# The Appraisal of Mesophilic Anaerobic Digestion and Thermophilic Aerobic Digestion in the Treatment of Municipal Sludge in Ireland

By Joanne Mc Guinness, BSc.



2003

MSc.

# THE APPRAISAL OF MESOPHILIC ANAEROBIC DIGESTION AND THERMOPHILIC AEROBIC DIGESTION IN THE TREATMENT OF MUNICIPAL SLUDGE IN IRELAND

BY JOANNE MC GUINNESS, BSC.

SUBMITTED TO:
THE INSTITUTE OF TECHNOLOGY SLIGO
FOR A MASTER OF SCIENCE DEGREE

**RESEARCH SUPERVISOR:**
**DR. JOHN BARTLETT, PH.D.**



SUBMITTED TO THE HIGHER EDUCATION AND TRAINING AWARDS COUNCIL, (JULY, 2003)

## DECLARATION

"I wish to state that this project is all my own work, except where acknowledgements have been made".

Joanne Mc Guinness
26-7-03

## ACKNOWLEDGEMENTS

I wish to thank Dr. John Bartlett, Head of Research and project supervisor, for his help and guidance during the course of research, and Dr. Eamonn Killilea (R.I.P), project supervisor.

Thank you to; Joe Kelly, Jim Sheridan and John Comiskey, Institute of Technology Sligo, Department of Civil Engineering, for your assistance in the day-to-day operation of the pilot plants.

Thank you to Professor Emer Colleran, Dr. Sean Pender (formally NUI Galway), and all the research students in the Department of Microbiology NUI Galway, for your assistance and hospitality while carrying out VFA analysis.

And sincere gratitude to: Dr. Enda Gibney, and Jane Lehany, Anthony Lang, and John Hession for baby-sitting the digesters at the weekends.

Many thanks to the following Local Authorities for allowing site access (where applicable) and provision of data:

- Kerry (especially Madeline Kelly in Killarney and John Galvin in Tralee)
- Leitrim (Director of Service Mr. John Mc Guinness, and Mr Terry Mc Dermot Manorhamilton treatment plant manager).
- Offaly (Tara Mitchell, for the provision of seed sludge)
- Louth (extended thanks to Mr. Tom Silkstone of Aquator Group Ltd.).
- Wicklow
- Kildare
- Dublin
- Meath
- Tipperary

Last but not least, Thank you Mum and Dad for your guidance, patience, and support (emotional and financial) over the last eight years of college, my boyfriend 'Phillip', sister 'Maria', Phelim, and lollie.

## ABSTRACT

It has been well documented that the optimum feedstock for anaerobic digesters consists of readily biodegradable compounds, as found in primary sludge or even a mixed substrate of primary and excess activated sludge. Due to the requirements of the Urban Wastewater Treatment Plant Directive of 1991, the quantities of secondary sludge generated is set to increase substantially. A pilot scale study was undertaken to evaluate the performance of both Mesophilic Anaerobic Digestion and Thermophilic Aerobic digestion in the treatment of secondary sludge. The results indicated that the anaerobic pilot scale digester achieved a greater solids destruction than the aerobic pilot plant averaging at 28 % T.S. removal verses 20% for the aerobic digester, despite the fact that secondary sludge is the optimum feedstock for aerobic digestion. This can, however, be attributed to the greater biomass yield experienced with aerobic systems, and to the absence of Autothermal conditions.

At present, the traditional technique of Mesophilic Anaerobic Digestion is in widespread application throughout Ireland, for the stabilisation of sewage sludge. There is only one Autothermal Thermophilic Aerobic Digester at present situated in Killarney, Co. Kerry.

A further objectives of the study was to compare full-scale applications of Mesophilic Anaerobic Digestion to ATAD. Two Sludge Treatment plants, situated in Co. Kerry, were used for this purpose, and were assessed mainly under the following headings; process stability, solids reduction, and economics. While both technologies achieved the same level of solids reduction on average, the ATAD plant in Killarney has the advantage of producing a 'Class A' Biosolid in terms of pathogen reduction, and can effectively treat double the quantity of sludge. In addition, economically the ATAD plant is cheaper to run, costing € 190 / t.d.s verses €211/ t.d.s. for the anaerobic digester in Tralee. An overview of additional operational Anaerobic Digestion Plants throughout Ireland is also presented.

TABLE OF CONTENTS

**LIST OF TABLES**

## LIST OF FIGURES



## LIST OF PLATES

# CHAPTER 1
# INTRODUCTION

## 1.1 "SLUDGE", A NEED FOR TREATMENT

Sewage sludge is a by-product of wastewater treatment processes. Sludge produced from municipal sewage treatment generally falls into one of two broad classifications: Primary or Secondary sludge.

Primary sludge is generated from the settlement of raw wastewater and is highly putrescible in nature, with a water content of approximately 94 – 98%. Secondary sludge is the settled product from the biological oxidation of organic matter. Secondary sludge is more stable than primary, but contains a larger volume of water, typically in the range of 97.5% - 98.5% and is therefore more difficult to dewater (Gray, 1999).

Sewage sludge is composed of mineral and organic matter, living organisms (bacteria and protozoa) and potentially pathogenic organisms or toxic compounds, which may restrict the disposal options available. Stabilisation prevents the putrefaction of the sludge during storage, destroys pathogens, causes partial destruction of volatile solids, and reduces the overall volume of sludge for disposal. There are numerous stabilisation techniques classified as: Biological, Chemical and Thermal. The most sustainable options, however, are those which will allow the re-use of the sludge in agriculture, as sewage sludge is a valuable commodity, rich in nutrients (Cheremisinoff, 2002).



## 1.2 THE EVOLUTION OF SLUDGE MANAGEMENT AND LEGISLATION IN IRELAND

Over the past decade, environmental regulations in relation to the treatment, disposal and reuse of Biosolids in Ireland have evolved at a rapid pace, for a number of reasons:

**(1) Due to stricter controls on the quality and quantity of sludge that can be applied to agricultural land.**

Up until 1991, the use of sewage sludge in agriculture was uncontrolled. It was in this year that EC directive 86/278/EEC (*On the protection of the environment and in particular the soil when sewage sludge is used in agriculture)* was brought into Irish legislation under SI 183 of 1991. This Directive regulates the quantity and quality of sewage sludge that can be applied to agricultural land, based on heavy metal

content, micro-pollutants and the amount of nutrients that can be safely applied to the soil.

**(2) A increase in the volume of sludge for treatment.**

EC Directive 91/271/EEC (Concerning Urban Wastewater Treatment) was brought into Irish legislation, under SI 419 of 1994. The Directive stipulates that secondary treatment should be facilitated in all towns with a p.e. of greater than 2000, by the year 2015, and also banned the dumping of sewage sludge at sea after 1998, increasing the overall quantity of sludge requiring treatment. Because this directive would have major implications in the future of sewage sludge generation and disposal in Ireland, the DoELG commissioned a study to determine the quantities, types, and best means of treatment for Biosolids ("Strategy Study on Options For the Treatment / Disposal of Sewage Sludge in Ireland") in 1992. This report concluded that, in 1993, approx. 37,500 tonnes of dry solids (tds) was produced in Ireland and estimated that this figure would increase to 129,795 tds by 2013, in - line with the implementation of the Urban Wastewater Treatment Directive (Bartlett and Killilea, 2001). To facilitate the management of such large volumes of sludge, the country was sub-divided into regions, proposed as sludge treatment centres or Hub Centres. Recommendations were made, based on available information, as to the sludge treatment technology most suitable for each Hub Centre.

**(3) The elimination of traditional disposal sites (landfill, and sea dumping).**

In addition to the ban on disposal of wastewater to the sea imposed by the implementation of the Urban Wastewater Treatment Plant Directive of 1991, the Dumping at Sea Act of 1996 (no. 14 of 1999) gave effect to the OSPAR convention of 1992, eliminating dumping of sludge at sea by 1999 (Bartlett, and Killilea, 2001). The disposal of sludge to landfill ceased following the introduction of The Waste Management Act of 1996 (no. 10 of 1996) and European Council Directive 1999/31/EC (*on the landfill of waste*). Furthermore, under The Waste Management (Planning) Regulations, 1997, S.I 137 of 1997, Local Authorities were required to quantify sludge production arising in their functional area, by means of a Waste Management Plan. The Waste Management Plan must include, options for the treatment, reuse and disposal of Biosolids.

These legislative regulations prompted the publication of a guidance document, for the use of Biosolids in agriculture in 1999 (*Code of Good Practice for the Use of Biosolids in Agriculture*). This is a mandatory code which provides advice to Biosolids producers in relation to:

- The treatment of Biosolids to produce a pasteurised product.
- Suitability of land, for Biosolids re-use.
- Transportation and spreading of Biosolids
- Nutrient management planning
- Quality control
- Liasing with the Customers

To achieve a pasteurised product, the 'Code' specifies the following treatment alternatives:

- Mesophilic anaerobic digestion with pre/post sanitation (Mean retention time of 12 days @ 35 °C, plus retention at >70 °C for 1 hr).
- Thermophilic anaerobic digestion (Mean retention time of 2 days @ > 55 °C or 4 days @ 50 °C).
- Thermophilic aerobic digestion.
- Composting (windrows or aerated piles).
- Alkaline stabilisation.
- Thermal drying.

The final product is considered to be pasteurised when a microbial standard of faecal coliforms <1,000 MPN per g of dry solids and Salmonella sp, < 3 MPN per 4g of dry solids, is attained.

This desired level of pathogen destruction is reflective of the U.S. EPA rule 503 regulations of 1994. Unlike the Code of Good Practice, compliance with the US EPA rule 503 is compulsory in the United States. In this case, Biosolids are classified into three categories, depending on degree of stabilisation achieved, which determines the final reuse or disposal route. In terms of pathogen reduction, a Class

'A' (Density of Faecal Coliforms in the biosolids must be less than 1,000 MPN per g T.S. or Salmonella sp. of less than 3 MPN per 4g T.S.) can be applied to agricultural land without restriction, Class 'B' (Faecal coliform content of less than 2,000,000 MPN per g T.S.) can be applied to land with restriction (e.g. soil injected) and, Class 'C' can not be spread on land, and is highly putrescible in nature.

### 1.3 Aims and Objectives of this Study

This study is concerned with two biological sludge stabilisation methods, Mesophilic Anaerobic Digestion and Thermophilic Aerobic Digestion. The aims of this project are as follows:

- To review the current state-of-the-art Mesophilic Anaerobic Digestion as a traditional sludge stabilisation technique, and the newer technology of Thermophilic Aerobic Digestion in Ireland.

- To Compare Mesophilic Anaerobic Digestion to Autothermal Thermophilic Aerobic Digestion, based on full-scale operations in County Kerry, (in terms of process stability, solids destruction and economics).

- To evaluate pilot plant performance in the treatment of secondary solids.

# CHAPTER 2
# A LITERATURE REVIEW OF ANAEROBIC DIGESTION

## 2.1 Definition of Anaerobic Digestion

Anaerobic digestion is a fermentative process involving the solubilisation and reduction of complex organic molecules, by a synergistic group of microorganisms, in the absence of oxygen. Anaerobic digestion produces stabilised Biosolids, reduces biomass quantity by partial destruction of volatile solids, and produces a useable gas as a by- product.

The composition of this biogas varies, and depends to a large extent on the nature of the feedstock, organic loading rate, the hydraulic retention time and the temperature of anaerobic decomposition. However, typically, biogas is composed of the following:

- Methane ($CH_4$) 55-56%
- Carbon dioxide ($CO_2$) 35 – 45%
- Nitrogen ($N_2$) 0-3%
- Hydrogen ($H_2$) 0-3%
- Hydrogen sulphide ($H_2S$) 0- 1% (Polprasert, 1996).



## 2.2 Microbiology of the Process

Anaerobic digestion occurs in a series of four interdependent stages, hydrolysis, acidogenesis, acetogenesis, and methanogenesis.

### A. Hydrolysis

Hydrolytic bacteria are a group of heterotrophic organisms, typically strict and facultatively anaerobic in nature. The hydrolytic bacteria are responsible for the solubilisation of complex organic macromolecules, such as protein, cellulose, lignin and lipids, into smaller monomers (glucose, amino acids, fatty acids and glycerol), which can be directly utilised by the next group of organisms (acidogenic bacteria) (Bitton, 1994). Hydrolysis is achieved by extracellular enzymatic degradation of the substrate, and has been identified as the rate limiting stage of anaerobic digestion (Tiehm *et al.*, 2001).

**B. Acidogenesis**

The acidogenic bacteria, are mainly obligate anaerobes, which convert sugars, amino acids and fatty acids to organic acids (volatile fatty acids, alcohols and ketones, (ethanol, methanol, glycerol, acetone), acetate, $CO_2$ and $H_2$. The resulting products of acidogenesis are ultimately determined by the types of bacteria present (Bitton, 1994).

**C. Acetogenesis**

The acetogenic population includes a number of bacterial groups capable of acetate production. The first group being sulfate reducing bacteria (SRB) (non obligate proton reducing bacteria), which produce acetate and hydrogen in a mixed culture (Marty, 1984). Numerous studies have focused attention on competition between SRB and methanogens in the anaerobic digester environment (Fukui *et al.,* 2001). However, a compatible relationship can exist in nutritionally rich environments, such as sewage sludge and animal wastes (Ueki *et al.*, 1992).

The second group is the obligate hydrogen producing acetogenic bacteria, which forms an important symbiotic relationship with the methanogenic population.

The third group is the homoacetogenic bacteria which produces acetate from fructose, but are also capable of oxidising hydrogen and reducing $CO_2$ into acetic acid, acting as competitors to methanogens or as synotrophic donors of acetate, hydrogen and $CO_2$ (Marty, 1984; Kotsyurbenko *et al.*, 2001).

**D. Methanogenesis**

Methanogens differ significantly from other bacteria. The most significant characteristic of the methanogen group is their ability to generate methane under strict anaerobic conditions (Lange and Adring, 2001).

In the anaerobic digester environment, there are three substrates which are important to the methanogenic population; acetic acid (which yields approx. 70% of the methane produced), hydrogen and $CO_2$ from which the remaining methane is formed. The

bacteria responsible for methane production can be sub-divided into the groups based on their substrate preference:

- Acetoclastic methanogens
- Hydrogen utilising methanogens (Polprasert, 1996).

The acetoclastic bacteria convert acetate to methane and $CO_2$: ($CH_3 COOH \longrightarrow CH_4 + CO_2$). Acetoclastic bacteria have a very slow generation time ($\mu$ max $\approx$0.04 hr) in comparison to the acidogenic population ($\mu$ max $\approx$1 hr) (Bitton, 1994). This is the reason that, during process instability, volatile fatty acids (VFA) may accumulate, which inhibits the activity of subsequent synergistic bacteria, resulting in pH decline. The pH does not affect the acidogenic bacteria themselves, however, until it drops below 4.5 (Marty, 1984).

The hydrogen utilising methanogens convert hydrogen and $CO_2$ to methane as follows: $CO_2 + 4H_2 \longrightarrow CH_4 + 2H_20$. In doing this, hydrogen partial pressure within the reactor is kept low, aiding in the conversion of VFA and alcohols to acetate, by the acetogenic bacteria (Bitton, 1994).

## 2.3 The Development of Anaerobic Digestion Technology

Anaerobic digestion of municipal sludges is not a new technology, having been in use since the end of the 19th century. The earliest applications of the technique were for the treatment of domestic sewage in Septic tanks and Imhoff tanks, and, over 100 years later, the technique is still in widespread use for the treatment of sewage sludge from urban wastewater treatment plants (Braber, 1995).

There are generally three types of reactors used for the anaerobic stabilisation of sewage sludge, due to the high solids content of the feedstock. These are:

- The standard rate digester
- The high rate digester

- The two stage digester

**A. Standard Rate Digester**

In standard rate digestion, sludge is fed intermittently to the digester and remains unmixed and unheated. As a result, four layers develop within the tank.

(i) Scum layer, which forms due to the flotation of light materials like fats, oils and grease carried to the surface by rising gas bubbles.

(ii) Supernatant layer

(iii) Actively digesting solids

(iv) Digested inert solids.

The supernatant and digested sludge are withdrawn by gravity to an overflow weir during feed administration. In the standard rate, or low rate digestion system as it is often called, the digestion period is long, ranging from a hydraulic retention time (HRT) of 30 – 60 days. Biogas generated during the process exits the reactor vessel through the top of the tank, where it is collected and used for electricity generation (Quasim, 1999).

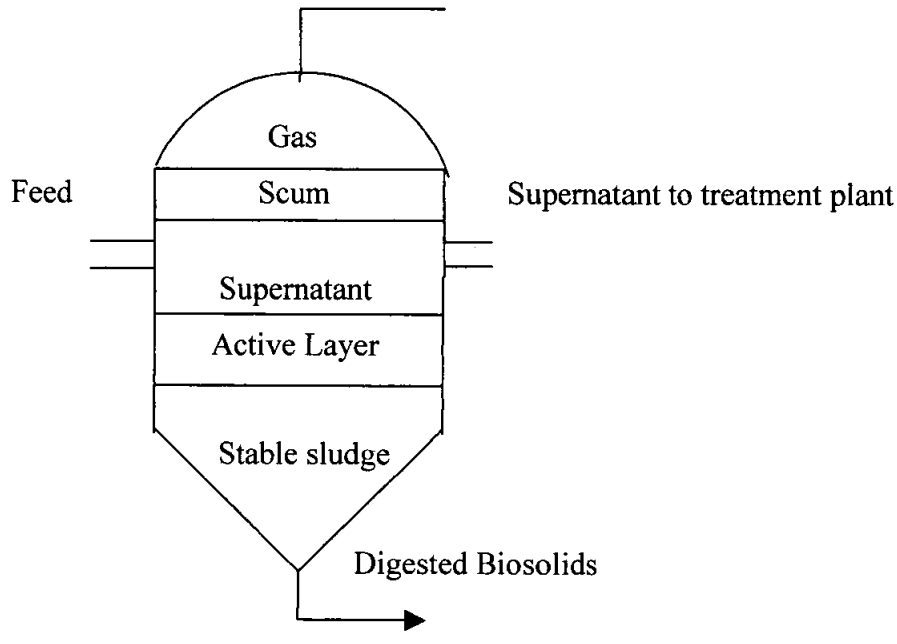



Figure: 2.1 The Standard Rate Digester

**B. The High Rate Digester / Continuously Stirred Tank Reactor (CSTR)**

In general, the reactor configuration in a high rate digester is quite similar to that of the standard rate digester. There are a number of modifications, which facilitate the application of a greater organic loading rate (OLR) and solids loading rate (SLR). These include the addition of heat by internal or external heat exchangers, and mixing by gas injection, recirculation or mechanical means. Typically, anaerobic digesters are operated in the mesophilic temperature range, 35°C, although thermophilic operation is set to increase in line with the stringent requirements of legislation. The provision of these key elements improves digestion efficiency and reduces HRT to 10 – 20 days (Lusk, 1999; Oleszkiewicz and Mavinic, 2002).

Sludge feed can be administered by batch or on a continuous mode. Continuous feeding is traditionally favoured to maintain constant conditions in the reactor (Tchobanoglous and Barton, 1991). However, Azbar *et al.,* (2001) contradicts this traditional feeding regime, in favour of the batch fed system, following pilot trials using a number of digestion systems. In their study, a batch fed single stage CSTR produced a higher quality effluent than a continuously fed single stage CSTR during the treatment of a range of test substrates. As with the standard rate system, gas in the high rate system exists the reactor through the roof of the tank, to the collection tank.

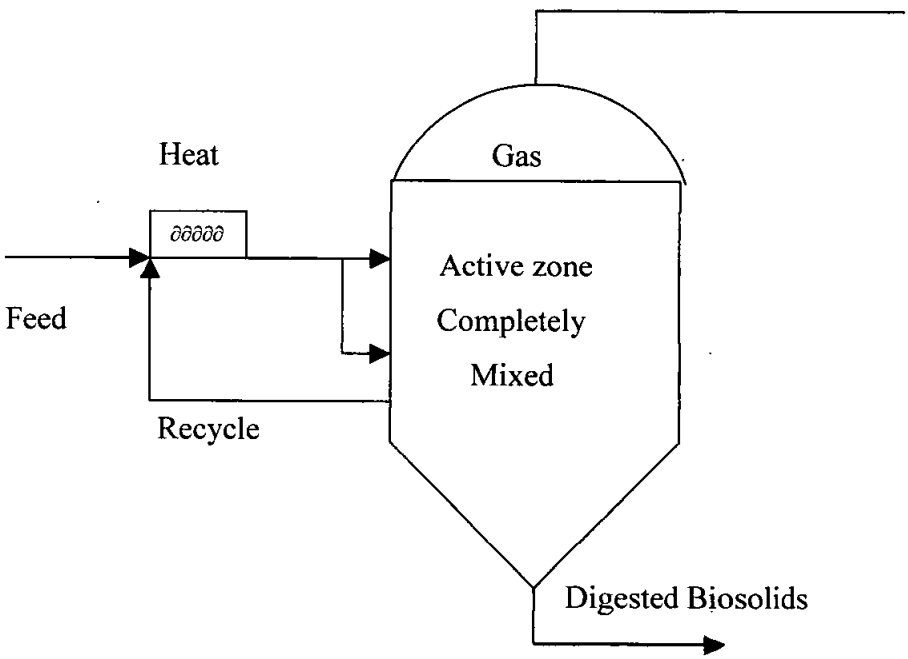



Figure: 2.2 The High Rate Digester

**C. Two - Stage Digester**

In this process, two digesters are operated in sequence. The first tank is a CSTR, in which active decomposition occurs, while the second tank is unmixed and receives the hydrolysed sludge from the first reactor. The second tank is used mainly for storage and supernatant formation. Both tanks are identical in design, to facilitate greater flexibility in operation. Gas is produced in both reactors and therefore gas collection must also be provided in each instance. In essence, the two-stage system consists of the high rate digester CSTR followed by the standard rate digester (Hammer, 1998).

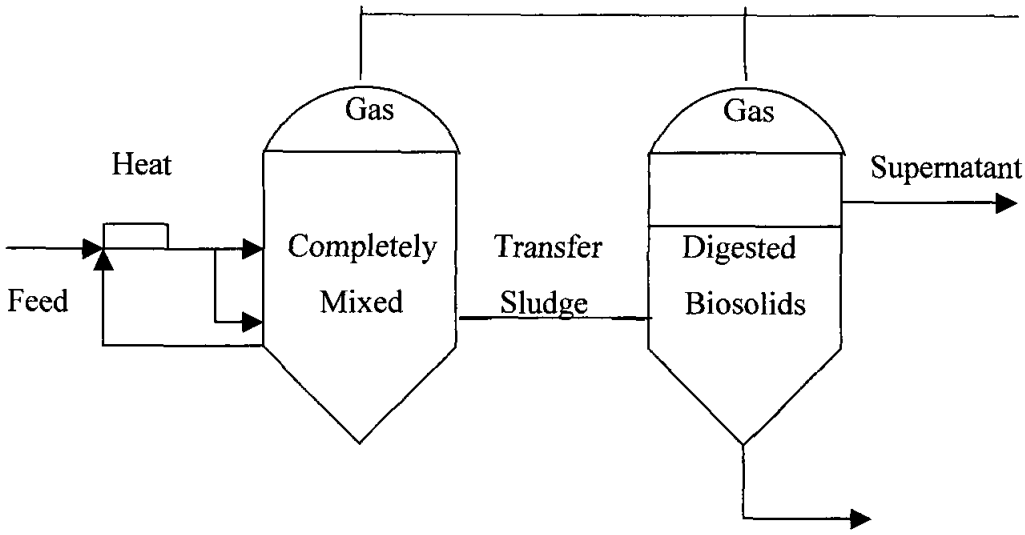



Figure: 2.3 The Two Stage Digester

## 2.4 Parameters for Optimum Digestion

### 2.4.1 Temperature

Bacteria are classified by three ranges of temperature at which they can grow. Generally, different strains of bacteria grow in the different temperature ranges. The ranges for growth are:

- Below 0 °C to 15 °C (Psychrophilic)
- From 15 °C to 45 °C (Mesophilic)
- From 45 °C. to 70 °C (Thermophilic)

The rate of digestion is temperature dependant. In the temperature range below 45 °C digestion becomes slower as temperature decreases, and mesophilic digestion virtually ceases at about 15 °C. However, the bacteria are not inactivated and digestion will take place if the bacteria are allowed sufficient time to acclimatise to environmental conditions. Above 50 °C, in the thermophilic temperature range (65 °C to 70 °C), digestion takes place at an increased rate, as substrate utilisation is increased. To date, the majority of research has focused on mesophilic digestion. It is accepted, however, that the bacteria concerned in thermophilic digestion are different species than those of a mesophilic digester. In general, the optimum temperature for mesophilic digestion is about 35 °C and for thermophilic digestion 55 to 60 °C (Hobson and Wheatley, 1993).

### 2.4.2 Feedstock Characteristics

Typically, anaerobic sewage sludge digesters are fed a mixed feedstock of primary and secondary sludges. Primary sludge contains faecal particles, toilet papers, cotton wool materials, and street debris. The majority of such paper materials are treated during manufacture to make then 'soluble', therefore the cellulose within these materials will be easily degraded in the digester (Hobson and Wheatley, 1993). Primary sludge is predominantly composed of biodegradable materials, thus a high proportion of the volatile suspended solids can be converted into biogas (55% – 65%) at mesophilic temperatures (De Souza Araujo *et al.*, 1998).

Secondary sludges, are composed mainly of finer faecal matter and microbial cells. Microbial cells from the secondary treatment stage contain starch-like storage polysaccharides, which form different cell structures. Mucopolysaccharides, from gut secretions are complex polysaccharides, which are believed to be non-degradable by anaerobic digestion, even if liberated by death and lysis of faecal bacteria or intestinal cells. As a result, it is considered that they do not contribute greatly to the overall biogas production (Hobson and Wheatley, 1993).

It was proposed by De Souza Araujo *et al.,* (1998), that there was a direct relationship between the composition of the organic materials of waste activated sludge (WAS) and the extent of biodegradability in an anaerobic digester. The volatile fraction of WAS was considered to be composed of an active fraction, formed by live material, and an inactive fraction which does not exhibit metabolic activity in an aerobic environment. The study involved the generation of five different sludges with different sludge ages, having active (live) sludge fractions from 16 to 76% of the volatile sludge concentration (calculated from the measured specific oxygen utilisation rates). The five different sludges were then fed to five corresponding anaerobic continuously stirred tank reactors, operating at a 20 day HRT. This study confirmed that, while a typical WAS may contain 65 – 70% Volatile solids (V.S.), not all of this matter is readily biodegradable, as digestion efficiency was 53% for the active sludge (live fraction 76% of the V.S.) and just 15% for the inactive sludge (live fraction 16% of the V.S.). Therefore, the success of anaerobic digestion depends largely on the sludge age of the feedstock.

### 2.4.3 Pre-treatment of Secondary Solids

While all organic compounds are biodegradable, the initial stages of anaerobic digestion are rate-limiting i.e. hydrolysis and solubilisation. However, there are a range of pre-treatment applications that can significantly increase the rate of biodegradation by physically, chemically or mechanically altering the microbial cell (cell lysis).

**A. Ozone**

Ozone is a very reactive oxidising agent. It reacts with the sludge compounds through two pathways, which occur simultaneously:

- Direct attack by electrophilic addition
- Indirect attack by free radicals produced by reaction with water and water constituents

The oxidation of sludge with ozone destroys microbial cell walls, dissolving cell cytoplasm in the sludge water. It splits water insoluble substances with high molecular weights into smaller water soluble and biodegradable fragments, and alters the sludge matrix (protein, lipid, polysaccharide) composition increasing the degree of biodegradability (Scheminski *et al.,* 2000; Weemaes *et al.,* 2000)

**B. Ultrasound**

Ultrasound is a pressure wave, when applied to sludge at a frequency of 20 – 40 kHz, the ultrasound travels through the medium releasing a vast amount of energy. This release of energy leads to the production of gas and vapour bubbles at high velocities, which grow and collapse rampantly (known as acoustic cavitation). The temperature and pressure achieved inside the collapsing bubbles leads to the generation of very reactive hydroxyl radicals (Wang *et al.,* 1999; Chu *et al.,* 2001; Tiehm *et al.,* 2001; Onyeche *et al.,* 2002).

Tiehm *et al.* (1997), concluded that, following ultrasonic treatment of sewage sludge, the COD in the aqueous phase increased (as organic substances from the sludge solids transferred to the aqueous phase). Particle size reduction was observed which corresponded to COD increase in aqueous phase, and the organic compounds transferred to aqueous phase were readily biodegradable, demonstrated by an increase in biogas production and volatile solids reduction. In addition, since there is a increase in temperature attained during the process, the microbial density of the sludge may also be affected. Insufficient investigations have been conducted in this area however, to confirm this suggestion ( Lafitte –Trouque and Forster, 2002).

**C. Mechanical Disintegration**

Mechanical disintegration is a method of physically disrupting the microbial cell wall barrier, releasing intracellular material for further biological metabolism. Two commonly used methods of mechanically disintegrating sludge is the ball mill (efficiency depends on a number of factors, such as; ball diameter, revolution speed, ball material and sludge concentration and quality) and cutting mill. Mechanical disintegration of waste sludge significantly enhances the soluble fraction of total COD (Baier and Schmidheiny, 1997; Wook Nah *et al.*, 2000).

**D. Thermal Hydrolysis**

Thermal Hydrolysis at high temperature and pressure (typically 130 $^{o}$C - 180 $^{o}$C for 30 mins at corresponding vapour pressure) alters microbial cells by breaking down the cell membrane releasing COD bound in the sludge, to the water phase as dissolved COD. Studies on thermal hydrolysis have highlighted many advantages of its use prior anaerobic digestion such as:

- Increase in gas yield
- Application of higher concentration of sludge to digesters, since the viscosity of the sludge is altered (i.e. sludge with 12% DS can be handled as a sludge of 5 –6 % DS)
- Improved dewatering after digestion
- Class 'A' Biosolid (in terms of pathogen removal) which can be spread directly to agricultural land (Weisz *et al.*, 2001; Oorshot, 2003; Panter, 2003; Sorensen, 2003).

**2.4.4 Potentially Toxic Compounds**

Hydrogen ion concentration, and several other compounds such as heavy metals and chloro-organic compounds, affect the rate of anaerobic digestion, even at very low concentrations. 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. Sulphide can be formed in the process due to the reduction of sulphate. Research indicates that the expected sulphide concentration in anaerobic sewage treatment systems may up to 50 mg/l, which is far lower than the minimum concentration for noticeable toxicity. Therefore, sulphide toxicity will normally not be a problem (Van Haandel and Lettinga, 1994).

One group of compounds that form an important part of sludge preparation are polyelectrolytes, yet little information exists on the effects of polyelectrolytes on subsequent sludge treatment unit operations, beyond that of dewatering. In a recent study by Chang *et al.*, 2001, the biodegradability of a flocculent polymer was investigated. To determine aerobic biodegradability, oxygen consumption was measured, while anaerobic biodegradability was measured by gas production, following inoculation of the cultures with the polymer at varying doses. It was concluded that the test polymer (Zetag 787) was subject to partial degradation by both aerobic and anaerobic cultures, as the pedant-groups of the polymer were hydrolysed and partially degraded. The removed portion was completely degraded anaerobically, but not aerobically, while the remaining acrylamide or acrylate polymer chain was not significantly degraded. Not all polyelectrolytes degrade in the same manner, however. In another study conducted by Uyanik *et al.* (2001), a range of polyelectrolytes was examined to determine the degree of inhibition imposed on anaerobic digestion. Specific methane activity (SMA) tests demonstrated that different types of polyelectrolytes caused various degrees of inhibition of anaerobic seed, (see Appendix A).

### 2.4.5 Mixing

Although some natural mixing occurs in an anaerobic digester, because of rising sludge gas bubbles and the thermal convection currents caused by the addition of heat, this level of mixing is not adequate to ensure stable process performance at high loading rates. It is therefore necessary that a mixing system is installed to create an homogeneous environment throughout the reactor.

There are a number of techniques to ensure adequate mixing within the digester, such as gas injection, mechanical mixing, and mechanical pumping (Quasim, 1999).

1. Gas Injection: there are two main types of gas injectors, described as unconfined and confined. In the unconfined system, gas is collected at the top of the reactor, compressed and then discharged through diffusers situated at the bottom of the reactor or through a series of radially placed top –mounted lances. Mixing is achieved by the releases of gas bubbles which rise to the surface, carrying and moving the sludge. In the confined system, while gas is again collected at the top of the reactor and compressed, it is then discharged in the digester through confined tubes. There are two main types of confined system, depending on the type of gas distribution system required, The Gas Lifter System and The Gas Piston System. Both the unconfined and confined gas injection systems are suitable for use in digesters with fixed, floating and gas holder covers.

2. Mechanical mixing: such mixers are usually installed into a shaft tube to promote vertical mixing. It is necessary to ensure preliminary treatment is in place to prevent fouling of propellers by rags and other transient materials. Low speed turbines are usually employed for mixing. Mechanical mixers are suitable for use in digesters with fixed or floating covers.

3. External pumped circulation: Sludge is pumped out of the digester and then returned by recirculation. Sludge is usually removed mid way in the digester and

can be returned to the base or surface of the reactor to break-up scum. This method requires high-energy input.

Numerous studies have been conducted on the importance of mixing in terms of digester performance, many of which having contradictory outcomes. For example, the importance of mixing was illustrated by Casey (1984), who outlined that inadequate mixing caused two main problems within the reactor:

a) Floatation of solids, due to biogas bubbles growing on the surface of the digesters solids which can cause buoyancy forces and thus solids form a surface floatation layer.
b) Settlement of heavier solids forming a bottom deposit layer.

Floatation is deemed to be the most detrimental to the digestion process, since the particles which float are centres of colonisation for methanogens. Casey also noted that significant mixing can occur in a full scale digester naturally without supplementation, due to biogas production and convective currents caused by heating, primarily due to the velocity of gas bubbles as they rise to the top of the reactor, thus mixing intensity is design dependent. Pilot scale digesters do not have the height required to ensure adequate natural mixing and thus inadequate mixing is apparent in such cases. However, Stroot *et al.*, (2001) also noticed contradictory information on the topic of digester mixing, and so conducted a pilot study to determine the implications of variations in mixing regime for the stability of mesophilic anaerobic digesters. The test substrates included: the organic fraction of municipal solid waste, primary sludge, and waste activated sludge (WAS). Six digesters were operated to compare performance under continuous mixing and reduced mixing levels at various loading rates and solids levels. The main outcome of this research was that continuous mixed digesters exhibited unstable performance at higher loading rates (propionate accumulation), while the minimally mixed digesters performed well for all loading rates evaluated. Results also showed that an unstable continuously mixed digester was quickly stabilised by reducing the mixing level.

### 2.4.6 Adequate Metabolism Time

Kiely (1997), defined hydraulic retention time (HRT) as: $\frac{\text{Working volume of reactor( L)}}{\text{Rate of sludge removal (L / D)}}$

And solids retention time (SRT) as: $\frac{\text{Mass of solids in reactor (kg)}}{\text{Rate of solids removed (kg/d)}}$

Both parameters are of critical importance to the anaerobic digestion process, as the retention time must be higher than the generation time of the slowest growing microorganism in the system, to prevent wash-out of the active biomass, and must also be long enough to achieve the required degree of volatile solids destruction (Dohanyos and Zabranska, 2001). The methanogens have the slowest generation time, when compared to the other members of the anaerobic digester microbial consortia, ranging from less than two days to more than 20 days at a mesophilic temperature (Malina and Pohland, 1992).

In a CSTR without recycle, the HRT and SRT are equal, ranging from 15 – 30 days typically (Tchobanoglous *et al.*, 2001). At the higher range of retention time, a safety factor against wash-out is permitted and maximum contact time is allowed, facilitating fermentation of slow digesting polysaccharide feedstocks.

## 2.5 Monitoring and Control of the Digester Environment

### 2.5.1 Sensors for Anaerobic Digestion

Anaerobic digestion process stability is considered to lag far behind that of its aerobic digestion (Feitkenhauer, 2001). Process instability may arise due to organic overloading, hydraulic overloading, toxins in the feedstock, and temperature fluctuations, which may result in the accumulation of intermediate by-products (VFA), leading to environmental changes within the digester and a shift in microbial populations. Considering the fact that methanogens have the strictest environmental and

nutritional requirements, such instability often leads to a decline in numbers. With inadequate process control, souring of the digester is inevitable. Fortunately, due to the complex nature of the process, which involves a series of interdependent microbial stages of degradation, and due to a good understanding of the process biology; instability, should it occur, can be foreseen by regular analysis of process intermediates and by-products, allowing remedial action to be taken. The analytical parameters chosen may be indirect indicators, such as the concentration of a metabolite in the digester, or a direct status indicator, such as the number of active microorganisms in the system. The most common parameters used to indicate process stability are indirect, including pH, alkalinity, VFA, biogas production and methane yield (Bjornsson *et al.*, 2001; Michaud *et al.*, 2001). However, a number of other parameters are used to evaluate overall removal efficiency, mainly solids analysis and chemical oxygen demand.

### 2.5.2 pH

pH represents the acidic nature of a liquid or the total concentration of hydrogen ions within the liquid (Fifield & Hains, 2000). It is used extensively as a monitoring tool of the anaerobic digestion process, and was considered the most important parameter by Irish plant operators in a survey conducted in 2000 (Scahill, 2000). Typically, anaerobic digesters operate at a narrow pH of 6.8 – 7.2. This is indicative of proper balance between the material entering and that discharged from the digester (W.P.C.F, 1987). Should the pH decrease below 6.0, the methanogenic bacteria are inhibited, characterised by a decline in methane production and VFA accumulation within the digester (Dohanyos and Zabranska, 2001).

### 2.5.3 Alkalinity

The alkalinity of a liquid is a measure of its capacity to neutralise acids. There are three major classes of materials that contribute to the alkalinity of a liquid: hydroxide, carbonate, and bicarbonate. In anaerobic environments salts of weak acids such as acetic, propionic, and hydrogen sulphide also contribute to the total alkalinity of the liquid (Sayer, 1999).

Alkalinity within the digester is of major importance during the acid phase of the anaerobic digestion process, to ensure adequate neutralisation or buffering of intermediate acids, ensuring optimum conditions for methanogenic bacteria. Typically, anaerobic digesters may have an alkalinity of 2000 – 6000mg/l as $CaCO_3$. However, should alkalinity fall below this level, a number of steps can be taken to alleviate imbalance. The first is to cease feeding until the pH increases to 6.8 (approx.), this action allows the methanogenic population time to consume the backlog of VFA. Secondly, chemicals may be added to increase the alkalinity, and thus the pH, within the digester artificially. Chemicals such as bicarbonates (which add bicarbonate alkalinity directly) or carbonate salts (which trap $CO_2$ from the gas and convert it to bicarbonate) are utilised. The addition of chemicals in the bicarbonate form is preferred, as precise additions can be achieved, unlike the carbonate salts, which must be added in small steps (to prevent the accumulation of insoluble calcium salts) to allow time for gas equilibrium to occur between each addition (Dohanyos and Zabranska, 2001).

### 2.5.4 Volatile Fatty Acids (VFA)

VFA such as: propionate, butyrate, iosobutyrate and acetate, are formed as intermediates during the anaerobic degradation of carbohydrates, proteins, and fats. Excess VFA can be inhibitory to the digestion process and must be managed. Typically, VFA concentrations ranging from 50 – 250 mg/l as acetic acid indicates a satisfactory balance between the methanogenic and acidogenic bacteria. However, should inhibition of methanogens occur, due to operational or environmental changes, a decrease in the rate of VFA destruction may occur, leading to accumulation within the system and a corresponding reduction in pH. Under conditions of imbalance, VFA's may reach concentrations of 2000 – 6000 mg/l and will not decrease until a neutralisation agent is added to increase pH to the required level (Sayer, 1999).

In terms of methanogenic bacteria, metabolism of short chain fatty acids are of vital importance. Acetate, for example, yields to approx. 75% of the methane produced

during digestion. Propionate and butyrate are important VFA's also, not only for the reason that they may be further converted into acetate and hydrogen, ultimately yielding methane, but also because the accumulation of these intermediate acids in the undissociated form retards the growth of several microbial species, which may cause a subsequent decrease in methane production (Aguilar, 1995).

Propionic acid in particular has been proposed as a valuable indicator of process performance. It has been noted to accumulate within the digester when the process has been subjected to shock loading, overloading, or during start-up (Gujer, 1983). A study conducted by Inanc *et al.,* (1996), proposed that propionic acid accumulation during process instability is due to shifts in the acidogenic bacterial population and end product distribution. Following pilot anaerobic digestion trials, microscopic examination of anaerobic biomass showed the bacteria present to be gram-positive rods, during periods of butyric acid accumulation, while during periods of propionic acid accumulation the bacteria were gram-negative rods. The study also suggested that the two- phase anaerobic digester configuration, where the acidogenic reactor is operated at a pH of 5 or less, could prevent propionic acid accumulation, as the propionic acid producing bacterial species were inhibited during pilot trials at pH 5.

### 2.5.5 COD (Chemical Oxygen Demand) and its Equivalence of Methane

COD is a measure of the amount of oxygen required to chemically oxidise organic matter in a sample. The amount of oxygen is measured directly in mg/l as the oxygen equivalent using a strong chemical oxidant. COD is analysed in 2 separate components, Total COD, which includes soluble and colloidal matter and Soluble COD consisting of the soluble fraction of anaerobic effluents which contains residual degradable and non or slowly degradable influent substrate, and intermediates products, such as; VFA, (Barker *et al.*, 1999). Soluble COD in itself is an important parameter, as it discloses information in regard to the extent of hydrolysis and solubilisation carried out by the acidogenic bacteria. (Maharaj and Elefsiniotis, 2001)

The quantity of methane produced per gram of COD removed can be easily determined for mass balance estimations. The COD equivalent of methane is:

$$CH_4 + 2\ O_2 \longrightarrow CO_2 + 2H_2O$$

From the above equation, it can be determined that for each mole of methane consumed (22.4 L @ 0 °C), two moles of oxygen equivalent are destroyed (64g). Therefore, 0.35 L (22.4 L / 64g) of $CH_4$ at 0 °C and 760 mm HG pressure (STP) is equivalent to 1 g COD destruction. At a mesophilic temperature however, the $CH_4$ equivalence is 0.395 L at 35 °C and one atmosphere {5.6 ft. 3@ 0 °C and 6.3 ft. 3@ 35 °C of methane is produced / pound COD destroyed} (Speece, 1996; Michaud *et al.*, 2001; Tchobanoglous *et al.*, 2001).

### 2.5.6 Solids

Total solids refers to the material residue left in a vessel following evaporation and drying of the sample in an oven at the designated temperature (APHA *et al.*, 1995).

The solids portion is further classified by ignition of the sample at 550 °C +/- 50 °C. The ignition process causes oxidation of the organic fraction of the solid matter, and is thus driven off by the extreme temperature (Volatile Solids). The ash remaining in the vessel represents inorganic matter or Fixed Solids (Tchobanoglous and Barton, 1991).
The extent of digestion is often measured by volatile solids reduction. Anaerobic digestion can achieve Volatile Solids destruction of 40 – 60% and an overall destruction in total sludge volume of 25 – 30 % (approx.) (Quasim, 1999).

### 2.5.7 Biogas Quantity and Quality

Digester gas analysis can provide valuable data on the process efficiency. Biogas composition depends largely on the raw material, organic loading and, time and temperature of decomposition (Polprasert, 1996).

Monitoring of the biogas quality can be used as a measure of digestion efficiency, as regular monitoring conveys deviations from typically obtained values of individual gas components, allowing the operator to take remedial action as soon as unstable conditions are noticed.

# CHAPTER 3
# A LITERATURE REVIEW OF AUTOTHERMAL THERMOPHILIC AEROBIC DIGESTION

## 3.1 History of ATAD

The ATAD process was first conceived in 1968 by Mr. Hurbert K. E. Fuchs, Sr., when he observed autothermal conditions during the aeration of agricultural manure. The first full-scale ATAD plant was commissioned in Vilsbiburg, Germany in 1977. Initially, the plants consisted of two-stage circular open topped tanks, later developing to totally enclosed and insulated vessels (Kelly, 1993). The technology has since developed in Europe, with in excess of 50 plants, mainly operating in Germany, Austria, Switzerland, Great Britain (Schwinning, 1993; Schwinning, 1996; Skjelhaugen, 1999). In the late 80's, ATAD was utilised in the US, following the recognition from the US EPA as a 'Technology to Significantly Reduce Pathogens' and allow unrestricted uses of biosolids on agricultural land (provided other legislative requirement have been met, including metals). In the year 2000, it was estimated that there were 35 ATAD systems operating in North America (Tchobanoglous *et al.*, 2001).

## 3.2 Biology of Thermophilic Aerobic Digestion

In aerobic digestion, organic matter is metabolised by the active thermophilic aerobic microbial consortia to form carbon dioxide, water, ammonia, and new microbial cells, In general, chemical changes brought about in an ATAD digester can be described by the following equation.

$$C_5H_7NO_2+5O_2 \longrightarrow 5CO_2+2H_2O+NH_3 + \text{energy}$$

(W.E.F., 1995)

There is little information regarding the microbial diversity of thermophilic aerobic reactors. Their biology differs from conventional activated sludge microflora in that nitrifying bacteria, floc-forming organisms, or protazoa and other life forms are not present. The reasons thermophilic bacteria fail to form a floc is unknown, however, contributing factors may include:

(1) lack of floc-forming species e.g. *Zooglea ramigera*,
(2) failure to achieve the proper physiological state conducive to floc formation
(3) physicochemical conditions inhibiting aggregation, and
(4) improper conditions to selectively favor floc formers (especially since scavenging protozoa are not present).

Because the main goal of thermophilic digestion is to reduce the level of organic matter in the waste stream, the predominant microorganisms are aerobic heterotrophs. Only certain species of organisms can proliferate in these reactors, those identified include *Bacillus, Thermus,* and actinomycetes. Thermophilic bacilli generally have complex growth requirements, and thrive well on the mixed substrate of sewage sludge, and therefore tend to be the dominant species (Lapara and Allemen, 1999).

Cell – free culture supernatants of proteases-positive *Bacillus* spp. can lyse vegetative bacterial cells. *Bacillus* spp. play an important role in the initial stage of thermophilic aerobic digestion. In the initial stage of the process, the mesophilic organisms present in the feedstock are subject to a temperature shock, once introduced into the reactor, which deactivates them. At the same time, the exo-enzymes (proteases) of the thermomphilic seed sludge cause lysis of the cell structure releasing the readily biodegradable matrix within the cell (Haner *et al.,* 1994). The subsequent stage of aerobic thermophilic digestion involves cryptic growth (endogenous respiration) which can be described by the utilisation of lysis products by active cells and once this carbon source has been eliminated, starvation of the active biomass occurs causing the organisms to digest their own protoplasm, leading to their subsequent death (Gaudy and Gaudy, 1980; Nasrin *et al.*, 2000). Cryptic growth essentially promotes biomass yield co-efficient minimisation as a result of the dissimilation of the lysis products. Both lysis products and extracellular metabolic products supplement the pool of soluble nutrients available to the process culture (Lapara and Alleman, 1999).

## 3.3 Fundamentals in Design and Operation of an ATAD

In order to achieve autothermal conditions, operational controls must typically be maintained as a closed loop system, i.e. it is necessary to control heat generation and heat lost from the system. The fundamental issues in operating an ATAD are:

- An adequate thickened sludge with an appropriate quantity of biodegradable volatile solids,
- An effective aeration and mixing system.
- A well insulated reactor, together with a low net heat loss system design, e.g. heat exchangers to recover heat from the outgoing hot sludge

### 3.3.1 Tank configuration

Due to the high rate of oxidation and lower HRT (5 – 9 days) experienced in ATAD, reactors are commonly smaller than those used for conventional aerobic digestion and those employed for anaerobic digestion. This smaller tankage requirement is a significant capital cost saving factor. Reactors are typically cylindrical in shape and constructed of carbon steel with an interior coaltar-based coating. Systems are located above ground, constructed on a concrete slab base (U.S. EPA, 1990). Insulation is an important consideration to retain the autothermal heat generated, and to maintain the digester's contents in the thermophilic temperature range. Mineral wool of approximately 10 cm thickness, is fitted around the tank walls for insulation, while styrene foam is used on the cover. Insulating to these specifications ensures the overall heat transfer co-efficient is 0.3 – 0.4 W /($m^2$ °C). The outside of the digester is covered with a metallic skin (corrugated steel or aluminium) to protect the insulation. Due to the nature of the aeration devices utilised in ATAD systems, the H/D ratio of the reactors must be higher than 1 and typically varies between 2- 5 to ensure complete mixing. An important factor of design includes a head space (free board) of approximately 0.5 and 1 m to allow for foam production (U.S. EPA, 1990).

### 3.3.2 Feed characteristics

As a minimum, 3% Total Solids (T.S.) may be fed to the reactor. T.S. in the range of 4 – 6% and 40 g/L COD or greater is recommended. A feedstock with less than 3% TS tends to contain too much water, and will not achieve autothermal conditions. While T.S. concentrations greater than 6% cause difficulty in mixing and aeration. The feed must contain a minimum volatile solids content of 75%, (U.S. EPA, 1990; W.E.F, 1995).

The degradation of organic matter, in the form of V.S., during the ATAD process is the main source of heat energy supplied to the reactor. Feedstocks have varying quantities and qualities (in terms of biodegradability) of volatile solids, depending on the unit operation from which they were produced. Table 3.1, shows typical values of heat output for a number of substrates. As the energy available from the breakdown of the volatile solids is used to heat the sludge within the reactor, the reduction in V.S. will indicate

- The heat output from the sludge
- The biodegradability of the particular sludge
- The efficiency of the process

As the quantity and quality of V.S., varies depending on the type of wastewater treatment process from which they were produced, so too does the amount of V.S. removed during the ATAD process, as can be seen from Table 3.2 (Stentiford, 2001). Typically, primary sludges are more readily biodegradable than biological sludges, and this is reflected in the percentage of volatile solids removed. From Table 3.2, one would conclude that primary sludge is a more suitable substrate for ATAD. However, Kelly (1995) contradicts this point in favour secondary sludge stating that while "*the volatile component of primary solids is made up of extracellular organics, it is not readily available as a pre-packaged substrate for thermophilic bacteria, as is the waste biomass*".

In sewage treatment plants operating at a low F/M, it is advisable the solids retention time in the aeration basin is kept below 15 days, to ensure that volatile solids reduction via endogenous respiration is kept to a minimum. The ATAD process can still be used on a low F:M substrate, however. If the heat output from the feed is found to be low (determined by V.S. % in the feed and biodegradability), an external preheat source via heat exchangers within the reactors could be applied (U.S. EPA, 1990; W.E.F, 1995).

**Table 3.1 Heat output from different wastes for each kg of volatile solids removed** (Stentiford, 2001)

| Material | Heat output (kJ / kg – VS removed) | Source |
|---|---|---|
| Organic fraction of municipal solid waste | 29,500 – 30,900 | Wiley 1957 |
| Mushroom compost substrate | 15,400 – 22,000 | Harper *et al.*, 1992 |
| Sewage sludge | 23,000 | Haug 1993 |
| | 21,000 | Andrews and Kambhu 1973 |

**Table 3.2 Volatile solids reduction in full sale ATAD plants with different sources of sludge.** (Stentiford, 2001)

| Sludge source | Volatile solids reduction% |
|---|---|
| Extended aeration | 25 – 35% |
| Primary + surplus activated | 30 – 56% |
| Primary + surplus activated + trickling filter | 43-66% |
| Surplus activated sludge | 25-40% |

### 3.3.3 Hydraulically controlled sludge age

During steady state conditions of the ATAD process, one of the most important variables is sludge age. In ATAD facilities, solids retention time is equal to hydraulic retention time, typically in the range of 5 – 9 days. When food supply, reactor volume, and feed rate are constant, sludge age in the system is expressed as :

$$\text{Sludge age} = \frac{Ve \times Xv}{Q \times X w}$$

Where Ve = reactor volume

Xv= concentration of reactor biomass

Q= flow or waste rate and

Xw = concentration of waste.

Since for this system the concentration of reactor biomass (Xv) and the concentration of waste (Xw) are equal, reactor volume (Ve) and flow (Q) are the control variables. Flow rate Q can be intermittent or continuous. When intermittent, the reactor is batch fed (once per day) or on an hourly basis. Digester volume (Ve) is the effective volume/ working volume. Adequate mixing of reactor is important, as a small reduction in an already small volume reduces the sludge age, increasing the likelihood of active biomass washout. Pre-thickening of the feed solids can facilitate the introduction of a greater organic loading rate [by feeding the same volumetric loading rate (V.L.R. l/l/d) but a higher organic loading rate (O.L.R. g.vs/d)] or can lower the flow rate [by feeding the same OLR but in a smaller volume (VLR)], which is an important sludge age control parameter (Kelly, 1995).

### 3.3.4 Feed Cycle

ATAD systems typically consist of two reactors in sequence. In larger centres of population, up to four reactors in sequence have been used (Oleszkiewicz, and

Manvinic, 2002). ATAD's are operated in either batch or semi batch mode. Batch mode is commonly applied in small treatment plants, whereby the system is completely emptied and filled every 8 days. Greater operator attention is required for systems operated continuously or on a semi batch mode. ATAD reactors that are operated in series are typically semi batch fed systems, in which a reverse – order filling procedure is applied. The typical operating strategy involves the following steps:

**(1)** At a predetermined time each day, aeration and mixing ceases

**(2)** Stabilised sludge is discharged from reactor II into the storage tank, which lowers the level in that reactor.

**(3)** Sludge is then allowed to flow by gravity or pumped from reactor I to reactor II, to replace that discharged.

**(4)** Raw sludge is pumped from the feed storage tank to reactor I until the volume in both tanks is equal. When both reactors are at operating level, aeration and mixing resume.

Daily feeding is typically completed in 30 min. to 1 hour, which ensures that both tanks are maintained at thermophilic temperatures (55 $^{o}$C) for at least 23 hr/day, ensuring the final product sludge is of Class A quality (U.S. EPA, 1990).

### 3.3.5 Effective aeration and mixing

#### (A) Oxygen take-up rate (OTR)

It has been well documented that oxygen saturation is affected by temperature, and it has been estimated that an increase in temperature from 20 – 55 $^{o}$C decreases oxygen saturation by 44% (i.e. at 20 $^{o}$C D.O. solubility in aqueous solution at po = 9.2 mg/l, whereas at 55 $^{o}$C = 5.15 mg/l)

There are numerous factors governing oxygen transfer rate such as :

- Air flow rate
- Bubble diameter
- Temperature

- Viscosity
- Tank geometry
- And wastewater composition.

The lower oxygen solubility leads to a smaller driving force and hence to a low oxygen take up rate (OTR). However, temperature also increases as the liquid viscosity and surface tension decreases, increasing the rate of oxygen transfer from gas to liquid phase, which may offset the smaller driving force, resulting in an overall positive effect on the OTR (Vogelaar *et al.*, 2000). Thus, facilitating the required dissolved oxygen concentration to the reactor may not be as problematic as lower solubility suggest (assuming 44% decrease in oxygen saturation at 55 °C, and including the fact that high rate oxidative nature of the process, would result in high BOD {$O_2$ consumption}. This was demonstrated by Jewell and Kabrick (1980) in which autothermal conditions prevailed in a pilot scale digester utilising two simple self – aspirating aerators, with an oxygen transfer efficiency exceeding 20%. Nonetheless, the provision of an adequate aeration system is one of the deciding factors in the success of the ATAD process, essential for autothermic conditions to prevail.

**(B) D.O. Concentration Monitoring**

It is believed that dissolved oxygen concentrations in excess of 0.5 mg/l are adequate in most ATAD systems. However, to date, there has been little published data in relation to this parameter. Few D.O. probes work at thermophilic temperatures and the accuracy of such instruments is likely to be questionable (due to the viscosity of the sludge, D.O solubility at thermophilic temperatures and low level of detection). In many instances, on-line monitoring of D.O. has not been possible; in such cases, oxidation- reduction potential (ORP) has been applied as a surrogate indicator. For the majority of the ATAD cycle, ORP values are typically in the range of - 50 to + 150 mV. An intermediate transient decline in ORP values (e.g. –100 to –300 mV) has however been noted during a periods of accelerated solids lysis and degradation (Pressley, 2001), following feed administration, when there is an excess of substrate awaiting utilisation and exerting a significant oxygen demand.

**(C) Mixing and aeration devices**

Aeration devices in ATAD are employed with the two fold purpose of aerating the reactor, and mixing the digester contents. The following aeration / mixing devices have been used in ATAD

- Aspirating aerators
- Combination recirculation pump / venturi arrangements and
- Turbine and diffused air

Aspirating aerators are the most common type of aerator used in ATAD systems. Generally, two aspirating aerators are side – mounted on the reactor. The angled installation of these aerators facilitates vertical downward mixing. Larger reactors commonly employ a third aerator, centre-mounted. The mixing motion in this case causes a circular horizontal flow pattern within the reactor. Aspirating aerators have the motor and bearings located outside the reactor, which reduces wear and tear on the device and is also useful for maintenance. Typical design parameters for this type of aeration device consists of:

- Specific power of 85 to 105 $W/m^3$ of active reactor volume,
- Air input of 4 $m^3/m^3$ of active reactor volume
- Energy requirement of 32 to 54 $MJ/m^3$ of solids throughput and
- Oxygen transfer efficiency of 2 kg/ $O_2$ /KWh.

For the above design criteria (4 $m^3/m^3/h$ air input) a feed sludge with V.S. range of 2.5 – 5% has an oxygen requirement of 1.5 $kg/O_2/kg$ V.S. destroyed (W.E.F, 1995).

**(i) Combination Recirculation Pump / Venturi Arrangements**

The Dutch aeration design of a combination recirculation pump and venturi arrangement was first used in the United Kingdom in 1987 and tested in Canada in 1989. A venturi aerator works by compressed air, which is fed to the venturi by a pressure regulator and gas flow meters. However, where recirculation is also employed externally, mounted pumps for circulation are used. In this case, sludge is drawn from the tank and pumped back through a nozzle and venturi which draws in ambient air. One advantage of this type of aerator is that the pump and venturi are located outside of

Where KT1 and KT2 are the reaction rates at temperatures T1 and T2 (°C), respectively, and Φ is a constant with typical values in the range 1.05 – 1.06. However, should the temperature become too high, the process becomes self-limiting and biological activity ceases, resulting in cooling of the reactor until the temperature eventually reduces to zero. This is represented in equation 2, where the first term on the right-hand side represents the increasing rate and the second term the reduction due to excessive temperature.

Equation 2: $KT1 = KT2 \{\Phi 1 (T1\text{-}T2 - \Phi 2 (T2 - T3)\}$

Where T3 is the upper temperature where inhibition begins to take place, and Φ1 and Φ2 are the increasing and decreasing temperature factors respectively. This equation shows that the rate will increase rapidly from ambient to 45 – 60 °C, above which it will decrease. It is thought that, above 65 °C, the rate drops rapidly to ambient (Stentiford, 2001).

In the majority of ATAD systems, the biodegradation of V.S. in the feed, results in the net release of energy, which heats the digester, and, therefore, is an obvious control factor (i.e. temperature can be reduced by reducing the OLR). In instances where this is not possible however, reactor cooling may be applied (if provisions for such a system exist at the facility). A number of other operational controls have been considered, such as reducing the system's temperature by reducing aeration. This method is not thought to be viable and may give rise to odours (Schwinning *et al.*, 1997).

### 3.3.7 pH

pH generally does not need to be controlled, as the thermophilic temperatures achieved during digestion ensures that nitrification is suppressed within the ATAD reactor, therefore the pH depression commonly experienced in nitrifying environments does not occur. In most cases, the pH is typically above 8.0, (when a feed of pH 6.5 is applied) (U.S. EPA, 1990). pH can in fact be used as (a) an indicator of autothermal

conditions and (b) as an early warning to unstable conditions. Should oxygen become limited due to organic overloading, pH may decrease, indicative of volatile fatty acids production during the fermentation of organics.

### 3.3.8 Foam production and control

Foam generation is a characteristic feature of the ATAD process. Foam has beneficial as well as negative effects on the overall process performance. The foam layer contains active biomass. All air that leaves the reactor must pass through the foam layer. In doing so, it contacts biologically active biomass that is entrained within the foam. The overwhelming biological activity occurring within the foam layer is believed to contribute to high oxygen utilisation efficiency. In addition, foam is also critical in the insulation of the reactor contents (Kelly, 1995). The majority of problems arise from excessive foam production. During start-up and process instability, excessive foam production has been noted at specific temperature thresholds. For example, at temperatures of 40 – 45 °C, foam is generated due to the break down of cellular proteins, lipids, oils and greases (Tchobanoglous *et al.,* 2001), as a response to a population shift of competitive bacteria from thermotolerant to true thermophilic strains. Biological foam is billowing and light brown in appearance. In the temperature range of 50 – 55 °C, foam production has been observed when a thin feed is added, which can generally be avoided by prethickening the feed to at least 5% T.S., and by using heat exchangers to heat the feed to thermophilic temperatures. Over-aeration also produces foam and rising air bubbles, which may cause solids floatation. Foam produced from over - aeration differs from biological foam in appearance, as it is frothy and matlike. Incorrect aeration can lead to partial solids washout. This foam -over leaves a characteristic brown mat on the ground with a feltlike appearance and feel.

There are a number of operational controls that can be applied to reduce foaming in the short term. Feeding always reduces foaming, which is possibly why foam is not normally found in excess in the first digester (Kelly, 1995). Shut down of aerators has been cited as a means to control excessive foam build – up. However, repeated shut down may induce oxygen deficient conditions, resulting in odours (Schwinning *et al.,*

1997). Most ATAD reactors incorporate mechanical foam breakers into the design. In the Fuchs system, for example, two mechanical foam cutters are installed in the reactor at fixed elevation. When the foam reaches a certain level in the tank (a sonar level transponder can be used to sense high foam) the foam cutters will automatically start to cut it up (U.S. EPA, 1990). A large freeboard allowance of 0.5 to 1.0 m is usually allocated in tank design (W.E.F, 1995). ATAD reactors without mechanical foam cutters commonly employ the use of sprays and defoamants.

### 3.4 ATAD Odour Generation and Control

Due to the thermophilic temperature of operation, ATAD results in the destruction of volatile solids, coupled with partial cell structure lysis. While both of these reactions are welcomed, the resultant release of odourous compounds is not. The odour produced from sludge digestion include: the reduced – sulfur compounds methyl and ethyl mercaptan (1 to 150 ppm), dimethyl disulfide and dimethyl sulfide (0.5 to 40 ppm) and ammonia concentrations up to 1400 ppm. These reduced sulfur compounds are products of protein breakdown. The presence of sulfur compounds is indicative of conflict between the aerobic and anaerobic nature of the process, further evidence of this is found in systems where the oxygen demand exceeds the oxygen supply. In such cases, high levels of VFA's have accumulated, which are known to be by - products of fermentation (Li *et al.*, 2002). Digesters exhibiting sulfurous odours have been described as first generation reactors, as these digesters are not truly aerobic. Second generation reactors as described by Pressley (2001) on the other hand, are more closely monitored and there is greater operational control, such as online oxygen utilisation rate (OUR) measurement and oxygen supply on demand. In typical reactors, the concentration of the ammonia and reduced sulfur compounds varies in relation to the feed cycle of the ATAD, and peaks in the production of these compound is often apparent. In order to determine peaks in production, monitoring should be conducted over a 24 hr period (to design for peak flows to the odour control system) (Pride, 2002).

### 3.4.1 Plant Management in Odour Control

There are a number of factors which can further add to the development of odours during ATAD, which include: sludge holding tank management practices, feed management practices, no or ineffective odour control, and excessively high operating temperatures (Schwinning *et al.*,1997).

**a) Storage Tank Management:**

In order to provide the digester with an adequate supply of V.S., prethickening is an essential component of the ATAD plant. Sludge holding tanks are also required to store the digested sludge until it can be applied to land. Consideration of the management of these systems can reduce odours. For example, prethickeners and sludge feed storage tanks should be managed to ensure that the feed is as fresh as possible on a daily basis. Post ATAD storage tanks have the twofold function of storage and cooling the sludge. Since ATAD digestion is at thermophilic temperatures, the sludge is hot when leaving the digester. By cooling the sludge, the residual biological respiration rate is slowed down, which reduces the potential for odour generation. Cooling can be achieved by mixing the holding tank or by a heat exchanger (Schwinning *et al.*, 1997).

**b) Temperature Control**

Control and maintenance of the operating temperature at 55 °C prevents the re-solubilisation of organics (by further cell lysis at extreme temperatures above 65 °C) (U.S. EPA, 1990). As previously mentioned, this cell lysis releases significant quantities of odorous compounds

**c) Odour Removal Systems.**

A problem exists with the removal of mercaptans, dimethyl disulfide, and dimethyl sulfide, as these particular compounds are insoluble in water, thus chemical scrubbing is ineffective. Activated carbon is also unsuitable, due to the high humidity of the ATAD air (100%, approx.). The insoluble nature and low odour threshold of these compounds ensures that they can persist in the environment, and can therefore spread over long distance from where they were produced (Pride, 2002). A number of systems have been

utilised for odour control including: soil biofilters, wet chemical scrubbers, bioscrubber towers, and trickling filter towers. All are effective if maintained and operated correctly. The most recent development in odour control is the biofilter system (Kelly, 1995), (in a biofilter, bacterial degradation together with adsorption onto the surface of the biofilm are of equal importance in the removal of insoluble compounds). Information from full scale plants in North America indicates that, while biofilters work best for soluble compounds (as these compounds are broken down on entry to the biofilm), longer residence times are required, however, for effective removal of mercaptans, dimethyl disulfide and dimethyl sulfide. Table 3.3 illustrates the maximum concentrations of odorous compounds found at full scale ATAD plant in North America, fitted with biofilter (from Ambio ltd) (Pride, 2002).

**Table 3.3: Concentration of Odorous Compounds Found at Full Scale ATAD Plant in North America.**

| | Long Sault Ontario | Cardinal Ontario | Mc Minnville Oregon | Franklin Indina | Princeton Indiana | Odour Threshold |
|---|---|---|---|---|---|---|
| Mercaptan (ppm) | ~10.0 | ~6.0 | 128 | ~8.0 | ~10.0 | 0.02 |
| Dimethyl disulfide (ppm) | ~1.0 | N/A | N/A | N/A | N/A | 0.002 |
| Dimethyl sulfide (ppm) | ~1.0 | N/A | 9.63 | N/A | N.A | 0.01 |
| Ammonia | 1150 | 900 | 1400 | 1200 | 400 | 10 |
| Year installed | 1994 | 1996 | 1997 | 2000 | 2000 | |
| ATAD supplier | FUCH'S | CBI-Walker | FUCH'S | FUCH'S | FUCH'S | |
| Empty bed residence time | 28 | 28 | 180 | 48 | 33 | |
| Air flow (cfm) | 1500 | 1500 | 3000 | 600 | 600 | |
| % of air flow from ATAD | 14 | 14 | 100 | 100 | 100 | |

Note : outlet concentrations of ammonia and reduced – sulfur compounds from the biofilters are generally below detection limits and, given the lack of offensive odour, must be in the range of the odour threshold (Pride, 2002).

There are a number of factors which need to be addressed when designing a biofiltration unit to ensure efficient removal of odourous compounds, such as maximum use of the bed volume, proper selection of filter material, and good humidity and pH control.

- **Max use of bed volume**: uniform delivery of air to the filter bed prevents development of anaerobic niches which may contribute to odour.

- **Filter material:** in order to keep the filter bed aerobic, a filter media that is resistant to compaction should be used.
- **Humidity and pH control:** there should be primary humidification prior to entering the filter bed and a secondary irrigation system in the bed itself, to wash out acid metabolites, which ensures pH is controlled in the bed and prevents acid accumulation (Pride, 2002).

### 3.5 Final Product Sludge Quality

Enteric pathogens from warm-blooded animals are successfully inactivated at temperatures greater than body temperature (37 °C). Thermophilic temperatures (55 °C) of operation are capable of inactivating approximately 99.9% of pathogenic organisms, and require a lower HRT than mesophilic processes (Krishna, 2000; G.V.R.D, 2000). Many studies have illustrated that the ATAD process destroys pathogens in accordance with the strict requirement of both Rule 503 US EPA, and the more recent and applicable to the Irish situation *Code of Good Practice in the Use of Sewage Sludge in Agriculture.* This means that the Biosolids produced from the ATAD process, can be applied to agricultural land, providing the sludge meets the other requirements of legislation, in regards to metal content for example. The sanitation of the product sludge is a direct reflection of plant management. Good operational control in regard to feeding regime and final product storage are critical factors in ensuring pasteurised biosolids do not get re-contaminated (Kelly, 1995; Kelly, 1997).

Time and temperature control is also critical, as demonstrated in a study conducted by Gantzer *et al.*, (2001), as nematode eggs were effectively eliminated by ATAD digestion at a temperature of 48 °C and above, but were detected when the temperature fell to 25 °C during the winter months of plant operation. In addition to meeting the required level of sanitation, the final sludge will appear well mixed, with a light to dark brown colour, and may be odourless or have a slightly musty odour. Volatile solids reduction is typically in the range of 40 – 60% (however, the actual V.S. reduction depends on the nature of feed solids) (Kelly, 1995).

### 3.5.1 ATAD Dewaterability

Little attention to date has been focused on the dewatering of ATAD biosolids, as the general practice throughout Europe is to spread the Biosolids on land in the liquid state, while the ATAD process is a relatively new to the US, first introduced in 1990. Full-scale applications of ATAD have reported poor biosolids dewaterability, with polymer dosages being 3-10 times (typically in the range of 44 to 50 kg/dry-ton total solids) more than that required for dewatering mesophilic aerobically digested sludge. This increase in polymer usage can significantly increase the overall cost of ATAD process. A limited number of studies on ATAD biosolids dewaterability have reported that the use of inorganic chemicals (ferric chloride) as a substitute for organic polymers can reduce the overall cost of dewatering (Kelly *et al.,* 2000, Murthy *et al.,* 2000b). Despite the successful implementation of this technique (the conditioning cost for ATAD treated biosolids has been reduced by 67%), at three full-scale ATAD installations in the USA; (College Station, Texas, Ephrata, Pennsylvania, and Princeton, Indiana), this technique has not been considered as a viable application in the long term, as the mass quantities of inorganic chemicals required to be added into ATAD biosolids were up to 5 times greater than that of organic polymers, increasing the weight of biosolids for final disposal. Such inorganic chemicals are corrosive in nature and need special handling and storage. Another technique proposed to reduce polymer addition and improve dewaterability involves the storage of ATAD biosolids for 25 days in mesophilic aerated tanks (Murthy *et al.,* 2000a). However, this procedure would result in the need for additional tankage and an increase in power usage, and therefore is considered not economically feasible (Zhou *et al.,* 2001).

The fundamental reasons for poor dewaterability are largely unknown, much like the microbiology of the process. It has been documented that floc forming organisms are absent from the digester's microflora, resulting in the dispersion of bacteria cells uniformly throughout the reactors contents, which is likely to be one of the determining factors in dewaterability. Pioneering research aimed at conveying greater insight into the effects of number of individual parameters on ATAD biosolids dewatering [such as feed composition, the production of extra-cellular bi-polymers (i.e. protein and

polysaccharides), temperature and sludge retention time], was conducted by Zhou *et al.*, 2001. The main findings of the work illustrates that:

- The composition of feed sludge affects dewatering properties. Thermophilic aerobic digestion of secondary sludge resulted in a significant reduction in dewaterability, while thermophilic aerobic digestion had less of an effect on the dewaterability when treating Primary sludge and Mixed Sludge.

- Sludge retention time was not a major factor in the dewaterability of thermophilic aerobically digested sludge.

- Rapid deterioration of thermophilic digested biosolids from Secondary Sludge was accompanied by rapid increase in amounts of soluble proteins, polysaccharides, and phosphate.

- Thermophilic digestion resulted in smaller and finer particles (biosolids flocs) than mesophilic digestion. These smaller particles could also be a cause for poor dewaterability and increased demands of polymers.

# CHAPTER 4
# MATERIALS AND METHODS

## 4.1 The Pilot Scale Study

The following section gives details of the pilot scale reactors configuration, seeding and set-up, and methods of analysis.

### 4.1.1 Reactor Configuration

Two 16 L. jacketed glass reaction vessels were made to order by AGB scientific, Dublin to required design specifications. Both aerobic and anaerobic reactors are identical in design for comparative operational assessment (Plate 4.1).

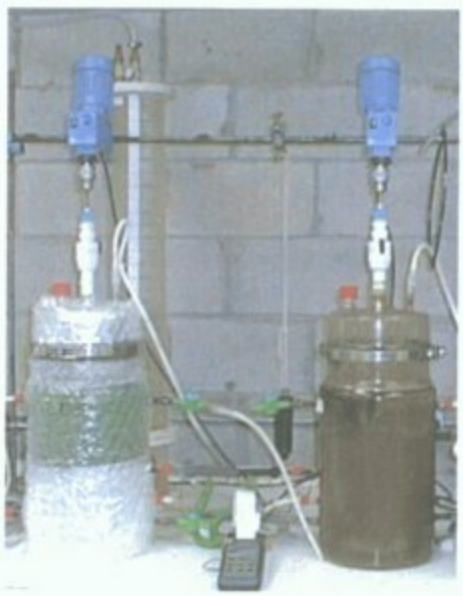


Plate 4.1 Pilot Scale Digesters in Operation

There are five portholes positioned at varying distances from the base of the unit. Portholes were supplied with HDPE screw caps for leak-proof closure, all of which were fitted with tubing & tubing clamps or taps to allow for later use as required, except for the second central screw cap, which was drilled through the centre to facilitate the housing of thermocouple probe "K type", which was connected to a Hanna printing and logging

thermometer (AGB) programmed to record and print the temperature of both reactors simultaneously every 3 hours.

The reaction vessels have removable glass lids attached to the base unit by a stainless steel flange. The lid has three portholes of various size, a 10mm port with screw cap, a 5 mm port with screw cap and the central porthole which is ground glass to allow the secure fitting of the "Cowie" stirrer shaft and guide. Mixing of the reactors was achieved by "IKA" overhead stirrers purchased from Lennox Lab supplies, Dublin.

### 4.1.2 Modification of Reactors for Specific Use.

**(A) Anaerobic Digester**

A 10 L. Memmet hot water bath was used to heat the water, which circulated through the double wall (jacket) of the reaction vessel. The water bath was thermostatically controlled and programmable, ensuring that the water was always maintained at the mesophilic temperature of 35 °C, supplied by Lennox Lab Supplies, Dublin. A submersible a (fountain) water pump, pumped the hot water from the bath to the reactor, entering the vessels jacket through the lower nippled porthole and exiting through the upper porthole, back to the water bath for reheating. The reactor was maintained at 35 °C +/- 1°C for the duration of the mesophilic trail.

**(B) Aerobic Reactor**

Air was supplied to the aerobic reactor through a porthole close to the base of the reactor; via an air compressor. Three different air compressors were used over the course of the study due to operational failure from continuous use. Air was diffused within the reactor through a handmade diffusion tube. Industrial diffusion tubes were too large for the reactor, and those used for fish tanks were too small and did not evenly supply air to the vessel. The diffusion tube consisted of plastic tubing, randomly perforated with pin holes. A loosely

sprung spring (also handmade) was put through the centre of the tubing to prevent it from collapsing. Both ends of the tubing were fitted to a T – Piece tubing connector to form a circular shape and was fitted to the inlet (glass tubing) at the bottom of the reactor, this device was successful in ensuring even distribution of fine bubbles within the reactor.

The 10mm porthole on the lid of the digester was modified by the addition of a detachable condensation trap (D.O was monitored daily through this porthole also), which was added to the vessel following the application of heat to the reactor. The trap consisted of a glass distillation apparatus, through which cold water was passing, cooling the air and reducing evaporation from the reactor. In addition, 2 L. of water was added to the reactor daily.

### 4.1.3 Anaerobic Digester Seeding and Start-up

Seeding was carried out using two seed sources. Digester sludge was obtained from a full-scale anaerobic digester in Tullamore, Co. Offaly. There was also a pilot scale reactor on site from which sludge was taken. Following collection of the samples, solids analysis was conducted and the seed was stored overnight in a refrigerator. It is common practice when transferring viable anaerobic biomass to a new reactor to pump directly to the unit while sparging Nitrogen through the headspace, in order to maintain anaerobic conditions. As there was no nitrogen on tap in the effluent laboratory in this case, inactivation of the anaerobic biomass by reduction in temperature was deemed the most appropriate method of ensuring a successful start-up.

The cooled 4.5 L. of digester sludge from the pilot scale reactor contributed a total of 52.074 g V.S., while 6.0 L. of Tullamore sludge was passed through a sieve to remove debris prior entering the reactor (Total V.S. from this source 77.736 g). 28g $NaHCO_3$ dissolved in 700 ml of tap water was also added to the 16 L. vessel in order to bring the liquid level to the designed working volume to 11.2 L. leaving 30% of the reactors volume void as headspace. The final V.S. content of the reactor was 11.6 g V.S. / L. reactor. The reactor was sealed and heated to 35 °C. On day three, the mixer was commissioned to

operate twice per hour at 15 minute intervals, at 60 RPM, which was later revised to operate continuously.

### 4.1.4 Seeding of the Aerobic Reactor

Pressed extended aeration sludge was obtained from Manorhamilton Sewage Treatment Plant in Co. Leitrim. The sludge was pressed via double belt press with the addition of an organic polyelectrolyte "Zetag 78N" to a total solids concentration of 11.6 %.

The pressed sludge was reconstituted to 4.9% T.S. with tap water, and mixed by overhead stirrer for 10 minutes, to break up the pressed sludge cake. The reconstituted sludge was sieved prior introduction to the reactor. 9.5 L. of sludge was used to seed the reactor. On completion, the unit was sealed and the aerator was switched on, to operate at continuous mode, delivering an excess of 1.5 – 2.0 mg/l D.O. to the sludge.

Mixing commenced immediately on a continuous basis at 60 RPM. The final Volatile solids concentration of the reactor was 33.18 g V.S / L. reactor. The digester was insulated using 6 inch " Rockwool". Heat was applied later by hot water bath, to maintain the digester temperature at 53 °C.

### 4.1.5 Feedstock Collection and Storage

On a weekly basis, a fresh sample of pressed sludge was obtained from Manorhamilton Sewage Treatment Plant. A 2.5 kg polyethylene sample bucket was employed for this task. The container was filled to the brim with fresh sludge and sealed. On arrival at the Water Engineering Laboratory (I.T. Sligo) the sample was analysed for Total, Volatile and Fixed Solids, and refrigerated until required.

### 4.1.6 Feedstock Preparation

The pressed sludge had a Total Solids content of approximately 12%. Each morning, a sample of the pressed sludge was weighed, proportional to the required Total Solids of the feed required, and reconstituted with tap water. The reconstituted sludge was mixed for 30 seconds by food blender and sieved to remove debris that could otherwise clog peristaltic pump tubing etc.

### 4.1.7 Digester Feeding

Both reactors were batch fed once per day. Sludge was fed to each reactor via peristaltic pump (Masterflex) using standard rate pump head and 6.7mm internal diameter tubing. Mixers were shut off prior feed administration. While feeding the anaerobic unit, effluent removal occurred simultaneously by simply opening the effluent porthole tap and allowing the required volume of effluent to be drawn off by gravity. In the case of the aerobic unit however, in addition to cessation of mixing, aeration was also stopped, effluent was gravity removed, and the digester was then fed. Total, Volatile and Fixed Solids analysis was performed daily on the feed to ensure reconstitution of the sludge was conducted accurately and to facilitate precise calculation of OLR and VLR. pH, and COD analysis was conducted on the first batch of reconstituted feed weekly.

## 4.2 Analytical Techniques

### 4.2.1 Volatile Fatty Acid

Volatile Fatty Acids were analysed using two different techniques. For routine monitoring on a daily basis Total VFA concentration of the anaerobic effluent was determined by the HACH method 8196. However, as individual VFA analysis provides further information on process stability, samples were pre-treated, frozen and later analysed by GC at NUI Galway. Both methods are presented below:

#### 4.2.1.1 VFA by HACH Spectrophotometer

The HACH method 8196 (0 – 2,800 mg/l as acetic acid) is based on the esterification of carboxylic acids and the determination of the esters by the ferric hydroxamate reaction. The method is specifically designed for application on digester sludge.

**Apparatus and Reagents**

Spectrophotometer (HACH model DR 2000), water bath, centrifuge, Ethylene glycol, 19.2N Sulfuric acid, Hydroxylamine Hydrochloride, 4.5 N Sodium Hydroxide Standard, Ferric Chloride Sulfuric acid solution.

**Procedure:**

Program number 770 for volatile acid expressed as mg/l acetic acid was entered into the spectrophotometer, and the wavelength was adjusted to 495 nm. 0.5 ml of deionised water was transferred to a clean 25ml sample cell (Blank). The anaerobic digester effluent sample was centrifuged at 13000 RPM for 10 minutes, the clear supernatant of which (0.5 ml) was placed in another sample cell. 1.5 ml of ethylene glycol was pipetted into each sample cell and mixed gently. 0.2 ml of 19.2 N sulfuric acid standard solution was added to each cell and swirled to mix. Timer 1 was set and both cells were heated in a boiling hot water bath for 3 minutes. Following heating, the sample cells were cooled under a running tap. 0.5 ml of Hydroxylamine Hydrochloride solution was pipetted to each and mixed. 2 ml of 4.5 N sodium Hydroxide standard solution was added to each cell and mixed. 10 ml of Ferric Chloride Sulfuric acid solution and 10ml of deionised water was added to each of the sample cells and mixed. Timer 2 was enabled, following a three-minute reaction period the blank was placed into the cell holder and zeroed. Then the prepared sample was placed in the cell holder, measuring the mg/l VFA expressed as acetic acid.

### 4.2.1.2 VFA via Gas Chromatography

Samples were centrifuged at 13000 RPM for 10 minutes to remove suspended matter. Following the addition of 25 µl concentrated orthophosphoric acid ($H_3PO_4$) to 1 ml of the supernatant, samples were frozen at – 18 °C, and defrosted prior analysis.

**Procedure:** Volatile Fatty acids (VFA) were analysed by gas chromatography using a Shimadzu GC-14B chromatograph with a hydrogen flame ionisation detector, fitted with a carbopack glass column. The column temperature was maintained at 175°C. The injection port and detector temperatures were 200 and 250°C, respectively. Prepared samples were injected onto the column using a Shimadzu AOC-20i auto injector equipped with a 5 µl syringe. VFA concentrations were calculated using a Shimadzu C-R6A chromatopac and a previously constructed standard curve.

### 4.2.2 COD Total and Soluble

Standard Methods (19th ed., 1995), No. 5220 (Reactor Digestion Method)

**Sample Preparation**

Sample for total COD (CODt) must be homogeneous in nature, comprising of both solid and liquid (soluble) fractions. The sample for soluble COD (CODs) analysis is centrifuged at 13,000 RPM for 10 minutes or may be filtered to remove suspended matter. COD may then be carried on the clear supernatant as this represents the soluble fraction of COD.

**Apparatus and Reagents**

COD reactor, HACH (DR 2000) Spectrophotometer, HR COD vials 0- 1500 mg/l, purchased from a HACH agent (Celtic Engineering).

**Procedure:**

Samples to be tested were diluted so that the COD range was between 0 – 1,500 mg/l. 2 ml of the raw or diluted sample was pipetted into a HR COD vial, and inverted to mix. This procedure was repeated using deionised water as a blank. Vials were then placed in the preheated COD heating block for 2 hours (timed) at 150 °C. Once cool, COD vials could be read in the spectrophotometer. Program number 435 for HR COD determinations was entered into the DR 2000, and the wavelength adjusted to 620 nm. The blank was placed in the cell holder and zeroed. Following calibration with the blank the sample vial was then read, results were displayed as mg/l COD.

### 4.2.3 Alkalinity

Standard methods (19th ed., 1995), 2320 B (Titration method)

**Apparatus and Reagents**

pH meter, Burette and stand, Magnetic stirrer, Stirrer bar.

**Procedure:**

Following the determination of the exact normality of 0.1 N standard sulfuric acid, 20 ml of digested effluent was titrated against standard sulfuric acid solution to an end point of pH 4.3. The following calculation was then applied to determine the alkalinity of the sample, expressed as mg $CaCO_3$ / L

$$\text{Alkalinity, mg } CaCO_3 / L = \frac{A \times N \times 50{,}000}{\text{ml sample}}$$

Where:

A = ml standard acid used

N = normality of standard acid

### 4.2.4 pH

Standard Methods (19th ed., 1995), No. 4500 (– $H^+$ B Electrometric Method)

**Apparatus and Reagents**

Delt OHM pH meter, Buffer solution

**Procedure**

Electrolytes were rinsed with distilled water and dried prior use. The meter was calibrated prior use with a buffer solution of a known pH, the buffer solution used in this instance was pH 6.98, which was closest to the pH of the sample to be measured.

### 4.2.5 Total Solids

Standard Methods (19th ed., 1995), No. 2540 D (Total Suspended Solids Dried at 103 – 105 °C)

**Apparatus**

Analytical balance, Pre-muffled porcelain crucibles, Desiccator, Forceps, Oven at 105 °C,

**Procedure**

A measured volume of sample was evaporated to dryness in a pre-muffled (15 min at 550 °C) and weighed porcelain evaporating dish, and subsequently dried in a oven at 103 – 105 °C, until at a constant weight was achieved. The total solids concentration of the sample was then determined in mg/l by the following calculation:

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

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

B = Weight of evaporation dish.

### 4.2.6 Fixed and Volatile Solids

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

**Apparatus**

Dried and weighed samples from T.S. analysis, Muffle furnace at 550 °C , Forceps, Desiccator.

**Procedure**

Following Total Solids determinations, the dried solids were placed in a muffle furnace until a constant weight was attained. The following calculations were then applied :

$$\text{mg Volatile Solids / L} = \frac{(A-B) \times 1000}{\text{Sample volume, ml}}$$

$$\text{mg Fixed Solids / L} = \frac{(B-C) \times 1000}{\text{Sample volume, ml}}$$

**Where:**

A = weight of residue + dish before ignition, mg

B = weight of residue + dish or filter after ignition, mg

C = weight of dish mg.

### 4.2.7 Digester Gas Analysis

A gas analyser was employed for daily gas analysis. The LMSx multi gas analyser purchased from Mason Technology, Dublin, uses an infra red methane and $CO_2$ detector

coupled with other gas and environmental sensors. On a daily basis, the gas analyser was connected directly to the gas line on the lid of the reactor and a 30 sec. grab sample was taken using an inbuilt gas-sampling pump. The gas analyser then displays % $CH_4$, $CO_2$ and $O_2$ of the biogas simultaneously on the LCD screen. Atmospheric pressure, borehole flow rate and site are also monitored. Results of analysis may be stored to memory with the current date and time. The memory allows the storage of 1000 readings from multiple sites and boreholes. The LMSx multi gas analyser also has a facility, which allows data to be downloaded to spreadsheet or other application using the optional serial lead and 'Siteman Program'.

### 4.3 REVIEW OF FULL-SCALE SLDGE TREATMENT PLANTS

Operational full-scale Mesophilic Anaerobic Digesters and Autothermal Thermophilic Aerobic Digestion plants throughout the Republic of Ireland were visited, and primary data were collected on the sludge treatment plants; design, equipment, operation and where possible results analysis. For the purpose of comparative evaluation of the Anaerobic Digestion process verses ATAD, operational data and results of analysis from Tralee (Anaerobic Digester) and Killarney (ATAD), in County Kerry were reviewed.

# CHAPTER 5
# RESULTS AND DISCUSSION

Table 5.1 Feeding Regime and Feedstock Composition (Anaerobic Pilot Scale Digester)

| Week no. | OLR (g. V.S./D) | VLR ( L/L/D) | HRT(days) | T.S. mg/l | T.S. % | V.S. mg/l | V.S. % | F.S.mg/l | F.S. % | CODt mg/l | CODs mg/l |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | ***Start - up*** | | | | | | |
| 2 | 9.32 | 0.83 | 25 | 38261 | 3.80 | 27185 | 71.40 | | 28.20 | 35442 | 414 |
| 3 | 9.13 | 0.81 | 33 | 39890 | 3.97 | 26944 | 67.50 | 12949 | 32.50 | 29830 | 344 |
| 4 | 9.56 | 0.85 | 33 | 43080 | 4.29 | 28207 | 68.90 | 13389 | 31.07 | 35196 | 310 |
| ***Average*** | **9.34** | **0.83** | **30** | **40410** | **4.02** | **27445.33** | **69.27** | **13169** | **30.59** | **33489** | **356** |
| | | | | | ***Steady state operation at a 33 day HRT*** | | | | | | |
| 5 | 10.01 | 0.89 | 33 | 41797 | 4.10 | 29551 | 70.71 | 12246 | 29.27 | 44731 | 614 |
| 6 | 9.87 | 0.88 | 33 | 42602 | 4.22 | 29119 | 68.30 | 13480 | 33.60 | 44162 | 507 |
| 7 | 10.75 | 0.96 | 33 | 46290 | 3.80 | 31723 | 68.30 | 14579 | 31.60 | 30000 | 700 |
| 8 | 10.81 | 0.96 | 33 | 46159 | 4.58 | 31890 | 68.90 | 14268 | 31.17 | 39265 | 425 |
| 9 | 10.16 | 0.90 | 33 | 47039 | 4.70 | 29980 | 63.70 | 17058 | 36.30 | 39265 | 425 |
| 10 | 10.5 | 0.94 | 33 | 45588 | 4.50 | 31002 | 67.90 | 14585 | 32.00 | 48520 | 686 |
| 11 | 9.33 | 0.83 | 33 | 41450 | 4.20 | 27535 | 66.43 | 13911 | 33.57 | 49600 | 860 |
| 12 | 9.61 | 0.86 | 33 | 42084 | 4.20 | 28361 | 67.00 | 13723 | 33.00 | 40840 | 732 |
| 13 | 11.12 | 0.99 | 33 | 49177 | 4.87 | 32819 | 66.60 | 16246 | 32.62 | 43360 | 610 |
| 14 | 11.12 | 0.99 | 33 | 49070 | 4.80 | 32805 | 66.90 | 16153 | 33.16 | 48450 | 750 |
| ***Average*** | **10.33** | **0.92** | **33.00** | **45126** | **4.40** | **30479** | **67.47** | **14625** | **32.63** | **42819** | **631** |
| | | | | | | | | | | | |
| 15 | 9.80 | 0.88 | 33 | 43881 | 4.40 | 28926 | 66.00 | 14954 | 34.30 | 50000 | 1100 |
| 16 | no feed & recirulated | | | nf | | | | | | | |
| 17 | no feed & recirulated | | | nf | | | | | | | |
| 18 | 10.16 | 0.90 | 33 | 45517 | 4.53 | 29985 | 65.80 | 15532 | 34.20 | 49000 | 790 |
| 19 | 10.75 | 0.96 | 33 | 48808 | 4.80 | 31725 | 65.00 | 17083 | 35.00 | 47531 | 621 |
| 20 | 10.75 | 0.95 | 33 | 49080 | 4.86 | 31716 | 64.20 | 17363 | 35.80 | 48200 | 697 |
| 21 | 10.28 | 0.95 | 33 | 45538 | 4.52 | 30338 | 66.70 | 15200 | 33.25 | 47450 | 690 |
| 22 | 27.84 | 2.48 | 15 | 52841 | 5.38 | 37332 | 68.80 | 16797 | 31.14 | 52950 | 700 |
| 23 | 25.61 | 2.28 | 15 | 51637 | 5.12 | 34355 | 66.47 | 17281 | 33.30 | 52301 | 689 |
| | | | | | | | | | | | |
| | | | | | ***Steady state operation at a 15 day HRT*** | | | | | | |
| *27 | 23.26 | 2.07 | 15 | 44618 | 4.41 | 31186 | 69.90 | 13430 | 30.12 | 47000 | 490 |
| 28 | 23.51 | 2.09 | 15 | 45330 | 4.53 | 31516 | 69.25 | 13284 | 30.70 | 46720 | 1028 |
| 29 | 22.03 | 1.96 | 15 | 42884 | 4.30 | 29544 | 68.50 | 13339 | 31.10 | 45880 | 978 |
| 30 | 22.50 | 2.10 | 15 | 46132 | 4.60 | 31570 | 68.50 | 14660 | 31.50 | 50200 | 1010 |
| 31 | 22.60 | 2.01 | 15 | 45477 | 4.53 | 30323 | 66.56 | 15153 | 33.43 | 48720 | 960 |
| 32 | 22.60 | 2.02 | 15 | 47347 | 4.60 | 30389 | 64.16 | 16957 | 36.13 | 47040 | 784 |
| 33 | 22.90 | 2.05 | 15 | 48913 | 4.85 | 30796 | 62.95 | 18117 | 37.05 | 53120 | 973 |
| 34 | 24.30 | 2.17 | 15 | 52276 | 5.20 | 32699 | 62.40 | 19550 | 37.45 | 55000 | 1011 |
| ***Average*** | **22.96** | **2.06** | **15.0** | **46622** | **4.63** | **31003** | **66.53** | **15561** | **33.44** | **49210** | **904** |

* Following vacation shutdown period

Table 5.2 Anaerobic Pilot Digester Effluent Analysis Results

| WK no. | pH | pH Range | VFA mg/l | VFA Range | T.S. mg/l | T.S. Range | V.S. mg/l | V.S. Range | CODt mg/l | CODt Range | CODs mg/l | CODs Range | $CaCO_3$ mg/l |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| *Start - up* | | | | | | | | | | | | | |
| 2 | 7.26 | 7.22 - 7.3 | 8.00 | 0-22 | 16990 | 5990-18456 | 9064 | 1740-10244 | 13109 | 810-12340 | 429 | 361-493 | 4587 |
| 3 | 7.19 | 7.13-7.26 | 206.00 | 0-93 | 19738 | 17524-21612 | 11068 | 9920-12905 | 13260 | 11500-14750 | 446 | 353-650 | 4799 |
| 4 | 7.07 | 6.94 - 7.19 | 7.30 | 0-28 | 23244 | 6088-24620 | 13564 | 2732-15036 | 18635 | 3350-21525 | 533 | 481-620 | 3573 |
| ***Average*** | **7.17** | | **73.77** | | **19991** | | **11232** | | **15001** | | **469** | | **4320** |
| *Steady state operation at a 33 day HRT* | | | | | | | | | | | | | |
| 5 | 6.97 | 6.7-7.08 | 10.00 | 0-48 | 2466 | 24252-25064 | 14502 | 13804-15056 | 20175 | 16935-22635 | 631 | 508-785 | 4300 |
| 6 | 6.97 | 6.81-7.07 | 35.50 | 1-77 | 25636 | 25168-26076 | 15024 | 14100-15524 | 22172 | 20790-24405 | 868 | 698-1091 | 4589 |
| 7 | 6.86 | 6.78-7.02 | 70.80 | 3-96 | 26938 | 24708-29320 | 16175 | 14200-18192 | 22929 | 18900-27270 | 922 | 570-1320 | 4845 |
| 8 | 6.97 | 6.88-7.01 | 60.00 | 0-146 | 28686 | 28020-29436 | 18376 | 17648-19056 | 28787 | 28260-29190 | 751 | 927-975 | 4739 |
| 9 | | | | | 31916 | | 19860 | | 27120 | | 725 | | |
| 10 | 7.06 | 6.94-7.24 | 128.50 | 50-181 | 31641 | 29772-33012 | 19726 | 19680-20708 | 30262 | 27090-32520 | 1288 | 799-1634 | 5102 |
| 11 | 7.06 | 7.05-7.09 | | | 31128 | | 18900 | | 28800 | | 695 | | |
| 12 | 7.03 | 7-7.07 | | | 35252 | 34836-35252 | 22182 | 21216-23148 | 29055 | 28590-29520 | 1367 | 1264-1471 | |
| 13 | 6.98 | 6.95-7.05 | 134.20 | 98-156 | 35544 | 34927-35292 | 22062 | 21584-22856 | 31140 | 29850-32550 | 1390 | 1144-1650 | 4985 |
| 14 | 7.00 | 6.84-7.06 | 226.50 | 198-255 | 36416 | 35952-36880 | 22790 | 22704-22876 | 35835 | 31770-31900 | 2014 | 1808-2220 | 5208 |
| ***Average*** | **6.99** | | **95.07** | | **28562** | | **18960** | | **27628** | | **1065** | | **4824** |
| | | | | | | | | | | | | | |
| 15 | 6.98 | 6.9-7.06 | 202.20 | 176-228 | 32999 | 20871-61103 | 20283 | 12826-37579 | 22645 | 19230-27545 | 1491 | 876-1994 | 5110 |
| 16 | 7.02 | 6.87-7.18 | | | | | | | | | | | |
| 17 | 6.97 | 6.9-7.04 | 207.00 | 168.-271 | | | | | | | | | |
| 18 | 7.07 | 7.05-7.11 | 182.00 | 153.-191 | 33694 | 32012-35513 | 20074 | 19102-21284 | 31220 | 29370-31590 | 1376 | 1204-2228 | 5601 |
| 19 | 7.06 | | 255.00 | | 54972 | | 32852 | | 39000 | | 1502 | | 4090 |
| 20 | 7.01 | 6.98-7.04 | 189.00 | 184.-192 | 38142 | 35453-42249 | 22698 | 21110-25152 | 38510 | 37530-38850 | 1639 | 1558-1782 | 4317 |
| 21 | 7.02 | 6.96-7.07 | 176.00 | 161.-198 | 36660 | 35988-37338 | 22081 | 21761-22603 | 34845 | 32940-36510 | 1997 | 1830-2252 | 4317 |
| 22 | 6.97 | 6.95-7.01 | 150.00 | 93-204 | 35402 | 34727-37297 | 22029 | 21456-22887 | 34800 | 32340-36210 | 1556 | 1408-2066 | 4545 |
| 23 | 7.01 | 7-7.03 | 138.50 | 119-164 | 34649 | 33344-36117 | 21722 | 20933-22661 | 34280 | 33930-48120 | 956 | 788-1042 | 4573 |
| | | | | | | | | | | | | | |
| *Steady state operation at a 15 day HRT* | | | | | | | | | | | | | |
| 27 | 7.01 | 6.94-7.11 | 221.70 | 103-326 | 38749 | 37830-39412 | 23251 | 22370-24126 | 36075 | 33630-37830 | 1710 | 1438-2244 | 4772 |
| 28 | 6.99 | 6.5-7.03 | 162.60 | 128-188 | 36267 | 31718-38637 | 22557 | 20046-24141 | 33870 | 31032-36210 | 1966 | 1404-2362 | 4542 |
| 29 | 6.94 | 6.83-7.01 | 242.00 | 198-279 | 35441 | 33776-37143 | 22293 | 21182-23320 | 34918 | 33408-35500 | 2068 | 1578-2350 | 4789 |
| 30 | 6.91 | 6.8-6.97 | 216.50 | 150-377 | 32803 | 30374-34841 | 20802 | 19269-22479 | 31990 | 31080-34500 | 2080 | 1902-2258 | 4680 |
| 31 | 6.99 | 6.8-6.91 | 269.50 | 266-273 | 37162 | 35056-40570 | 23014 | 21157-26130 | 36390 | 31140-41160 | 2193 | 1884-2474 | 4700 |
| 32 | 6.85 | 6.75-6.9 | 201.00 | 165-237 | 36079 | 35155-37642 | 22170 | 21690-22961 | 34665 | 31560-38610 | 2055 | 1692-2412 | 4650 |
| 33 | 7.05 | 7-7.1 | 196.50 | 184-209 | 39447 | 37298-42903 | 24106 | 22565-26821 | 39550 | 35870-39550 | 2407 | 2338-2602 | 5020 |
| 34 | 7.06 | 7.05-7.07 | 289.00 | 260-318 | 44441 | 41971-47954 | 26828 | 25203-28828 | 40350 | 38670-42360 | 2943 | 2388-3964 | 4890 |
| ***Average*** | **6.97** | | **224.85** | | **37549** | | **23128** | | **35976** | | **2178** | | **4755** |

**Table 5.3 Biogas Composition**

| WEEK no. | $CH_4$ % | CH4 % Range | $CO_2$ % | $CO_2$ % Range | $O_2$ % | $O_2$ % Range |
|---|---|---|---|---|---|---|
| | | | ***Start - up*** | | | |
| 2 | 62.5 | 61-64 | 31.5 | 31-32 | 1.1 | 0.8-1.5 |
| 3 | 61.3 | 61-64 | 30.8 | 30-33 | 0.8 | 0-1.1 |
| 4 | 61.4 | 60-64 | 31.7 | 31-32 | 0.0 | 0-1.3 |
| ***Average*** | **61.7** | | **31.3** | | **0.6** | |
| | | | ***Steady state operation at a 33 day HRT*** | | | |
| 5 | 65.8 | 63-67 | 30.2 | 26-33 | 0.0 | 0 |
| 6 | 64.1 | 61-66 | 30.5 | 26-32 | 0.6 | 0-1.7 |
| 7 | 62.6 | 61-67 | 32.6 | 31-35 | 0.2 | 0-1.4 |
| 8 | 60.2 | 59-63 | 34.5 | 34-35 | 0.7 | 0-1.6 |
| 9 | 61.4 | 61-64 | 35.0 | 34-36 | 0.0 | 0 |
| 10 | 61.6 | 60-64 | 34.6 | 33-36 | 0.4 | 0-1 |
| 11 | 63.8 | 63-65 | 34.0 | 34 | 0.8 | 0.7-0.9 |
| 12 | 64.3 | 64-65 | 35.0 | 35 | 0.3 | 0-0.8 |
| 13 | 65.0 | | 33.6 | | 0.0 | |
| 14 | 64.8 | 64-66 | 29.8 | 33-34 | 0.1 | 0-0.7 |
| | **63.4** | | **33.0** | | **0.3** | |
| | | | | | | |
| 15 | 57.8 | 55-64 | 31.7 | 27-34 | 0.2 | 0-0.6 |
| 16 | 58.0 | 57-59 | 32.6 | 31-32 | 0.3 | 0-0.7 |
| 17 | 57.8 | 56-59 | 30.0 | 32-3 | 0.0 | 0-0.7 |
| 18 | 52.5 | 44-59 | 18.1 | 28-32 | 3.5 | 0 |
| 19 | 24.7 | 1.4-48 | 28.3 | 7.6-22 | 0.4 | 0-15.4 |
| 20 | 57.6 | 57-59 | 30.7 | 27-30 | 0.5 | 0-0.7 |
| 21 | 55.2 | 53-57 | 34.4 | 30-32 | 0.2 | 0-1.1 |
| 22 | 62.1 | 57-64 | 34.0 | 33-36 | 0.1 | 0-0.7 |
| 23 | 65.3 | 63-66 | 33.6 | 34 | 0.3 | 0-0.5 |
| | | | | | | |
| | | | ***Steady state operation at a 15 day HRT*** | | | |
| 27 | 55.4 | 51-57 | 32.1 | 25-37 | 1.2 | 0-0.9 |
| 28 | 62.8 | 60-64 | 34.6 | 34-36 | 0.0 | 0 |
| 29 | 64.6 | 62-67 | 34.0 | 33-36 | 0.0 | 0-0.6 |
| 30 | 65.3 | 64-66 | 32.1 | 33-35 | 0.0 | 0 |
| 31 | 65.0 | 64-67 | 33.6 | 27-34 | 0.4 | 0 |
| 32 | 66.0 | 61-66 | 33.6 | 32-35 | 0.3 | 0-1.5 |
| 33 | 64.8 | 62-66 | 32.5 | 32-34 | 0.2 | 0-0.8 |
| 34 | 63.3 | 58-66 | 34.0 | 25-35 | 0.0 | 0-0.9 |
| ***Average*** | **63.4** | | **33.3** | | **0.3** | |

## Removal Efficiency of the Anaerobic Pilot Scale Reactor

**Figure 5.1 Influent and Effluent Total Solids Over Time, and % Removal**

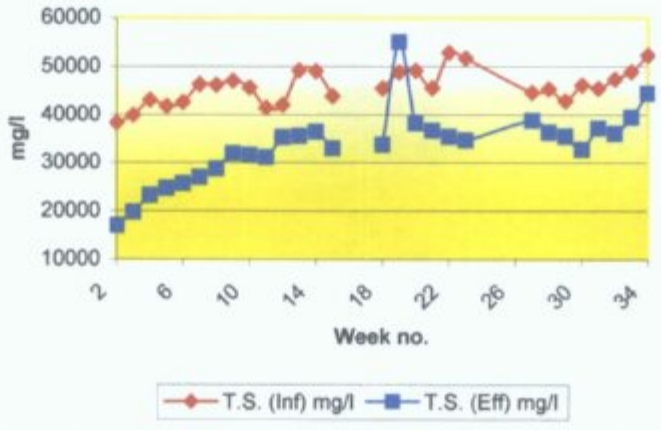



**Figure 5.2 Influent and Effluent Volatile Solids Over Time, and % Removal**

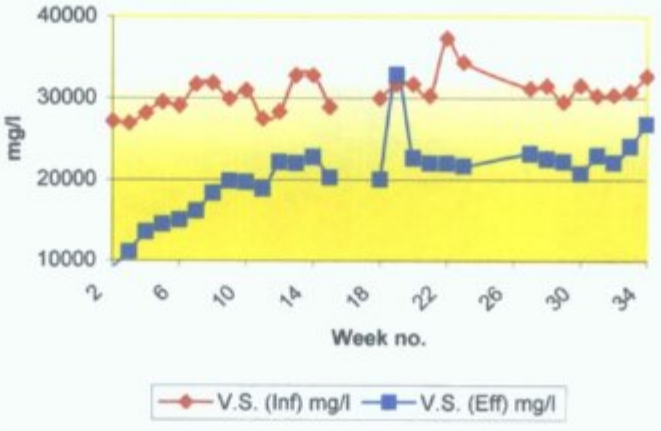
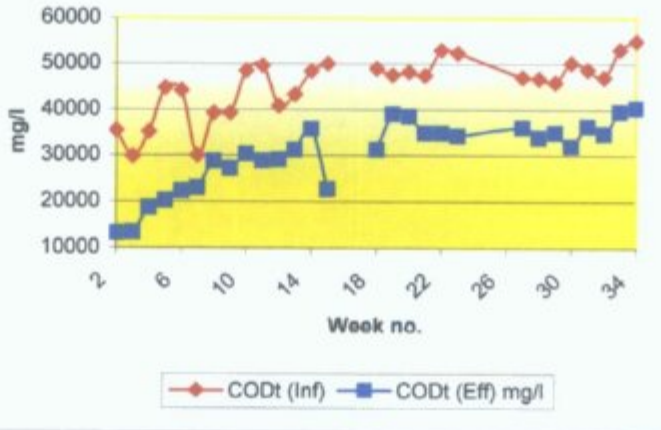
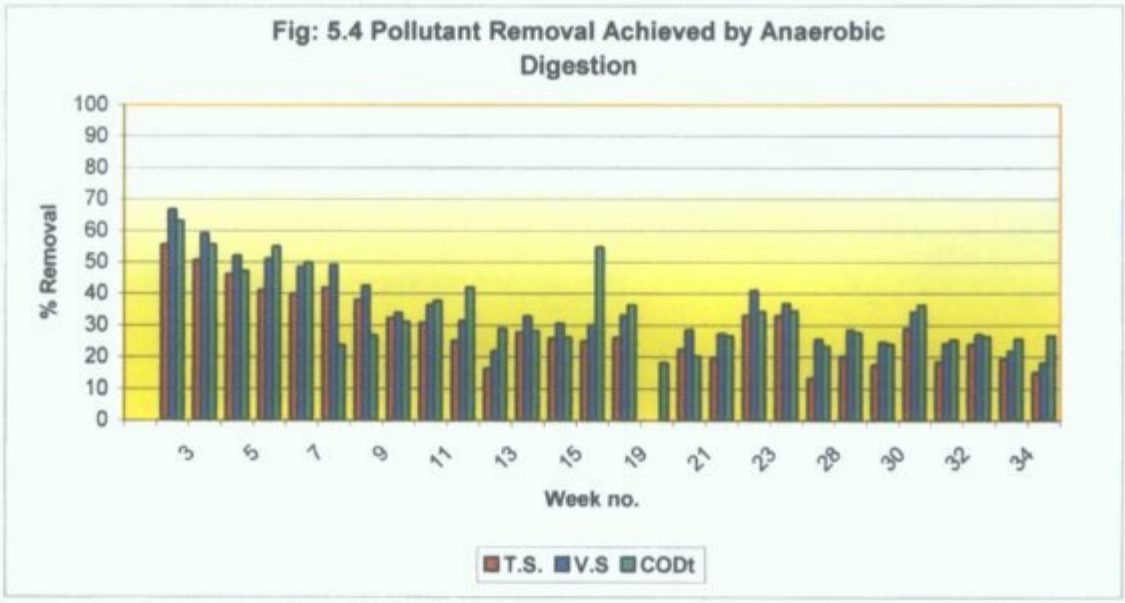



**Figure 5.3 Influent, Effluent CODt Verses Time**





## 5.1 Operational Performance of the Pilot Scale Anaerobic Digester

**a) Start – up**

The digester was seeded as per materials and methods section 4.1.3. The digester contents reached 35 °C (mesophilic temperature) overnight, the mechanical mixer was commissioned on day 3, set to operate for 15 min / hr at 60 r.p.m. The digester was allowed to acclimatise for the duration of week one. On week two, the digester was fed for the first time, at a mean OLR of 9.32 g VS/d, a VLR of 0.832 l/l/d and an HRT of 33 days.

During week two, the Total Solids, Volatile Solids and CODt of the effluent varied greatly. The T.S. ranged from 5990 – 18450 mg/l, V.S. from 1740 – 10244 mg/l, and total COD from 810 - 12340 mg/l, while pH averaged at 7.26 which is within the optimum range for the operation of anaerobic digesters. It was concluded that the variations in sample consistency was due to insufficient mixing and settlement. On the last day of week two, the mixing frequency was increased to 15 minutes every 30 minutes to eliminate settlement. Volatile Fatty Acid production was negligible, results ranging from 0 – 22 mg/l as acetic acid (average 8 mg/l), indicating that there was little hydrolysis of the feedstock, and that residual VFA in the seed sludge had been utilised (initially measured at 181 mg/l).

A rapid start-up was evident by the quality of biogas produced during this time, with the average $CH_4$ content being 62.5% by volume. The percentage COD removal during the week was 63%, Total Solids removal was 55.6%, and Volatile Solids removal was 66%. This apparent removal efficiency was due to the retention of solids in the digester rather than complete utilisation of the feedstock, which is clearly apparent from the variations in the consistency of effluent samples. pH decreased slightly to 7.19.

Week 3: Results continued to be variable, indicating that some settlement was still occurring. CODt, T.S. and V.S. removal of 55%, 50% and 59 % respectively, was achieved representing a decline slightly in comparison to week 2. pH again decreased, to 7.07.

Week 4: Mixing frequency was increased again, the mixer was allowed to run continuously at 60 r.p.m.

**b) Stable Operation at a 33 Day HRT**

Week 5: Alteration of the mixing frequency ensured that effluent analysis results became more consistent, covering a much narrower range. For example, Total Solids ranged from 24,252 mg/l to 25,064 mg/l. CODt removal was 54%, T.S. removal was 41% and V.S. were reduced by 50%. The average methane content of the biogas was at its highest so far (65.8% $CH_4$), indicating that the methanogenic bacterial population had acclimatised to the feedstock (having previously been fed a mixture of primary and secondary sludge in Tuallmore). VFA was low, averaging at just 10 mg/l, suggesting that once the fatty acids were made available they were utilised by the methanogens. pH decreased to a mean of 6.97. pH had decreased steadily since start-up. The most likely reason for this was that the bicarbonate alkalinity added at start-up was being partially removed during feeding, and not due to process instability, as VFA levels were consistently low.

Conditions within the reactor during week six were similar to week 5, though removal efficiency continued to decline slightly (COD removal was 49%, T.S. removal 39.8% and V.S 48%). The Methane content of the Biogas averaged at 64.1% with $CO_2$ at 30.2%.

Week 7: COD removal was significantly lower at 23%. This result can be explained by the fact that the influent had an unusually low COD of just 30,000 mg/l, which may be due to irregularities during the wastewater treatment process, (such as: operation at a long sludge age, or omission of a high strength industrial wastewater discharge from the influent stream during this time).

Overall, the digester was not adversely affected due to this low COD feed, as T.S. removal was 41.8% and V.S. removal was 49%. pH decreased slightly to 6.89, corresponding to an increase in VFA, which averaged at 70.8 mg/l. As the COD of the feedstock was so low, it is possible that residual undigested solids were also utilised during this period, as hydrolysis is the rate limiting stage of the anaerobic digestion process.

Week 8: pH increased to 6.97 and the average VFA was 60 mg/l. T.S. removal declined to 37.9% as did V.S. removal to 42%, while CODt removal was 26% The gas

composition was monitored during the week revealing $CH_4$ content of 60.2% and $CO_2$ of 34.5%.

For the duration of week nine, sodium bicarbonate was added to the feedstock at a rate of 5g/l, pH had declined in the weeks previous. Gas composition was recorded with an average $CH_4$ content of 61.5% and $CO_2$ of 35.5% . CODt removal was poor at 30.9%, T.S. removal was 32.1% and V.S. removal was 33%.

Week 10: Typically, the biogas contained 61.6% $CH_4$ and 34.6% $CO_2$. While COD removal increased to 37.6%, T.S. removal was 30.6% and V.S. removal was 36% .VFA increased to 128.5mg/l, and pH increased to 7.06 due to the addition of sodium bicarbonate for the duration of week 9. The increase in VFA may indicate that the methanogenic microbial consortium were under stress, but not to an extent that would cause concern, especially since the biogas composition was reflective of the previous weeks figure, the hydrolysis of retained solids may also have been a contributing factor.

WEEK 11: The $CH_4$ content of the biogas increased to 63.8% and $CO_2$ of 34% indicating re-stabilisation of synergistic bacteria. COD removal improved to 41.9% , but T.S. removal was poor at 24.9% and V.S. was 31%. pH remained stable at 7.06.

Week 12: The $CH_4$ increased to 64.3% and $CO_2$ of 35%. The COD removal, was 28%, T.S. removal was at 16.2% and V.S. of 21.7% was achieved. pH remained at 7.06.

Week 13: The biogas was composed of 65% was methane and 33.6 % $CO_2$. COD removal was only 28%, T.S. removal was 27.7% and V.S. removal was 32%. VFA had risen slightly (from 128.5 mg/l week 10 to 134.2 mg/l); this had a knock on effect on pH, which decreased to 6.98.

Week 14: CODt removal was 26%, T.S. removal was 25% and V.S. removal was 30.5%, the methane content of the biogas was normal at 64.8% and $CO_2$ of 29.8%.

**c) Operational issues: Mixing and Air shock**

During week 15, effluent samples varied in consistency from very thin to extremely thick, Total Solids ranged from 20,871 mg/l to 61,103 mg/l throughout the course of the week, therefore all parameters varied greatly. Volatile Solids ranged from 12,826 mg/l to 37,579 mg/l and COD total from 19,230 mg/l to 27,545 mg/l. VFA results were

of a narrower range 176 mg/l to 228 mg/l (average 202.2 mg/l). The composition of the biogas fell to 57.8% $CH_4$ , $CO_2$ of 31.7% and $O_2$ of 0.17%. This left 10.4% of the biogas unaccounted for, which may be attributed to water vapour or due to a shift in the population of bacteria present, possibly due to competition between sulphite reducing bacteria and the methanogens. Due to variable effluent samples the average COD removal was 54.71% T.S was 24.8% and V.S was 29.87% . By the end of the week, stratification within the digester became apparent. Visually, only the bottom half of the digester was actively mixing. The top half was thick and gel - like in appearance with gas pockets visible. For the duration of weeks 16 and 17 the digesters contents was recirculated from bottom to top using a 'Masterflex' peristaltic pump at a rate of 140 ml/min (the tubing was primed prior use to prevent air from entering the digester), to enhance contact time between the active biomass and the retained solids at the top of the digester, facilitating the utilisation of retained solids. biogas composition didn't vary greatly from that of the previous week ($CH_4$ 58%, $CO_2$ 32.6% and $O_2$ 0.28%). Samples were taken throughout week 17 for pH and VFA, as feeding was to recommence when VFA reduced. The pH was 6.97 while VFA averaged at 207 mg/l, which actually increased since the last recording week 15 (202.2 mg/l) this indicates that undigested residual solids which had accumulated throughout the trial were being utilised during non feeding also confirming slow hydrolysis of secondary solids and ineffective mixing. The gas composition for this week was as follows: 57.8% $CH_4$, 30% $CO_2$ and 0 % $O_2$.

Week 18: The digester was fed as normal and recirculated for the duration of the week. VFA reduced to 182 mg/l and pH increased to 7.07 this is an indication that recirculation aided in digestion allowing maximum contact between the anaerobic bacterial population and the incoming secondary solids. The COD removal was 36.2% T.S. removal was 26% and V.S. removal was 33% which was an improvement on the results achieved from week 15. The $CH_4$ content of the biogas decreased to 52.1%, CO2 was 30 % and O2 was 0%, one explaination for this result is that the methanogenic bacteria were acclimatising to conditions imposed by recirculation (i.e. improved contact between the bacteria and the feedstock, thus a higher concentration of feed would have been made available to the bacteria). Having recorded irregularities in biogas composition for the second time, once off analysis for total sulphur was conducted to determine if sulphite reducing bacteria were competing with the

methanogens for substrate. This was not the case however, as results yielded 0 mg/l Total sulphur.

Week 19: In week 19 there was an air shock caused by cracked tubing, which resulted the displacement of 2 lt. of the digesters contents, visually the digesters contents appeared grey in colour with the surface scum being almost white (Appendix B).

For the duration of the week, the digester was fed, but no effluent was removed to replace the quantity displaced. The feeding regime and biogas composition for this week can be viewed in Appendix C. An effluent sample was taken mid week to assess the effect. The results were as follows: pH was 7.06, this was attributed to the dispersion of $CO_2$ gas from the digester (which is one of the reasons for the rapid analysis of pH when a sample is taken from a active digester), VFA increased to 255 mg/l indicating that the methanogens were inhibited to some degree by the oxygen shock, the extent of which will become more apparent with time. COD removal was 17.9%. Total Solids exiting the digester was higher than in the influent (Feed T.S. 48,804 mg/l while the effluent T.S. was 54,972 mg/l) thus V.S. concentration was similar (Feed V.S 31,725 mg/l and effluent V.S was 32,852 mg/l). At this early stage, the air shock seemed to have cause flotation of solids. The average methane content of the biogas was 24% (but ranged from 1.4 to 48 % by the end of the week), $CO_2$ 18.12 % (ranged from 7.6 to 20%) and $O_2$ 3.5% (ranged from 1.2% to 15.4 ). By the end of the week the methanogenic population appeared to be re-acclimatising as the methane content of the biogas was increasing on a daily bases.

Week 20: The overhead stirrer was used as the sole means of digester mixing. The methane content increased to an average of 57.6%, 30.7 % $CO_2$ and 0.4% $O_2$, pH was 7.01 and VFA decreased to 189 mg/l indicating that the bacterial population were quickly recovering from the air shock. COD removal was 20.1%, T.S. removal was 22.3% and V.S. removal of 28.4% was achieved.

Week 21: The percentage methane decreased to 55.2% and $CO_2$ increased to 34.4% suggesting that acetogentic bacteria were the most dominant form at this time, and that the methanogenic bacteria were more adversely affected. COD removal efficiency improved at 26.5%, T.S. removal was 19.5% And V.S. removal was 27%. The pH was almost neutral at 7.02 and VFA again decreased to 176 mg/l on average.

Week 22: At this stage of the study the HRT was decreased to 15 D for two reasons: to increase the flow through the reactor thereby removing inactivated microbes (from the air shock) and secondly to be more representative of full-scale operation. Typically, the OLR was 27.84g VS/D and VLR was 2.48 L/L/D. With immediate effect, the methane content increased to 62.1%. VFA reduced to 150 mg/l despite pH decreasing to 6.97, COD removal also improved to 34.2%, T.S to 33% and V.S. removal to 40.9 %.

Week 23: This was the final week of feeding prior a shut down period for Christmas. For the duration of the week the methane content increased to 65.2% and $CO_2$ was 33.6%. pH was 7.01 and VFA again decreased to 138.5 mg/l. CODt removal and T.S. removal was almost identical to the previous week (34.4% and 32.9 % respectively) but V.S. removal decreased to 36%. At the end of the week the digester was shutdown (Heating, feeding and mixing ceased) for a 3-week period, during this period the temperature of the reactor decreased to a low of 6.2 $^{o}$C. The shut down period, would provide valuable information regarding the flexibility of anaerobic digestion.

Week 27: On the first day back after the holiday period the digester was heated. The following day the temperature was up to 35 $^{o}$C and mixing was re-commissioned. Feeding commenced on the third day, the OLR, VLR and HRT was maintained as before shutdown.

Stratification was apparent in the digester, to a much greater extent than that noticed on week 15. In this instance, there were four zones visible. At the top of the reactor there was scum, dark brown in colour with a volume of 2.2 L., under this there was black zone volume was 2 L., with visible gas pockets, (presumably this was the methanogens niche). Then there was a brown zone of fine sludge, with no gas pockets or visible mixing, the volume here was 2 L., and finally, at the bottom, the sludge was similar in appearance to the previous zone but was actively mixing, the volume here was 5 L. Prior to initial feeding after shutdown, duplicate samples were taken at varying stages of the reactor (top, middle and bottom, see Appendix D), at the top of the digester the T.S. was 6.2%, in the middle T.S. was 4.1% and at the bottom T.S. was 4.3%. At a glance, these results would illustrate that at the top of the reactor light floatable matter had accumulated, the middle of the digester was the active zone of degradation, and at the bottom fixed solids may have accumulated. However VFA analysis suggests a different situation. At the bottom of the digester, where the feed is administered, the VFA

concentration was the highest at 589 mg/l indicating that this is the zone of hydrolysis, acidogenesis, and acetogenesis, as VFA are the intermediate by –products. In the middle of the digester, the VFA concentration decreased to 326 mg/l representing a transition or cross – over of the bacterial population as the substrate moves towards the methanogens. At the top of the vessel the VFA concentration was considerably lower at 268 mg/l, indicative of the utilisation of VFA and their conversion to methane.

The sludge from the top of the reactor was gel -like in texture, which suggests that the polyelectrolyte used to dewater the sludge at Manorhamilton sewage treatment plant was accumulating in the reactor. A product data sheet for the polyelectrolyte in question "Zetag 78N" from General Chemicals LTD. in Dublin, states that *" Zetag while biodegradable, the rate of degradation is relatively slow. Assessed by means of the closed bottle tests they tend to give a measured BOD (within 28 days) which is about 40 – 60% of their COD, thus it is not rated as being readily biodegradable"* (Ciba Specialty Chemicals P/L, 1998). In a study conducted on the effect of polymer addition on granulation in an anaerobic baffled reactor, a range of polyelectrolytes were tested for specific methanogenic activity tests, the results of this study revealed Zetag to be more inhibitory than a number of other commercially available polymers (Uyanik, Sallis and Anderson, 2001).

Week 27: Due to the zonation within the digester and consistency of the effluent, effluent could only be removed from the middle porthole; this was the case up until the second last week of operation. During week 27, the mean CODt removal was 13.2% T.S. removal was 25% and V.S. removal 23%. These results reflect the three week shut-down. It is most likely that after a period of dormancy the anaerobic bacterial population required time to re-acclimatise. The pH remained neutral, despite VFA being 221 mg/l.

**d) Stable operation at a 15D HRT**

During week 28, the methane content of the biogas improved to 62.8% and $CO_2$ was 34.6%, which was a marked improvement on the previous week. COD removal was 27.5% (ranged from 31,032 to 36,210 mg/l), T.S. removal was 20% (ranged from 33776 to 37143 mg/l) and V.S. removal was 28% (ranged from 21,182 to 23,320 mg/l),and there was a significant reduction in VFA to 162.6 mg/l.

Week 29: The methane content of the biogas also increased to an average of 64.6% by volume and $CO_2$ was 34%. COD removal was 23.8% (ranged from 33,408 – 35,500 mg/l), T.S. removal was 17.4% (ranged from 33,776 – 37,143 mg/l), and V.S. removal was 24.5% (ranged from 21,182 – 23,320 mg/l). VFA increased to 242 mg/l indicating that there was an increase in the rate / extent of hydrolysis and that the acetogenic bacteria were more active in comparison to the methanogens, while pH decreased slightly reflective of VFA production to 6.94.

Week 30: CODt removal increased to 36.2% (range 31,080 – 34,500 mg/l) T.S. removal increased to 28.9% (range 30,374 to 34,841 mg/l) and V.S. removal increased to 34% (range 19,269 to 22,479 mg/l). The Biogas was composed of 65.3% methane and 32.1% $CO_2$. These results indicate that the methanogens had become more active. VFA was reduced to 216.5 mg/l.

On week 31 COD removal decreased to 25% (ranged from 31,140 to 41,160 mg/l) T.S. removal was 18.3% (ranged from 35,056 to 40,570 mg/l) and V.S. was 24.1% (ranged from 21,157 to 26,130 mg/l). Again due to mixing difficulties large variations were noted in the consistency of the effluent samples. Biogas production consisted of 65% methane and 33.6% $CO_2$. The pH remained almost neutral, but VFA increased to 269.5 mg/l which is the highest value for VFA to date. This was due to the hydrolysis of retained solids and not process instability as the methane content of the biogas was high.

Week 32: COD removal was 26% (range 31,560 to 38,610 mg/l) T.S. removal was 23% (range 35,155 to 37,642 mg/l) and V.S. removal was 27% (range 21,690 to 22,961 mg/l). VFA decreased to 201 mg/l and pH decreased to 6.85.

Week 33: There was no significant change in digester performance during week 33, except for the fact that the Total Solids concentration of the effluent increased ranging from 37,298 to 42,903 mg/l, the average being 39,447 mg/l. Again, the methane content of the biogas was high at 64.8% and $CO_2$ of 32.5%. VFA was 196.5 mg/l and pH increased to 7.05

Table 5.4 Feeding Regime and Feedstock Compositon (Aerobic Thermophilic Pilot Scale Reactor)

| Week | OLR g.vs/d | VLR g.vs/l/d | HRT (days) | T.S. mg/l | T.S. % | V.S. mg/l | V.S. % | F.S. mg/l | F.S. % | CODt mg/l | CODs mg/l | pH |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 1 | | | | | | | | | | | | |
| 2 | 12-18 | 1.28 | 30 | 57253 | 5.7 | 38537 | 67.23 | 18716 | 32.17 | 59307 | 647 | 6.92 |
| 3 | 13.26 | 1.40 | 30 | 62142 | 6.18 | 41967 | 67.52 | 20173 | 32.59 | 606040 | 777 | 6.60 |
| | **12.72** | **1.34** | **30** | **59698** | **5.94** | **40252** | **67.38** | **19445** | **32.38** | **332673** | **712** | **6.76** |
| 4 | 28.03 | 2.95 | 12 | 52967 | 5.29 | 35432 | 66.86 | 17536 | 33.14 | 62720 | 573 | 6.85 |
| 5 | 28.39 | 2.99 | 12 | 51960 | 5.35 | 35888 | 68.97 | 16072 | 30.03 | 49520 | 172 | 6.93 |
| 6 | 31.02 | 3.27 | 12 | 58461 | 5.85 | 39220 | 67.21 | 23502 | 32.78 | 59453 | 626 | 6.91 |
| 7 | 33.59 | 3.54 | 12 | 63533 | 6.35 | 42460 | 66.84 | 21073 | 33.16 | 60686 | 586 | 6.50 |
| | **30.26** | **3.19** | **12** | **56730** | **5.71** | **38250** | **67.47** | **19546** | **32.28** | **58095** | **489** | **6.80** |
| 8 | 33.02 | 3.48 | 12 | 62240 | 6.22 | 41741 | 67.07 | 20499 | 32.92 | 61908 | 629 | 6.80 |
| 9 | 33.79 | 3.56 | 12 | 61721 | 6.16 | 42714 | 69.20 | 19007 | 30.79 | 62880 | 797 | 6.90 |
| 10 | 31.65 | 3.33 | 12 | 58490 | 5.83 | 40010 | 68.37 | 18480 | 31.62 | 60720 | 781 | 7.01 |
| 11 | 31.15 | 3.28 | 12 | 58143 | 5.8 | 39380 | 67.73 | 18762 | 32.26 | 58160 | 567 | 6.97 |
| | **32.40** | **3.41** | **12** | **60148** | **6.00** | **40961** | **68.09** | **19187** | **31.90** | **60917** | **694** | **6.92** |
| 12 | 31.00 | 3.26 | 12 | 58228 | 5.82 | 39191 | 67.30 | 19037 | 32.70 | 59360 | 487 | 7.0 |
| 13 | 33.15 | 3.49 | 12 | 61040 | 6.09 | 41914 | 68.67 | 16443 | 31.32 | 61600 | 646 | 6.74 |
| 14 | 33.82 | 3.56 | 12 | 61939 | 6.19 | 42754 | 69.02 | 19185 | 30.97 | 63920 | 840 | 6.81 |
| 15 | 37.78 | 3.98 | 12 | 60590 | 6.04 | 47757 | 68.81 | 18817 | 31.18 | 64080 | 1053 | 7.04 |
| 16 | 32.86 | 3.46 | 12 | 61005 | 6.06 | 41543 | 68.10 | 19462 | 31.90 | 65120 | 1120 | 6.95 |
| | **33.72** | **3.55** | **12** | **60560** | **6.04** | **42632** | **68.38** | **18589** | **31.61** | **62816** | **829** | **6.91** |
| 17 | 40.31 | 4.24 | 10 | 61349 | 6.14 | 42434 | 69.09 | 18964 | 30.91 | 67105 | 1156 | 7.02 |
| 18 | 40.73 | 4.29 | 10 | 62232 | 6.23 | 42879 | 68.91 | 19353 | 31.10 | 68200 | 1302 | 6.97 |
| 19 | 40.57 | 4.27 | 10 | 61499 | 6.15 | 42708 | 69.45 | 18791 | 30.55 | 66900 | 1752 | 7.02 |
| 20 | 40.30 | 4.24 | 10 | 60716 | 6.07 | 42419 | 69.87 | 18296 | 30.13 | 66240 | 2021 | 6.92 |
| 21 | 39.76 | 4.19 | 10 | 60719 | 6.07 | 41853 | 68.94 | 18866 | 30.95 | 58400 | 1525 | 7.00 |
| | **40.33** | **4.25** | **10** | **61303** | **6.13** | **42459** | **69.25** | **18854** | **30.73** | **65369** | **1551** | **6.99** |
| 22 | 50.33 | 5.30 | 8 | 61191 | 6.12 | 42398 | 69.29 | 18793 | 30.71 | 60500 | 1430 | 6.93 |
| 23 | 50.60 | 5.33 | 8 | 61721 | 6.17 | 42628 | 69.07 | 19093 | 30.93 | 65320 | 1345 | 6.98 |
| 24 | 50.49 | 5.31 | 8 | 61234 | 6.13 | 42532 | 69.41 | 18752 | 30.60 | 71600 | 1641 | 6.90 |
| 25 | 50.20 | 5.28 | 8 | 61403 | 6.14 | 42291 | 68.88 | 19111 | 31.12 | 73360 | 2031 | 6.87 |
| 26 | 50.19 | 5.38 | 8 | 61304 | 6.13 | 42484 | 69.30 | 18820 | 30.70 | 74021 | 2140 | 6.88 |
| | **50.36** | **5.32** | **8** | **61370** | **6.14** | **42467** | **69.19** | **18914** | **30.81** | **68960** | **1717** | **6.91** |

**Table 5.5 (A) : Aerobic Thermophilic Pilot Digester Effluent Analysis Results During Start - up**

| | Week | pH | pH Range | CODt mg/l | CODt Range | CODs mg/l | CODs Range | T.S mg/l | T.S Range | V.S mg.l | V.S % | V.S Range | Alkalinity |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Mesophilic Temperature Zone | 1 | 6.99 | 6.94 - 7.03 | 51573 | 52080 - 52406 | 772 | 700 - 818 | 49974 | 48727 - 50968 | 33857 | 67.75 | 33183 - 34561.5 | 4493 |
| | 2 | 6.97 | 6.97 - 7.01 | 50280 | 49120 - 53440 | 541 | 409 - 723 | 48109 | 47633 - 48991 | 31988 | 66.49 | 31409 - 32397 | 4542 |
| | 3 | 7.26 | 6.64 - 7.85 | 43260 | 41680 - 46000 | 1369 | 738 - 1952 | 44641 | 42072 - 46823 | 29166 | 65.33 | 26011 - 33090 | 4721 |
| **Mean** | | **7.07** | | **48371** | | **894** | | **47575** | | **31670** | **66.57** | | **4586** |
| Thermotolerant Temperature Zone | 4 | 7.85 | 7.74 - 7.97 | 39768 | 37640 - 42690 | 1530 | 1422 - 1942 | 41892 | 40290 - 43676 | 25353 | 60.52 | 24038 - 26513 | 4672 |
| | 5 | 7.70 | 7.64 - 7.76 | 36896 | 35840 - 37400 | 1671 | 1125 - 2085 | 41559 | 38411 - 43294 | 25253 | 60.76 | 23039 - 26401 | 4640 |
| | 6 | 7.59 | 7.52 - 7.67 | 32124 | 29230 - 34840 | 903 | 723 - 1240 | 36371 | 34218 - 40496 | 22000 | 60.49 | 20460 - 24359 | 4737 |
| | 7 | 7.41 | 7.32 - 7.52 | 35840 | 30040 - 40640 | 1498 | 933 - 2796 | 47784 | 39745 - 50926 | 28490 | 59.62 | 28571 - 30206 | 4624 |
| **Mean** | | **7.64** | | **36157** | | **1400** | | **41902** | | **25274** | **60.32** | | **4668** |
| Thermophilic Temperature Zone | 8 | 7.20 | 7.05 - 7.35 | 41557 | 39160 - 44560 | 1123 | 861 - 1374 | 51739 | 49694 - 53752 | 30112 | 58.20 | 29263 - 31414 | 5047 |
| | 9 | 7.36 | 7.3 - 7.35 | 40985 | 40360 - 42120 | 1238 | 930 - 1815 | 50611 | 50317 - 51091 | 29709 | 58.70 | 28435 - 30451 | 4965 |
| | 10 | 7.37 | 7.35 - 7.38 | 42003 | 40780 - 42600 | 966 | 856 - 1128 | 50205 | 49878 - 50415 | 29173 | 58.11 | 28830 - 29402 | 4835 |
| | 11 | 7.40 | 7.39 - 7.42 | 43068 | 41320 - 45760 | 1141 | 870 - 1395 | 50638 | 50259 - 51395 | 30011 | 59.27 | 29728 - 30632 | 4754 |
| **Mean** | | **7.33** | | **41903** | | **1117** | | **50798** | | **29751** | **58.57** | | **4900** |

**Table 5.5 (B): Aerobic Thermophilic Pilot Digester Effluent Results During Variations of HRT**

| HRT | Week | pH | pH Range | CODt mg/l | CODt Range | CODs mg/l | CODs Range | T.S mg/l | T.S Range | V.S mg.l | V.S % | V.S Range | Alkalinity |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 12 | 12 | 7.45 | 7.4 - 7.5 | 42730 | 41520 - 44160 | 1176 | 956 - 1395 | 49812 | 49045 - 50392 | 29294 | 58.81 | 28723 - 30257 | 4868 |
| 12 | 13 | 7.43 | 7.41 - 7.47 | 42840 | 41640 - 43720 | 955 | 885 - 1134 | 50562 | 49793 - 50867 | 29911 | 59.16 | 29113 - 30419 | 4672 |
| 12 | 14 | 7.44 | 7.41 - 7.5 | 42697 | 42120 - 43120 | 1067 | 930 - 1172 | 50735 | 49785 - 51787 | 29766 | 58.67 | 29108 - 30465 | 4689 |
| 12 | 15 | 7.43 | 6.86 - 7.72 | 36933 | 31280 - 42320 | 1788 | 1196 - 2640 | 42476 | 39095 - 50862 | 25339 | 59.65 | 20250 - 30107 | 3565 |
| 12 | 16 | 7.49 | 7.41 - 7.55 | 35639 | 35040 - 35920 | 1300 | 1166 - 1392 | 41872 | 40773 - 42348 | 25008 | 59.73 | 24237 - 25451 | 4235 |
| **Mean** | | **7.45** | | **40976** | | **1124** | | **48245** | | **28495** | **59.06** | | **4623** |
| 10 | 17 | 7.51 | 7.48 - 7.54 | 38317 | 38000 - 38722 | 1430 | 1390 - 1798 | 44506 | 43630 - 45553 | 27202 | 61.01 | 25385 - 27641 | 4530 |
| 10 | 18 | 7.48 | 7.43 - 7.55 | 41487 | 40352 - 42240 | 1314 | 1235 - 1432 | 46493 | 45228 - 47292 | 27872 | 59.95 | 27094 - 28649 | 4710 |
| 10 | 19 | 7.53 | 7.49 - 7.55 | 41287 | 41185 - 41425 | 1307 | 1270 - 1370 | 48036 | 47126 - 48991 | 24471 | 50.94 | 27968 - 29791 | 4645 |
| 10 | 20 | 7.75 | 7.73 - 7.77 | 41579 | 40440 - 42360 | 1302 | 1198 - 1398 | 48451 | 47017 - 49942 | 29170 | 60.20 | 28267 - 29744 | 4589 |
| 10 | 21 | 7.75 | 7.71 - 7.77 | 41438 | 37400 - 42432 | 1369 | 1176 - 1803 | 49048 | 48299 - 49942 | 29498 | 60.14 | 29137 - 30160 | 4645 |
| **Mean** | | **7.604** | | **40822** | | **1344** | | **47307** | | **26743** | **56.53** | | **4624** |
| 8 | 22 | 7.75 | 7.73 - 7.77 | 42765 | 42492 - 43210 | 1329 | 1160 - 1250 | 48022 | 46777 - 51950 | 29245 | 60.90 | 28367 - 30128 | 4733 |
| 8 | 23 | 7.75 | 7.72 - 7.76 | 45535 | 44560 - 46020 | 1249 | 1182 - 1291 | 50076 | 46421 - 51950 | 30336 | 60.58 | 28480 - 31009 | 4861 |
| 8 | 24 | 7.54 | 7.51 - 7.55 | 45961 | 45792 - 46250 | 1273 | 1256 - 1301 | 51658 | 51454 - 51950 | 31239 | 60.47 | 31038 - 31467 | 4732 |
| 8 | 25 | 7.52 | 7.49 - 7.55 | 45870 | 45094 - 47520 | 1142 | 1004 - 1263 | 51730 | 51344 - 52386 | 32030 | 61.92 | 31343 - 32319 | 4565 |
| | 26 | 7.55 | 7.51 - 7.57 | 50681 | 50026 - 51986 | 1095 | 1002 - 1166 | 53072 | 52610 - 53919 | 32360 | 60.97 | 31886 - 33262 | 4621 |
| **Mean** | | **7.622** | | **46162** | | **1218** | | **50912** | | **31042** | **60.97** | | **4702** |

* : Results omitted from weekly average

**Table 5.6: Aerobic Pilot Digester Operational Conditions**

| Week | HRT | DO | Temp | Min | Max | S.D. |
|---|---|---|---|---|---|---|
| 1 | | 1.99 | 16.63 | 15.74 | 17.7 | 1.06 |
| 2 | | 2.55 | 15.5 | 14.57 | 16.35 | 0.65 |
| 3 | | 2.71 | 28.64 | 1.57 | 32.1 | 5.47 |
| | **30** | **2.42** | **20.26** | **15.34** | **22.05** | **2.39** |
| 4 | | 2.3 | 47.59 | 46.9 | 48.1 | 0.43 |
| 5 | | 2.17 | 47.35 | 46.34 | 48.04 | 0.63 |
| 6 | | 2.11 | 48.20 | 47.17 | 49.12 | 0.60 |
| 7 | | 2.41 | 48.89 | 48.10 | 48.80 | 0.27 |
| | **12** | **2.25** | **48.01** | **47.13** | **48.52** | **0.48** |
| 8 | | 1.35 | 53.09 | 52.60 | 53.40 | 0.30 |
| 9 | | 2.20 | 52.98 | 52.48 | 53.65 | 0.40 |
| 10 | | 2.50 | 53.02 | 51.76 | 53.63 | 0.61 |
| 11 | | 2.25 | 52.96 | 50.95 | 53.63 | 0.95 |
| | **12** | **2.08** | **53.01** | **51.95** | **53.58** | **0.57** |
| 12 | | 2.50 | 53.00 | 52.39 | 53.63 | 0.53 |
| 13 | | 2.30 | 52.64 | 50.97 | 52.93 | 0.59 |
| 14 | | 2.40 | 52.55 | 50.70 | 53.60 | 1.08 |
| 15 | | ND | 55.93 | 50.86 | 60.80 | 3.60 |
| 16 | | ND | 53.21 | 51.67 | 53.77 | 0.43 |
| | **12** | | **53.47** | **51.32** | **54.95** | **1.25** |
| 17 | | ND | 52.96 | 51.57 | 53.73 | 0.65 |
| 18 | | ND | 52.94 | 51.14 | 53.84 | 0.81 |
| 19 | | ND | 52.80 | 50.45 | 53.80 | 1.04 |
| 20 | | ND | 52.52 | 50.05 | 53.80 | 1.18 |
| 21 | | ND | 52.74 | 50.98 | 53.44 | 0.79 |
| | **10** | | **52.79** | **50.84** | **53.72** | **0.89** |
| 22 | | ND | 52.84 | 51.87 | 53.68 | 0.63 |
| 23 | | ND | 52.73 | 51.45 | 53.53 | 0.67 |
| 24 | | ND | 52.96 | 52.60 | 53.70 | 0.38 |
| 25 | | ND | 52.89 | 51.90 | 53.60 | 0.48 |
| 26 | | ND | 52.72 | 52.65 | 54.00 | 0.50 |
| | **8** | | **52.83** | **52.09** | **53.70** | **0.53** |

* Air compressor failure

## Removal Efficiency of the Thermophilic Aerobic Pilot Scale Reactor

Figure 5.5 Influent and Effluent Total Solids Over Time and % Removal

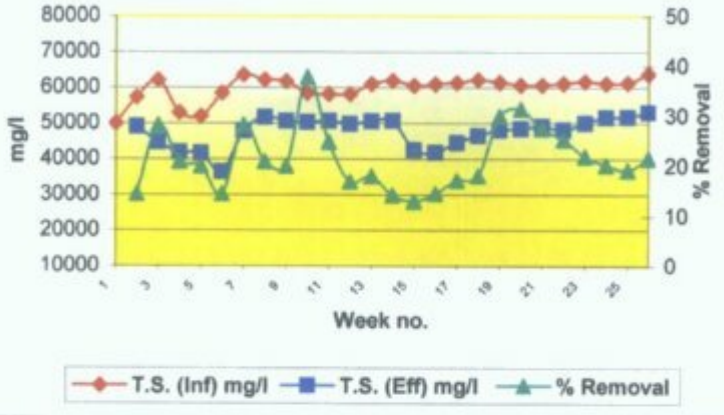



Figure 5. 6 Influrent and Effluent Volatile Solids Over Time, and % Removal

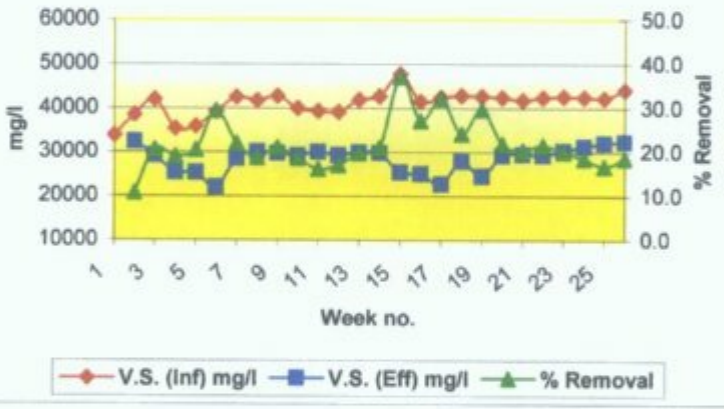
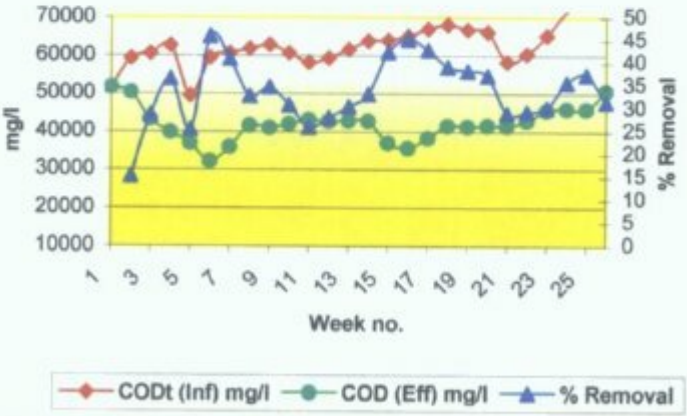



Figure 5.7 Influent and Effluent CODt Over Time, and % Removal



## 5.2 Operational Performance of the Thermophilic Aerobic Pilot Scale Reactor

### (Start – up) Operation at a Mesophilic Temperature

The start – up process of a full scale ATAD is rapid, reaching a temperature of 55 °C in less than 14 days (Kelly, 1995), and does not require the transplantation of active biomass from an operational reactor to the new system. The pilot scale reactor in this case was seeded, in accordance with recommendations made by Kelly (1995). The pilot reactor containing 10.5 L. of activated sludge at 5% T.S. was then sealed, mixed and aerated. The D.O. at this stage was maintained at 2.1 mg/l and the temperature was 16.6 °C on average. The digester was not fed for the first week of operation, and was monitored closely. Samples were taken for T.S., V.S. and COD reduction, while temperature was measured at intervals of 30 min. During the first week of operation the digesters contents became more concentrated on a daily basis due to evaporation, enhanced by the force of aeration and mixing. CODt increased from 4.87 to 5.17%. While the general mass of V.S. correspondingly increased from 33,183 – 34,605 mg/l during this time, the actual percentage of V.S. decreased from 68 – 66.9% of the T.S., within the initial five days of operation. This decrease indicates that a small quantity of the solids were being utilised by the aerobic biomass, during this period of acclimatisation and non – feeding. Microbial activity is further evident, when the CODs concentration is taken into account, as CODs decreased from 884 to 723 mg/l. The CODs, free and available for metabolism, has most likely stemmed from that originally present in the seed sludge, rather than newly produced (by hydrolysis) soluble matter as the extent of hydrolysis would have been limited.

In the second week of operation, the digester cooled slightly and the temperature averaged at 15.5 °C. Feeding commenced at a relatively high HRT of 30 days and a total solids concentration of 5.7%, as suggested by Kelly (1995). Evaporation was apparent on a daily basis but was minimal, at approximately 150 to 200 ml/day. Water was added to the digester, proportionate to the quantity evaporated and the digester was left undisturbed, aerating and mixing for 5 minutes prior to effluent removal and feeding. This procedure was continued until reactor shutdown. Results of analysis were similar to the previous week. pH was neutral and alkalinity was 4542 mg/l. Once again the CODs decreased to 540.5 mg/l on average. T.S. was reduced by 16%, V.S. was reduced by 17.2% and CODt removal was 12%. The actual removal at this stage is

almost certainly negligible as the CODs decreased significantly indicating that there was little hydrolytic action occurring and that the microbial population were dependent on readily available CODs to sustain life.

It was apparent by week 3 that the reactor was not going to self heat, as the reactor should have increased in temperature by 2 –3 °C / day whilst unfed, and increased in temperature by a further 8 – 10 °C overnight once feeding has commenced at a low feed rate (Kelly, 1995). This did not occur however and it was decided to apply heat to the vessel to achieve thermophilic conditions. A water bath was employed to circulate the warm water at 32 °C around the double wall of the reaction vessel. The contents of the reactor increased to 28 °C, which is still within the mesophilic range.

This initial temperature rise was slow in order to avoid thermal shock to the microbes. In order to limit heat loss from the system, the vessel was stripped of its original lagging and a new lagging jacket of Rockwool at 6 inch thickness surrounded by black plastic was fitted securely to the reactor. Auto-heating did not occur, but there were a number of noticeable effects following the application of heat to the reactor. For instance, the pH of the effluent increased to 7.26 (ranging from 6.64 – 7.85) coupled with a marked increase in $CaCO_3$ to 4721.20 mg/l (due to metabolism –generated alkalinity). pH is a feature of thermophilic conditions, as the temperature achieved during digestion suppresses nitrification within the reactor, ensuring that the pH remains relatively alkaline in nature (U.S. EPA, 1990). It is possible that exothermic conditions were present at this stage, and were masked by the cooling effects of aeration and poor heat retention. Hydrolysis of the sludge was apparent for the first time since start-up, as the CODs increased from 738 (week 2)– 1952 mgl (week 3). It appears that the increase in temperature stimulated the microbial population thereby enhancing hydrolysis of the feed sludge.

There was a significant increase in the rate of evaporation during this week. To prevent this, a condensation trap was installed to collect the condensate returning it to the vessel, at various periods during a 24 hr period, by timed pump. The quantities lost by evaporation increased from 150 – 200 ml the previous weeks to 2 L. on average, during this time, which may also be indicative of heat generation through respiration and metabolism. The condensate trap consisted of a tube from the lid of the reactor

connected to a Bunchner flask, which was sitting in a bucket of cold water. The condensate was analysed for CODt, which was found to be approximately 80 mg/l. T.S. removal during week 3 was 28.2 %, V.S. 30.5 % and CODt was 28.6%, which has almost doubled that of the previous week.

**(Start up): Operation in the Thermotolerant Temperature Range**

In the following week of operation (week 4) the feed rate to the digester was increased to 28.03g V.S./d, with a reduction in HRT from 30 days to 12 days. The temperature of the reactor was brought into the thermotolerant zone at 47.59 °C. In response to the temperature increase, the D.O. concentration fell from 2 mg/l on average to 0.5 mg/l and below. The air compressor was changed from the Stanley unit (capacity of 25 L.) to a 170 L. capacity unit as the Stanley compressor was at its maximum output. ATAD's have been known to operate at dissolved oxygen concentrations, as low as 0.5 mg/l (Pressley, 2001). However, it was decided in this case to keep the D.O. greater than 2 mg/l, to prevent odours and anoxic conditions, which are thought to occur following feed rate adjustments and temperature increases (Jowill *et al.*, 2002).

Condensation increased with the higher temperatures and the condensate trap was redesigned to trap the evaporate. A glass distillation column was fitted to the lid of the reactor, held in place by two clamps. Cold tap water flowed through the column on a continuous basis. This improved the situation significantly. For the remainder of the trial, 2 L. of water was added to the reactor daily by timed dosing pump.

For the period of operation in the thermotolerant temperature zone, the pH of the digester effluent averaged at 7.64. The extent of hydrolysis improved, as COD soluble averaged at 1400 mg/l. The increase in hydrolysis of the cells was due to a combination of factors: firstly the alteration of the digester temperature may have killed or inactivated mesophilic organisms present and secondly, the temperature increase stimulated the growth and activity of thermophilic organisms. The percentage of V.S. within the digester decreased significantly from that found during mesophilic operation. For example V.S. percentage during mesophilic operation averaged at 66.55% but decreased to 60.3% during operation in the thermotolerant range. Solids removal was best towards the later stages of operation in the thermotolerant temperature range,

following acclimatisation to environmental conditions. On average 24.6% of the T.S. was removed, 27.5% of V.S. was removed and 37.3 % of Total COD was removed.

**Operation in the Thermophilic Temperature Range**

The digester was maintained at a HRT of 12 days and an average D.O. concentration of 2.08 mg/l. The temperature was increased to 53 °C, which is specified as one of the time – temperature conditions to achieve a Class A Biosolid in terms of pathogen reduction (> 120 hr retention at 53 °C) (U.S. EPA, 1990). Following alteration of the digesters temperature, the D.O. concentration decreased, the average for the first week of operation at 53 °C was 1.35 mg/l. For the remainder of this phase in the trial the D.O. was maintained above 2 mg/l. The rate of evaporation correspondingly increased due to the temperature and air flow alteration. The concentration of soluble COD in the digester effluent decreased significantly to an average of 1117 mg/l during thermophilic operation. This may be due to a transition phase, where the microbial population moved from predominantly thermotolerant, to thermophilic. Removal efficiency was 14.6 % for T.S. and 28.81% for V.S., while Total COD removal was just over 30% (31.11%) lagging behind the removal efficiency experienced during thermotolerant operation. The pH decreased to 7.33, on average.

**Digester Performance during Alteration of Loading Rate**

**12 Day HRT**

With the digester fully operational and acclimatised to the operating temperature of 53 °C, the next phase of the trial focused on the performance of the digester during variations in HRT, in order to determine the optimum feed rate / HRT for the system. The reactor was maintained at the HRT of 12 days and an average D.O. of 2.4 mg/l (week 12 –14 measurements).

During week 15, the air compressor failed on two separate occasions. The first failure occurred overnight. As the forced condensation stopped with cessation of aeration, the automatic condensate replacement resulted in an excess of liquid in the digester and some dilution. Also, as the air forced cooling effect had ceased, the reactors contents

increased to the temperature of the water jacket (70 °C). Soluble COD in the effluent doubled, from 1,196 mg/l to 2,640 mg/l, this can be attributed to cell lysis as a result of heat shock.

While it is reasonable to assume that the reactor became depleted of oxygen, there was no noticeable odour, above the normal VFA odour (sweet fermention odour). VFA was measure at 181 mg/l as acetic acid. The pH fell to 6.86 during the period of compressor failure, but had returned to 7.72 by the end of the week. Alkalinity was measured at 3,565 mg/l $CaCO_3$ at the time of low pH. The digester settled quickly after the first air compressor replacement, returning to 53 °C and, in fact, climbing to 56.6 °C for the next two days, when there was a second air compressor failure. In the six hours following the second breakdown, temperature increased to 60.3 °C. After the second replacement, temperature returned to 53 °C, and was stable at that for the remainder of the trial period. It is possible that the short self-heating stage following the first compressor breakdown and replacement (temperature increased from 53 – 56.6 °C) was a result of utilisation of excess soluble COD produced during the initial temperature shock.

From week 15, reliable D.O. measurement was not possible as D.O. meters are unable to measure the D.O. accurately in such low concentrations. The digester was also qualitatively assessed daily for changes in sludge colour and odour, to highlight the development of an oxygen deficient environment, while pH would continue to provide valuable information on the health of the system. In terms of removal efficiency the digester performed well, despite failure of the air compressor (omitting week 15 results, due to dilution) the T.S. reduction was 20.35% while V.S. reduction was 31.17% and CODt was 34.23%.

**10 Day HRT**

The HRT was reduced from 12 days to 10 days which had a knock on effect on the OLR to the digester, averaging at 40.33g V.S./d for this period. The rapid turnover rate of the thermophilic aerobic digester facilitated by the short HRT and SRT of the system, ensured that the system remained largely unaffected by the shocks encountered during the previous stage of the trail (HRT 12D). Removal efficiency was good, achieving T.S. removal of 22.8%, V.S. removal of 37.19%, and total COD removal of 37.17 %. This was the best removal efficiency since start-up. Alkalinity and pH also increased

following the utilisation and neutralisation of VFA's previously encountered, with values in the range s of 4624 mg/l $CaCO_3$ and 7.6 respectively. VFA was measured and, at 28 mg/l as acetic acid, was found to have been reduced from the period of temperature shock and oxygen deficiency.

**8 Day HRT**

In the final stage of the trial, the HRT was reduced to 8 days, which is typical of a full-scale reactor. The OLR increased correspondingly to 50.36 g V.S./d. The pH remained stable averaging at 7.62, while alkalinity increased to 4,702 mg/l $CaCO_3$ on average, indicating that oxygen limitation was occurring despite increasing the BOD load to the reactor. The percentage T.S. removal, however, decreased to 17.65% on average, V.S. removal decreased to 27.46%, while CODt removal decreased slightly to 32.87%.

## 5.3 Evaluation of Pilot Plant Performance in the Treatment of Secondary Sludge

### Removal efficiency

The success of sludge digestion is typically assessed quantitatively by means of net solids reduction (T.S. removal) and V.S. removal. Anaerobic sewage sludge digesters are maintained at a HRT ranging from 15 – 30 days on average (Tchobanoglous and Barton, 1991).

In the initial stage of the trial, the pilot scale anaerobic reactor was operated at a HRT of 33 days and an average OLR of 10.33 g VS/D at a T.S. concentration of 4.5%. During this period, an average T.S. removal of 37%, V.S. removal of 38% and CODt removal of 37% was achieved. However, following a reduction in HRT to 15 days, which increased the OLR to 22.9 g VS/D approx. (at 4.6% T.S.), solids began to accumulate within the digester and removal efficiency decreased to an average of 19% for T.S., 25% for V.S. and 26% for CODt.

As secondary sludge is produced following the biological oxidation of wastewater, its composition varies greatly from that of primary sludge, which is composed mainly of fine faecal matter. The majority of secondary sludge is composed of bacterial cells and organic constituents, which could not be treated sufficiently (due to a short retention time) in the wastewater treatment process.

Proteins, fats, and short chain carbohydrates are all easily oxidised during activated sludge process. The compounds less amenable to oxidation however, are carried forward for treatment with the sludge. These compounds are polysaccharide in nature and may include; cellulose, hemicellulose, lignin, pectin, and starch.

The first stage of digestion is hydrolysis, and is performed by means of extracellular enzymes. The hydrolytic organisms present are substrate specific, for instance celluolytic bacteria such as Clostridia, Ruminococcus, Butyrivibrio, Bacteroides and Cellobacterium, are genera commonly observed in sewage sludge digesters. These organisms play a predominant role in the hydrolysis of polysaccharides, but cellulotic bacteria are often in fewer numbers than the acidogenic organisms and thus hydrolysis

of these compounds is rate limiting. Starch, on the other hand, is more susceptible to hydrolysis. The main genera capable of hydrolysing starch include Bacteroides, Micrococcus, Bacillus, Clostridium and Pseudomonas (Marty, 1984).

The rate limitation of hydrolysis was clearly apparent during operation of the pilot scale anaerobic digester, as greater removal efficiency was observed at the longer HRT of 33 days. At the mesophilic temperature of 35°C and a 33 day HRT the Biosolid produced is reflective of a class 'B' sludge with respect to V.S. removal (U.S. EPA, part 503). However, operation of a digester at a HRT of 33 days in practice requires the provision of larger reactors, increasing the capital cost of treatment.

Thermophilic aerobic digestion proceeds at a much faster rate than anaerobic digestion, due to the rapid synthesis of biomass within the reactor. The high temperature of operation, has a number of beneficial effects on digestion, including:

- Increased substrate utilisation rate, as microbial metabolic activity increases with temperature.
- Aids in mixing, as the viscosity of the sludge is reduced.
- Increased rate of oxygen diffusion (even though oxygen saturation decreases).

All of these factors allow operation of a thermophilic aerobic reactor at a much lower HRT, typically in the range of 5 – 9 days, (Kelly 1995).

The aerobic pilot scale reactor was operated at three different hydraulic retention times during the course of the trial, 12, 10, and 8 days. Decreasing the HRT increased the OLR to the digester, which ranged from 33.72g V.S./d, 40.33g V.S./d and 50.36g V.S./d, respectively.

At the 12 day HRT, approx. 22% T.S. removal, 34% V.S. removal and 35% CODt removal was achieved. There was slightly better removal achieved at the 10 day HRT which resulted in 23%T.S. removal 37% V.S. removal and 37% CODt removal. However, the largest variation in removal efficiency was experienced when the HRT was decreased to 8 days, as T.S. removal was just 17%, V.S. removal 22% and CODt

removal was 32%. These results indicate insufficient time for metabolisation and hydrolysis of the substrate. CODt removal suggests that the pool of readily available CODs within the digester was being utilised and thus the active biomass had not gone into the endogenous respiration stage, as there is ample substrate available. The endogenous respiration phase is where the net reduction in solids typically occurs.

**Process Stability**

The conversion of volatile matter to methane, carbon dioxide, water and new biomass by anaerobic means, is a sequential process involving the interaction of a synergistic group of microorganisms. The nature of the process is such that the by –products of degradation produced by one group of organisms forms the substrate for another (Aguilar *et al.*, 1995).

Under stable operation, the production of these intermediate by-products is equivalent to their utilisation. However, alteration of operational and environmental parameters such as HRT or feedstock composition can have a profound effect on the production of intermediate by-products, especially Volatile Fatty Acids (VFA), as the specific growth rate of the acid forming organisms is greater than that of the methanogens ($\mu$ max ~ 1 / hr, and $\mu$ max ~ 0.04 / hr, respectively) (Bitton, 1994). VFA production in the pilot scale anaerobic digester was monitored by two different techniques.

- HACH Spectrophotometer, as an early warning of possible instability (expressed as mg/l acetic acid).
- Gas Chromatograph, to evaluate the overall health of the system based on specific VFA's present. (results presented in Appendix E)

At the HRT of 33 days, VFA's were low, averaging at 95mg/l acetic acid. As expected, the VFA concentration increased following alteration of the HRT to 15 days, averaging at 225mg/l as acetic acid. This is due to the doubling of the OLR to the digester. These results indicate that VFA's were not been utilised quickly enough by the methanogenic population and were starting to accumulate within the digester. However, on review of the concentrations of specific VFA's present, this is not the case, as at a HRT of 33

days the concentration of acetate averaged at 1.97 mg/l, propionate 1.32 mg/l and butyrate 2.11 mg/l. when the HRT was decreased to 15 days, these particular VFA concentrations were similar, with 2.99 mg/l acetate, 0 mg/l propionate, and 0.21 mg/l butyrate on average. It is concluded, based on the fact that a VFA concentration up to 224 mg/l as acetic acid was recorded, the individual constituents of this total VFA figure for secondary sludge are different to those normally produced in anaerobic digesters treating primary sludge. Thus, while propionate, butyrate and acetate are deemed good indicators of process stability in the digestion of primary sludge (Aguilar *et al.,* 1995), they are not good performance indicators in the treatment of secondary sludge. VFA's suitable as indicator of process performance could be determined by analysis for a broader spectrum of VFA's in future studies. Biogas composition, in this instance, is of greater significance, as inhibition of methanogenic activity would be reflected by the methane content of the biogas. On consideration of biogas composition, the process was seen to be stable when operated at 33 day HRT and at a 15 day HRT, as the biogas contained 63% methane on average.

VFA's are not typically produced in aerobic environments, except under oxygen deficient conditions. VFA was measured however; during operation of the pilot aerobic plant following the application of heat to the digester to ensure the system had not become oxygen limited. The result was 0 mg/l as acetic acid. VFA was repeated when the air compressor failed, prompted by the presence of a sweet odour from the reactor. The results in this case were 180 mg/l as acetic acid. Following aeration and re – stabilisation of the reactor, repeat analysis was conducted to ensure VFA production has subsided, the result of which was negligible. These results are in accordance with the findings of Li *et al.*, (2002) in which VFA was found to accumulate in systems were the oxygen demand exceeded the oxygen supply.

**Influencing Factors** (Feedstock biodegradability)

Removal efficiency is dependant on the biodegradability of the feedstock. Typically, feed solids should contain greater than 70% V.S. for efficient anaerobic digestion (in terms of biogas yield) (Malina, 1992). Similarly, 70% V.S. is deemed as adequate to achieveand maintain autothermal conditions within the aerobic reactor.

In terms of secondary solids, not all of the volatile solids presented to the reactors, are readily biodegradable. It has been estimated by De Souza Araujo *et al.,* (1998) that an activated sludge plant operating at a sludge age of 10 days, produces a sludge with 64% V.S., but only 44% of this volatile matter is readily biodegradable / live (estimated by means of specific oxygen utilisation rates). When this feedstock was subjected to mesophilic anaerobic digestion, at a 20 day HRT, a T.S. reduction of just 26% and V.S. reduction of 29% was achieved.

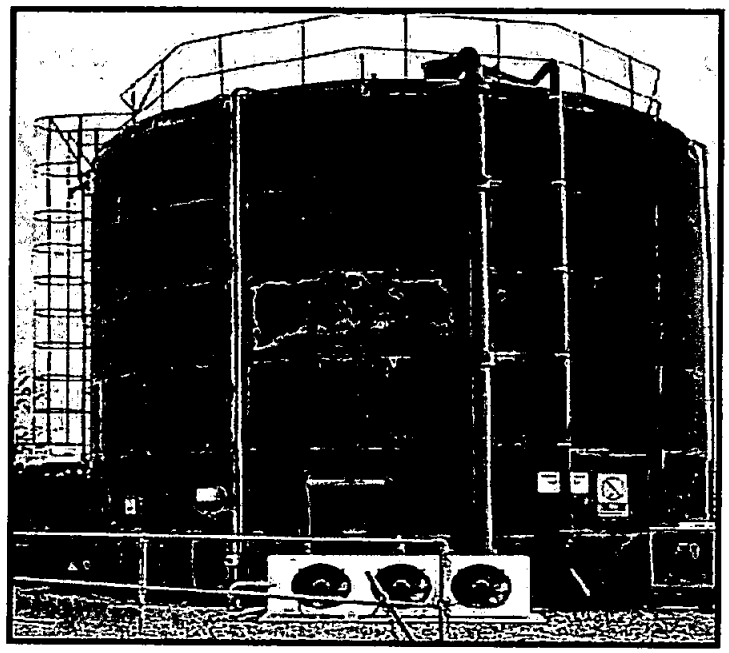
While the percentage removal via ATAD was not investigated in this study it is presumed that the live portion of the V.S. would be an important consideration in the biodegradability of the feed solids by aerobic means also. The Water Pollution Federation U.S (1995), advises that solids retention time is maintained at less than 15 days in the aeration basin of the wastewater treatment plant to keep endogenous respiration to a minimum.

The presence of toxins was discussed in section 2.2.4. However, it is important to point out on the topic of polyelectrolytes that, while Zetag 787 was found to be more amenable to degradation by anaerobic means in comparison to aerobic digestion, the possibility of a polyelectrolyte accumulating within an mesophilic anaerobic digester is more feasible, as the retention time is substantially longer than that of a ATAD, settlement or flotation and accumulation of such substances is further avoided in ATAD due to the temperature of operation, and the vigorous mixing and aeration system within the reactor.

## 5.4 Review of Full Scale Sludge Treatment Plants

The following section describes the current use of Mesophilic Anaerobic Digestion and ATAD in Ireland.

### 5.4.1 Sludge Treatment Plant Summary: Tullamore, Co. Offaly



**History**

| | |
|---|---|
| Treatment Technology: | Mesophilic Anaerobic Digestion |
| Year of Construction: | 1989 |
| P.E.: | Design 16,000, actual 14,000 |
| **Feedstock character:** | Primary and Secondary sludge produced on site. |
| Volume/ day: | 17 $m^3$ approx |
| HRT: | 20 days approx. |
| Feeding system | Continuous |
| **Plant and Equipment** | |
| No. Of Digesters | One |
| Volume: | 330 $m^3$ |
| Heating system: | Internal heat exchanger |
| Hot water boilers: | X 2, (1 propane gas and 1 methane) |
| CHP unit: | 2 gas engines (1 duty, 1 stand-by) |
| Gas Holder | 1 Bell over water system, 10 $m^3$ |
| Mixing system: | Gas injection |
| Final product storage: | 1 Open top storage tank |
| Final sludge disposal / reuse | Belt pressed to 17% T.S. approx., and land spread |

**Tullamore Sludge Treatment Plant**

**Primary Treatment**

2 Primary clarifiers (Size: 14m diameter, side wall depth of 2 m). System desludges automatically to the picket fence thickener.

2 Primary Biofilters (9.8 m diameter, 1.8 m deep, total volume of media 272$m^2$)

**Secondary Treatment**

2 Aeration basins (size 12m X 12m X 2.6 m deep, freeboard 1100 mm, liquid capacity of 750 $m^3$), operated in series. Ferric chloride added at inlet to aeration basin.

2 Secondary sludge clarifiers (size: 14 m Diameter, 3 m deep sidewall and 15 % floor slope). Desludged on a continuous basis by hydrostatic head to the sludge screw pump inlet chamber. Sludge drawn off to thickener.

**Sludge Preparation**

**Primary Clarifier**

Settled sludge is pumped to the sludge reception chamber, before being pumped to picket fence thickener (size: 5.8m in diameter, 4.5 m deep with rotary floor scrapers and picket frame, the tank is a covered concrete unit. Sludge continuously pumped to the PFT via automatic desludging equipment.

**Secondary Sludge:**

Initially the secondary sludge was mixed with the primary sludge but since it was only 0.7 % T.S. it resulted in a thin feed. In 1997 during up –grading of the plant two 2 new belt presses were acquired, all WAS is not belt pressed before being fed to reactor.

**Sludge Treatment**

There is one anaerobic digester, which is constructed above ground. The reactor has a volume of 330$m^3$, and is constructed from glass- coated steel panels on a concrete base. (7.6 m diameter, and 7.3 m in height). The reactor is externally insulated with twin skinned glass reinforced plastic (g.r.p) and polyurethane foam panels. The reactors roof is a fixed structure, constructed with insulated g.r.p.

**Digester Feeding**

The thickened sludge at 4% T.S. approx. is pumped (by two feed pumps one duty and one standby) separately from the PFT and belt press to the digester, in small batches at varying times throughout the day. The inlet port is situated on the roof of the digester, and consists of a tube projecting beneath the sludge surface level to a distance of 1000 mm, to prevent gas loss. The inlet is fitted with level probes to detect high-level sludge in the digester. Effluent removal occurs automatically during feeding, as the digested sludge passes to overflow weir (g.r.p assembly bolted to the side of the tank) to the down pipe through a 200 mm square opening in the tank.

**Digester Heating**

There is a hot water internal heat exchanger at base of tank. There are three methods of heating the water.

(1) There is one gas boiler operated on methane (used if CHP is not in use)

(2) One gas boiler operated on propane (used for start up or in case of failure of methane gas supply)

(3) The CHP unit which uses methane used to generate electricity, heat recovered from the exhaust and cooling the oil system satisfies the heating requirements of the digester.

The Digester is fitted with 8 internal annular heat exchangers, fabricated in epoxy coated steel, fixed to the digester base in circular array, each located above stirring pipes. Heat exchanger: 610 mm diameter, 1200 mm high, exchangers are fed from 2" supply and return water main. Digester is fitted with one water pump on the heating water circuit, which is designed to operate at all times, regardless of whether the boilers are operating.

**Mixing**

Performed 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 ratchet, which moves one portion every minute, to give one minute of stirring for each stirring pipe while the gas pump is

running. Gas pump operates automatically, controlled by gas pump timer on the main digester control panel. Manual operation is also possible.

**Gas Storage**

Gas is removed from the roof of the digester and flows to gasholder. Which is a bell over water system (twin – skinned g.r.p construction, on a concrete base). The gasholder has a small capacity of 10 $m^3$ as it was designed more for pressure regulation, gas buffering and gas filtering than storage.

Gas taken from the splash trap which is a large chamber at the top of the digester (to prevent particulate matter and water from entering the gas lines and gas holder). It 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, to provide some cooling of the gas and condense out some of the vapour contained in the gas. The outlet pipe turns up inside the trap to draw from its highest point.

The low points of the gas lines to digester, boiler, and gas pump are fitted with a condensation trap, next to the gas holder. This is an A.B.S water fitted trap that collects condensation from gas before it enters the boilers, pumps and engines.

There is a gas vent pressure release valve mounted on the top of splash trap in the centre of the digester roof, to allow gas venting in case of emergency.

**Energy Generation**

The CHP unit has an output of 21 kW, cannot operate independent from the ESB supply. Power generated is fed back to the main control panel, reduces the demand from the ESB grid. A flare stack was installed in 1997 but is rarely used.

**Final Sludge Storage**

The digested sludge storage tank is a rectangular open tank 24m x 7.5 m x 2.9 m deep with 300 mm freeboard, capacity 529 $m^3$. Sludge is dewatered to 17% T.S. by belt press and spread on land.

**Odour Control**

An odour control system was installed in 1997. This consisted of sealing the primary tank desludging units, the trickling filters, the picket fence thickener and emergency sludge holding tank, once collected, the odorous air is drawn off by extraction fans to biofilter units (shell system)

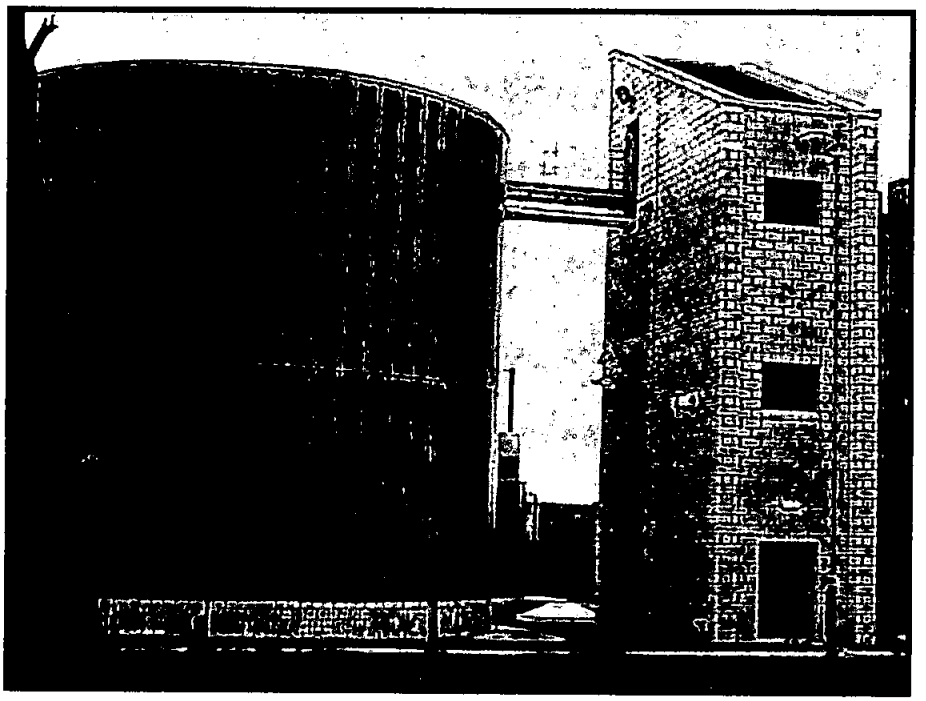
**Analysis**

Raw and digested sludge analysis was conducted twice per week up until 1998. However, since 1998 only temperature and pH has been monitored daily, as the digester is considered to be in stable operation.

**Table 5.7 Tullamore Analysis Results (Average 6/'96 – 8/'98)**

| **Feedstock** | | **Digester Sludge** | |
|---|---|---|---|
| pH | 6.50 | pH | 7.33 |
| T.S. % | 3.57 | T.S. % | 1.94 |
| | | T.S. removal % | 46 % |
| | | VFA mg/l acetic acid | 75 |
| | | Alkalinity mg/l $CaCo_3$ | 4,370 |
| | | Biogas composition: | |
| | | $CH_4$ % | 59 |
| | | $CO_2$ % | 41 |

### 5.4.2 Sludge Treatment Plant Summary: Greystones, Co. Wicklow



**History**

| | |
|---|---|
| Treatment Technology: | Mesophilic Anaerobic Digestion |
| Year of Construction: | 1996 |
| P.E.: | 13,000 (30,000 design) |
| **Feedstock character:** | Primary and Secondary sludge produced on site |
| Volume/ day: | 18 $m^3$ approx / digester. |
| HRT: | 30 day approx. |
| Feeding system | Continuous |

**Plant and Equipment**

| | |
|---|---|
| No. Of Digesters | One |
| Volume: | 570 $m^3$ / reactor |
| Heating system: | External heat exchanger |
| Hot water boilers: | X 1, (Dual fired, operates on biogas and propane) |
| CHP unit: | 1 gas engine |
| Gas Holder | 1 bell over water system (capacity 16.4 $m^3$) |
| Mixing system: | Gas injection |
| Final product storage: | 1 secondary digesters (stirred tank, with odour control) |
| Final sludge disposal / reuse | Dewatered by 1 decanter to 22% T.S., and land spread |

**Greystones Sludge Treatment Plant**

**Primary Treatment**

There are three primary settlement tanks, with provision for a fourth. Each 17.3 m diameter, when the sludge blanket level (indicated by probe) is detected the bellmonth drops and the hydrostatic pressure feeds the sludge to a sump.

**Secondary Treatment**

There are six aeration cells only three in use,(12m long, 12m wide and 6 m deep), aerated by fine bubble diffused aeration. The cells are used in sequence; the tanks are fitted with D.O. probes and MLSS probes at the outlets. Building No. 2 houses the air blowers, used to supply the aeration tanks with air. There are three secondary settlement tanks (20.2m diameter), with provision for a fourth. When the sludge level blanket level is detected the bellmonth drops and the hydrostatic pressure feeds the sludge forward to a collection sump.

**Sludge Preparation**

From the primary sludge sump the sludge is pumped to one of the two-covered PFT. PFT are glass-lined tanks, 4.2m in diameter and 4.0 m in height. The primary sludge is thickened to 4 – 5% T.S. The WAS is pumped from the secondary clarifiers to building No. 3, where it is thickened by two rotary drum thickeners. Solids are thickened to 5 – 6% T.S. Both sludges are then pumped to one of two-covered sludge mixing tanks (only one commissioned), and blended together. The tanks are glass lined 7.68 m in diameter and 2.8 m in height.

**Anaerobic Digester:**

There is one anaerobic digester, with a capacity of 570 $m^3$, (4.2 m in diameter and 8.4 m in height, with fixed roof).

The reactor is constructed from glass lined reinforced steel, the sidewalls and roof of the reactor is insulated with 200 mm of rockwool.

**Digester Feeding**

The digester is automatically fed once every hour (approx 0.79 $m^3$/hr). The feed is pumped from the mixed sludge storage tank to the digester by mono (positive

displacement pumps 1 duty and one standby) pump with a capacity of 15 $m^3$/hr. The feed sludge enters at the recirculation line before the external heat exchanger.

**Digester Heating**

The digester is heated by means of an external concentric-tube hot water heat exchanger unit. The unit consists of four heat exchangers, two on the bottom and two on the top. Each exchanger is 1.5 m long with a 225 mm diameter pipe. The sludge flows in sequence through each exchanger starting at the bottom and flows 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 an opposite direction. Heat exchanger has to be cleaned once every 6 months, which means that the digester is without heat on that particular day for 4 hours approx. the sludge is circulated constantly to the heat exchanger from the digester by two mono positive displacement pumps (1 duty and 1 standby) with a capacity of 37 $m^3$ /hr.

There is only one boiler on site and no back up. This boiler is dual fired run on methane or propane. The hot water is then circulated to the heat exchanger by centrifugal type hot water pumps (1 duty and 1 standby). Two temperature probes in the digester control the boiler. The temperature probes can also be used to indicate mixing, as readings from probes should be constant to indicate good mixing.

**Digester Mixing**:

Mixing is achieved by gas injection. There are 8 gas pipes, which enter through the top of the digester and travel down the side of the digester walls, finally turning out towards the centre of the digester. A gas compressor, outside the digester, compresses the gas to the 8 pipes. The quantity of gas entering the digester is controlled by solenoid valves, located on each of the pipes at the entrance to the digester. Opening and closing of these solenoid valves is controlled by an air compressor stored in the sludge building.

The gas injection system works as follows: Two-gas valve are opened and the gas is compressed into the digester for a predetermined period, mixing that part of the tank. Prior to the closing of these valves the adjacent two valves are opened and gas is compressed and expelled to these two valves. Then the two other valves cutout. Thus 4

valves are on together just for a short period. This pattern continues round the 8 valves and continues for 24 hr / day.

**Gas Storage**

There is one gas holding tank, bell over water type, with a storage capacity 16.44 $m^3$. A level switch controls the gas level in the holding tank. When the gas level reaches 95% of the holder capacity, the flare stack cuts in and flares the gas until the tank is back to 75% capacity. There is also a breathing safety valve. Prior to entering the gas holding tank the gas goes through a condensation pot, to remove moisture. Since the gas is used in the boiler, it again goes through another condensation pot. The pots are cleaned once per week, as, if they become blocked, gas will be trapped in the digester (should this situation occur, the digester is fitted with a pressure release valve and vacuum valve located at the top of the digester). The gas is not scrubbed or filtered which causes problems for the gas compressor, causing scale build up and requires regular cleaning.

**Energy Generation**

There is a CHP unit on site but it has not been commissioned.

**Final Sludge Storage**

The storage tank is situated beside the digester, which can also be described as a secondary digester, the tank is sealed with activated carbon units at the top for odour control, it is unheated, unmixed and provides 15 day storage. Sludge is then dewatered by decanter to 22% T.S. and is spread on land.

**Odour Control:**

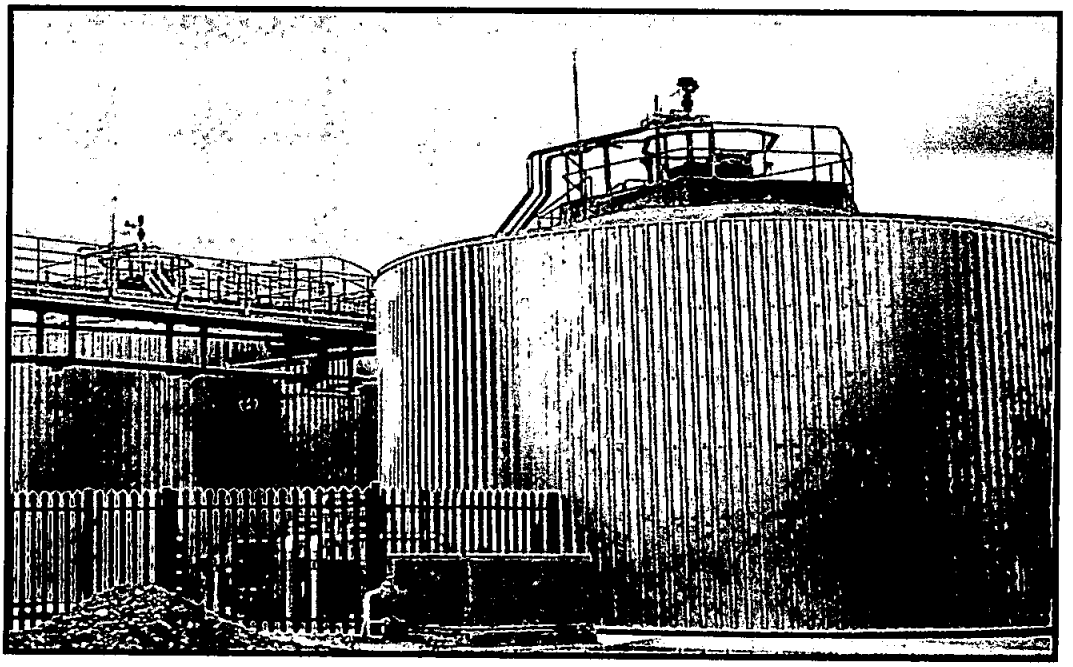
Odorous gas from PFT and Mixing tank is passed through a shell biofilter.

**Analysis:** Online pH probes, 2 temperature probes one in upper and one in the lower section of the tank (measured daily). Temperature probe in recirculation line (measured daily). Biogas volume and quality measured daily all other parameters 2 –3 times per week.

**Table 5.8 Greystones Analysis Results: 12/'02 – 2/'03**

| **Feedstock** | | **Digester Sludge** | |
|---|---|---|---|
| pH | 7.12 | T.S. % | 4.08 |
| T.S. % | 6.4 | T.S. removal % | 36 |
| | | VFA mg/l acetic acid | 244.87 |
| | | Alkalinity mg/l $CaCo_3$ | 5,300 |
| | | $NH_3$ mg/l | 1,353 |
| | | COD mg/l | 5,576 |
| | | Biogas composition: | |
| | | $CH_4$ % | 63 |
| | | $CO_2$ % | 34 |
| | | Biogas vol. $m^3/d$ | 221.5 |

### 5.4.3 Sludge Treatment Plant Summary: Clonmel, Co. Tipperary



**History**

| | |
|---|---|
| Treatment Technology: | Mesophilic Anaerobic Digestion |
| Year of Construction: | 1998 |
| P.E.: | 70,000 |
| **Feedstock character:** | Primary and Secondary sludge produced on site. |
| Volume/ day: | 80 $m^3$ approx. |
| HRT: | 20 day approx. |
| Feeding system | Continuous |

**Plant and Equipment**

| | |
|---|---|
| No. Of Digesters | Two |
| Volume: | 800 $m^3$ / reactor |
| Heating system: | External heat exchangers |
| Hot water boilers: | X 2, (Dual fired, operates on biogas and propane) |
| CHP unit: | 2 Dual fired |
| Gas Holder | 1 bell over water system (capacity 200 $m^3$) |
| Mixing system: | Gas injection |
| Final product storage: | 1 storage tank (same spec. as digester but unmixed and unheated) |
| Final sludge disposal / reuse | Dewatered by 2 double belt presses, landspread. |

**Clonmel Sludge Treatment Plant**

**Primary Treatment**

There are two Primary settlement tanks diameter (10.6 m), two trickling filters, two intermediate clarifiers.

**Secondary Treatment:**

One square aeration tank containing 4 aeration cells, aerated by surface aerators, followed by two clarifiers.

**Sludge Preparation**

Primary and intermediate sludge is pumped to a PFT, which is a covered concrete tank with a sludge capacity of 185$m^3$. Secondary sludge is thickened in a sludge thickener in the sludge treatment building. Both sludges are transferred to a blending tank, with vertical axis agitator to blend the incoming sludge. The blending tank has a sludge capacity of 35 $m^3$.

**Anaerobic Digester**

There are two reactors constructed from concrete with a sludge capacity 800 $m^3$/ reactor. Reactors are 10m in height, of which 4m is underground. The internal diameter is 10m. The walls of the reactors are 400 mm thick and the roof is 200 mm thick. Digesters are insulated with 100 mm of mineral wool.

The digesters operate in parallel. To date one digester is operating to full capacity with the second operation to half capacity. Digesters are fed automatically every 2 hours for 30 min; the feed pumps are controlled by the PLC installed in the control panel. The pumps cease operation if, for some reason, there is a low level in the blending tank or high level in the digester – feeding pit.

**Digester Feeding**

Blended sludge is pumped forward for 30 min every 2 hrs to the digester. Each digester has 1 duty and 1 standby pump; the pumps are progressive cavity type, with an average flow rate range of 2.4 to 12$m^3$/hr. On occasion, grease is added to the digester through separate grease loading pumps. Sludge feed is mixed with the recirculated heated sludge

in the feeding pit located at the top of the digester, and then conveyed by gravity to the sludge mass.

**Digester Heating**

Digesters are heated by external heat exchangers. Sludge is taken from the bottom is recirculated through the hot water heat exchanger and returned to the digester. The heat exchangers are jacket pipe type, in which hot water is pumped counter current to the sludge flow through a concentric pipe surrounding the sludge pipe. Horizontal centrifugal pumps are used for recirculation of sludge from the digester to the heat exchanger and back to the digester inlet pipe. The water is heated by 2 CHP units and / or by dual fired boilers (2) operated on methane or propane gas.

**Digester Mixing**

Mixing is achieved by gas injection. Compressed biogas is injected through a series of 4 drop pipes, located on the roof, descending vertically to the bottom of the digester. Large gas bubbles pass through the sludge mass, and mix the digesters contents. Biogas is recirculated by rotary vane compressors. Prior to the gas entering the compressor, it passes through a gravel filter to remove particulates and condensation. Each injection pipe is fitted with a solenoid valve, for sequential distribution of gas in the pipes. Mixing is performed 24 hrs/ d.

**Gas Storage**

Biogas is recovered from the dome at the top of the reactor. Some of the biogas is diverted to storage and some is recycled for mixing.

As previously mentioned, biogas is passed through a gravel filter and from here to one common gas storage tank serving both reactors. The gasholder is a concrete tank, 8 m in diameter and 4m in height, with a monolithic constructed mild dome with vertical guides. The total capacity of the tank is 200 $m^3$, this is approx. 6 hours storage. The tank has a floating cover with breathing safety valve. There is a flare stack located beside the gasholder, which automatically flares gas when a high level is reached on the gas holder.

Gas for use in the boilers is purified by gas scrubber to remove $H_2S$ gas. Element of the gas scrubber include:

- Washing tower
- Water storage and washing pump
- Sodium hydroxide storage tank and dosing pump
- Heating system in the sodium hydroxide storage pump.

**Energy Generation**

There are two CHP units on site one duty and one standby. Heat is also recovered from the flue gas discharge and cooling water / oil system. An emergency cooling system has been installed on the water pipe from the heat exchanger in case of the digester been heated above 35 °C.

**Final Sludge Storage**

Digester effluent travels by gravity to a holding tank, the holding tank is the same dimensions of the digesters, but is unheated and unmixed. Since gas is given off during storage the tank is fitted with safety overpressure valve (with solenoid valve). Sludge is gravity fed via telescopic valves form the storage tank to the dewatering building, where it is dewatered by two double belt presses, and spread on land

**Odour Control**

Gas from the PFT and sludge-blending tank is diverted to a shell biofilter unit.

**Analysis**

On-line temperature probe, pH measured daily, all other analysis approx. 2- 3 times per week

**Results:** Not available.

## 5.4.4 SLUDGE TREATMENT PLANT OF OSBERSTOWN, CO. KILDARE.

**History**

| | |
|---|---|
| Treatment Technology: | Mesophilic Anaerobic Digestion |
| Year of Construction: | 2001 |
| P.E.: | 80,000 |
| **Feedstock character:** | Primary and Secondary sludge produced on site and imported |
| Volume/ day: | 131.2 $m^3$ / digester approx |
| HRT: | 10 days approx. |
| Feeding system | Continuous |

**Plant and Equipment**

| | |
|---|---|
| No. Of Digesters | Two |
| Volume: | 1317 $m^3$ / digester |
| Heating system: | External heat exchanger |
| Hot water boilers: | X 2, (Dual fired, operates on biogas and natural gas) |
| CHP unit: | 2 Dual fired (Natural gas and Biogas) |
| Mixing system: | Gas injection |
| Final product storage: | 2 tanks (capacity 916 $m^3$ / tank) |
| Final sludge disposal / reuse | Dewatered, by 2 belt presses, applied to non agricultural land |

**Osberstown Sludge Treatment Plant**

**Construction and Commissioning: 1999- 2001**

P.E: 80,000

The work was the first "design and build" operation conducted by a local authority in Ireland.

Stage I: initial plant 1981

Stage II: upgrade to current level

Stage III: upgrade to P.E 130,000 due to start 2004 (+ Sludge handling Hub center)

**Primary Treatment**

2 Primary settlement tanks (24 m x 5m depth), Volume 1,696 $m^3$ / tank.

**Secondary Treatment**

There are four 40 m diameter circular basins, with fine air diffusion grids. Each basin, during a normal flow cycle, operates for four hours, therefore repeating the cycle six times a day. The process works on the "fill and draw" principle and is thus not a continuous process. The four hour cycle consists of a two hour fill-aerate sequence, followed by a fill-settle sequence and finally the decant phase. Thus, at any one time each of the basins is in one of the above sequences. A dissolved oxygen probe, located near the skimmer arms, monitors the D.O. in the basin continuously.

**Sludge Treatment**

The entire liquid and sludge treatment process is controlled by a SCADA system. MCC 3 and 4 are located in the Sludge Building. MCC3 controls the heat exchangers, CHP units, boilers and digestion plant. MCC 4 controls the imported sludge, buffer tanks, drum thickeners and sludge presses.

There are three sources of sludge: primary sludge, surplus activated sludge (SAS) and imported sludge from outside treatment plants. The Primary, SAS and Imported Sludge are pumped to two buffer tanks before being thickened by the Drum Thickeners. Imported sludge is received from the smaller municipal treatment plants by tanker and is screened before being pumped to the buffer tanks. In order to allow imported sludge into the system the tanker details must be entered into the SCADA computer before it will open the imported sludge screen valves. The mixed buffer sludge has a

polyelectrolyte added to thicken it from approx.1% T.S. to 6% T.S. The thickened sludge is then pumped to a digester feed holding tank. The thickened sludge is reduced in solids content upon digestion from 6% T.S. to approx. 3% T.S.

**Digesters**

There are two digesters each with a capacity of 1,317 $m^3$. Typically, the digesters are operated at a HRT of 10 days.

**Digester Heating**

Operating temp 34 °C approx. The feed sludge is heated, by passing the sludge through a heat exchanger. The heat exchanger is a circular unit and contains two pipes, one for the thickened sludge, and one for hot water. Either the Biogas/Oil burners or the Combined Heat and Power (CHP) units heat the water to 70 °C. Returned sludge from the digesters is also combined with the feed (thickened) sludge before passing through the heat exchangers. The heated sludge is fed to the digesters at a controlled rate, of 55 $m^3$ / hour.

**Digester Mixing**

There are three gas-mixing compressors, two are on a duty basis the other as a common standby. The gas is drawn off the top of the digester, compressed, reintroduced to the top of the digester through a seven-way manifold, and passed down through seven diffusers at the base of the tank. The excess biogas is drawn off and stored in the gasholder. The gasholder helps maintain a constant system pressure from the irregular gas production by the digesters. The gas in the holding tank is boosted by two gas boosters before being used as a fuel in the CHP and boiler units. A number of safety features are built into this system. On the digesters there are pressure and vacuum release valves, which operate if too much pressure builds up in the digesters or if a vacuum develops for some reason. Similarly, the gasholder contains pressure and vacuum release elements. The gasholder is connected to a flare stack, which ignites excess methane not required by the boilers or CHP units.

**Final Product Storage**

The digested sludge exits to the two holding tanks each with a capacity of 916 $m^3$, giving a 20-day holding capacity. From the holding tanks the digested sludge is fed to 2

dewatering presses, and is thickened to approx. 20% T.S. Pressure switches on the skips detect when the skip is full and automatically the system switches from a full skip to an empty one, ensuring a continuous process. The sludge is disposed of by land spreading at present but within the next five years a sludge handling facility will be built on site to further treat the dewatered sludge. No decision has been made to date as to what method will be used. From the Kildare County Council Waste Management Plan 2000 some of the options include lime stabilisation, pyrolysis or heat treatment with energy recovery (Cronin, 2002).

**Results of Analysis:** Not Available

## 5.4.5 SLUDGE TREATMENT PLANT OF NAVAN, CO. MEATH.

**History**

| | |
|---|---|
| Treatment Technology: | Mesophilic Anaerobic Digestion |
| Year of Construction: | 2001 |
| P.E.: | 40,000 (Stage I) |
| **Feedstock character:** | Primary and Secondary sludge produced on site and imported sludge |
| Volume/ day: | 35 $m^3$ approx / digester. |
| HRT: | 20 day approx. |
| Feeding system | Continuous |

**Plant and Equipment**

| | |
|---|---|
| No. Of Digesters | Two |
| Volume: | 800 $m^3$ / reactor |
| Heating system: | External heat exchangers |
| Hot water boilers: | X 2, (Dual fired, operates on biogas and propane) |
| CHP unit: | 1 methane |
| Gas Holder | 1 bell over water system (capacity $m^3$) |
| Mixing system: | Gas injection |
| Final product storage: | 2 secondary digesters (stirred tanks capacity 760 $m^3$ each) |
| Final sludge disposal / reuse | Dewatered by 2 centrifuges to 18 – 25% T.S. and lime treated, it is then spread on tillage land. |

**Navan Sludge Treatment Plant**

Year of Construction / commissioning: 2001 - 2002

Design parameters

| | Stage I | Stage II |
|---|---|---|
| P.E | 40,000 | 60,000 |
| Flow L/S | 104.17 | 156.25 |
| DWF | 2 | 3 |
| BOD | 2,400 KG/D | 3,600 KG/D |
| SS | 2,880 KG/D | 4,320 KG/D |

**Primary Treatment**

There are two primary clarifiers (Size: 22m diameter, capacity 1000$m^3$)- sludge at 2% T.S. is desludged automatically, at a rate of 107$m^3$ /d).

**Secondary Treatment**

Two aeration basins with 4 cells, each cell with a capacity of 1,800 $m^3$. There are 2 secondary settlement tanks, diameter of each is 28 m, with a capacity of 800 $m^3$ / tank. WAS is either pumped to the sludge reception chamber at a T.S. concentration of 0.5 – 1% or returned to the aeration basins. The RAS is dosed with ferric sulphate at a rate of 19 – 20 / hr.

**Primary Clarifiers**

Settled sludge is pumped to sludge reception chamber system, before being pumped to 2 concrete picket fence thickener (size: 7m in diameter, 4.3 m deep with rotary floor scrapers and full depth picket frame). Sludge is continuously pumped to PFT via automatic desludging equipment.

**Secondary Sludge**

Secondary sludge is pumped to the sludge reception chamber, and it is then pumped to the dewatering room where it is thickened by two rotary drum thickeners. The feed rate to the thickener is 1 – 15 $m^3$ / hr, therefore 170 – 200 $m^3$/ d of sludge can be produced. Polyelectrolyte Zetag 7898 is used to aid in dewatering at a rate of 180 – 340 L / hr.

**Imported Sludge**

Imported sludge is received at the sludge reception chamber and, depending on its consistency (i.e. Primary of secondary sludge), it is thickened as stated above. At present, the Hub centre is only taking in liquid sludges, as the plant has no facilities on-site to accept, or reconstitute dry sludge cakes (8 % T.S. and above). Dry sludge cake produced in the County is currently being sent to a local composting farm where it is treated by a windrow system, mixed with wood chips, sold and used for building up road verges.

**Sludge Treatment**

There are two digester constructed above ground, construction form glass- coated steel with concrete base. Each digester has a sludge volume of 800 $m^3$. The reactors are insulated externally, with twin skin glass reinforced plastic (g.r.p) and polyurethane foam panels. Tank roof is fixed, using insulated g.r.p construction.

**Digester Feeding**

Thickened sludge pumped (by two feed pumps 2 duty and one standby) separately from PFT and rotary drum thickener to the digester in small batches at varying times throughout the day.

**Digester Heating**

Heating is achieved by a jacketed hot water external heat exchanger. Two ways of heating water

(1) 2 Dual fired boiler, which can be run on propane or methane (used if CHP is not in use)

(2) 1 CHP, uses methane to generate electricity

**Mixing**

Gas injection. 6 gas pipes located near the bottom of the digester, spread evenly around the diameter of the tank. A gas compressor outside the digester, compressed the gas to the 6 pipes. The quantity of gas entering the digester is controlled by pneumatic valves, located on each of the pipes at the entrance to the digester. Opening and closing of these pneumatic valves is controlled by an air compressor stored in building located between the two gasholders.

Gas storage: Biogas is removed from the roof of the digester and flows to gasholder, which is a bell over water system.

**Energy Generation**

The CHP unit has an output of 85 kW, cannot operate independent from the ESB supply. Power generated is fed back to the main control panel. It reduces the demand from the ESB grid. A flare stack has been provided on site to burn off excess biogas.

**Final Sludge Storage and Treatment**

Digested sludge spills from the digesters to the (2) secondary digesters. The tanks are unheated, sealed units, with a stirrer for mixing gas released is diverted to gas collection tanks. Each tank has a capacity of 760 $m^3$.

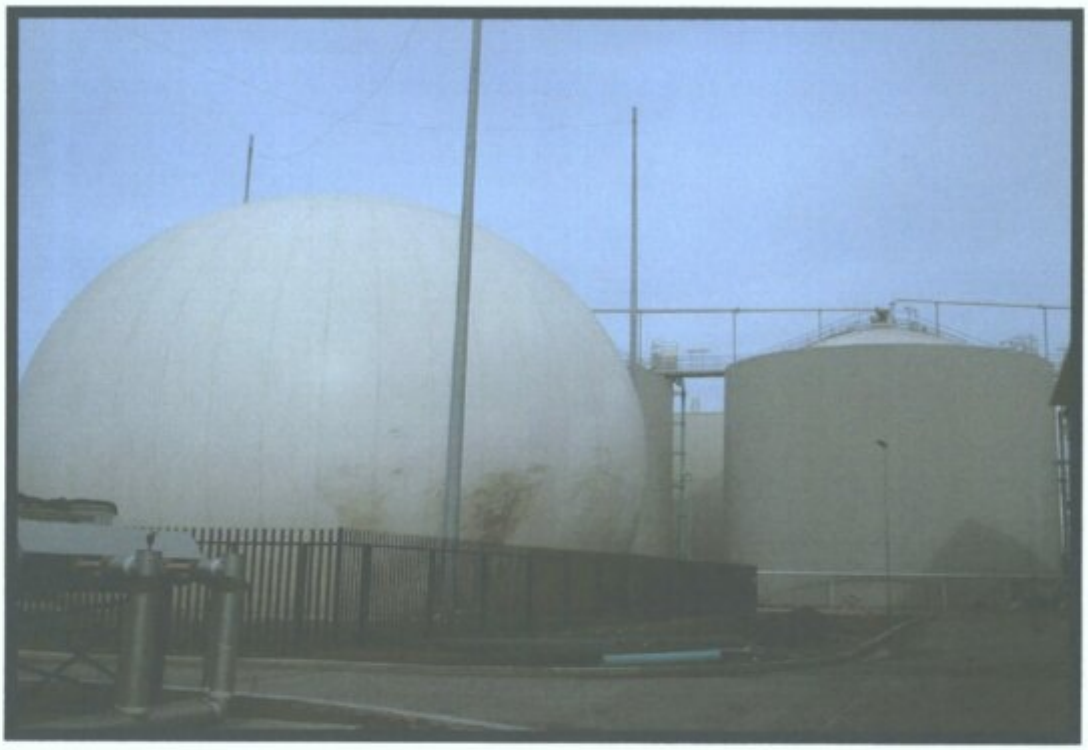
The digested sludge, at 3% T.S. approx., is stored for 15 days in the secondary digesters. It is then dewatered by 2 centrifuges, to 18 – 25% T.S. with the aid of polyelectrolyte and receives further treatment by the lime stabilisation. Prior to lime stabilisation the dewatered sludge is passed through another heat exchanger to heat the sludge to 73 °C. The liming process further increased the T.S. of the sludge to approx. 40%. At present 70 tonnes of sludge is produced per year, however 15% of this weight is composed of lime. The finished product sludge is then transported to a local tillage farm, where it is stored on site and spread twice per year on tillage land. The estimated cost of the lime process at Navan is € 120 / tonne. It has been proposed that the plant will soon be accepting Alum sludge from water treatment plants in County Meath, this sludge will be received to a separate reception chamber and will be fed into the secondary digestion tanks, and treated and disposed of as per the digested sludge.

**Analysis:** Three times per week.

**Table 5.9 Navan Analysis Results (7/'02 – 9/02)**

| **Digester 1.** | | | | **Digester 2.** | | | |
|---|---|---|---|---|---|---|---|
| **Feedstock** | | **Digester Sludge** | | **Feedstock** | | **Digester Sludge** | |
| T.S. % | 6.3 | T.S. % | 4.90 | T.S. % | 5.95 | T.S. % | 2.71 |
| V.S.% | 63.38 | T.S. removal % | 22% | V.S. % | 68.66 | T.S. removal % | 54% |
| COD mg/l | 1,684 | V.S.% | 53.9 | COD mg/l | 4657 | V.S.% | 51.60 |
| pH | 6.91 | V.S. removal % | 34% | pH | 5.72 | V.S. removal % | 66% |
| Feed rate $m^3$ | 19.32 | pH | 7.63 | Feed rate $m^3$ | 16.52 | pH | 7.11 |
| | | VFA mg/l acetic acid | 273 | | | VFA mg/l acetic acid | 263.2 |
| | | Alkalinity mg/l $CaCo_3$ | 5225 | | | Alkalinity mg/l $CaCo_3$ | 2597 |
| | | COD mg/l | 5665 | | | COD mg/l | 2,524 |
| | | Biogas composition | | | | Biogas composition | |
| | | $CH_4$ % | 63.5 | | | $CH_4$ % | 58.74 |
| | | $CO_2$ % | 35.3 | | | $CO_2$ % | 38.81 |

### 5.4.6 Sludge Treatment Plant Summary: Ringsend, Dublin



| | |
|---|---|
| **History** | |
| Treatment Technologies: | Thermal Hydrolysis<br>Mesophilic Anaerobic Digestion<br>Thermal Drying |
| Year of Construction: | 2002 |
| P.E.: | Design 1.7 million, (capacity could be increased to 1.9 million, corresponding to population increase over the next 20 years) |
| **Feedstock character:** | Hydrolysed primary and secondary sludge |
| Volume/ day: | 849 $m^3$ /d |
| HRT: | 15 days approx. |
| Feeding system | Continuously fed to the digester |
| **Plant and Equipment:** | |
| **Anaerobic Digester** | |
| No. Of Digesters | Three |
| Volume: | 4,250 $m^3$ / reactor |
| Heating system: | No internal / external heat exchangers, temperature is maintained at 35 °C, by continuous feed administration at 40 °C from the hydrolysis stage, and good insulation of the tanks |
| Mixing | 1 Mechanically mixed draft tube/ tank |
| Gas collection system | 1 Double membrane gas bag (Capacity 5000$m^3$) |
| Final product storage: | Storage sump sludge is fed forward to Thermal drying plant |
| Final sludge disposal / reuse | Sludge is marketed as 'Biofert' soil conditioner and fertiliser. |

**Ringsend Sludge Treatment Plant (Dublin Bay Project)**

The Dublin bay project is the largest wastewater treatment facility of its kind in Ireland and, in some aspects, the most advanced plant in the world. The overall project involves three elements; upgrading the existing wastewater treatment plant at Ringsend, building a pumping station at Sutton and laying the submarine pipeline under the Bay to bring wastewater from Sutton to Ringsend. The treatment plant will serve the Dublin area to Dun Laoghaire, the west of Dublin and parts of Meath, and north Dublin including the airport and Portmarnock region.

Plant p.e. is 1.7 million, 1.2 million from domestic and 0.5 million industrial. The overall capacity of the plant could be increased to 1.9 million in accordance with predicted population increases over the next 20 years.

**Treatment Plant Upgrading**

The majority of the upgrading work has been completed, however sections of the plant are under going commissioning.

**Preliminary Treatment**

The influent is screened to remove fat, oils and grease (later added to the anaerobic digester) and undergoes grit removal, to remove coarse particles, which may damage equipment.

**Primary Treatment**

The raw wastewater is diverted to 12 primary 'Lamella' settlement tanks arranged in 2 banks of six tanks, each lamella tank is 40 x 12 x 5 M in size. The lamella tank design was chosen due to space restrictions on site. Settled sludge at 3% T.S. is automatically pumped by progressive cavity pump to two mixing tanks were it meets the secondary sludge.

**Secondary Treatment**

Biological oxidation of the wastewater occurs in 24 sequential SBR fine bubble diffused aeration tanks; the each tank is 52 x 39 x 7 m in size. Due to space restrictions the tanks

are housed in a two – storey stacked structure. The final effluent passes through a UV filter prior discharge to Dublin bay, while the settled secondary sludge at 0.8% T.S. is automatically pumped via progressive cavity pump to two mixing tanks where it joins up with the primary sludge.

**Sludge Treatment**

The Ringsend plant is state of the art in regard to sludge treatment facilities, incorporating a number of technologies such as Thermal Hydrolysis, Anaerobic Digestion, and Thermal Drying to produces a 'Class A' Biosolid which can be re-used in agriculture.

**Feed Preparation and Treatment**

The primary and activated sludge collected in the 2 sludge mixing tanks is dewatered by belt press with the aid of polyelectrolyte to 14% T.S. It is then subjected to batch thermal hydrolysis in which the sludge is heated to 165 $^{o}$C at 8 bar pressure for 30 minutes. Some dilution occurs during hydrolysis and the sludge now at 12 % T.S. It is subsequently passed through a heat exchanger to cool the sludge to 40 $^{o}$C, and fed directly to the anaerobic digesters for further treatment (solids reduction and gas production). For the hydrolysis to be economically viable, the process is largely dependent on the gas produced from the anaerobic digestion process. The hydrolysis building houses two dual fired boilers operated on natural gas and methane, 4 combined heat and power (CHP) units each with an output of 1000 kW, and there are also two heat exchangers which recover heat from the exhaust of the CHP units.

Just as the anaerobic digestion process contributes to hydrolysis, so too does hydrolysis to anaerobic digestion, as the hydrolysis of the feed facilitates the application of a higher feed solids to the reactors. The digesters are sequentially fed on a continuous basis, at present the digesters are operated on a HRT of 15 days. Approximately 849 $m^3$/d is fed to the digesters by two sets of duty/ standby progressive cavity pumps, as the digesters are fed, effluent removal occurs simultaneously by gravity through the hopper and finally to a sludge sump. Typically, the feed is 12% T.S., which is almost unheard of in anaerobic digestion processes for a number of reasons such as, the high organic loading (double the feed typically fed to anaerobic digester) could contribute to volatile fatty acid overload, theoretically.

However, since the feed is maintained at this level, the digester would, in time acclimatise to high loading without difficulty. Pumping and heating such dense sludge would also pose a problem for typical unit operations. In this case, however, hydrolysis complements the anaerobic digestion process in many ways; firstly pumping of the sludge at 12% T.S. is not as problematic as one might have thought, as the sludge enters the digester at 40 °C and therefore the viscosity of the sludge is greatly reduced and imitates that of a much thinner sludge. Pumping is achieved by progressive cavity pump. Secondly the temperature of the feed sludge is adequate to ensure that the digester does not require an additional heat source, which is an economic advantage of pre - hydrolysing the feed. Furthermore, hydrolysis of the feed causes destruction of the sludge floc releasing the biodegradable organics for utilisation in the digester. Not only does the digester achieve volatile solids reduction of 56 percent (with a feed V.S. of 79% typically) but an additional advantage is that the quantity of biogas produced correspondingly increases, it has been perceived that 60% of the energy required for the entire plant is supplied by the anaerobic digesters.



**Anaerobic Reactors**

There are three reactors each with a working volume of 4250 $m^3$, constructed of concrete, which is approximately 200 'mm in thickness. The vessels have a hopper bottom located below ground. The reactors walls and roof are insulated with 150 mm of rockwool, followed by an external GRP cladding, the insulation provided ensures that the temperature within the reactor does not fall by more than 0.5 °C within 24 hr. Each digester is fitted with a spray head in the gas off -take to prevent foam entering the gas system.

**Digester Dimensions**

| | |
|---|---|
| Internal diameter | 17.7 m |
| Side wall depth | 15.4 m |
| Hopper depth | 5.5 m |
| Hopper angle | 35 ° |

**Digester Heating**

As previously mentioned, the digesters are not heated beyond that supplied by the continuous application of digester feed at 40 °C, the reactor temperature is maintained successfully at 37 °C by this method. There are three temperature probes located within each digester to monitor the temperature, which is conveyed to a SCADA system. However, should the temperature fall unexpectedly, unlike in most plants where the boiler or CHP would automatically cut into increase the temperature, in this instance temperature in the digester could only be altered, by feeding at a higher temperature.

**Digester Mixing**

The reactors were designed to facilitate adequate mixing throughout the tank, each digester is mechanically mixed by draft tube, which operates on a continuous basis.

**Gas Collection System**

The expected gas production per day is 35,000 $m^3$. Gas produced during the process is taken from the roof of the digesters and piped to a common gasholder. The gasholder provides buffer capacity for the CHP system, auxiliary steam boilers and new sludge drying plant. The gas holder is a double membrane gas bag, in which the outer membrane is constantly filled with compressed air and the inner membrane stores the biogas, the outer membrane keeps constant pressure on the inner membrane, which semi pressurises the system. The storage capacity is 5000$m^3$. Gas enters and leaves the gasholder *via* condensation traps, all gas lines fall towards the condensation traps. The gasholder is self-contained with level monitors and pressure sensors. The system aims to fully utilise all gas produced, but, in the event of excess gas production, there is a flare stack is on site.

**Flare Design**

| | |
|---|---|
| Flare type | Natural draught concealed burner type |
| Maximum gas flow rate | 1750 N $m^3$/h |
| Chamber Diameter | 2450 mm |

**Digested Sludge Storage / Treatment**

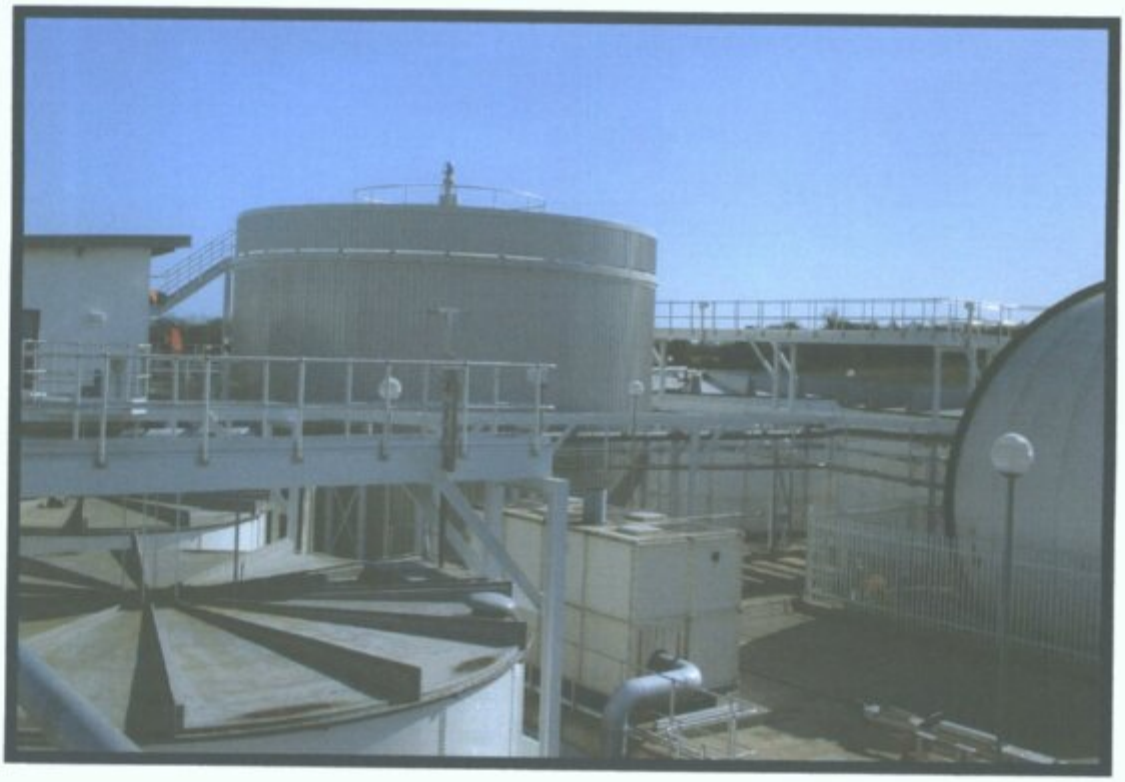
Digester effluent flows to a communal sludge sump, which is undersized, overflowing on occasion. The sludge undergoes further treatment by means of thermal drying.

The digester effluent at approximately 6% T.S. is centrifuged with the aid of polyelectrolyte to 14% T.S. Once dewatered it is then pumped to the Cambi thermal dryer where it undergoes air drying at 450 $^{o}$C approx. the final product specification is for a granular soil like compound of 91 % D.S. approx., but during plant visits the final product was cotton wool like in texture.

**Table 5.10 Ringsend Analysis Results 20-3-03**

| **Feedstock** | | **Digester Sludge** | |
|---|---|---|---|
| T.S. % | 6 | T.S. % | 4 |
| V.S. % | 83.3 | T.S. removal % | 33 |
| pH | 5.9 | V.S.% | 75 |
| | | V.S. removal % | 40 |
| | | pH | 7.9 |
| | | **Thermally dried :** | |
| | | T.S. % | 92.5 |
| | | V.S. % | 65 |

### 5.4.7 Sludge Treatment Plant Summary: Drogheda, Co. Louth



**History**

| | |
|---|---|
| Treatment Technology: | Mesophilic Anaerobic Digestion |
| Year of Construction: | 2001 |
| P.E.: | 67,773 |
| **Feedstock character:** | Primary and Secondary sludge produced on site. |
| Volume/ day: | 120 $m^3$ approx |
| HRT: | 15 days approx. |
| Feeding system | Continuous |

**Plant and Equipment**

| | |
|---|---|
| No. Of Digesters | One |
| Volume: | 1800 $m^3$ (second digester is proposed for the future) |
| Heating system: | External heat exchanger |
| Hot water boilers: | X 1, (Dual fired, operates on biogas and natural gas) |
| CHP unit: | 1 Dual fired |
| Gas Holder | 1 double skinned gas bag (570 $m^3$) |
| Mixing system: | Gas injection |
| Final product storage: | 3 tanks sealed with mixers |
| Final sludge disposal / reuse | Dewatered, followed by lime treatment, applied to non agricultural land |

**Drogheda Sludge Treatment Plant**

**Inlet Works**

The treatment plant serves a p.e. of 67,773, which includes flows from Bryanstown pumping station, East Meath rising main, Newtown pumping station, scum pumping station, supernatant pumping station and storm flow from storm tanks. The design DWF of the inlet is 28,183 m / hour. Influent is screened through rotary belt screens with a 5 mm mesh size, followed by grit removal by coarse bubble diffusers. Fats, oils, and grease (FOG) accumulated during grit removal is transferred by means of grease skimmers to a grease chamber. A set of positive displacement pumps transfers FOG to the anaerobic digester.

**Primary Treatment**

Flow from the inlet is split between two primary settlement tanks. Primary settlement removes 30% BOD and 60% S.S. Desludging of the primary settlement tanks is controlled by means of actuated valves in the primary pumping station. The settled sludge at approx 3% T.S. is pumped to the sealed primary picket fence thickener.

**Secondary Treatment**

The flow is distributed equally over four aeration tanks, the tanks are plug flow systems, configured for full nitrification and partial denitrification of the effluent, by means of incorporation of an anoxic zone. Each basin has an axial flow recirculation pump between the anoxic and aerobic zone (separated by a baffle wall), for internal recirculation of mixed liquor for denitrifcation. Air is supplied to the aerobic section of the tank by fine bubble diffused aeration system. The mixed liquor enters the two clarifiers at the central hopper, through an inlet pipe and the flow is discharged through a submerged bellmouth, located just below the top water level. In order to avoid short-circuiting, the flow is dispersed radially by a diffusion drum. The surplus activated sludge, at 1% T.S. approx. is pumped by three pumps (two duty one standby) to the activated sludge picket fence thickener, which is also a sealed unit.

**Imported Sludge**

Liquid imported sludge can be accepted at the plant. Typically, the imported sludge is received at the sludge reception chamber, where it undergoes 5 mm screening for the

removal of gross solids. The sludge is then pumped forward to the primary picket fence thickener

**Feed Preparation**

Once the respective sludges have been thickened, the sludge is pumped forward to a blending tank and mixed prior entry to the digester. The tank is 9 m in diameter and 4.6 m high, the total volume of the tank is 289.6 $m^3$, but the typical working volume is 165.5 $m^3$ . Mixing is achieved by means of a submersible mixer. The blended sludge at this stage is approximately 4 % T.S. Ferric chloride is added to the digesters recirculation line at a dose of 20 mg/l to reduce the quantity of hydrogen sulphide gas produced, as $H_2S$ is extremely toxic and corrosive. The addition of ferric chloride causes the formation of insoluble salts, which precipitates within the digester and exits the tank in the outgoing effluent, therefore minimising the possibility of CHP and boiler damage. Approximately 120 $m^3$ is fed to the reactor on a continuous basis daily, maintaining a HRT of 15 days within the reactor.

**Anaerobic Digester**

The digester is 14m diameter, 14.37 m high (11.7m of which is above ground) constructed of glass fused to steel panels erected on a concrete base. The tank has an external insulation layer. The working volume of the digester is 1800$m^3$.

**Temperature of Operation**

The digester is maintained at a temperature of 35 °C by means of Alfa Laval water to sludge counterflow spiral heat exchangers, in which the feed sludge is heated and fed to the reactor through the recirculation line. Hot water is supplied to the heat exchanger by a dual fired hot water boiler (380 kW Beeston Broxley cast iron sectional boiler) which operates on biogas or natural gas, and a CHP unit 499 kW which is also dual fired, provides an estimated 162 kW of continuous electrical output and a heat output of up to 275 kW. The electrical power produced by the CHP unit is fed back to the main control panel in the air blower building, where the generated power can be used in parallel with the main incoming electrical supply. The system is designed to be self supporting, except in extremely cold weather where the process can be supplemented with natural gas. In such circumstances, the boiler is designed to operate in series with the CHP to maintain the required output. As well as heating sludge for the digestion process,

domestic heating for the main administration and sludge building is provided by the CHP or Boiler. Excess heat generated in the system can be dumped by means of a thermostatically controlled fan cooled radiator.

**Mixing**

Mixing of the system is by gas recirculation. Gas is taken from the head of the reactor and recirculated at a rate of 440$m^3$ /hour through a vertical lance system, with 25 delivery tubes. The opening and closing of the gas valves is controlled by pneumatic valves, the valve open five at a time to cause hectic mixing within the digester. Mixing is carried out by means of a 22 kW bush mink rotary claw blower. Two blowers are provided to work on a duty and standby basis.

**Gas Collection**

Typically 79.8 $m^3$ of biogas is produced per hour, while the methane concentration is not measured routinely, the $CO_2$ is normally 36% and $H_2S$ is 200 ppm, measured by Dragger tubes. Gas storage is provided by means of a double membrane 570$m^3$ gasbag. The Sattler double membrane gasholder is constructed form three highly resistant, fabric reinforced membranes mounted on a reinforced concrete base. A bottom membrane inflates and deflates with gas production or demand. The exterior shell acts as a resilient shell that enables a support blower to maintain constant pressure between the outer and inner membrane. This constant pressure gives the working pressure of the gas system. Flame arrestors and condensate traps are provided on gas lines as required. Flame arrestors are provided where flame propagation is a potential problem. Condensation traps are provided on low points of the gas lines assisting in protecting downstream equipment from clogging, corrosion, or water hammer. A condensation trap is provided immediately downstream of the digester to remove moisture from the gas as it cools leaving the digester. Gas lines are laid at appropriate falls to accommodate adequate condensate drainage. Over pressure is protected against by means of a vacuum – pressure relief valve complete with a flame arrestor, mounted on the digester roof, and a glycerol filled relief valve provided in the gas bag area.

**Flare Stack**

A gas flare has been provided which consists of a flare tip complete with a windshield, pilot igniter fitted inside the windshield, an automatic ignition system, and a control panel to give automatic control and monitoring of the system.

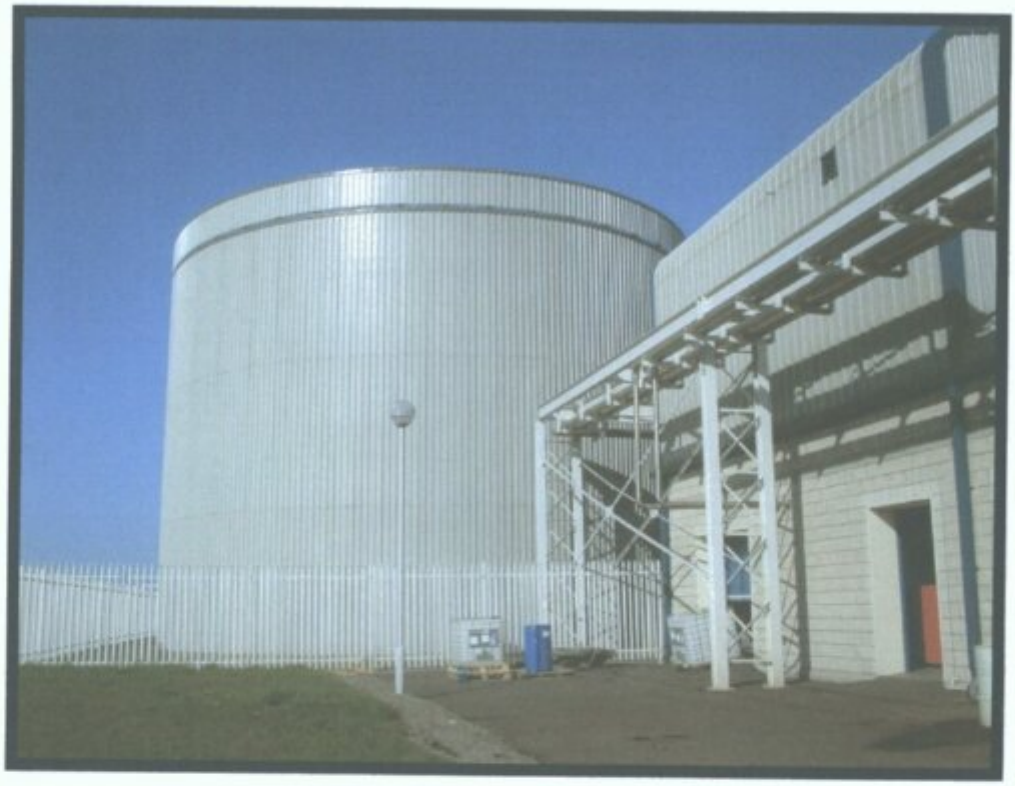
**Final Sludge Storage**

Digester effluent overflows to the draw off weir simultaneously during feeding, and flows by gravity to one three sludge holding tanks. Each tank has a capacity of 1924 $m^3$, and incorporates a mixer. The digester sludge is held in storage for 15 days, thus the system is designed so that one tank is filling, one tank emptying and one tank is allowed to stand for 15 days. The final sludge is considered by the plant manager as a 'Class C', however; mesophilic anaerobic digestion achieves a Class B Biosolid in terms of pathogen reduction, thus one can assume that the sludge contains high concentrations of heavy metals. In order to obtain 'Class B' status the digester effluent is further treated by dewatering followed by lime stabilisation (unheated) to a final TS concentration of approximately 30.6%, the sludge is then applied to non-agricultural use land.

**Table 5.11 Drogheda Analysis Results 11/'01 – 4/'03**

| **Feedstock** | | **Digester Sludge** | |
|---|---|---|---|
| T.S. % | 3.6 | T.S. % | 1.87 |
| | | T.S. removal % | 48 |
| | | *pH* | 6.92 |
| | | VFA mg/l acetic acid | 52.68 |
| | | Alkalinity mg/l $CaCo_3$ | 4,372 |
| | | Biogas vol. $m^3/d$ | 773.54 |

### 5.4.8 Sludge Treatment Plant Summary: Dundalk, Co. Louth



**History**

| | |
|---|---|
| Treatment Technology: | Mesophilic Anaerobic Digestion |
| Year of Construction: | 2001 - 2002 |
| P.E.: | 180,000 (Current 110,000 2/3 from industry) |
| **Feedstock character:** | Secondary sludge produced on site |
| Volume/ day: | 110 $m^3$ / digester approx |
| HRT: | 20 days approx. |
| Feeding system | Continuous |

**Plant and Equipment**

| | |
|---|---|
| No. Of Digesters | Two |
| Volume: | 2200 $m^3$ / digester |
| Heating system: | External heat exchangers |
| Hot water boilers: | X 2 , (Dual fired, operates on biogas and natural gas) |
| CHP unit: | 1 Dual fired |
| Gas Holder | 1 Double membrane gas bag (capacity 17202 m3) |
| Mixing system: | Gas injection |
| Final product storage: | 4 tanks (submerged mixers in each tank) |
| Final sludge disposal / reuse | Dewatered, by 2 belt presses to 20 % T.S., and disposed to landfill. |

**Dundalk Sludge Treatment Plant**

The plant consists of a two –stage system of biological oxidation. The incoming flow is split into three and passes through fine screens and a grit trap prior aeration. There are three primary aeration basins 18600 mm x 22500 mm x 5000 mm deep, each basin has a single aeration lane (5800 mm wide x 22500 mm long). Aeration is achieved by fine bubble diffused aeration. Following primary aeration, 50% of the organic load has been removed. The Mixed liquor is fed forward to two primary clarifiers 30000 mm in diameter side wall 15000 mm and 3732 mm deep, where the mixed liquor is allowed to settle (producing a sludge). The supernatant flows to three secondary diffused aeration basins 50000 mm x 22740 mm x 5000 mm deep, each with a single aeration lane of 7180 mm in width x 22740 mm in length, for further oxidation. The mixed liquor is again fed forward to two clarifiers for sedimentation, and the final effluent is discharged to the bay.

**Feed Preparation**

Settled solids from the primary clarifiers and the secondary clarifiers are pumped forward to the correspondingly named picket fence thickeners, through an actuator chamber, which essentially allows the operator to decide which thickener is to be used. The picket fence thickeners are sealed tanks constructed from glass lined steel panels on a reinforced concrete base. The tanks are circular, with a hopper bottom; A larger volume of sludge is produced from primary aeration stage as the picket fence thickener in this case is larger in size, as can be seen from the following:

| | **Primary aeration PFT** | **Secondary aeration PFT** |
|---|---|---|
| Internal diameter | 21340 mm | 16220 mm |
| Side wall depth | 3518 mm | 3668 mm |
| Overall depth | 4898 mm | 4675 mm |

The sludge is thickened to approximately 4% T.S. by PFT, from the respective PFT the sludge is fed directly to the digesters without prior blending. The sludge is heated, dosed with ferric chloride to reduce $H_2S$ concentration in the gas and administered to the digester on a continuous bases at a rate of approximately 180 $m^3$/d, which facilitates a HRT of 20 days. The plant, as a designated hub centre for Co. Louth, will accept imported sludge in the future, but, at present, there is no reception facility on site for such sludges.

**Digesters**

There are two digesters on site, but at present just one is in operation. The digesters are constructed from glass lined steel panels on a reinforced concrete base, and insulated with 100 mm of polyurethane foam and an outer cladding. The working volume of each digester is 2200 $m^3$ approx. The digester lid is fitted with a pressure release valve in case of emergency.

Digester dimensions

| | |
|---|---|
| Circular hopper bottom | 45 $^{o}$ slope |
| Internal diameter | 18770 mm |
| Side wall depth | 15130 mm |
| Overall depth | 19400 mm |

**Digester Heating**

The digester is maintained at the mesophilic temperature of 35 $^{o}$C, by an external heat exchange system. There are two (one per digester) sludge to water spiral heat exchangers, capable of heating up to 70 $m^3$ / hr. Typically, the sludge entering the heat exchanger is composed of sludge recirculated from the digester and injected feed sludge, thus the incoming temperature is 30 – 35 $^{o}$C, the sludge on exit from the heat exchanger ranges in temperature from 35. 9 – 40.9 $^{o}$C that successfully maintains the internal digester temperature at 35 $^{o}$C. There are two water-to-water heat exchangers, the incoming temperature of the water being 80 $^{o}$C and the outgoing temperature 65.8$^{o}$C.

Hot water required for operation of the heat exchangers is derived from two dual fired boilers each supplying 275 kW; the boilers can be run on methane or natural gas. There is one CHP unit, 300 kW, which converts methane gas to electricity, the electricity produced is fed back to the main control panel of the plant.

The digesters have also been supplied with an emergency cooling system, consisting of two water-to-water heat exchangers, which will automatically ditch heat should the digester temperature rise above 39 °C in the summer months.

**Mixing**

Mixing of the system is by gas recirculation. Gas is taken from the head of the reactor and recirculated through a vertical lance system, with 25 delivery tubes. The opening and closing of the gas valves is controlled by pneumatic valves, the valve open randomly, five at a time, within the digester.

**Biogas Collection**

Biogas produced during the process exits through the lid of the digesters and flows to a communal double membrane gasbag, the volume of the gasbag is 17020 $m^3$. The double membrane gasholder is constructed form three highly resistant, fabric reinforced membranes mounted on a reinforced concrete base. A bottom membrane inflates and deflates with gas production or demand. The exterior shell acts as a resilient shell that enables a support blower to maintain constant pressure between the outer and inner membrane. This constant pressure gives the working pressure of the gas system. Condensate traps are provided on all gas lines. The volume of gas produced is not measured, and the gas is not analysed for methane content beyond a flame test conducted daily

**Flare Stack**

A gas flare has been provided on site to burn off excess biogas safely.

**Final Sludge Storage**

Digester effluent overflows to the draw off weir simultaneously during feeding, and flows by gravity to one four sludge holding tanks complete with submersible mixers, of

the following dimensions:

| | |
|---|---|
| Diameter | 18.72 m |
| Slope of tank bottom | 8 $^{o}$ |
| Height straight side | 4.2 m |
| Height pitcher roof | 1.65 m |
| Depth of hopper bottom | 1.32 m |
| Concrete stand | 0.15 m |
| Overall height above ground | 6.07 m |

The final sludge is dewatered to 20 % by belt press and disposed of to landfil.

**Table 5.12 Dundalk Analysis Results 9/'02 – 3/'03**

| **Feedstock** | | **Digester Sludge** | |
|---|---|---|---|
| T.S. % | 3.88 | pH | 7.44 |
| V.S. % | 69.36 | VFA mg/l acetic acid | 67.59 |
| V.S. % | 69.36 | T.S. % | 2.59 |
| | | V.S. % | 57.09 |
| | | T. S. removal % | 33 |
| | | V.S. removal % | 40 |

### 5.4.9 Sludge Treatment Plant Summary: Tralee, Co. Kerry

**History**

| | |
|---|---|
| Treatment Technology: | Mesophilic Anaerobic Digestion |
| Year of Construction: | 1998 |
| P.E.: | Design 42,000, actual 32,000 |
| **Feedstock character:** | Designated Hub Centre, but at present treating primary and secondary sludge produced on site only. |
| Volume/ day: | 23 $m^3$ approx |
| HRT: | 20 days approx. |
| Feeding system | Continuous |

**Plant and Equipment**

| | |
|---|---|
| No. Of Digesters | Two (one commissioned and in use) |
| Volume: | 400 $m^3$ / digester |
| Heating system: | External heat exchanger |
| Hot water boilers: | X 2, (Dual fired Methane and propane) |
| CHP unit: | 1 gas engines |
| Gas Holder | 2 Bell over water system, 25 $m^3$ |
| Mixing system: | Gas injection |
| Final product storage: | 2 Open top storage tank |
| Final sludge disposal / reuse | Centrifuged to 22% T.S. approx., and land spread typically |

**Tralee Anaerobic digester**

Tralee is a costal town with a population of 25,000 approx. and increasing. The town, like Killarney is a popular tourist location. Up until 1998 there was no sewage treatment provided and wastewater was discharged directly to the sea.

**Design of Anaerobic digestion Plant**

The sludge Hub centre was designed for a p.e. of 42,000. The system consists of two anaerobic digesters each with a capacity of 400m$^3$. To date only one digester has been commissioned. The reactors are cylindrical vessels, 8.54 m diameter and 8 m in height, of which 4 m is underground, the above ground section is constructed of glass lined section steel, while the underground sub-structure is constructed of concrete. The walls and roof of the reactors are insulated with 50 mm polyurethane foam, and rigid GRP encapsulates the reactors to prevent external corrosion. The decision to submerge half of the reactors volume underground was not only to mitigate the visual impact of the plant on the environment but also for the purpose of further insulation.

**Digester Heating**

Plate:5.1 Heat exchanger unit

Plate:5.2 Gas boilers

Plate:5.3 CHP unit

The reactors are maintained at the mesophilic temperature of 35 °C approx. Heating is achieved by means of external heat exchangers 'Concentric – Tube hot water heat exchangers'. The sludge is pumped by sludge recycle pump from the reactor through the heat exchanger, this system would therefore also aid in mixing of the digester. The hot water used in the heat exchanger can be obtained from the boiler or CHP unit. Biogas is

taken from the gas-holder and compressed by two rotary vane compressors and is used to fuel the CHP unit (Combined heat and power) which heats the water. However, if biogas production is insufficient the dual fired boilers are operated on propane gas to heat the water. Typically, both systems operate in sequence as the dual fired boiler automatically comes into operation when biogas quantity is insufficient. Temperature probes within the reactors control the digesters heating, through the SCADA computer based monitoring system.

Figure 5.8 Tralee Anaerobic Digester Temperature of Operation




**Digester Mixing**

Plate: 5.4 External view of gas distribution pipes

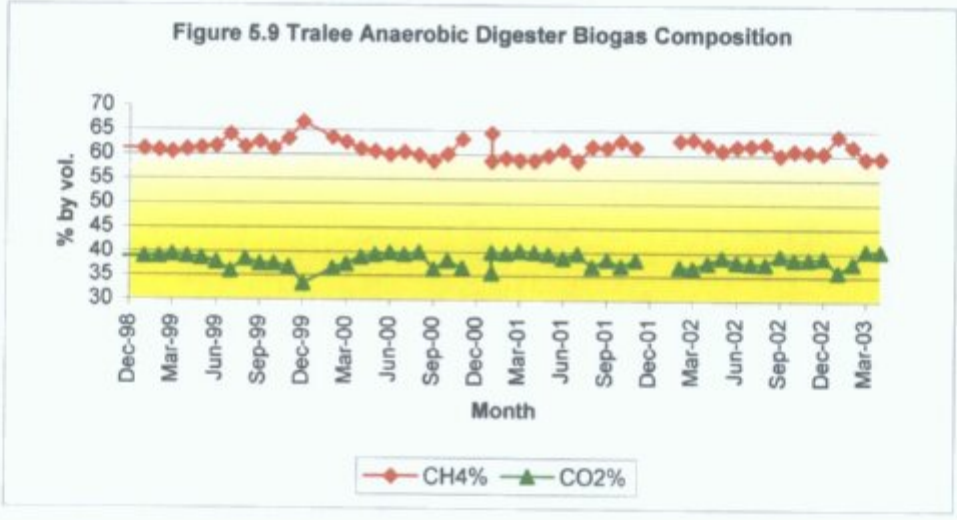
The sludge digestion tanks are mixed using compressed biogas. A distribution pipe enters at the top of the digester and drops to the bottom of the tank, where the compressed biogas is evenly distributed through a series of six separate unconfined gas mixing pipes situated around the circumference of the digester.

### Biogas Collection

Biogas generated during the process leaves the digester through the roof under its own pressure, and is diverted to a gas holding tank. There are separate gas holding tanks for each of the digesters, each with a 25m$^3$ capacity. The gas holders are the typical bell-over-water design equipped with breathing safety valves and level switches which are connected to a flare stack, to burn off excess biogas if the tank is full. Each holder has three condensate pots, one coming from the digester to the gas holder, one before entering the boiler, and finally one for gas leaving the gas holder before entering the other gas holder (in case of emergency).

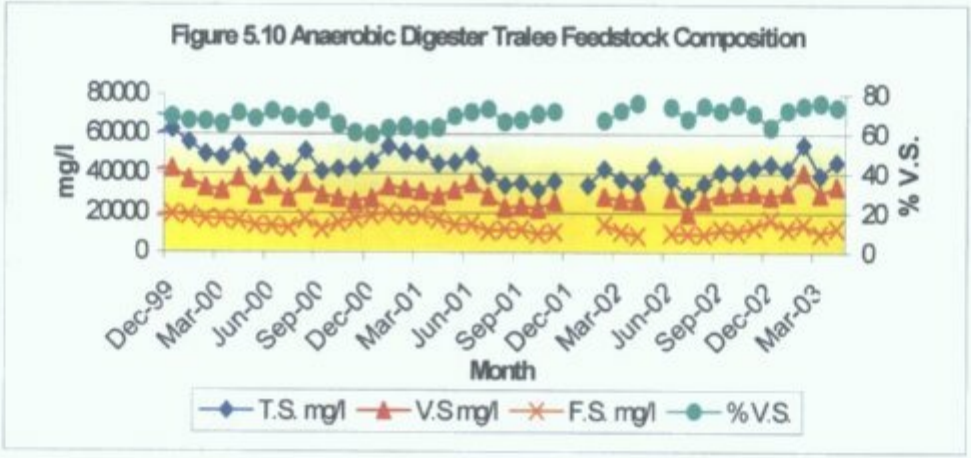
### Biogas Measurement

The biogas quality is measured daily by LMSX multi gas analyser, the quantity however is not measured on a regular basis, but the theoretical biogas yield for the system is estimated to be 622m$^3$/d.




### Feed Preparation

The anaerobic digester at present treats sludge produced on site only. The feedstock is a mixed substrate, which includes sludge from two primary settlement tanks and secondary sludge from 2 trickling filters and 3 diffused aeration tanks. Following settlement, both sludges are pumped to an open top picket fence thickener with 300m$^3$ capacity where they are gravity thickened to 4-6% T.S. Typical feedstock composition is illustrated in figure 5.9.




**Feed / Effluent Withdrawal Regime**

The designed flow through of the digester ranges from 26.6 $m^3$ /day (HRT 15 D with 1 reactor in operation) to 54 $m^3$ / day (HRT 14.8 D with 2 reactors in operation). However, anaerobic digesters can safely operate at HRT of 12 D and therefore up to 66.6m3 /D could be applied. The digesters are operated as single stage CSTR. Sludge is fed to the digesters on a time-controlled basis. The feed lines are equipped with flow measurement. Feed enters the digester in the recirculation line after the heat exchanger; therefore cold sludge combines with the heated sludge before entering the digester. Digested effluent is displaced by the incoming sludge and flows by gravity to the storage tanks where it is stored for a minimum of 14 Days prior to dewatering.

The average tonnage of dry solids produced in Tralee averages at 329.34 tds. Figure 5.10, and 5.11 conveys influent and effluent quality and percentage removal for the plant at Tralee.

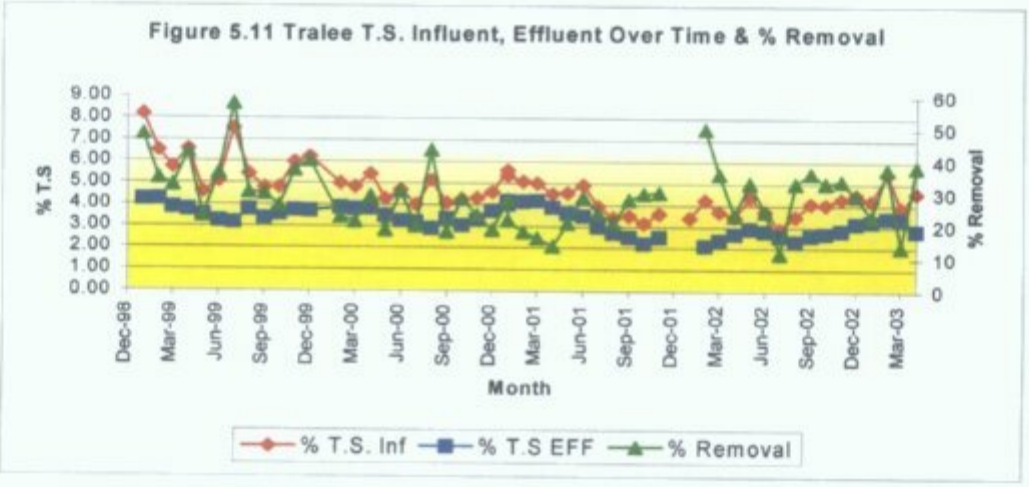


Figure 5.12 Tralee V.S. Influent, Effluent Over Time & % Removal

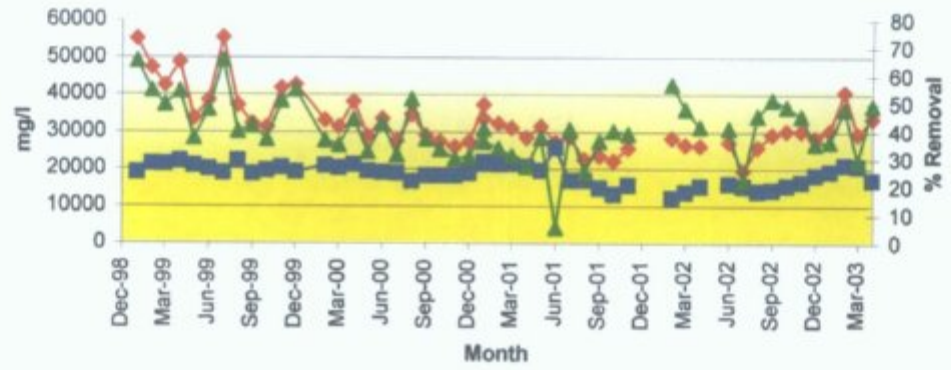
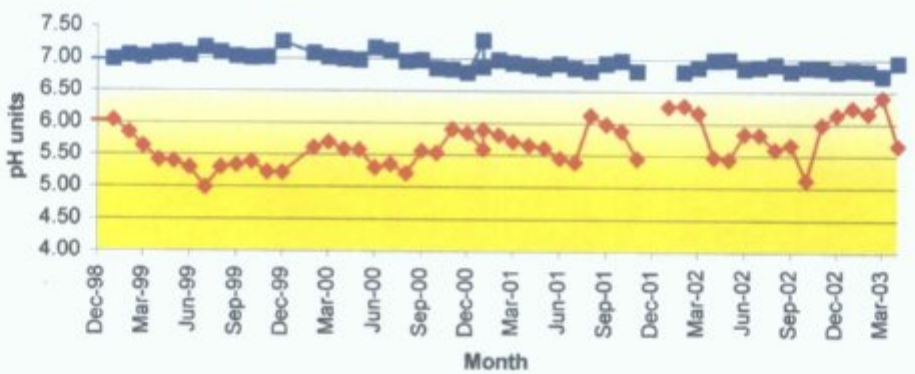



Figure 5.13 Tralee pH Influent, Effluent Over Time






**Dewatering and Disposal**

Digested sludge is pumped from the storage tank to the dewatering room. The effluent is dewatered by one of two centrifuges operating in alternation. The sludge is dewatered to 22% TS with the aid of polyelectrolyte Zetag 64 and is conveyed to a covered skip, where it awaits transportation for reuse on agricultural land.

**Cost of Treatment**

| | |
|---|---|
| Capital cost (estimate) | €3,868,166 |
| **Operational Costs:** | |
| ESB usage 384 kW/d | €16,044 / annum |
| Propane gas | €10,400 / annum |
| Dewatering of final product sludge | € 20,500 / annum |
| Transport to agricultural land | €22,500 / annum |
| **Total Cost of Treatment** | €69,544 / annum |
| Cost of Treatment per Tonne Dry Solids | € 211.16 |

Table 5 .13 (A) Anaerobic Digester Tralee Results of Analysis

**Feedstock characteristics and operating parameters**

| Date | Feed rate $m^3/d$ | HRT (days) | Temp °C | pH | %T.S. | T.S range | %V.S. | COD mg/l | $NH_3$ mg/l | $CaCO_3$ mg/l |
|---|---|---|---|---|---|---|---|---|---|---|
| Dec-98 | 4.83 | 82.76 | 34.2 | 5.70 | | | | 7900 | 398 | |
| Jan-99 | 6.03 | 66.29 | 32.4 | 6.03 | 8.18 | 6.59 - 10.43 | 67.34 | 3453 | 153 | 5179 |
| Feb-99 | 10.14 | 39.44 | 31.7 | 5.85 | 6.50 | 5.05-7.33 | 72.78 | 4824 | 292 | 5259 |
| Mar-99 | 19.26 | 20.77 | 30.6 | 5.63 | 5.73 | 3.62-7.12 | 74.26 | 5682 | 349 | 4067 |
| Apr-99 | 18.25 | 21.92 | 29.7 | 5.42 | 6.57 | 5.13-7.23 | 74.35 | 8632 | 417 | 3858 |
| May-99 | 16.40 | 24.39 | 29.8 | 5.40 | 4.59 | 3.76-5.44 | 73.30 | 7351 | 352 | 2872 |
| Jun-99 | 16.62 | 24.07 | 30.5 | 5.30 | 5.08 | 4.45-5.69 | 75.61 | 7012 | 186 | 1920 |
| Jul-99 | 5.39 | 74.19 | 33.7 | 4.99 | 7.53 | 5.12-9.15 | 73.47 | 11946 | 239 | 2864 |
| Aug-99 | 18.45 | 21.68 | 32.1 | 5.31 | 5.42 | 3.9-8.07 | 68.57 | 8872 | 270 | 1995 |
| Sep-99 | 18.00 | 22.22 | 29.5 | 5.34 | 4.78 | 4-6.01 | 67.48 | 8140 | 220 | 1457 |
| Oct-99 | 14.91 | 26.83 | 27.6 | 5.39 | 4.81 | 4.5-7.3 | 64.96 | 7493 | 229 | 2002 |
| Nov-99 | 14.25 | 28.07 | 25.5 | 5.23 | 5.97 | 3.51-6.7 | 70.05 | 9738 | 265 | 1525 |
| Dec-99 | 6.95 | 57.55 | 31.9 | 5.22 | 6.22 | 4.3-6.8 | 68.59 | 11503 | 561 | 2544 |
| Jan-00 | 13.97 | 28.64 | 33.4 | 5.59 | 5.60 | 5.2-8.8 | 66.62 | 9167 | 239 | 2080 |
| Feb-00 | 19.79 | 20.22 | 33.5 | 5.62 | 4.99 | 4.1-6.4 | 66.25 | 10365 | 201 | 1776 |
| Mar-00 | 20.10 | 19.90 | 35.1 | 5.70 | 4.83 | 3.6 -5.8 | 64.95 | 11732 | 224 | 1987 |
| Apr-00 | 22.37 | 17.88 | 35.1 | 5.59 | 5.41 | 4.3-6.2 | 70.33 | 8798 | 214 | 2266 |
| May-00 | 21.56 | 18.55 | 35.0 | 5.58 | 4.28 | 4-4.7 | 67.46 | 11182 | 221 | 2364 |
| Jun-00 | 21.54 | 18.57 | 34.6 | 5.30 | 4.69 | 3.99-6.29 | 71.40 | 4790 | 167 | 1440 |
| Jul-00 | 22.39 | 17.87 | 35.3 | 5.36 | 4.00 | 3.81-4.58 | 68.98 | 5470 | 189 | 1183 |
| Aug-00 | 23.59 | 16.96 | 35.0 | 5.21 | 5.14 | 4.6- 6 | 67.49 | 5710 | 84 | 1089 |
| Sep-00 | 20.95 | 19.09 | 34.8 | 5.56 | 4.10 | 2.88-4.7 | 71.10 | 4096 | 225 | 1287 |
| Oct-00 | 22.78 | 17.56 | 34.8 | 5.53 | 4.25 | 3-6.58 | 64.66 | 3079 | 149 | 1287 |
| Nov-00 | 22.93 | 17.44 | 36.1 | 5.90 | 4.31 | 3.65-4.94 | 60.40 | 1440 | 69 | |
| Dec-00 | 23.02 | 17.38 | 34.8 | 5.85 | 4.60 | 4.24-5.06 | 59.47 | 2677 | 82 | 2397 |
| Jan-01 | 23.15 | 17.28 | 34.8 | 5.89 | 5.36 | 4.73-6.37 | 62.58 | 3808 | 125 | 2768 |

Table 5.13 (B) Anaerobic Digester Tralee Results of Analysis

| | Digester Sludge | | | | | | | | Biogas | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Date | %T.S. | T.S. range | % V.S. | COD mg/l | $NH_3$ mg/l | pH | VFA mg/l | $CaCO_3$ mg/l | $CH_4$ % | $CO_2$ % | $m_3$/d |
| Dec-98 | | | | 4655 | 1181 | 6.99 | 357 | | 66 | 34 | |
| Jan-99 | 4.22 | 4.04-4.37 | 45.14 | 4770 | 1273 | 6.98 | 343 | 7807 | 61 | 39 | 246 |
| Feb-99 | 4.24 | 3.93 - 4.24 | 50.41 | 4718 | 1611 | 7.05 | 96 | 8634 | 61 | 39 | 294 |
| Mar-99 | 3.85 | 3.71-4.01 | 55.28 | 5106 | 1564 | 7.02 | 134 | 7928 | 60 | 39 | 400 |
| Apr-99 | 3.75 | 3.65-3.84 | 59.09 | 4742 | 1761 | 7.07 | 194 | 7554 | 61 | 39 | |
| May-99 | 3.53 | 3.21-3.89 | 59.23 | 4448 | 1536 | 7.08 | 80 | 6995 | 61 | 39 | |
| Jun-99 | 3.23 | 3.09-3.38 | 61.91 | 3908 | 1134 | 7.04 | 86 | 5220 | 62 | 38 | |
| Jul-99 | 3.17 | 2.99-3.4 | 59.52 | 5063 | 1248 | 7.16 | 568 | 5786 | 64 | 36 | |
| Aug-99 | 3.76 | 3.51-3.92 | 59.14 | 5494 | 1197 | 7.09 | 474 | 5701 | 62 | 38 | |
| Sep-99 | 3.34 | 2.5-3.63 | 55.74 | 3223 | 990 | 7.03 | 453 | 4729 | 63 | 37 | |
| Oct-99 | 3.54 | 3.2 -3.8 | 55.21 | 2578 | 1469 | 7.01 | 383 | 4830 | 61 | 38 | |
| Nov-99 | 3.74 | 3.5-3.9 | 54.63 | 3015 | 1443 | 7.02 | 438 | 5751 | 63 | 37 | |
| Dec-99 | 3.72 | 3.4-5.5 | 51.62 | 3771 | 1913 | 7.26 | 333 | 7321 | 67 | 33 | |
| Jan-00 | 4.01 | 3.7-6.51 | 54.72 | 4037 | 1963 | 7.28 | 240 | 7381 | 64 | 36 | 35 |
| Feb-00 | 3.85 | 3.7-4 | 54.12 | 4121 | 2017 | 7.08 | 245 | 6300 | 63 | 37 | |
| Mar-00 | 3.79 | 3.6-3.91 | 53.56 | 4784 | 1641 | 7.02 | 465 | 6208 | 63 | 37 | |
| Apr-00 | 3.82 | 3.7-3.95 | 55.16 | 4669 | 1638 | 6.99 | 252 | 5695 | 61 | 39 | |
| May-00 | 3.47 | 3.2-3.9 | 55.80 | 4610 | 2185 | 6.98 | 384 | 5379 | 61 | 39 | |
| Jun-00 | 3.24 | 3.1-3.34 | 58.79 | 2907 | 1043 | 7.17 | 354 | 4701 | 60 | 40 | |
| Jul-00 | 3.20 | 3.04-3.17 | 58.93 | 3239 | 947 | 7.12 | 277 | 4428 | 60 | 40 | |
| Aug-00 | 2.90 | 1.69-3.22 | 57.64 | 3050 | 917 | 6.95 | 213 | 4346 | 60 | 40 | |
| Sep-00 | 3.35 | 2.77-3.89 | 54.33 | 2275 | 739 | 6.98 | 99 | 3918 | 59 | 36 | |
| Oct-00 | 3.03 | 2.82-3.07 | 59.93 | 2226 | 1009 | 6.85 | 53 | 3942 | 60 | 38 | |
| Nov-00 | 3.30 | 2.76-3.83 | 54.99 | 2438 | 890 | 6.83 | 40 | 4395 | 63 | 37 | |
| Dec-00 | 3.73 | 3.6-3.83 | 50.61 | 2590 | 818 | 6.78 | 45 | 5099 | | | |
| Jan-01 | 4.18 | 3.96-4.43 | 50.91 | 3233 | 1049 | 6.87 | 129 | 5340 | 59 | 40 | 253 |

Table 5 . 13 (C) Anaerobic Digester Tralee Results of Analysis

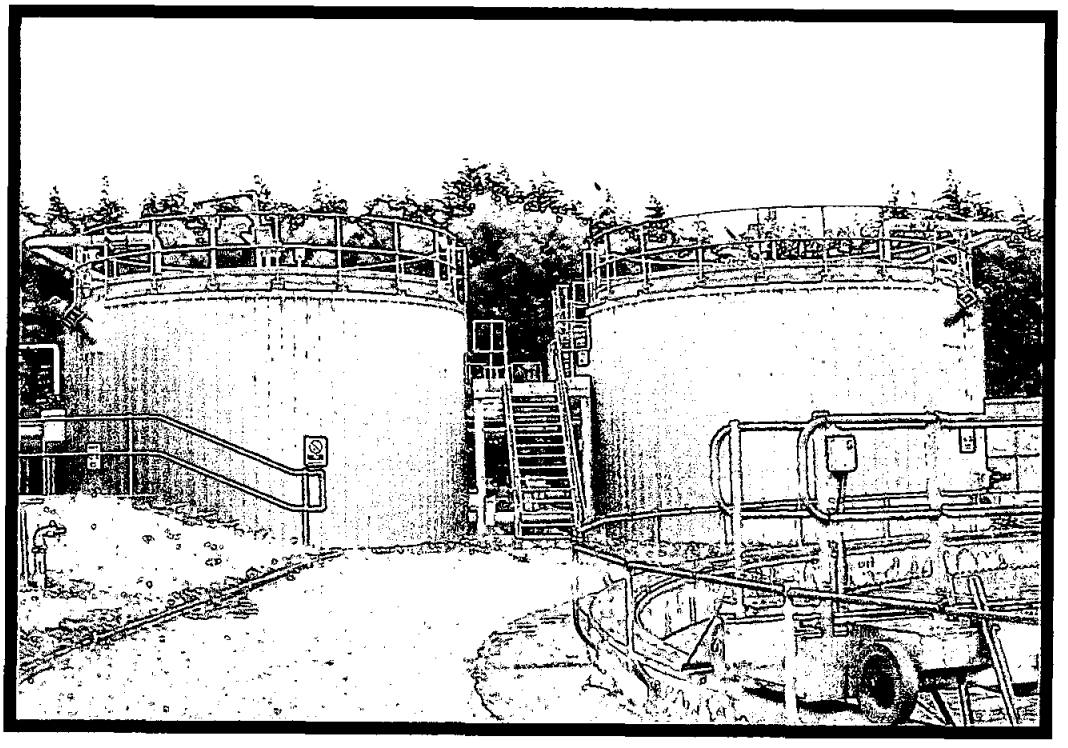
**Feedstock characteristics and operating parameters**

| Date | Feed rate m3/d | HRT (da)ys | Temp | pH | %T.S | T.S range | %V.S | COD mg/l | NH3 mg/l | CaCO3 mg/l |
|---|---|---|---|---|---|---|---|---|---|---|
| Feb-01 | 21.99 | 18.19 | 35.0 | 5.82 | 5.09 | 4.7-5.73 | 63.42 | 3993 | 171 | 2247 |
| Mar-01 | 22.88 | 17.49 | 35.5 | 5.71 | 5.00 | 4.45-5.63 | 62.08 | 3655 | 128 | 2720 |
| Apr-01 | 22.57 | 17.72 | 35.5 | 5.65 | 4.51 | 4.05-5 | 63.14 | 2445 | 119 | 2076 |
| May-01 | 16.33 | 24.50 | 35.2 | 5.61 | 4.56 | 4.11-4.96 | 69.00 | 5610 | 115 | 1824 |
| Jun-01 | 14.53 | 27.54 | 35.0 | 5.45 | 4.93 | 2.26-7.69 | 71.00 | 6025 | 173 | 959 |
| Jul-01 | 21.20 | 18.87 | 34.8 | 5.39 | 3.94 | 3.2-4.74 | 72.57 | 3882 | 92 | 884 |
| Aug-01 | 22.26 | 17.97 | 34.6 | 6.12 | 3.42 | 2.25-7.27 | 65.88 | 997 | 58 | 983 |
| Sep-01 | 13.58 | 29.46 | 34.9 | 5.98 | 3.51 | 1.83-4.76 | 66.90 | 4691 | 208 | 1331 |
| Oct-01 | 7.33 | 54.55 | 36.9 | 5.88 | 3.17 | 1.75-4.19 | 69.89 | 3862 | 173 | 1417 |
| Nov-01 | 18.56 | 21.56 | 32.7 | 5.45 | 3.61 | 3.14-4.89 | 71.13 | 2983 | 449 | 891 |
| Dec-01 | 23.90 | 16.73 | | | | 2.91-4.26 | | | | |
| Jan-02 | 14.03 | 28.50 | | 6.26 | 3.42 | 1.71-7.5 | | | | |
| Feb-02 | 7.64 | 52.34 | 32.8 | 6.28 | 4.23 | 1.67-5.29 | 66.78 | 1491 | 54 | 1373 |
| Mar-02 | 7.71 | 51.85 | 35.7 | 6.16 | 3.70 | 1.66-5.16 | 71.56 | 3609 | 126 | 1467 |
| Apr-02 | 11.28 | 35.47 | 34.2 | 5.48 | 3.47 | 3.33-3.81 | 75.58 | 7328 | 220 | 1294 |
| May-02 | 15.83 | 25.26 | 35.7 | 5.44 | 4.37 | 4.13-4.5 | | 3930 | | |
| Jun-02 | 19.14 | 20.89 | 31.9 | 5.84 | 3.70 | 3.19-4.59 | 73.56 | 2982 | 142 | 1056 |
| Jul-02 | 14.60 | 27.40 | 34.6 | 5.83 | 2.91 | 1.63-3.86 | 67.41 | 5108 | 147 | 1092 |
| Aug-02 | 14.33 | 27.91 | 34.8 | 5.60 | 3.49 | 2.69-4.17 | 74.07 | 4125 | 104 | 1616 |
| Sep-02 | 21.73 | 18.41 | 34.0 | 5.67 | 4.09 | 1.87-9.02 | 71.63 | 3511 | 171 | 1024 |
| Oct-02 | 17.42 | 22.96 | 32.7 | 5.13 | 4.04 | 3.55-5.82 | 74.82 | 3647 | 339 | 993 |
| Nov-02 | 18.45 | 21.68 | 35.0 | 5.99 | 4.29 | 3.62-4.87 | 70.18 | 2787 | 124 | 1095 |
| Dec-02 | 23.22 | 17.22 | 33.8 | 6.14 | 4.49 | 4.23-4.85 | 63.22 | 2709 | 218 | 1335 |
| Jan-03 | 22.00 | 18.18 | 35.4 | 6.24 | 4.22 | 3.18-7.03 | 71.78 | 1668 | 74 | 1286 |
| Feb-03 | 23.43 | 17.07 | 34.4 | 6.16 | 5.48 | 4.61-7.89 | 74.12 | 2617 | 90 | 1237 |
| Mar-03 | 24.53 | 16.30 | 29.7 | 6.41 | 3.89 | 1.88-6.1 | 75.67 | 2859 | | 1597 |
| Apr-03 | 23.17 | 17.27 | 31.4 | 5.67 | 4.57 | 4.41-4.75 | 73.33 | 3529 | | 992 |
| **Mean** | **18.37** | **26.36** | **33.5** | **5.89** | **4.68** | | **70.64** | **5706** | **209** | **1920** |

Table 5 . 13 (D) Anaerobic Digester Tralee Results of Analysis

| | Digester Sludge | | | | | | | | Biogas | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Date | % T.S. | T.S range | % V.S. | COD mg/l | NH3 mg/l | pH | VFA mg/l | CaCO3 | CH4% | CO2% | m3/D |
| Feb-01 | 4.16 | 4.06-4.36 | 51.07 | 4329 | 1300 | 6.98 | 133 | 6277 | 59 | 40 | 231 |
| Mar-01 | 4.18 | 4.02-4.28 | 50.87 | 4507 | 1045 | 6.93 | 119 | 6057 | 59 | 40 | 242 |
| Apr-01 | 3.88 | 3.67-4.18 | 53.24 | 3948 | 1068 | 6.90 | 92 | 5488 | 59 | 40 | 164 |
| May-01 | 3.61 | 3.45-3.73 | 54.14 | 2725 | 936 | 6.86 | 81 | 5065 | 60 | 40 | 298 |
| Jun-01 | 3.50 | 3.16-4 | 73.72 | 4542 | 1037 | 6.92 | 145 | 4134 | 61 | 39 | 317 |
| Jul-01 | 2.99 | 2.65-3.41 | 56.52 | 3814 | 941 | 6.85 | 226 | 3795 | 59 | 40 | 210 |
| Aug-01 | 2.70 | 2.6-2.83 | 62.34 | 2372 | 625 | 6.81 | 245 | 3109 | 62 | 37 | 89 |
| Sep-01 | 2.52 | 2.31-2.76 | 58.97 | 3438 | 889 | 6.93 | 503 | 2917 | 62 | 39 | 122 |
| Oct-01 | 2.21 | 2.06-2.53 | 59.69 | 3712 | 1061 | 6.97 | 300 | 5190 | 63 | 37 | 51 |
| Nov-01 | 2.51 | 2.33-2.64 | 62.29 | 2851 | 1105 | 6.81 | 458 | 3094 | 62 | 38 | 139 |
| Dec-01 | | 2.25-2.71 | | | | | | | | | |
| Jan-02 | | 1.84- 2.85 | | | | | | | | | |
| Feb-02 | 2.11 | 1.92 - 2.35 | 57.57 | 2189 | 841 | 6.80 | 477 | 3233 | 63 | 37 | |
| Mar-02 | 2.36 | 2.14-2.91 | 58.20 | 3242 | 1015 | 6.87 | 174 | 3514 | 63 | 37 | |
| Apr-02 | 2.65 | 2.26-2.96 | 57.51 | 4688 | 1294 | 6.98 | 130 | 3854 | 62 | 38 | |
| May-02 | 2.91 | 2.86-2.98 | | 5135 | | 6.99 | 145 | | 61 | 39 | |
| Jun-02 | 2.78 | 2.63 - 2.98 | 57.47 | 2841 | 927 | 6.85 | 246 | 3245 | 62 | 38 | |
| Jul-02 | 2.57 | 2.27-2.7 | 59.40 | 2581 | 1436 | 6.88 | 365 | 5122 | 62 | 38 | |
| Aug-02 | 2.32 | 2.09 - 2.65 | 60.29 | 2896 | 823 | 6.92 | 448 | 3270 | 62 | 38 | |
| Sep-02 | 2.59 | 2.19 - 2.97 | 54.85 | 2783 | 750 | 6.82 | 550 | 2881 | 60 | 39 | |
| Oct-02 | 2.69 | 2.38-4.03 | 57.42 | 3165 | 796 | 6.88 | 412 | 3523 | 61 | 39 | |
| Nov-02 | 2.83 | 0.81 - 3.05 | 57.92 | 3301 | 1002 | 6.87 | 309 | 3291 | 61 | 39 | |
| Dec-02 | 3.16 | 3.05 - 3.25 | 57.69 | 3190 | 924 | 6.82 | 248 | 3600 | 61 | 39 | |
| Jan-03 | 3.24 | 3.02-4.12 | 59.43 | 2689 | 786 | 6.84 | 220 | 3041 | 64 | 36 | |
| Feb-03 | 3.41 | 3.28 - 3.64 | 61.78 | 2554 | 889 | 6.83 | 314 | 3358 | 62 | 38 | |
| Mar-03 | 3.36 | 1.58 - 6.1 | 61.98 | 2859 | | 6.75 | 433 | 3571 | 59 | 41 | |
| Apr-03 | 2.81 | 1.27 - 3.27 | 60.50 | 3129 | | 6.95 | 688 | 3640 | 60 | 40 | |
| **Mean** | **3.41** | | **56.61** | **3738** | **1231** | **6.95** | **278** | **4889** | **61** | **39** | **206.06** |

### 5.4.10 Sludge Treatment Plant Summary: Killarney, Co. Kerry



**History**

| | |
|---|---|
| Treatment Technology: | Autothermal Thermophilic Aerobic Digestion |
| Year of Construction: | 2001 |
| P.E.: | Design 42,000 actual 18,000 fluctuates |
| **Feedstock character:** | Secondary sludge produced on site/ Imported sludge. |
| Volume/ day: | 23 $m^3$ approx |
| HRT: | 9.1 days approx. |
| Feeding system | Batch fed once per day |

**Plant and Equipment**

| | |
|---|---|
| No. Of Digesters | Four |
| Volume: | 110 $m^3$ / tank |
| Heating system: | None, system is autothermal maintaining a temperature of 55 °C approx. |
| Aeration | Each tank has 2 spiral aerators and 1 circulatory aerator |
| Mixing | Facilitated by aerators |
| Odour control | Bord na Mona 'Monoshell' biofilter system |
| Final product storage: | 4 sealed tanks supplied with mixers |
| Final sludge disposal / reuse | Liquid sludge is spread on land |

**Plant Description**

The ATAD plant consists of four reactors in order to treat varying quantities of sludge, an important consideration considering Killarney town experiences population peaks during the summer months, due to a thriving tourism trade. The Hub centre facilitates the treatment of imported sludge from other treatment plants in Co. Kerry.

The design p.e. of the plant is 42,000 and flow through of the ATAD system ranges from 12$m^3$/d (with 1 reactor in operation) to 64.7 $m^3$/d ( with all 4 reactors in operation).

**ATAD Design Criteria**

The ATAD reactors were designed by FUCHS GmbH, Germany.

**Descriptive data**

| | |
|---|---|
| 1. Design capacity: | Up to 42,000 p.e. |
| Sludge flow rate: | 6.2 – 64.7 $m^3$/d. |
| Solids content: | 6% |
| Volatile solids content: | 70 % of TS (minimum) |
| **2. Reactor Description** | |
| Hydraulic retention time : | 7 – 10 days Currently 9.1 days |
| Number of reactors: | 4 |
| **Dimensions** | |
| Diameter: | 6.83 m |
| Sludge Depth: | 3.00 m |
| Working volume: | 110 m |
| Wall height: | 4.00 m |
| Freeboard: | 1.00 m (measured at reactor wall) |
| **Insulation** | |
| Walls: | 100mm PTFE insulation |
| Roof: | 100mm PTFE insulation |
| Floor: | 100mm PTFE insulation |
| **3. Mechanical equipment** | |
| 4 spiral aerators abbr: SA, | type WBL-V (Reactors 1 A and 1B) |
| 4 spiral aerators abbr: SA, | type WBL-IV (Reactors 2 A and 2B) |
| 4 circulatory aerators abbr:CA, | type UBL-IV |
| 16 foam controllers abbr: FC, | type SSc/0 |

**Aeration and Mixing**

Aeration is continuous. Reactors have been fitted with two spiral aerators, side mounted opposite each other, on the wall of the reactor. The spiral aerators are composed of an air –cooled a.c motor, a hollow shaft running on bearings in the motor, immersed at an angle into the sludge. The propeller generates an angular flow to the bottom of the tank and in conjunction air is sucked through the hollow shaft, producing fine bubbles and intensive turbulence. Since reactor 1 A and reactor 1 B are the primary reactors(in the two-stage train of operation), the spiral aerators are larger at 5.5 kW each, as these reactors receive the feed sludge which is highly concentrated, exerting a higher oxygen demand. In reactors 2 A and 2 B the spiral aerators are 4 kW each.

In addition, mixing and aeration is enhanced by circulation aeration. A circulation aerator is centrally mounted to the lid of each reactor. Each of the circulation aerators are of equal size 4 kW, and consists of an air cooled vertically mounted stirrer motor, an air pipe co-axial to the drive shaft, an impeller with radial vanes and an immersed end, as well as directional flow device. The impeller sucks the sludge to be aerated from underneath and sucks the air required for aeration through a pipe from above, which combines with the sludge, distributing it radially on all sides along the flow directional device, outward creating fine bubble diffusion.

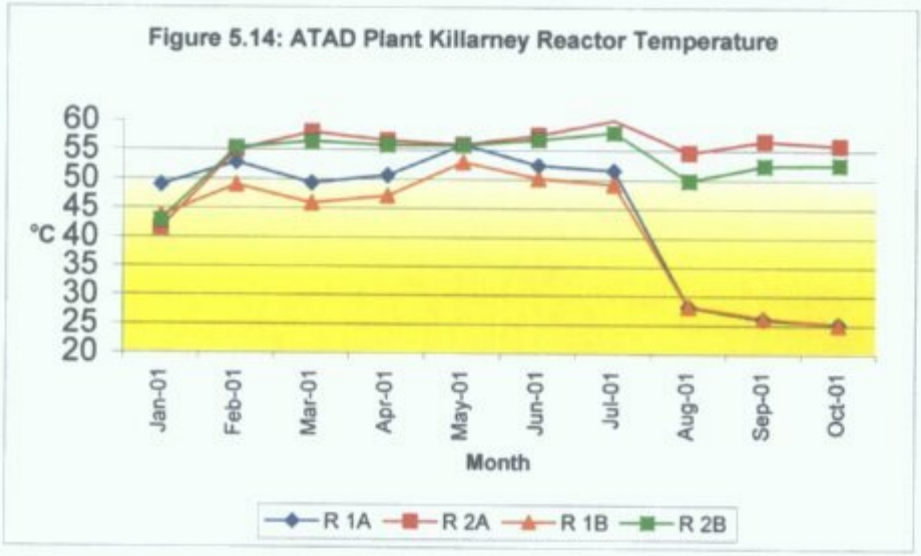
**Foam Control**

Foam control is of critical importance in the operation of an ATAD plant, as excessive foam can lead to loss of solids from a reactor, if uncontrolled. Each reactor has been fitted with four mechanical foam cutters. Supplied by Fuchs, the foam control system consists of an in-house horizontally mounted a.c. motor (0.75 kW) and a stainless steel blade (four sickled) which is connected to the drive shaft of the motor. The level of the foam layer is measured by a pressure sensitive depth gauge, once the foam layer reaches 0.4 m approx. the foam cutters automatically starts and the blades rotate rapidly, cutting through the foam, physically breaking it up.

**Temperature of Operation**

The reactors are operated in the thermophilic temperature range. The digesters have not been fitted with internal or external heat exchangers. Heating of the system relies totally on the generation of heat from the metabolism of organic solids, and is therefore

'Autothermal'. The design temperature of operation is 55 °C. However, depending on the operating sequence, the temperature in the primary reactors tends to be lower (mesophilic to thermotolerant temperature range) than that of the subsequent reactors, due to the introduction of large quantities of cold feed (as can be seen from figure 5.12). Each reactor has been fitted with two temperature probes. Temperature is monitored and controlled by a SCADA system. Since the process becomes self-limiting at excessive temperatures (70 °C), a cooling water system comes into operation when the temperature goes above an upper set limit of 60 °C. When the reactors have cooled down to 55 °C the cooling water is turned off. There is no system in place however, to heat the digesters should they fall below 55 °C, in this instance the operating temperature could only be controlled by alteration of Volatile Solids concentration of the feedstock.




**Piping**

The digester piping arrangement provides highly flexible operation. It is possible to operate one, two, three, or four reactors. A feed cycle can run automatically or manually, controlled by a SCADA system. The feeding sequence is carried out on a discontinuous (batch) mode once / day. There are seven modes in which the system can be run on depending on the volume of sludge available for digestion, which are as follows:

| Mode | Mode of operation | Reactors | Max. Flow |
|---|---|---|---|
| 1 | Single stage | 2 A | 12 m3/d |
| 2 | Single stage | 2 B | 12 m 3/d |
| 3 | Two stage | 1A & 2A | 24 m3/d |
| 4 | Two stage | 1 B & 2 B | 24 m3/d |
| 5 | Three stage | 1 B, 2 B & 2 A | 36 m3/d |
| 6 | Three stage | 1 A, 2 A & 2 B | 36 m3/d |
| 7 | Running in parallel | 1 A & 2 A, 1 B & 2 B | 64.7 m3/d (HRT 6.2 days) |

**Feed Preparation**

Plate: 5.5 TAD Feed Tank

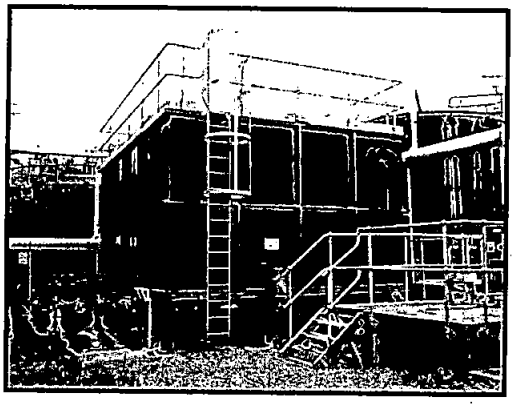
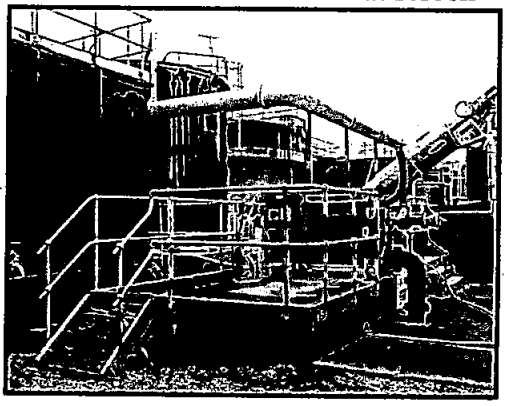


Plate: 5.6 Huber Rotamat screen

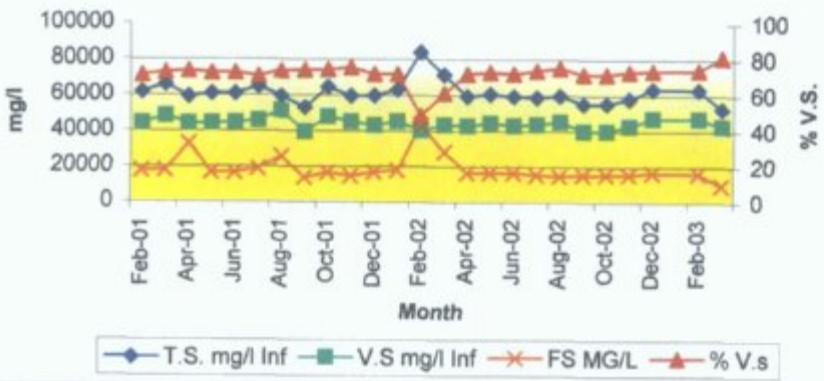


**Locally (on-site) Produced Sludge**

Feed solids content is critical in the operation of a TAD system. Typically the feed T.S. is maintained at 6%. To date, the ATAD facility has not received imported sludge on a regular basis, and operates on sludge generated on-site from the wastewater treatment plant, consisting of two oxidation ditches built 1974 p.e. 9,000 / tank; 3 aeration basins built 1984 p.e. 12,000 in total, and 8 diffused aeration tanks, built 1997, p.e. 12,000, in total. The waste sludge is pumped to two picket fence thickeners (160 $m^3$ / PFT) and preliminarily thickened to 3 % T.S. After a short retention time of approx. 3 hours the sludge is pumped to a sludge sump and from here directly to the belt press, with automatic polyelectrolyte dosing system, a wash water system and air extraction system. The sludge is thickened to 6%T.S. with the aid of Zetag 67 polyelectrolyte, and is then directed to the TAD feed storage tanks, which precede the ATAD reactors. There are

two feed tanks each with a working volume of 33 $m^3$, each tank has a mixer, a level sensor and are also equipped with an air extraction system. A problem came to light with the feed tanks, as the mixers are unsuccessful in adequately mixing the feed. This resulted in stratification of solids, which lead to irregularities in the feed consistency, i.e. on alternating days feed may be highly concentrated or very low in solids. It is thought that nitrification during the wastewater treatment process was also a causative agent in this case (as the nitrogen gas may have caused floatation of solids). To mitigate further occurrence of this problem the retention time in the picket fence thickener is now controlled at 3 hrs, and the level of the submerged mixer is manually alternated between high and low levels in the TAD feed tank on a daily basis, to aid in mixing.

Figure 5.15 ATAD Killarney Feedstock Composition




**Imported Sludge Preparation**

Imported sludge usually arrives at the hub centre in a pre-thickened state, and is therefore above 6% T.S. In this instance, the imported sludge must be diluted. The sludge is off-loaded in the imported sludge reception area, and is pumped to a dilution tank, which is a concrete structure with a mixer, level sensor, sludge diversion pipe and air extraction system. Dilution water is added proportional to the volume required to dilute the sludge feed to 6% T.S., the mixer starts automatically or manually and is controlled by the level sensor within the tank. The unscreened sludge is discharged to the screening area through the sludge diversion pipe. The system consists of a Huber Rotamat screen with wash water and level sensors. The screen removes particles 5 mm in diameter from the sludge. The sludge then flows by gravity to the imported sludge sump, which is located below the ground. The concrete tank has 2 submersible pumps (1 duty and 1 stand-by) and a level sensor. The sludge is then pumped to the imported sludge storage tank, which has a working volume of 33 $m^3$. It is a closed tank, with

access ladder, a level sensor and 2 feed forward pumps (1 duty and 1 stand-by), a mixer and air extraction system. The tank can receive sludge not only from the imported sludge sump but also from directly from the belt press.

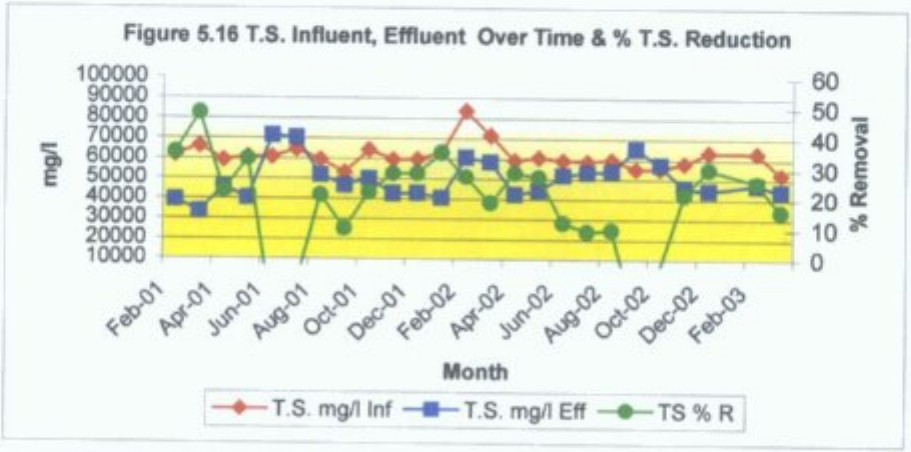
**Feed and Effluent Discharge**

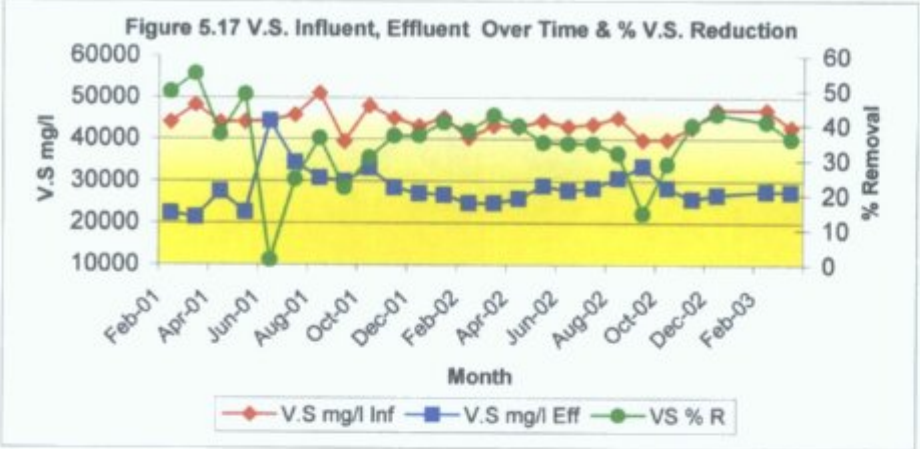
Feeding of the reactor occurs once per day in either automatic or manual mode. Typically, the system is operated automatically and relies on pressurised level sensors within the reactors as the main relay for the process. Taking the simplest mode of operation as an example R 2 'A' single stage process (Mode 1). At a predetermined time each day, all 3 aerators in the reactor stop, the outlet valve will open, one of the product feed forward pumps (1 or 2) will pump the desired quantity of digested sludge to the selected sludge storage tank, (there are four product storage tanks, 370 $m^3$ each, made of carbon steel, each of which is provided with a mixer, level sensor and air extraction system). The outlet valve will close, and one of the four ATAD batch feed pumps will come on and fill the reactor to 100% level. The aerators re-commence operation. During feed withdrawal and administration there is a down time of 1 hour.

**Final Sludge Quality**

The average tonnage of dry solids produced year in Killarney averages at 546.62 tds. Figure: 5.14 and 5.15 illustrates typical influent and effluent characteristics and the percentage removal.

Figure 5.16 T.S. Influent, Effluent Over Time & % T.S. Reduction






## Odor Control

Plate: 5.7 Ammonia Scrubber

Plate: 5.8 Monoshell Unit

There are 2 'Monoshell' biofilters on site, supplied by Bord na Mona. The first extracts and treats 3300 $m^3$ of air / hour at 1200 pa, from the ATAD reactors and product sludge storage tanks. The second extracts and treats 3800 $m^3$ of air / hour from the belt thickener, sludge pressroom, dilution tank, imported sludge storage tank and TAD feed tanks.

Each biofilter contains 33 $m^3$ of inoculated media, *Mytilus Edulis* or Mussel shells, packed on a support of recycled plastic pallets, each unit has two centrifugal fans (1 duty and 1 stand-by) which extracts air through the biofilter and out through a stack to the atmosphere. The filter media effectively neutralises sulphuric acid which is a by-product produced during the biological oxidation of hydrogen sulphide.

**Cost of Treatment**

| | |
|---|---|
| Capital cost (estimate) | €3,555,300 |
| **Operational Costs:** | |
| Thickening of Feed Sludge | €12,808 / annum |
| ESB usage 1162.12 kW/d | €46,877 / annum |
| Transport to agricultural land | €45,000 / annum |
| **Total Cost of Treatment** | €103,957 / annum |
| Cost of Treatment per Tonne Dry Solids | €190.18 |

Table 5.14 ATAD Killarney Results of Analysis

| Feedstock composition and operational parameters | | | | | | | | Digester Sludge | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Date | Feed rate $m^3/d$ | HRT (days) | T.S.% | T.S% Range | % V.S. | %F.S. | pH | T.S.% | T.S. Range | V.S.% | V.S. mg/l | pH |
| Feb-01 | 22.78 | 9.10 | 6.18 | 5.55 - 6.83 | 71.35 | 28.65 | 5.76 | *3.925* | *3.68 - 4.17* | *56.69* | *22250* | *8.92* |
| Mar-01 | 22.80 | 9.10 | 6.61 | 5.62 - 7.98 | 72.94 | 27.06 | 5.75 | 3.34 | 1.52 - 4.57 | *63.77* | 21300 | 8.88 |
| Apr-01 | 20.86 | 9.10 | 5.93 | 5.47 - 6.26 | 73.66 | 26.34 | 5.85 | 4.58 | 2.9 - 6.04 | *60.04* | 27500 | *8.06* |
| May-01 | 23.80 | 9.10 | 6.09 | 5.63 - 6.44 | 72.60 | 27.40 | 6.07 | 4.03 | 3.53 - 4.39 | *55.75* | 22467 | 8.22 |
| Jun-01 | 25.66 | 9.10 | 6.10 | 5.62 - 6.47 | 72.83 | 27.17 | 6.28 | 7.15 | 6.09 - 8.2 | *62.07* | 44350 | 7.84 |
| Jul-01 | 26.41 | 9.10 | 6.47 | 5.86 - 7.5 | 70.80 | 29.20 | 6.48 | 7.05 | 4.44 - 8.4 | *48.88* | 34440 | 8.29 |
| Aug-01 | 24.32 | 9.10 | 5.96 | 1.68 - 7.37 | 73.61 | 26.39 | 6.78 | 5.18 | 4.06 - 5.8 | *59.41* | 30760 | 7.73 |
| Sep-01 | 24.70 | 9.10 | 5.32 | 4.28 - 6.38 | 73.99 | 26.01 | 6.67 | 4.68 | 4.5 - 4.9 | *64.01* | 29933 | 8.40 |
| Oct-01 | 23.03 | 9.10 | 6.46 | 4.28 - 6.38 | 74.38 | 25.62 | 6.13 | 5.01 | 4.5 - 5.91 | *66.13* | 33156 | 8.17 |
| Nov-01 | 24.38 | 9.10 | 5.98 | 5.69 - 6.25 | 75.45 | 24.55 | 6.10 | 4.28 | 3.96 - 5.04 | *66.42* | 28400 | 8.84 |
| Dec-01 | 25.58 | 9.10 | 5.99 | 5.85 - 6.23 | 71.93 | 28.07 | 6.25 | 4.28 | 4.25 - 4.3 | *63.38* | 27125 | 8.83 |
| Jan-02 | 25.32 | 9.10 | 6.32 | 5.97 - 6.12 | 71.93 | 28.07 | 5.98 | 4.08 | 3.94 - 4.16 | *65.44* | 26700 | 8.80 |
| Feb-02 | *24.82 | 9.10 | 8.39 | 6.98 - 10.28 | 48.57 | 51.43 | 6.05 | 6.09 | 4.12 - 8.16 | *40.59* | 24725 | 9.03 |
| Mar-02 | *25.58 | 9.10 | 7.16 | 5.96 - 8.2 | 61.36 | 38.64 | 5.89 | 5.87 | 4.39 - 7.69 | *42.11* | 24700 | 9.06 |
| Apr-02 | 25.47 | 9.10 | 5.95 | 5.82 - 6.08 | 72.10 | 27.90 | 5.72 | 4.24 | 4.11 - 4.33 | *60.86* | 25780 | 8.94 |
| May-02 | 24.58 | 9.10 | 6.09 | 5.79 - 6.22 | 72.93 | 27.07 | 5.89 | 4.42 | 3.83 - 4.52 | *65.21* | 28800 | 8.82 |
| Jun-02 | 25.10 | 9.10 | 5.96 | 4.46 - 6.38 | 72.26 | 27.74 | 6.20 | 5.20 | 3.87 - 7.65 | *53.24* | 27683 | 8.98 |
| Jul-02 | 29.16 | 9.10 | 5.90 | 5.38 - 609 | 73.81 | 26.19 | 6.31 | 5.34 | 3.98 - 6.44 | *53.05* | 28350 | 9.03 |
| Aug-02 | 31.61 | 9.10 | 6.00 | 5.9 - 6.12 | 75.36 | 24.64 | 6.25 | 5.41 | 4.53 - 6.72 | *56.72* | 30700 | 8.83 |
| Sep-02 | 27.45 | 9.10 | 5.54 | 5.23 - 5.84 | 71.91 | 28.09 | 6.38 | 6.54 | 3.94 - 8.8 | *51.32* | 33556 | 8.70 |
| Oct-02 | 26.84 | 9.10 | 5.56 | 5.15 - 5.86 | 71.87 | 28.13 | 6.30 | 5.75 | 3.9 - 7.31 | *49.33* | 28375 | 8.59 |
| Nov-02 | 24.83 | 9.10 | 5.83 | 5.75 - 5.97 | 73.34 | 26.66 | 6.10 | 4.57 | 4.07 - 5.88 | *56.26* | 25700 | 8.93 |
| Dec-02 | 23.27 | 9.10 | 6.38 | 6.11 - 6.77 | 73.89 | 26.11 | 5.93 | 4.45 | 4.39 - 4.53 | *60.12* | 26725 | 9.03 |
| Feb-03 | 22.07 | 9.10 | 6.34 | 5.78 - 6.96 | 74.15 | 25.85 | 5.75 | 4.72 | 4.02 - 5.15 | *58.37* | 27550 | 9.00 |
| Mar-03 | 23.77 | 9.10 | 5.26 | | 81.37 | 18.63 | 5.70 | 4.41 | | *62.13* | *27400* | 9.00 |
| **Mean** | **24.96** | | **6.00** | | **73.51** | **26.51** | **6.10** | **4.89** | | **59.06** | ***28652*** | **8.67** |

* Silt present in sludge from road construction works

## 5.5 A Comparison of Mesophilic Anaerobic Digestion in Tralee and Autothermal Thermophilic Aerobic Digestion in Killarney.

### Solids Destruction

Solids destruction in sludge digestion is important for a number of reasons. Firstly, sludge is composed of highly putrescible matter, which, if untreated would putrefy and decompose, posing a threat to the environment and human health. Secondly, the extent of solids destruction is of economic importance in terms of the quantity of sludge for disposal (cost of transportation etc).

The total solids reduction achievable by anaerobic means, in theory ranges from 25 – 30% (Quasim, 1999). In practice, the anaerobic digester in Tralee exceeds this level, with 31% T.S. removal. In terms of V.S. removal, the anaerobic digester achieves 44.8%, which is at the lower end of the predicted V.S. removal range of 40 – 60% (Quasim, 1999).

The ATAD in Killarney removes an average of 21.2% T.S., which is significantly less than that achieved in Tralee. This is primarily due to the rate of biomass synthesis within the aerobic reactor, which can be from 5 – 20% greater than that of anaerobic processes (Speece, 1996). In relation to V.S. removal, the ATAD in Killarney achieves 36.6% removal on average, this is typical of examples cited in literature of ATAD systems treating waste activated sludge, which ranges from 25 – 40% removal, depending on the biodegradability of the feedstock (Stentiford, 2001).

### Process Stability

Control of digester pH can be problematic in anaerobic digesters, particularly during time of process instability. The synergistic microbial consortia present within an anaerobic digester have an optimum pH in the range of 6.8 – 7.2 (W.P.C.F, 1987). Above or below this pH and digestion is severely curtailed. pH within the digester is dependant on VFA production and can not be considered in isolation, as during periods of instability, caused by organic overload for instance, VFA's may accumulate within the digester. This is because the acidogenic and acetogenic organisms have a faster rate

of biomass generation (μ max approx. 1 hr) in comparison to the methanogenic population (μ max approx. 0.04 hr) (Bitton, 1994), and these organisms will thrive on the excess feed source producing volatile acids quicker than the methanogens can utilise them.

In Tralee, while the pH of the feed sludge is slightly acidic in nature (typically 5.9), the digester manages to maintain a neutral pH of 6.95, on average, due to metabolism - generated alkalinity from the digestion of proteins, nitrogenous compounds and the salts of weak acids (acetic, propionic, and $H_2S$) (Speece, 1996; Sayer, 1999). As pH is affected by the production of VFA, when the VFA concentration increases significantly in Tralee, sodium bicarbonate is added to the reactor. This is a precautionary measure, as the alkalinity of anaerobic digesters typically falls with in a range of 2000 – 6000 mg/l as $CaCO_3$, which is sufficient to buffer the effects of a decrease in pH (Dohanyos and Zabranska, 2001). The alkalinity of the anaerobic digester in Tralee falls within this range at 4889 mg/l $CaCO_3$ on average.

Overall, the anaerobic digester in Tralee appears to be relatively stable, regardless of variations in organic loading. While the feed solids concentration fed to the digester averages at 4.68% T.S. (s.d. of 0.71%), the quantity (load) fed to the reactors varies greatly from (day to day &) month to month, as can be seen from table, (5.13 A and C), which in turn effects the extent of VFA production within the digester. Typically, the VFA concentration averages at 278 mg/l, as acetic acid but varies greatly. The overall stability of the digester is not severely restricted by VFA production, as the quality of biogas remains consistent (61.48% $CH_4$ and 39% $CO_2$ on average). It has been noted however, that during periods of VFA accumulation, that feeding of the digester ceases, (to allow utilisation of the VFA present), and once the digester has stabilised feeding re-commences at even greater volume to treat the back log of sludge awaiting treatment. This action remedies the problem in the short – term but results in a never-ending loop of VFA accumulation. Therefore maintaining consistency of the feedstock is of vital importance, this is difficult to achieve with a picket fence thickener however. It would be beneficial to the plant to install a sealed feed storage tank with mixer to enhance the consistence of the feedstock, as unlike the ATAD plant in Killarney, Belt Pressing of the sludge is not recommended due to the possible accumulation of polyelectrolyte within the anaerobic digester, as demonstrated during pilot plant trials. ATAD has an

advantage over anaerobic digestion in this regard, as while polyelectrolytes are less biodegradable under aerobic conditions, accumulation within a digester is not a factor due to the short HRT ensuring rapid flow through digester.

In terms of ATAD operation, there are few indicators of digester stability beyond T.S. /V.S. removal. The most significant indicator of imbalance from overload for example is temperature, as temperatures in excess of 65 °C may arise if the overload is organic in nature, whilst a sudden decrease in temperature may occur if the overload experienced is hydraulic. Typically, the effects of such imbalance are sudden and little can be done to mitigate the resultant effects, beyond close monitoring of feed rate and concentration to prevent overloading, (cooling may be used if overload is due to organic concentration). pH directly correlates to temperature, due to the suppression of nitrification imposed by the high temperature maintained in an ATAD system, thus the pH tends to be alkaline. The pH of the digested sludge in Killarney averages at 8.7.

Other possible indicators of imbalance, not monitored in Killarney include oxygen take-up rate (OTR), and VFA. High VFA concentrations have been encountered in ATAD's operating at low D.O. levels and are thought to develop due to conflict between the aerobic and anaerobic organisms present within the digester, but no significant work has been conducted to distinguishing between the VFAs present or to determine which individual VFA's are indicative of instability, as has been done in the case of anaerobic digestion. In The ATAD system, it is most likely that the presence of VFA's would not be highlighted to any great extent by pH reduction, as observed in the anaerobic process, as the extent of aeration scrubs $CO_2$ from the head of the digester, thus preventing pH depression (Speece, 1996). From pilot scale trials CODs was noted as a possible indicator of Autothermal conditions, as following transition from 16 °C to 28 °C approx the CODs increased significantly, as did pH, which is a characteristic feature of Thermophilic conditions. CODs could be utilised on a regular basis to indicate the extent of hydrolysis in an ATAD plant, thus reduction in CODs would indicate at an early stage poor biodegradability of the feed solids (reduction in cell lysis), and even irregularities in feed consistency (If V.S content of the feed reduces then so too would the CODs).

The feed rate (24.96 $m^3$/d) and feed concentration (6%T.S.) administered to the ATAD in Killarney is typically constant. Variations in peak flows in ATAD's are mitigated by the fact that additional digesters can be put in to operation during such times. This is the case in Killarney, therefore the feed rate to the reactor is often unaffected as the additional flow is spread equally between the digesters. The ATAD in Killarney is stable in regard to operation, but there does appear to be considerable variations in the consistency of the Total Solids concentration exiting the reactors (see T.S. ranges). This suggests inadequate mixing within the digester, one reason for this may be the smaller spiral aerators used in the secondary tanks (chosen because of lower oxygen demands at this stage).

**Temperature of Operation**

A comparison operating temperature is not appropriate, as one system is mesophilic, and the other thermophilic.

The anaerobic digester in Tralee is heated by means of an external heat exchanger, which maintains the reactor temperature at 33.5 °C on average. At present in Ireland, mesophilic anaerobic digestion is used widely. National sludge treatment policy *(The Code of Good Practice in the Use of Biosolids in Agriculture)* states that a 'Class A' Biosolids (Pathogen reduction of <1,000 MPN.g D.S. for fecal coliforms, and <3 MPN. 4g D.S. for salmonella sp.) should be achieved for sustainable management of Biosolids, (i.e. land spread). Mesophilic Anaerobic Digestion cannot produce a Class A Biosolid, without a pre / post pasteurisation stage, therefore the Tralee plant produces a 'Class B' Biosolid (U.S. EPA Rule 503).

The high temperature of operation in an ATAD, ensures the technology complies with both the requirements of the Code of Good Practice and the U.S. EPA Rule 503 ('Class A' Biosolid) in terms of pathogen reduction. The ATAD in Killarney maintains an average temperature of 55 °C in the secondary reactors. It is apparent that there was a change in feeding regime from the initial operating conditions. From start – up to August of 2001 all rectors were operating in the thermophilic zone but, from this period on, the temperatures in the primary reactors fell into the mesophilic zone (operating at 25 – 30 °C). This suggests that the during start – up operation the reactor were fed on a

continuous basis, and that the feeding regime was changed to a batch fed process of once per day during august 2001.

**HRT**

The HRT of the anaerobic digester in Tralee averages at 26 days approximately.
In Killarney the HRT is maintained at 9.1 days on average, it is for this reason that ATAD's are typically cheaper to construct than anaerobic digesters, as the digesters can effectively be a third to half the volume of an anaerobic digester treating the same quantity of sludge.

**Economic Assessment**

There is a preception that ATAD is more expensive to operate and maintain than anaerobic digestion, due to the extent of aeration required. The cost incurred in the treatment of sludge at Killarney is €190.18/ t.d.s. and at Tralee is €211.16 / t.d.s. In this case, the ATAD in Killarney is more economical as the sludge treatment plant is treating on almost twice (546.62 t.d.s. / year) that of the anaerobic digester in Tralee (329.34 t.d.s. / year).

Both plants are designed for a p.e. of 42,000. The low sludge production yield in Tralee can be explained by the fact that the wastewater treatment plant in Tralee includes primary treatment and secondary treatment (trickling filters and diffused aeration). The majority of organic matter in this instance is removed from the waste stream during primary treatment and the secondary treatment by trickling filter. During primary settlement readily biodegradable faecal matter is settled out of suspension, removing up to 60% of the solids (20% of the organic load). The trickling filter system removes a further 45% of the organic load with minimal sludge production (sloughing off just once / twice per year). As a result, a substantial portion of the organic load has been treated prior to the diffused aeration stage; therefore the sludge production from this stage is also minimal. The sludge treatment plant in Tralee appears to be oversized for the quantity of sludge it is producing on a daily basis. The digester has been in operation since 1998 and, to date, only one of the two digesters has been commissioned, despite

the fact that the plant is treating up to 75% of its design p.e. (design p.e. 42,000, current p.e. 36,000).

The final sludge from the ATAD plant in Killarney is transported to land in liquid form. The cost of dewatering has not been considered in this cost estimation, however, polyelectrolyte dosage for the dewatering of ATAD sludge can be from 3 – 10 times greater than that required for anaerobic sludge (Kelly *et al.*, 2000).

**Summary**

The ATAD plant in Killarney has proven to attain a higher quality of sludge (Class A), than the mesophilic anaerobic digester in Tralee, at a shorter HRT of 9.1 days. Economically the ATAD plant supersedes the anaerobic digester both in terms of capital cost and, cost of operation. Furthermore, due to the complex nature of the anaerobic digestion process, in terms of the extent of operating skill required, the ATAD plant is once again at an advantage and is flexible in operation, should the flow to the plant increase addition reactors can be brought on-line, and be fully operational in a relatively short period time (two weeks approx.).

In light of this study it is surprising that the ATAD plant in Killarney is the only sewage sludge digester of its type in Ireland, but this may be due to the fact that ATAD is still considered as a new technology. In the authors opinion, ATAD is an excellent means of achieving a pasteurised Biosolid product suitable for re-use in agriculture, in line with the specifications of the '*Code of Good Practice in the use of Biosolids in Agriculture*', and, while anaerobic digestion achieves greater solids reduction in terms of overall mass of sludge for disposal, unless the anaerobic digestion process is thermophilic (or with a pre or post pasteurisation stage) it can not attain 'Class A' status, and final disposal of such sludge will eventually become problematic.

# CHAPTER 6
# CONCLUSIONS AND RECOMMENDATIONS

## 6.1 Conclusions and Recommendations

Following the operation of the pilot scale aerobic and anaerobic reactors and the appraisal of full-scale sludge treatment plants throughout Ireland, the main findings of this project are as follows:

- In the treatment of Secondary Sludge, the pilot scale anaerobic digester achieved greater T.S., V.S. and CODt reduction on average of 28%, 31.5% and 32% than that of the aerobic thermophilic pilot plant which achieved 20% T.S. removal, 31% V.S. removal and CODt removal of 34.6%, despite the fact that secondary sludge is the optimum feedstock for aerobic digesters. This is due to the greater biomass growth rate of aerobic systems.

- Following reductions in HRT the removal efficiency of both pilot plants reduced, indicating hydrolysis to be problematic not only for the anaerobic digestion process but also for aerobic digestion, and is also indicative of poor biodegradability of the feedstock.

- Volatile Fatty Acids identified as indicators of process performance in primary sludge digesters, such as propionate, butyrate, and acetate, are not deemed as good indicators of process stability in the treatment of secondary sludge. As following VFA analysis by GC, VFA's were consistently low even during operational difficulties. This may be attributed in part, to poor biodegradability of the feedstock (i.e. slow rate of hydrolysis, would result low concentrations of VFA, and there subsequent utilisation by the biomass). However, as constituent VFA's are typically substrate specific, future analysis on a broader spectrum of VFA is recommended to determine which are predominantly produced under stressful conditions with a test substrate of secondary solids.

- From the operational difficulties encountered in the operation of the anaerobic pilot scale reactor it is apparent that mixing is vital in the anaerobic digestion of secondary sludge. Polyelectrolyte should not be used as the sole means of thickening a feedstock for anaerobic digestion due to possible accumulation

within the digester. Oxygen, which has been identified as a inhibitor of anaerobic digestion process, did not have a long-term detrimental effect on digestion as the methane content of the biogas increased to 55 % within five days, following an air shock.

- Strict control of feedstock consistency and flow, is of critical importance in maintaining harmony between the synergistic groups of bacteria within an anaerobic digester. Maintaining feedstock consistence to the anaerobic digester in Tralee is problematic, to over-come this, the installation of a sealed feed storage tank with mixer would be beneficial to process stability.

- Due to the shorter HRT of the ATAD plant in Killarney double the quantity of sludge can be treated, averaging at 546 t.d.s / year in Killarney and 329 t.d.s. / year in Tralee.

- At present Biosolids producers are encouraged to achieve a 'Class A' Sludge, this is not a mandatory requirement however. Future legislative requirements are likely to become more stringent. Due to the high temperature of operation attained by ATAD systems a 'Class A' sludge is achieved in terms of pathogen reduction, whereas mesophilic anaerobic digestion plants nationwide would require expansion to include per or post sanitation to achieve a 'Class A' Biosolid, which will add to the overall cost of treatment.

- The ATAD plant in Killarney is more cost effective to operate per t.d.s, than the mesophilic anaerobic digester in Tralee (€190/t.d.s ATAD verses €211/ t.d.s. Anaerobic digestion), partially due to the fact that the ATAD plant is treating double the quantity of sludge per year on average.

# REFERENCES

Aguilar, A., Casas, C., and Lema, J.L. (1995) *Degradation of VFA by Differentially Enriched Methanogenic Cultures Kinetics and Inhibition.* Water Research. Elsevier Science. **29** (2), pp. 505 – 509.

APHA, AWWA, WEF. (1995) *Standard Methods for the Examination of Water and Wastewater 19th Ed.*, American Public Health Association, Washington, D.C.

Azbar, N., Ursillo, P., Speece, R. (2001) *Effect of Process Configuration and Substrate Complexity on the Performance of Anaerobic Processes.* Water Res. **35**(3), 817 – 829.

B. Inanc, B., S. Matsui, S., and Ids, S. (1996 ) *Propionic Acid Accumulation and Controlling Factors in Anaerobic Treatment of Carbohydrate, Effects of H2 and pH.* Water Science and Technology. Elsevier Science. **34** (5-6), pp 317 – 325.

Baier, U. and Schmidheiny, P. (1997) *Enhanced anaerobic degradation of mechanically disintegrated sludge.* Water Science and Technology. IAWQ. **36** (11), pp. 137 – 143.

Barker, D.J., Mannucchi, G.A., Salui, S., and Stuckey, D.C. (1999) *Characterisation of Soluble Residual COD in Anaerobic Wastewater Effluents..* Water Research. Elsevier Science **33** (11), pp 2499 – 2510.

Bartlett and Killilea, (2002) *Characterisation, Treatment and Sustainable Reuse of Biosolids in Ireland.* In Conference proceedings of Sludge Management Entering the 3rd Millenium, Industrial Combined Water, and Wastewater Residues. Taipei, 25-28 March 2002. IWA.

Bitton, G. (1994) *Wastewater Microbiology.* Wiley, New York.

Bjornsson, L., Murto, M., Jantsch, T.G., and Bjornsson, M., (2001) *Evaluation of New Methods for the Monitoring of Alkalinity, Dissolved Hydrogen and the Microbial Community in Anaerobic Digestion.* Water Research. Elsevier Science. **35** (12), pp 2833 – 2840.

Braber, K. (1995) *Anaerobic Digestion of Municipal Solid Waste: A Modern Waste Disposal Option on the Verge of Breakthrough.* Biomass and Bioenergy. Elsevier Science. **9** (1 –5), pp. 365 – 376.

Casey, T.J. (1984) *Requirements and Methods for Mixing in Anaerobic Digesters.*In: Anaerobic Digestion of Sewage Sludge and Organic Agricultural Wastes. e.d. Bruce, A.M., Katsirs, A.K., and Newman, J.P. Proceedings of Seminar. Elsevier Applied Science, London.

Chand, L.L., Raudenbush, D.L., and Dentel, S.K. (2001) *Aerobic and Anaerobic Biodegradation of a Flocculant Polymer..* Water Science and Technology. IWA. **44**

( 2-3), pp, 461 – 468.

Cheremisinoff, N.P. (2002) *Handbook of Water and Wastewater Treatment Technologies.* Butterworth- Heinemann, Boston.

Chu, C.P., Chang, B., Liao, G.S., Jean, D.S. and Lee, D.J. (2001) *Observations on changes in ultrasonically treated waste-activated sludge.* Water Res. Elsevier Science ltd. **35** (4), pp 1038 – 1046.

Cronin, F. (2002) *Osberstown Wastewater Treatment Plant.* http://www.osberstownw.ie

De Souza Araujo, L., Catunda , P. and Van Haandel A.C. (1998) *Biological Sludge stabilization. Part 2 ; Influence of the composition of WAS on anaerobic stabilization.* Water Science and Technology. Elsevier Science ltd. **24** (3), pp 231-236.

Dohanyos and Zabranska. (2001) *Sludge to Biosolids: Processing, Disposal, Utilization.* e.d: Spinosa, L., and Vesilind, P.A. IWA , London.

E.P.A., U.S. (1990) *Environmental Regulations and Technology, Autothermal Thermophilic Aerobic digestion of Municipal Wastewater Sludge.* Technomic Inc, Lancaster .

Feitkenhauer, H., Von Saches, J. and Meyer, U. (2001) *On – line Titration of VFA for the Process Control of Anaerobic Digestion Plants.* Water Research. Elsevier Science. **36**, pp.12 – 218.

Fifield and Hains, (2000) Environmental Analytical Chemistry 2nd Ed. Blackwell Science.

Fifield and Hains. (2000) *Environmental Analytical Chemistry* 2nd e.d. Blackwell Science .

Fukui, M., Jung, S., and Urushikuni, Y. (2001) *In situ Substrates for Sulfidogens and Methanogens in Municipal Anaerobic Sewage Digesters with Different Levels of Sulfate.* Water Res. **34** (5), 1515 – 1524.

G.V.R.D. (2000) *The Biosolids report, Recycling biosolids to soil.* **1.,** May. G.V.R.D. http://www.gvrd.bc.ca/nutrifor/publications/BR_pathogens.pdf.

Gantzar, C., Gaspard, P., Galvez, L., Huyard, A., Dumouthier, N., & Schwartzbrod, J. (2001) *Monitoring of Bacterial and Parasitological Ccontamination During Various Treatment of Sludge.* Wat Res. **35** (16), 3763 – 3770.

Gaudy, A., & Gaudy, E. (1980) *Microbiology for Environmental Science and Engineering.* Mc Graw Hill.

Gray, N.F. (1999) *Water Technology – An Introduction For Environmental Scientists*

*and Engineers.* Arnold & Wiley, London.

Gujer, (1983) *Conversion Process in Anaerobic Digestion.* Water Science Technology. **15,** pp.127 – 167.

Haandel, V and Letting, A. (1994) *Anaerobic Sewage Treatment: A Practical Guide for Regions with a Hot Climate.* Wiley, England.

Hammer, M. (1998) *Water and Wastewater Technology 3rd e.d.* Prentice.

Haner, A., Mason, C.A. & Hamer, G. (1994) *Death and lysis During Aerobic Thermophilic Sludge Treatment: Characterization of Recalcitrant Products.* Wat. Res., **28** (4), 863 –869.

Hobson, P.N. and Wheatley, A.D. (1993) *Anaerobic digestion, Modern Theory and Practice.* Elsevier Applied Science, London.

Jewell, W.K., & Kabrick, R.M. (1980) *Autoheated Aerobic Thermophilic Digestion with aeration.* Water Pollut. Control Fed., **52** (512).

Kelly, H.G. (1995) *What's in a Name? – Flexibility.* Water Environment and Technology Magazine, July 1995

Kelly, H.G.(1989) *Aerobic Thermophilic Digestion or liquid composting of Municipal Sludges.* Environmental Engineering, Proceedings of the 1989 Specialty Conference/ EE Div. / ASCE, Austin, Texas, July. Ed.. Malina, J.F.

Kelly, H.G., Frese, H., Gibb, A., and Zhou, J. (2000) *Dewatering Improvements ofThermophilically Digested Biosolids Through Addition of Metal Salts at the Salmon Arm.* Proceedings of CSCE, Water Pollution Control Centre, June 7-10, London, Ontario, 241-246.

Kelly, H.G., Jennings, C., Sivyer, D., Robinson, L., Frese, H. & Cale, C. (1995) *Hot Tips for Autothermal Aerobic Digestion- A step – by- step guide to startup and operation of an ATAD system.* Water Environment Federation, **12** (5).

Kelly, H.G., Melcer,.H., Mavinic, D.S. (1993) *Autothermal Thermophilic Aerobic Digestion of Municipal Sludges: A one – year, Full-scale Demonstration Project.* Water Environment Federation.

Kelly, H.G., Warren, R. (1997) *Autothermal Thermophilic Aerobic Digestion Design.* Proceedings Vol. 1, Environmental Engineering Conference, CSCE ASCE, July 22- 26, Edmonton, Alberta, Canada. Ed. Stanley, S.J., Ward, C.J.W., & Smith, D.W.

Khalili, N.R, Chaib, E., Parulekar, S.J., & Nykiel, D.(2000) *Performance Enhancement of Batch Aerobic Digester via Addition of Digested Sludge.* Journal of Hazardous

Materials (B76), 91 –102.

Kiely. (1997) *Environmental Enginering.* Mc Graw and Hill, Berkshire.

Kotsyurbenko, O., Glagolev, A., Nozhevnikova, N., and Conrad, R. (2001) *Competition Between Homoacetogenic Bacteria and Methanogenic Archaea for Hydrogen at Low Temperatures.* FEMs Microbiology Ecology **38**,153 – 159.

Krishna, R.P., Hyungjin, K., & Cheunbarn, T. (2000) *Aerobic Thermophilic and Anaerobic Mesophilic Treatment of Swine Waste.* Wat. Res. **34** (10), 2727-2753.

Lafitte - Trouque, S. and Forster, C.F. (2002) *The use of ultrasound and γ- irradiation as pre-treatments for the anaerobic digestion of waste activated sludge at mesophilic and thermophilic temperatures.* Bioresource Technology. Elsevier Science. **84,** pp 113 – 118.

Lange, M., and Adring, B. (2001) *A Comprehensive Study into the Molecular Methodology and Molecular Biology of Methanogenic Archaea.* FEMs Microbiology Reviews, **25**, 553 – 571.

Lapara, T.M. & Allemen, J.E. (1999) *Thermophilic Aerobic Biological Wastewater Treatment.* Wat. Res., **34** (4).

Li, J.Z., Donald S. Maninic, S.M., & Kelly, H.G. (2002) *Batch and Continuous Feeding of Thermophilic Aerobically Digested Sludge Supernatant as a Carbon Source for Biological Nutrient Removal.* Environ Eng. Sci. (1), 213 –224.

Lusk, P. (1999) *Latest Progress in anaerobic digestion- Part I.* In: Biocycle Magazine. U.S.

Maharaj, I., and Elefsiniotis, P. (2001) The Role of HRT and Low Temperature on the Acid –phase Anaerobic Digestion of Municipal and Industrial Wastewaters. **76,** pp 191 – 197.

Malina, J.F., and Pohland, F.G. (1992) *Design of Anaerobic Digesters for the Treatment of Industrial and Municipal Wastes.* Technomic Publications, Lancaster.

Marty, (1984) *Microbiology of anaerobic digestion.* In: Anaerobic Digestion of Sewage Sludge and Organic Agricultural Wastes, ed. Bruce, A.M., Kouzeli – Katsirs, A., and Newman, P.J. Elsevier Applied Science Publications, London .

Michaud, S., Bernet, N., Buffiere, P., Roustan, M. and Moletta, R. Michaud. (2001) *Methane Yield as a Monitoring Parameter for the Start-up of Anaerobic Fixed Film Reactors.* Water Research. Elsevier Science. **36** (March), pp.1385 – 1391.

Murthy, A., Novak, S.N., Holbrook, J.T., & Surovik, R.D. (2000a.) *Mesophilic Aeration of Autothermal Thermophilic Aerobically Digested Biosolids to Improve Plant Operations*. Water Environment Research, **72** (4), 476-481.

Murthy, S.N., Novak, J.T., & Holbrook, R.D. (2000b) *Optimising Dewatering of Biosolids From Autothermal Tthermophilic Aerobic Digesters (ATAD) Using Inorganic Conditioners*. Water Environment Research, **72** (6), 714-721.

Nah, W., Kang, Y.W, Hwang, K., and Song, W. (2000) *Mechanical pre-treatment of waste activated sludge for anaerobic digestion process*. Water research. Elsevier Science ltd. **34**(8) pp 2362 – 2368.

Oleszkiewicz , J.A., & Mavinic , D.S.(2002) *Wastewater Biosolids : An Overview of Processing, Treatment, and Management*. J. Environ Eng. Sci. (1), 75 –88.

Oleszkiewicz, J.A., and Mavinic, D.S. (2002) *Wastewater Biosolids: An Overview of Processing, Treatment, and Management*. Journal of Environmental Engineering Science. 1, 75 – 88. 2002. NCR Research Press

Onyeche, T.I., Schlafer, O., Bormann, H., Schroder, C., Sievers, M. (2002) *Ultrasonic cell disruption of stabilized sludge with subsequent anaerobic digestion.* Ultrasonics. Elsevier. **40**, (31- 35).

Oorschot, R., (2003) *Future Direction in Biosolids Management.* http://www.cambi.com/publications/sludge/default.htm

Panter, K. (2003) *Heat Treatment and Thermal Hydrolysis.* Simon – Hartley Cambi . http://www.cambi.com/publications/sludge/default.htm

Polprasert, (1996) *Organic Waste Recycle Technology and Management* 2nd ed. Wiley and Son, Chinchester .

Polprasert, C. (1996) *Organic Waste Recycling Technology and Management 2nd Ed.* John Wiley, England.

Pressley, R.L. & Eloff, J. (2001) *Second generation Autothermal Thermophilic Aerobic Digestion: Conceptual Issues and Process Advancements.*

Pride, C. (2002) *ATAD, Odours, and Biofilters.* Florida Water Resources Journal, April.

Quasim, (1999) *Wastewater Treatment Plants; Planning Design and Operation 2nd e.d.* Technomic Publications, Lancaster.

Sayer and Mc Carthy, (1999) *Chemistry For Environmental Engineers*. Parkin

Scahill,C. (2000) *The Appraisal of Anaerobic Digestion in Ireland to Develop Improved Designs and Operational Practice*. Research Masters of Engineering, School of

Engineering, Institute of Technology Sligo.

Scheminski, A. Krull, R. and Hempel, D.C.(2000) *Oxidative treatment of digested sewage sludge with ozone.* Water Science and Technology. IWA Publishing. **42** (9), pp 151 – 158.

Schwinning, H.G., Deeny, K., & Fuchs, L.(1993) *ATAD : An Effective PFRP Alternative.* WEFTEC, Water Environment Federation 66th Annual Conference and Exposition, Anaheim, California, USA October 3 –7.

Schwinning, H.G., & Fuchs Gas und Wassertechnik GmbH. (1996) *Autoheated Thermophilic aerobic digestion of industrial sewage sludge.* WEFTEC ,Water Environment Federation, 69th Annual Conference and Exposition, October 5 –9 1996, Dallas Convention Centre Texas USA.

Schwinning, H.G., Fuchs Gas-und Wassertechnik GmbH, Deeny, K.J., KC Environmental, Inc. Sun-Nan Hong, Kruger. (1997) *Experience with Autothermal Thermophilic Aerobic Digestion (ATAD) in the United States.* Water Environment Federation, 70th Annual Conference and Exposition, WEFTEC, Chicago, Illinois, October 18 – 22.

Skjelhaugen, O.J.(1999) *Thermophilic Aerobic Reactor for processing organic liquid wastes.* Wat. Res. **33** (7), 1593 – 1602.

Sorensen, J., Tholstrup, G., and Andreasen , K. (2003) *Anaerobic Digestion and Thermal Hydrolysis to Reduce Production of Sludge in WWTPs.* http://www.cambi.com/publications/sludge/default.htm

Speece, R.E. (1996) *Anaerobic Biotechnology for Industrial Wastewater.* Vanderbilt University. Archae Press, Nashville.

Stroot, P.G., Mc Mahon, K.D., Mackie, R.I., and Raskin, L. (2000) Ana*erobic Co-digestion of Municipal Solid Waste and Biosolids Under Various Mixing Conditions: I, Digester Performance.* Water Research. Elsevier Science. **35** (7), pp 1804 – 1816

Stroot, P.G., Mc Mahon, K.D., Mackie, R.I., and Raskin, L. (2001) *Anaerobic Co-digestion of Municipal Solid Waste and Biosolids Under Various Mixing Conditions: II, Microbial Population Dynamics.* Water Research, Elsevier Science. **35** (7), pp 1817 – 1827.

Tchobanoglous, G., and Barton, F. (1991) *Wastewater Engineering; Treatment, Disposal, and Reuse 3rd e.d.* Mc Graw Hill International, Civil Engineering Series. Metcalf and Eddy Inc.

Tecobanoglous, G., and Burton, F.L., and Stensel, D (2001) *Wastewater Engineering; Treatment, Disposal, and Reuse, 4th Ed.* Mc Graw Hill International, Metcalf and Eddy Inc.

Tiehm, A., Nickel, K., and Neis, U. (1997) *The use of ultrasound to accelerate the anaerobic digestion of sewage sludge.* Water Science and Technology. Elsevier Science ltd. **36**. (11), pp. 121-128.

Tiehm, A., Nickel, K., Zellhorn, M. and Neis, U. (2001). *Ultrasonic Waste Activated Sludge Disintegration For Improving Anaerobic Stabilization.* Water Res., **35** (8), 2003 – 2009.

Ueki, k., Ueki, A., Takahashi, K., and Iwatsu, M. (1992) *The Role of Sulfate Reduction in Methanogenic Digestion of Municipal Sewage Sludge.* J.Gen. Appl. Microbiol., **38**, 195 – 207.

Uyanik, S., Sallies, P.J., Anderson, G.K.(2001) *The Effect of Polymer Addition on Granulation in an Anaerobic Baffled Reactor (ABR) Part I.* Water Research. Elsevier Science.

Uyanik, S., Sallies, P.J., Anderson, G.K.(2002) *The Effect of Polymer Addition on Granulation in an Anaerobic Baffled Reactor (ABR) Part II: Compartmentalization of Bacteria Populations.* Water Research. Elsevier Science. **36** (4), pp 944 – 955.

Vismara R. (1985) *A Model for Autothermic Aerobic Digestion. Effects of Scale Depending on Aeration Efficiency and Sludge Concentration.* Wat. Res. **19** (4), 441 – 447.

Vogelaar, J.C.T., Klapwijk, A., Van Lier, J.B., & Rulkens, W.H. (2000) *Temperature effects on the oxygen transfer rate between 20 and 55 °C.* Wat. Res. **34** (3),1037 – 1041.

W.P.C.F. (1987) A*naerobic Sludge Digestion Manual of Practice no. 16, 2nd e.d.* Water Pollution Control Federation .

Wang, Q., Kuninobu, M., Kakimoto, K., Ogawa, H.I., Wang, Y.W. (1999) *Upgrading of Anaerobic Digestion of Waste Activated Sludge by Ultrasonic Pretreatment.* Bioresource Technology. Elsevier Science. **68**, pp. 309 – 313.

Water Environment Federation, (1995) *Wastewater Residuals Stabilisation - Manual of Practice.* Water Environment Federation.

Weemaes, M., Grootaerd, H., Simoens , F., and Verstraete, W. (2000) *Anaerobic digestion of ozonized Biosolids.* Water Research. Elsevier Science ltd. **34** (8), pp 2330 – 2336.

Weisz, N., Fjprdside, C., Kepp, U., and Solheim, O.E. (2001) *Thermal Hydrolysis is one Answer To The Current Trends in Sludge Management.* In: Sludge Management Entering the 3rd Millennium Industrial, Combined Water and Wastewater Residues; Conference proceedings, Grande Hotel, Taipei, Taiwan 25 – 28 March 2001. IWA International Water Association.

Zhou, J., Kelly, H.G., Mavinic, D.S., & Ramey, W.D. (2001) *Digestion Effects on Dewaterability of Thermophilic and Mesophilic Aerobically Digested Biosolids.* Proceedings of WEFTEC 2001, October 13-17, Atlanta, Georgia, USA.

# APPENDIX A
# POLYELECTROLYTE INHIBITION TRIALS

# APPENDIX A POLYELECTROLYTE INHIBITION TRIALS

| | | Kymene SLX-2 Cationic polyamide-epichlorohydrin | | | Reten 320 Cationic 1.2-dichloroethane resin | | | Zetag 51 Modified polyacrylamide | | |
|---|---|---|---|---|---|---|---|---|---|---|
| Jar test results | Dose (ml 500 ml) | Turbidity | App. Colour | True colour | Turbidity | App. Colour | True colour | Turbidity | App. Colour | True colour |
| | 1 | 6.4 | 90 | 10 | 7.4 | 100 | 15 | 14.7 | 240 | 75 |
| | 5 | 6.2 | 80 | 10 | 6.8 | 80 | 10 | 12.6 | 170 | 70 |
| | 10 | 9.8 | 120 | 30 | 6.5 | 60 | 15 | 13.5 | 150 | 60 |
| | 20 | 11 | 150 | 35 | 4.4 | 60 | 10 | 11.5 | 130 | 45 |
| | 50 | 11.1 | 160 | 35 | 5.1 | 50 | 5 | 13 | 150 | 40 |
| SMA | ml.g.VSS | SMA (kg CH4-CODkg VSS.d) | | | SMA (kg CH4-CODkg VSS.d) | | | SMA (kg CH4-CODkg VSS.d) | | |
| | 0.05 | 1.53 | | | N/A | | | N/A | | |
| | 0.1 | 1.51 | | | 1.18 | | | N/A | | |
| | 0.2 | N/A | | | 1.01 | | | N/A | | |

| | | Zetag 63 Cationic acrylamide copolymer | | | Zetag 78FS40 Cationic polyacylamide | | | Zetag 87 Cationic acylamide copolymer | | |
|---|---|---|---|---|---|---|---|---|---|---|
| Jar test results | Dose (ml 500 ml) | Turbidity | App. Colour | True colour | Turbidity | App. Colour | True colour | Turbidity | App. Colour | True colour |
| | 1 | 12.2 | 150 | 45 | 8.4 | 120 | 55 | 10.6 | 140 | 30 |
| | 5 | 8.9 | 120 | 30 | 7.9 | 100 | 50 | 8.3 | 110 | 25 |
| | 10 | 8.5 | 90 | 30 | 8.5 | 100 | 50 | 8.2 | 90 | 30 |
| | 20 | 8.6 | 90 | 15 | 8.9 | 90 | 35 | 8 | 100 | 25 |
| | 50 | 7 | 80 | 15 | 6.5 | 65 | 15 | 6 | 50 | 10 |
| SMA | ml.g.VSS | SMA (kg CH4-CODkg VSS.d) | | | SMA (kg CH4-CODkg VSS.d) | | | SMA (kg CH4-CODkg VSS.d) | | |
| | 0.1 | N/A | | | 1.14 | | | N/A | | |
| | 0.2 | N/A | | | 0 | | | 0.96 | | |
| | 0.5 | N/A | | | N/A | | | 0 | | |

(Uyanik et al, 2001)

SMA is specific metanogenic activity. N/A is non - applied. Unit of Turbidity is NTU. Unit of colour is couler unit determined by Nessleriser with standard colour discs. Kymene SLX-2 resin is available as ready to use 13.0% solids aqueous solution. Reten 320 resin was a ready to use 25% solids solution. Zetag product range was required to be prepared for laboratory tests by dissolving in water to provide 0.5% solid or liquid solutions of the product.

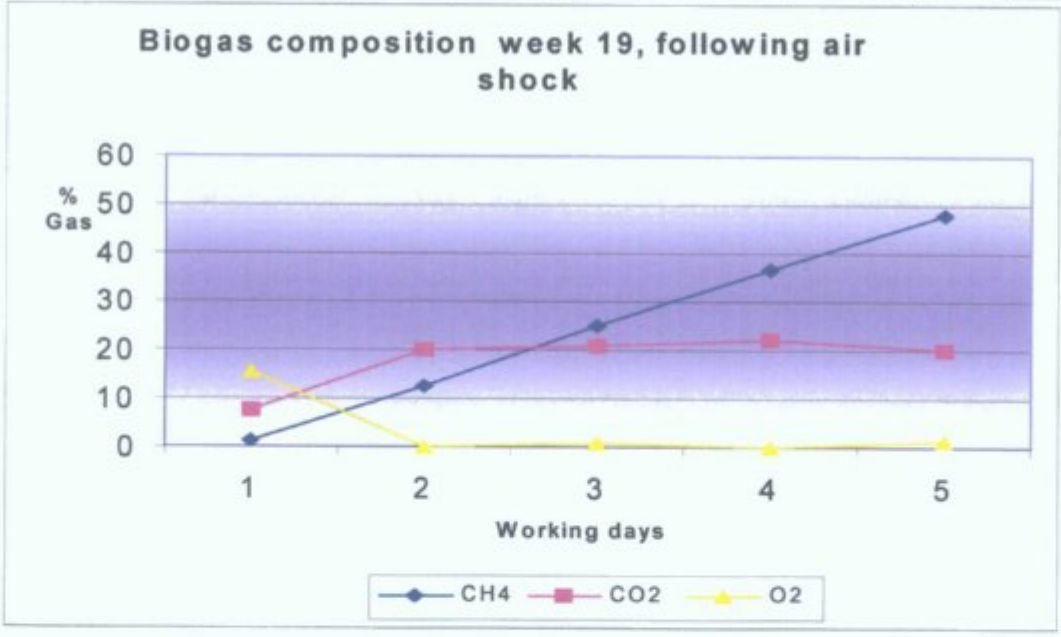
# APPENDIX B
## ANAEROBIC DIGESTER APPEARANCE FOLLOWING AIR SHOCK

## Appendix B Anaerobic Digester Appearance Following Air Shock

# APPENDIX C
## (ANAEROBIC DIGESTER)
## WEEK 19 -AIR SHOCK, EFFLUENT ANALYSIS AND FEEDING REGIME

## APPENDIX C (ANAEROBIC DIGESTER) WEEK 19 -AIR SHOCK, EFFLUENT ANALYSIS AND FEEDING REGIME

| Day | Feeding regime | pH | VFA mg/l | $CaCO_3$ mg/l | T.S mg/l | T.S % | V.S mg/l | VS % | F.S mg/l | FS % | COD T mg/l | COD S mg/l |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 1 | Fed - No EFF | | | | | | | | | | | |
| 2 | Fed - No EFF | | | | | | | | | | | |
| 3 | Fed - No EFF | | | | | | | | | | | |
| 4 | Fed - Sample | 7.06 | 255 | 4090 | 54972 | 5.49 | 32852 | 60 | 22120 | 40.3 | 39000 | 1502 |
| 5 | Fed - No EFF | | | | | | | | | | | |
| 6 | Fed - No EFF | | | | | | | | | | | |



# APPENDIX D
## (ANAEROBIC DIGESTER)
## RESULTS OF SAMPLING FOLLOWING PERIOD OF SHUT DOWN

## APPENDIX D (ANAEROBIC DIGESTER) RESULTS OF SAMPLING FOLLOWING PERIOD OF SHUT DOWN

| Sample | T.S | V.S | F.S | VFA mg/l | CODt mg/l | CODs mg/l | pH Units |
|---|---|---|---|---|---|---|---|
| Top | 6.2% 62,144 mg/l | 63.0 % 39,169 mg/l | 37.0% 22,975 mg/l | 268 | 69,250 | 2534.5 | 6.97 |
| | 6.2% 62,287 mg/l | 62.6 % 38,991 mg/l | 37.4% 23,296 mg/l | | | | |
| **Average** | **6.2% 62,215 mg/l** | **62.8% 39,080 mg/l** | **37.2% 23,135 mg/l** | | | | |
| Middle | 4.1% 41,342 mg/l | 60.0 % 25,085 mg/l | 40.0 % 16,257 mg/l | 326 | 38130 | 1962 | 6.94 |
| | 4.1% 41,430 mg/l | 61.0 % 26,729 mg/l | 39.0 % 16,142 mg/l | | | | |
| **Average** | **4.1% 41,385 mg/l** | **60.5% 25,907 mg/l** | **39.5% 16,199 mg/l** | | | | |
| Bottom | 4.4 % 43,546 mg/l | 61.0 % 26,729 mg/l | 39.0 % 16,816 mg/l | 589 | 43440 | 2922 | 6.87 |
| | 4.3 % 43,438 mg/l | 61.8 % 26,887 mg/l | 38.0 % 16,551 mg/l | | | | |
| **Average** | **4.4% 43,492 mg/l** | **61.4% 25,808 mg/l** | **38.5% 16,683 mg/l** | | | | |

# APPENDIX E
## ANAEROBIC DIGESTER VFA CONCENTRATION BY GC ANALYSIS

## APPENDIX E: ANAEROBIC PILOT SCALE DIGESTER VFA CONCENTRATIONS BY GC ANALYSIS (mg/l)

| Week | Acetate | Acetate Range | Propionate | Propionate Range | Butyrate | Butyrate Range |
|---|---|---|---|---|---|---|
| 1 (seed) | 34.28 | 0 | | | 0 | |
| 2 | 0 | 0 | | | 0 | |
| 3 | 0 | 0 | | | 0 | |
| 4 | 8.184 | 0-40.92 | 4.81 | 0-24.07 | 0 | |
| **33 Day** | **HRT** | | | | | |
| 5 | 0 | | 0 | | 1.69 | 0 – 6.79 |
| 6 | 0 | | 0 | | 1.69 | 0-6.78 |
| 7 | 0 | | 5.155 | 0-20.62 | 3.02 | 0-12.05 |
| 8 | 11.97 | 0-47.9 | 8.11 | 0-32.44 | 16.41 | 0-65.67 |
| 9 | 0 | | 0 | | 0 | |
| 10 | 7.8 | 0-15.6 | 0 | | 0 | |
| 11 | 0 | | 0 | | 0 | |
| 12 | 0 | | 0 | | 0 | |
| 13 | 0 | | 0 | | 0 | |
| 14 | 0 | | 0 | | 0 | |
| **Mean** | **1.97** | | **1.32** | | **2.11** | |
| 15 | 18.21 | 0-31.58 | 0 | | 0 | 0-15.98 |
| 16 | 17.64 | 0-30.04 | 0 | | 14.82 | 0-15.98 |
| 17 | 15.58 | 0-42.11 | 0 | | 10.8 | 0-22.61 |
| 18 | 18.95 | 0-35.15 | 0 | | 3.58 | 0-17.92 |
| 19 | 0.1 | 0-0.146 | 0.028 | 0 –0.196 | 0 | 0 |
| 20 | 0.025 | 0-0.21 | 0 | | 0.14 | 0-0.7 |
| 21 | 0 | | 0 | | 0 | |
| 22 | 0 | | 0 | | 0 | |
| 23 | 0.2 | 0 – 1.2 | 0 | | 0 | |
| **15 Day** | **HRT** | | | | | |
| 27 | 0 | | 0 | | 0 | |
| 28 | 7.61 | 0-15.37 | 0 | | 1.69 | 0-6.76 |
| 29 | 8.165 | 0-16.33 | 0 | | 0 | |
| 30 | 0 | | 0 | | 0 | |
| 31 | 0 | | 0 | | 0 | |
| 32 | 0 | | 0 | | 0 | |
| 33 | 0 | | 0 | | 0 | |
| 34 | 8.185 | 0-16.37 | 0 | | 0 | |
| **Mean** | **2.99** | | **0** | | **0.21** | |
| | | | | | | |
| | | | | | | |
| | | | | | | |