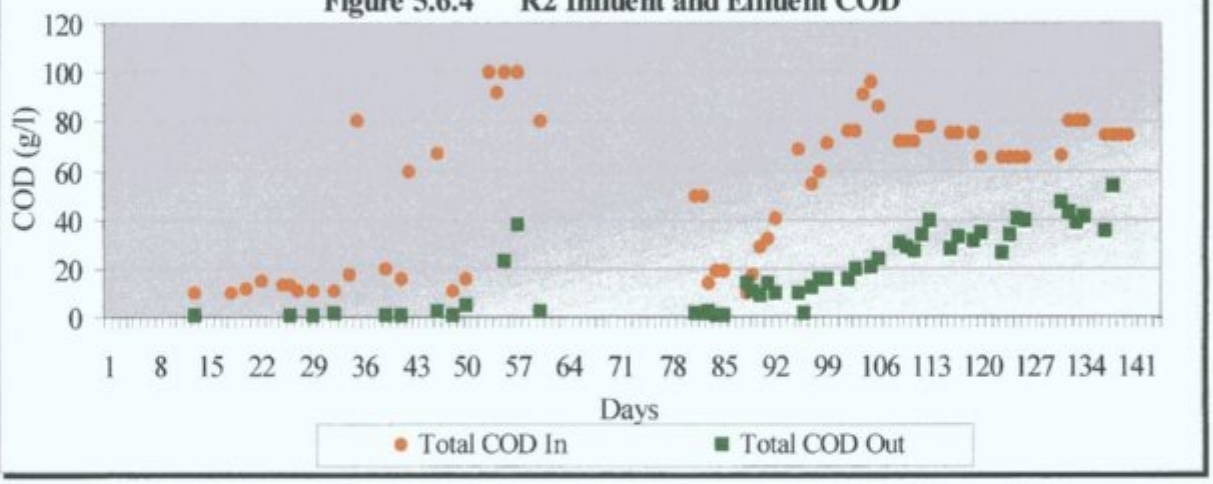
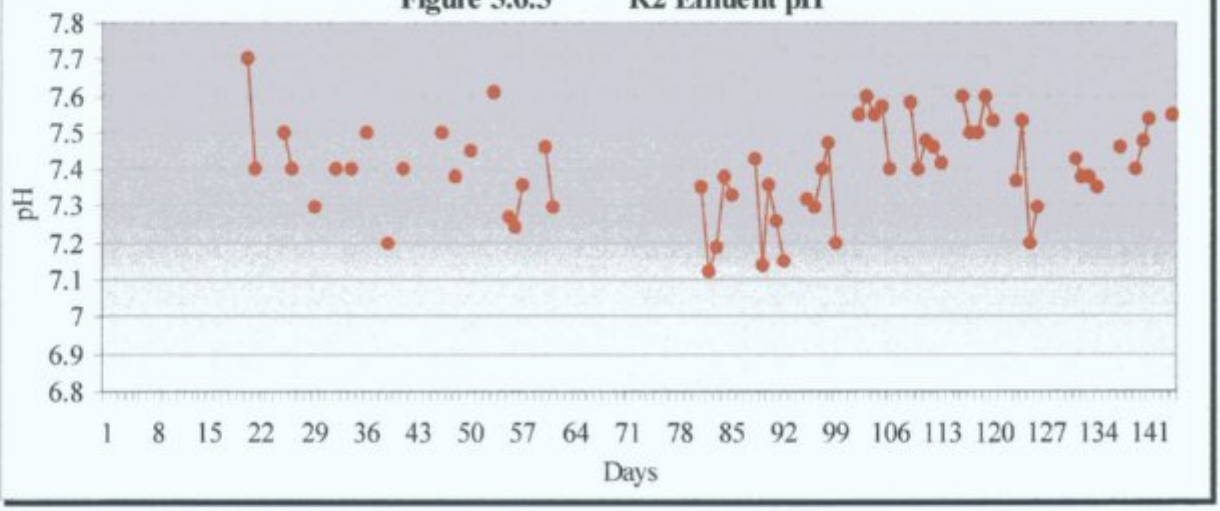
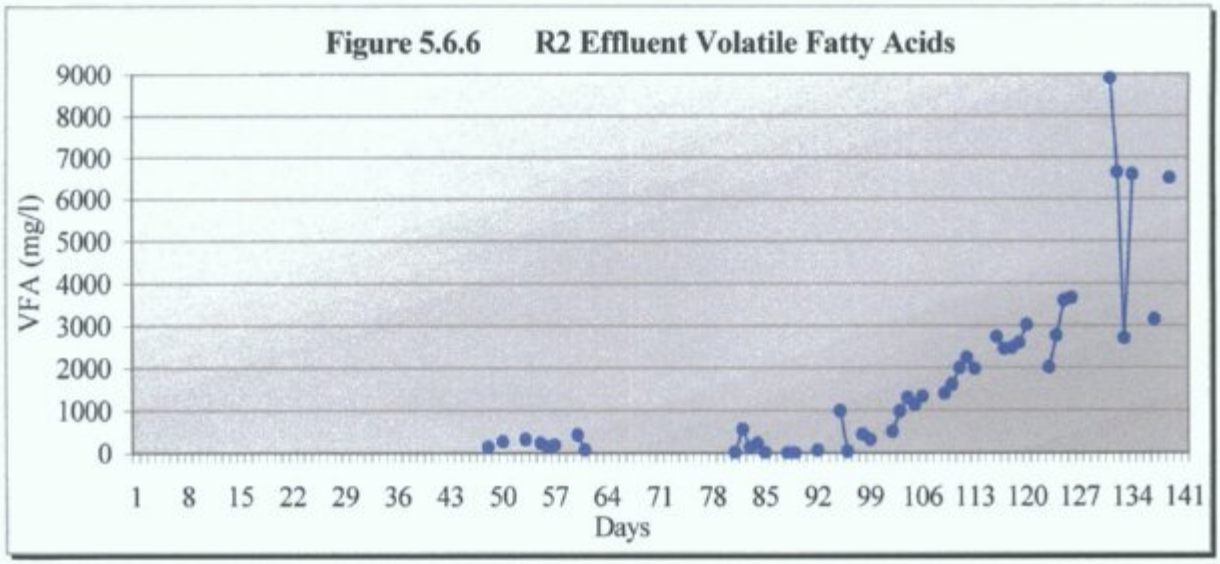


Figure 5.6.5 R2 Effluent pH





5.6.1 PERIOD VII, DAYS 109-139

The feed regime, quantity and composition were the same as for R1 throughout the study period. The performance of R2 was also similar to R1 until day 124, when R1 exploded. Variations in operating conditions occurred between the two reactors at this time, as a direct result of the explosion.

The mean biogas production in R2, was 0.31 l/l/d, which was a decrease of 0.08 l/l/d compared with the previous period (Figure 5.6.1 & Table 5.6.1).

The mean percentage methane in R2 was 63.9%, which was an increase from the previous period. From day 109 to 124 the average percentage methane was 65.7%, varying very little. On day 126 the percentage CH_4 acutely dropped to 39% (Figure, 5.6.2). There was no obvious reason for this except perhaps due to VFA inhibition in the digester leading to the start of the souring process. However, the following day (day 125) the percentage CH_4 had

increased again to 67%. On day 130 the percentage CH_4 was 56%, which was significantly low. This was attributed to the opening of the digester on that day to remove the scum layer that had built up on the top of the digester contents. While carrying out this process some air would have filtered into the digester contents. As the methanogens are extremely sensitive to oxygen this would have caused the inhibition of some methane-producing bacteria. However the process recovered and the following day the percentage CH_4 was 63%. The average percentage CH_4 for the remainder of this period was 65%.

The COD removal rate was 49.4%, which would suggest that 3.1 l CH_4/d should have been produced, however only 0.93 l CH_4/d was produced. This figure would be equivalent to only 14.5% COD removal. Either the reactor only removed 14.5% COD by digestion or the some biogas was escaping from the biogas collection system.

The solids removal rate was 48.8%. This would be closer to the expected figure of 50% than observed during the pervious periods.

The VFA concentrations in the effluent increased rapidly during this period. The mean value was 3504 mg/l, which is well above the value for optimum performance (Figure 5.6.6). However the alkalinity value was increasing with the VFA, therefore keeping the pH stable. The mean pH was 7.45 (Figure 5.6.5), which is perfect for optimum performance. The mean alkalinity was 6663 mg/l.

During this period, problems occurred with the recirculation equipment. This was directly related to the composition of the animal slurry and the build up of the scum layer in the top of the digester. On days 109, 112 and 127 the recirculation pipe became blocked with scum and

the pump stopped working. On day 130, the scum layer was removed from the top of the digester the same as for R1 (Section 5.5.7.1). This scum layer was similar shown for R1 in plates 5.1 and 5.2.

5.6.2 Decommissioning of Reactor 2

Feeding was discontinued to R2 on day 140. The recirculation and heating was stopped on this day. The heating jacket was removed from around the outside of the reactor. It was now possible to see into the digester contents and the layering was obvious. Again between day 128 and day 140 the scum layer had formed to a depth of 300mm on the top. From the bottom to about half way up the digester it was apparent that the solids were thicker, therefore confirming that settlement had taken place. The reactor was allowed to cool down for the weekend before further tests were taken.

On day 144, the digester was opened and the 250 mls of a scum layer was again scooped out. The solids content of this was measured. Then the remaining contents in the reactor were stirred well and a sample was taken so as the solids could be measured (Table 5.6.2).

Table 5.6.2 Solids in R2 at the Start and Finish of the Pilot Study

Row	Sample	Total solids g
1	Day 1 sludge sample (seed) TS	121.9
2	Cumulative TS entered R1	639.8
3	Cumulative TS exited R1	253.5
4	Day 144 sludge sample TS	302.6

The total solids removed were 32.1%. These figures also suggest that 180.7g solids were retained in the digester, thus the low solids removal. A reason for this trapping of solids may have been due to the fact that R2 was not mixed and solids were settling to the bottom of the reactor.

From Table 5.6.3 it is obvious that there was quite a large percentage of organic matter (volatile solids) and solids in the scum layer (75.5%).

TABLE 5.6.3 R2 COMPOSITION OF SLUDGE AT THE START AND FINISH OF THE PILOT STUDY INCLUDING THE SCUM LAYER

Sample	Total Solids g (%)	Volatile Solids g (%)	Fixed Solids g (%)
Scum layer	52.87 (100%)	39.93 (75.5%)	12.93 (24.45%)
Day 144 reactor contents	249.75 (100%)	160.36 (64.2%)	89.39 (35.79%)
Day 1 reactor contents	121.93 (100%)	71.68 (58.78%)	50.78 (41.64%)

5.7 Operation of the Mechanically Mixed Digester (Reactor 3)

5.7.1 Period I, Start-up, 1-12 days

The reactor was seeded and commissioned as in section 5.3. Once the digester reached optimum temperature (~35°C) the stirrer was commissioned, (day 4), and the contents of the digester was mixed slowly for 15 minutes every hour. Table 5.7.1 summarises the results obtained for Reactor 3.

The biogas production was recorded from day 4 (Figure 5.7.1) and the mean volume produced for this period was 0.07 l/l/d. The composition of the biogas was recorded from day 6 (Figure 5.7.2) and the mean percentage methane (CH_4) for this period was 71.7% (Table 5.7.1).

Figure 5.7.1 R3 Biogas Production

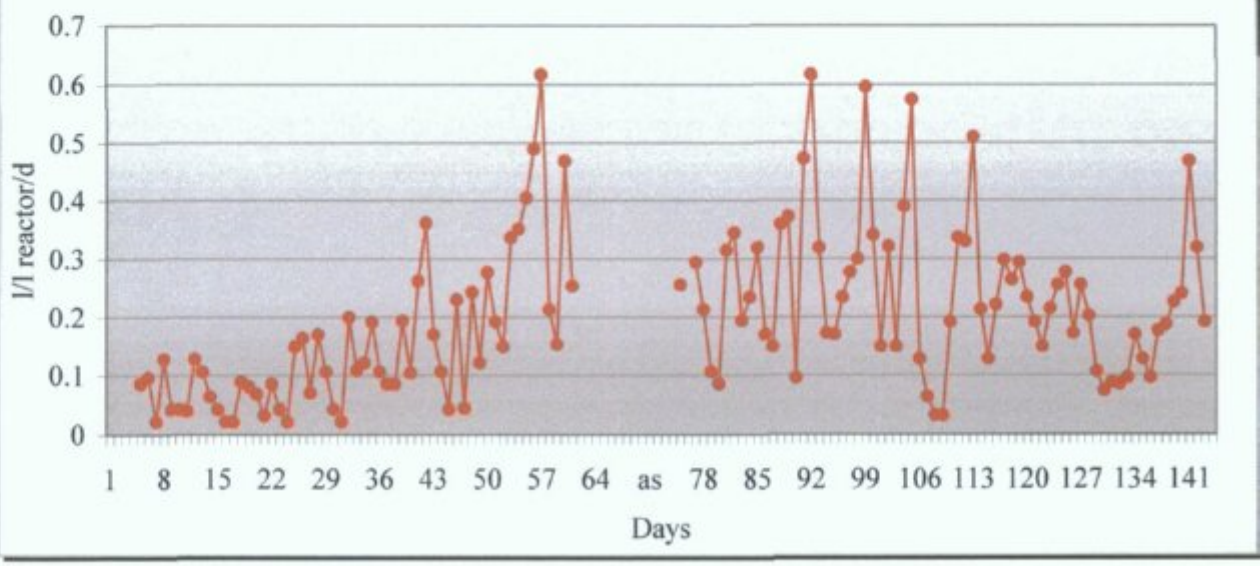
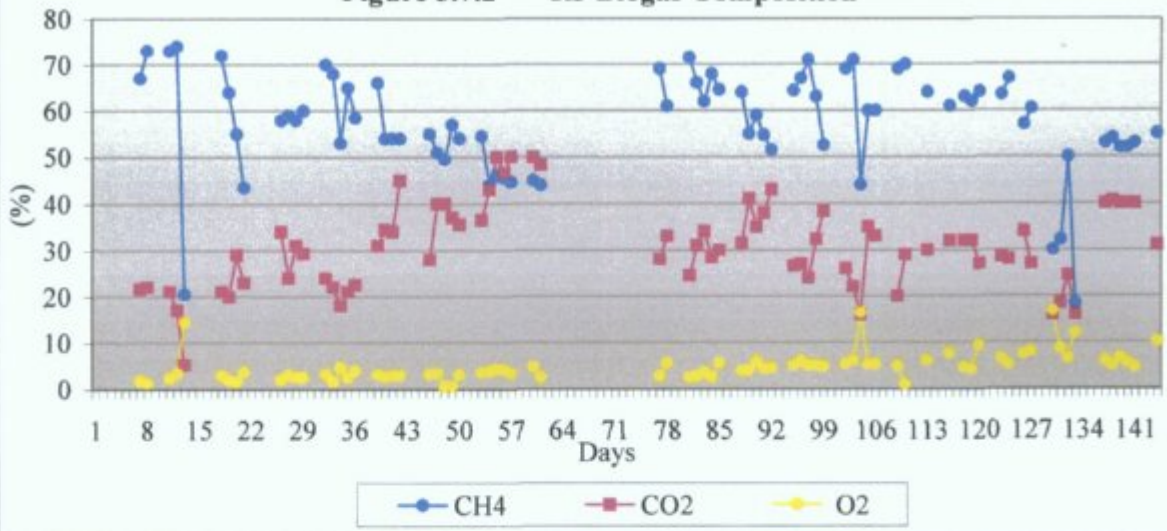


TABLE 5.7.1 REACTOR 3 RESULTS SUMMARY TABLE

Period (Days)	Organic Loading Rate (OLR) gCOD/d		COD (effluent) g		Retention Time (RT) days	Volumetric Loading Rate (VLR) gCOD/l/d		Biogas Volume l/l/d		Methane Composition %		PH (effluent)		VFA (effluent) mg/l		Total Solids Removal %		COD Removal %	
	M	*S.D.*	*M*	*S.D*	*M*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*	*M*	*S.D.*
1-12								0.07	0.04	71.7	3.2								
13-34	3.8	0.57	1.26	0.1	15.5	0.81	0.12	0.08	0.05	56.7	13.8	7.4	0.23			27.7	24.3	66.8	20.8
35-62	5.13	1.5	0.91	0.56	56.2	1.09	0.32	0.23	0.14	52.3	6.7	7.21	0.19	365.7	548	50.8	43.6	82.2	34.4
63-75																			
77-94	5.4	1.83	1.26	0.6	26	1.15	0.39	0.27	0.14	62.2	6.2	7.22	0.14	130.1	178	13.3	34.7	76.6	27.7
95-108	14.9	4.06	3.1	1.7	24.9	3.19	0.86	0.27	0.17	63	8.5	7.38	0.17	691.8	481	78.7	4.43	79.2	2.47
109-140	18.3	4.64	4.68	2.08	19.9	3.9	0.99	0.1	0.17	54.8	13.9	7.38	0.12	2986	1408	58.0	11.2	74.4	10.6
140-144								0.33	0.14										

M = Mean S.D. = Standard Deviation

Figure 5.7.2 R3 Biogas Composition



5.7.2 Period II, days 13-34

The feed regime, quantity and composition were the same as for R1 and R2, for this period. Figure 5.5.3, in section 5.5.2, shows the organic loading rate and feed composition for the days the reactors were fed.

The mean biogas production was 0.08 l/l/d with a mean percentage methane equal to 56.7% (Table 5.7.1). The biogas volume decreased initially after the first feed and the percentage CH_4 dropped dramatically to 20.5%. This was expected as the microorganisms, especially the methanogens, would initially be inhibited by the substrate and would need time to acclimatise to the new conditions. On day 18, the digester was fed again, the biogas production for that day was 0.09 l/l/d (Figure 5.7.1) and the percentage CH_4 was 72% (Figure 5.7.2). This result indicated stabilised conditions for the microbial consortia. For the next 3 days both the biogas production and the percentage CH_4 decreased gradually until day 21 when only 0.03 l/l/d of biogas was produced and only 43.7% of this was CH_4. This indicated that acclimatisation of

the biomass to the substrate was not yet optimised. A higher volumetric loading rate was applied on day 22, to compensate for no reactor feeding over the weekend. From days 25 to 34, both the biogas production and the percentage CH_4 increased slightly indicating greater reactor stability.

The mean COD removal was 66.8% over this period (Table 5.7.1). This figure would suggest that 0.89 l CH_4/d should have been produced however, only 0.66 l CH_4/d was actually produced, indicating that only 16% COD was removed by digestion or a quantity biogas was escaping from the collection system (see Sample Calculation, section 5.5.2). The solids removal was 27.7%. Figures 5.7.3 and 5.7.4 shows the solids in the influent and effluent and the COD respectively.

The mean pH value was 7.4, which would be considered optimum for reactor operation (Table 5.7.1).

Figure 5.7.3 R3 Influent and Effluent Solids

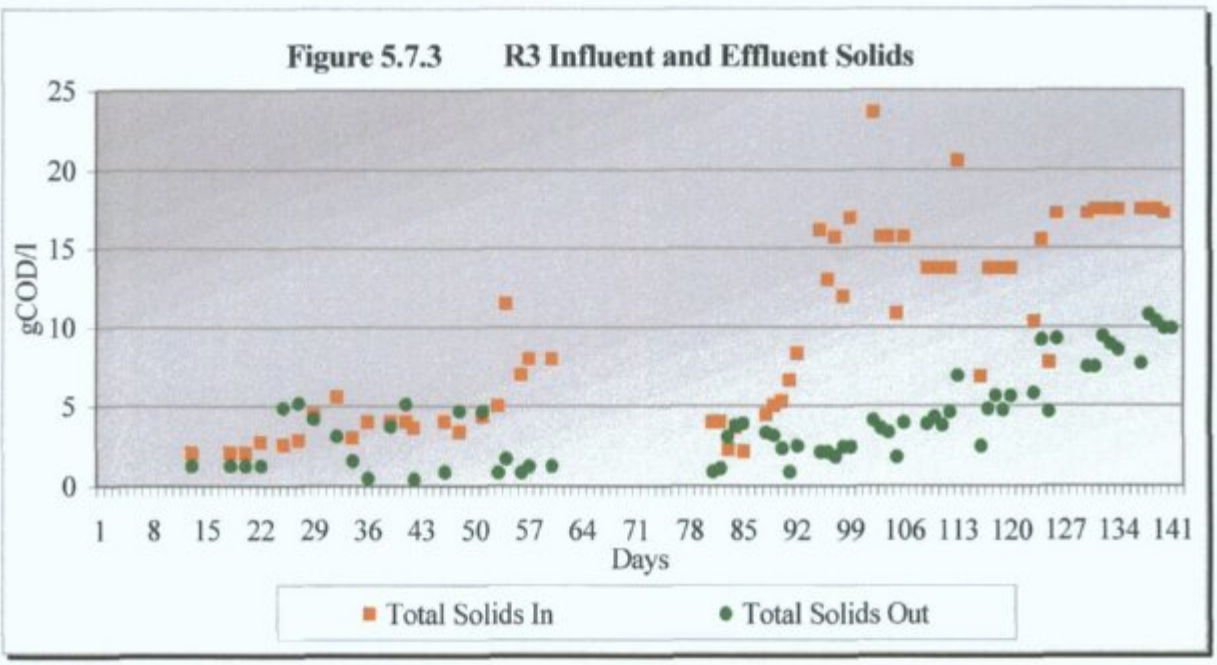
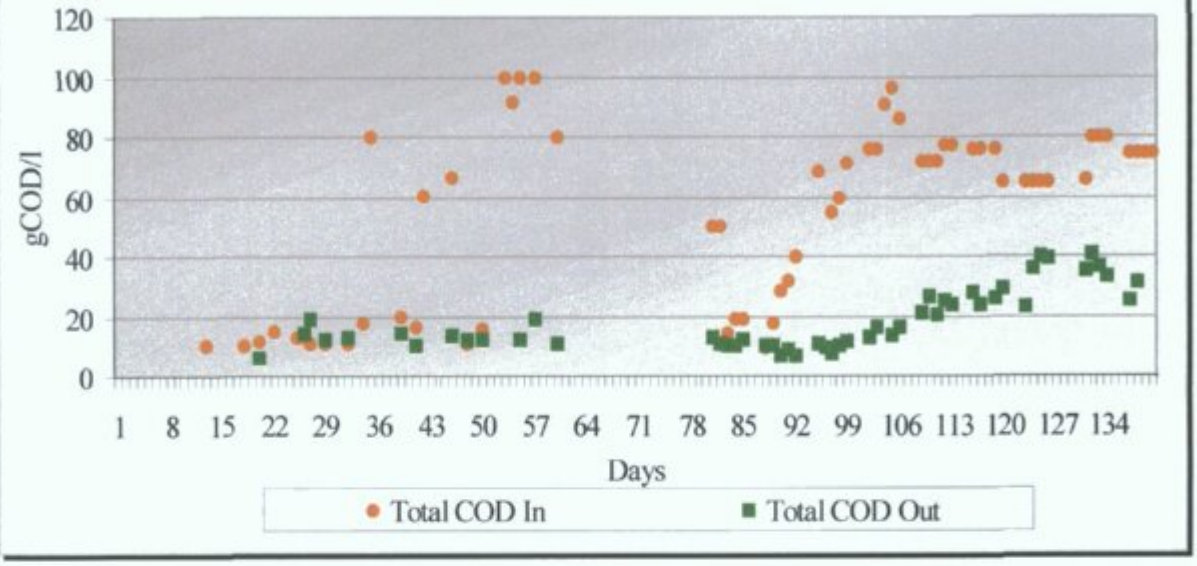


Figure 5.7.4 R3 Influent and Effluent COD



5.7.3 PERIOD III, DAYS 35-62

The feed regime, quantity and composition were the same as for R1 and R2, for this period (Section 5.5.3 & Figure 5.5.3).

When the secondary sludge was added, on days 48 and 50, there was no apparent increase or decrease in the biogas volume (Figure 5.7.1). Whey was added again from day 53. When this substrate was added consecutively for 3 days, (days 53, 54 & 55), the biogas production increased. This increase could also be attributed to the increase of the volumetric loading rate from 1.06 to 1.7 gCOD/l/d.

The mean percentage methane for this period was 52.3% (Table 5.7.2). This was a decrease on the pervious period. On day 35, methane composition was 65% but dropped to 58.5% on day 36. This may have been because the methanogens were acclimatising to their new

conditions. In general, the percentage methane was low and may be attributed to the composition of the whey. When the volumetric loading rate increased towards the end of the period the percentage methane decreased to an average of 44.8% (Figure 5.7.2). This would suggest that the methanogens were not coping with the increase in organic load and that inhibitory concentrations of the VFA were building up.

There was a definite increase of VFA and the mean for the period was 365.7 mg/l (Table 5.7.1). The last sample before an extended shutdown period, on day 61, shows that they had risen to 1707 mg/l, leading to definite cause for concern (Figure 5.7.5). However, these levels dropped again during the shutdown period. While this increase in the VFA was observed, the pH stayed above 7, with a mean value of 7.21, however this was a decrease of 0.23 on the pervious period (Figure 5.7.6). The increase in the VFA shows that the methanogens are not converting the intermediate acids as fast as they were being produced. A build up of these VFA would lead to a drop in pH and consequently inhibit the methane-producing bacteria eventually leading to reactor souring and process failure. From the results, it was apparent that the methanogens had been inhibited somewhat and the cause would be the significant build up of the VFA, which would be directly connected to the quantity of organic matter entering the digester. The methane-producing microorganisms have a much slower growth rate than their former acid-producing microorganisms. Similar high levels were not recorded for R1 and R2. This could be attributed to the mixing of R3 leading to less retention of solids including entrapment of the slow growing methanogenic bacteria. The low solids concentration and mixing may be leading to washout in R3 that did not occur in R1 and R2.

The COD removal was calculated to be 82.2% for this period. This is equivalent to 1.4 l CH_4//d. This is how much CH_4 should have been produced if 82% COD was digested,

however only 0.56 l CH_4/d was produced (see Sample Calculation, section 5.5.2). This figure would suggest that only 31% COD was removed by digestion, either this was the case or biogas was escaping somewhere from the gas collection system. The solids removal was 50.8%.

5.7.4 Period IV, days 63-75

As with R1 and R2, reactor 3 was shutdown on day 62, to allow for a holiday shutdown period.

5.7.5 Period V, days 77-94

The feed regime, quantity and composition were the same as for R1 and R2, for this period. During this second stage of the study recirculation was introduced in reactor 3 along with the mechanical mixer.

The mean biogas production for this period was 0.27 l/l/d. This was the largest quantity of biogas produced so far in the study. On days 88 and 89, when the whey was added, the biogas production increased slightly, however on day 83 when the primary sludge and whey was added, the biogas volume dropped to 0.19 l/l/d (Figure 5.7.1). A reason for this could be that the microorganisms were acclimatising to the new substrate. By day 85, the biogas volume showed a gradual increase to 0.33 l/l/d. On days 88 and 89, the biogas production continued to show a steady increase, however on day 90 the volume was only 0.1 l/l/d. This was because the gas valve on the gas collection chamber was closed by accident over night and the gas could not escape into the gas collection tank. In this case the gas escaped under

pressure out through the base sealing section (Figure 2.5, Chapter 3). As the volumetric loading continued to increase on days 81 and 92, so did the biogas production, with gas production reaching 0.62 l/l reactor on day 92. This was the largest amount of biogas produced since feeding of the reactor commenced.

The mean percentage methane production for this period was 62.2 %. The CH_4 production was good for days 74 to 81, of this period only dropping slightly on day 83 (this would have been caused by the change of feed, slightly inhibiting the methanogens). The following week, days 88 to 92, the percentage CH_4 showed a gradual decrease until it reached 51.5% on day 92 (Figure 5.7.2). This decrease could be related to the gradual increase in the organic loading. Despite the drop in percentage methane the mean VFA value of the effluent was 130 mg/l (Figure 5.7.5) with an average pH of 7.22 (Figure 5.7.6). The latter two parameters indicated reasonable process stability.

COD removal for this period was 76.6% (Table 5.7.1). This indicates that 1.4 l CH_4/d should have been produced, however only 0.79 l CH_4/d was produced, indicating that only 42% COD was actually removed or biogas was escaping (see Sample Calculation, section 5.5.2).

The solid removal was 42.7%, which is significantly high for the present process conditions (i.e. low solids in influent). Figures 5.7.3 and 5.7.4 shows the solids entering and leaving and the COD the digester.

5.7.6 PERIOD VI, DAYS 95-108

The feed regime, quantity and composition were the same as for R1 and R2, for this period (Section 5.5.6 & Figure 5.5.3).

The mean biogas production recorded for this period was 0.27 l/l/d (Table 5.7.1). This was the same as for the previous period. The initial biogas volume on day 95 was 0.17 l/l/d (Figure 5.7.1), which was the same as day 94. From days 95 to 99 the biogas volume showed a rise of up to 0.6 l/l/d. On day 102, the biogas volume recorded was less and decreased further on day 103 to 0.15 l/l/d. However, these lower gas volume recordings may have been due to gas pipe clogging at the top of the digester due to scum. At this stage, day 102, the scum layer was touching the lid of the digester causing the blockage of this gas outlet. To solve this problem an extra 300mls was removed from the digester to allow the scum layer to come away from the top of the digester. On days 104 and 105, the biogas production recording increased significantly to 0.57 l/l/d. However on day 106 the biogas production dramatically decreased to 0.13 l/l/d. There was no obvious reason for this unless the digester had failed, due to some type of shock.

The mean percentage methane recorded for this period was 63 %. It was initially high for the first 3 days and even reached 71% on day 97 (Figure 5.7.2), which would suggest an affinity of methanogens for the animal slurry. However, on days 98 and 99, the percentage methane recorded began to drop and reached 52.6% by day 99. By the following week the percentage CH_4 had increased again to 69% and 71% on days 102 and 103 respectively. On day 104, the percentage CH_4 decreased acutely to 44%. The only explanation for this was that some air might have entered the digester when the blockage in the gas pipe was being relieved,

causing inhibition to the methanogens due to O_2. After 2 days the percentage CH_4 recovered to 60%.

The percentage COD removal for this period was 79.2%, which was equivalent to 4.1 l CH_4/d production. From the mean biogas production and methane composition, only 0.19 l CH_4/d was actually produced. However some of this biogas was lost when the gas pipe was blocked. Nevertheless, the 79% COD removal figure probably came about because of the increase in organic loading rate. This new organic loading rate would take a while to impact the contents of the digester.

The solids removal rate was equal to 82.2%. A reason for this high removal result may relate to the increase in solids concentration in the feed entering the digester (Figure 5.7.3).

The mean VFA value was 691.8 mg/l (Table 5.7.1), which is a significant increase on the pervious period (Figure 5.7.5). The main increase was between days 102 and 106, when it increased to an average of 1083 mg/l. This increase was certainly a cause for concern. However, on examination of the pH values, it was observed that the mean pH for this period was 7.38, which was well above neutral (Figure 5.7.6). This would suggest that even though the VFA were increasing they were not causing the pH to decrease, which would in turn inhibit the methanogens. Therefore, this would indicate that there was enough alkalinity being produced within the digester to keep the digester contents at the optimum pH value. The digester was exhibiting self-buffering characteristics. However, the high VFA concentrations in the effluent, suggested diminishing process stability due to the slurry feed.

Figure 5.7.5 R3 Effluent Volatile Fatty Acids

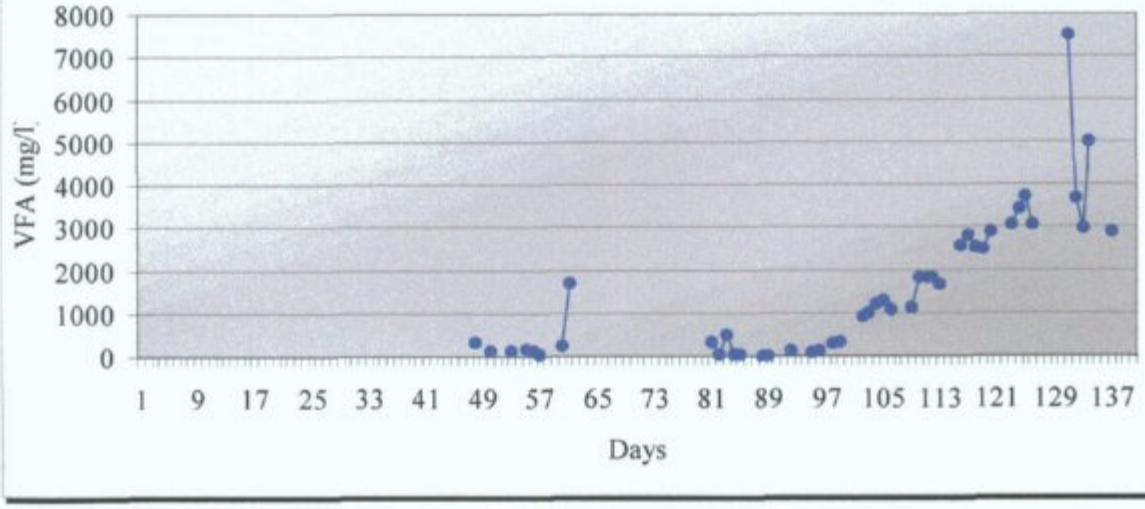
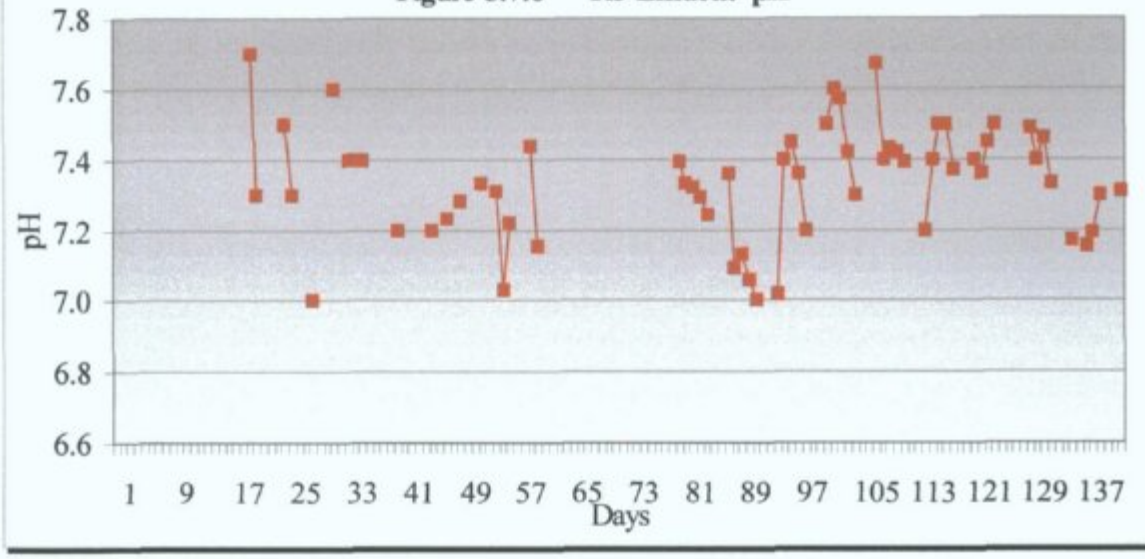


Figure 5.7.6 R3 Effluent pH



5.7.7 PERIOD VII, DAYS 109-139

The feed regime, quantity and composition were the same as for R1 and R2, for this period (Section 5.5.7 & Figure 5.5.3).

The mean biogas production decreased to 0.1 l/l/d compared with the previous period (Figure 5.7.1). Initially, the reactor seemed to have been shocked from the previous period. However, when the substrate was added the biogas volume increased gradually until it reached 0.51 l/l/d on day 113. From days 116 to 126 the biogas production decreased to a mean of 0.21 l/l/d. A reason for this lower biogas production was that on five different occasions, overnight the gas outlet in the digester became blocked stopping the gas from entering into the gas collection tank where the biogas was measured. The scum layer, as already discussed, caused this problem.

On day 130, the scum layer was removed with as little as possible agitation to the digester contents. This scum was similar to that observed in reactor 1 (Plates 5.1 and 5.2). The digester was then sealed and the recirculation started again. For the next few days the biogas production was very low. The following week (days 137 to 140) the biogas volume gradually increased from 0.18 to 0.24 l/l/d (Figure 5.7.1).

The mean percentage methane production was 54.8%. From days 109 to 124 the average CH_4 concentration was 64.8% however, for the remainder of the period this average decreased to 45.9% (Figure 5.7.2). This was presumably caused by oxygen shock to the anaerobic bacteria when the digester was opened to remove the scum layer.

As well as oxygen shock, it was observed that the VFA concentrations were increasing significantly (Figure 5.7.5). The mean for this period was 2986 mg/l which is well above the acceptable level of between 200-800mg/l. Despite the high VFA concentration pH levels remained above 7. The mean pH was 7.38, however it can be seen that towards the end of this period it did seem to be dropping slightly (Figure 5.7.6). The lowest pH value was 7.15, on

day 139. The alkalinity was obviously sufficient to maintain a neutral pH. However, it can be seen that as the VFA increased so did the alkalinity. This suggests that the slurry feed has good buffering characteristics.

The percentage COD removal for this period was 74.4% suggesting that 4.7 l CH_4/d should have been produced. However, only 0.25 l CH_4/d was recorded as an unknown quantity of the biogas was lost when the gas outlet became blocked several times. Thus, this biogas production calculation for this period is not accurate. The mean solids removal was 55.1% (Table 5.7.1).

5.7.8 DECOMMISSIONING OF REACTOR 3

Feeding was discontinued to R3 on day 140. The recirculation, heating and stirrer systems were switched off on this day. The heating jacket was removed from around the outside of the reactor. It was now possible to see into the digester contents. The digester contents were not as stratified as that observed for digesters 1 and 2, therefore the stirring device was successful at keeping the solids in the digester well mixed. Between day 128 and day 140 a second scum layer had formed to a depth of 300mm from the top of the digester contents. The reactor was allowed to cool down for the weekend before further tests were carried out.

On day 144, the digester was opened and the 250 mls of a scum layer was again scooped out which was the same quantity removed from all three reactors. The solids content of this was measured. Then the remaining contents in the reactor were stirred and a sample was taken for solids measurement. Table 5.7.2 shows the results of the final solids taken.

TABLE 5.7.2 SOLIDS IN R3 FROM START AND FINISH OF THE PILOT STUDY

Row	Sample	Total solids g
1	Day 1 sludge sample (seed) TS	121.93
2	Cumulative TS entered R1	639.8
3	Cumulative TS exited R1	267.95
4	Day 144 sludge sample TS	232.1

The total solids removed for the entire study were 36.2% solids removal. A certain quantity of solids and organic matter were being retained in the reactor and not removed. A net retention of total solids over the 144-day operation was 109.17g. This was not expected because the reactor was mechanically mixed, however this figure was considerable less than in the other two reactors, indicating that the mixer was of some benefit.

TABLE 5.7.3 R3 COMPOSITION OF SLUDGE AT THE START AND FINISH OF THE PILOT STUDY INCLUDING THE SCUM LAYER

Sample	Total Solids g (%)	Volatile Solids g (%)	Fixed Solids g (%)
Scum layer	63.75 (100%)	45.5 (71.37%)	18.26 (28.69%)
Day 144 reactor contents	168.35 (100%)	111.57 (66.3%)	56.77 (33.72%)
Day 1 reactor contents	121.93 (100%)	71.68 (58.78%)	50.78 (41.64%)

From Table 5.7.3 it is obvious that there was a large percentage of organic matter (volatile solids) in the scum layer (71.37 %), therefore organic matter was being trapped in the reactor.

5.8 COMPARISON OF REACTORS PERFORMANCE

From the results obtained for each reactor previously discussed the following section compares the performance of each reactor in terms of feeding, mixing, scum formation, solids removal etc.

5.8.1 EFFECT OF SUBSTRATE COMPOSITION ON THE PERFORMANCE OF THE REACTORS

In terms of substrate composition on reactor performance, all 3 reactors displayed the same patterns. From Figure 5.8.1, it was observed that whey exhibited the best biogas production rates in terms of gCOD whey to l biogas/l reactor/d (l/l/d). When whey was added in a greater organic loading rate than the secondary sludge (days 13 & 18, 35 - 47 and 52-65) the biogas production increased considerably than when secondary sludge was added in larger organic loading rates. This was apparent for all three reactors. For the second stage of the study, whey was added with sludge until day 109, after this the biogas production decreased significantly, especially in reactors 1 and 2, however this decrease may also be related to the increase in organic loading discussed later in this section.

For all 3 reactors, the percentage methane decreased noticeably towards the end of the first stage of the pilot trials, even though the biogas production was increasing (Figure 5.8.2). This may have been due to the increased loading rate or the whey composition (Figure 5.8.3). This was most apparent for Reactor 3 (mechanically mixed), when the CH_4 dropped to between 40 and 50% on day 44. The whey may be lacking in essential nutrients for the biomass, which the sludge feed would have had naturally.

Figure 5.8.1 Comparison of Substrate Composition with Biogas Production

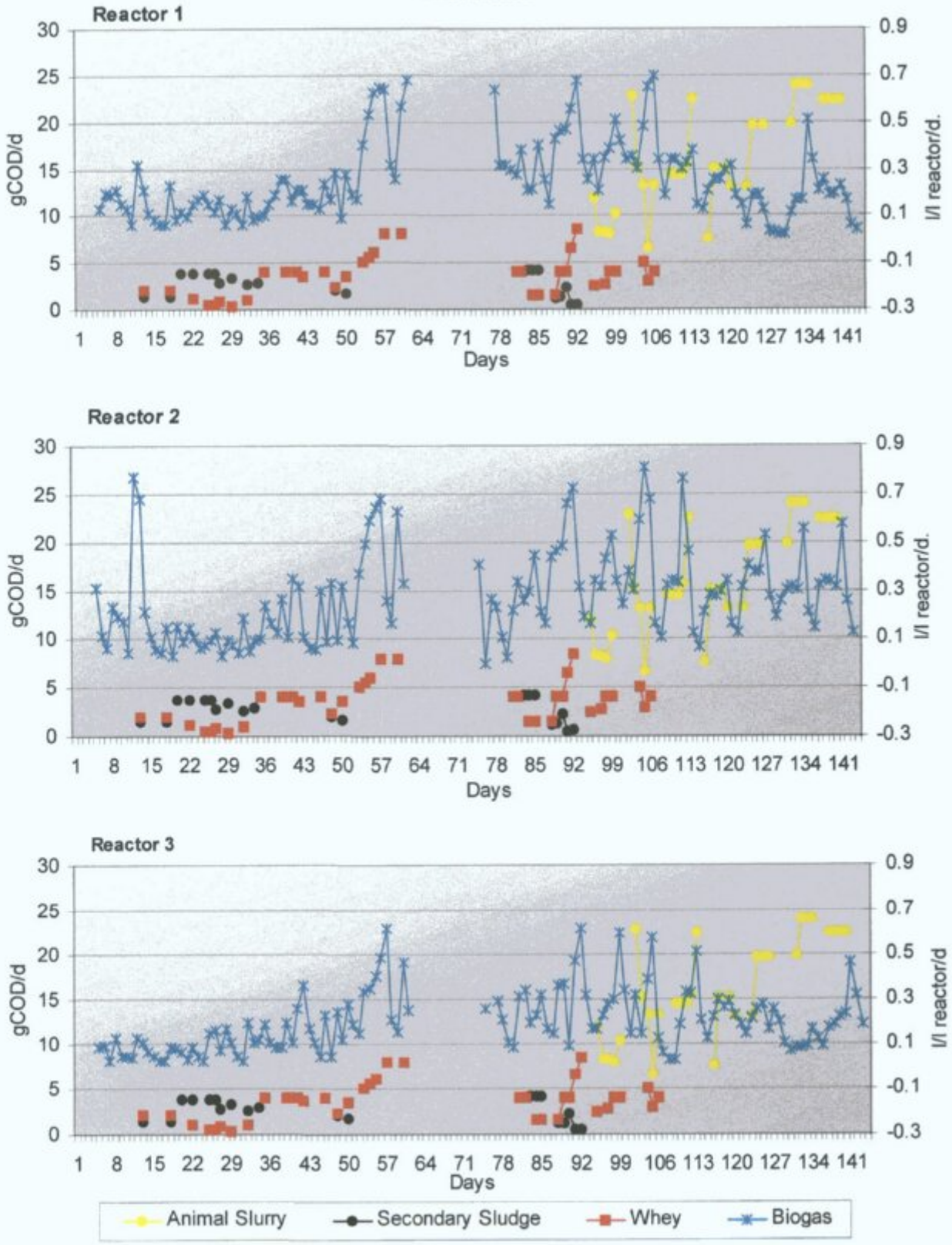


Figure 5.8.2 Comparison of Biogas Production with Methane Production

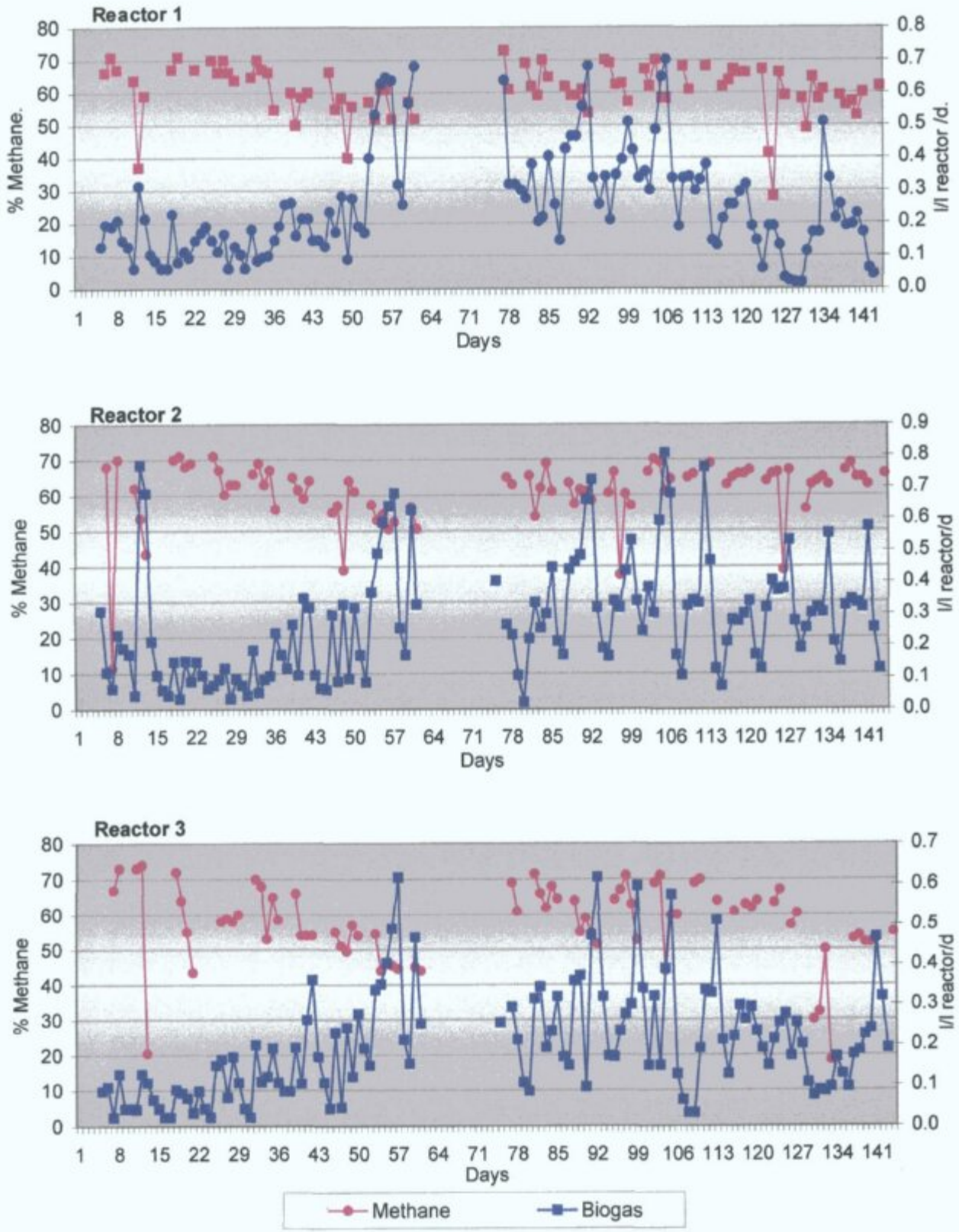
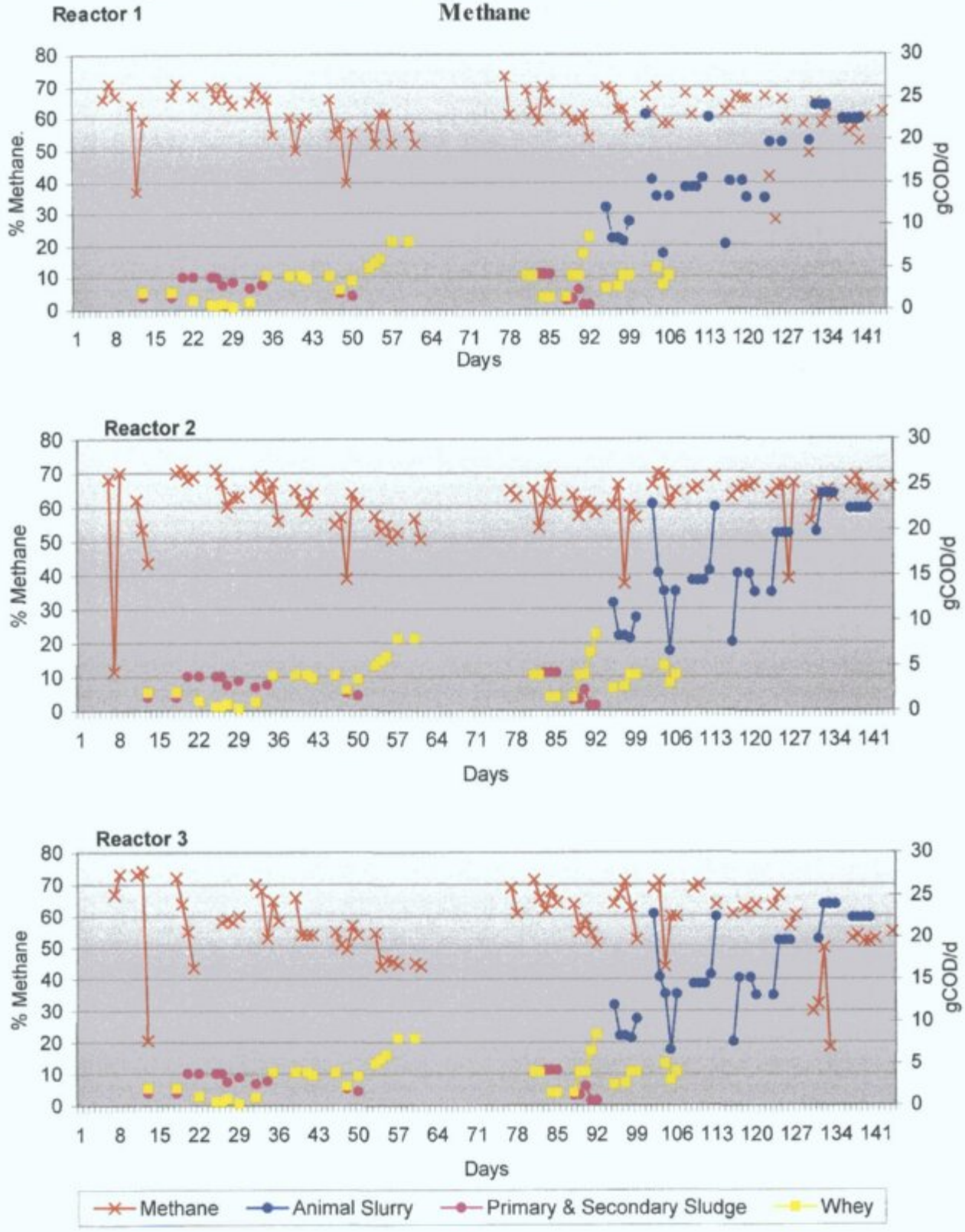


Figure 5.8.3 Comparsion of Substrate Composition with Percentage Methane



In the second period the percentage methane in reactor 1 and 2 was in the expected range of 60 - 70% except when the methanogens were shocked with the new substrate (animal slurry) and on a separate occasion, when there was an oxygen shock. Therefore the methane-producing microorganisms sustained the increase in organic loading. This would suggest that the decrease in methane during the first stage of the study, in all reactors, was caused by the composition of the whey and not the increase in organic loading rate. Indeed this was evident again in the second stage, when whey was added in a greater quantity than the sludge, on days 89 to 93, all of the reactors showed a decrease in methane production.

5.8.2 DEGRADABILITY OF SEWAGE SLUDGE AND ANIMAL SLURRY

Much of the slurry from animals is excretal and the slurry formed is basically similar in chemical composition to the primary sludges of a municipal sewage plant. Not all the compounds, of similar chemical analysis in different sludges are equally available for bacterial metabolism. Some analysis of sludges and slurries are given in Table 5.8.1. (Hobson, 1990). Primary sludge contains faecal particles and pieces of toilet papers and other paper and cotton wool materials. Most of these papers have being chemically treated to make them soft and soluble and the digester bacteria will easily degrade the cellulose in them.

Cellulose in sewage sludge is mainly residues of prepared and cooked vegetables used as human food, along with tissue papers and other materials made of delignified, manufactured, plant fibre. Such lignin free material is relatively easily degraded by microbes (Hobson, 1990). More complex microorganisms from human faeces or aerobic treatment plants (activated sludge) can contain starch-like storage polysaccharides as well as more complex polysaccharides in different cell structures. Gut secretions contain complex

mucopolyccharides. Some of these complex polysaccharides may be virtually resistant to degradation in the digestion process. Overall, they probably do not contribute to a large proportion of the gas generation in digesters. In sewage sludge digestion lipids (fats) contribute to a large proportion of the gas produced (Hobson & Wheatley, 1993).

Most animal wastes contain more fibre than sewage sludge. The cellulose and hemicellulose in animal excreta, while of similar general analysis to sewage sludge, contain higly-lignified residues of the leaves, stalks and grain husks of animal feeds, which have resisted microbial degradation in the animal rumen or hind-gut. Since lignin is non-degradable anaerobically and its presence prevents degradation of plant carbohydrates these fibres are much more resistant to further microbial degradation than are the fibres in the sewage sludge (Hobson, 1990). The most important source of gas in digestion of faecal material from farm animals is the residues of the vegetable matter (carbohydrates, and volatile fatty acids formed by fermentation of carbohydrates) from the foodstuffs passing through the gut. Depending on the animal and feed, the plant residues in animal faeces will tend to form a poorer substrate for the digester bacteria than the residues in human faeces and require a longer retention time for degradation. Farm waste may also contain animal bedding which are relatively difficult to degrade. (Hobson & Wheatley, 1993).

During the present pilot trials the digesters were mainly fed secondary and primary sludge mixed with whey from days 13 to 92. The performance of all digesters was as expected in terms of biogas production and methane composition, with the only variations occurring when high quantities of whey was added, as discussed previously in section 5.8.2. The animal slurry was introduced on day 95. Initially biogas production and methane composition was good. However, as a direct result of the physical composition of the animal slurry the digester

performance decreased gradually for all reactors. A thick scum layer formed, trapping organic matter in the digesters and causing blockages to frequently occur in the pipes and eventually causing the explosion in reactor 1. Mechanical mixing did not appear to ease the problem of scum formation as a significant scum layer developed in the mechanically mixed reactor also. While the microorganisms initially coped with the animal slurry composition it was quickly observed that the animal slurry would need longer retention times as the physical composition was difficult to degrade. However, this was not impossible, because of its large plant residues and grassy composition. A scum removal system would have been of great benefit, eliminating the need to open the reactors allowing oxygen to enter the digesters. These problems did not occur when the primary and secondary sewage solids were entering the digesters.

TABLE 5.8.1 THE COMPOSITION OF SOLIDS FROM DAIRY CATTLE SLURRY AND SEWAGE WASTES (% DRY WEIGHT)

Waste from	N	Fat	ADF	NDF	Lignin	Ash
Fattening Cattle	1.6	2.9	44.5	60.3	11.8	8.2
Dairy Cattle	2.3	6.5	41.0	56.7	13.8	14.1
Pigs	7.4	13.7	23.6	45.2	8.1	14.0
	Cellulose	Hemicellous	Lignin	Fat	Protein	Ash
Sewage sludge	3.8	3.2	5.8	34.4	27.1	24.1

where *N - organic nitrogen (in compounds such as protein)*

ADF - Acid Detergent Fibre (cellulose and lignin).

NDF - Neutral Detergent Fibre (cellulose, hemicellulose and lignin).

(Hobson, 1990)

5.8.3 Effect of Irregular Feeding on the Performance of the Reactors

The irregular feeding pattern at start-up had the obvious effect of producing irregular volumes of biogas on a daily basis (Figure 5.8.1). In order to ensure uniform biogas production, it is obviously preferable to feed the digester at frequent intervals. For the first half of this study the reactors were fed every second day and were not fed over the weekend. This is apparent in biogas production from the Figure 5.8.1, as the biogas volumes also decreased every second day and at the weekends. During the second part of the study the digesters were fed every day but not during the weekends. Therefore, if a decrease in biogas production occurred during the week, it was not due to the lack of feed.

5.8.4 Effect of Organic Loading Rate on the Performance of the Reactors

All reactors were fed the same feed on the same day at the same organic loading rate (OLR). The quantity of biogas produced was dependent on the organic loading rate. This can be seen for all three reactors in Figure 5.8.1. As the OLR increased the quantity of biogas produced increased. However, towards the end of the study, as the OLR increased the biogas production decreased for all 3 reactors. This would suggest that the consortium of microorganisms could not cope with an organic loading rate of over 15 gCOD/d. The organic loading entering the reactors was continued at this high loading rate to see if the digesters would fail or acclimatise to their new loading rate. To achieve very high loading rates requires close monitoring of pH, volatile acids concentration, alkalinity and often trace metal sufficiency, because the process can become unbalanced in a short time at high loading rates (Speece, 1996). It should be noted that the pilots were operated for a short period (144 days) and with a variety of different feeds. If just one feed was used reactor stability and greater

OLRs may have been achieved. It is also possible that after more days operation VFA levels may have dropped as observed in the full-scale plants. Despite the increase in OLRs leading to diminishing stability actual reactor souring did not occur at the higher loading rates. It was apparent however that increasing the OLR would have led to further reactor instability for all reactors.

The behaviour of the solids reduction varied slightly from reactor to reactor as the OLR was increasing. Reactor 1 and 2 generally shared the same pattern, however reactor 3 (mixed reactor) showed significant variations. From days 13 to 61 of the study reactor 1 and 2 showed large solids reduction. This was caused by the retention of solids by settlement, which had occurred in the reactors due to lack of mixing. In reactor 3, the solids reductions were somewhat lower and were more representative of the digestion process. Some retention by settlement also occurred in this reactor however at a lesser quantity than that of the other reactors. This was because this reactor was mixed and the solids were evenly graded throughout the digester. On several days the reactors achieved no solids reduction whatsoever. This was because when the whey was being added on its own it had a very low solids content (lower than the contents of the digester).

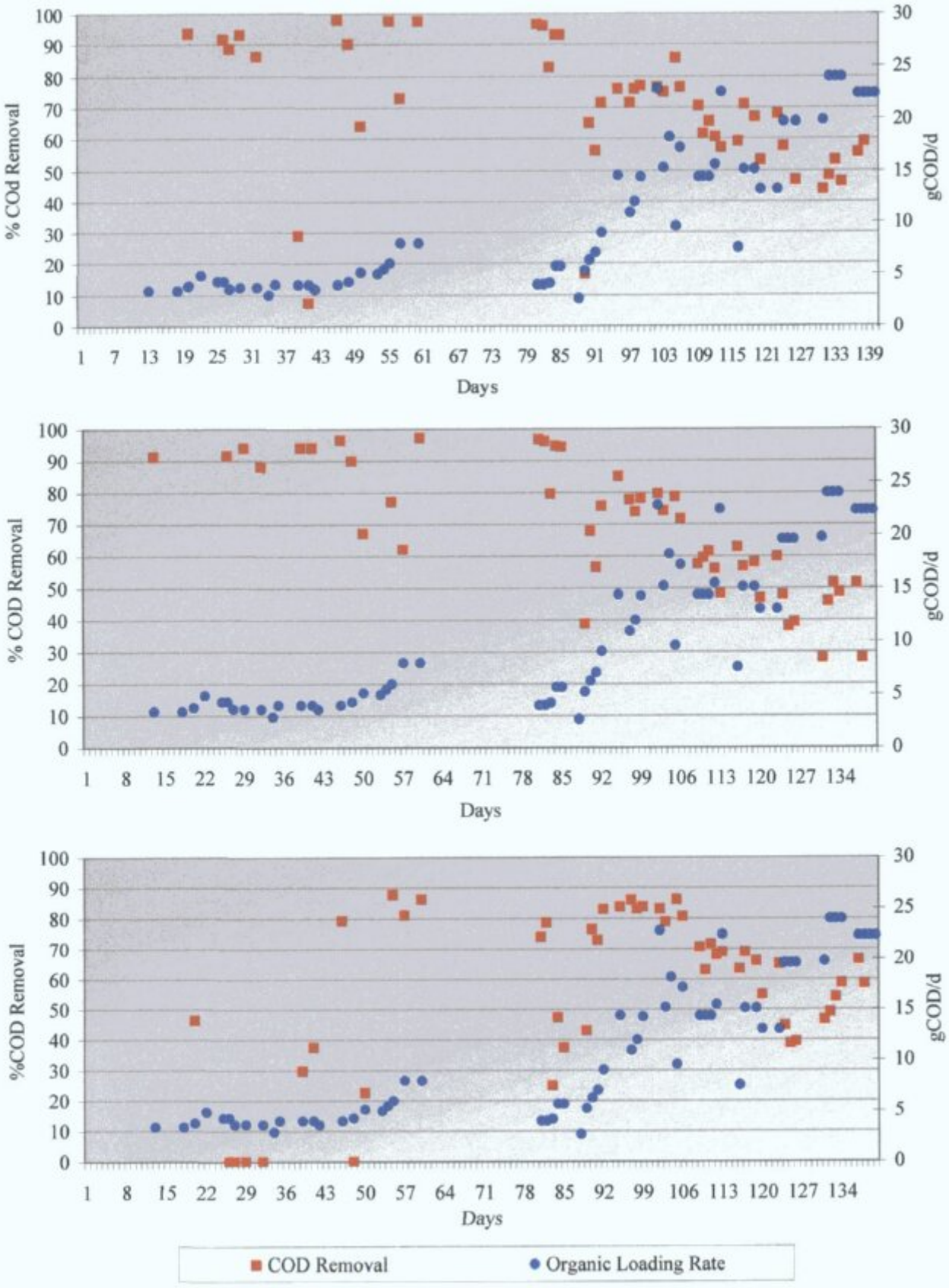
From results obtained from Tullamore sewage treatment plant it was shown that, the fewer solids there were entering the digester the less solids reduction occurred. This was observed when the picket fence thickener was decommissioned for repairs and the solids entering the digester were not thickened. These solids entered the digester at approximately 1.22% and exited the digester at approximately 1.23% therefore very little removal. Once the picket fence thickener was working again the solids removal increased to between 30-40%.

From days 109 - 140 of the trials the percentage solids removal in each of the reactors decreased with the increase in solids entering the digesters. For example percentage total solids removal average was 54%, 48% and 54% for R1, R2 and R3 respectively, for days 109 to 140, whereas the higher performances of 67%, 74% and 67% were obtained respectively for days 95-108. This reduction commenced at the time the animal slurry was added on its own to the digesters. The animal slurry had a much higher solids content than the other feeds. The digestion process became under stress as less solids would have been digested, which was indicted through the decrease in biogas production. However, on average the solids removal was much higher (days 95 - 140) than would be expected in an anaerobic digestion process for all reactors because of its decreasing performance. This may be due to the fact the digester was not operating long enough to show the true solids removal due to building of solids in the reactors. It is expected that a change in the anaerobic process takes one complete retention time for acclimatisation. Even more time may be required in this case due to then physical composition of the animal slurry making it difficult to degrade.

The scum formation trapping solids and leading to poor contact between the substrate and the biomass would also inhibit digestion performance considerably.

In all reactors the COD removal can be seen to decrease when the organic loading rate increases especially when the animal slurry was added (Figure 5.8.4). However this could be due to the fact that the process was coming under stress due to the new feed and higher loading rate and as a direct result biogas decreased therefore indicating there was very little organic matter being digested. This would also take one complete retention period or longer to adjust and show true COD removal rates. After one retention period passed (day 128) all three reactors showed an increase in COD removal (Figure 5.8.4).

Figure 5.8.4 Comparsion of COD Removal with Organic Loading Rates



5.8.5 Effect of Scum Formation on the Performance of the Reactors

It was obvious that the scum layer was caused by the composition of the animal slurry. This animal slurry had a very grassy texture, which caused considerable stress to the feed/recirculation pumps once feeding of the animal slurry commenced. In order to reduce the grass content of the animal slurry, it was sieved before entering the digester. This may have solved the feed problem, however it did not abate the scum formation. Scum started to form on the top of the digester a few days after the slurry was added, and by two weeks the layer was 50mm thick floating on the top of the reactor contents. This large mass of scum reduced the effective capacity of the digester and therefore the retention period and interfered with mixing and recirculation of the tank contents. The scum layer was removed on day 130 and two days after the scum layer was removed it was noticed to be forming again and at the end of the study a similar amount of scum was removed.

All three reactors formed the same quantity of scum. The total solids content of this scum layer was 60.3g, 52.8g and 63.7g and the volatile solids content of these solids was 47.5g, 40g and 45.5g for reactor 1, 2 and 3 respectively. Therefore solids and organic matter were being trapped and retained in the digester by the formation of this scum layer, as was evident in the decrease of solids and COD removal rates once the animal slurry was added to the digesters. From these results it can be seen that there was no significant difference between the mixed reactor and the unmixed reactors when it came to scum formation.

All of the reactors exhibited a decrease in methane when the reactors were opened to remove the scum layer. This was caused by oxygen getting in to the reactors. After the reactors were closed, reactor 1 and 2 seemed to recover well, however in reactor 3 the methane content

remained low (average 55%). The oxygen shock may have inhibited more methane-producing microorganisms in R3 than in the other reactors, taking it longer to recover. The mixing of R3 may have resulted in oxygen getting to more sections of these reactors than the others making the oxygen shock more significant for R3.

5.8.6 Overall Performance of the Reactors

In terms of reactor performance none of the reactors reached optimum performance. If this was the case the reactors would be producing biogas approximately 2-3 times their reactor volume and the percentage methane would be approximately 70%. However it was not the intention of this study to reach optimum performance conditions. The time span of the study was short and the reactors were fed a variety of different feed compositions. Thus, these reactors never reached optimum stable conditions.

From the summary tables (Tables 5.5.1, 5.6.1 and 5.7.1), it can be seen that, the reactor which produced most biogas was reactor 2 with an average of 0.28 l/l/d. Reactor 1 (also unmixed with recirculation) was next and very close behind reactor 2 producing 0.27 l/l/d. This is not surprising as both R1 and R2 were of identical configuration with identical feeding regimes. Reactor 3 produced an average of 0.19 l/l/d, a considerable quantity lower than reactor 1 and 2. This may be partially due to the unsuitability of the low solids whey feed to the mechanically mixed anaerobic digester. These entrapped solids in the unmixed R1 and R2 would have retained more methane forming microorganisms.

All the reactors almost produced the same percentage of methane approximately 60.6%. The COD removal rates were basically the same for all reactors, at 72.7%, 74.6% and 76.4%,

respectively. These figures were high and not very representative of typical COD removal rates expected of the process performance, partly due to settlement (especially for R1 & R2) and the formation of the scum layer. Based on the relationship 1g of COD stabilised = 0.35 l CH_4 (Parkins & Owen, 1986), the theoretical volume of methane produced was calculated. In all reactors the actual measured methane production per unit of organic matter destroyed was less than the theoretical values. Possible reasons for this include: only a fraction of the organic matter was converted to new bacterial cells (microbial growth); and not all of the gas produced is accurately measured and captured by systems used in the field; and solids retained by settlement in reactors but not digested.

The solids removal over the entire 144-day of the trials varied slightly. Reactor 1, 2 and 3 removed 36.9%, 32% and 36.2% total solids, respectively. Approximately 50% removal would be expected from a digester operating well (Hammer, 1996; & Metcalf & Eddy, 1991). However, solids reduction in anaerobic digestion will vary from plant to plant regardless of the efficiency of the digestion process, because of the feed sludge characteristics as well as the upstream processing (Parkins & Owen, 1986). From the above results reactors 1 and 3 removed the same quantity of solids whereas reactor 2 removed less. All reactors retained solids. Reactors 1, 2 and 3 retained 163.2g 180.7g and 109.17g solids, respectively. This would suggest that reactor 3 digested the most solids when the retained solids are taken into account, as reactor 3 retained less solids then reactor 1. This was expected, as reactor 3 was the mechanically mixed reactor. Reactor 1 and 2 retained more solids then reactor 3, as they were not mechanically mixed.

In conclusion reactor 1 and 2 seemed to have performed better then reactor 3, in terms of biogas production. However, reactor 3 removed slightly more solids then R1 by digestion

because it was mixed allowing improved contact between the microorganisms and the difficult to digest solids. Many potentially digestible solids may have remained trapped and undigested in the bottom of Reactors 1 and 2.

5.9 INFORMATION GIVEN BY DIFFERENT PARAMETERS

5.9.1 MONITORING PARAMETERS OF THE ANAEROBIC PROCESS

Several parameters have been suggested as stress indicators. Some of the most commonly used indicators include measurement of biogas production, biogas composition, pH, volatile solids destruction and VFA concentrations (Ahring *et al.*, 1995). All of the pre-mentioned parameters were used to monitor process performance, in the present pilot trials. In general, most of these indicators are suitable for detecting gradual changes. However, pH, volatile solids reduction and gas composition are often too slow for optimal detection of sudden change (Anelidaki & Ahring, 1994). One of the main reasons for the set-up of these pilot trials was to confirm this statement by observing which of the above parameters detects process imbalance at the earliest stage.

Biogas production, methane composition of the biogas production, pH range and VFA were all believed by plant operators and in the literature (Bjornsson *et al.*, 1997; Ahring *et al.*, 1995; Rozzi *et al.*, 1997) to give good warning signs of process instability. Therefore, all of these parameters were monitored very carefully throughout the pilot trials. For days 13 to 60 of the trials all three reactors were operating successfully, however not optimally. The pH range and the VFA were both within their ranges. Towards day 60 the percentage methane in the biogas started to decrease gradually in all three reactors. This reduced methane

production would cause less energy yield, less growth of the crucial methanogens and less organic pollutant reduction in the reactors and was a concern. It was thought that the composition of the substrate caused this reduction in methane production. Organic loading rate was ruled out because the VFAs had not built up in the digesters, which in turn did not effect the pH range. However, the VFA concentration in the effluent were increasing.

The reduction in methane at this time was attributed to the feed composition. Whey powder was being added on its own. The composition of this whey was believed to be the cause of inhibition of the methane-production microorganisms (Figure 5.8.3). This belief was substantiated by investigations carried out by Kelly and Switzenbaum (1984). They confirmed that trace nutrients influence reactor performance and concluded that the effect was more pronounced than temperature influence (as methanogens are sensitive to temperature change). During their investigations whey powder supplemented with nitrogen and phosphorus was still nutrient-limited by Ni, Fe, or Co or some combination of these elements (Speece, 1996). Therefore the whey powder was deficient in both micronutrients and macronutrients. Methanogens are especially inhibited by trace metal (micronutrients) deficiency therefore this decrease in methane production was probably due to a lack of micronutrients. Nutrient deficiencies also elevate VFA concentrations (Speece, 1996), which was the case in these pilot trials as the VFA in each reactor began to increase, however not at alarming concentrations. The methane production was normal when secondary and primary sludge was added with the whey as long as the quantity of whey did not exceed the quantity of the sludge. This indicated that the sludge being added had sufficient quantities of nutrients.

Methane production also decreased, in almost all cases, when a change in substrate occurred and when the oxygen shock occurred (Figure 5.8.3).

From days 95 - 140 of the pilot trials the organic loading rate was intentionally raised considerably for all reactors. This was partly to see how the process would cope with the increased loading rate and to see which monitoring parameter would indicate process instability first, as the process can become unbalanced in a short time at high organic loading rates (Speece, 1996). It was obvious (Figure 5.9.1) that as the organic loading rate was increased the VFA concentration increased simultaneously. Volatile fatty acids accumulation reflects a kinetic uncoupling between acid producers and consumers and is typical for stress situations (Speece, 1996). This increase in VFA concentration tends to reduce the pH, eventually leading to process failure. As the VFAs were increasing the pH was monitored very closely. No decrease in pH was apparent until the VFA were ~ 3000mg/l and then the pH dropped slightly down to 7.2. This was still well above neutral and by itself could not be considered as a cause for concern. Alkalinity was then measured and it can be seen that the increase in alkalinity corresponded to the increase in VFA (Figure 5.9.2). This relationship was keeping the pH above neutral. Towards the end of the study the methane composition had decreasing slightly. This was attributed to oxygen shock and not the high VFA. However the VFAs may have compounded the problem of the anaerobic microorganisms dealing with the oxygen shock.

Figure 5.9.1 R1 Comparsion of Volumetric Loading Rate with Volatile Fatty Acids

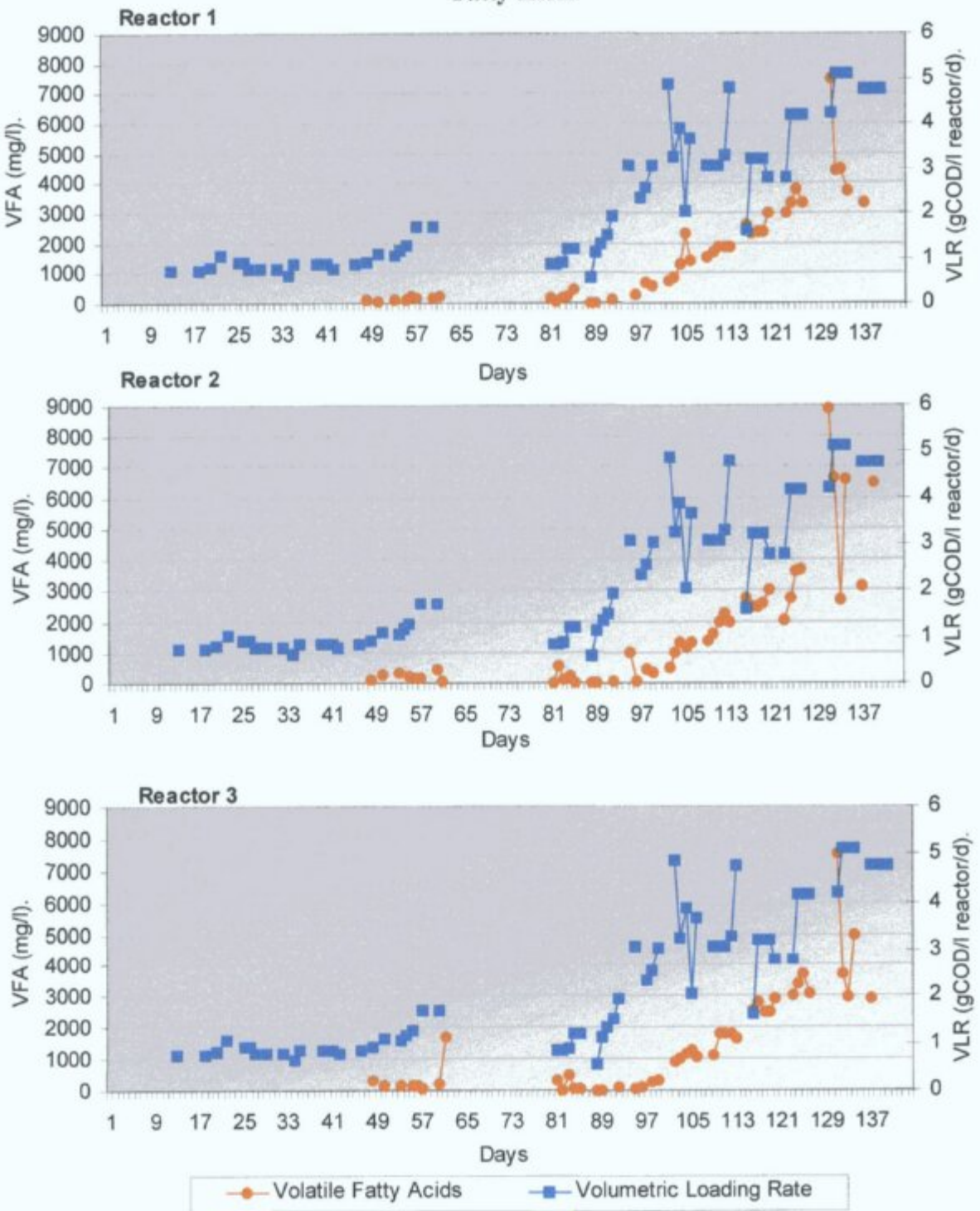
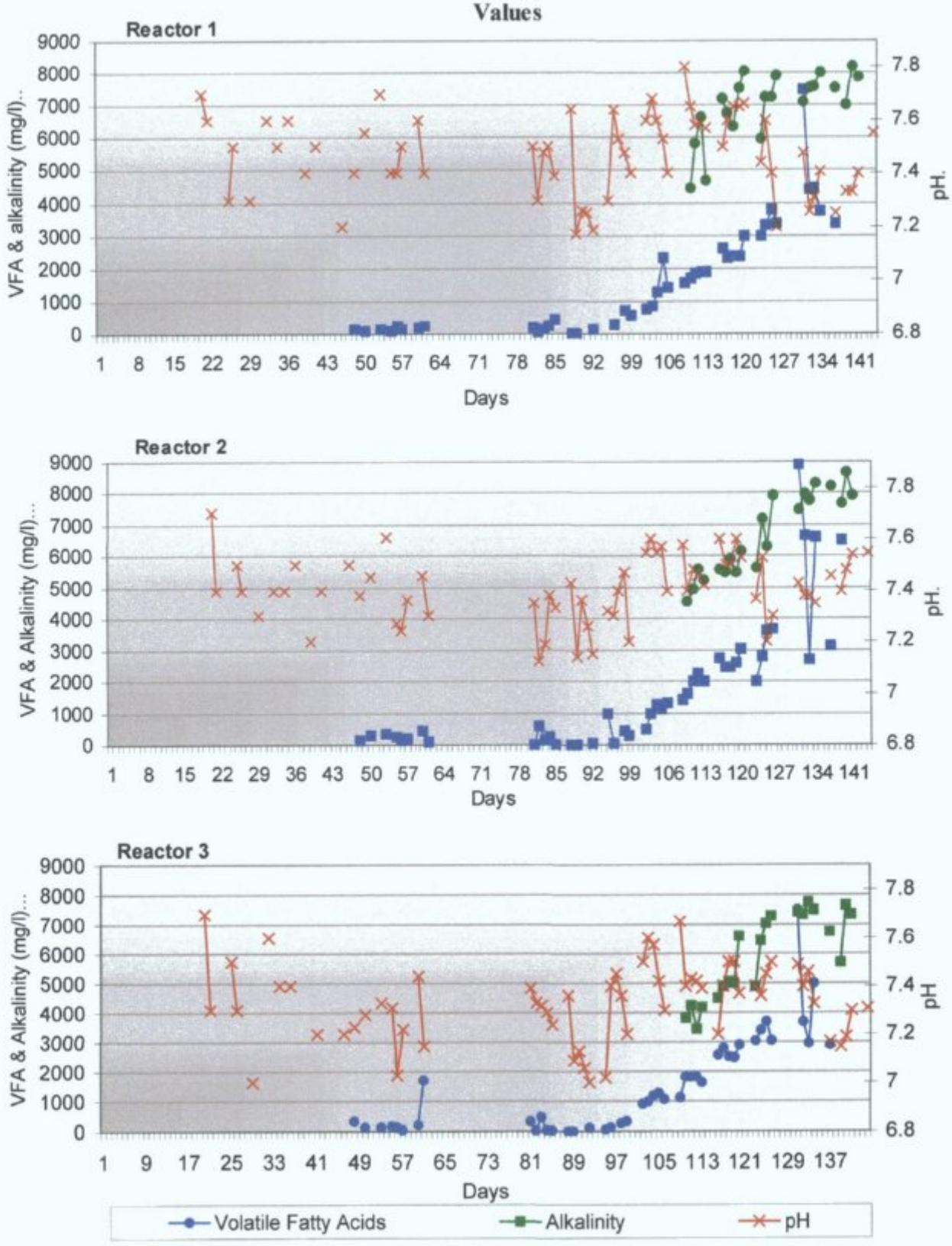


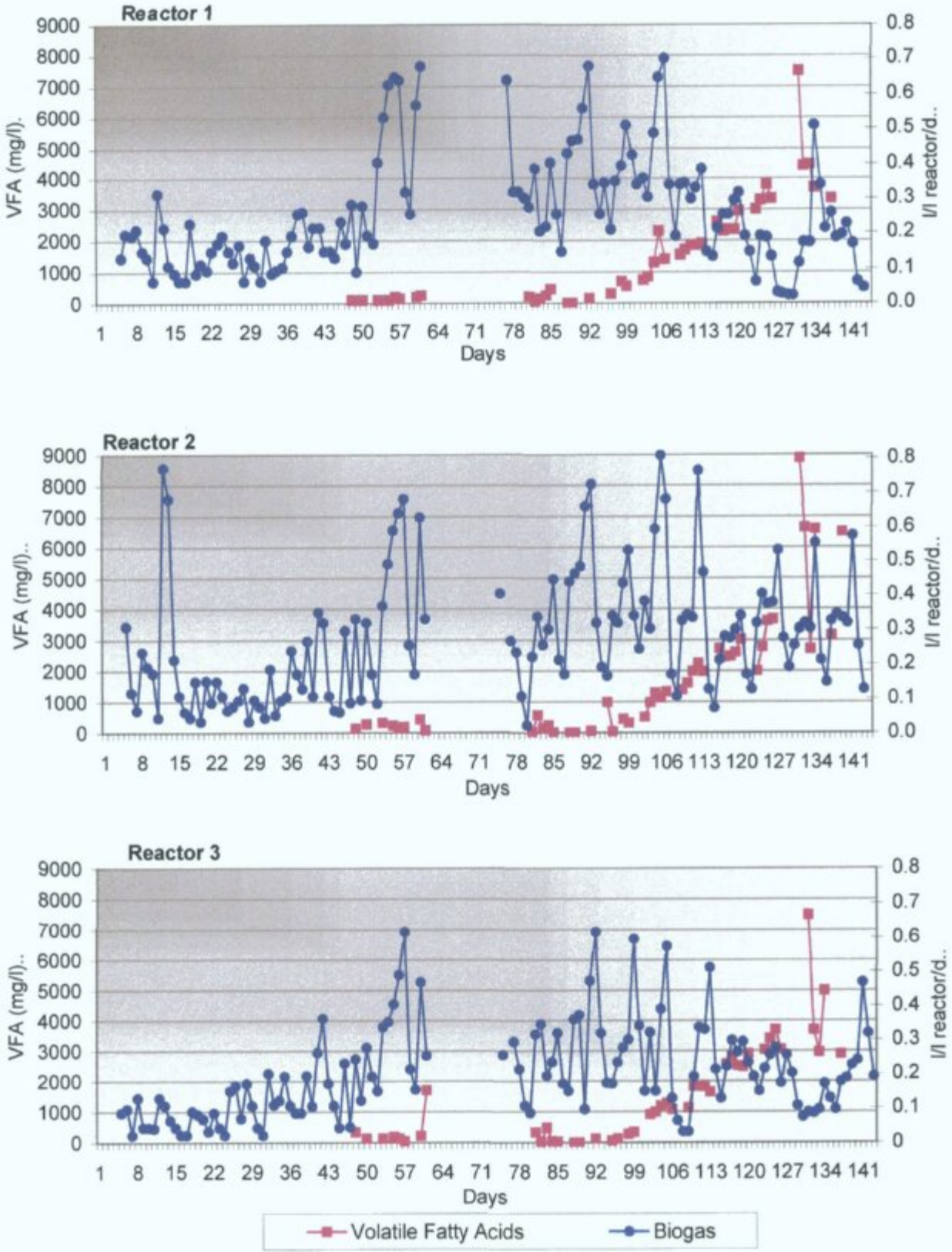
Figure 5.9.2 Comparsion of Volatile Fatty Acids and alkalinity with pH Values



The biogas production decreased in all three reactors before the oxygen shock occurred on day 130, indicating that the build up of acids had caused adverse effects, even if it did not bring the pH value down. At this stage the reactors were souring. Another reason for this decrease in biogas production could have been the formation of a thick scum layer on the top of the digester contents in all three reactors. After this scum layer was removed, all reactors suffered the effects of oxygen shock, however within a few days the biogas production in all three reactors increased significantly as if they were recovering, even though the VFA concentration in the effluent were still high (Figure 5.9.3). However, towards the end of this stage the VFA concentration showed signs of dropping in each reactor. Reactor 2 (unmixed) showed the best biogas production throughout this stage.

The reactors obviously suffered from the effects of oxygen shock and the scum layer formation. However, there are two more likely other possibilities for the increase in VFA, before the pre-mentioned disturbances occurred, which could have contributed to the low biogas production. These possibilities are the toxic effect and the nutrient deficiency effect. The toxic effects of high VFA concentrations on the anaerobic digestion process have been studied and reported by several authors (Ahring & Westermann, 1988; Gorris *et al.*, 1989; Gourdon & Vermande, 1987), and the resulting drop in pH is generally considered to be the main cause of the toxicity (Hill, 1982, Mosey & Fernandes, 1984). Several studies have shown that high concentration of VFA in themselves have no adverse effect on the biogas process (Boone, 1980, Gourdon & Vermande, 1987). This was not the case with the three reactors in this pilot study. All three reactors showed biogas volumes decrease as the VFA increased (Figure 5.9.3). However, it has been shown that degradation of propionate and butyrate is inhibited by acetate (Ahring & Westermann, 1988, Kasper & Wuhrmann, 1978).

Figure 5.9.3 Comparsion of Biogas Production with VFA



This may have occurred in the three pilot reactors as the acetate is the abundant acid produced in the acid-production stage and as the VFA concentration increased so did the acetate quantity eventually inhibiting the degradation of the other acids, leading to low or no biogas production. The specific role of the individual VFAs in the overall anaerobic process has been the subject of debate and is still not completely understood (Ahring *et al.*, 1995). Nutrient deficiency is vital in guaranteeing optimal methane gas production. Speece investigated nutrient deficiency in cattle waste by monitoring a thermophilic anaerobic process treating cattle slurry. The volatile acids concentration became "stuck" at approximately 12,000 mg/l. Gas production was directly related to whether the reactor was fed as gas production ceased with no decrease in volatile acids noted, when feeding was suspended. However iron supplementation brought about a dramatic increase in gas production with an accompanying decrease in the VFA concentrations. Ironically, the symptoms of nutrient deficiency and toxicity are essentially identical (reduced rates of biogas production and elevated concentrations of volatile acids) but it has normally been favoured to assume that toxicity in the feed stream was responsible for high VFA (Speece, 1996).

So which of the above-explained effects caused process imbalance in the pilot scale anaerobic digesters? With toxicity the pH would have been expected to drop below neutral, thus inhibiting the methane gas production. However the pH did not decrease below neutral (enough alkalinity was produced by the slurry feed to counteract the VFA concentrations) and the gas production still declined. With nutrient deficiency in the animal slurry, the percentage methane in the biogas production only would have expected to drop like in the first stage of the study. However, the total biogas production decreased, maybe this was because the methanogens were severely inhibited by VFA levels and longer times would be required for the reactors to stabilise at these higher loading rates.

CHAPTER 6

DISCUSSIONS AND CONCLUSIONS

6.1 PLANT INVESTIGATIONS DISCUSSIONS AND CONCLUSIONS

In relation to the operation of full-scale anaerobic digestion plants, the plant operators considered temperature and pH to be the most important parameters in terms of process control, and literature support this conclusion (Malina, 1992). Sudden changes in pH or temperature could have disastrous consequences for process stability and therefore must be maintained at all costs. The findings of this research support the installation of an alarm system to indicate when pH or temperatures fall outside the desired range in conjunction with on-line probes.

pH control at the five plants, operating in Ireland, is manual and usually there is no requirement for supplementation of the feed with lime. However, pH control is cumbersome when required. An on-line pH probe on the feed line to the digester with control of a lime or NaOH dosing pump would ensure a more stable pH control. One digester gave poor process performance at start-up when operated at a low pH. Lime addition enhanced reactor performance and aided start-up.

Dry solids concentrations and biogas volumes give reliable data on process performance evaluation but are not significant in terms of providing alarms when the process is stressed.

Where laboratory methods are used, other than those outlined in APHA standard methods, occasional cross-referencing with APHA standard methods assays might provide more reliable results especially in terms of VFA and alkalinity measurement.

The installation of on-site operational stand-by equipment is recommended, including boilers, heat exchangers, pumps etc. Such provisions allow for greater flexibility and reduced down-time during maintenance.

Activity and toxicity tests, such as those outlined by Colleran *et al.*, (1992), are not carried out at any of the on site laboratories. In many cases, operation decisions were made by visual inspection of incoming feed and operator intuition. For example, reducing feed where "strong chemical smells" were noted. Such observations would be greatly aided by back up activity tests and toxicity tests. In addition, these tests would supply supplementary information on possible feeds to be used for co-digestion (gas yields, risk of inhibition etc.).

Mesophilic anaerobic digestion with storage is no longer considered sufficient for sterilisation of sludge for disposal to land. The incorporation of thermophilic pre or post treatment would provide the required sludge disinfection.

Currently there are five anaerobic digestion plants constructed for the treatment of sewage solids in Ireland. Four of these digesters, Tullamore, Greystones, Clonmel and Tralee, treat all sludge generated at the plant, and solids removal and gas production rates compare favourably with results published by other Authors (Li *et al.*, 1996). The fifth digester, Buncrana, has been decommissioned for upgrading.

6.2 PILOT TRIALS DISCUSSIONS AND CONCLUSIONS

After the full-scale anaerobic digestion plants were researched, the findings were investigated further by comparing the results with the results obtained from operating laboratory scale anaerobic digesters and an intensive literature research.

Important features of a good process indicator are its ability to detect imbalance at an early stage and its ability to reflect the metabolic state of the system directly. It is also important that the relative change of the parameter following a perturbation (process disturbance) is significant compared to background fluctuations and analysis uncertainties (Speece, 1996).

It was obvious from the results of the laboratory trials (Chapter 5) that the VFA was the best parameter to show the early signs of process imbalance. The VFA concentrations were excessively high, in all three digesters, days before the biogas production decreased. Ahring *et al.*, (1995), accepts that, for a long time it has been recognised that the VFA concentration is one of the most important parameters for the accurate control of anaerobic digestion (Chynoweth & Mah, 1971; Fischer *et al.*, 1983; Hill & Bolte, 1989; McCarty & McKinney, 1961). This study supports these findings.

Many investigations have correlated the process stability to the concentrations of individual VFA concentrations in the reactor (Hill *et al.*, 1987; Kasper & Wuhrmann, 1978; Hill & Holmberg, 1988; Varel *et al.*, 1977) and not the accumulated (absolute) VFA concentrations. Acetate concentrations higher than 13mM (~780mg/l expressed as acetic acid) have been suggested to indicate imbalance (Hill *et al.*, 1987). Propionate has been suggested by some investigations to be better indicators of process instability (Kasper & Wuhrmann, 1978, Varel

et al., 1977). Hill (1982), proposed that the propionate/acetate ratio should be below 1.4. However, from the many different levels of VFAs found in different reactor systems, it can be concluded that it is not feasible to define an absolute VFA level indicating the state of the process. Different anaerobic systems have their own "normal" levels of VFA, determined by the composition of the substrates digested or by the operating conditions (Angelidaki *et al.*, 1993, Ahring *et al.*, 1995).

Ahring *et al.*, (1995), recommends the measurement of individual VFA. Her investigations found that butyrate and its iso-form, isobutyrate provided the best indictor of process stress, detecting process imbalance within two days. The propionate/acetate ratio parameter was not a good early indicator of process stress and was not useful in detecting imbalance within two days. She concluded that VFA accumulation in an anaerobic reactor, where the concentrations of the individual VFAs usually are below 50 mM (3000mg/l expressed as acetate acid), may be considered mainly as a warning and not as a cause of imbalance. The results justify the use of individual VFAs as a process performance indicator. The concentrations of the individual VFAs after perturbation did not reach inhibitory levels. Therefore, it should be stressed that it is the relative changes of the VFAs that are used as indicators for process perturbation, not the absolute concentrations. However, the method of measuring individual VFA concentrations is very expensive and generally would not be available to sewage treatment plant operators. From the current study it can be seen that the accumulative value for VFA was a good enough indication of process upset.

The pH was not a good early process indicator in these trials as the production of alkalinity was adequate to keep the pH above neutral. Rozzi (1991), found that at low initial alkalinity in the process, the decrease in pH will be larger at the increased VFA concentrations caused

by overload. Therefore, pH is more suited to monitoring digesters fed on wastes which produce low bicarbonate alkalinities during treatment. Wastewater containing carbohydrates, VFA and lipids would not generate bicarbonate alkalinity at all when anaerobically degraded (Anderson & Yang, 1992). Bjornsson *et al.*, (1997), also found that the use of pH monitoring can only be recommended in systems where the buffering capacity is low. They concluded that it was important to make careful investigations before choosing an indicator parameter for monitoring reactor status, since the metabolic pattern of mixed cultures obviously change when the microorganisms are immobilised.

The methane composition of the biogas in all three reactors also showed process imbalance when possible nutrient deficiency was occurring, due to the composition of the whey feed (Chapter 5, Section 5.8.1). The percentage CH_4 gradually declined and the biogas production showed no sign of disturbance. In fact, the biogas production increased towards the end of the stage in all three reactors (Figure 5.8.2). In this case the VFAs were not built up enough to indicate process instability, however they were beginning to increase in all three reactors. The percentage methane detected this inhibition first. Therefore the methane composition of the biogas is an important process indictor as well as VFA, depending on the sort of process imbalance materialising. The methane yield also showed other process disturbances for example a change in substrate and oxygen shock. However, in some cases the significant level of change was small and the yield recovered after a day or two, while the VFA remained high, making an evaluation of the process based only on measurements of methane yield doubtful. Measurements of the methane production rate, was not evaluated as a possible monitoring parameter by itself, since changes could reflect the actual loading of the reactor and not only the state of the process as reported by Ahring *et al*, 1995. It would be recommended that the VFA and the methane production be monitored together. The VFA

concentrations showed the early indication of process imbalance, whereas the percentage CH_4 in the biogas observed an inhibition possibly caused by nutrient deficiency. It is important to note that when the feed composition inhibited the methane-producers in the first stage of the study there were no warning signs by other parameters except the percentage CH_4. The biogas production remained high and the VFA and pH stayed within expected ranges. Therefore, it is important that the methane composition of the biogas is measured as well as the VFA in order to observe certain shocks.

From the results presented it can be seen that the three pilot anaerobic digesters set-up, all three displayed similar operation characteristics. They possessed much the same trends in parameter measurement and mechanical problems. Reactor 3 was the mechanically stirred reactor and this reactor was not as efficient as the other two reactors in terms of biogas production, even though it was operated with the same feed and temperature as the other two reactors, the only difference being the contents were being mechanically mixed. The nature of the feed used may have contributed to the lower performance of the mixed reactor. Whey powder because of it low solids content and high soluble content is more suitable for treatment using retained biomass reactors (e.g. UASB). Biomass washout occurred for the mixed digester that did not occur for the unmixed reactors, which entrapped more solids (and perhaps attached methanogens) than the mixed counterpart.

There were four pilot scale anaerobic digesters on site. Three of them were commissioned and used in the study. The other one was used for spare parts which proved to be essential in the daily running and maintenance of the pilot plant, and would suggest the vital importance of standby equipment in full scale anaerobic digestion plants.

Anaerobic digestion of sewage solids is quickly becoming an established technology in Ireland. The purpose of this research is to ensure that the experiences and lessons gained at one plant will be of assistance to other plant designers and operators as many more anaerobic digesters are scheduled for the coming years.

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