DESIGN AND PERFORMANCE STUDIES OF PILOT SCALE INCLINED TUBULAR DIGESTERS OPERATING ON CATTLE SLURRY

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ABSTRACT

DESIGN AND PERFORMANCE STUDIES OF PILOT SCALE INCLINED TUBULAR DIGESTERS OPERATING ON CATTLE SLURRY

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Two pilot scale (1 and $2m^3$) inclined tubular digesters were designed and constructed adjacent to a large dairy unit. The digesters were instrumented to monitor automatically gas production, temperatures and the electrical energy used for heating. Operation of the digesters was reliable. Aspect ratio, angle of inclination, retention time and type of slurry could be varied independently, and the effect of these parameters on digester performance was investigated in five sets of operating conditions during a 20 month period. These parameters had the following values : Aspect ratio, 10 and 20 ; Retention time, 15 and 20 days ; Temperature 25 and 35 degrees Centigrade ; Feedstock, whole and mechanically separated slurry.

Gas yields ranged from 0.282 - 0.318 m^3/kg VS added with whole slurry at 35 °C. The maximum gas yield was with a digester inclination of 20° to the horizontal, aspect ratio of 10, and retention time 20 days.

For a digester inclination of 20°, gas yields (m^3/kg VS added) from whole slurry were higher at an aspect ratio of 10 than 20 for both 25 and 35°C operating temperatures. At 10° inclination, gas yields were marginally lower at an aspect ratio of 10 for both whole and separated slurry.

Stability of the digestion process was greatest at 20° inclination, this being attributed to the more uniform temperature distribution arising from mixing caused by the evolved biogas, and convection currents.

Tracer studies which were designed to characterize the flow of feed components through the digester, showed actual retention times to be less than the nominal values at the times they were carried out. With separated slurry, the flow of soluble components was very similar to a completely mixed system.

Energy production from the digesters was investigated, digester heating requirement exceeding gross energy production, primarily because of the large heat losses which occurred due to the experimental nature of the plant. It was predicted that this would be improved in farm scale versions.

A technical and economic appraisal established that the design could be scaled up to farm scale (>$100m^3$), but major problems would arise from the need to support a reactor of this size at the optimum inclination of 20°. This was shown in the capital construction cost estimates. Comparisons with the more common continuously stirred tank reactor were made.

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CHAPTER 1

INTRODUCTION

1.1 The Anaerobic Digestion Process

Anaerobic digestion is a naturally occurring process in which the breakdown of organic matter is carried out by bacteria in the absence of oxygen. Related processes occur naturally in a range of environments such as anaerobic muds in marshes and bogs, soils and in the digestive tract of ruminant animals. A gaseous mixture of methane and carbon dioxide, commonly referred to as "biogas" is produced. The microbial population involved is complex, and is comprised predominantly of strictly anaerobic bacteria.

The process has been applied, under controlled conditions, for the treatment of a wide variety of agricultural, domestic and industrial wastes, but often for different reasons. The principle benefits ascribed to the process are energy production, pollution reduction, solids destruction, odour control, reduction of pathogens and, for certain feedstocks, an improvement of the fertilizer value. It is difficult to quantify the monetary value of all these benefits. In the sewage industry for example, the reduction of solids, odour and pathogens are the main criteria on which the process is assessed.

In an agricultural situation, application of the anaerobic digestion process would generally be judged from the standpoint of pollution reduction, energy production and utilization, and possibly odour control.

1.1.1 Historical

The anaerobic digestion process has been taking place naturally for thousands of years in the anaerobic muds in marshes and bogs. Marsh gas was discovered by Shirley in 1667 (Tietjen, 1975), and was also identified by Van Helmont as early as 1630.

During the early part of the 19th century, Davy collected methane from experiments with cattle manure kept in a vacuum (Tietjen,1975). The first recorded digestion plant was built in a leper colony in Bombay, in 1859 (Hull-Bang Dao, 1974), and the biogas used for heating and lighting. Much of the early development work on the anaerobic digestion process was carried out in India, and developments have been continuing (Meynell,

1976) since the nineteenth century. There has always been considerable interest in the process from developing countries, on account of the renewable source of energy made available via anaerobic digestion, from locally available feedstocks, particularly cow dung.

The septic tank was developed in the middle of the nineteenth century, utilising the process for the treatment of human excreta but with no attempt being made to collect the gas produced. However, an important step in the development of the process was made in the 1890's when Cameron designed a septic tank in which gas was collected (Hobson et al, 1974) and used for street lighting in Exeter.

In the Western world, the primary applications of the anaerobic digestion process continued to be for the treatment of sewage. Heating or mixing of the digester contents was uncommon.

Another step forward in the development of digester design came during the early twentieth century, when heating was introduced to digesters to speed up the fermentation process, and reduce the digester volume. The gas produced by the process was regarded as a useful by-product. It was not until the first of the so called "energy crises" that biogas came to be regarded as valuable and strategic. The first energy crisis resulted from World War II, and this stimulated considerable interest in renewable forms of energy, and several hundreds of digesters were built on farms in Europe, particularly in France, to try and alleviate the problems of fuel shortages. However, many of the estimations of the amount of gas which could be produced from agricultural wastes were wildly optimistic, and it was realised that in many instances the energy inputs for digesters operated in typical central European climates were similar to the energy outputs, under good conditions.

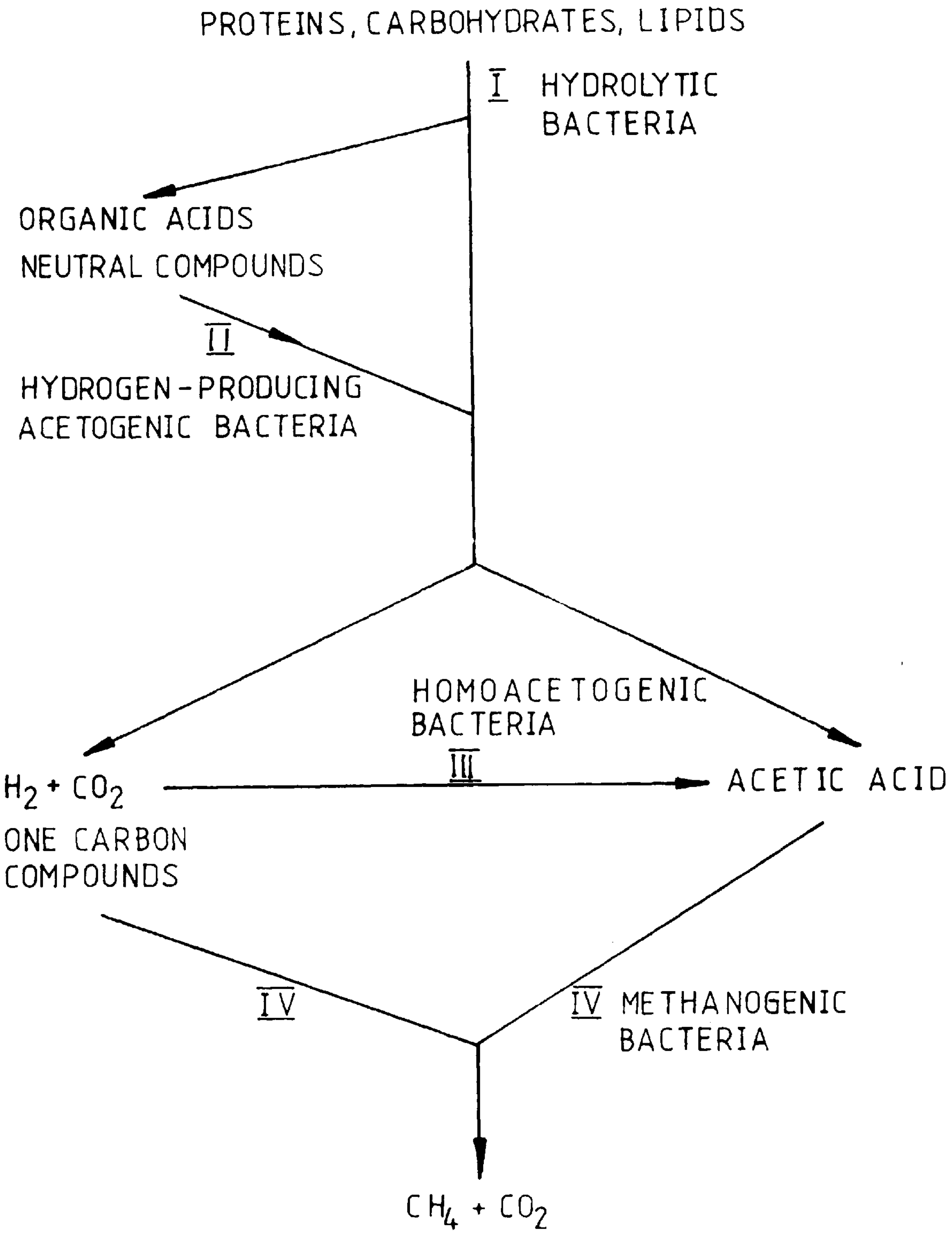
Many thousands of digesters have been installed in the Western world during the twentieth century, mainly for sewage treatment. However, an increasing number of digesters are now being built to treat industrial wastes. The main criteria for installation is usually pollution reduction rather than gas production as such.

In countries such as India, China and Korea, the development of digesters has proceeded along different lines to the Western world. There are two main reasons for this. Firstly, these countries have much warmer climates and so less energy is needed for digester heating. In many instances such digesters are un-heated. The second resaon relates to the social and economic structure of these countries. Digestion technology has been kept simple so that large numbers of people can take advantage of it. Digesters are frequently sized in "family" units, and take the waste from the household and any livestock associated with it. The small amount of gas produced is used for cooking, and the digested effluent returned to the land as a fertilizer. In these countries, financial viability is determined by the ability of outputs (gas and slurry) to substitute for fuels, fertilizers or feeds which were previously purchased for money (Stuckey, 1983a). The status of anaerobic digestion technology in developing countries has been critically reviewed by Stuckey (1983b).

1.1.2 Microbiology

The anaerobic digestion process is a bacterial fermentation by which organic matter is converted to methane and carbon dioxide by a mixed culture of micro-organisms in the absence of oxygen. The interdependency of the different bacteria is complex, and the microbiology of the process is usually discussed in terms of four metabolic groups of bacteria or stages of fermentation (Zeikus, 1980; Hobson and Richardson, 1985). These groups are shown schematically in Figure 1.1.

FIGURE 1.1 Schematic of the anaerobic digestion process showing the four major metabolic groups of bacteria involved.



Zeikus, 1980

The hydrolytic and fermentative bacteria (Group I) hydrolyze materials such as lipids, protein and polysaccharides to individual monomers, which in turn are fermented to various intermediates.

The obligate hydrogen-producing acetogenic bacteria (Group II) metabolize propionic, butyric and possibly other end products of the first stage to additional hydrogen, carbon dioxide and acetic acid.

The homoacetogenic bacteria (Group III) synthesise acetate from hydrogen and carbon dioxide.

The methanogenic bacteria (Group IV) catabolize the end products, mainly acetate, carbon dioxide and hydrogen produced jointly by the other two groups, to the terminal products. Methane is formed by either the decarboxylation of acetate or the reduction of carbon dioxide (Hobson & Richardson, 1985).

The bacteria of anaerobic digesters have not been characterized fully and detailed studies have been relatively few (Hobson & Richardson, 1985). It has not yet been shown whether species of bacteria are common to all digesters, but the reactions seem to be basically the same, although the nature of the feedstock determines which of the hydrolytic and fermentative reactions predominate. It is commonly found with agricultural wastes that hydrolysis is the rate-limiting step (Bryant, 1979; Ferrara et al, 1984).

The scheme separating the three main stages of fermentation cannot be extended to a discussion of their metabolism, since the efficient metabolism of each group is dependent on the others. The kinetics of the anaerobic digestion of farm wastes are considered by Hobson (1983).

When a digester population builds up from its natural innoculum, the hydrolytic and fermentative group becomes established first followed by the slower-growing methanogenic bacteria (Hobson & Richardson, 1985). During this start-up period is the main and possibly the only time when bacterial imbalance can lead to digester failure. When imbalance occurs, intermediates in the reaction pathways build up and cause secondary action which will inhibit or kill some or all of the bacteria.

The rates of bacterial reactions increase overall with temperature. There are two temperature ranges in which different groups of bacteria, the mesophilic and thermophilic, have optimum growth. These are approximately 20-40°C and 45-70°C (McInerney et al, 1980). Farm digesters are generally operated in the mesophilic range at approximately 35°C.

All the biological reactions are affected by pH, and particularly by acidic values, the general optimum being about neutrality or very slightly alkaline (c. pH 7-7.5). A combination of a balanced fermentation and methanogenesis, with ammonia and other ions in the digester fluid, keeps a farm digester at this pH in most circumstances (Hobson & Richardson, 1985).

1.1.3 Engineering considerations

The design of an anaerobic digester and the engineering associated with it depends upon the type and volume of waste it is required to process, together with other factors such as the geographical, environmental and social conditions applicable to the particular situation (Horton, 1980). There is no single design which can be considered as ideal for universal application. However, there are underlying principles which are applicable to the design of all digesters.

The main function of an anaerobic digestion plant is to provide optimum conditions for digestion to take place. The digestion "tank" must be capable of maintaining anaerobic conditions, and be chemically resistant to the material being digested, and the products of digestion.

Most digesters can be classified in two ways - the method by which they are fed (continuously or batch) and the microbial environment that is chosen (primarily a function of temperature and mixing). Additionally, digesters can be classified according to the way in which the active biomass is immobilized, if at all. Different digester designs are discussed in Section 1.2. Discussion here will be limited to continuous digesters, operating at mesophilic temperatures.

To feed a digester, knowledge is needed of the handling properties of the feedstock. The simplest method of feeding is by gravity, where the level of the contents inside the digester is below the level at which the waste is produced and handled. Many digesters are pump fed, the most common types of pump for this purpose being centrifugal, diaphragm and positive displacement. Any pump chosen should be able to handle the feedstock reliably, and associated pipework should be of the greatest diameter possible to minimise blockages. Pipe runs should be kept as straight as possible, again to minimize blockages.

The size (volume) of a digester is determined by the volume of material to be digested, and the desired retention time. Long retention times imply large and therefore costly reactors, but digestion is more complete. Short retention times enable smaller reactors to be built, and an increase in the rate of gas production, but a decrease in the total amount of gas produced from a given volume and lower solids destruction from a given volume of waste. Digester stability is reduced at short retention times, and the chance of bacterial washout is greater. There is an economically optimum retention time for most situations. In designing a digester it is usually necessary to monitor the volume of waste being produced each day, and to carry out batch digestion tests on representative samples in order to determine the optimum retention time. It is usual to build a digester to operate at a retention time greater than the optimum, to allow for any increase in the volume of waste to be treated.

Digester shape wil influence capital cost, mixing and heat transfer within the reactor, and heat losses. Traditional designs of digesters for sewage treatment centred aroung a large reinforced concrete tank, requiring civil engineering expertise in design and construction. More recently, prefabricated glass-enamelled steel tanks have become common in the sewage industry (Noone & Brade, 1982). Such tanks have the advantage of quick assembly (days), and less dependence on civil engineering contractors. The construction methods for prefabricated tanks are now well established, and they are probably the simplest form of digester to construct.

Plug-flow digesters have been built from a wide variety of materials, and orientation of the digester is usually horizontal. The use of an earth lined trench with flexible cover has been pioneered by Jewell *et al*, (1980). Other plug flow digesters have utilized concrete as the main material of construction.

Digester heating

Digester heating is required to raise the feed temperature to that of the digester, and to replace heat losses, primarily through the digester walls. There are four main ways of heating a digester. These are by hot water recirculation, direct steam injection, heating of influent only and the use of external heat exchangers.

Recirculating hot water through a closed loop of internal heating pipes is a widely used method for digester heating. The heating pipes initiate internal mixing by means of thermal convective movement. The main factors influencing efficiency of this system are the pipe surface area, temperature of recirculated water, flow rate of water and thermal conductivity of the pipe material.

Direct steam injection has been used as a method of digester heating, primarily with sewage sludges (Noone & Brade, 1982). Advantages of this system include very simple internal heating apparatus, and mixing from the steam injection.

Heating of digester influent only eliminates all heating aparatus from inside the digester. It is practiced on some sewage works where the feedstock has been treated by a pasteurisation stage, involving heating to $70^{o}C$ (Bruce & Oliver, 1986).

In an external heat exchanger, slurry is recirculated through an external pipe system, and a heat source transfers heat to the slurry. An advantage of this system is that all components are located outside the digester for ease of maintenance and repair. The major disadvantage is the high energy requirement for pumping the slurry.

Whatever heating system is chosen, it must be able to maintain digester temperature under the most severe weather conditions expected. Heat should be uniformly distributed within the digester, as temperature stratification would otherwise occur, leading to the formation of cold plugs or regions of reduced metabolic activity.

Hot water for digester heating is usually obtained from the combustion of biogas in a boiler, or alternatively from the cooling water of an engine in a combined heat and power (CHP) unit. Natural gas or LPG can also be used to fuel boilers for digester heating, and in some industrial applications, waste heat from other processes is available. The use of electric heating (immersion heaters) is expensive but reliable and convenient to use.

The heating system must be such that the contents can be heated to the desired operating temperature on start up. In some instances, this means a secondary heating system is required. If the normal method of heating was by burning biogas, such a system would be needed until the digester was producing sufficient biogas itself to fuel the boiler. Secondary systems are commonly fuelled by gas, or are electric.

Digester mixing

Digesters are commonly mixed, although the precise amount and degree of mixing required is not accurately known (Casey, 1986). Mixing is provided to achieve a uniform temperature distribution within the digester, prevent scum and crust formation, and eliminate settlement within the reactor. Mixing is commonly achieved by compressing the biogas produced and recirculating it through the liquid contents of the digester. Other methods include the use of paddles, and screws in draft tubes. In some digester designs, forced mixing is unnecessary. This is because self-mixing from the evolution of the biogas produced mixes the contents sufficiently.

Safety considerations

Safety measures are a vital part of any digester design. The dangers which surround a digester can be divided into three groups:

1. Physical dangers associated with tanks, ladders, walkways, manholes and machinery with moving parts;

2. Fire and explosion hazards due to ignition of mixtures of biogas and air;

3. Toxic and asphyxiant hazards due to breathing biogas containing hydrogen sulphide.

(Meynell, 1985)

A code of practice concerning safety in and around digesters has been produced by BABA (1982). Whilst this code of practice is not mandatory, many purchasers of digesters, particularly Water Authorities, insist that it is followed. Similarly, responsible manufacturers and designers also follow it.

At some stage it will probably be necessary to shut the digester down either for routine maintenance or repair. A well-engineered digester will enable this to be done easily, and will normally enable the bulk of the contents to be pumped or drained out. Access hatches should be provided for entry of personnel, and possibly conveyors for removal of grit deposits on the digester base. Shut-down of a digester should be carried out with care, and system design should be such that implosion of the tank will not occur. Digester shut-down, for other than planned maintenance, should only occur very rarely for a well engineered digester. Whilst a digester is shut down, alternative and probably expensive ways of dealing with the waste need to be found. Any digester system must receive repair and maintenance, if it is to continue to function correctly.

Control and monitoring

All digesters feature some form of control. The extent to which this control is automated is usually a function of the system cost. Most agricultural digesters are fairly simple, and will continue to function provided feeding is regular, and temperature is maintained at or near the target operating temperature. Feeding can be easily automated with the use of pumps and timers. The simplest indicator that the digestion process is stable, comes from monitoring the biogas production. In the simplest cases, a recording of daily biogas production would suffice. Additional information is gained from measuring the gas composition (%methane and % hydrogen sulphide). On larger units, some form of wet chemical analysis on the digester feed and contents would be carried out.

All digesters should have a comprehensive operating manual detailing the normal operating procedure, and the steps to be taken if performance is not as expected. Equipment for the remote monitoring and control of digesters is presently being developed. This takes the form of a simple datalogger which records information from the digester, and a simple computer which can be remotely accessed from a central office (Stafford and Spensley, 1986).

1.2 Digester types

The main function of an anaerobic digestion plant is to provide optimum conditions for digestion to take place. The conditions are:

1. To continuously provide digester bacteria with nutrients, and to remove the metabolism products from the viable biomass.

2. To ensure optimum retention times for the solid and soluble components of the organic feedstock.

3. To prevent uncontrolled accumulation of solids in the digester, either as sediment or scum.

4. To provide an even distribution of temperature throughout the digester.

There is no one single design which is able to meet these criteria for all feedstocks, due to large differences in the physical and chemical composition of substrates which are digested. For example, early digester designs were for the treatment of solid feedstocks such as cattle manure and batch digesters were widely used (Wheatley, 1978). Digesters have since been built for the treatment of soluble wastes, which required different digester designs. Between these two extremes, there are many digester types, and these are considered according to the following categories:

Type of digester

Batch

High rate

Floc-based

Attached-film

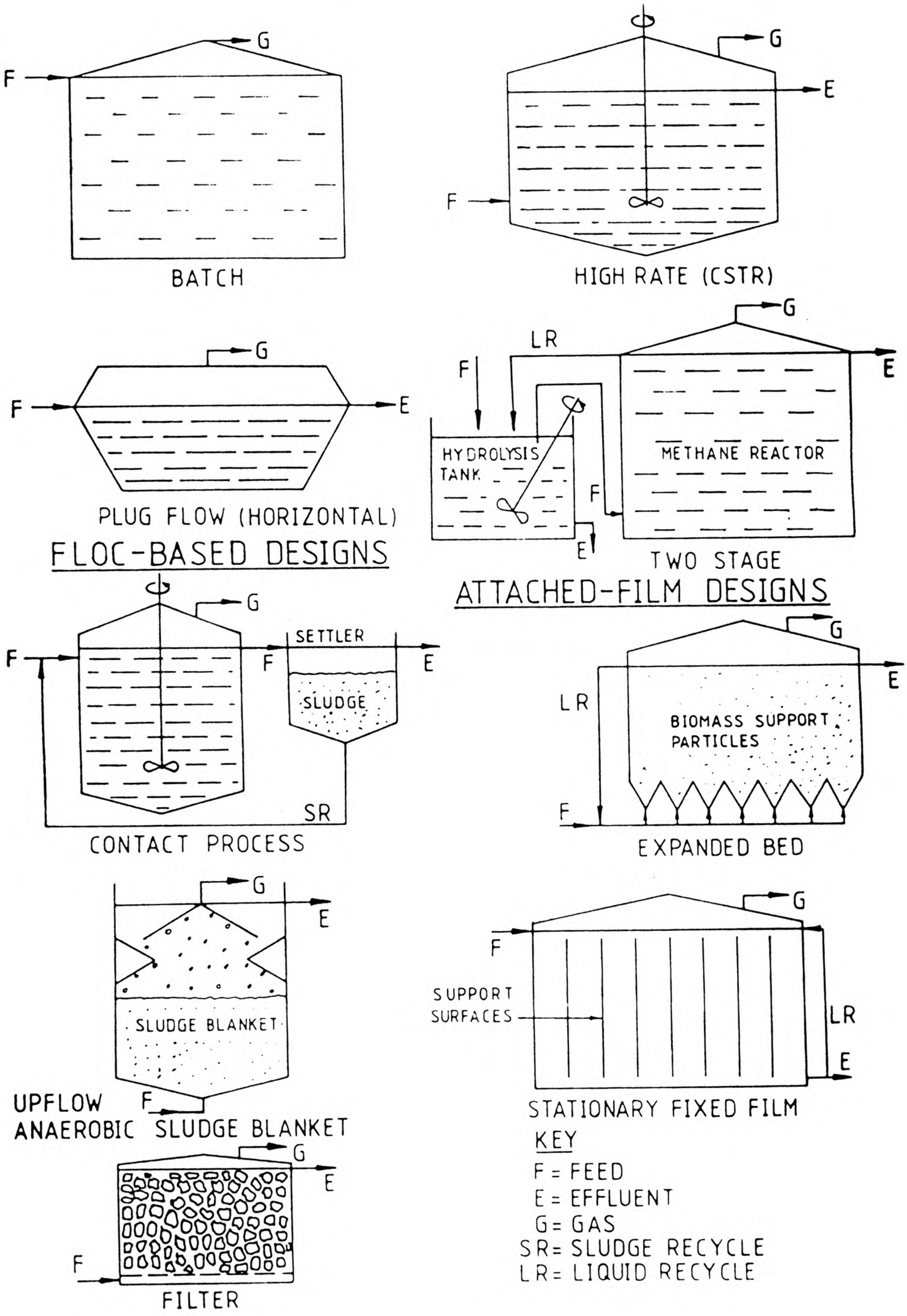
Multi-stage

Plug flow

1.2.1 Batch digesters

This is the simplest type of digester, and it operates on a dis-continuous basis. The material to be digested is placed in the reactor (Figure 1.2) and a quantity of seed material added to start the digestion. The reactor is then sealed, and the contents are left to digest anaerobically for up to 6 months (Meynell, 1976). Although this type of reactor is cheap and simple to build, loading and unloading of the contents can be difficult and therefore expensive. Gas production is uneven, being very low initially, rising to a peak and declining to a low level again, which may be sustained

FIGURE 1.2 Schematic of the main types of anaerobic digesters.



for several months. Daily gas production can be made more even by building more than one batch digester, and filling them in a sequential order. These are referred to as linked-batch digesters. Several hundred batch digesters were built in Europe (Horton, 1980) during World War II, feedstocks being dung, straw and vegetable tops and cuttings. There were both heated and unheated versions. This design is presently being tested with working volumes of 30-250m^3 (Rozzi and Passino, 1985), the most common feedstock being cattle manure. The design can be financially attractive where an existing manure pit is used as the reactor, and no external heating is supplied. The main disadvantages with this design are the irregular gas production (where the gas is required), and difficulties in loading and unloading of the digester.

The main practical uses of this design are found in laboratory or pilot scale studies, assessing the digestibility of a particular waste, before a full scale unit is built. The very simplicity of the design means that several units can be operated side by side in a laboratory, and the performance of full scale digesters fed on the same feedstock predicted.

1.2.2 High-rate digesters

A high-rate digester is, by definition, heated and mixed (Hobson et al, 1981). Many digesters built for sewage treatment were originally constructed without any heating or mixing, and the term "high-rate" is used to differentiate from these, and to signify that a particular digester is heated and mixed. An alternative and less commonly used definition of "modern high-rate processes" is given by Lettinga (1984), who defines such processes as being based on some form of bacterial sludge immobilization. The definition of Hobson is more widely adopted, and will be used here.

The purpose of mixing a high-rate digester is to keep the digester contents homogoneous, at an even temperature and to prevent scum formation. The most common applications of high-rate digesters are in the treatment of sewage sludges, animal slurries and very concentrated (COD >30,000 mg/l) wastes (Mosey, 1981).

Feedstock is added, and an equal volume of digester contents are removed one or more times per day. It is preferable to feed the digester frequently with a small volume of fresh waste, since this minimises the maximum heating requirement, and helps keep the contents homogenous. This mode of operation resembles that of the continuous stirred tank of chemical reactor theory, and this design is commonly referred to as being of the "CSTR" type.

In this digester design, all components of the feedstock remain in the digester for the same period of time i.e. the Hydraulic Retention Time (HRT) and Solids Retention Time (SRT) are the same. High active biomass concentrations cannot be achieved since all biomass production is discharged as effluent. The CSTR is ideal for soluble wastes where the desired retention time of all components is the same, but not for heterogenous wastes such as animal slurries. This is because solids are digested more slowly than soluble components of the feedstock, the degradation of solids is usually the overall rate-limiting reaction in the anaerobic digestion of farm wastes (Hobson, 1983).

Digester tanks were traditionally made of reinforced concrete (Noone and Brade, 1982), the most common material now being vitreous-enamelled steel tanks, these having the advantages of good corosion resistance and ease of assembly.

Mixing was frequently by mechanical stirrers (Rozzi and Passino, 1985) but gas mixing, where a proportion of the biogas produced is compressed and discharged through nozzles in the digester base, is becoming more common. Disadvantages of mechanical mixing include the forces transmitted to the digester tank, and the difficulty of keeping mixer shaft seals gas-tight. Additionally, where wastes are fibrous or contain large quantities of rags as in some sewage, then the mixer can become fouled.

The performance of high-rate digesters can be improved in some instances by changing the operating mode of the digester mixing and feeding (Callander and Barford, 1983). They were able to increase the retention of both active biomass and digesting solids, by stopping digester mixing for 1 hour before feeding. A supernatant liquid low in active biomass and digesting solids is produced, and discharged when feedstock is added to the digester. This operating mode will only lead to increased performance where considerable settling of the digester solids contents occurs in a period of 1 hour. In slurries of >8% TS, settling is negligible and it is unlikely that performance would be improved by operating digester mixing and feeding in this way.

1.2.3 Floc-based digester designs

Floc-based designs are digesters where the natural tendency of biomass to form aggregates (flocs), which are large enough to be separated from waste water by settling, is encouraged. This increases the digester biomass concentration, by retaining biomass within the digester for a longer time than the waste being digested. Biomass can exist in a variety of forms, the main ones being individual cells, colonies, flocs, pellets and biofilms. The formation of any particular structural aggregate depends on several factors including the size range of cells within the microbial population, and the location of each cell relative to other cells and the medium, for example at a gas/liquid interface (Stronach et al, 1986). The formation of different forms of biomass is dicussed comprehensively by Stronach et al (1986). The term "floc-based" is used here to characterize digester designs where flocculation, the aggregation of dispersed cells, is encouraged. The main digester designs following these principles are:

a) Anaerobic contact process

b) Upflow anaerobic sludge blanket (UASB)

c) Anaerobic filter

1.2.3.1 Anaerobic contact process

In this design, effluent passes from the digester (Figure 1.2) to a settlement tank where the flocculated biomass settle out and are returned to the digester via a recycle pump. This type of plant is used to treat medium-strength wastes (2000-20,000 mg COD/l) at very high hydraulic loading rates (Mosey, 1981).

The main problems with contact digesters are in the separation and concentration of biomass flocs prior to their return to the digester. This is a function of the particular sludge, since some sludges have good settling characteristics. Problems are also caused by gas bubbles adhering to the biomass flocs entering the settlement tank (Rozzi and Passino, 1985). The problem has been overcome with the use of flotation separating devices, or by temporary inhibition of the methanogenic activity during the settling phase. Such inhibition can be induced by a rapid temperature drop (5-10^{o}C).

1.2.3.2 Upflow anaerobic sludge blanket (UASB)

This design uses the same basic technology as upflow floc blanket processes for the removal of suspended solids from potable water. Unlike the contact digester where biomass is removed from and re-cycled to the digester, the UASB retains biomass in the digester either as flocs or dense granules.

Figure 1.2 shows a UASB digester. Feed, which is fed into the digester through a series of pipes on the digester base, rises slowly through the granular sludge bed where most of the biomass is found (Rozzi and Passino, 1985). Above the sludge bed is the sludge blanket where the preliminary separation of gas bubbles from the sludge flocs and granules occurs. The rising gas maintains the biomass in this region in a fluidised state (Callander and Baford, 1983). At the top of the digester, there is a baffle arrangement which separates the gas, solid and liquid fractions. This quiescent zone is free of gas bubbles, so encouraging the settling of bacteria.

The design is conceptually very simple, consisting of a reactor tank with a baffled roof. However, to function effectively, operation and in particular the first start-up of the reactor are critical (Lettinga, 1984). In the initial start-up, low density seed sludge components are allowed to wash out of the reactor, while active biomass with better settling properties is retained.

The process has been developed by Lettinga in Holland since 1971 (Lettinga _et al_, 1980), initially for the treatment of sugar beet waste waters. Wastes which have been successfully treated with UASB reactors (100-5000m^3) include those from processing sugar beet, potatoes, maize starch and brewery wastes (Lettinga _et_ al, 1980; Callander and Barford, 1983). More than 20 commercial applications of the process have been made.

The design is most suited to treating soluble wastes of low total solids concentrations (<1%). High loading rates (>25 kg COD/m^3 day at 30°C) have successfully been used with sugar beet and potato processing wastes, with liquid retention times as short as 3 hours (Lettinga et al, 1980). Removal efficiencies of 70-95% COD have been obtained (Rozzi and Passino, 1985).

1.2.3.3 Anaerobic filter

The anaerobic filter consists of a reactor packed with a support medium (Figure 1.2). Anaerobic filters can be either upflow or downflow, and were first investigated extensively by Young and McCarty (1969).

The support medium was originally provided to act as a fixed surface for biomass attachment. However, it has been observed that a major portion of the biomass is not attached to the support medium, but is in flocs in the interstitial spaces (Callander and Barford, 1983). Performance of anaerobic filters has been improved through the use of high voidage (0.8 - 0.9) packing materials. The first materials to be used for packing materials were stones of 25-40mm diameter, but other materials providing an increased surface area for bacterial attachment such as PVC, polyester and glass beads are now more common (Van Den Berg and Kennedy, 1981). The design has successfully been used to treat vegetable processing wastes of 1-10% TS.

1.2.4 Attached-film digester designs

In attached-film digester designs, large surface areas for bacterial attachment are provided within the reactor, and the volume occupied by the media minimised. The main form of biomass in these designs is as biofilms. The main digester designs based on these principles are the expanded bed, the fluidised bed and the stationary fixed-film types.

1.2.4.1 Expanded and Fluidised bed

These two designs are very similar, the main difference being in the degree of bed expansion. The basic design concept is that of a reactor filled with a fine granular material (0.5mm diameter) to which bacteria adhere in a thin film. The support material is partly fluidised by the upward flow of recycled digester contents to give a bed expansion of 10-20% for expanded bed processes, compared with 30-100% expansion for fluidised beds (Callander and Barford, 1983). A disadvantage with the fluidised bed reactor is the large amount of energy required to recycle the effluent and keep the bed fluidised. This energy requirement can be minimised by using support particles of low density.

There are relatively few expanded or fluidised bed reactors in commercial operation, development work is mostly at the laboratory and pilot scale (Rozzi and Passino, 1985). Neither of these two processes are subject to clogging (Jewell, 1980) and full scale applications with organic slurries such as animal manures and sewage sludge should be feasible.

1.2.4.2 Stationary fixed-film

The principles of this design are similar to those of the expanded and fluidized bed i.e. provision of a large surface area for bacterial attachment. However, in this design, the bacterial film grows on a stationary, vertical support material within the reactor. Both upflow and downflow modes of operation have been used. Support materials which have been used include PVC sheeting, glass, needle punched polyester and clay drainage pipes (Van Den Berg and Kennedy, 1981). These materials have given surface areas for bacterial attachment ranging from $50\text{-}250m^2/m^3$ reactor volume.

1.2.5 Multi-stage digesters

Digesters have been designed with more than 1 reaction vessel, to separate the acidification and methanogenesis stages of the digestion process. The reactors are usually connected in series, this approach allowing environmental conditions related to both phases to be optimised (Cohen, 1983). The microbial species active in the acidification stage are physiologically and metabolically very different from the obligatorily anaerobic acetogens and methanogens (Barry *et al*, 1982). Two stage processes have substantial benefits over conventional single stage high rate digestion systems (Ghosh and Klass, 1978). The main advantage of a two-phase digestion system is that the maximum biomass activity of the methane forming step is typically three times that of a conventional single-phase system (Callander and Barford, 1983). Digester stability is also improved.

Disadvantages of multi-reactor systems include greater capital cost for the process volume (2 or more reactors), and the associated necessary instrumentation for monitoring and control to maintain optimum conditions in each reactor, and in the system as a whole.

1.2.6 Plug flow digesters

There is theoretically no axial mixing in a true plug flow reactor. If this criteria were met, then it would be impossible for digestion to take place since there would be no innoculation of the incoming feed material.

The term "plug flow" is established nomenclature in chemical engineering, and many reactors of this type have been built. However, the term has been widely applied to digesters operating on primarily agricultural wastes, where the digestion vessel is tubular and of horizontal orientation. The successful operation of such reactors is dependent on axial mixing and solids retention. This type of reactor is more accurately described as a displacement type digester.

The first reported plug flow digester was that of Buswell and Boruff (1933), designed for the continuous fermentation of fibrous materials. A similar system was designed by Reinhold and Noack in Darmstadt, and became known as the "Darmstadt System" or "Fermentation Canal Process" (Horton, 1980).

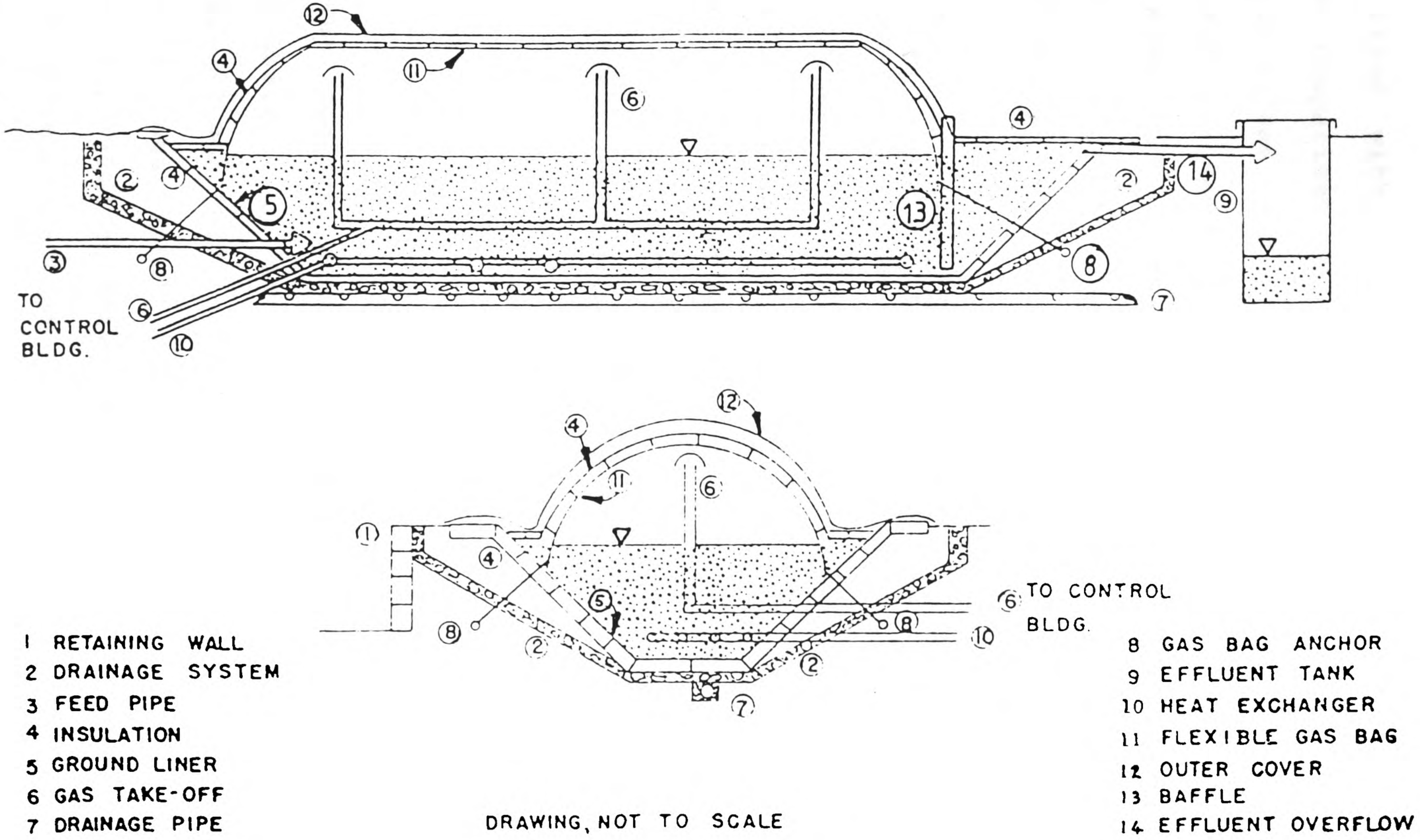
In both systems, an agitator on a longitudinal shaft the same length as the reactor, was operated several times per day. Heating was by steam injection. No operational data is available. The construction of several "Darmstadt" type systems on farms has been reported by Tietjen (1975).

During the late 1950's, 2 plug flow digesters were built on a farm in South Africa (Fry, 1974), to minimize the pollution arising from more than 1000 pigs. The design and subsequent construction were carried out by the farmer working quite independently of any other organisations. His rationale for choosing horizontal orientation are unclear. The two digesters were constructed from reinforced concrete with a common central wall. Scum accumulation was a major problem and the digesters were eventually shut down. To try and overcome the problem of scum formation, Fry (1974) constructed a "scum drag" in each digester. In operation, the level of the digester contents was lowered, a hatch on the digester end was opened, and a rake on chains pulled the scum through this hatch. Provision was also made to remove sediment from the digester in a similar manner. Whilst providing a means of removing scum, crust and sediment from the digester, the mechanical complexity of the design was increased. The digester design did not seek to minimize scum formation in any way. Also, the digesters were inoperative whilst this operation was carried out. There is little quantitative data associated with this work.

A prototype plant of this type was shown at the 1963 Royal Agricultural Show (Anon, 1963), but it is not known if any digesters of this type were installed on farms in the UK. Martin and Dale (1978) report the construction of a similar unit at Custer, Michigan operating on beef cattle waste. Only limited performance data is available for that installation.

A comprehensive study on the technical, practical and economic feasibility of using anaerobic digestion on small dairies (40 and 100 cows), and medium sized beef feedlots (1000 cows) was carried out at Cornell University by Jewell et al (1976). A subsequent study (Jewell et al, 1978) had the development of a simplified low cost reactor design for small scale operations as one of its main priorities. The design concept identified was that of an unmixed digester, with all slurry flows being by gravity. A schematic of the full scale plug flow reactor system developed by Jewell is shown in Figure 1.3. The use of soil-supported structures, and inexpensive materials which could be erected by a farmer was also considered necessary. An extensive programme of research and development was carried out, at laboratory, pilot and full scale ($38m^3$). The final design concepts

FIGURE 1.3 SCHEMATIC OF FULL SCALE PLUG FLOW REACTOR SYSTEM



JEWELL et al, 1982

featured earthen basins that could be graded, insulated and lined with a hypalon rubber material. In a full scale comparison with a completely mixed reactor of similar volume ($35m^3$), the plug flow reactor maintained a consistently higher level of performance with respect to gas production, and solids destruction (Jewell et al, 1981a). The feedstock was "as received" dairy cow manure of 10-12% TS.

The major weaknesses in the designs of Jewell are in the anchoring of the flexible digester cover, and the small range of feedstocks and bedding which can be handled. In early designs the digester operating pressure was 200mm water gauge (Jewell et al, 1980). However, several failures of the digester roof occurred. The present recommendation is for digester operating pressure to be kept below 25mm water gauge (Jewell et al, 1982).

The digester design has operated satisfactorily on cattle waste at an "as received" 10-12% TS concentration. However, if the feedstock is diluted, perhaps by addition of rain and wash water, the viscosity of the substrate is lowered and crust formation and sedimentation take place. This problem is accentuated when bedding materials of either straw or wood chips are in the feedstock, since they float readily and accumulate at the digester surface. It was found that flotation of bedding materials was not a problem when reactor temperature was 25°C, due to the increased viscosity of the digester contents, compared with the more common operating temperature of 35°C. At an operating temperature of 25°C, pollution reduction by the digester will be less than at 35°C.

The constraints regarding bedding material and influent solids concentration pose severe limitations on the widespread use of this design. These findings were confirmed by the work of Hills (1983) who concluded that a range of 12-15% TS influent concentrations were necessary for satisfactory operation with beef feedlot manures. However, Chen _et al_ (1983, 1984) found scum accumulation in plug flow digesters to be the same for 7-9% TS, and 10-12% TS dairy manure collecting using a flush system. This apparent anomaly was theorized by Chen _et al_ (1984) to be due to the removal of water soluble binding agents during the flushing and subsequent solid-liquid separation process. They recommend that plug flow digesters should not be used with flushed dairy manure. Flush cleaning of dairy barns and pig housing was first introduced to the United States in the 1970's (Moore and Miner, 1985). It is attractive compared with more conventional scraped manure systems due to its low labour requirement, ease of automation and superior cleaning (Hermanson, 1985).

Similar design concepts to those used by Jewell feature in the plug flow digester described by Steinsberger and Shih (1984) operating on poultry waste. Again, earth lined trenches were used, but the primary and secondary digesters were horizontal, and made from a cylindrical plastic liner. The liner was a flexible PVC material made from Red Mud Plastic (RMP). This material is claimed to be cheaper, show greater thermal stability and weather resistance, compared to other compounded PVC materials (Hao et al, 1979). Since the reaction vessel is cylindrical, there is no possible source of gas leaks between the 2 piece liner design used by Jewell. The digesters are covered with a wooden structure.

Plug flow digesters have been built in the USA as practical, commercial installations, as well as the research and development units previously described. Examples of commercial plug flow digesters include those described by Gilman and Bennett (1984), Martin and Lichtenberger (1981) and Scheller (1982). Cournoyer et al (1984) report the of a mixer (gas recirculation) in a digester to minimize scum formation with a feedstck of 8% TS dairy manure. Whilst solving the scum formation problem, this detracted from the simplicity and hence low construction and operating costs of the plug flow digester. As such, the reactor is more accurately described as a mixed plug flow digester.

The number of full scale digester installations (plug flow and completely mixed) on farms on the USA in 1984 was estimated to be 84, and in Canada 13, by Wentworth (1984).

Plug-flow digesters have been built in Europe, mainly as experimental facilities. In a recent survey, Demuynck _et al_ (1982) identified 66 plug flow units operating on agricultural wastes in the EEC and Switzerland.

Some of the more different designs are to be found in Switzerland and are described by Wellinger (1985). All horizontal below-ground digesters are made of concrete in Switzerland. This is a requirement of that country's water pollution legislation. There are three main types of sunken digester. In the first type, a rectangular or circular tank is mixed intermittently with a paddle stirrer, and this design is again a "mixed" plug flow type. The second type of digester is an improved version of the "Darmstadt" design. A baffle is incorporated in the digester base to minimise short-circuiting of the slurry flow, and a series of paddles are attached to the central rotating shaft at the effluent end to remove scum and sediment. In the third type, manure is collected below the slatted floors of livestock buildings, and the digester replaces the usual manure storage pit. This system, which is un-heated, is really a semi-batch system. Due to its location beneath the livestock housing, heat is obtained from the warm air there. Additionally, much of the heat in the fresh manure is retained and digester temperature seldom drops below 15°C (Wellinger, 1985). At such temperatures, psychrophilic bacteria predominate and the process is low rate. However, since Swiss law requires farms to have 100 days manure storage capacity, this system offers a level of waste treatment at a cost comparable to traditional manure storage. Net gas production is reported to be comparable with digesters operating under mesophilic conditions, since there is no separate digester heating system.

The first full-scale ($324m^3$) plug flow digester to be built in the UK for agricultural wastes was designed by Helix Multi Professional Services (Dodson, 1981). The design is similar to that developed by Cornell University, except that concrete walls and a GRP roof are used. The roof consists of 9 interlocking GRP caps, held in place by concrete locking weights. The "roof" provides gas storage for the digester. The caps can be removed individually to inspect the digester or to carry out repairs and maintenance. The digester operates on scraped dairy manure at approximately 11% TS (Crocker, 1985). All slurry flows are by gravity. Even though the digester had a gas recirculation system for mixing, this was ineffective and temperature distribution throughout the reactor was non-uniform (Friman, 1984). The installation of a pre-heating chamber for the feedstock improved the temperature profile in the digester (Friman, 1986).

A below ground, plug flow digester ($30m^3$) has been built at University College, Cardiff, and operated on pig slurry (Stafford and Etheridge, 1982). The main advantages claimed for this cylindrical unit made from GRP are quick installation, simplicity of operation and aesthetic considerations. Little quantative data has been reported on this work.

The use of plug flow digesters for the treatment of sewage was investigated by Casey and Power (1984), who showed that a stable high-rate digestion process could be operated at retention times in the range 12.5-20 days, in a $3.1m^3$ digester. There had previously been little interest in the use of plug flow reactors for the treatment of sewage. This work was initiated to see if old sedimentation tanks 29m square by 2.4m deep could be converted to plug flow digesters at modest cost. It is not known if the process was scaled up to this application.

1.2.7 Inclined tubular digesters

Inclined tubular digesters are displacement type digesters where the reaction vessel is inclined at an acute angle to the horizontal. The feed is introduced to the digester at the lowermost end of the reactor using a gravity displacement system, or pump. The main advantages of inclined over horizontal digesters are that the surface area where scum and crusts can form is minimised; no forced mixing is necessary; gas yields per unit mass of VS added are usually greater than from comparable CSTR's due to increased retention of solids.

1.2.7.1 Laboratory scale

Applications of the inclined tubular digester have been reported by Buhlert et al (1981), Hawkes et al (1981) and Floyd (1984).

The work described by Buhlert et al (1981) was for the treatment of wine stillage and pea blancher wastewater (20,000 mg COD/litre). The digester had an aspect ratio of 22.45, and a total volume of 2.91. Digester inclination was not reported. This design was selected since the washout of suspended sludge has been shown to limit the maximum rate of treatment in a conventional mixed anaerobic process. The digester removed solids with high efficiency, and removal of sludge bacteria was minimal.

Work by Hawkes <u>et al</u> (1981) also showed an inclined digester to retain solids when operated on a 1% TS piggery waste. Gas yields per unit mass of VS added were greater than those reported for CSTR systems. Further work on this digester design was reported by Floyd (1984). Experiments were conducted with pig slurry ranging from 1-10% TS, solids retention being greater with slurry of low solids content. Digester inclination was 16-20° to the horizontal, and the formation of scum was not a serious problem. However, it was necessary to incorporate a manual agitator at the effluent end (uppermost) of the digester to break up surface scum and crusts intermittently.

1.2.7.2 Other studies

A $2m^3$ inclined (10-15°) plug flow reactor was used in studies on the digestion of maize and potato slops by Braun and Huss (1982). The reactor incorporated an electrically driven blade stirrer, so the reactor was really a mixed plug flow system. The overall performance of the reactor was similar to completely mixed systems, and no specific advantages were claimed for the design.

An inclined tubular digester of approximately $2m^3$ was made from a butyl rubber bag supported on a concrete ramp at the University of Nottingham (Wilton, 1985). The bag failed several times and the project was eventually abandoned. No operational data was reported.

A full-scale ($95m^3$) inclined (15^o) plug flow reactor has been constructed in Austria (Himmel, 1984). The reactor was constructed from 8 prefabricated, reinforced concrete sections, and was buried below ground. The substrate used was mostly pig manure (2.6-5.6% TS), although a small amount of dry chicken manure (74.1% TS) was also added. The digester included a hydraulic jet system for breaking up scum and crust layers, and a mixing system using gas recirculation was also included. The digester functioned satisfactorily for approximately 1 year, but scum and sediment accumulation forced its shutdown. Trouble had also been experienced making effective seals between adjacent concrete sections. During the shutdown period, a submersible mixer pump was incorporated in the upper third of the reactor, and this is operated for 8 minutes every 2 hours. The digester is now a mixed plug flow system. No performance data is reported. This digester was constructed without any preliminary studies at the laboratory or pilot scale, and it was rather ambitious to embark on such a design at so large a scale.

1.3 Farm scale anaerobic digestion

Anaerobic digesters have been installed on farms for one or more of the following reasons:

a) As a means of pollution control, possibly to meet legislative requirements

b) As a source of renewable energy

c) Improved slurry management

A survey of biogas plants operating on agricultural wastes had identified 378 full scale plants and 42 pilot scale plants in the EEC and Switzerland at the end of 1982 (Demuynck *et al*, 1984). Switzerland accounted for 108 of these digesters. Tables 1.1 and 1.2 show the distribution of biogas plants in the EEC and Switzerland, and the number of biogas plants per type of agricultural waste.

The relative importance of the different reasons for installation of biogas plants has changed. In the early

TABLE 1.1 Distribution of anaerobic digesters in the European Economic Community and Switzerland according to the type of waste treated.

Country	Number of plants per type of waste				
	Agricultural wastes	Energy crops	Domestic refuse	Industrial wastes	Total
Belgium	25			10	35
Denmark	23			6	29
France	74		5	15	94
F.R. Germany	75		10	12	97
Greece	4			1	5
Ireland	5			4	9
Italy	63	1	1	13	78
Holland	22		11	23	56
United Kingdom	21		9	5	35
Switzerland	108				108
TOTAL	420	1	36	89	546

Demuynck et al, (1984)

TABLE 1.2 Number of biogas plants per type of agricultural waste

Country	Number of plants per type of waste			
	Liquid or semi-solid	Solid	Mixed	Total
Belgium	21	1	3	25
Denmark	18		5	23
France	32	32	8	72
F.R. Germany	57	10	7	74
Greece	2		2	4
Ireland	5			5
Italy	58		5	63
Holland	14	1	7	22
United Kingdom	11	1	9	21
Switzerland	73		31	104
TOTAL	291	45	77	413

Demuynck et al, 1984

1970's, energy production was seen as a major incentive due to the first of the so called energy crises. However, many of the estimates of the contribution which anaerobic digestion could make to meeting national energy requirements were wildly optimistic. This was partly due to bad design and construction of digesters, but more importantly there was a lack of knowledge regarding performance of digesters under farm-scale conditions. Anaerobic digester systems were generally evaluated on the value of the biogas produced, since in most instances this was the only product from the process which could be assigned a definite value.

Performance was often not up to design predictions because the TS content of digester feedstocks was far less than expected. As an example, with liquid fed pigs, the mean TS slurry content is usually about 6%, whereas in practice it was commonly found to be about 2% (Nielsen, 1985). This dilution was attributed to ground water entering collection pits, leaking drinkers, and other sources of water e.g. rainwater, entering slurry handling systems. This had a two-fold effect on digester performance. Specific gas yields are higher at greater solids concentrations, and the dilution means additional energy is required for digester heating.

Whilst many farmers and designers saw the biggest problem as producing energy, in reality the greatest difficulties were in watching energy supply and demand. On a typical dairy farm, the major requirement for energy occurs at milking time, usually twice per day. However, a digester produces energy on a continuous basis 24 hours a day, so some form of gas storage is usually necessary. Also, the greatest net energy production from a digester is in the summer months when ambient temperatures are highest. However, the greatest demand for energy on the farm is usually in the winter months when ambient temperatures are lower, and digester heating requirement is greatest.

In some livestock production systems, the amount of waste available (i.e. collectable) varies with the time of year. This is the case with over-wintered dairy cows where apart from during the winter, the only waste which could be collected would be that from collecting yards prior to milking, and parlour washings. It may be economical in some instances to "import" another material such as poultry waste to maintain digester operation. This has been carried out at the farm scale (Oliver *et al*, 1986).

Small numbers of farm digesters continued to be built in the UK, either as "one-off" prototypes built by individual farms or as collaborative ventures between farmers and manufacturers (Friman, 1986). This situation meant that there was no "standard" agricultural digester design, and experience from design and operation of full scale plants accumulated relatively slowly. Grant aid for the installation of farm digesters (other than for research) was not available at this time. Performance of such digesters was often not up to design predictions (Chesshire, 1986). This can largely be attributed to poor digester management by farmers, and a lack of basic knowledge of the process. Frequently the gas yields from farm digesters have been considerably lower than results obtained under laboratory or experimental conditions, where there is usually far greater control of the amount and TS content of the slurry.

Digesters have been installed on intensive pig enterprises for the reduction of odour (Voermans, 1985). In some EEC countries, particularly Holland and Italy, the installation of a digester or other treatment system is necessary to meet legislative requirements regarding odour emissions, and effluent quality standards (Demuynck et al, 1984). Digester installations are particularly attractive financially where the cost of alternative treatment methods is high, and where a producer could be forced to close an animal production unit due to legislative requirements. Situations such as this are most likely to arise where large livestock units, with limited land or other disposal route for livestock waste, are situated close to residential areas.

Increased fertilizer value of digested compared with untreated slurry is a benefit often ascribed to the anaerobic digestion process, but rarely supported with experimental results. Organic nitrogen compounds in the feedstock are thought to be converted into forms more readily available to plants by the anaerobic digestion process. There are contradictory reports in the literature. Field plot trials carried out in Denmark over 3 years comparing untreated and digested slurry, indicated there was no significant difference in the crop yields from the different slurries (Nielsen, 1985). However, a 20% improvement in the efficiency of utilisation of slurry nitrogen from digested, compared with untreated slurry was found by Pain (1986). In this instance, the plot trials were carried out over 2 years. It is not known if this increased availability was due to physical or chemical effects. Digested slurry will be of lower total solids content than untreated slurry, and this dilution effect could lead to increased availability of slurry nitrogen, rather than anaerobic digestion *per se*. Any such benefits would be very much dependent on slurry management subsequent to digestion, particularly application of slurry at the optimum time.

There have been 20 anaerobic digesters constructed on farms in the UK since 1976 (Friman, 1986). Of these digesters, 8 with capacities $>50m^3$ are operational, 4 treating cattle slurry and 4 treating pig slurry. There are 4 digesters with capacities $<50m^3$ treating high solids FYM, beef cattle manure, cow manure with vegetable wastes and pig slurry. Of the 20 digesters, 3 full scale plants are not operational, 3 have been dismantled and 2 full scale plants are shortly to be commissioned. In addition to these digesters on private farms, a further 4 digesters were constructed for research and development purposes by government bodies.

Grants towards the installation of digesters on farms became available to farmers from MAFF, in certain circumstances from 1985 (MAFF, 1985). These grants are awarded under the Agriculture Improvement Scheme, and are designated "Farm environment and energy-saving grants".

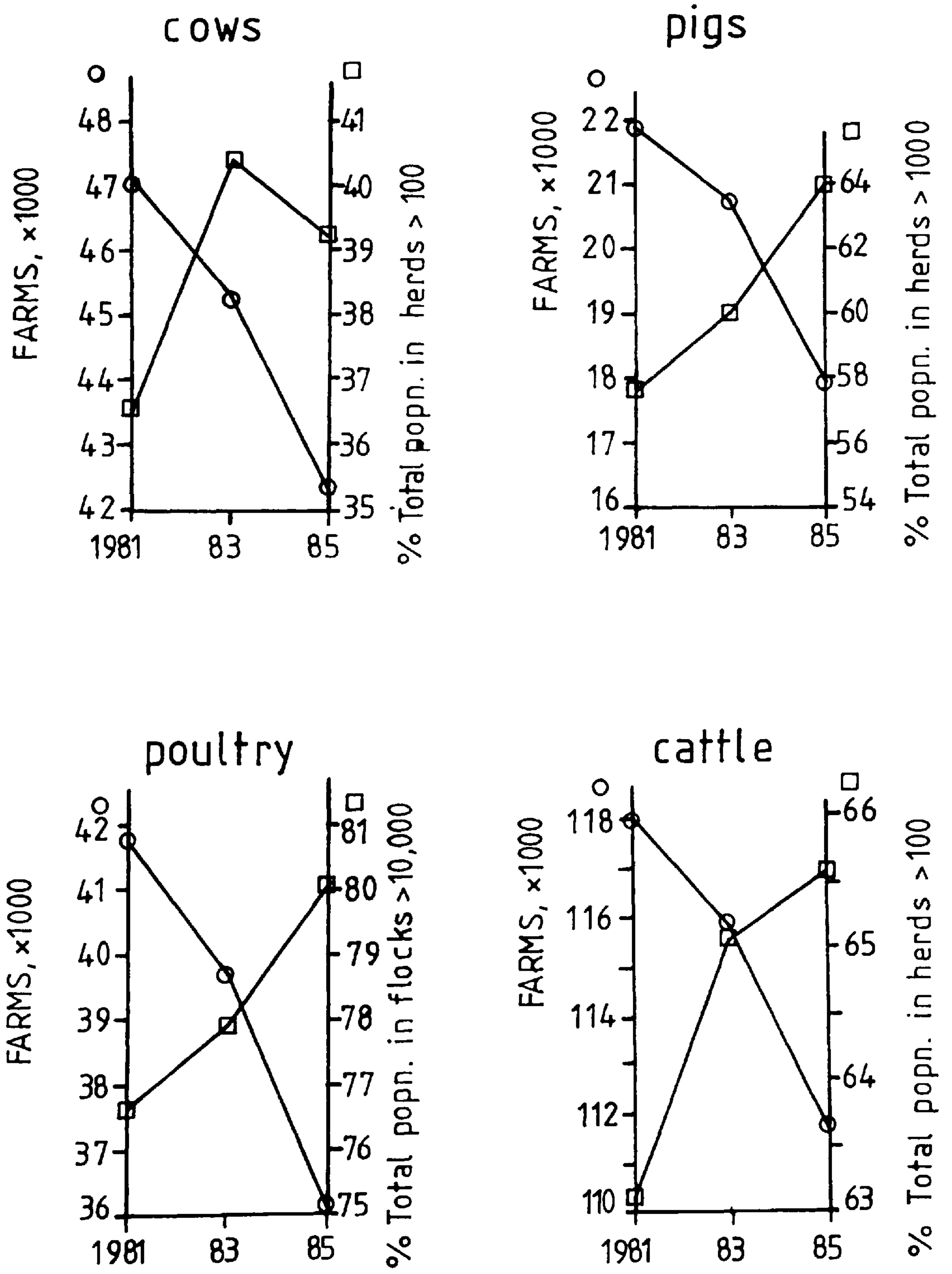
1.4 Livestock wastes in England and Wales

The main function of livestock farming is to produce meat or some other product such as milk or eggs. As a result of the production process, by-products arise. The most obvious of these is livestock excreta, which is often designated as a "waste", implying it has little or no value. Livestock waste needs to be carefully managed to obtain the maximum fertilizer value, and to avoid pollution. Traditionally, this was not difficult to achieve since livestock units generally had enough land associated with them for land disposal of their manures. However, livestock production units have increased in size to achieve economies of scale, and as urban areas have encroached on farms, management of manures has become more difficult. Figure 1.4 shows how for each of the 4 livestock types presented, the total number of farms with each type of livestock has decreased during the period 1981-85. However, during this same period, the percentage of the total population of each animal type kept in large groups has increased, with the exception of dairy cows during the period 1983-85 when there was a slight decrease. Similar trends were found by Grundey (1980) for livestock numbers over the period 1970-78.

There are a variety of disposal or treatment routes available to the farmer, the most common being storage followed by subsequent land spreading. Processes particularly suited to production units with small amounts of land close to residential areas include aeration and anaerobic digestion.

The volume and type of waste produced by an animal depends on its size, age, breed and housing system (MAFF, 1983). Various estimates have been made of the amount of livestock wastes produced in the UK (MAFF, 1981;

FIGURE 1.4 Population of dairy cows, pigs, poultry and cattle, 1983-85 (England and Wales).



MAFF, 1986a

O'Callaghan _et al_, 1971), the most comprehensive being by Larkin _et al_ (1981a). In that study, livestock numbers were taken from the June Agricultural Census for 1976 (Larkin _et al_, 1981a), and the approximate period of housing for each class of livestock was determined. Values for the daily production of waste by each livestock category were obtained from the literature, and the total collectable livestock waste production is summarised below:

Livestock waste production in England and Wales

	England and Wales kT dry matter
Cattle: Dairy	2,176
Beef	1,242
Pig wastes	832
Poultry wastes	1,264
Total	5,514

Larkin _et al_, 1981b

It must be recognised that the above estimates are of the total dry matter production. Depending on the type of animal housing, the mass of waste collected will generally be greater, due to the inclusion of bedding material and water. Rainfall should be excluded from animal slurry wherever possible, since it increases the volume, and hence cost, of material to be handled as slurry. Rainwater (from buildings) should be handled separately as clean water. Estimates of the daily volume and moisture content of excreta produced by livestock are given in Table 1.3. The actual volume collected is very site specific.

The largest category of wastes are those arising from dairy cattle. In 1985, there were 6767 herds of more than 100 cows, and these made up 39.2% of the total dairy cow population (MAFF, 1986a).

1.5 Objectives of present study

The objectives of the present study were:

1. To investigate the performance of pilot scale inclined tubular digesters operating on cattle slurry, with regard to pollution reduction and energy production.

TABLE 1.3 Approximate amounts of excreta produced by livestock.

Type of livestock	Body weight kg		Amount of excreta (faeces & urine or droppings) litre/day		Moisture content of excreta %
	Range	Approx mean	Range	Best estimate	
1 Dairy cow	450-560	500	32-54	41	87
1 Beef bullock	200-450	400	19-28	27	88
1 Pig - dry meal fed	20-90	50	2.0-5.5	4.0a	90
1 Pig - liquid fed (water:meal ratio 2.5:1)	20-90	50	2.0-5.0	4.0a	90
1 Pig - liquid fed (water:meal ratio 4:1)	20-90	50	4.0-9.0	7.0a	94
1 Pig - swill fed	20-90	50	-	14.0a	98
1 Pig - whey fed	20-90	50	9.0-15.0	14.0a	98
1 Dry sow	-	125	-	4.5	90
1 Sow and litter to 3 weeks	-	170	-	15.0	90
1000 Laying hens	1800-2300	2000	100-140	114	75
1000 Broilers (and litter)	100-2000	-	56-63	68kg	30

(a) Amounts of excreta produced over liveweight range 20 - 90 kg i.e. production per pig place.

MAFF, 1983

2. To establish the effect of, and interation between, retention time, aspect ratio, angle of inclination and temperature on digester performance.

3. To determine the effect of scale up on material flow.

4. To appraise technical and economic viability of the design for farm scale application.

CHAPTER 2

DESIGN OF PILOT SCALE DIGESTERS

2.1 Theoretical Considerations

The concept of a "plug flow" digester is by no means a new one. Indeed many digesters of this type have been built and are operational around the world. An idealized plug-flow reactor has been defined (Jewell *et al*, 1980) as follows:

1. There is no intermixing or interaction between particles.

2. Each increment of fluid passes through as an independent cross section or plug.

3. The velocity of fluid throughout the reactor vessel, and at all points in any cross section of the stream, is constant.

4. Each fluid particle is retained in the reaction vessel for a time equal to the theoretical retention time.

5. Microbial activity, substrate concentration, temperature, or any other physical, chemical, or biological property must be constant for all points in any cross section of the stream.

If the above criteria were met, it would be impossible for digestion to take place since the incoming feed would not be innoculated from the bacterial population already in the reactor. In practice, ideal plug flow does not occur due to back-mixing, settling, scum formation or short-circuiting. The type of design under consideration is more accurately described as a displacement type digester.

A major problem with horizontal displacement digesters is the formation of scum and crusts at the gas-liquid interface, particularly where the Total Solids content of the feedstock is low (<6%) as is commonly the case with cattle and pig slurries. Scum (floating material) and crust (dried feedstock) accumulate in most reactor designs and reduce the efficiency of the digestion process by reducing the effective working volume of the reactor. The retention time is reduced and bacterial washout will eventually occur. Digesters are designed so that scum and crust formation is minimised, but the complexity of the design is increased.

To overcome these problems, the inclined tubular digester investigated in this study was proposed in 1979 by Hawkes and Horton from the Polytechnic of Wales, and a laboratory-scale study was carried out by Floyd (1984). The digester is a modified form of a horizontal displacement digester, but the digestion vessel is tubular, and inclined at an acute angle to the horizontal, as shown in Figure 2.1. Thus, the main advantages of a horizontal displacement type digester are retained, and more importantly the exposed surface area of the digester contents where scum and crusts can form is minimised. The inclined tubular digesters used in the present study are shown in Plate 1, and Plate 2 shows the completely mixed reactors (2 * 125m^3) which were on the same site. The relationship between the exposed surface area and the angle of inclination for a tubular digester is shown in Figure 2.2. By minimising the area in which any scum and crust can form, it is mechanically simpler to remove.

A further advantage with the inclined tubular design arises because sedimentation is several times faster in an inclined tube, than in a vertical tube of the same height (Hill <u>et al</u>, 1977). This can be explained by recognising that in any tube, the particles in a liquid settle vertically until they meet an upward facing surface. Obviously in an inclined tube, the maximum

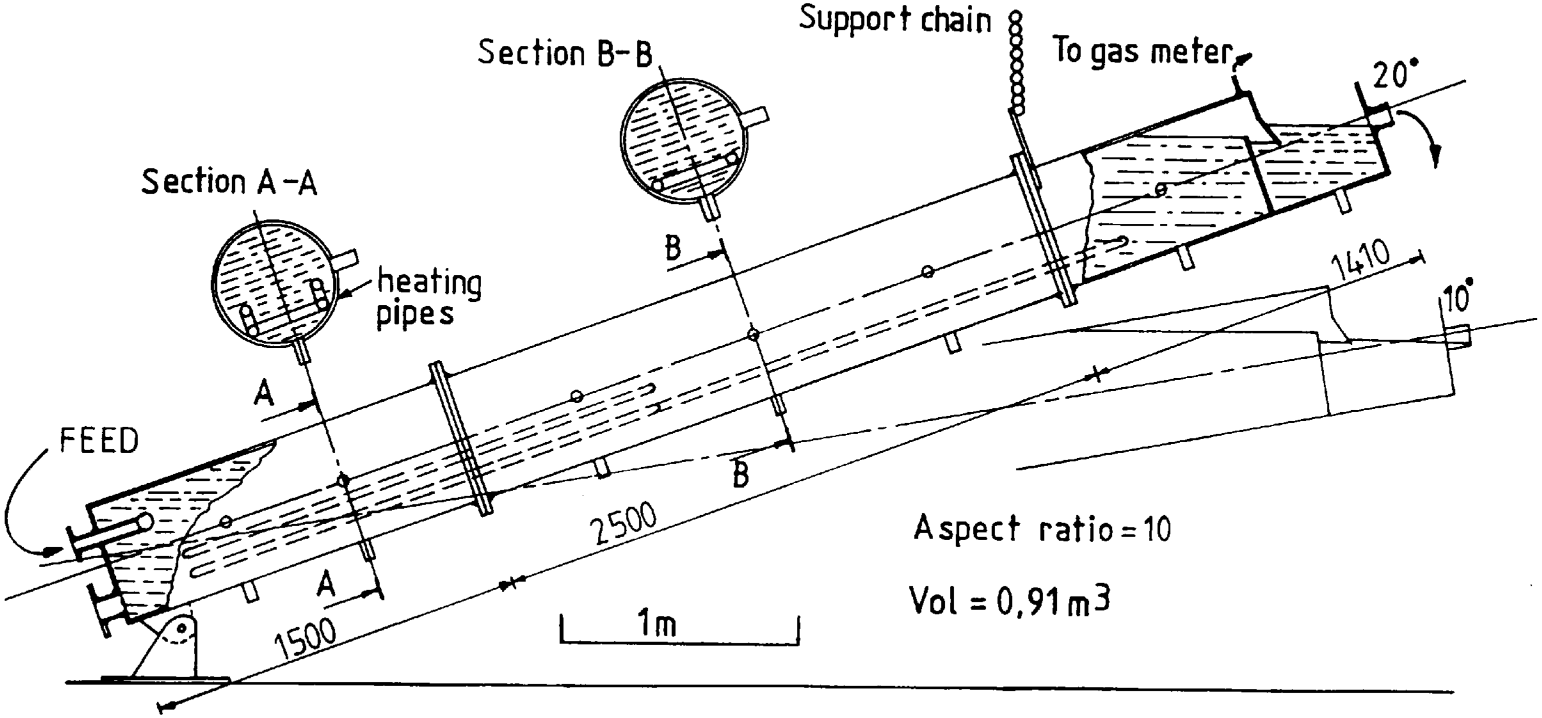


Figure 2.1 INCLINED TUBULAR DIGESTER

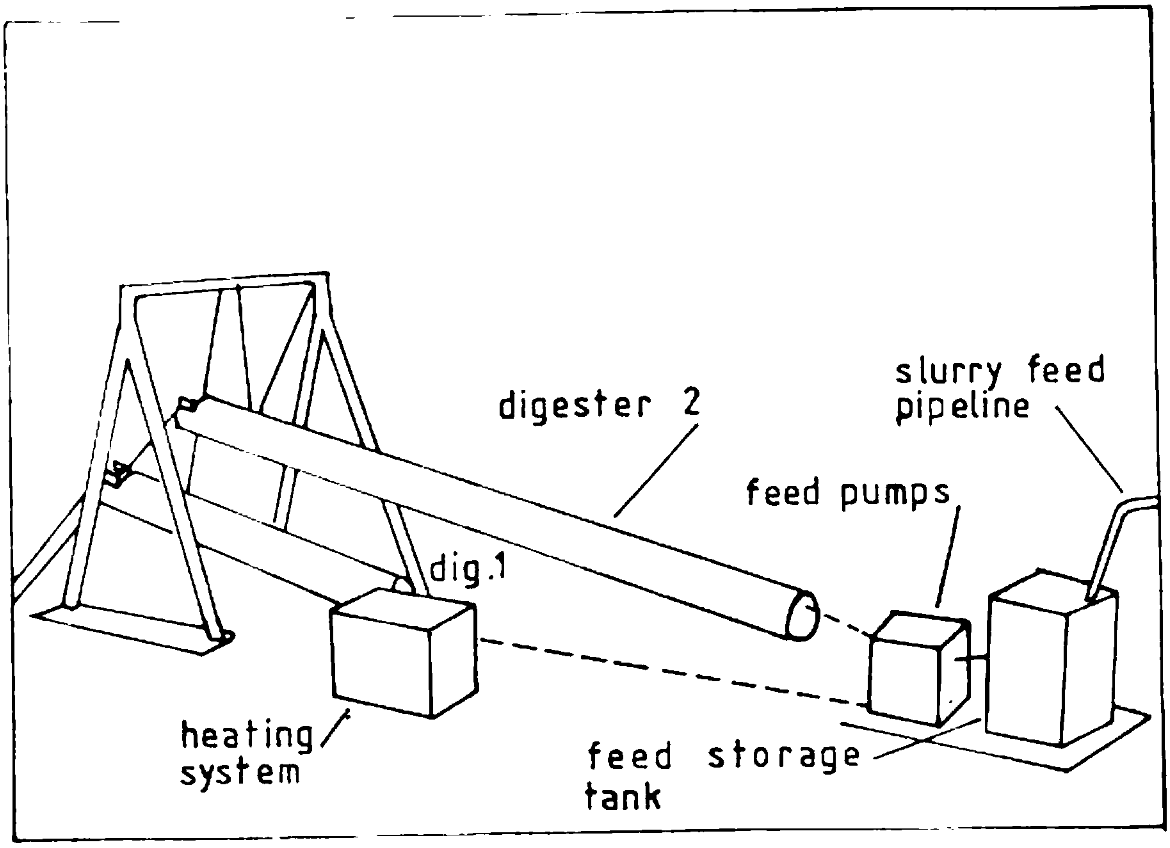


PLATE 1–KEY

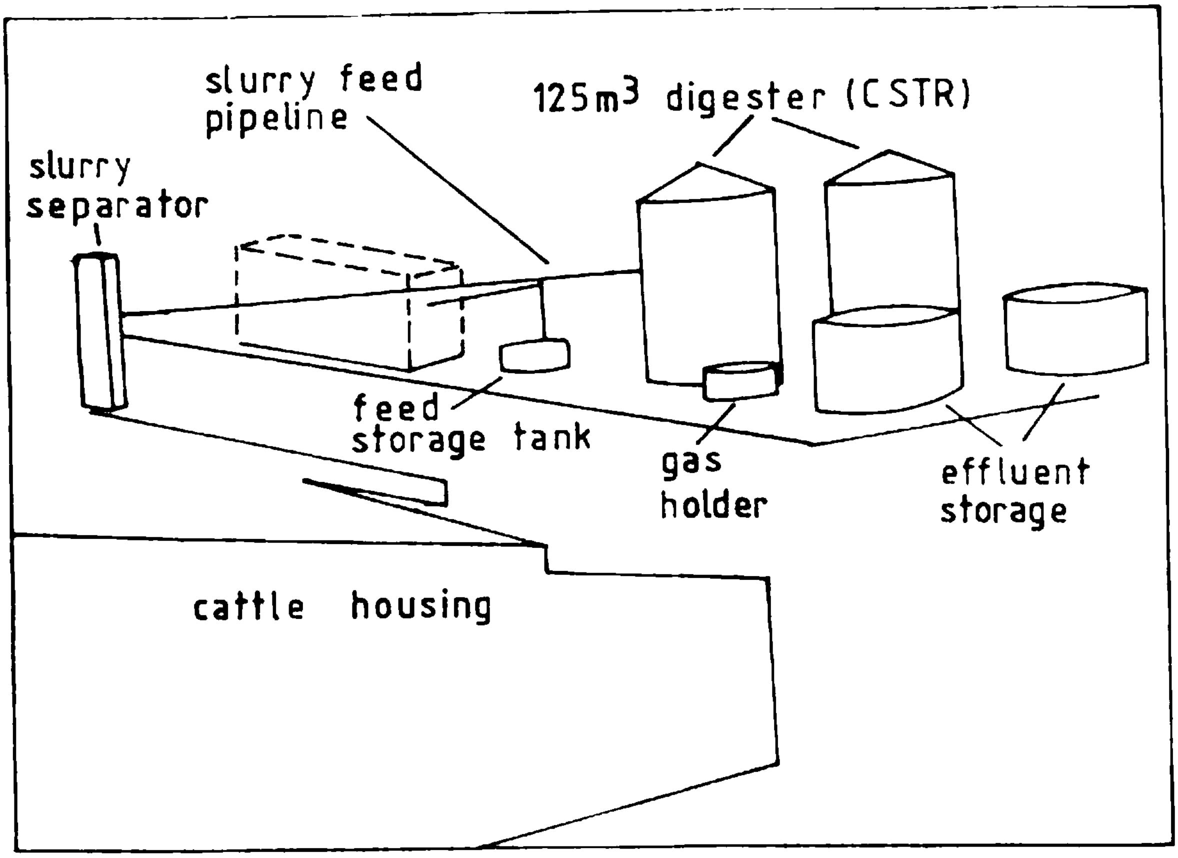
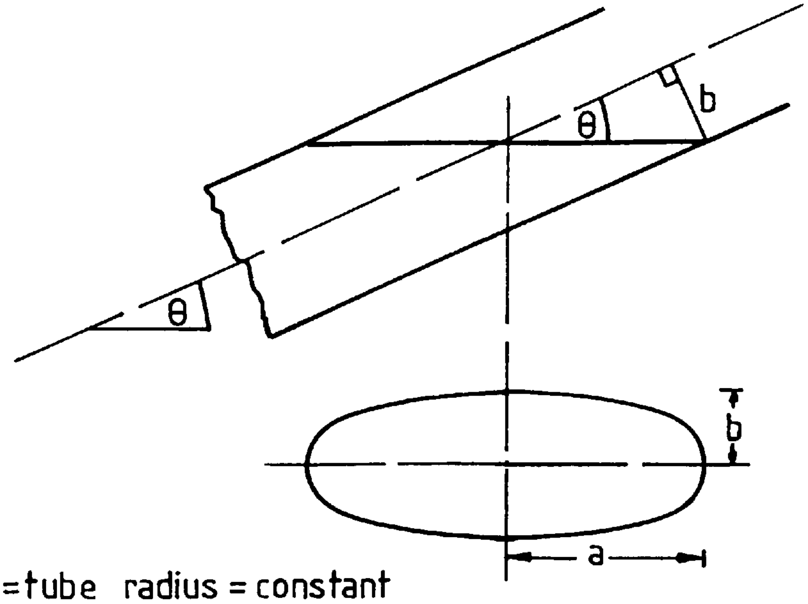


PLATE 2–KEY

PLATE 1. Pilot scale inclined tubular digesters.

PLATE 2. Farm and pilot scale anaerobic digestion research facility at AGRI.

FIGURE 2.2 Relationship between the surface area of liquid in an inclined pipe and the angle of inclination.

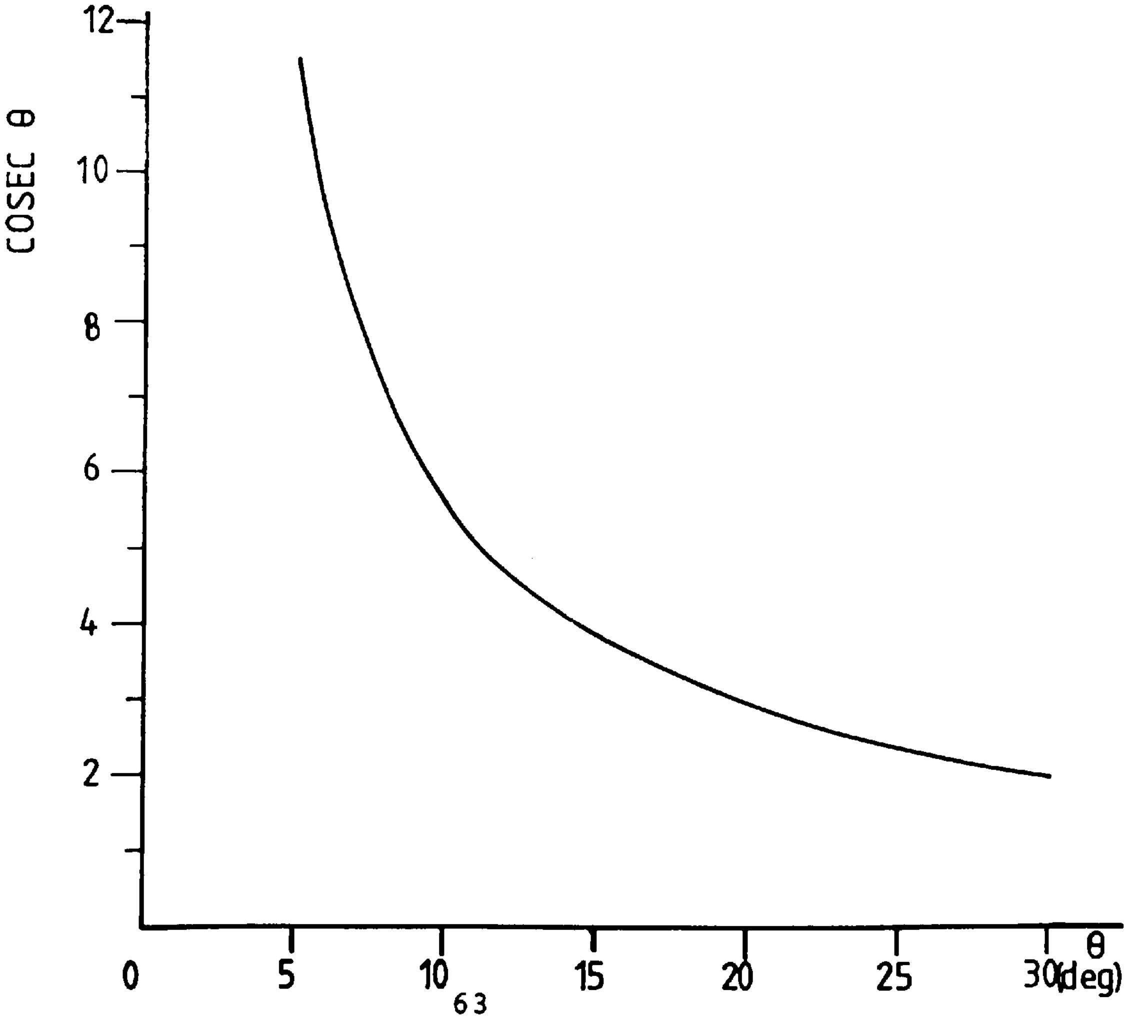


b = tube radius = constant

Area of an ellipse = πab.

$a = \frac{b}{\sin\theta} = b \operatorname{cosec}\theta$

$\therefore$ Surface area $= \pi b^2 \operatorname{cosec}\theta$



distance through which a particle can settle is very much less than in a vertical tube of the same length. Consequently more particles will settle from an inclined tube, compared with a vertical tube. This is advantageous to the digestion process since solid particles are less degradable than soluble or small particulate material, and should ideally stay in a reactor for longer than the liquid stream.

Gas bubbles produced during the anaerobic digestion process rise vertically, and become concentrated in the region bounded by the concave wall forming the upper part of the tube. These gas bubbles may collect or combine with other gas bubbles as they rise vertically, and also as they travel along the tube towards the liquid / gas interface. Solid particles attached to gas bubbles may also be transported in this way, or, become detached from the gas bubble and settle within the reactor.

The process of sedimentation in inclined tubes is further speeded up by convection currents (Kinosita, 1949). As particles settle, convection currents cause the space they occupied to be filled, and sedimentation occurs at an increased rate due to the decreased particle concentration.

Whilst increased rates of sedimentation are advantageous for degradable material, the reverse is true for non-degradable components, particularly grit which is found in many animal slurries. Consideration needs to be given to how grit will be removed in any digester design. With the inclined tubular design, deposition of grit is greatest at the lowermost end of the digester, from where it can be easily removed by a suitably sized valve, or other mechanical system. The ability to remove grit from the digester easily was important, since accumulation reduces effective working volume.

2.2 Digestion vessel

The digestion vessel is heated (Sec. 2.2.3) internally, but unlike other high-rate digesters, it was not mixed. This represented a simplification from the viewpoint of design, and an attraction from capital and operating cost considerations.

In earlier work by Floyd (1984), three laboratory scale digesters with volumes of 13-15.3l were used. These were constructed from perspex tube and sheet.

Much greater working volumes were obviously needed to study this digester design at a scale intermediate between laboratory and farm-scale units. Initially there was going to be one digester, at the approximate size of a small farm-scale unit (20-30m^3). However, meaningful comparisons between work at this scale and the laboratory scale would have been difficult. It was decided instead to construct two digesters, with a working volume of 1-2m^3. At this scale they were representative of a farm-scale unit, yet not so far removed from the previous laboratory work to make comparisons meaningless. This represented an increase in volume of 100-150 over the laboratory digesters of Floyd (1984). A scale-up factor of this order is commonly used in the chemical engineering industry for first scale up of a process.

2.2.1 Aspect ratio

In earlier work by Floyd (1984), laboratory scale inclined tubular digesters with aspect ratios (length/diameter) of 15 and 6 were used. Selection of optimum aspect ratio was important because it dictated the shape and cost of the digestion vessel. Aspect ratios of 10 and 20 were selected for the present study, because they were considerably different from the value of 15 which Floyd (1984) used for two of his three digesters.

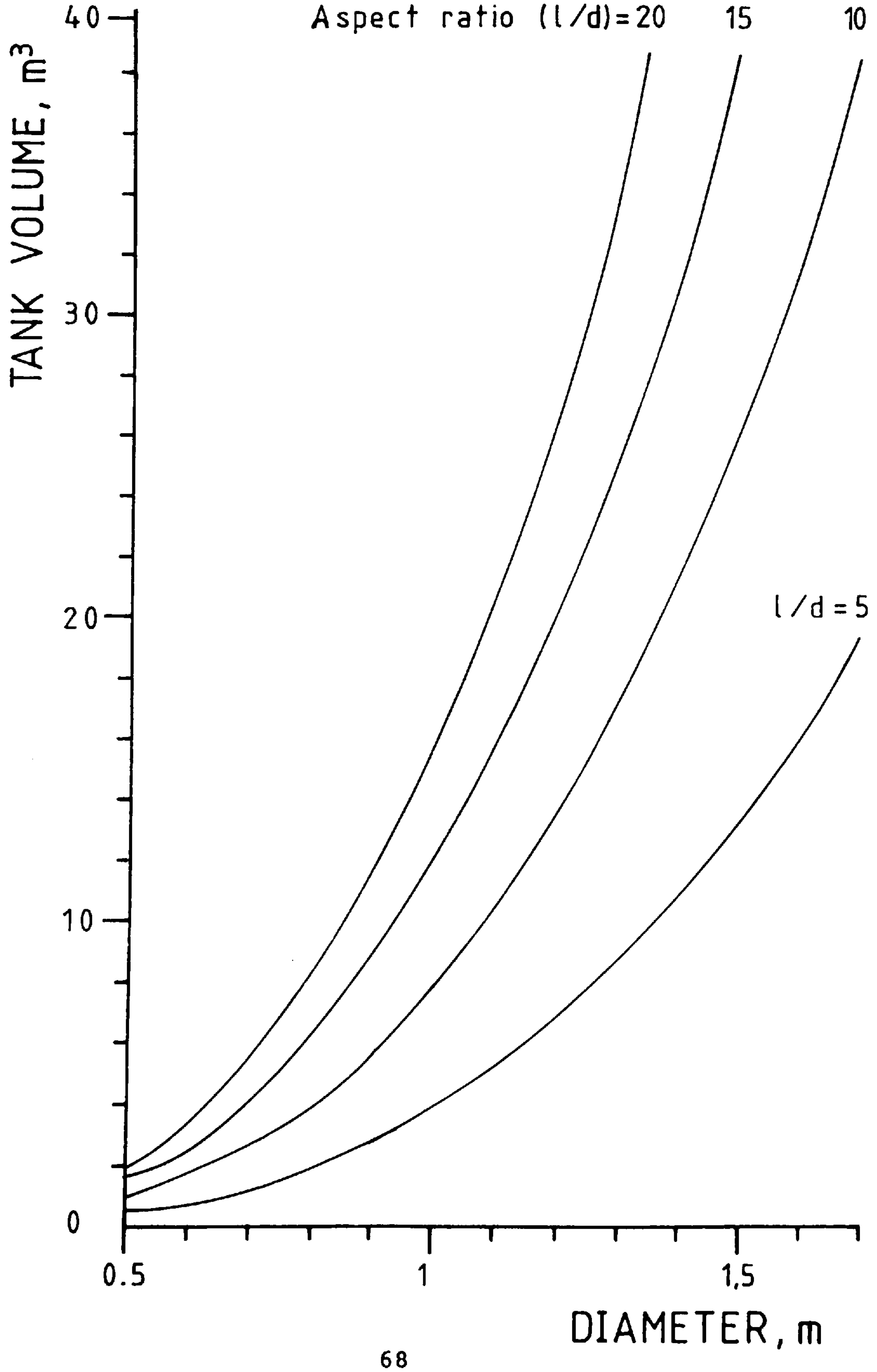
There are two alternative means of designing digesters to compare different aspect ratios. The digesters could either be of the same length and different diameters, or vice-versa. The greatest flexibility in terms of possible configurations was provided by a modular construction, using tube of constant diameter in flanged lengths, so this was the option chosen. This overall design, including support-frame, was simpler than a design using different diameters. A dis-advantage with this arrangement was that it did not allow the effect of diameter to be studied.

The way in which the volume of a cylindrical tank varies with diameter for a constant aspect ratio is shown in Figure 2.3. Having decided on the working volume and aspect ratios required, the actual pipe diameter was determined to be approximately 0.5m. The actual diameter used was 0.488m (internal), this being a readily available size.

2.2.2 Materials of construction

For a below-ground digester construction of fixed volume and angle of inclination, materials which could be used for the reactor included concrete, steel, plastic, GRP, rubber and PVC. However, the digesters were built above-ground due to their being experimental, and requirements for ease of sampling, maintenance, repair

FIGURE 2.3 The volume of a cylindrical tank as a function of diameter for constant aspect ratio.



and dismantling.

Having decided that the digesters were to be above-ground, and with variable aspect ratio and angle of inclination, the choice of materials for the reactor was limited. Although any of the above materials could have been used, the cost and degree of complexity associated with each material varied widely. Steel (ASTM A106 Grade B) was selected as the main material of construction due to its ready availability, and the ease with which pressure-tight joints and modifications can be made.

2.2.3 Heating system

The target operating temperature of the digesters was in the mesophilic range of 20-40°C. During each experimental run, the temperature was controlled at a fixed point within this range. The two main components of a digester's heating requirement are the heat needed to raise the temperature of the feed slurry to that of the digester, and the heat needed to replace that which is lost through the digester walls.

2.2.3.1 Influent heating

The heat required, Q, is given by the equation:-

$$Q = M\ C_p\ \Delta T$$

where M = Mass flowrate (kg/day)

C_p = Specific heat capacity of slurry (J/kg °C)

ΔT = Difference between ambient and digester temperature

Maximum digester volume = $2m^3$

Shortest likely retention time = 10 days

$$\text{Maximum flowrate} = \frac{2}{10}\ \frac{m^3}{day} = 0.2m^3/day$$

The density of slurry is taken as $1000kg/m^3$

Specific heat capacity assumed to be that of water, (4200J/kg°C)

Then $Q = 0.2 \times 1000 \times 4200 \times \Delta T$

It can be assumed that the digester feed temperature will closely approximate ambient temperature, since the feed slurry was stored in an above ground PVC tank for a period of at least 12 hours. For design purposes, a worst case of 0°C ambient was used.

For a digester operating temperature of 35°C, $\Delta T = 35-0$

Therefore $Q = 0.2 \times 1000 \times 4200 \times 35$

$= 29.4$ MJ/day (worst case)

$$29.4 \text{ MJ} = \frac{29.4 \times 10^6}{24 \times 3600} \text{ J/s} = 340 \text{ W continuous}$$

2.2.3.2 Heat losses

Anaerobic digesters are insulated to minimise the heat loss by conduction through the digester walls. An "economically optimum" thickness of insulation can be calculated from a consideration of insulation material costs, the heat loss, and the operating cost of the digester heating.

The heat loss is calculated from the equation,

$$Q_w = U_w A_w \Delta T$$

where Q_w = Heat loss through walls (W)

U_w = Heat transfer coefficient for digester walls ($W/m^2\ {}^oC$)

A_w = Area of walls below liquid level (m^2)

ΔT = Temperature gradient across digester walls (${}^oC$)

The greatest difficulty was in calculating U_w. In an insulated digester, each layer resists the heat transfer. The total thermal resistance was the sum of the respective resistances to heat transfer produced by the "water" film inside the digester, the digester walls, the insulation, and the air film outside the digester.

The heat transfer coefficient for the digester walls is the reciprocal of the sum of the individual resistances to heat transfer.

$$\frac{1}{U_w} = \frac{x_1}{k_1} + \frac{x_2}{k_2} + \frac{1}{h_1} + \frac{1}{h_2} + \frac{1}{h_3}$$

where x_n = thickness of a particular layer

k_n = Thermal conductivity (W/m^oC) for a particular layer

h_1 = Film coefficient of air outside the digester ($W/m^{2o}C$)

h_2 = Film coefficient of gas inside the digester ($W/m^{2o}C$)

h_3 = Film coefficient of sludge to wall inside digester ($W/m^{2o}C$)

h_4 = Coefficient of conductance of gas in the space above the liquid ($W/m^{2o}C$)

For a digester of 0.5m diameter, 10m long and made from 11mm steel, externally insulated (50mm thick), ignoring heat losses through the tube ends, then:-

$$\frac{1}{U_w} = \frac{0.011}{54} + \frac{0.050}{0.043} + \frac{1}{28} + \frac{1}{150}$$

$$\frac{1}{U_w} = 1.2053$$

Film coefficients and k_{wall}
-Stafford et al, 1980

$K_{insulation}$
- Rockwool Ltd., 1981

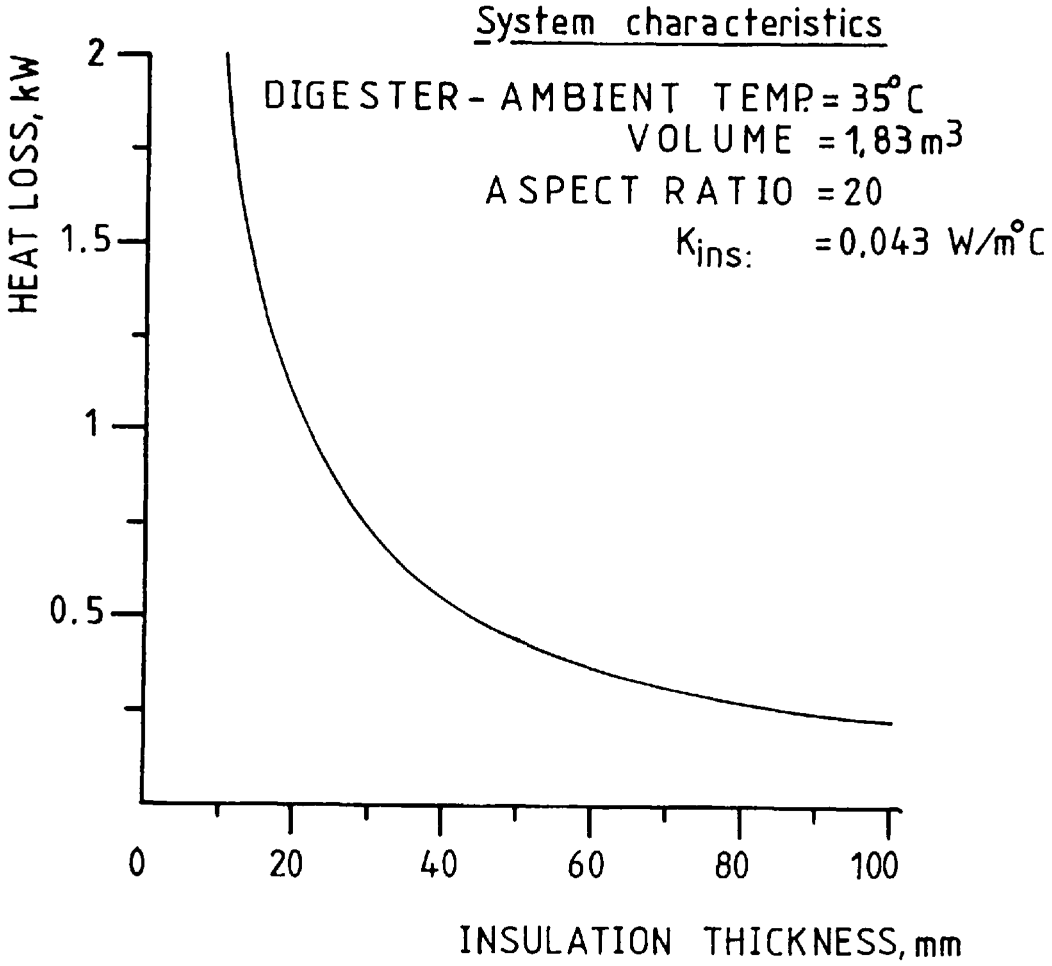
$$U_w = 0.8296 \text{ W/m}^{2\,o}\text{C}$$

$$Q_w = 0.8296 \times 15.70 \times (35-0) = 455.86\text{W}$$

(Worst case, assuming digester temperature = 35°C,
ambient = 0°C.)

The major resistance to heat flow is from the insulating material. Figure 2.4 shows how the heat loss (Q_w) varies with insulation thickness. The major resistance to heat flow is from the insulating material. The digesters were insulated with 50mm Lamella Mat (Rockwool Ltd.), and clad with 22 s.w.g. aluminium. Removable boxes of the same materials were fitted around

FIGURE 2.4 The effect of insulation thickness on heat loss.



the flanges joining adjacent digester sections. The insulation has a thermal conductivity "k" value of 0.043 W / m^{o}C at 50^{o}C (Rockwool Ltd., 1981).

2.2.3.3 Heat exchanger design

Having evaluated the estimated heating requirement for the digester, the heat exchanger was sized.

It is generally agreed (Hawkes, 1985; Persson et al, 1979) that the surface temperature of a digester heat exchanger should be kept at or below 50^{o}C. At greater temperatures, the surface will become dried out resulting in a loss of efficiency.

The obvious place for the heat exchanger was at the lower end of the digester, where fresh feed at ambient temperature was introduced. Because of the convective currents within the digester, it was thought unnecessary for the heat exchange surface to be evenly distributed throughout the digester. It was decided to use heating coils situated at the lowermost digester end, arranged in a multipass configuration.

Since the digesters were only fed twice daily, the flow of slurry past the heat exchanger was low, and the system was neither a true parallel flow or counter-flow arrangement. However, the system most closely approximates to parallel flow.

The governing equation for the design of a heat exchanger is:

$$Q = UA\,\Delta T_m \qquad (1)$$

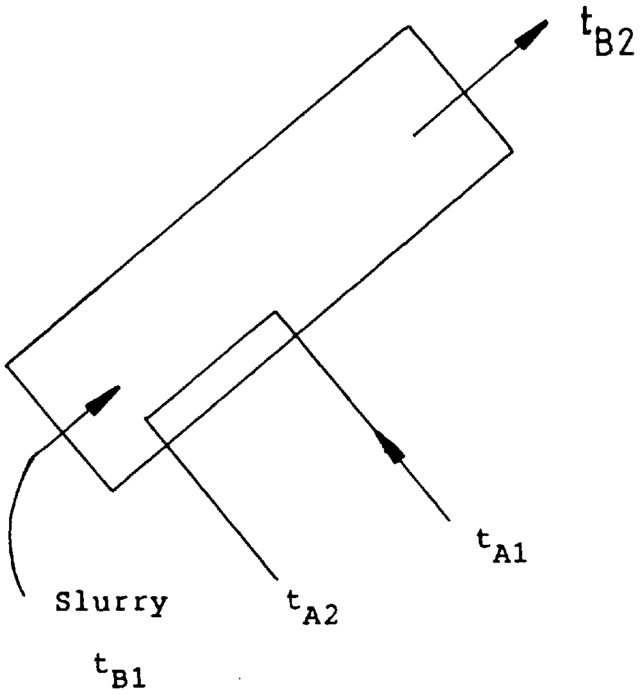
where Q = Heat added to system

U = Overall heat transfer coefficient ($W/m^2\,^oC$)

A = Surface area of heat exchanger

ΔT_m = Logarithmic mean temperature difference (^oC)

The system is shown diagramatically below:



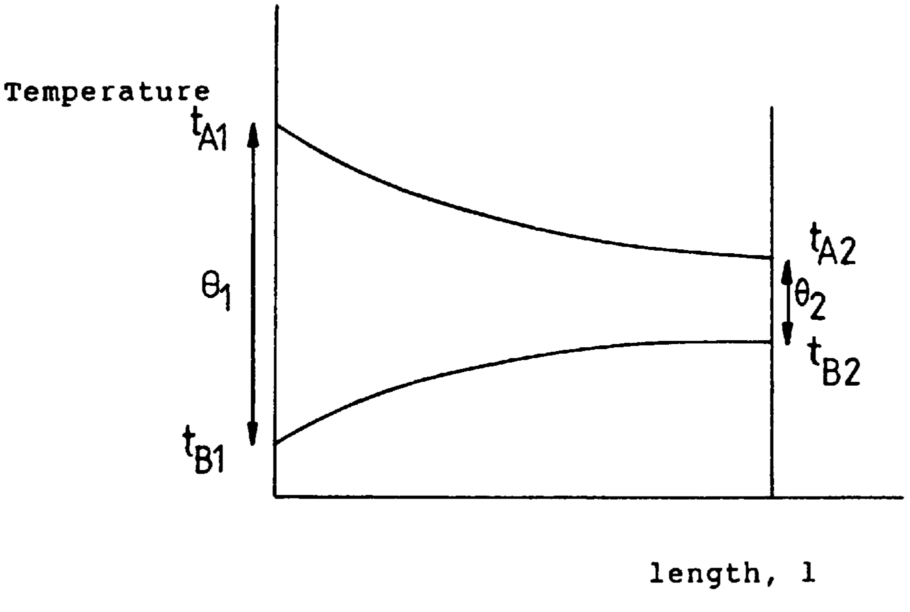
The inlet and outlet temperatures were fixed at 50°C and 47°C, respectively.

$$\Delta t_m = \frac{\theta_1 - \theta_2}{\log_e \theta_1/\theta_2}$$

where $\theta_1 = t_{A1} - t_{B1} = 50 - 35 = 15$

$\theta_2 = t_{A2} - t_{B2} = 47 - 35 = 12$

$$\Delta T_m = \frac{15 - 12}{\log_e 15/12} = 13.44$$



Parallel flow arrangement

It was difficult to determine the overall heat transfer coefficient accurately, since the film coefficients were dependent on factors such as the physical nature of the sludge, and flow conditions both inside and outside the heat exchanger. In operation, scales and films of dirt build up on heat exchange surfaces and reduce the efficiency of heat transfer. Allowance for such deposits can be made if the fouling factors are known. They are best determined experimentally. Fouling factors were not included in the calculation for the overall heat transfer coefficient, since no values for pipes in similar heat exchange systems could be found.

The overall heat transfer coefficient (OHTC) is found from the equation $\frac{1}{U} = \frac{1}{h_A} + \frac{x}{k} + \frac{1}{h_B}$

where h_A = Film coefficient of water to wall of tube ($W/m^{2\,o}C$)

h_B = Film coefficient of sludge to wall of tube ($W/m^{2\,o}C$)

x = tube thickness (m)

k = Thermal conductivity of tube ($W/m^{o}C$)

The resistance of the tube is usually small, and is often ignored.

Because of the difficulty in estimating the film coefficient for the sludge to the wall (Orth, 1981), an overall heat transfer coefficient was used. Values of the OHTC given by Perry (1963) for a copper coil, circulating hot water, immersed in slurry, range from 250-1200W $W/m^2\ ^oC$. For design purposes, a conservative value of $150W/m^2\ ^oC$ was used.

Referring back to equation (1),

$$Q = UA\ \Delta T_m$$

The terms U and ΔT_m were evaluated.

The heat required by the system for influent heating was evaluated as 29.4 MJ/day. Theoretically, that design would have taken 24 hours to raise the influent to the desired operating temperature.

Assuming that the influent must be heated to the target operating temperature in four hours, then,

$$\text{Rate of heating} = \frac{29.4}{4}\ \text{MJ} = 7.30\ \text{MJ/hr (for 4 hrs/day)}$$

$$\text{Actual rate} = \frac{7.35 \times 10^6}{60 \times 60} = 2041\text{W (for 4 hrs/day)}$$

$$\text{Therefore } A = \frac{Q}{U\Delta T_m}$$

$$A = \frac{2041}{150 \times 13.44} = 1.01 m^2$$

For a circular pipe, surface area $A = \pi dl$

Therefore length required, $l = \frac{1.01}{\pi d}$

For a nominal bore of 25mm,

$$l = \frac{1.01}{\pi \times 0.025} = 12.85 m$$

The calculations have so far assumed the heater to be 100% efficient. A typical efficiency of 70%, leads to a pipe length of $\frac{12.85}{0.7} = 18.35 m$

Similarly, to replace the heat lost through the digester walls:

$$Q_w = 455 W$$

$$Q = UA \Delta Tm$$

$$A = \frac{495}{150 \times 13.44} = 0.245 m^2$$

For a pipe of nominal bore 25mm, length required,

$$l = \frac{0.245}{\pi \times 0.025} = 3.12 m$$

Assuming an efficiency of 70%, $l = \frac{3.12}{0.7} = 4.46m$

In digester 1 (vol = 0.91 m^3), the heat exchanger consisted of a main section 8m long, concentrated at the lower end of the digester, and a secondary loop of 6m distributed throughout the whole digester.

In digester 2 (vol = 1.83 m^3), the main loop was 15m long, and the secondary loop was 15m long, distributed throughout the whole digester.

The loops were made from 25mm nominal bore (outside diameter 33mm) mild steel pipe, joined with malleable iron fittings (screwed).

The sizing of the heat exchangers was only very approximate, and in the situation it was prudent to oversize both from an efficiency point of view and also because the form of the temperature distribution throughout the digesters was not known.

2.2.4 Angle of inclination

The experimental programme included investigating the effect of angle of inclination on digester performance. Earlier work by Floyd (1984) and Himmel (1984), inclined the digestion vessel at 15-20^o to the horizontal. At these inclinations, the design approximates closely to the horizontal displacement type, and the digester is intended to be a development of that. Steeper angles of inclination (>30°) could have been used, but the engineering work, and ultimately cost, would have been greater, particularly in full-scale digesters of this design. Advantages of shallow (<30°) angles of inclination include distribution of digester weight over a larger area, and aesthetic considerations whereby the digester can be partly or completely buried, if required.

In the present study it was decided that angle of inclination should be variable between 0 and 20° to the horizontal. If steeper angles had been used then the design would not be a close approximation to the horizontal displacement digester, rather it would simply be an un-mixed tank of vertical orientation.

2.2.5 Feeding and discharge system

The two main factors affecting choice of a pump are flow rate, and the nature of the feed. Feeding rate varies between 0.1-0.2m^3 slurry/day, and the feedstock was fibrous in nature, requiring pipework of at least 50mm diameter to avoid frequent blockages. The above flow rates are low for pumps which are able to satisfactorily handle the size and nature of the feedstock. Whilst there was a wide range of pumps available, those which could have met these design criteria were primarily developed for specialist applications in the chemical engineering industries, and as such were prohibitively expensive for the present project.

Digester feeding was carried out twice per day. This meant that a far wider range of standard pumps could be considered for digester feeding. Positive displacement pumps were identified as being the most suitable for this application, primarily on account of their reliability, and repeatability enabling the standard pump to be used as a metering pump, in conjunction with a timer. Additionally, these pumps acted as "non-return" valves when not pumping, and eliminated the need for any such valves, although other valves were obviously necessary to isolate the pump for maintenance purposes.

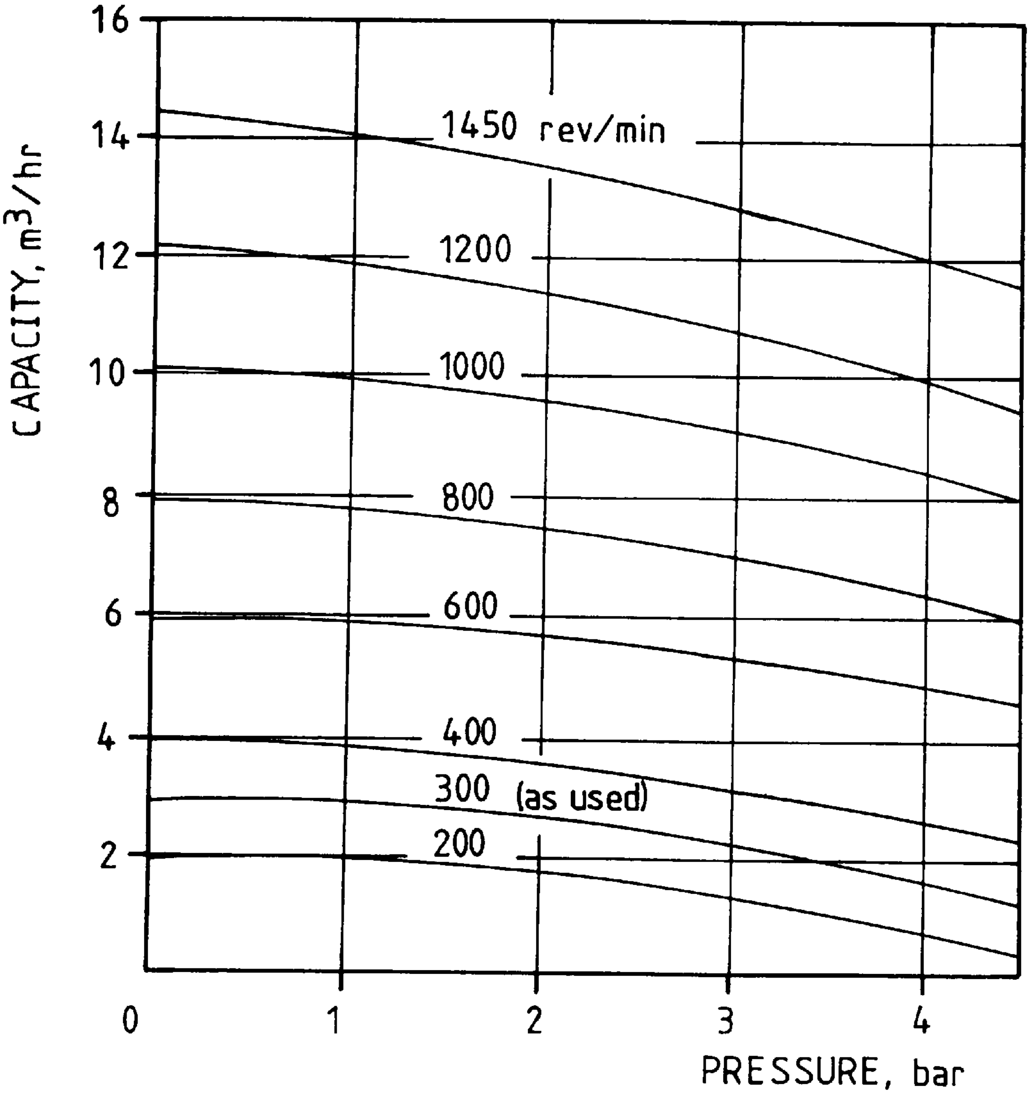
Both digesters were pump fed from a common $1m^3$ storage tank, fitted with a mechanical mixer operating at 16 rpm. The use of one feed storage tank was preferred since both digesters were operated on the same feedstock. Any variations between different batches of feed were, therefore, the same for each digester.

A 50mm scroll and flexible stator pump (Mono Pumps Ltd.) was used to feed each digester. The pumps were sized to minimise the likelihood of any blockages occurring due to the fibrous material in the feedstock which had a maximum 10% Total Solids (TS) content. Additionally, the pump speed was reduced by a factor of 3 to 300 rpm because of the viscous and abrasive nature of the feedstock, using a vee belt drive arrangement. Since capacity is approximately proportional to the speed of the pump, this meant that the pump could be used as a metering pump with more accuracy, compared with the standard unit. Figure 2.5 shows how pump capacity varies with speed.

The digesters were fed twice daily (1000, 2200 hrs). The pumps were calibrated to deliver the required amount of slurry and a timer (Omron Ltd.) was used to repeat this action automatically. The actual amount fed by the pump was checked monthly by measuring the volume of slurry discharged during a feed cycle and the pump being re-calibrated if necessary.

The feed was introduced to the digesters through a 50mm pipe on the bottom end plate of each digester. A twin elbow fitting (malleable iron) attached to the end of this pipe minimised any tendency of the feed slurry to short-circuit the digester. After passing through the twin elbow, the feed impinged on the side walls of the

FIGURE 2.5 Relationship between speed and capacity for MONO MD60 pump.



CURVE NO:
1569-m

MONO PUMPS LTD., 1982

digester, and caused a limited amount of mixing at the base of the digester during the time that the pumps were on.

Digested slurry was discharged from the digester at the uppermost end, over a weir, as shown in Figure 2.6. Slurry flowed over the weir when the feed pumps were operated. This arrangement, which was designed as a flooded weir, proved very simple and reliable, and it had the advantage of not requiring a separate pump for discharge. The liquid inside the digester was maintained at a constant level by the weir which was the only interface between the digester contents and the atmosphere. The weir design had to maintain anaerobic conditions in the digester at all times. The chosen arrangement did this, even when a large volume of slurry was removed from the digester, perhaps for sampling.

The weirs on each digester were modified during the shutdown in February 1985. The new arrangement is shown in Figure 2.7. A viewing port was made in the top end plate of each digester at this time. These modifications were made because slurry was crusting in the effluent boxes and forming a fibrous mat, making it difficult for the digested slurry to flow out of the digester. On several occasions the effluent box had overflowed, and at other times the gas outlet pipeline had become blocked

FIGURE 2.6 Digester effluent weir (original).

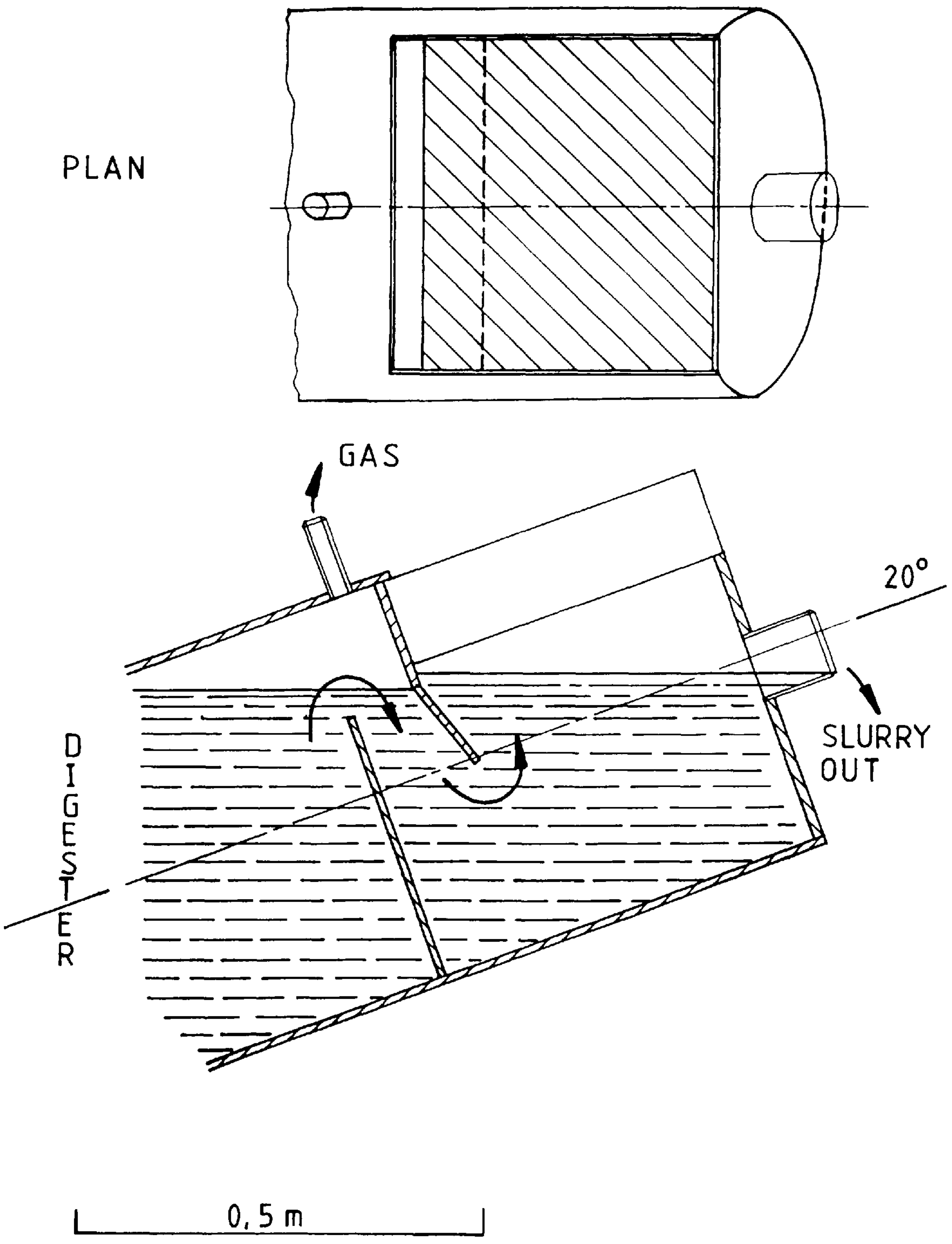
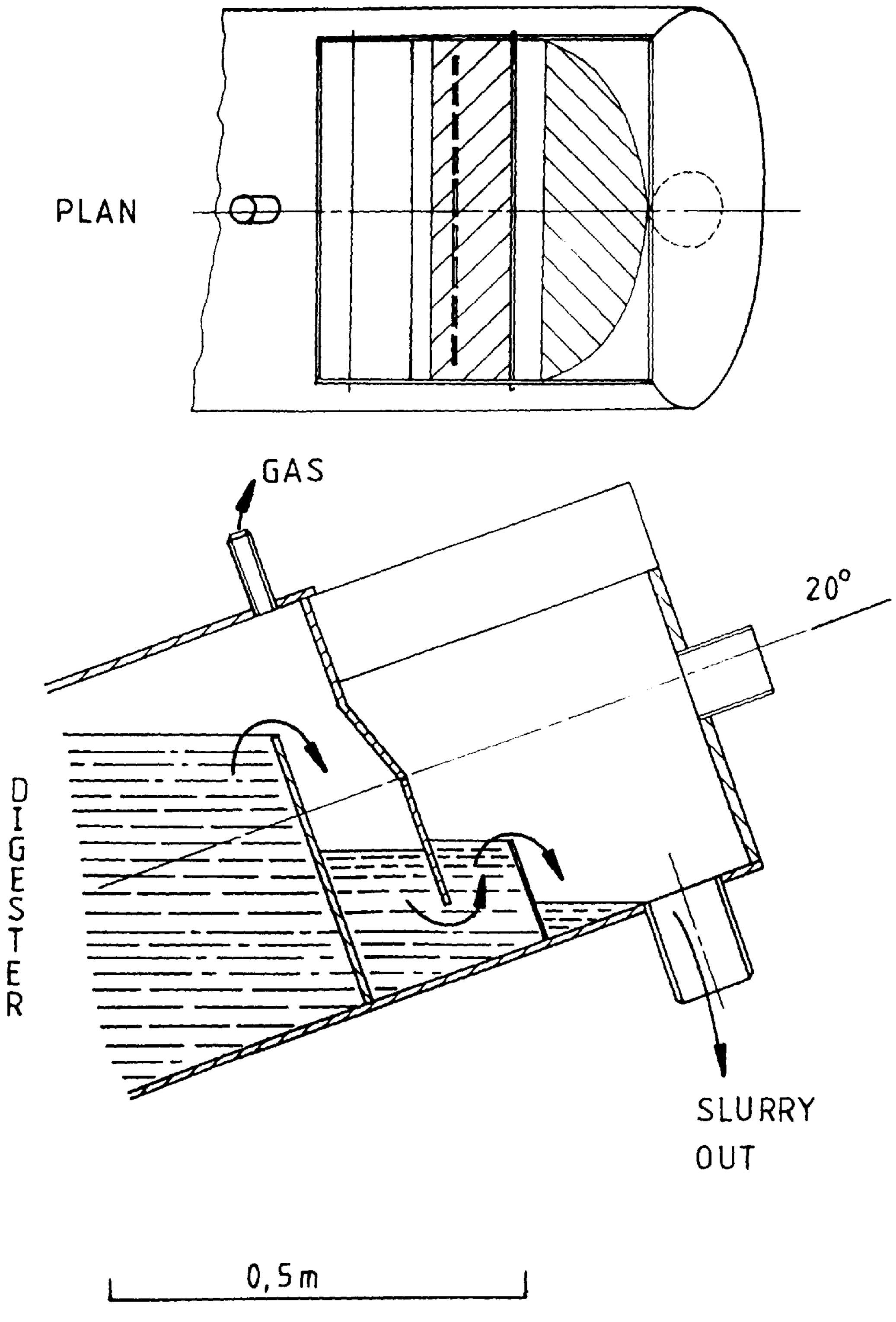


FIGURE 2.7 Digester effluent weir (modified).



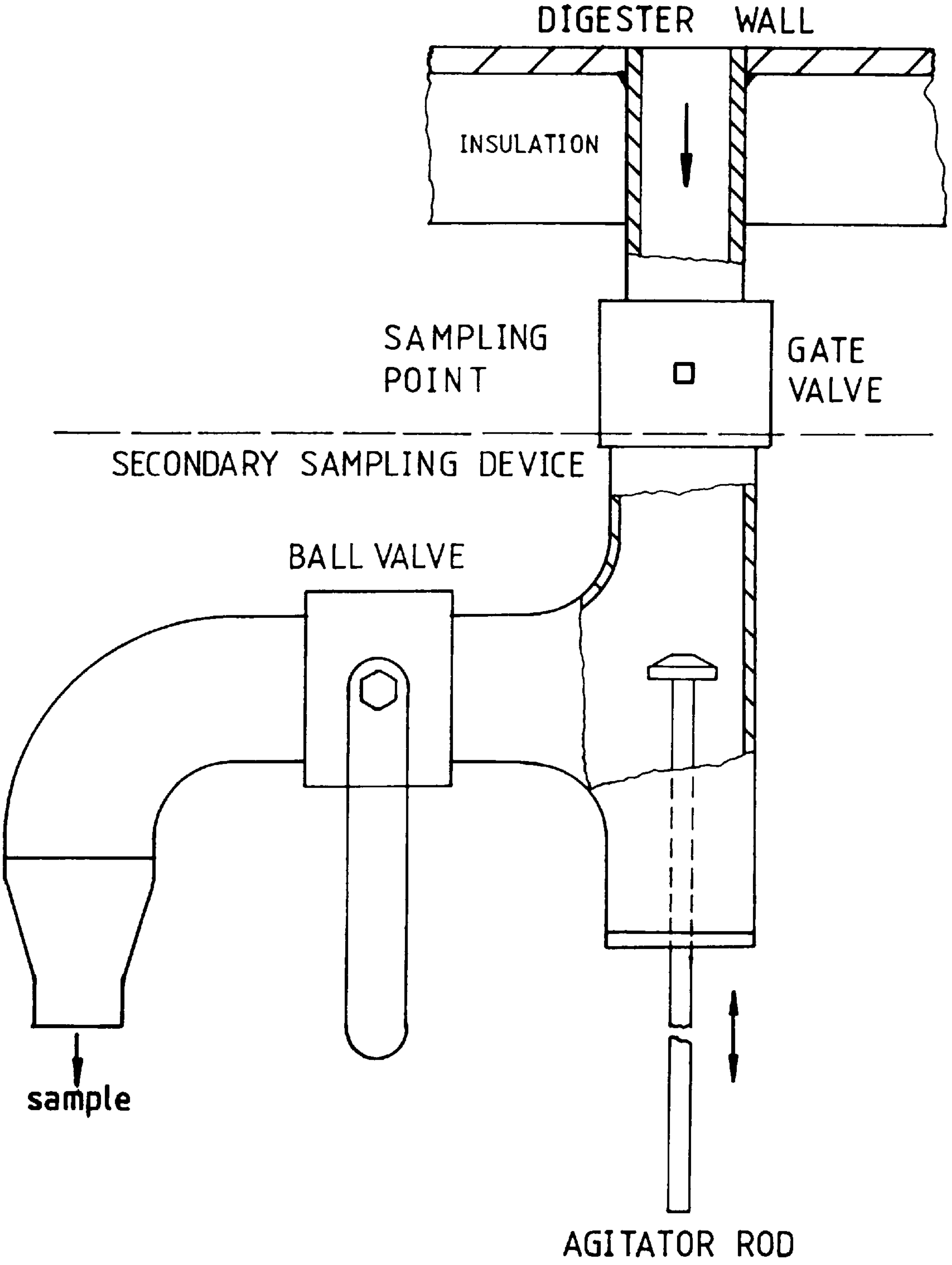
with discharged slurry. The modified weir arrangement increased the headspace in the digester, and the volume of slurry in the effluent boxes was considerably reduced.

2.2.6 Sampling points

Sampling points were fitted to each digester to obtain samples of digesting slurry for routine analysis and tracer studies (see Section 3). These sampling points were on the base and one side of each digester, spaced 1m apart, except for the lowest two valves on each digester which were 0.5m apart. The type of valve used for each sampling point was a 40mm gate valve, chosen on account of its cheapness compared with other valve types of a similar size.

A major problem with the gate valve in these circumstances is that the valve cannot be fully opened and closed quickly (compared with a butterfly or ball valve). As a result it is difficult to take small, representative samples. To overcome this problem a secondary sampling device employing a ball valve was constructed, and is shown in Figure 2.8. This device was used in conjunction with the original gate valves, and enabled a more representative sample to be taken, and eliminated any restriction to material-flow caused by a partly opened valve.

FIGURE 2.8 Digester secondary sampling device.



2.3 Instrumentation

The digesters are instrumented to monitor automatically some aspects of plant performance including gas production, temperature and the electrical energy used for heating.

2.3.1 Gas metering

The amount of gas produced by a digester installation such as this was typically 1 reactor volume per day. With the digester volumes of 1-2m^3 used, gas flows were expected to be in the range of 40-80 1/hour. A typical composition of biogas would be 60% methane, 40% carbon dioxide and hydrogen sulphide at trace concentrations of up to 1000 ppm.

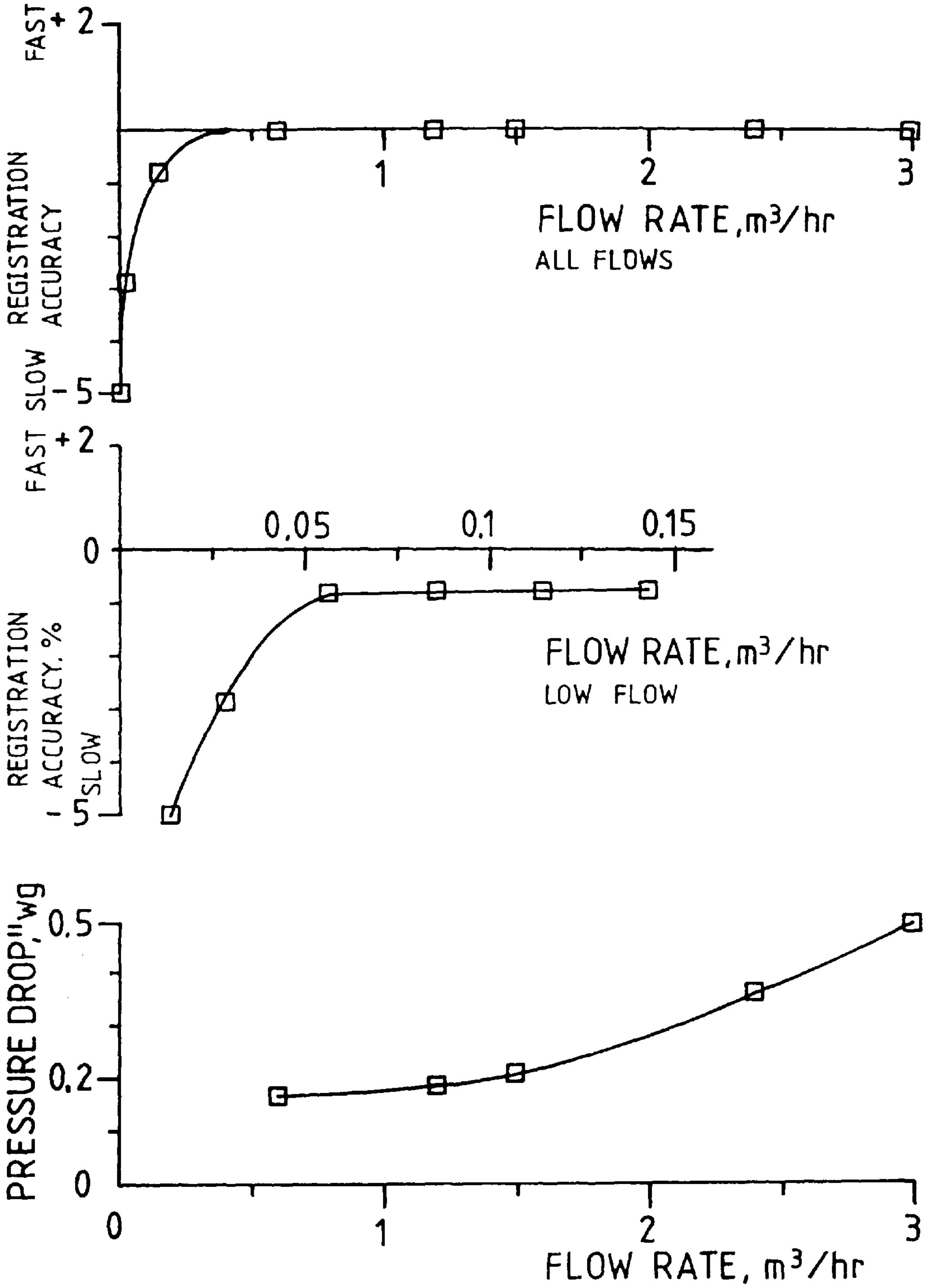
When selecting a gas meter, accuracy and reliability are of the utmost importance. There were several meters which were designed to measure flow-rates similar to these, but none were designed to be operated with hydrogen sulphide, which is very corrosive, in the gas stream. The most accurate meter would have been either a "wet" gas meter or rotameter, but these are both liable to fail when hydrogen sulphide is present in the gas stream. They are both expensive to purchase and maintain.

By far the most common type of gas meter is the domestic type. These had the advantage of being cheap, but are normally rated at 6 m^3/hour and above. However, a version rated at 3 m^3/hour is made, and its accuracy verified at the flows expected by testing in the manufacturer's laboratory (Thorn EMI). Figure 2.9 is the result of one such test, and it shows the registration accuracy of the meter to be correct to within -5%, at the low flows tested.

Daily gas production from each digester was measured using a standard domestic gas meter (U3, Parkinson Cowan Ltd), rated at $3m^3$/hour. This was the lowest rated domestic gas meter produced in the United Kingdom. A problem commonly found with this type of meter is corrosion of the flexible diaphragms, which were traditionally made of leather, by hydrogen sulphide. The meters used were of polymer diaphragm construction, and were more resistant to corrosion.

Careful siting of the gas meters was important to minimise any condensation in the gas meter itself. A water trap was placed in the gas pipeline immediately adjacent to the gas meter, and this was drained periodically, as necessary. A gas sampling point was inserted in the gas pipeline immediately before the gas meter.

FIGURE 2.9 Registration accuracy of Parkinson Cowan U3 gas meter.



THORN EMI, 1983

A pulsed output unit was designed and built to convert the mechanically measured gas production into an electrical signal. There were no such units commercially available. The Health and Safety Executive have expressed concern (West, 1984) at the design and construction of the pulsed outputs on some gas meters because of the low (5V) voltage present inside the index cover. This problem was avoided in the present design by mounting the pulsed output externally to the index face.

An infra-red emitting diode and silicon phototransistor sensor were housed in a moulded package. The sensor responds to the radiation emitted from the infra-red source when the silvered zero is within the field of view. Possible problems with ambient illumination were overcome with an infra-red transmitting filter. Pulses were sensed by a Schmitt trigger remote from the gas meter, and then fed into the datalogger in digital form. The construction of the pulsed output unit was such that it allowed the gas meters to be read directly with a greater degree of accuracy, than by the pulsed output unit. The proving roller on the meter is 10 1, and the zero is silvered as standard. Readings correct to the nearest litre can be obtained.

2.3.2 Temperature measurement and control

The temperature of the digesters was sensed using copper/constantan (Type T) thermocouples. Previous workers (West, 1984; Friman,1985a) working with slurry have noted corrosion problems with thermocouples made from these materials when they are simply spot welded at the end. A further dis-advantage with such a simple design is that liquid slurry can get between the thermocouple wires and their protective casing. Any hydrostatic head of slurry above the thermocouple will force slurry along the wire to where the casing ends, typically at the measuring instrument.

In the light of this experience, it was decided to use thermocouple probes (Labfacility Ltd.). These were 50mm long and consisted of a single hot junction in the silver-soldered end of a stainless steel sheath. The conductors were glass-fibre insulated to the flexible extension cable. Probe diameter was 3mm and all welded construction was employed.

The thermocouple probes were placed in compression glands which were previously screwed into the digester wall, along the uppermost edge. The lowermost thermocouple was 0.75m from the bottom end plate of the digester, and the interval between thermocouples was 1.25m. The numbered leads from the thermocouples ran to the flanged joints of the digesters, where they were joined to extension cables by minature Type T thermocouple connectors. Such connectors, which were made of similar material to the thermocouples, were necessary to avoid measuring errors caused by spurious e.m.f.'s.

The thermocouples and associated wiring were fitted before the digesters were externally insulated to protect them from the weather and moisture ingress.

On each digester, one of the thermocouple probes was used as the sensor for controlling the heating system. This sensor will be referred to as the "control thermocouple".

The signal from the control thermocouple was fed into the datalogger, and was monitored in the same way as the other thermocouples. In addition, it was scanned every 10 minutes by the logger, and if the reading was below a preset level, equivalent to the target operating temperature, then an alarm output channel was activated. The output channels were part of the Digital Input/Output Module (Solartron Ltd., 35302B). When an alarm condition was detected, a solid state relay was activated from the internal floating 5V supply. The relay's 240V contacts were "made" and the digester's electric heating was switched on. When a logger scan indicated that the temperature of the control thermocouple was above the present level, the heating was switched off.

2.3.3 Datalogging

A Solartron 3530 Data Logging system was used for monitoring and control of the digesters. The system was being used for another project on the same site, but proved extremely suitable for the present application.

The datalogger was a very versatile instrument, and up to 200 channels could be defined for measurement. A maximum of 9 programmes can be defined on the instrument, and each programme can have up to 8 tasks. Within each task, up to 200 channels can be defined. The programmes were stored on a magnetic tape cassette, and were loaded into the datalogger's non-volatile memory when required for use. One programme was used for the present study, and the tasks are described below:

TASK NUMBER	DESCRIPTION
8	Detected the switching on (by a separate timer) of a pump which occurred at 2 hourly intervals. This also stimulated TASK 7.
7	Read gas meters. Output goes to cassette
6	Detects events e.g. failure of a device in the control panel. Continuously on
5	Temperature scan of all thermocouples. Done every 6 hours. Stimulated by an internal timer. Output goes to cassette
4	BLANK
3	Analogue alarms. Channels were scanned every hour and any alarm conditions detected (high, low or outside defined band) were output to paper tape on the front of the logger
2	BLANK
1	Digester temperature control. Every 10 minutes, the control thermocouple of each digester was scanned and when it was below the preset level, then the digester heating was switched on

The datalogging system was menu based, the logger being informed of the type of sensor e.g. Type T thermocouple, via programmed function keys on the front of the instrument. There was no need for supplementary amplifiers for thermocouples, as with some other systems. Inputs to the logger were in either analogue or digital form. A particular advantage with this instrument was that digital inputs were converted into real or "user" units. The number of pulses received for any particular channel was recorded by a count function built into the logger, and when interrogated, this number of pulses was multiplied by the appropriate conversion factor to give the actual value. This function was scanned every two hours, output being written to an integral magnetic cassette, and the register was zeroed. Similarly, the instrument can also display real or "user" units from analogue inputs, e.g. temperature values, from thermocouples.

The logger was mains operated, but in the event of power failure a battery backup (24V DC) maintained all logger functions. When the 24V supply failed, an internal battery maintained the date and time until power was restored. The provision of the 24V backup was very important because it enabled the logger to carry on functioning as before. This meant that no data shoudl be lost when power failure occurred, and any control done by the logger continued. This was obviously only applicable when the other equipment did not depend on mains voltage for operation. Power failure was recorded as an event, and the time of occurrence and duration were recorded. A low voltage supply (5V) was provided by the logger, and it was used for pulse recording and in the switching of 0V contacts.

The Datalogger was linked to a Digital MINC PDP-11 mini-computer with 64K of RAM, via an RS232 serial interface. A computer programme (OFLOGB) was available for controlling the logger and transferring data from the magnetic tape cassette to the computer, for more permanent storage on floppy disk. This programme was run at intervals of approximately 3 days. Data incoming during transfer to the computer was temporarily stored in a logger buffer memory, which was interrogated at the end of the main transfer period. The only time that incoming data would be lost was during the short (75 seconds maximum) period when the logger tape was being erased. Data was stored on 8" floppy disks in direct access files for further processing by an existing suite of programmes. A separate file was set up each day for "channels" and "events".

CHAPTER 3

MATERIALS AND METHODS

3.1 Slurry

The digesters were constructed adjacent to a 400 cow dairy unit, and manure was obtained from this source. The unit was operated as 2x200 cow herds of lactating British Friesian dairy cows which were fed a maize silage based diet, and were housed throughout the year. Slurry was obtained from the half of the unit where the cows were bedded on sawdust.

Cubicle passageways were scraped twice daily into a below-ground cross-channel. This channel also received some parlour and collecting yard washings. A mechanical scraper in this channel transported slurry into a $10m^3$ reception pit. This scraper was operated daily. A submersible pump (Flygt Ltd.) was used to mix the reception pit and either to pipe slurry to a mechanical separator or transport whole slurry to the $1m^3$ digester feed tank. This tank was stirred continuously to keep the contents relatively homogenous, and at a uniform temperature. A thermocouple probe sensed the temperature of this tank.

3.1.1 Slurry separation

Slurry was separated with a commercial two stage roller press machine (Farrow Irrigation Ltd.). A conventional brushed screen is used for the primary stage, followed by a roller-pressing stage. The second stage consists of a semi-circular screen, with 3mm holes, over which passed a series of brushes and spring-loaded rollers. The separator was mounted on a 4m high gantry, and separated slurry flowed by gravity to the digester feed tank.

3.2 Total Solids determination

Total solids (TS) were determined by drying 200g samples to constant weight at 105°C (Anon, 1972) in an industrial oven. Samples were placed in aluminium trays and weighed when cool before and after drying.

3.3 Volatile Solids determination

Volatile solids (VS) were determined by combusting dried solids in a muffle furnace (Carbolite Eurotherm) at $500^{o}C$ for at least 30 minutes (Anon, 1972). Samples were placed in weighed porcelain crucibles and weighed when cool before and after combustion. The solids lost during combustion are termed "volatile solids", and approximate to the organic and hence bio-degradable proportion of the total solids. Volatile solids are expressed here as a percentage of the total solids, and are correctly referred to as the "Total Volatile Solids" (TVS).

3.4 Volatile Fatty Acid determination

Volatile fatty acids (VFA) are organic acids containing 6 or less carbon atoms. They are produced by the acid-forming bacteria during the anaerobic digestion process, and are also called "short-chain fatty acids". The VFA concentration is expressed in terms of acetic acid. The VFA concentration is often monitored as an indicator of digester performance and stability.

3.4.1 By Steam Distillation

The total volatile fatty acid (TVFA) concentration was determined by steam distillation of prepared samples in a Markham Still. During distillation, the lower carboxylic acids in the sample were released from solution. The distillate was collected and titrated against standard alkali. This method does not differentiate between the fatty acids present in the slurry.

A 10ml test tube of slurry was centrifuged at 2500 rpm for 10 minutes. 5ml of the supernatant was mixed with an equal volume of acidified magnesium sulphate in a test tube. The tube was inverted slowly, and centrifuged at 2500rpm for 10 minutes. 2ml of the supernatant was distilled in a Markham Still until at least 75ml of distillate was collected. The distillate was titrated against 0.025M sodium hydroxide, using phenol red as an indicator. During the titration, nitrogen was bubbled through the distillate in order to make the end point more stable. The distillate was stirred by a magnetic stirrer. Without the nitrogen, carbon dioxide from the atmosphere dissolved into the distillate changing the pH, and making the end point difficult to determine.

3.4.2 By Gas Chromatography

The individual volatile fatty acid concentrations (C_2 - C_6) were analysed by gas chromatography, using a Pye 304 chromatograph, by the method of Williams et al, (1984). Samples were prepared by centrifuging and adding 0.2ml of a solution of 2.0g/l crotonic acid internal standard in 2.4M tribasic phosphoric acid (H_3PO_4) to each millilitre of supernatant. Samples of 1ul were injected onto a 25m quartz capilliary column coated with Free Fatty Acid Phase, by a Grob injection head and an auto-injector. The operating temperatures were : injection head, column oven and detector block; 200, 130 and 250^{o}C, respectively. The carrier gas, injection purge and septum purge flow rates of helium were 1.0, 10 and 1.0ml/min, respectively. The hydrogen and air flows to the flame ionization detector were 40 and 400 ml/min, respectively.

3.5 Gas Composition

The carbon dioxide and hydrogen sulphide concentrations in the gas were measured weekly. A gas sample of 100ml was drawn through a detector tube (Draeger Safety), into a Draeger Multi Gas Detector 21/31 (Draeger Safety), which is a hand-operated bellows pump. In operation, the bellows were compressed and the detector tube inserted into the gas stream. Gas was drawn through the tube for the "opening period" which was the time needed for the release of the compressed bellows, ending when the arrestor chain was fully tensioned. Gas concentration was proportional to the length of the tube discolouration.

The Draeger tubes used were:

	Reference No.	Measuring Range
carbon dioxide 5%/A	CH20301	5-60% vol
hydrogen sulphide 100/a	CH29101	100-2000ppm
hydrogen sulphide 0.2%/a	CH28101	0.2-7% vol

3.6 Tracer studies

The inclined tubular digester is a form of plug flow reactor, where feed moves sequentially through the digester from the lowermost (inlet) to the uppermost (discharge) end. True plug flow is unlikely to occur, but the manner in which solid and soluble components of the digester feedstock move through the reactor is of interest since it directly determines the suitability of a reactor design for a particular feedstock.

To study the movement of particulate, soluble and microbial components through the reactor, markers or "tracers" which move with the component in question are needed. A tracer should not interfere with the anaerobic digestion process, and it should not take up a large part of the digester volume. The background concentration of any tracer material already in the digester feed should be low, or the resulting interference will make subsequent analysis more difficult. Similarly, any tracer must not be added at a concentration which is toxic to the anaerobic digestion process.

Many different markers have been used in studies on the movement of feed through animals (Uden et al, 1980) and much of this expertise, particularly from studies with ruminants, is directly aplicable to tracer studies in anaerobic reactors, since in both cases conditions are anaerobic, and similar groups of bacteria exist. Markers which have been commonly used in animal studies include water insoluble minerals such as chromic oxide (Cr_2O_3), various dyes, particles of plastic, rubber and other materials, water soluble materials such as polyethylene glycol, the cobalt chelate of ethylenediaminetetraacetic acid (Co-EDTA), the chromium chelate of ethylenediaminetetraacetic acid (Cr-EDTA) and particulate bound markers including the rare earths Ytterbium (Yb) and Lanthanum (La), (Ellis et al, 1982). Radioisotope techniques have also been used in digester mixing studies (Zoltek and Gram, 1975).

Many of the above markers have been shown to have application in tracer studies with animals. However, the cost and complexity of marking the feed dose precludes their use in larger "reactors", such as in the present study, and also in much larger industrial plants. Some markers require complex analyses which may be very accurate, but which could not be justified on a large scale.

Tracer studies in municipal sewage digesters have concentrated on determining the actual hydraulic retention time for digesters, and calculating the associated active or working volume (Monteith and Stephenson, 1981; Tenney and Budzin, 1972). The most commonly used tracer was sodium fluoride, movement of the fluoride anion being followed by an ion specific electrode. This method is very simple, and has the advantage that tracer concentrations can be measured in minutes, since no wet chemical analyses are necessary.

Many other materials are available in fluoride or chloride forms, and these could have been used equally well and analysed using a specific ion electrode, or, similarly, virtually any metal compound could be used, and, knowing the atomic weight, be analysed by atomic absorption spectrophotometry.

There are several reported tracer studies on digesters operating on agricultural wastes. Fanfoni et al (1978) used a chromium tagged fibre to follow the movement of solids, and soluble salt tracers of Co Li-EDTA.3H_2O and sodium chloride (Na Cl) to determine liquid retention time in a plug flow digester (5m^3) fed 10-12%TS cow manure. These tracer studies showed the average retention time to be about 2/3rds of the calculated value, and the solid and liquid retention times were the same. The cobalt tracing technique described by Fanfoni et al (1978), was used by Jewell et al (1980) in tracer studies on full scale (34m^3) plug flow reactors operating on 10-12%TS cow manure. In this work only 1 tracer (liquid) was used because of the close relationship between solid and liquid retention times which had been shown in the work of Fanfoni et al (1978).

Petersen (1984) and Friman (1985b) report the use of lithium chloride (Li Cl) as a tracer in studies on different plug flow digesters. Each digester was fed on cow manure of approximately 10%TS. The advisability of using 1 liquid tracer on a 10%TS feedstock is questionable, since no information is gained on how solids move through the reactor.

Laboratory scale tracer studies on inclined tubular digesters have been carried out looking at the movement of liquid components by Hawkes et al (1981), and on solid and liquid components by Floyd and Hawkes (1986).

Chromium-tagged particles showed digester solids to move in well dispersed plug flow, and a $MgCl_2$ tracer showed soluble components of the slurry were mixed throughout the whole digester volume in 25% of the hydraulic retention time or less.

Hamad et al (1983) working with a Chinese digester design of $5m^3$, fed on water-buffalo dung at 9-11%TS used a plastic material of specific gravity 1.05 as a tracer. The plastic beads were chosen as a tracer in that situation because of the simplicity in determining the exit tracer concentration. The plastic beads in the effluent were collected by sieving and washing, followed by drying and weighing to determine the tracer concentration.

3.6.1 Selection of tracers

Two separate tracer experiments were carried out with the inclined tubular digesters to investigate the manner in which feedstock moved through the reactors. In the first expeiment, chromic oxide was used as the tracer and Cr was analysed by atomic absorption spectrophotometry. Chromic oxide has been criticised as a tracer for use as a specific flow marker in animal studies by Uden et al (1980) and Ellis et al (1982) on account of its lack of association with either particulate or water soluble components. However, it has been widely used as an inert dilution marker, and it was chosen for the initial tracer experiment since the gas production from the digesters was declining rapidly (see Figure 4.2) and the main information required was an estimate of the actual hydraulic retention time. The digester feedstock at this time was whole cattle slurry.

The second tracer experiment was conducted during a period of relatively stable digester operation (see Figure 4.2) on separated cattle slurry. Two markers (chromic oxide and lithium chloride) were used to try and distinguish between the movement of solid and liquid components. Lithium chloride had previously been used in digester tracer studies by Friman (1985b), Petersen (1984) and Noone and Brade (1982).

3.6.2 Methods

The tracer technique used required that all the tracer be introduced as a slug dose. The simplest way of achieving this was to dose a tank of slurry, containing slightly more than the volume of slurry to be fed, with markers, and for this tank to be directly connected to the feed pumps. This arrangement ensured that all the markers entered the digesters at the same time, and the possibility of any residual tracer entering the normal feed tank and subsequently entering the digesters was eliminated.

In Tracer Experiment 1, chromic oxide was the only marker used. The actual volume of slurry to be fed to both digesters was 68.5l. Chromic oxide M100 (British Chrome and Chemicals Ltd., Stockton-on-Tees), in the form of a green powder, was added to the feed slurry to give a concentration of 8g Cr_2O_3/l slurry, a total of 100l of slurry being dosed in this way. Assuming the digesters were completely mixed, then the concentration of chromic oxide in each digester =

$$\frac{68.5l \times 8 \text{ g/l}}{(0.91+1.83) \times 1000} = 200 \text{ mg/l} = 200 \text{ ppm}$$

The actual concentration of chromium (Cr) is 200 x 0.68 = 136ppm, since the molecular weight of Cr is 52, and O is 16.

The slurry was stirred rigorously as the chromic oxide was added to the dosing tank. The feed pumps were then operated as normal, the volume of slurry discharged from each digester was collected and measured. The discharged slurry was stirred rigorously, and 3 x 1l samples were taken from each tank. After these samples had been taken, the tank collecting the effluent from each digester was emptied and washed out. These samples were then mixed to give 1 sample per digester feed, for each digester. Samples were taken each time the digester fed (1000, 2200 hrs). After the dosed slurry had been fed to the digester, the normal digester feed tank was re-connected, after ensuring that all residual chromic oxide had been removed.

The above procedure for the addition of markers, was repeated during Tracer Experiment 2. Chromic oxide (British Chrome and Chemicals Ltd.) was added at the same rate of 8g Cr_2O_3/l slurry. The total feed volume was 901 (RT = 15 days), and, assuming complete mixing, the concentration of chromic oxide in each digester =

$$\frac{90 \times 8}{(0.91+1.83\times1000} = 263 \text{ mg/l} = 263\text{ppm}.$$

The concentration of Cr is 263 x 0.68 = 179ppm.

Hydrated lithium chloride (BDH Chemicals Ltd., Poole), dissolved in water, was added at a rate of 1.25g Li $Cl.H_2O$/l slurry. Assuming complete mixing, the concentration of Li $Cl.H_2O$ in each digester =

$$\frac{90 \times 1.25}{(0.91+1.83)\times1000} = 41.1 \text{ mg/l} = 41.1\text{ppm}$$

The concentration of Li is $41.1 \times \frac{7}{60.41} = 4.76$ppm

(Mol. Wt. of Li = 7, Li $Cl.H_2O$ = 60.41).

Effluent samples (2x200g) were dried to constant weight at 105°C. The material remaining was ground in a Moulinex coffee grinder and stored in air-tight bags prior to further analysis.

3.6.2.1 Assay for chromium

The wet digestion method of Stevenson and de Langen (1960) was used for the determination of chromium sesquioxide (Cr_2O_3). After oven-drying at 105°C, samples were ashed and digested with a sulphuric / phosphoric acid mixture to bring the chromic oxide into solution, mostly as dichromate. Addition of potassium bromate solution oxidised the remaining chromic compounds to dichromate. The mixture was made up to 100ml with water, and the chromium concentration was measured (atomic absorption) by direct aspiration (without dilution) into a nitrous oxide - acetylene flame. The lamp current was 25 mA, the absorption line and slit width were 429.0 nm and 0.7 nm respectively. An integration time of 3.5 sec was used. Working standards of potassium dichromate equivalent to 2 - 150 ug / cm^3 Cr_2O_3 were used, and re - calibration was carried out after every 10 samples. Reproducibility of the method is high (Stevenson and de Langen, 1960), to within 2 %.

3.6.2.2 Assay for lithium

Wet digestion of oven - dried samples was carried out as for the chromium samples. The digest was analysed by atomic absorption spectrophotometry. An oxidising air / acetylene flame was used, with an absorption line set at 670.8 nm, and a slit width of 0.7 nm.

3.7 Dataprocessing

Numerical data from the Solartron datalogger was recorded onto magnetic floppy disk by the Digital MINC computer. Information of this type was gas meter readings, and digester temperatures. The gas production data, which was logged every 2 hours, was printed out regularly and checked.

Results from the laboratory analyses of slurries were recorded on special purpose sheets. This information was transferred to magnetic floppy disk for futher processing using an Acorn BBC Model B computer. The programming language used (BBC Basic) is easy to follow and very versatile. A number of programmes were written for calculation and graphical presentation of digester performance data. Other programmes were written to process the data on energy production, tracer studies and scale up of the digester design.

The BBC Model B computer was chosen for subsequent dataprocessing, mainly because of its versatile programming language. Additionally, data could easily be transferred from this computer to the Super Mini VAX system at the Institute (AGRI), which had many standard software packages on it. It was initially thought that these may be of use to the project. However, they were not used extensively and the majority of the processing was done on the BBC Model B computer.

CHAPTER 4

RESULTS OF STUDIES WITH PILOT SCALE DIGESTERS

4.1 Experimental programme

The main objective of the experimental programme was to establish the effect of a range of operational and design parameters on digester performance in terms of biogas production and changes in slurry composition.

A series of experiments were carried out to investigate the relationships between retention time, angle of inclination, aspect ratio, temperature and type of slurry. Digester performance was assessed from daily biogas production, biogas yields, gas composition, changes in total and volatile solids content, and volatile fatty acid concentration.

These experiments were conducted over the period March 1984-November 1985 (20 months). Table 4.1 indicates the sub-periods during which particular combinations of operational and design parameters were applied. At all times the angle of inclination was the same for both digesters. Similarly, the aspect ratio of Digester 1 was 10, and Digester 2 was 20.

TABLE 4.1 SUMMARY OF DIGESTER OPERATING CONDITIONS

Period	Operating conditions 1984/5	Days	Inclination (° to horiz)	RT (days)	Temp. (°C)	Slurry type
1	Start up and commisioning. Feb 13-April 26	74	20	20	35	whole
2	Operation at 20 day RT. April 26-June 18 EXPERIMENTAL RUN 1 (April 26 - June 9)	54 45	20	20	35	whole
3	Operation at 15 day RT. June 18-Aug 31 EXPERIMENTAL RUN 2. (July 18-Aug 31)	75 45	20	15	35	whole
4	Operating temperature reduced to 25°C (Aug 31-Oct 10)	41	20	15	25	whole
5	Operation at 15 day RT. Oct 10-Dec 3 EXPERIMENTAL RUN 3 (Oct 10-Nov 23)	77 45	20	15	25	whole
6	RT increased to 20 days. Dec 3-Jan5, 1985	34	20	20	25	whole
7	Shut-down and digester modifications (Jan 5 - Mar 18)	73	-	-	-	-
8	Start up and operation at 20 day RT. (Mar 18 - Aug 20) EXPERIMENTAL RUN 4 (May 8-June 21)-Dig 1 EXPERIMENTAL RUN 4 (June 8-July 10)-Dig 2	156 45 33	10	20	35	whole
9	Shut down Aug 20-Sep 4	16	-	-	-	-
10	Start up and operation at 15 day RT. Sep 4 - Nov 12 EXPERIMENTAL RUN 5 (Sep 19-Nov 2) Shut-down. Nov 12, 1985.	70 45	10	15	35	separated

NOTE : The aspect ratio of Digester 1 was 10, and Digester 2 was 20.

Period 1: Start up and commissioning

The assembled digesters were filled with water to check for leaks, and test the heating and associated control system. All instrumentation was checked at this time.

The digesters were 2/3rds filled with digested cattle slurry, obtained from a farm scale digester operating on the same site. The remainder of each digester was then filled with fresh cattle slurry, and daily feeding of the digesters commenced.

Period 2: Operation at 20 day RT, 35°C and 20° to horizontal.

EXPERIMENTAL RUN 1

Data for Experimental Run 1 was obtained during the first 45 days, the digesters having reached steady state operating conditions.

Period 3: Operation at 15 day RT, 35°C and 20° to horizontal

EXPERIMENTAL RUN 2

The retention time for each digester was reduced from 20 to 15 days, all other parameters remaining unchanged. The last 45 days of this period were used as Experimental Run 2.

Period 4: Operating temperature reduced from 35 to 25°

The temperature of each digester was reduced from 35 to 25°C on Day 1, and the digesters allowed to adjust to this change in operating conditions.

Period 5: Operation at 15 day RT, 25°C and 20°
to horizontal
EXPERIMENTAL RUN 3

Experimental Run 3 was conducted during the first 45 days of this period.

Period 6: Retention time increased from 15 to 20 days

Digester retention time was increased from 15 to 20 days and this period was for the digesters to adjust to this change in operating conditions. However, despite seemingly adequate insulation and frost protection measures, it was not possible to operate the digesters beyond the end of this period and they were shut-down on account of prolonged severe weather conditions.

Period 7: Digester shut-down and modifications

The digesters were emptied and cleaned. The effluent weir on each digester was modified, since they had not been operating entirely satisfactorily.

Period 8: Start up, and operation at 20 day RT, 35^{o}C, and 10^{o} to horizontal

EXPERIMENTAL RUN 4

The digesters were started up again, using the procedure as described in "Period 1". Angle of inclination was 10^{o} to the horizontal, compared with 20^{o} previously. When steady state conditions had been achieved, data for Experimental Run 4 was collected. The performance of Digester 2 was somewhat unstable, and a period of 33 days was used to collect data.

On completion of Experimental Run 4, the performance of both digesters declined, and they were both unstable. In order to investigate this falling off in performance a tracer study using chromic oxide as the marker was carried out. Additionally, it proved possible to measure individual fatty acid concentrations on an intermittent basis. Up to this time, only the total volatile fatty acid concentration had been measured.

Period 9: Digester shut-down

The digesters were shut-down, due to their very low gas production. This was because there was considerable accumulation of fibrous material in the digesters, and a recent tracer study had the actual retention time to be approximately 2 days, whilst the theoretical value was 20 days.

The digesters were emptied and washed out. It was not necessary to dismantle the digesters in order to empty them. Digester configuration remained the same.

Period 10: Start up, and operation at 15 day RT, 35°C, and 10° to horizontal. Separated slurry.
EXPERIMENTAL RUN 5

The digesters were started up according to the procedure described in "Period 1", except that the seed used was separated digested slurry. Retention time was 15 days, and Experimental Run 5 of 45 days duration was conducted towards the end of this period. The feedstock was separated cattle slurry. A tracer study using lithium chloride and chromic oxide as the markers was carried out during this period. Upon completion of this tracer experiment, the digesters were emptied and shut down.

4.1.1 Digester operation

The digesters were constructed on a farm site, and were designed to operate with a minimal amount of daily labour input. Each digester was fed by an electrically driven feed pump (Mono Pumps. Ltd.) at 1000 and 2200 hours daily, that was controlled by an electric timeswitch. A separate electric timer (Omron Ltd.) contolled the period of operation of each feed pump. When the digesters were fed , a volume of slurry equal to the feed volume was discharged over the effluent weir and into an adjacent slurry lagoon. The digester feed was stored in a tank of $1m^3$ capacity, and was filled at intervals of approximately 3 days by an overhead slurry pipeline.

The digester heating system was monitored and controlled by a datalogger, which also recorded digester temperature at 6 hourly intervals. Gas production was measured with domestic bellows type gas meters, and signals from the attached pulsed output units were recorded every 2 hours. All information recorded by the datalogger was transferred to the Digital MINC computer for more permanent storage on magnetic floppy disk.

Twice weekly, samples of feed and discharge slurries were collected from each digester and analysed for total solids, volatile solids, and volatile fatty acids. On one day each week the gas and electric meter readings were recorded manually, and the gas composition (carbon dioxide and hydrogen sulphide) measured. At this time the digesters were checked over visually, and examined for any damage or mal-function. The water trap in the gas line by each gas meter was removed and emptied weekly. A log was kept of any equipment mal-functioning, together with the recording of any other events which could affect digester performance e.g. power failures.

In general, the digesters functioned reliably. Some problems were encountered with the formation of crusts in the effluent box on each digester. This meant that digested slurry was not able to flow over the discharge weir. On two occasions, gas meters had to be replaced due to slurry entering the gas pipeline, and ultimately the meters themselves. In the short term, this problem was minimised by the manual removal of crusted fibre from the effluent boxes. The problem was finally overcome with the re-design of the effluent weir.

4.1.2 Digester Performance Criteria

There is no one parameter which can be satisfactorily used to express digester performance. Rather, from the routine analyses made, additional parameters are calculated, and when considered together, digester performance can be assessed satisfactorily.

The parameters measured routinely were the TS, VS and VFA of the feed and discharge slurries, daily gas production and ambient, feed and digester contents temperatures.

Total Solids: This parameter indicates the amount of solid material in a sample. However, the measurement is of all solids present, including those which are not biodegradable such as sand and grit which do not contribute to gas production. A waste which is low in TS e.g. 1-2% will be large in volume terms. Generally, the TS should be kept as high as possible. Wastes of 12-14% are at the upper limit of acceptability. Where the TS is greater than this, dilution with water or the use of a digester designed for high solids concentrations, will be necessary.

Volatile Solids: The solids of the feed material is usually only partly degradable, and this portion is known as the volatile solids. In many wastes the biodegradable portion is referred to as the VS, since they are roughly equivalent to the amount volatilised on heating (Section 3.3).

Volatile Fatty Acids: These are fatty acids containing 6 or less carbon atoms which are soluble in water and which can be steam-distilled at atmospheric pressure. In general, the higher the VFA concentration, then the greater will be the gas production. However, very high levels of VFA (>10,000ppm), perhaps caused by addition of silage liquor, can lead to failure of the digestion process.

In many situations, the VFA concentration (feed and discharge), is monitored as an indicator of digester performance and stability. It is usually only necessary to monitor the total VFA (TVFA) concentration. Raised levels of digester VFA may be an effect, not a cause of methanogen inhibition. In situations where digester performance is decreasing, then an analysis of the individual VFA could provide the explanation. It has been shown (Peck et al, 1986) that the concentration of the branched VFA and propionic acid represent more sensitive monitors of digester stability than acetic acid or the TVFA concentration.

Gas production: This may be expressed as m^3/day or, more usefully, as m^3 gas/m^3 reactor/day. The latter expression is still insufficient for a useful comparison to be made unless the RT, TS and temperature are also known.

Gas yields: m^3/kg TS added, m^3/kg VS added.
These gas yields are a useful indicator of digester performance. They have theoretical maximum values which cannot be obtained in practice, unless RT, Loading Rate, Temperature and VS content are optimised.

Because these gas yields are expressed per unit weight of TS or VS added, they are useful for digester design purposes, since the volume of waste to be treated is usually known, and likely gas production can be estimated.

The problem of not knowing how much of the TS are biodegradable still applies when quoting figures as m^3/kg TS added. However, with animal wastes coming from the same source regularly (daily), then the VS usually remain a very constant proportion of the TS.

Gas yields expressed as m^3/kg VS added are a measure of the degree of digestion of the organic matter. It is the volume of gas generated per unit weight of organic matter added to the digester.

Gas yields: m^3/kg TS destroyed, m^3/kg VS destroyed.
These parameters represent the amount of gas that is produced from the destruction of unit weight of TS or VS. They are useful indicators of the efficiency of the

digestion process, because they are calculated from both the input and output TS and VS values. Gas yields expressed on a "destroyed" basis are always greater than those expressed on an "added" basis. This is because the solids "added" will not all be broken down, except under idealised conditions. Gas yields per unit weight of TS or VS destroyed should be relatively constant, for a particular feedstock. This is because a certain volume of gas is produced from the destruction of unit weight of feedstock constituents (fats, lipids, proteins etc.), whatever the digester. These gas yields may vary due to different amounts of biomass being formed, or the formation of intermediates which are not used. This would be seen in increased VFA concentrations, and any effects are likely to be small in the present study.

Gas yields of m^3/kg VS destroyed are usually a very constant proportion of m^3/kg TS destroyed values.

% TS reduction: This parameter represents the reduction in TS content occurring during the digestion process. It is therefore a measure of how much (by weight) the solids in the feed have been reduced. This parameter is of particular interest to the sewage treatment industry, since anaerobically digested sludge is often tankered (by road) to land for final disposal. Volume reductions are of great significance in this instance. A reduction in the solids content of a feedstock makes it easier to handle by pumping.

% VS reduction: This parameter represents the reduction in VS or organic content occurring during the digestion process. It is a good measure of how much the polluting power of the feedstock has been reduced.

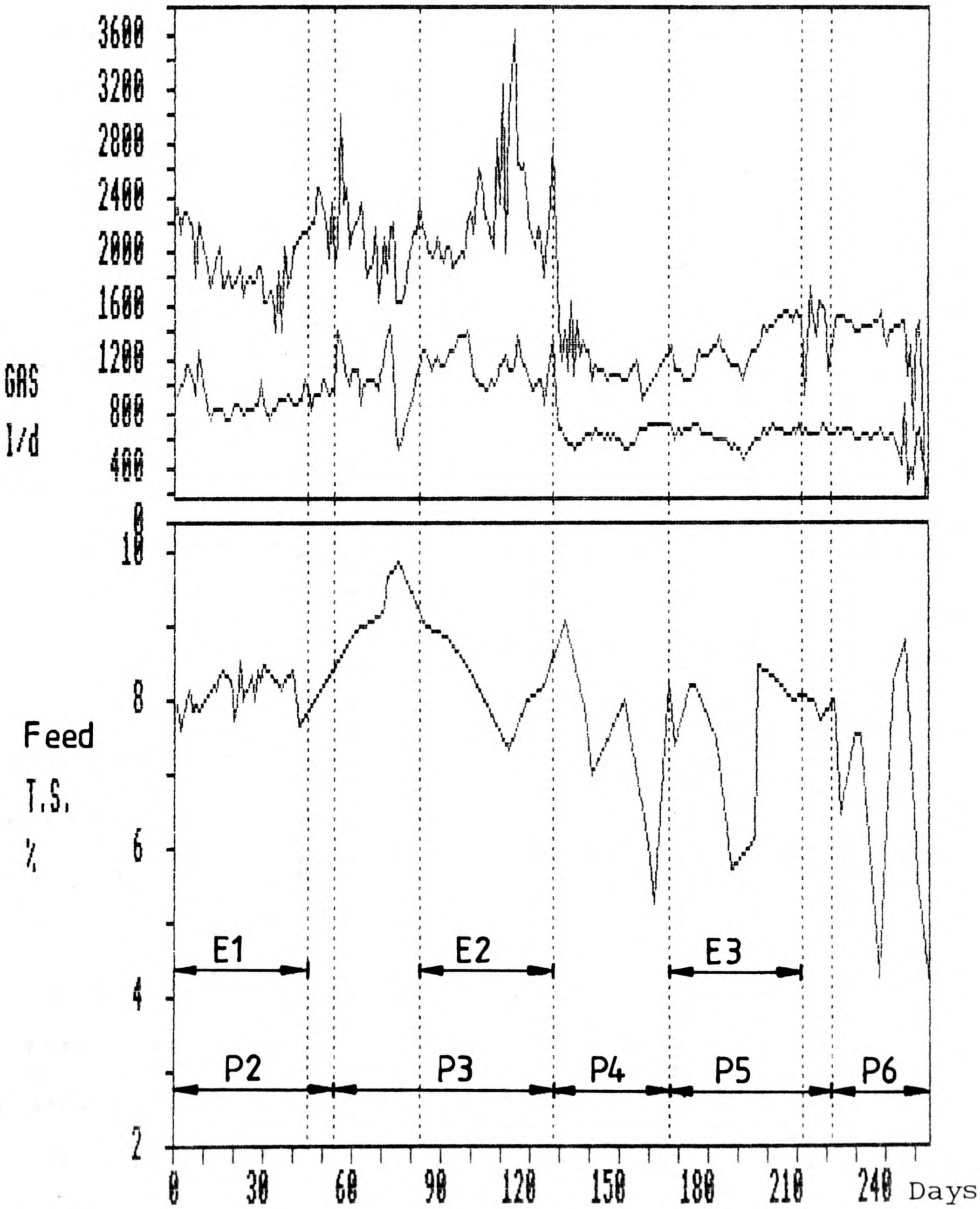
%VFA reduction: This parameter also gives a good indication of how much the polluting power of the waste has been reduced. The parameter needs interpreting with some caution because VFA occur in the feedstock, and they are also produced during the digestion process.

Loading rate (kg VS/m^3 reactor / day: This is the amount of volatile solids added per cubic metre of digester per day. It takes into account the digester retention time and %TS content of the feed material.

4.2 Results

The digesters were operational for the period March 1984-November 1985 (20 months). This can conveniently be divided into two sections, namely from commissioning up to January 5, 1985 (Periods 1-6), and March-November, 1985 (Periods 8-10). The daily gas production from each digester, along with the total solids of the feed during these periods are shown in Figures 4.1 and 4.2 respectively.

FIGURE 4.1 Periods 2-6. Daily gas production and feed total solids content.



P = Period number

E = Experimental Run number

There were five periods when the performance of the digesters was relatively stable, and these are taken as being "Experimental Runs". A standard length run of 45 days was used here, so that results were collected over the same time period, and the same period in time. Experimental Run 4 for digester 2 was of 33 days duration, because the digester was relatively unstable, and there was not a stable period of 45 days. Results from Experimental Runs 1-5 are presented in Figures 4.3-4.10, and Figures 4.13 and 4.14. Results for gas production are from daily readings, VFA concentrations and TS content are the results from twice weekly analyses. Each Experimental Run was divided into 3 equal, successive periods, the mean and standard deviation for digester performance data being calculated from these.

Figure 4.1 shows a strong correlation between the daily gas production from each digester, for Periods 2-6. The period marked E1 is used as an Experimental Run, daily gas production being relatively constant during this period. At the end of Period 2 (P2), the retention time was reduced from 20 to 15 days. In the period up to the start of Experimental Run 2 (E2), gas production fluctuated widely but the trend was upwards. The pronounced fall in gas production just before the start of Experimental Run 2 (E2) was due to problems with the feedback system used to control digester temperature.

At the end of Period 3, digester operating temperature was reduced from 35 to 25^{o}C. Gas production fell sharply as expected, and stabilised within several days. The Period 4 (P4) was allowed for the digesters to adjust to this change in operating conditions.

Experimental Run 3 (at 25^{o}C) followed the Period 4.

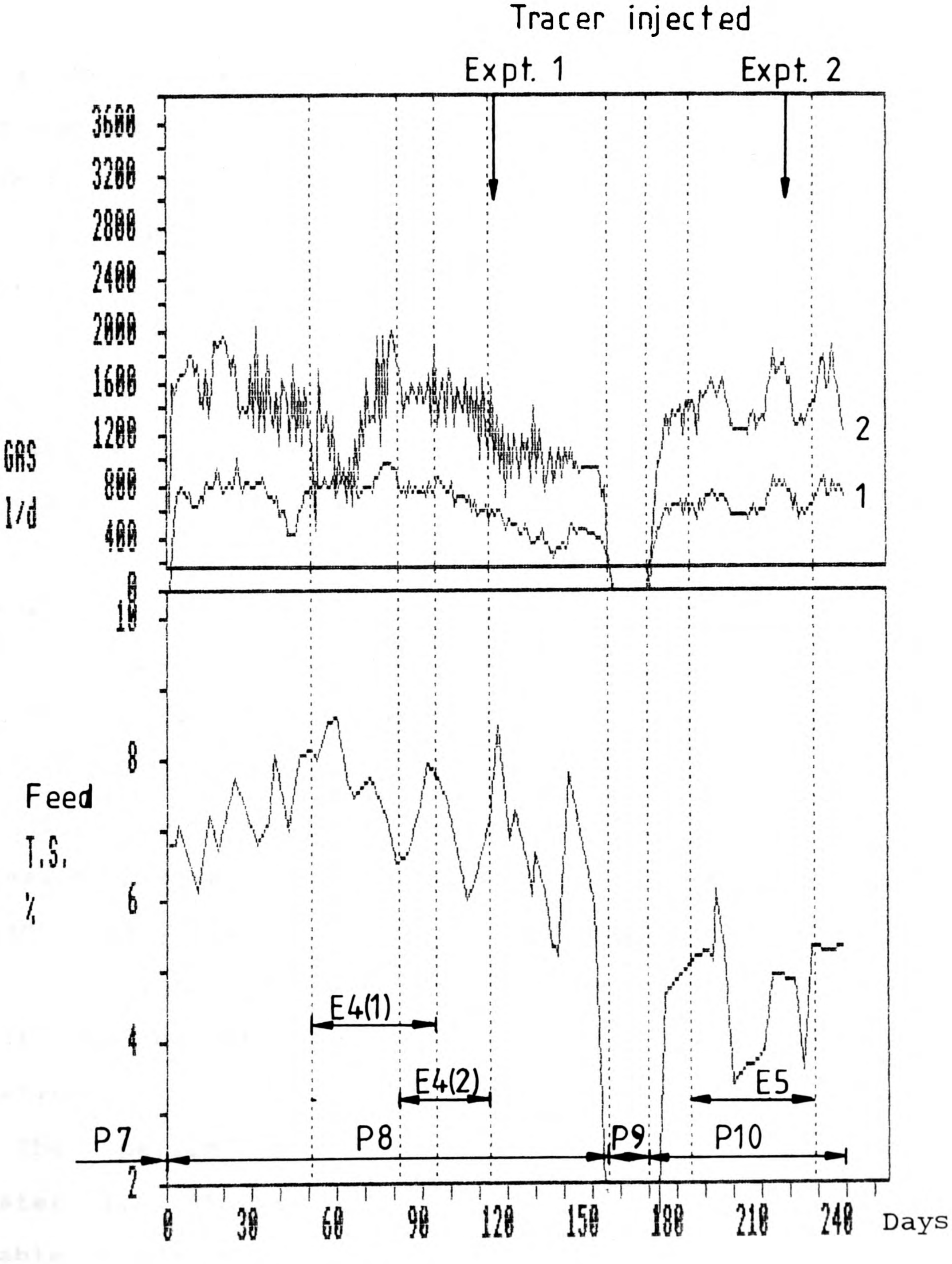
The nominal retention time was increased from 15 to 20 days at the end of Period 5, and Period 6 was for the digesters to adjust to this change in operating conditions. However, despite seemingly adequate insulation and frost protection measures, it was not possible to operate the digesters beyond the period indicated by P6 (January 5, 1985), and they were shut down on account of prolonged severe weather conditions.

The effluent weirs on the digesters had not been operating entirely satisfactorily up to this time, and the opportunity to modify them was taken during this period of digester shut-down (see Section 2.2.5).

Figure 4.2 shows daily gas production for each digester and feed TS content during Periods 8-10.

At the start of Period 8 (March 1985), following modifications to the effluent weirs and checking of the pilot scale plant, the digesters were started up again. The same configuration of aspect ratios was maintained

FIGURE 4.2 Periods 8-10. Daily gas production and feed total solids content.



P = Period number

E = Experimental Run number

(1) = Digester 1

(2) = Digester 2

(D1=10, D2=20), but the angle of inclination was reduced from 20° to 10°. Figure 4.2 shows a strong correlation between daily gas production from each digester.

However, the variation in gas production from Digester 2 on a daily basis is much greater. Examination of the longitudinal temperature profile within Digester 2 showed there to be considerable variations. In order to try and achieve a more uniform temperature distribution, the position of the "control thermocouple" was changed from being 5.75m from the lower end plate of the digester, to 3.25m on 30 April 1985 (Day 44 in Figure 4.2). For the 25 days prior to this, the mean overall digester temperature for Digester 2 was 35.31 °C (SD ±1.25). During this time, temperatures within the raector ranged from 27.2 - 47.6 °C. After changing the position of the control thermocouple, the mean overall digester temperature during Experimental Run 4 (Digester 2) was 36.2 °C (SD±0.83). Changing the thermocouple position reduced temperature variation within the digester, but increased the mean overall digester temperature. It is thought that the lower angle of inclination (10° compared with 20°) significantly reduced the convection currents within the digester, and caused greater temperature variation.

The period E4(1) was Experimental Run 4 with Digester 1. At that time, Digester 2 was relatively unstable, and a different period E4(2) was used to evaluate digester performance.

It can be seen from Figure 4.2, that from the end of E4(2), the trend in daily gas production from both digesters is downwards. In order to investigate the reasons for this, a tracer study using chromic oxide was carried out. The tracer was mixed with the digester feed and introduced as a single pulse close at the end of E4(2). Up to this time, only the TVFA concentration in the feed and discharge slurries had been measured. It then proved possible to measure individual VFA concentrations on an intermittent basis, and this showed the concentration of propionic acid to be greater than acedic acid. Also, VFA concentrations in the discharge from each digester had been high (2000-3000ppm), suggesting short circuiting of the feed. This was confirmed with the results of the tracer study which showed retention time to be approximately 2 days, instead of the nominal 20 days. Consequently, the digesters were shut down at the end of Period 8 (P8).

During Period 9 (P9), the digesters were drained and washed out with water. The digester configuration was not changed.

At the start of Period 10, the digesters were started up with digested separated cattle slurry. Separated slurry was used for the duratio of the experimental period. Experimental Run 5 (E5) was used to gather information for this set of operating conditions.

A second tracer experiment was carried out during this period, lithuim chloride and chromic oxide being used as the tracers.

Upon completion of the second tracer experiment, the digesters were emptied and shut down, at the end of Period 10 (P10).

4.2.1 Experimental Run 1

Each digester was inclined at 20^o to the horizontal, and the aspect ratio of digesters 1 and 2 was 10 and 20 respectively. The nominal retention time was 20 days. The feedstock was whole dairy cow slurry.

The daily gas production, and feed and discharge TS and TVFA are presented graphically in Figure 4.3 (Digester 1) and Figure 4.4 (Digester 2). This information is summarised for both digesters in Table 4.2.

The mean TS content of the feed was 8.13% (SD±0.24), and the VFA concentration was 6034ppm (as acetate), (SD±1081). Gas production (D1) was relatively stable at $1.001m^3/m^3$ reactor day (SD±0.1267), a failure in the heating system (power cut) being responsible for the marked variations from days 5-10. The mean TS of the discharge was 6.07% (SD±0.33).

FIGURE 4.3 Experimental Run 1. Digestion of whole slurry in Digester 1.

Daily gas production with feed (F) and discharge (D) TS and VFA

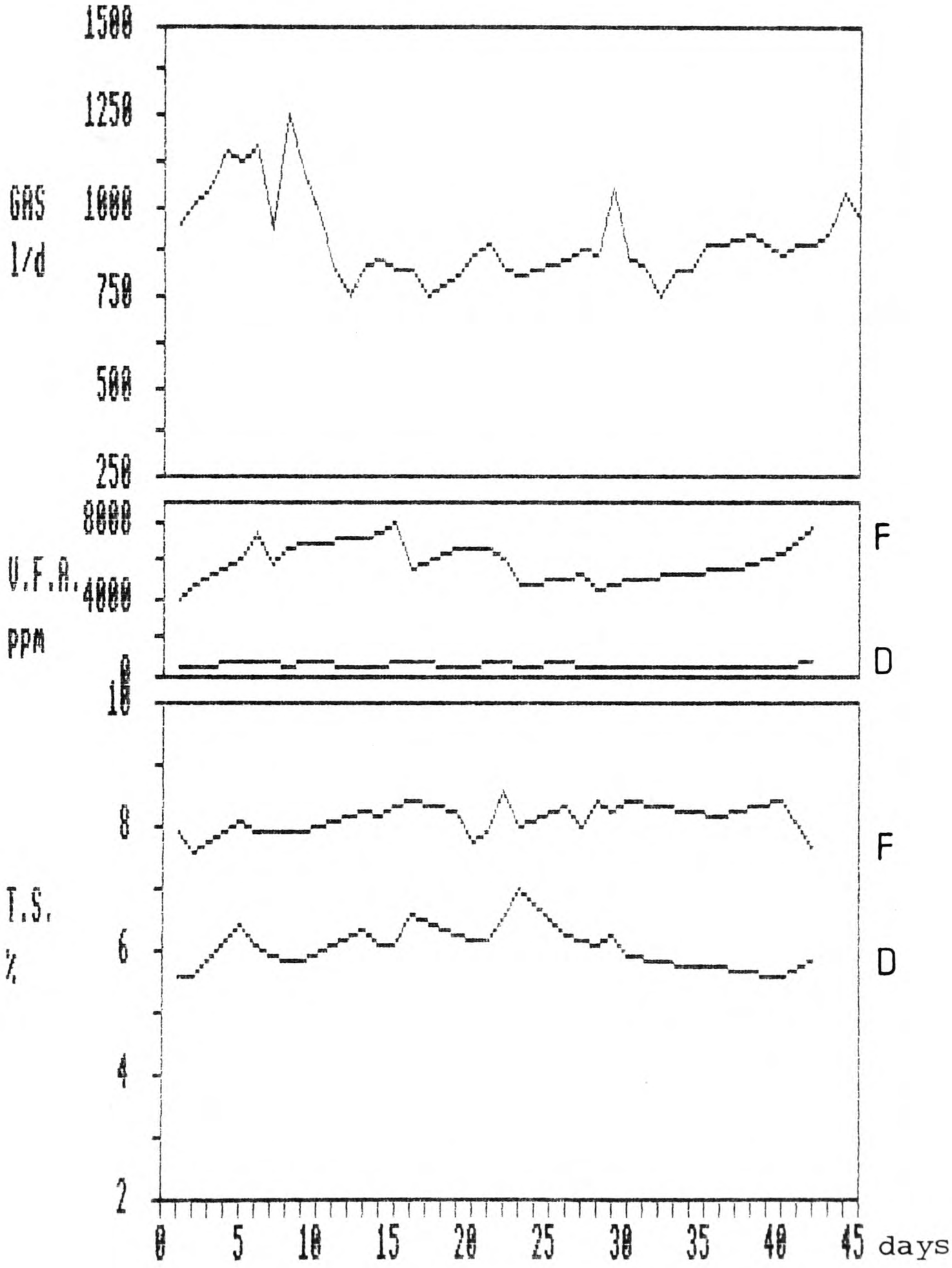


FIGURE 4.4 Experimental Run 1. Digestion of whole slurry in Digester 2.

Daily gas production with feed (F) and discharge (D) TS and VFA

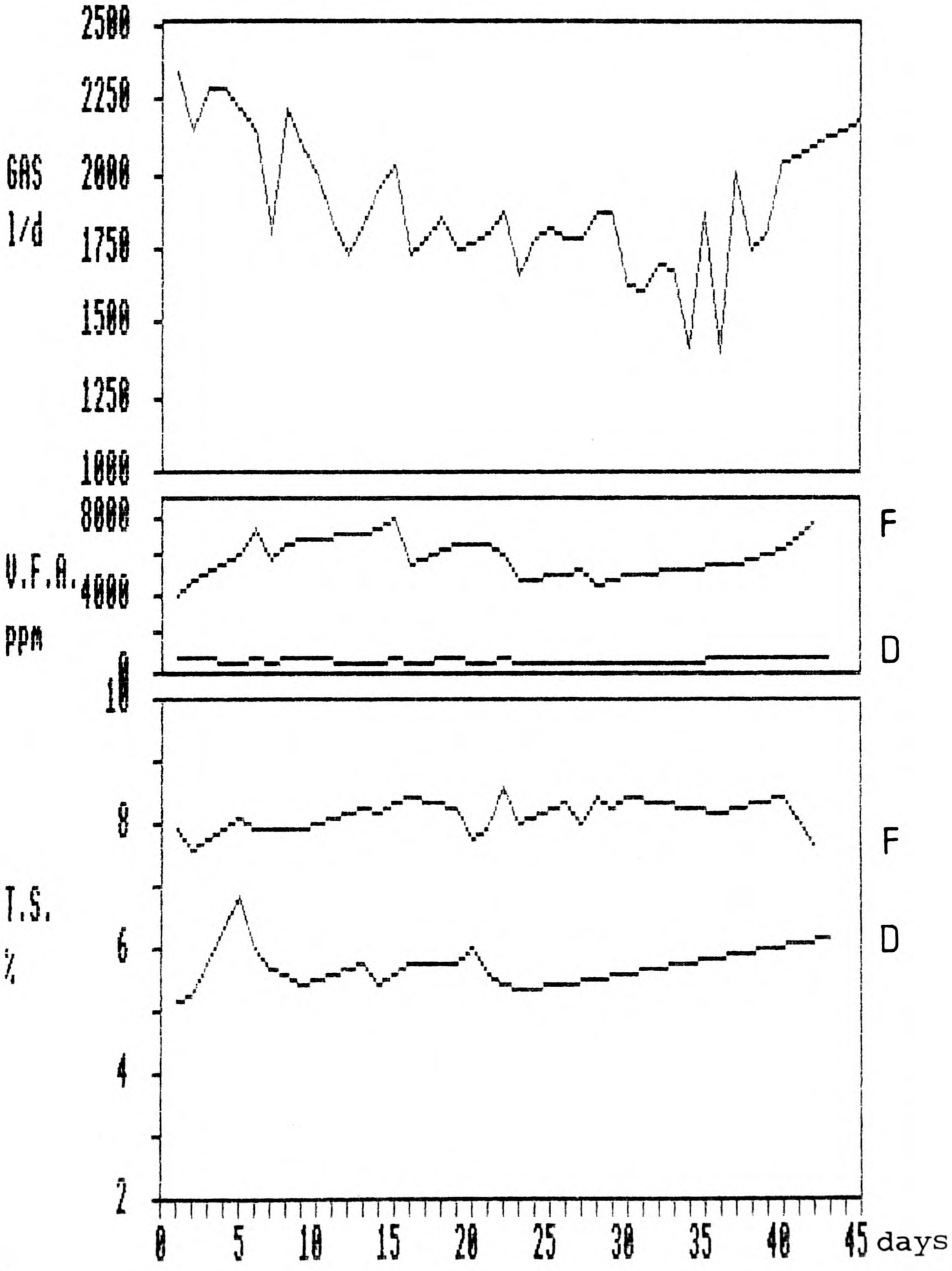


TABLE 4.2 Experimental Run 1. Summary of digester performance data.

April 26 - June 9, 1984		Digester 1	Digester 2
Aspect ratio		10	20
Temperature (SD), °C		34.61(0.40)	34.77(0.50)
Nominal RT (days)		20	20
FEED	TS, %		
	Mean (SD)	8.13 (0.24)	
	VS (% of TS)		
	Mean (SD)	83.24 (1.52)	
	VFA, ppm		
	Mean (SD)	6034 (1081)	
DISCHARGE	TS, %		
	Mean (SD)	6.07 (0.33)	5.69 (0.38)
	VS (% of TS)		
	Mean (SD)	82.59 (0.96)	82.55 (1.19)
	VFA, ppm		
	Mean (SD)	673 (94)	667 (93)
GAS YIELDS	m^3/day		
	Mean	0.911	1.906
	SD	0.1153	0.2201
	m^3/m^3 reactor / day		
	Mean	1.001	1.042
	SD	0.1267	0.1203
	m^3/ kg TS added		
	Mean	0.265	0.240
	SD	0.0245	0.0213
	m^3/ kg VS added		
	Mean	0.318	0.288
	SD	0.0244	0.0209
	m^3/ kg TS destroyed		
	Mean	1.025	0.852
	SD	0.1470	0.1309
	m^3/kg VS destroyed		
	Mean	1.200	1.004
	SD	0.1474	0.1613
REDUCTIONS	% TS (SD)	26.1 (3.35)	28.4 (3.46)
	% VS (SD)	26.8 (3.73)	29.1 (3.97)
	% VFA (SD)	88.9 (0.91)	88.9 (0.38)

Volatile solids remained a fairly constant proportion of the feed and discharge TS at 83.24% and 82.59% (D1), 82.55% (D2), respectively.

Gas production from D2 was similarly stable at 1.042m^3/m^3 reactor day (SD±0.1203). A similar failure occurred in the heating system, causing the variations in gas production between days 5-10. Gas production varied in an oscillating pattern, the wavelength being approximately 3 days, the amplitude of the oscillations being variable and not showing any trend.

The mean temperature of Digester 1 was 34.61°C (SD±0.40), and Digester 2 was 34.77°C (SD±0.50).

In Digester 1, the mean temperature was slightly lower (34.43°C, SD±0.69) in the half of the reactor closest to the lower end plate of the digester where feed was introduced, compared with the upper half of the digester (34.79°C, SD±0.25). This same pattern was observed in Digester 2 (l/d=20), where the mean temperature in the lower half was 34.67°C (SD±0.90), and in the upper half 34.88°C (SD±0.48).

The gas yields (m^3/kg TS added) give a useful indicator of digester performance. The gas yield from D1 was the greatest at 0.265m^3/kg TS added (SD±0.0245), compared with 0.240 (SD±0.0213) from D2.

The same trend is found as expected in the gas yields expressed as m^3/kg VS added, since the VS were a fairly constant proportion of the TS. The gas yield for D1 was 0.318 (SD±0.0244), and D2 was 0.288 (SD±0.0213). These values are a measure of the degree of digestion of the organic matter.

The gas yields m^3/kg TS destroyed represent the amount of gas that would be produced from the destruction of unit weight of TS. Gas yield for D1 was 1.025 (SD±0.1470), and D2 was 0.852, (SD±0.131). The slurry used was the same for both digesters. These differences are primarily due to the fluctuations in gas production from Digester 2, and the measured increase in discharge TS from Digester 2 from Day 23 onwards. These are useful indicators of the efficiency of the digestion process since they are calculated from both input and output TS values.

Similarly, the gas yields (m^3/kg VS destroyed) follow the same trend, for the same reasons, D1 being 1.200 (SD±0.1474), and D2 was 1.004 (SD±0.1613).

The % TS reduction was greater from D2 at 28.41% (SD±3.46), than D1 at 26.08% (SD±3.35). The % VS reduction was similar, being 29.06% (SD±3.97) for D2, and 26.78% (SD±3.73) for D1. These are good measures of how much the polluting power of the feedstock has been reduced.

The % VFA reduction was very similar in each digester, being 88.91% (SD$\pm$0.91) and 88.92% (SD$\pm$0.38) for Digesters 1 and 2 respectively.

4.2.2 Experimental Run 2

Each digester was inclined at 20° to the horizontal, and the aspect ratio of digester 1 was 10 and digester 2 was 20. The nominal retention time was 15 days. The feedstock was whole dairy cow slurry.

The daily gas production, and feed and discharge TS and TVFA are presented graphically in Figure 4.5 (Digester 1) and Figure 4.6 (Digester 2). This information is summarised for both digesters in Table 4.3.

The mean TS content of the feed was 8.12% (SD$\pm$1.67), and the VFA content was 4931 ppm (SD$\pm$775). The daily gas production (D1) varied between 1.07 and 1.54 m^3/m^3 reactor day, the mean value being 1.279 (SD$\pm$0.1420). Gas production varied in an oscillating manner, successive peaks occurring at intervals of approximately 17 days. Gas production from D2 was generally more stable than from D1, apart from during the Period Days 25-33. During that time, gas production oscillated widely above the mean value for the run of 1.253 m^3/m^3 reactor day

FIGURE 4.5 Experimental Run 2. Digestion of whole slurry in Digester 1.

Daily gas production with feed (F) and discharge (D) TS and VFA

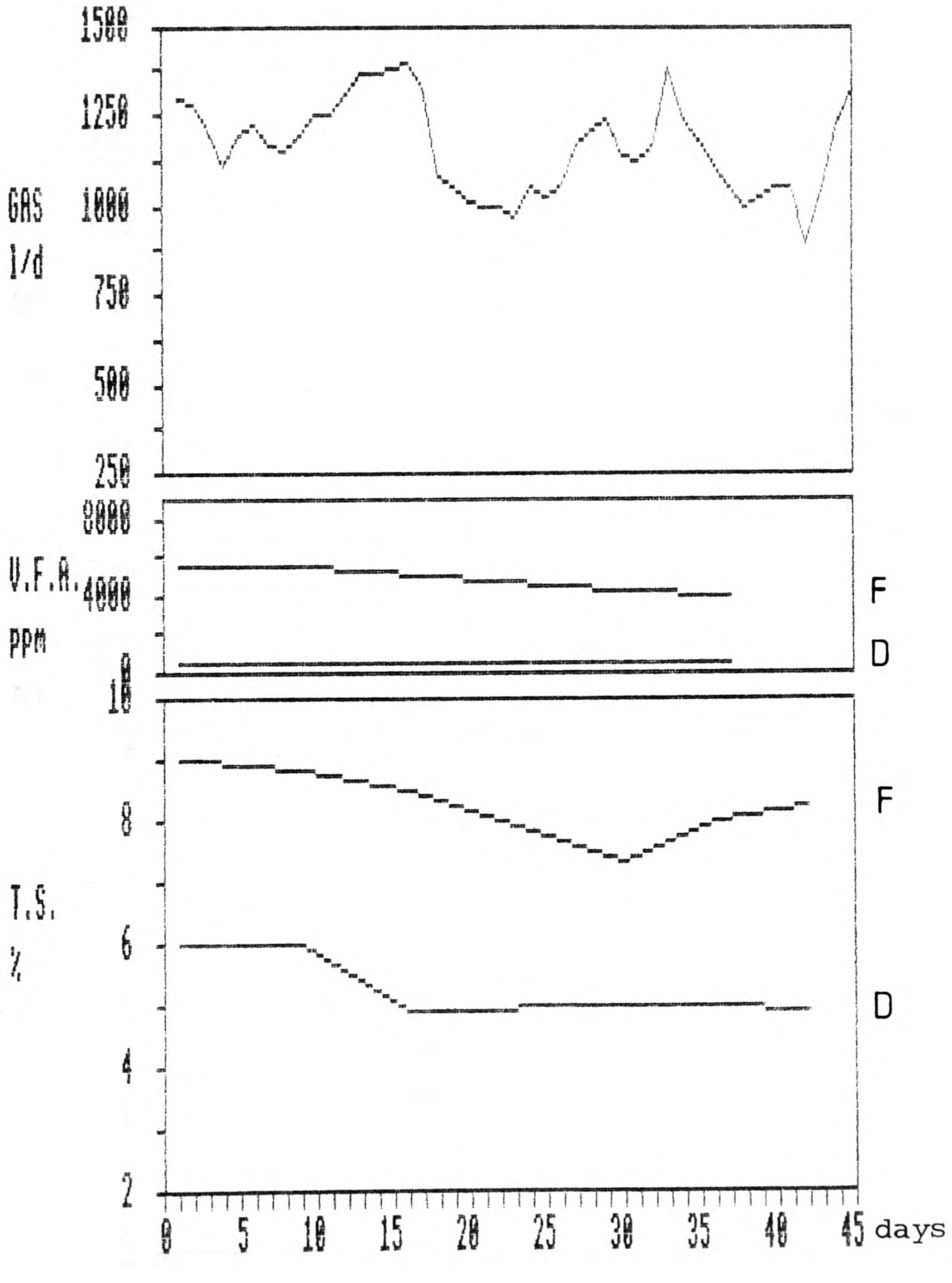


FIGURE 4.6 Experimental Run 2. Digestion of whole slurry in Digester 2.

Daily gas production with feed (F) and discharge (D) TS and VFA

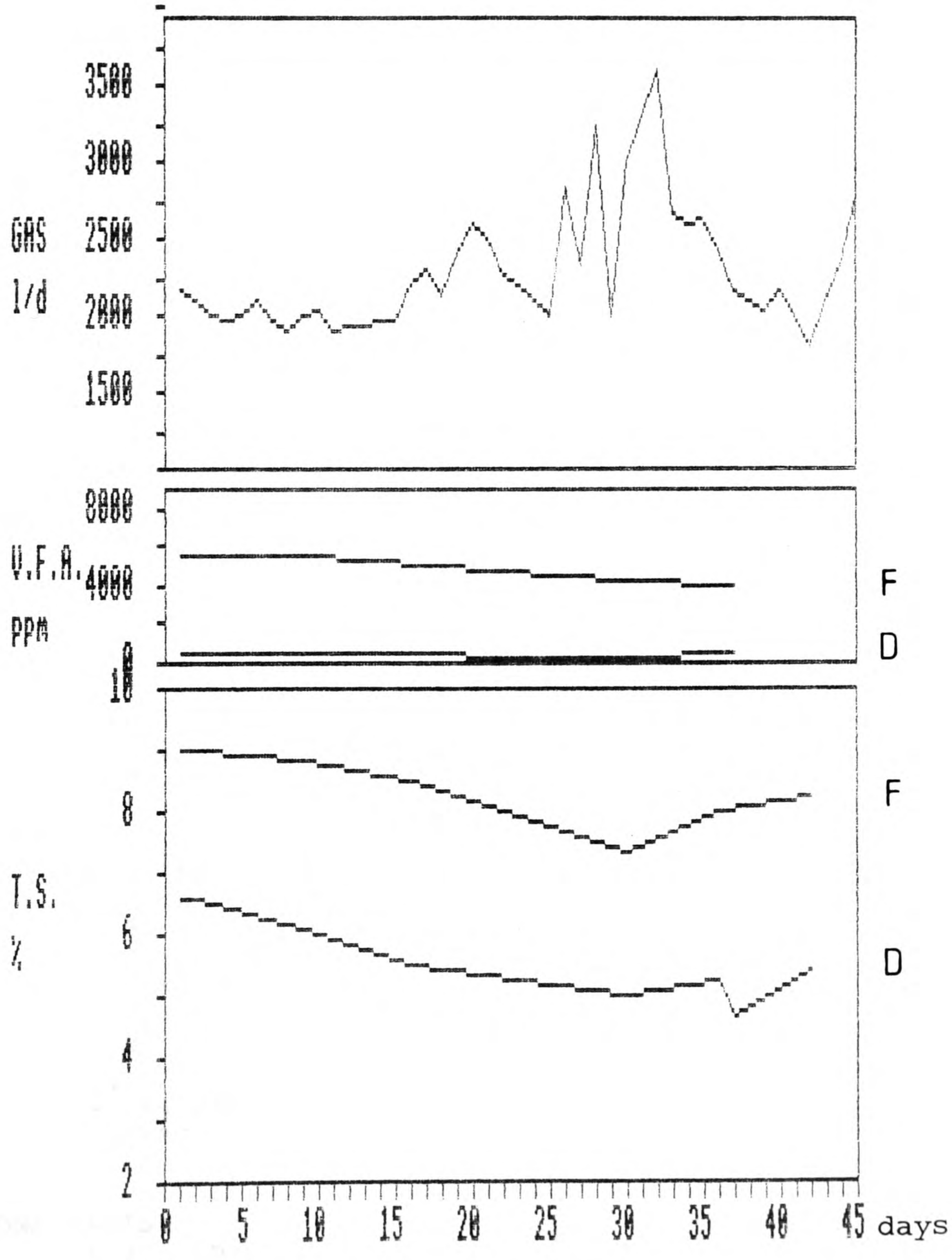


TABLE 4.3 Experimental Run 2. Summary of digester performance data.

July 18 - Aug 31, 1984		Digester 1	Digester 2
Aspect ratio		10	20
Temperature (SD), ^{o}C		34.7(0.40)	35.1(0.50)
Nominal RT (days)		15	15
FEED	TS, %		
	Mean (SD)	8.12 (1.67)	
	VS (% of TS)		
	Mean (SD)	80.91 (3.11)	
	VFA, ppm		
	Mean (SD)	4931 (775)	
DISCHARGE	TS, %		
	Mean (SD)	5.42 (0.64)	5.64 (0.66)
	VS (% of TS)		
	Mean (SD)	80.69 (0.80)	80.95 (0.91)
	VFA, ppm		
	Mean (SD)	578 (95)	469 (56)
GAS YIELDS	m^3/day		
	Mean	1.164	2.293
	SD	0.1292	0.4032
	m^3/m^3 reactor / day		
	Mean	1.279	1.253
	SD	0.1420	0.2203
	m^3/ kg TS added		
	Mean	0.245	0.226
	SD	0.0108	0.0441
	m^3/ kg VS added		
	Mean	0.308	0.282
	SD	0.0138	0.0423
	m^3/ kg TS destroyed		
	Mean	0.719	0.723
	SD	0.0725	0.1196
	m^3/kg VS destroyed		
	Mean	0.956	0.968
	SD	0.2155	0.1096
REDUCTIONS	% TS (SD)	34.2 (2.60)	31.1 (1.95)
	% VS (SD)	33.1 (5.95)	29.6 (6.59)
	% VFA (SD)	87.9 (3.09)	89.9 (2.57)

(SD±0.2203), and reached a maximum of 1.98 m^3/m^3 reactor/day.

Figures 4.5 and 4.6 show the general trend of discharge TS decreasing as gas production increased, showing that digesting solids are being converted to gas.

Levels of VFA in the discharge remained relatively constant at 578 ppm (SD±95) for D1, and 469 ppm (SD±56) for D2.

Gas yields (m^3/kg TS added, m^3/kg VS added) were greater for D1 than D2. The gas yield of 0.719 m^3/kg TS destroyed (SD±0.0725) for D1 was very similar to the value for D2 of 0.723 (SD±0.1196). Similarly, the gas yields per kg VS destroyed were very similar, the value for D1 being 0.956 (SD±0.2155), and D2 was 0.968 (SD±0.1096).

The %TS and %VS reduction were greater for D1 than D2.

The %VFA reduction for D2 (89.89%, SD±2.57) was very slightly greater than for D1 which was 87.89%, SD±3.09.

The mean digester temperature for D1 during the run was 34.7°C, SD±0.55. Temperature fluctuations in D2 were very similar, the mean digester temperature being 35.3°C, SD±0.60.

4.2.3 Experimental Run 3

Each digester was inclined at 20^o to the horizontal, and the aspect ratios of Digester 1 and Digester 2 were 10 and 20 respectively, the target operating temperature of each digester being 25^oC. The nominal retention time was 15 days. The feedstock was whole dairy cow slurry.

The daily gas production, and feed and discharge TS and TVFA are presented graphically in Figure 4.7 (Digester 1) and Figure 4.8 (Digester 2). This information is summarised for both digesters in Table 4.4.

From Figure 4.7, it can be seen that there is a strong correlation between daily gas production, VFA fed, and TS fed and discharged. The correlation between gas production and TS fed appears stronger than that between gas production and VFA fed.

The mean daily gas production from each Digester was relatively stable, being $0.706m^3/m^3$ reactor / day (SD±0.0687) for D1, and 0.703 m^3/m^3 reactor / day (SD±0.0867) for D2.

The mean TS, VS and VFA added were similar to those in previous runs at 7.66% (SD±0.83), 85.13% (SD±1.98) and 4809 ppm (SD±917) respectively.

FIGURE 4.7 Experimental Run 3. Digestion of whole slurry in Digester 1.

Daily gas production with feed (F) and discharge (D) TS and VFA

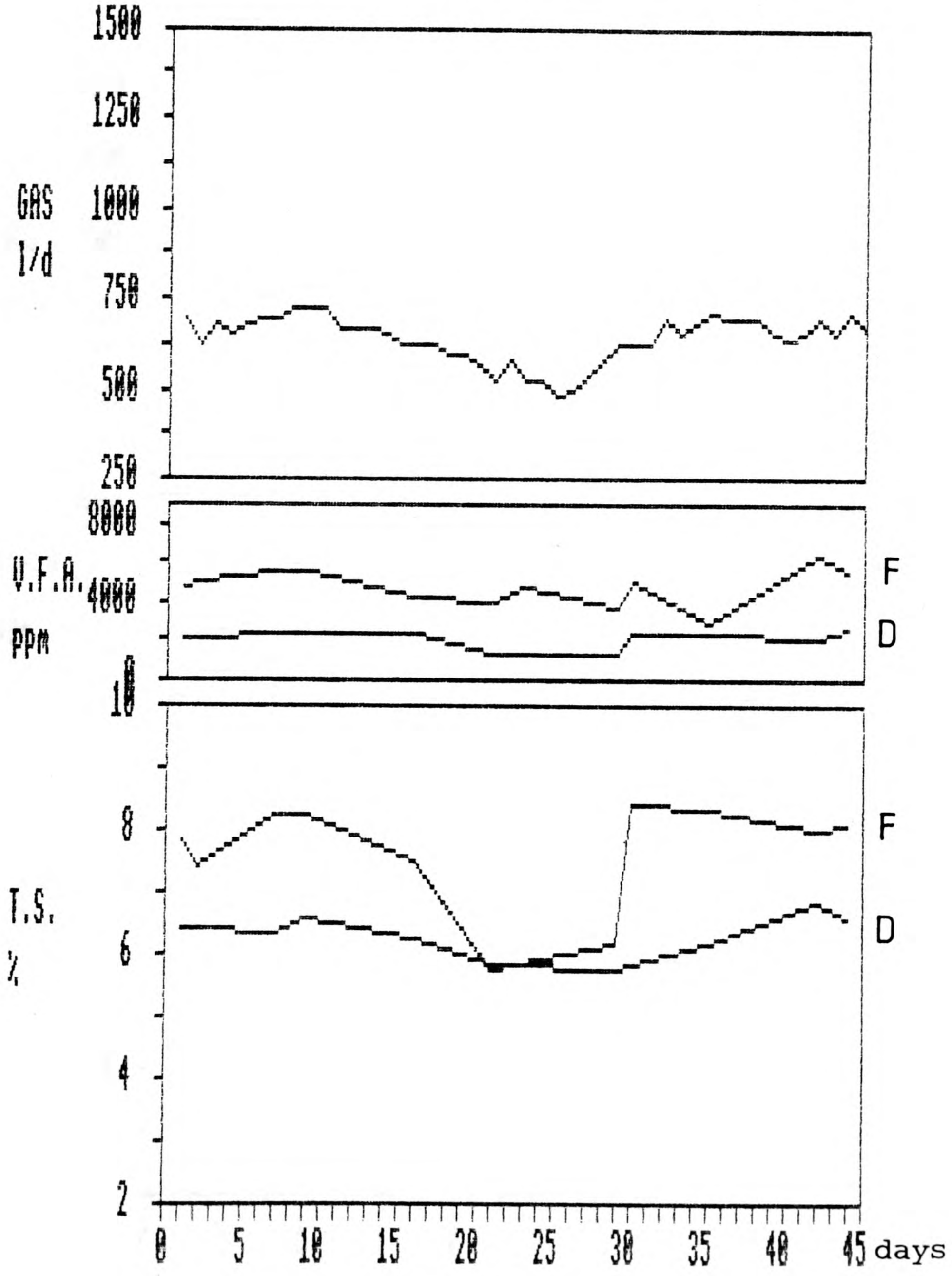


FIGURE 4.8 Experimental Run 3. Digestion of whole slurry in Digester 2.

Daily gas production with feed (F) and discharge (D) TS and VFA

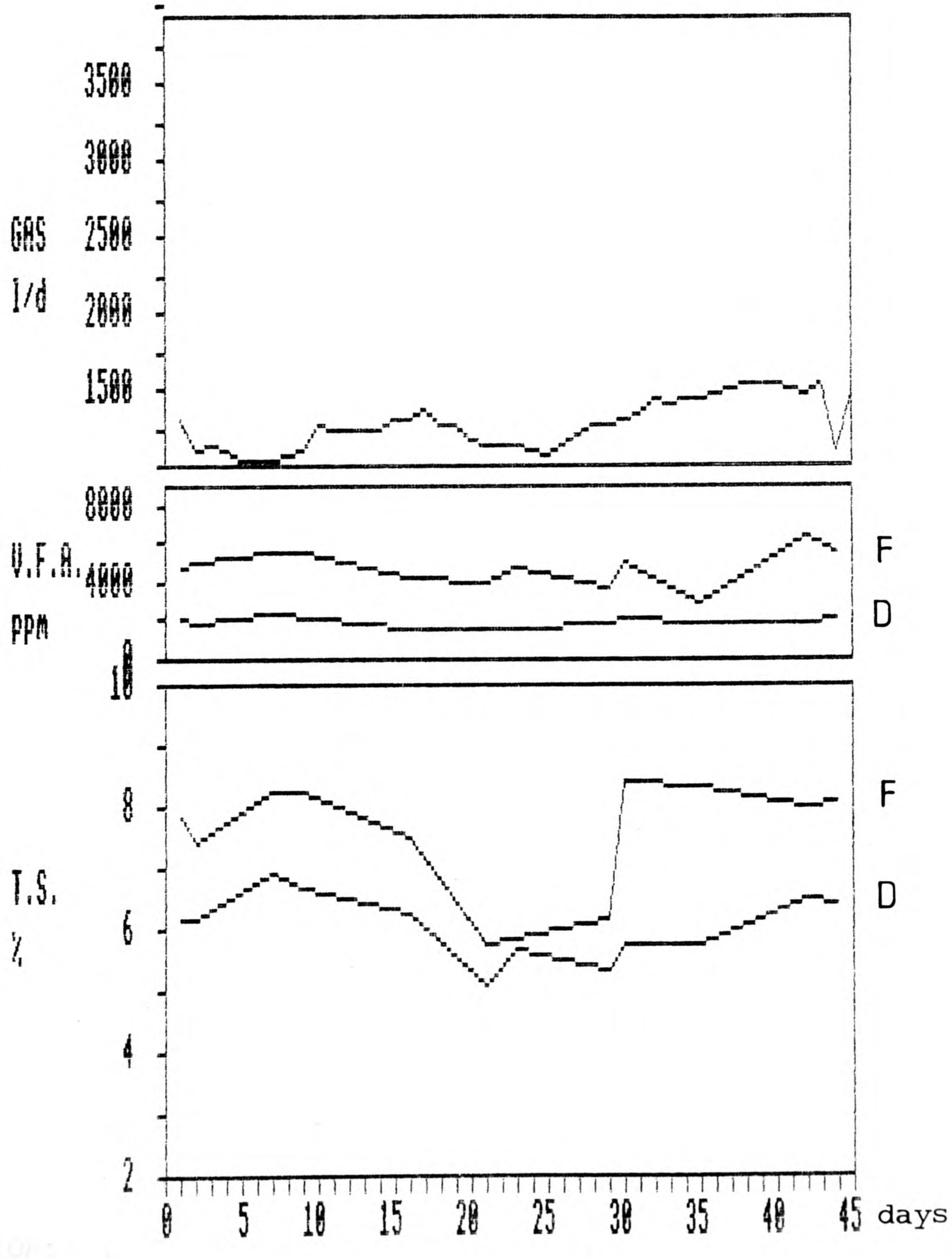


TABLE 4.4 Experimental Run 3. Summary of digester performance data.

Oct 10 - Nov 23, 1984		Digester 1	Digester 2
Aspect ratio		10	20
Temperature, °C (Nominal)		25	25
Nominal RT (days)		15	15
FEED	TS, %		
	Mean (SD)	7.66 (0.83)	
	VS (% of TS)		
	Mean (SD)	85.13 (1.98)	
	VFA, ppm		
	Mean (SD)	4809 (917)	
DISCHARGE	TS, %		
	Mean (SD)	6.29 (0.34)	6.06 (0.52)
	VS (% of TS)		
	Mean (SD)	83.87 (0.85)	83.46 (0.74)
	VFA, ppm		
	Mean (SD)	2093 (444)	1938 (263)
GAS YIELDS	m^3/day		
	Mean	0.642	1.287
	SD	0.0625	0.1587
	m^3/m^3 reactor / day		
	Mean	0.706	0.703
	SD	0.0687	0.0867
	m^3/ kg TS added		
	Mean	0.133	0.130
	SD	0.0049	0.0137
	m^3/ kg VS added		
	Mean	0.156	0.153
	SD	0.0065	0.0162
	m^3/ kg TS destroyed		
	Mean	0.720	0.631
	SD	0.0360	0.0373
	m^3/kg VS destroyed		
	Mean	0.796	0.694
	SD	0.0255	0.0429
REDUCTIONS	% TS (SD)	18.4 (1.34)	20.7 (3.35)
	% VS (SD)	19.6 (1.17)	22.1 (3.51)
	% VFA (SD)	56.3 (3.01)	59.5 (1.70)

There were relatively small reductions in TS and VFA during this run. The mean TS was 6.29% (SD±0.34) for D1 and 6.06% (SD±0.52) for D2. The mean VFA in the discharge was 2093 ppm (SD±444) for D1, and 1938 ppm (SD±263) for D2.

The gas yields from each digester were very similar at 0.132m^3/kg TS added (SD±0.0049) for D1, and 0.130 m^3/kg TS added for D2. Similarly, gas yields per unit weight of VS added were 0.156 (SD±0.0065) and 0.152 (SD±0.0162) for Digesters 1 and 2 respectively.

The gas yields were greater in D1 than D2, averaging 0.720 (SD±0.0360) and 0.631 (SD±0.0373) for Digesters 1 and 2 respectively. Similarly, gas yields per unit weight of VS destroyed were 0.796 (SD±0.0255) and 0.694 (SD±0.0429) for Digesters 1 and 2 respectively. These gas yields should theoretically be the same. The differences here can probably be accounted for by the higher mean discharge TS from Digester 1 compared with Digester 2. The greater standard deviation of the TS discharge from Digester 2 accentuates these differences.

The %TS and %VS reductions were similar from Digesters 1 and 2.

The %VFA reduction in each digester was very similar at 56 and 59% for Digesters 1 and 2, respectively.

4.2.4 Experimental Run 4

Each digester was included at 10^{o} to the horizontal (all previous runs were at 20^{o}), and the aspect ratios of Digesters 1 and 2 were 10 and 20 respectively. The nominal retention time was 20 days and target operating temperature was $35^{o}C$. The feedstock was whole dairy cow slurry.

The daily gas production, and feed and discharge TVFA are presented graphically in Figure 4.9 (Digester 1) and Figure 4.10 (Digester 2). This information is summarised for both digesters in Table 4.5.

Figure 4.2 shows that during Period 8, daily gas production from Digester 2 was fluctuating widely. It was not possible to identify a period of 45 days when the daily gas production from both digesters was stable, so two different but overlapping periods are used to evaluate performance. These are shown in Figure 4.2.

The mean daily feed TS, VS and VFA values were very similar for both the periods in this Experimental Run.

The mean discharge %TS was greater for Digester 1 (5.94%, SD±1.22) than Digester 2 (4.89%, SD±0.77).

The mean daily TVFA in the discharge was 1367 ppm (SD±257) for Digester 1, and 2725 ppm (SD±734) for

FIGURE 4.9 Experimental Run 4. Digestion of whole slurry in Digester 1.

Daily gas production with feed (F) and discharge (D) TS and VFA

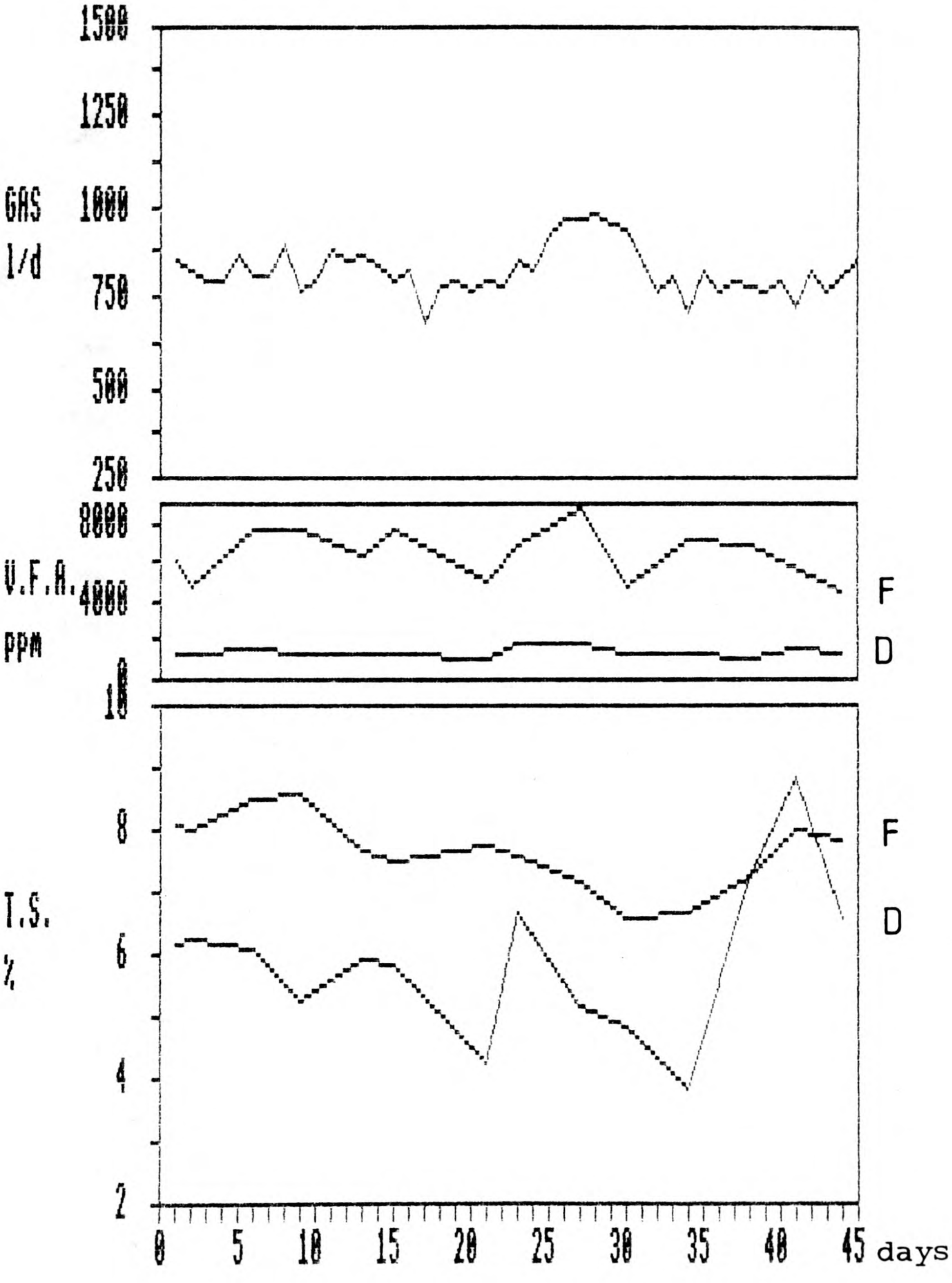


FIGURE 4.10 Experimental Run 4. Digestion of whole slurry in Digester 2.

Daily gas production with feed (F) and discharge (D) TS and VFA

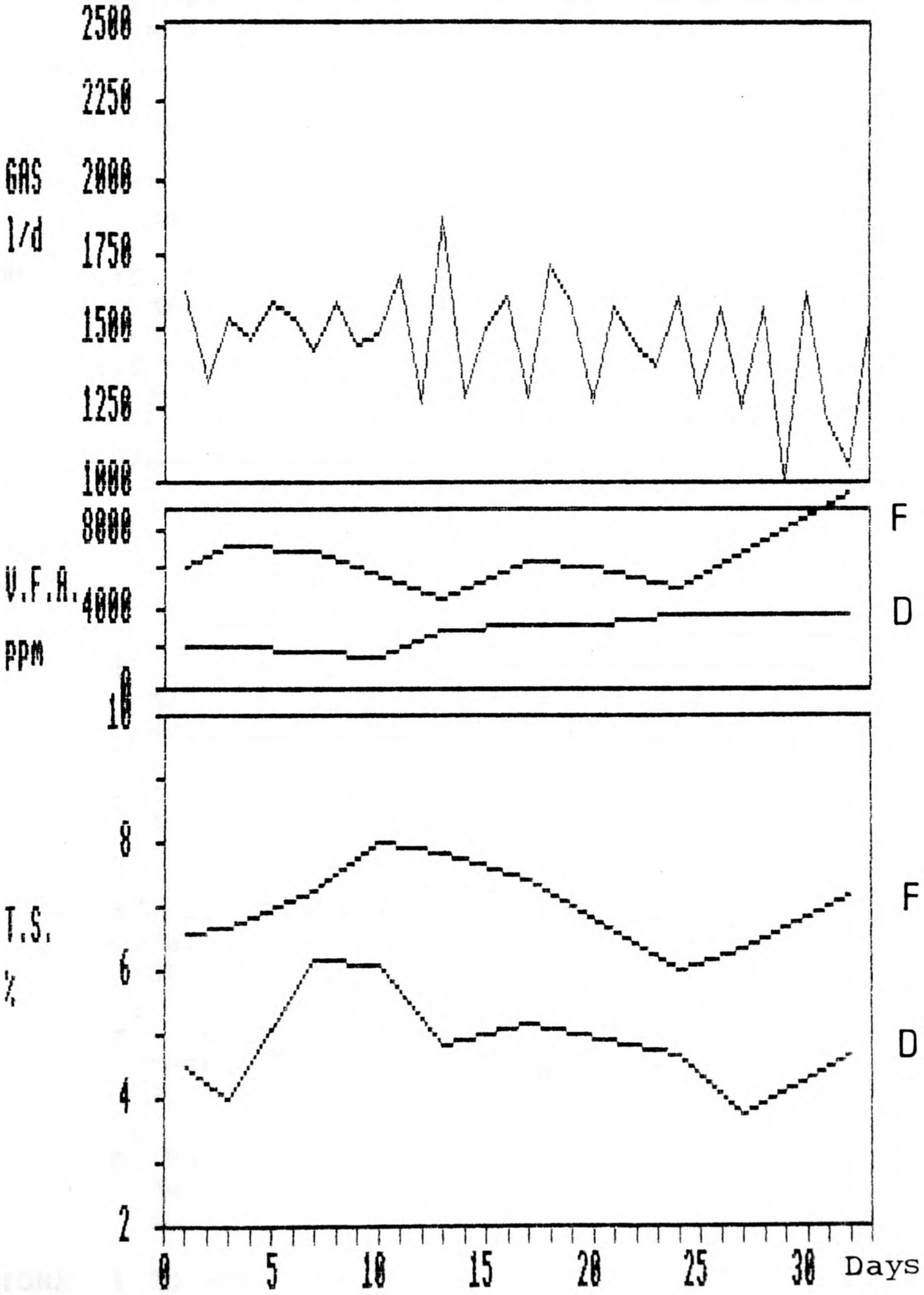


TABLE 4.5 Experimental Run 4. Summary of digester performance data.

		Digester 1 8/5/-21/6/85	Digester 2 8/6-10/7/85
Aspect ratio		10	20
Temperature (SD), °C		33.26(0.76)	36.20(0.83)
Nominal RT (days)		20	20
FEED	TS, %		
	Mean (SD)	7.66 (0.58)	7.03 (0.62)
	VS (% of TS)		
	Mean (SD)	81.02 (1.61)	79.86 (1.53)
	VFA, ppm		
	Mean (SD)	6513 (1327)	6459 (1490)
DISCHARGE	TS, %		
	Mean (SD)	5.94 (1.22)	4.89 (0.77)
	VS (% of TS)		
	Mean (SD)	79.00 (3.64)	75.73 (4.57)
	VFA, ppm		
	Mean (SD)	1367 (257)	2725 (734)
GAS YIELDS	m^3/day		
	Mean	0.826	1.466
	SD	0.0644	0.1896
	m^3/m^3 reactor / day		
	Mean	0.907	0.801
	SD	0.0730	0.1036
	m^3/ kg TS added		
	Mean	0.228	0.228
	SD	0.0181	0.0088
	m^3/ kg VS added		
	Mean	0.282	0.286
	SD	0.0236	0.0119
	m^3/ kg TS destroyed		
	Mean	0.827	0.753
	SD	0.0623	0.1312
	m^3/kg VS destroyed		
	Mean	0.958	0.848
	SD	0.0670	0.1518
REDUCTIONS	% TS (SD)	27.6 (0.31)	30.8 (4.39)
	% VS (SD)	29.4 (0.73)	34.4 (5.60)
	% VFA (SD)	78.6 (2.37)	56.1 (15.05)

Digester 2. These values are both very high. The VFA level for Digester 2 of 2725 ppm was exceedingly high and fluctuating widely as indicated by the SD of 734 ppm.

The gas yield (m^3/m^3 reactor day) of D1 was 0.907 (SD±0.0730), compared with 0.801 (SD±0.1036) for D2.

Gas yields per unit weight of TS and VS added were, however, the same for both digesters.

Gas yields per unit weight of TS and VS destroyed were greater for D1 than D2. However, the standard deviations of these parameters for D2 were approximately twice those of D1. Digester 2 was particularly unstable during Period 8 (Fig. 4.2) which included this Experimental Run. Different, but overlapping periods in time were used to evaluate the performance of Digesters 1 and 2. Tracer studies (Section 4.4) conducted after Experimental Run 4 for Digester 1 showed there to be considerable accumulation of solids in both digesters. The raised levels of VFA also suggest that short-circuiting within the digester is taking place. The gas yields were evaluated using a nominal retention time of 20 days, although the actual retention time was probably considerably less.

The low %VFA reduction from Digester 2 (56.08%, SD±15.05) compared with Digester 1 (78.56%, SD±2.37) is due to the level of VFA in the discharge from Digester 2 being very much greater than from Digester 1.

Both digesters were extensively sampled on Day 26 of Experimental Run 4 (Figure 4.10), and analysed for the individual volatile fatty acids. This was after the period of Experimental Run 4 for Digester 1. Figure 4.11 shows the concentration of acetic and propionic acids at different points along the reactor for Digester 1. The concentration of acetic was greter than propionic acid at all times for samples taken from the sampling valves on the side of the digester. Samples taken from the sampling valves on the base of Digester 1 showed the same trends, but absolute values were less. This suggests there was less metabolic activity along the base of the digester, compared with the digester region at and above the side sampling valves.

Figure 4.12 shows acetic and propionic acid concentration profiles for Digester 2, samples being taken at the same time as above. For the first 5m of digester length, the concentration of acetic was greater than propionic acid (samples from side sampling valves). However, in the top 5m of digester length, the concentration of propionic was greater than acetic acid. For the samples taken from the sampling valves on the base of Digester 2, the concentration of propionic acid was greater than acetic at 5 of the 6 sampling positions.

FIGURE 4.11 Concentration of acetic and propionic acids in Digester 1.

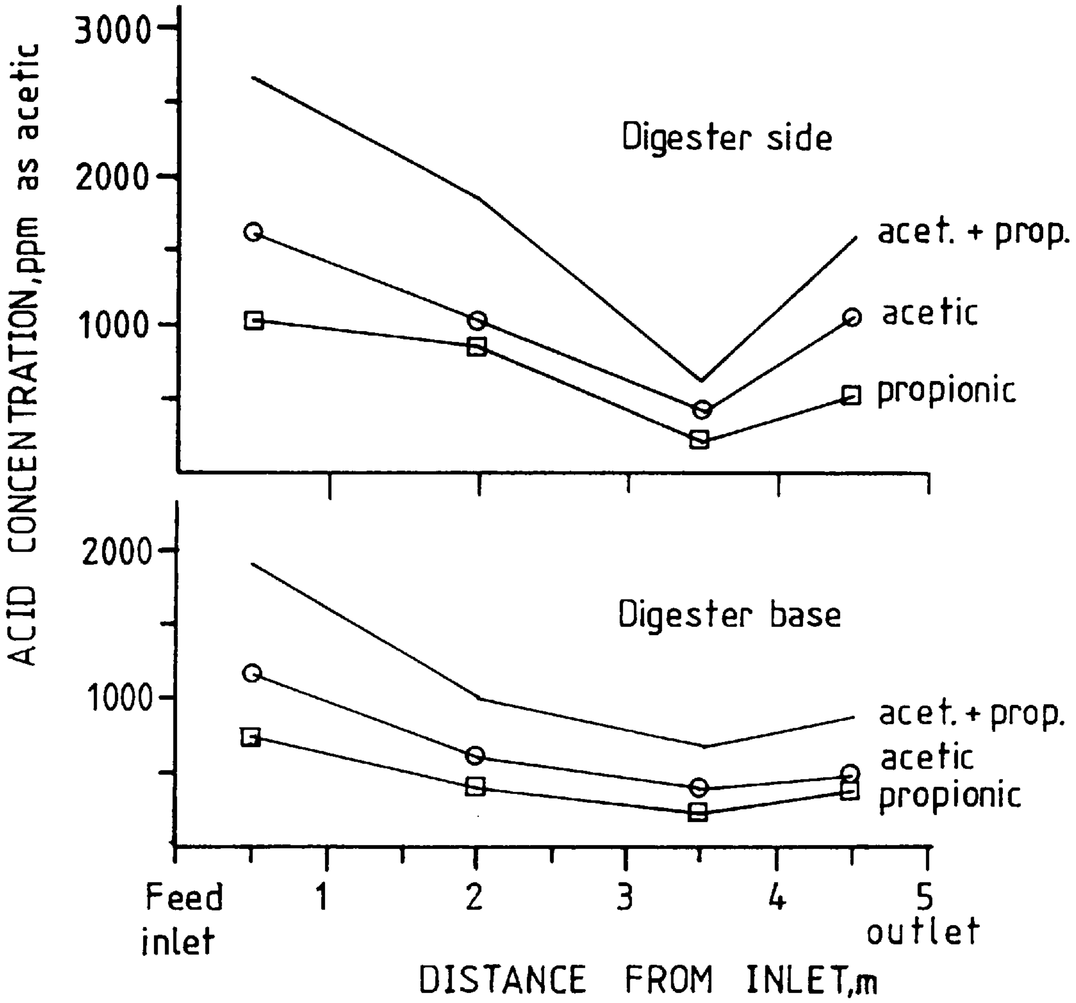
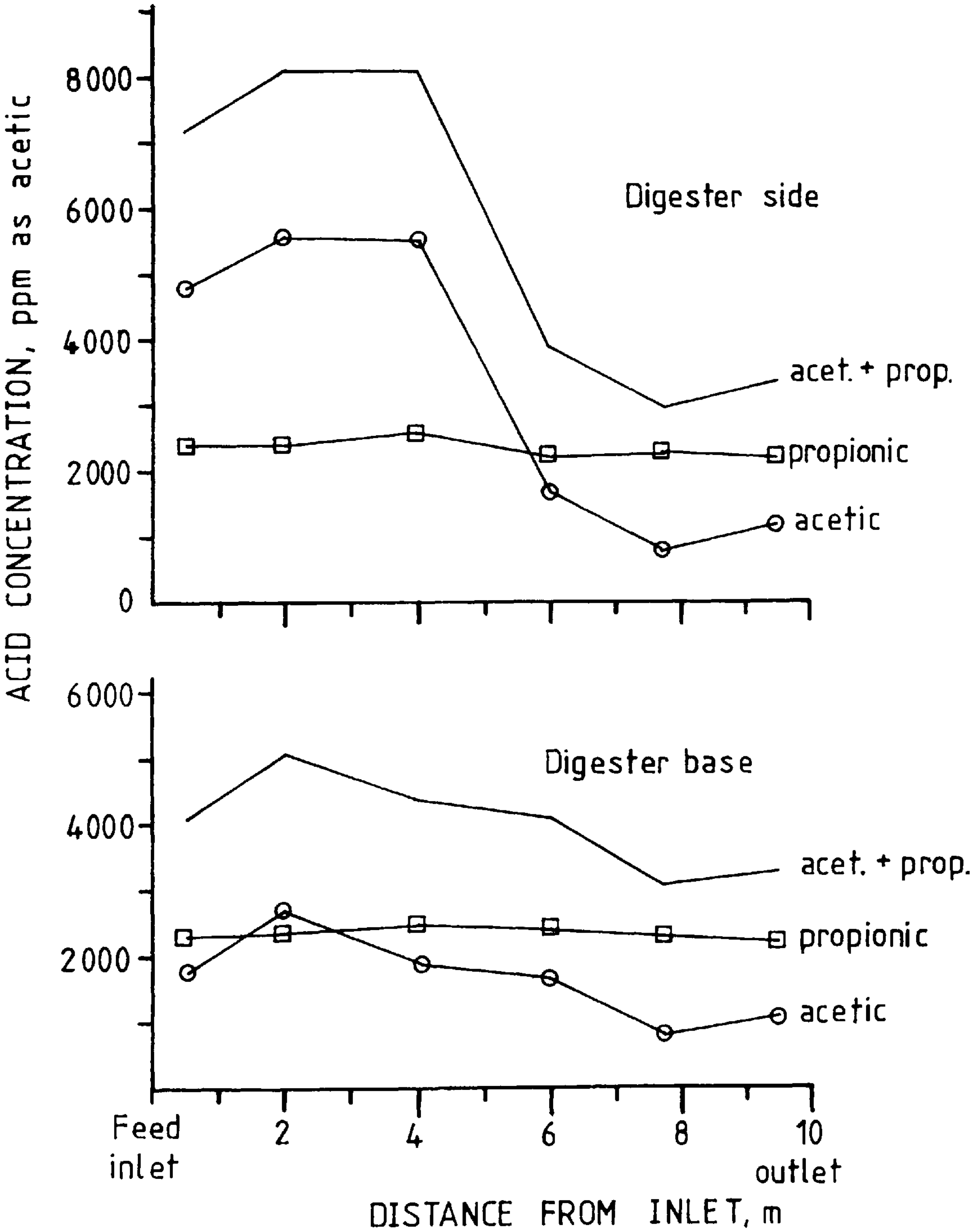


FIGURE 4.12 Concentration of acetic and propionic acids in Digester 2.



Partial failure of methanogenesis from raised hydrogen concentrations can raise the proportion of propionic acid in the VFA (Hobson *et al*, 1981). This increased concentration of propionic acid inhibits methanogenesis further, and the inhibition can become, in effect, "autocatalytic" (Hobson *et al*, 1974).

4.2.5 Experimental Run 5

Each digester was inclined at 10° to the horizontal, and the aspect ratios of Digester 1 and Digester 2 were 10 and 20 respectively, the target operating temperature of each digester being 35°C. The nominal retention time was 15 days, and the feedstock was separated dairy cow slurry.

The daily gas production, and feed and discharge TVFA are presented graphically in Figure 4.13 (Digester 1) and Figure 4.14 (Digester 2). This information is summarised for both digesters in Table 4.6.

In Figures 4.13 and 4.14, gas production can be seen to increase with increased levels of VFA and TS fed. The TS discharged was reduced in the periods following the peaks in gas production.

FIGURE 4.13 Experimental Run 5. Digestion of separated slurry in Digester 1.

Daily gas production with feed (F) and and discharge (D) TS and VFA

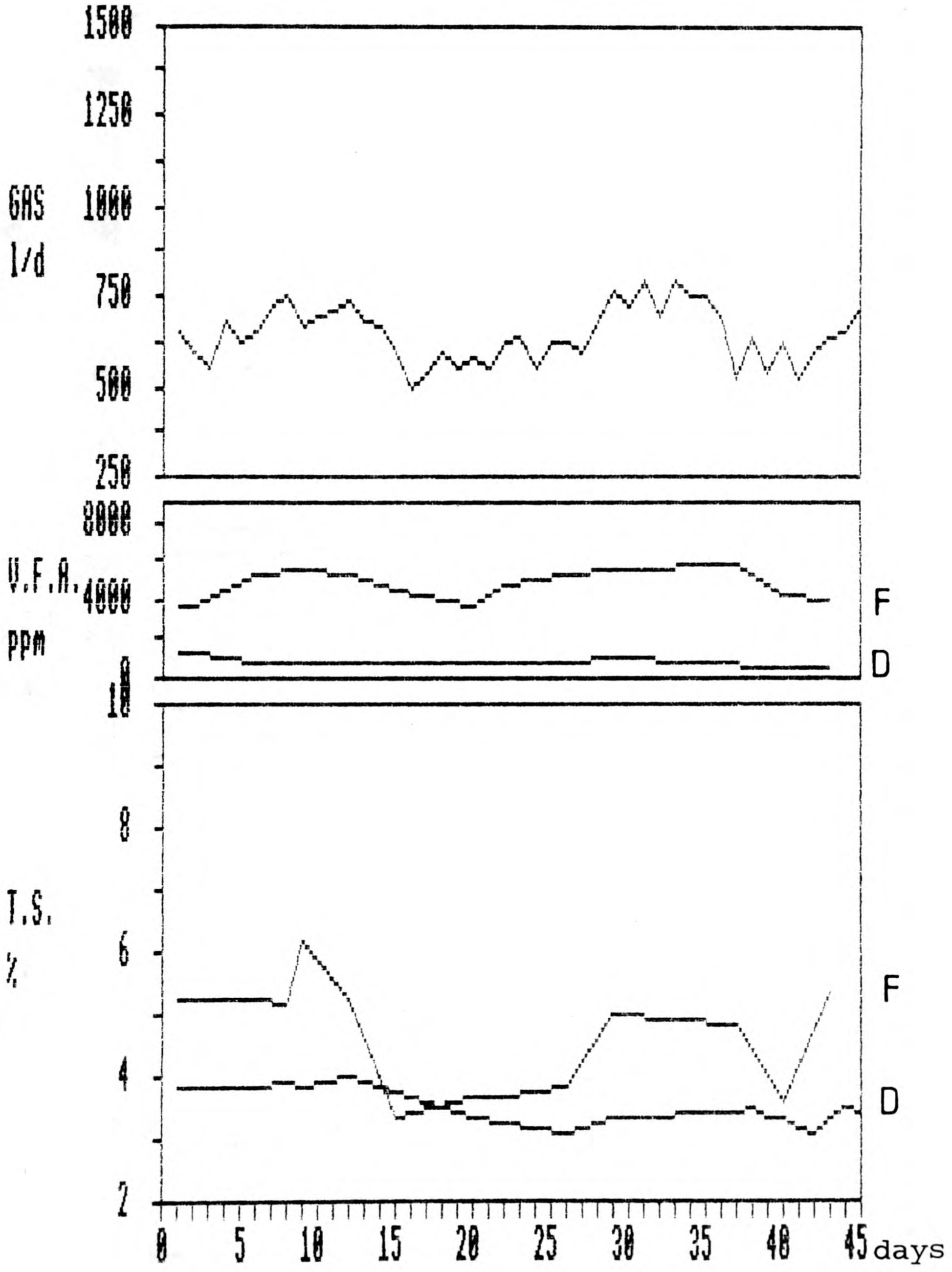


FIGURE 4.14 Experimental Run 5. Digestion of separated slurry in Digester 2.

Daily gas production with feed (F) and discharge (D) TS and VFA

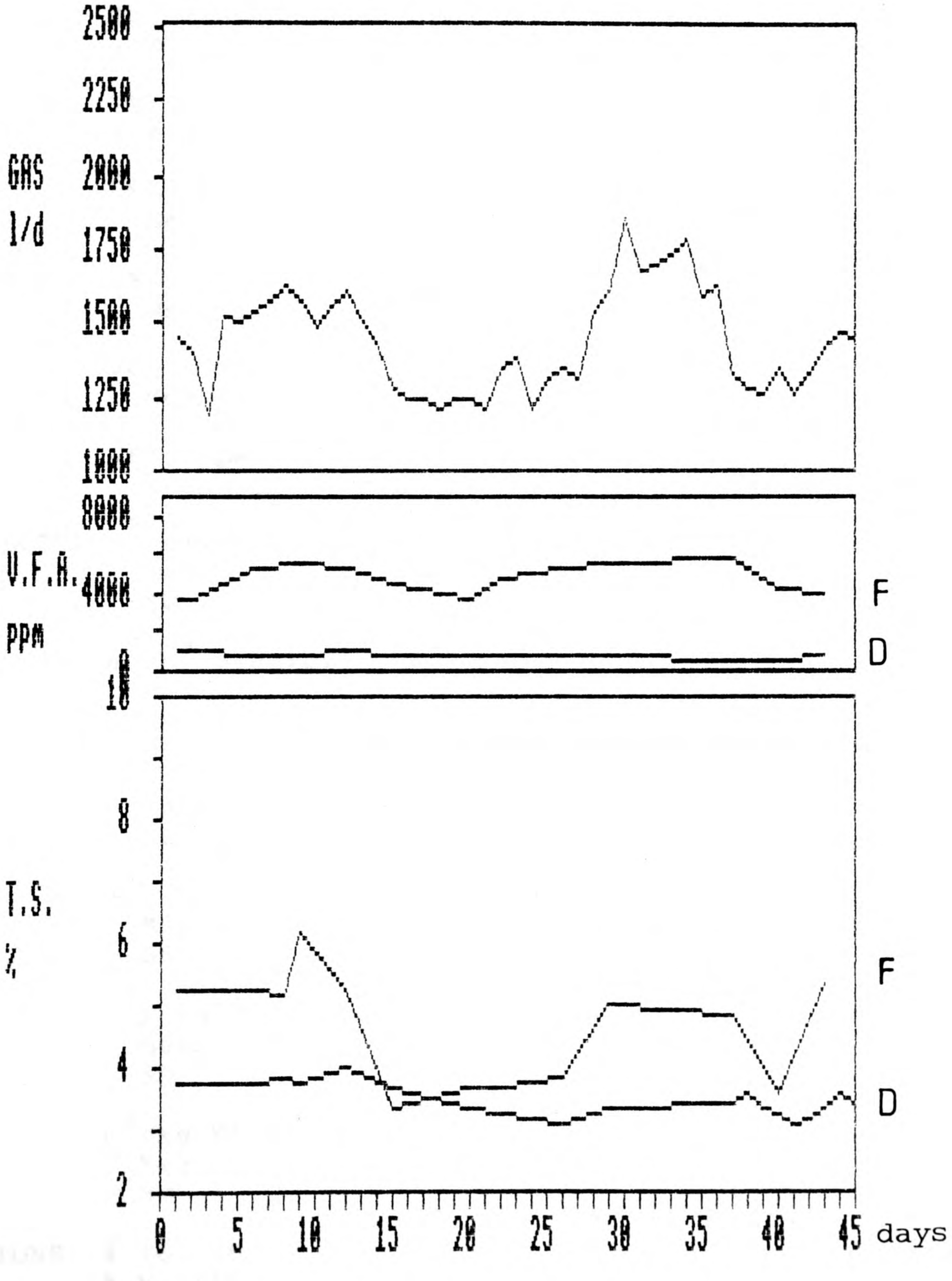


TABLE 4.6 Experimental Run 5. Summary of digester performance data.

Sep 19 - Nov 2, 1985		Digester 1	Digester 2
Aspect ratio		10	20
Temperature (SD), °C		34.57(0.26)	34.41(0.34
Nominal RT (days)		15	15
FEED	TS, %		
	Mean (SD)	4.69 (0.83)	
	VS (% of TS)		
	Mean (SD)	77.06 (1.79)	
	VFA, ppm		
	Mean (SD)	4749 (739)	
DISCHARGE	TS, %		
	Mean (SD)	3.52 (0.28)	3.49 (0.26)
	VS (% of TS)		
	Mean (SD)	70.67 (1.65)	70.39 (1.92)
	VFA, ppm		
	Mean (SD)	843 (199)	780 (188)
GAS YIELDS	m^3/day		
	Mean	0.649	1.444
	SD	0.0766	0.1710
	m^3/m^3 reactor / day		
	Mean	0.713	0.789
	SD	0.0842	0.0934
	m^3/ kg TS added		
	Mean	0.226	0.244
	SD	0.0196	0.0170
	m^3/ kg VS added		
	Mean	0.293	0.316
	SD	0.0274	0.0243
	m^3/ kg TS destroyed		
	Mean	0.981	1.043
	SD	0.2088	0.2416
	m^3/kg VS destroyed		
	Mean	0.997	1.074
	SD	0.1936	0.2585
REDUCTIONS	% TS (SD)	23.5 (3.58)	23.9 (3.95)
	% VS (SD)	29.9 (3.59)	30.2 (5.05)
	% VFA (SD)	84.2 (3.76)	84.6 (4.13)

The separation process had obviously reduced the mean feed %TS, the value for this run being 4.69% (SD±0.83). The mean values of %VS and VFA at 77.06% (SD±1.79), and 4749 ppm (SD±739), respectively, are also lower due to the separation process.

The mean discharge %TS, %VS and VFA were very similar from each digester.

The gas yield (m^3/m^3 reactor day) was greater from D2 at 0.789 (SD±0.0934), compared with D1 at 0.713 (SD±0.0842).

The gas yields from D2 were marginally but consistently greater from D2, at 0.244m^3/kg TS added (SD±0.0170), compared with 0.226 m^3/kg TS added (SD±0.0196) from D1. The same trend was observed in the gas yields per unit weight of VS added.

The gas yields per unit weight of TS and VS destroyed were similar from both digesters. Gas yields per unit weight of TS destroyed were 0.981 (SD±0.2088) and 1.043 (SD±0.2416) for Digesters 1 and 2 respectively. Similarly gas yields per unit weight of VS destroyed were 0.997 (SD±0.1936) and 1.074 (SD±0.2585).

The %TS, %VS and %VFA reductions were very similar for both digesters.

Digester performance during this run was very stable as shown by Figures 4.13 and 4.14.

The mean digester temperature for D1 during the run was 34.57°C (SD±0.26), and for D2 it was 34.41°C (SD±0.34).

4.3 Energy production

The gross energy production (measured) from each digester (kWh/d) was calculated from the gross biogas production (m^3) multiplied by the mean calorific value of the biogas (MJ/m^3) at s.t.p. (15.55°C, 1.016 bar).

The $\%CO_2$ of the biogas was measured weekly using draeger tubes, and it was assumed that methane made up the remainder of the gas volume

i.e. $\% \ CH_4 = 100 - \% \ CO_2$

The gross calorific value of methane at STP is 36.95 MJ/m^3 (Perry, 1963). The gross calorific value (MJ/m^3) of biogas was determined by multiplying $\% \ CH_4$ by 36.95.

Table 4.7 shows mean values per experimental run of gas composition, feed and bulk digester temperatures. Gas production and the electricity used for digester heating during experimental runs are also presented. The

TABLE 4.7 Energy production and usage during Experimental Runs 1 - 5.

Experimental run	1		2		3		4		5	
Digester	1	2	1	2	1	2	1	2	1	2
% methane	-	-	59	60	60	59	61	58	64	65
(SD)	-	-	3.2	3.5	4.7	3.6	4.3	1.9	1.3	3.4
hydrogen sulphide, ppm	-	-	985	900	1860	1460	1300	1450	2700	2643
(SD)	-	-	212	300	586	461	173	122	827	759
Feed temp., ^{o}C	10.3	10.3	17.9	17.9	-	-	13.1	15.4	12.4	12.4
(SD)	2.42	2.42	2.62	2.62	-	-	2.5	3.4	3.70	3.70
Digester temp., ^{o}C	34.6	34.8	34.7	35.3	25	25	33.3	36.2	34.6	34.4
GAS PRODUCTION										
Total for run, m^3	40.98	85.79	52.36	103.19	28.9	57.92	37.15	48.38	29.21	64.98
Mean daily value, l	911	1906	1164	2293	642	1287	826	1460	649	1444
ELECTRICITY FOR DIGESTER HEATING										
Total for run, kWh	867	1086	491	640	524	725	742	676	729	973
Mean daily value kWh	19.3	24.13	10.91	14.2	11.6	16.1	16.5	20.5	16.2	21.6
Dig. 2 / Dig 1. heat requirement	1.25		1.30		1.38		1.24		1.33	

digester feed and bulk temperatures were logged every 6 hours, and the values presented are daily mean values. Similarly, gas production was recorded daily and both the total gas production during each experimental run, and the mean daily value is presented.

The average methane content in the biogas was relatively constant at 59-61% during experimental runs 2, 3 and 4. During experimental run 5 it averaged 64 and 65% for digesters 1 and 2 respectively. Variations in the hydrogen sulphide concentration were more pronounced, the average values being 900-1860 ppm in experimental runs 2,3 and 4. In experimental run 5, average values were 2700 and 2643 ppm for digesters 1 and 2 respectively.

The digester configuration was always such that the aspect ratio, volume and surface area of Digester 2 was twice that of Digester 1. Angle of inclination and retention time were the same for each digester. It was to be expected therefore that the electrical energy used for heating Digester 2 would be twice that of Digester 1. Table 4.7 shows that this is clearly not the case, Digester 2 heating requirement ranging from 1.24-1.38 times greater than that of Digester 1. Spot checks on the total amount of electricity being used for digester heating for periods of between 1-7 days were made during experimental runs. The results are presented below and again show the Digester 2 heating requirement to be 1.20 - 1.55 times greater than that of Digester 1.

Exptl. Run	Length of check (hours)	Energy used for heating (kWh/d)				ΔT (^oC)	RT days
		Dig 1	Dig 2	D1+D2	D2/D1		
1	53.5	16.6	24.7	41.3	1.49	24.4	20
2	52.5	9.1	14.1	23.2	1.55	13.5	15
	24	10.0	12.0	22.0	1.20	15.5	15
	30.0	11.1	14.3	25.4	1.29	15.4	15
3	32.5	9.6	14.8	24.4	1.54	-	15
	23.5	10.2	14.3	24.5	1.40	-	15
4	169.5	17.7	24.9	42.6	1.41	22.4	20
5	166	17.8	22.8	40.6	1.28	19.6	15

4.3.1 Net energy

The net energy for each digester was calculated as follows:

Net energy = Gross energy - Q_s - Q_w

Gross energy (MJ/m^3) = Gross gas production x calorific value of biogas

Q_s = Influent heat requirement (i.e. heat needed to raise temperature of feed slurry to digester bulk temperature
(See Section 2.2.3.1)

The equation for Q_s can be simplified to:

$$Q_s = \frac{1.167 \times \text{Digester volume } (m^3) \times \Delta T (kWh/d)}{\text{Retention time (days)}}$$

ΔT = Digester bulk temperature - Feed slurry temperature (^{o}C)

Q_w = Heat lost by conduction through the digester walls.
(See Section 2.2.3.2)

In calculating Q_w, the heat transfer coefficient for the digester walls, U_w, was determined to be 0.830 W/m^2 ^{o}C.

Table 4.8 shows daily mean values per run of the temperature difference, calculated wall losses and influent heat requirement, and gross and net energy (calculated). It was assumed that the biogas was 60 % methane. The measured values of electricity used for digester heating (kWh/d) are also presented, quoted as mean daily values per experimental run. Temperature data was not available for Experimental Run 3.

TABLE 4.8 Measured and predicted digester heat balance for Experimental Runs 1 - 5.

EXPERIMENTAL RUN	DIGESTER	(MEASURED) BULK-AMB TEMP ˆT °C	P WALL LOSSES Q_w kWh/d	P INFLUENT HEATING Q_s kWh/d	Q_w/Q_s	P GROSS ENERGY kWh/d	P NET ENERGY kWh/d	P TOTAL HEATING Q_w+Q_s kWh/d	MEASURED TOTAL HEATING kWh/d
1	1	24.3	3.67	1.29	2.84	5.61	0.64	4.96	19.3
	2	24.5	7.40	2.62	2.82	11.7	1.72	10.02	24.1
2	1	16.8	2.5	1.19	2.13	7.17	3.45	3.71	10.9
	2	17.1	5.18	2.44	2.12	14.1	6.50	7.62	14.2
3	1	-	-	-	-	-	-	-	11.6
	2	-	-	-	-	-	-	-	16.1
4	1	20.2	3.05	1.07	2.85	5.17	1.05	4.12	16.5
	2	20.7	6.26	2.21	2.83	8.73	0.25	8.47	20.5
5	1	22.1	3.34	1.57	2.13	4.26	-0.65	4.91	16.2
	2	22.0	6.65	3.13	2.12	9.64	-0.15	9.78	21.6

P=predicted

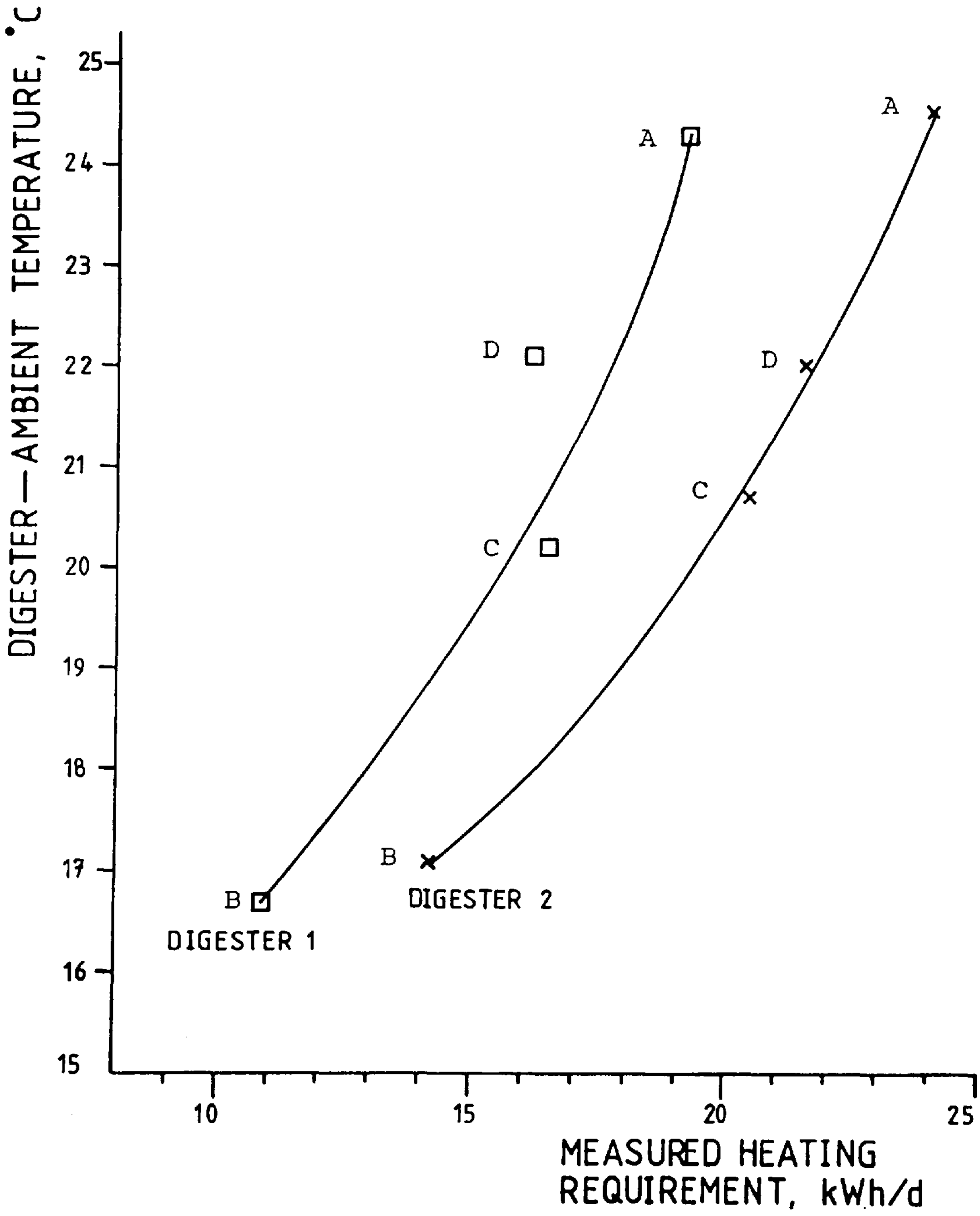
There are wide discrepancies between the predicted and measured digester heating requirement as shown in Table 4.8. The measured value is greater than the predicted, by at least 100%, except for Digester 2, Experimental Run 2 where the difference is 86%. Table 4.8 shows that the greater of the two components making up the digester heating requirement is replacing the heat lost through the digester walls (Q_w). The ratio of Q_w/Q_s is 2.84 and 2.13, at retention times of 20 and 15 days respectively. The measured digester heating requirement was in all cases greater than the gross energy produced by the digesters. This means that gas production from the digesters is insufficient to maintain digester temperature.

The way in which total digester heating requirement (measured) varies with temperature difference (ΔT) is shown in Figure 4.15. Each point represents the mean daily value per experimental run of the heating requirement.

4.4 Tracer experiments

The amount of information that can be gained from a tracer experiment depends mainly on how the tracer was introduced to the reactor, and the sampling regime employed. In experiments looking at the effect of mixer

FIGURE 4.15 Measured digester heating requirement during Experimental Runs.



Experimental Run	Letter
1	A
2	B
3	-
4	C
5	D

performance in digesters, then the digester would be sampled at several locations and the interval between successive samples would be minutes.

In the present experiments, it was required to establish the actual retention time, and the active working volume of the digester. These parameters can be determined from a plot of tracer concentration against time, and a plot of the cumulative amount of the tracer originally added that has left the digester, against time. These are commonly referred to as "E" and "F" curves, or "washout curves". Figures 4.16 and 4.17 are plots of the tracer concentration, and the cumulative amount of tracer in the digester effluent depicting theoretical conditions for a completely mixed reactor, dispersed plug flow and perfect plug flow.

In perfect plug flow, no tracer would be detected in the effluent until 1 retention time after tracer addition, and the tracer concentration would be the same as it was in the dosed feed added to the digester. All the added tracer would be discharged at that time, and no tracer would be detected subsequently. The time when the tracer appeared would be equal to both the actual and theoretical retention time, and the active volume would be equal to the digester volume.

FIGURE 4.16 Tracer concentration in digester effluent for theoretical conditions.

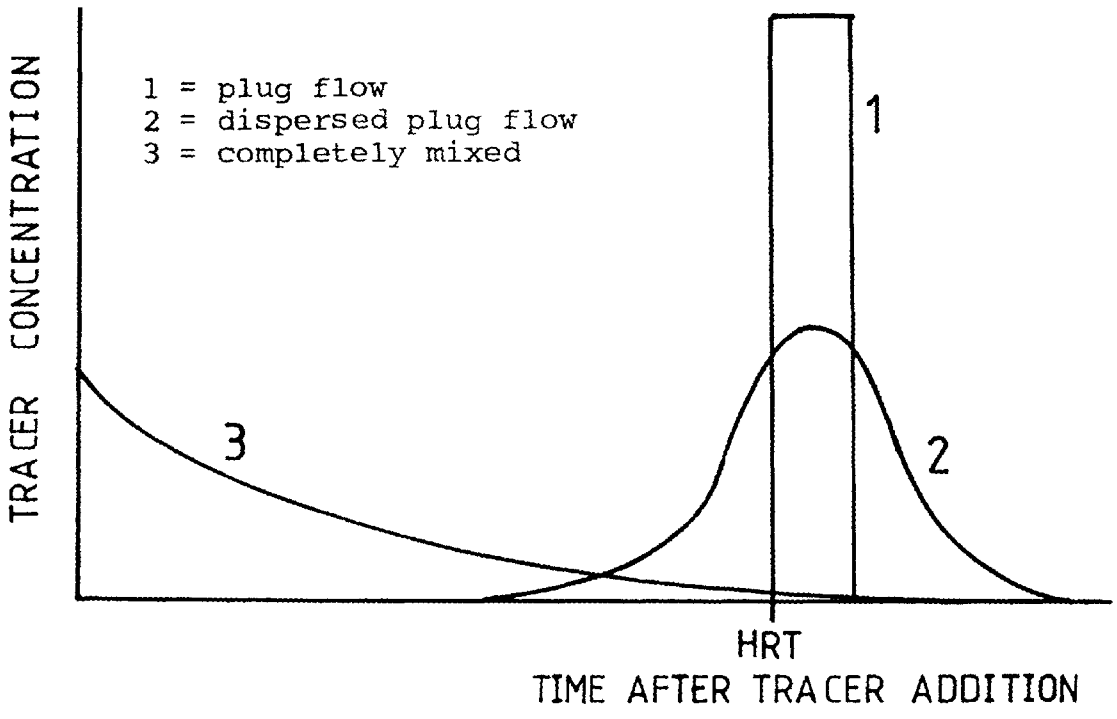
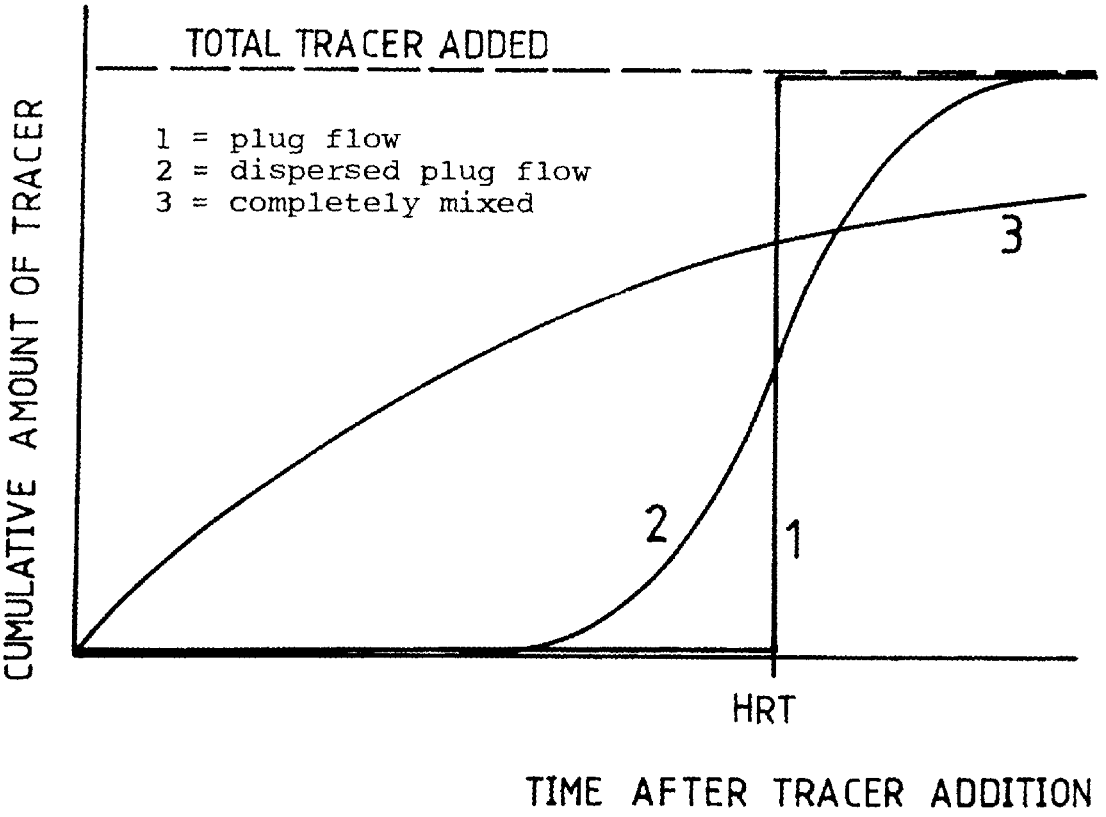


FIGURE 4.17 Cumulative tracer in digester effluent for theoretical conditions.



Dispersed plug flow is more likely to occur. The peak tracer concentration again occurs after 1 retention time, but tracer is detected both before and after this peak. The maximum amount of tracer is expelled from the reactor at the time of peak tracer concentration, which is also the actual retention time. The maximum rate of output is depicted by the inflection point of curve 2 in Figure 4.17. The cumulative washout curve will eventually reach its theoretical maximum value.

In a completely mixed reactor, tracer concentration reduces from an initial peak according to the equation (for continuous feeding):

$$C = C_o \, e^{-t/t_o}$$

where: C = tracer concentration at any time t

C_o = initial tracer concentration at time $t = o$

t = any time t

t_o = hydraulic retention time.

There is a series of possible washout curves ranging from perfect plug flow to the completely mixed situation. When the working volume of the digester has been reduced due to accumulation of solid material, then the peak tracer concentration will be reached before the theoretical retention time, and the washout curves will fail to approach their theoretical limits.

The background level of Cr and Li was measured on 7 occasions during Experimental Runs 1 and 2. The mean concentration of Cr was 0.87 ppm, and Li was 0.50 ppm. Since these concentrations were so low in comparison with the concentrations of added Cr and Li, these background concentrations were ignored in the subsequent analysis.

4.4.1 Tracer experiment 1

The concentration of chromium (Cr) in the digester effluent for Digesters 1 and 2 is shown in Figure 4.18. The cumulative amount of Cr discharged from Digesters 1 and 2 is shown in Figure 4.19.

Figure 4.18 clearly shows that the maximum Cr concentration in the effluent for both digesters was reached 2 days after the dosed feed was introduced. There was obviously a large inactive volume within each digester, and the actual retention time for the components marked by the Cr was 2 days.

The washout curve for complete mixing of digester contents for a retention time of 20 days is shown on Figure 4.18. The Cr concentration was greater than that predicted for a completely mixed system during days 1-6.5 for Digester 1, and days 1-6 for Digester 2. The high (5440 ppm) initial concentration of Cr which was

FIGURE 4.18 Tracer Experiment 1. Concentration of chromium in effluent for Digesters 1 and 2.

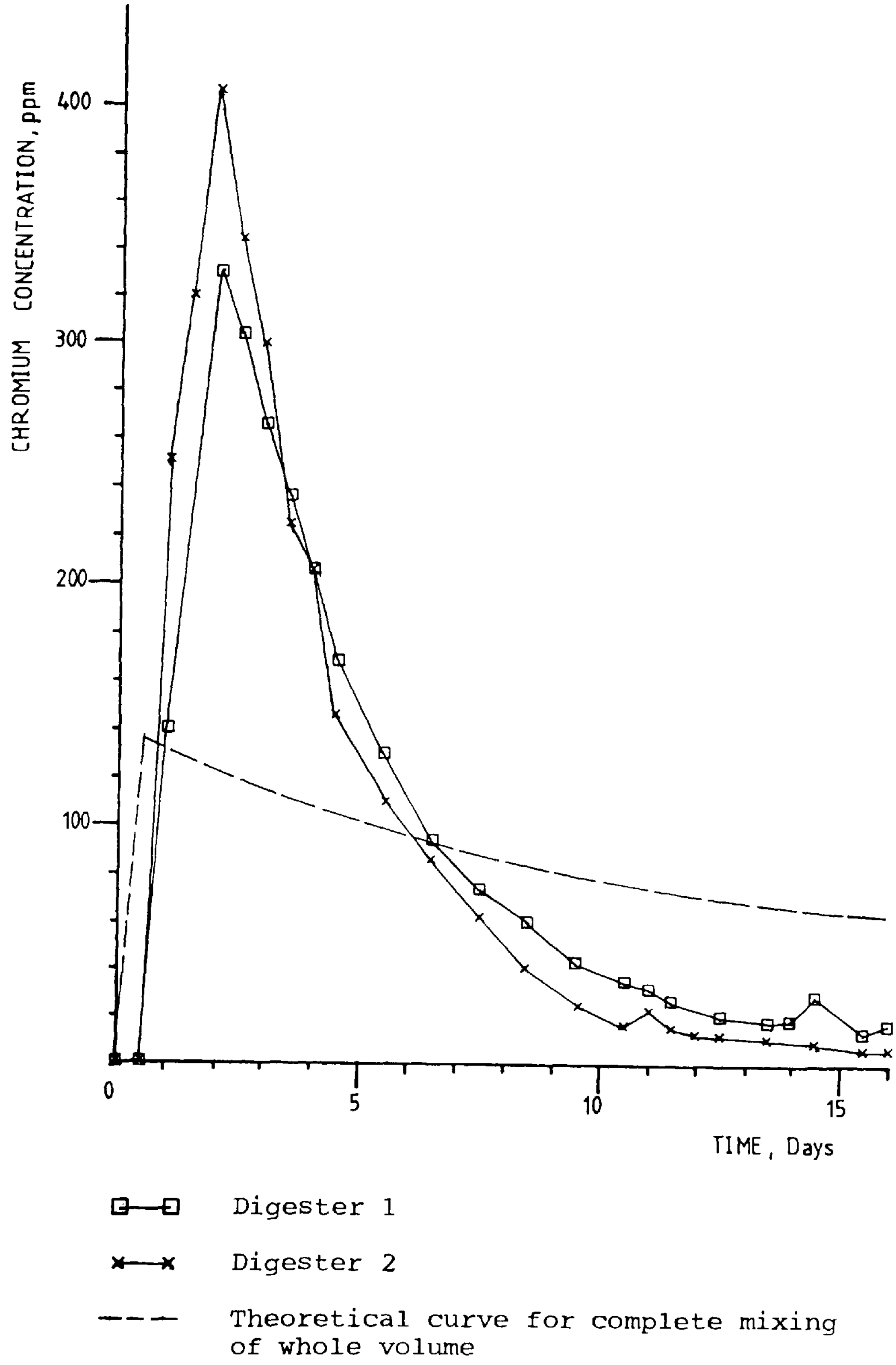
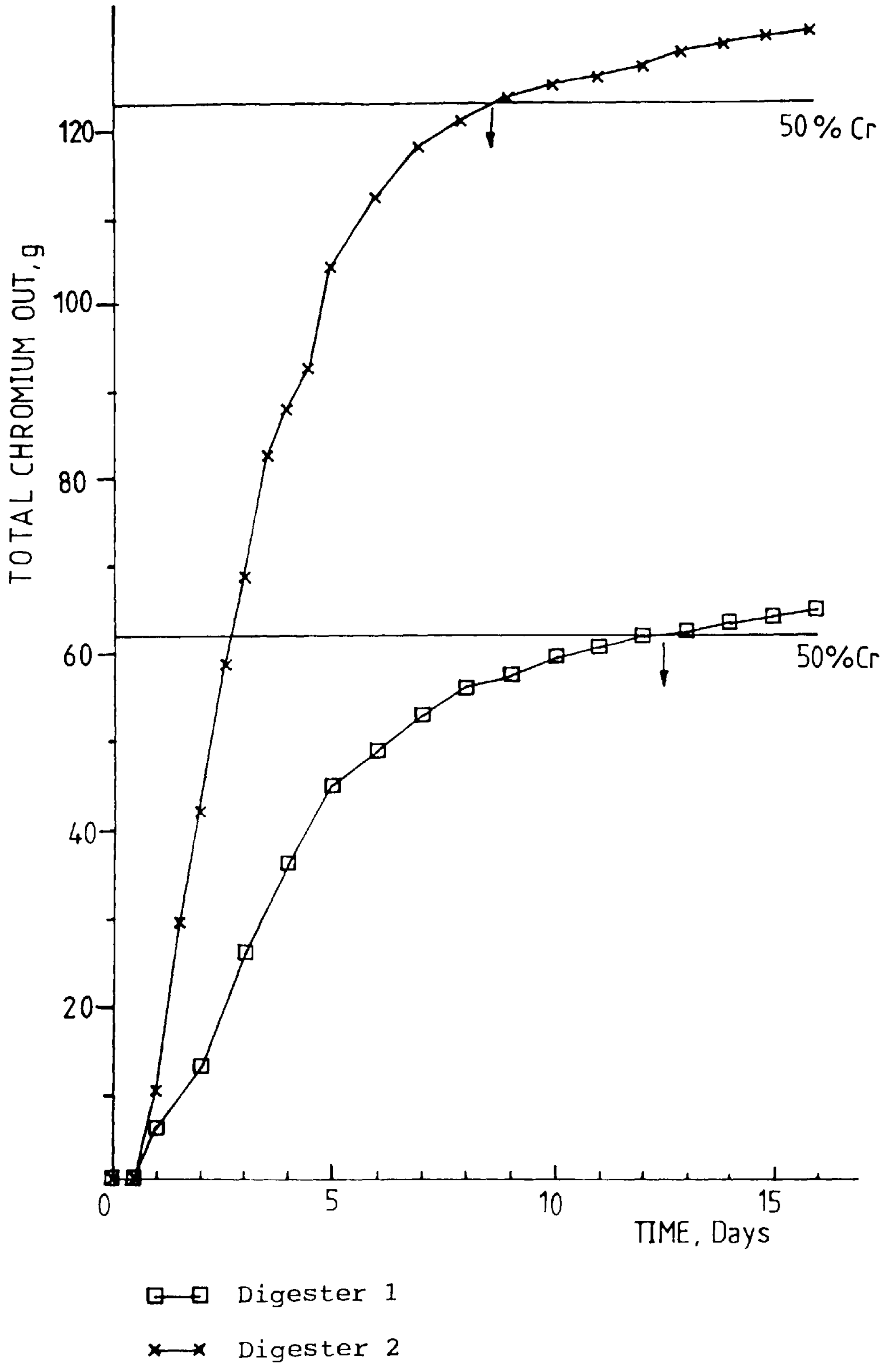


FIGURE 4.19 Tracer Experiment 1. Cumulative amount of chromium discharged from Digesters 1 and 2.



introduced in the feed "plug" has obviously not been dispersed evenly throughout the reactor, as shown by the high Cr concentrations up to day 6.5. The higher Cr effluent concentration from Digester 2 compared with Digester 1, implies that the greatest % reduction in working volume occurred in Digester 2. This is further confirmed by examination of the cumulative or "F" curves in Figures 4.19. From Figure 4.19 it can be seen that it took 12.5 and 9 days for 50% of the Cr to be discharged from Digesters 1 and 2 respectively. Therefore, components moving through the digester in a similar way to Cr, had a shorter <u>actual</u> retention time in Digester 2, than in Digester 1.

After 50% of the Cr had been recovered from each digester, the amounts of Cr recovered per day were very small, as shown by the gradient of the lines in Figure 4.19. The cumulative washout curves failed to approach their theoretical limits, indicating that a large amount of Cr had been retained in each digester, and is no longer in circulation.

4.4.2 Tracer experiment 2

The markers used during this experiment were chromic oxide and lithium chloride. Digester feedstock was separated cattle slurry, and the nominal retention time was 15 days.

The concentration of Cr and Li in the effluent from Digesters 1 and 2 are shown in Figures 4.20 and 4.23 respectively.

Digester 1 Aspect ratio = 10

Figure 4.20 shows the maximum concentration of both Li and Cr to occur 2 days after injection. The shape of the 2 curves is very similar. The curve for Li is very close to the theoretical washout curve (for Li) assuming complete mixing of the digester contents, for a 15 day retention time. As Li is a soluble tracer, this shows soluble components of the digester feedstock are well mixed throughout the whole digester. Figure 4.21 shows Li concentration plotted on a log scale, against time (arithmetic scale). The equation of the line was calculated to be :

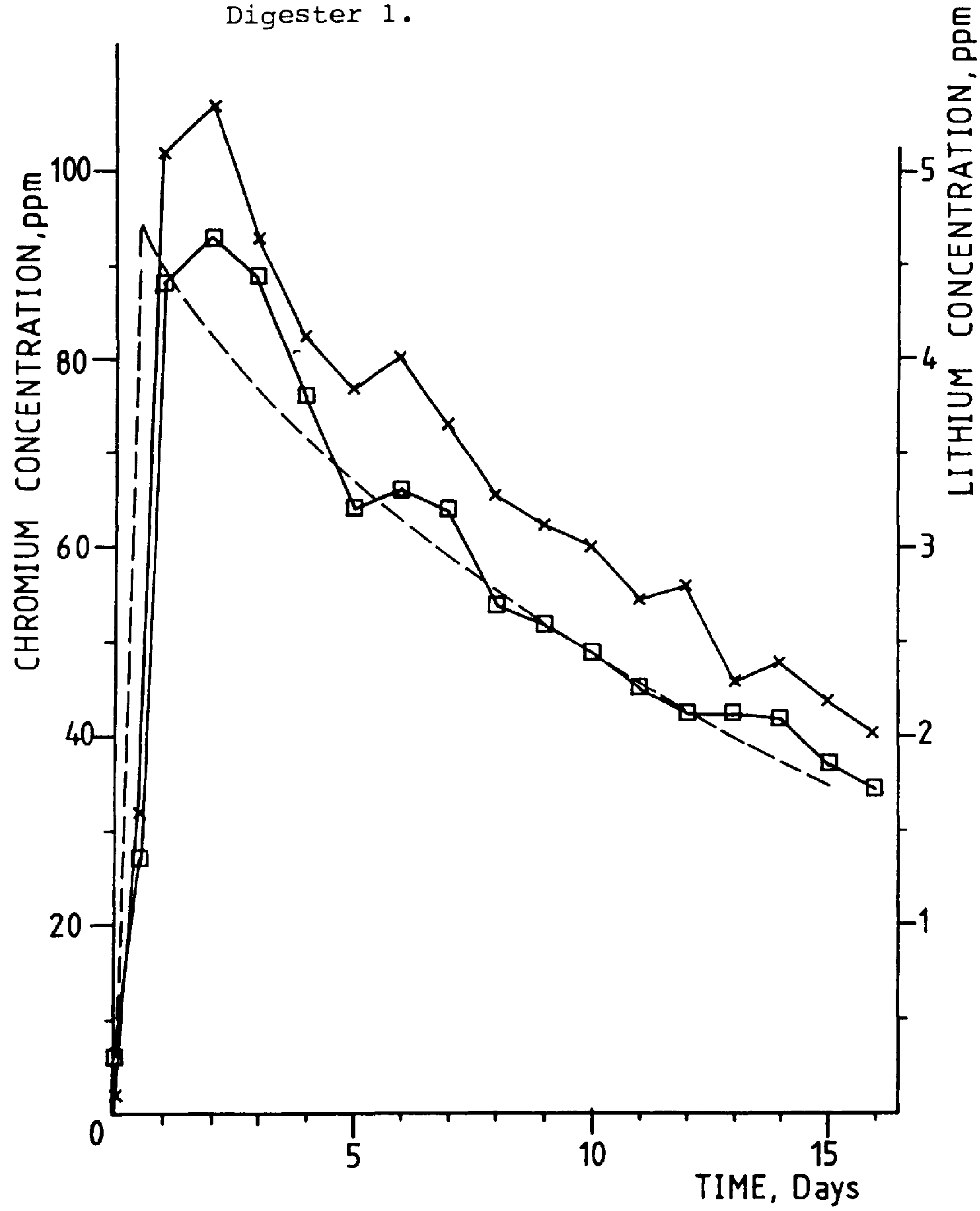
$$C=4.91\ e^{-0.066t}$$

where C = Tracer concentration at any time t
t = Time (days) after injection of tracer.

$$\text{The actual retention time} = \frac{1}{0.066} = 15.2 \text{ days}$$

Figure 4.22 shows the cumulative washout curve for Li and Cr, 50% of the Li being discharged within 9 days. By the end of the experiment (16 days), 72.3% of the initial Li had been recovered.

FIGURE 4.20 Tracer Experiment 2. Concentration of chromium and lithium in effluent from Digester 1.



×—× chromium

⊡—⊡ lithium

— — — theoretical curve for complete mixing of lithium throughout the whole volume

FIGURE 4.21 Tracer Experiment 2. Concentration of lithium (log scale) in effluent from Digester 1.

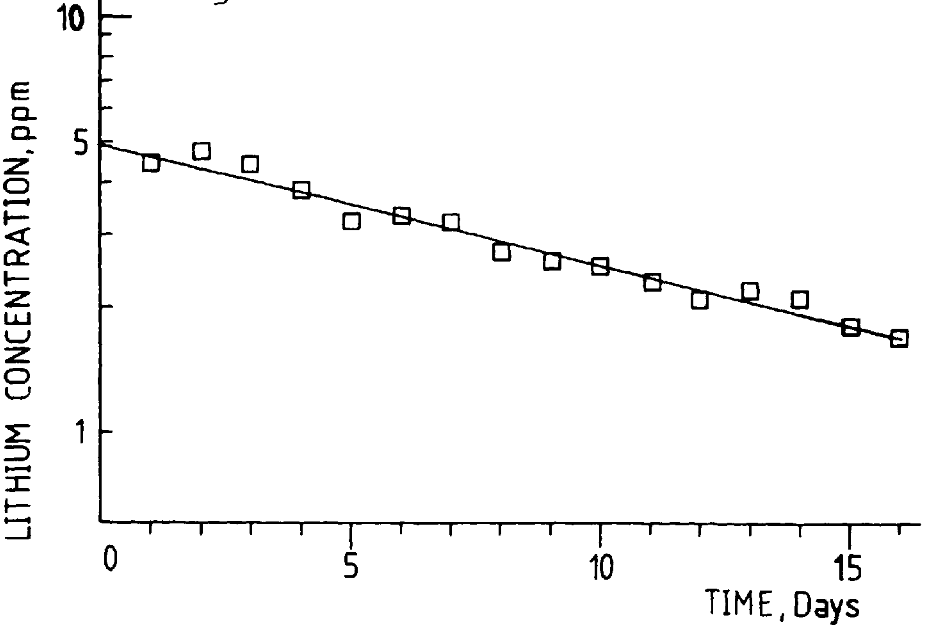
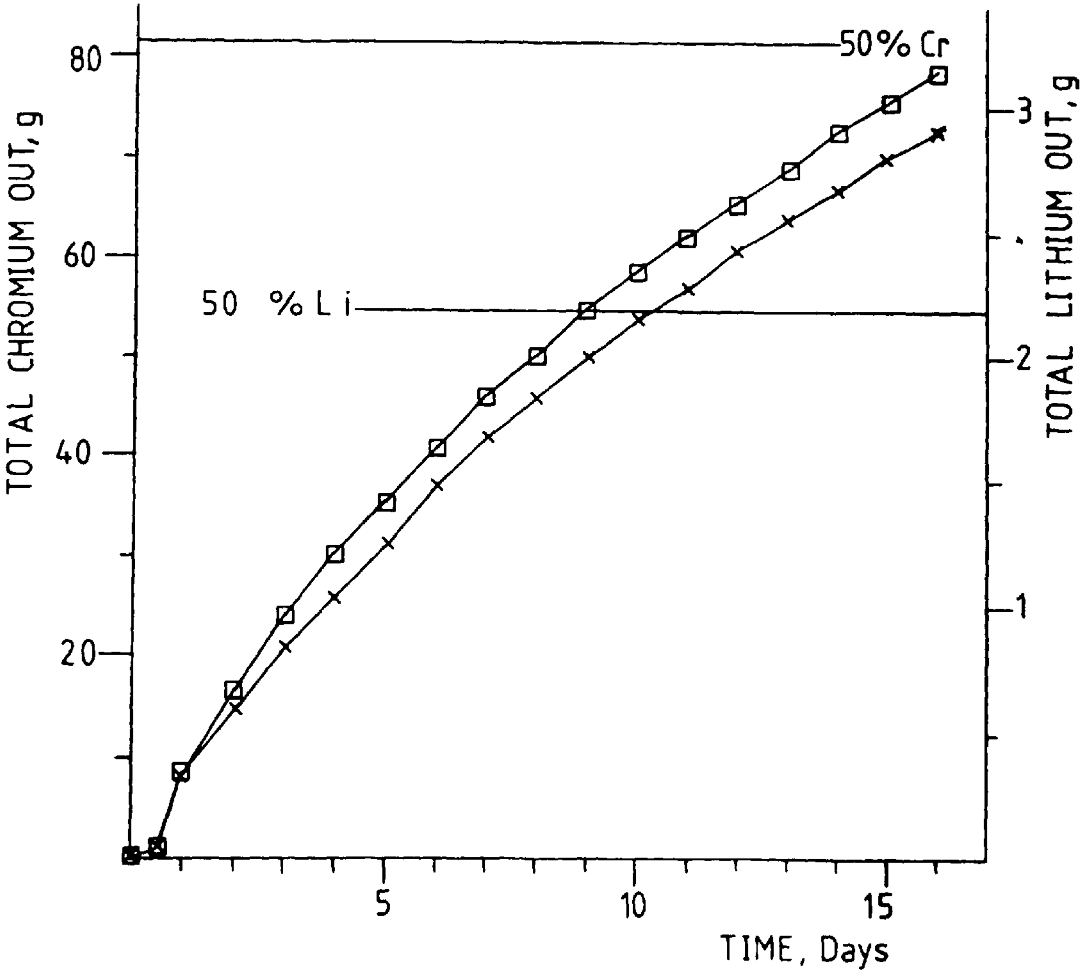


FIGURE 4.22 Tracer Experiment 2. Cumulative amount of chromium and lithium discharged from Digester 1.



The movement of Cr through the digester was similar to Li. Although the shape of the washout curve for Cr was similar to that of Li, it was always less than the theoretical washout curve for Cr, assuming complete mixing. This implies that some Cr moves through the digester like Li and soluble components of digester feed. Since measured Cr concentrations never reached theoretical values for a completely mixed system, a large amount of Cr is obviously being retained in the digester and passing out of circulation. The total amount of Cr recovered was 44.6% of the initial dose. This is not suprising considering the high specific gravity of chromic oxide (5.1), and the separated slurry digester contents, the mean effluent TS being 3.6% during the tracer experiment.

Digester 2 Aspect ratio = 20

The movement of Cr and Li marked components was very similar to that in Digester 1. Figure 4.23 shows the movement of Li to closely approximate the washout curve for a completely mixed digester, at a 15 day retention time. The peak Li and Cr concentrations occurred after 1 day, showing this digester to be better mixed than Digester 1.

Figure 4.24 shows Li concentration plotted on a log scale, against time (arithmetic scale). The equation of

FIGURE 4.23 Tracer Experiment 2. Concentration of chromium and lithium in effluent from Digester 2.

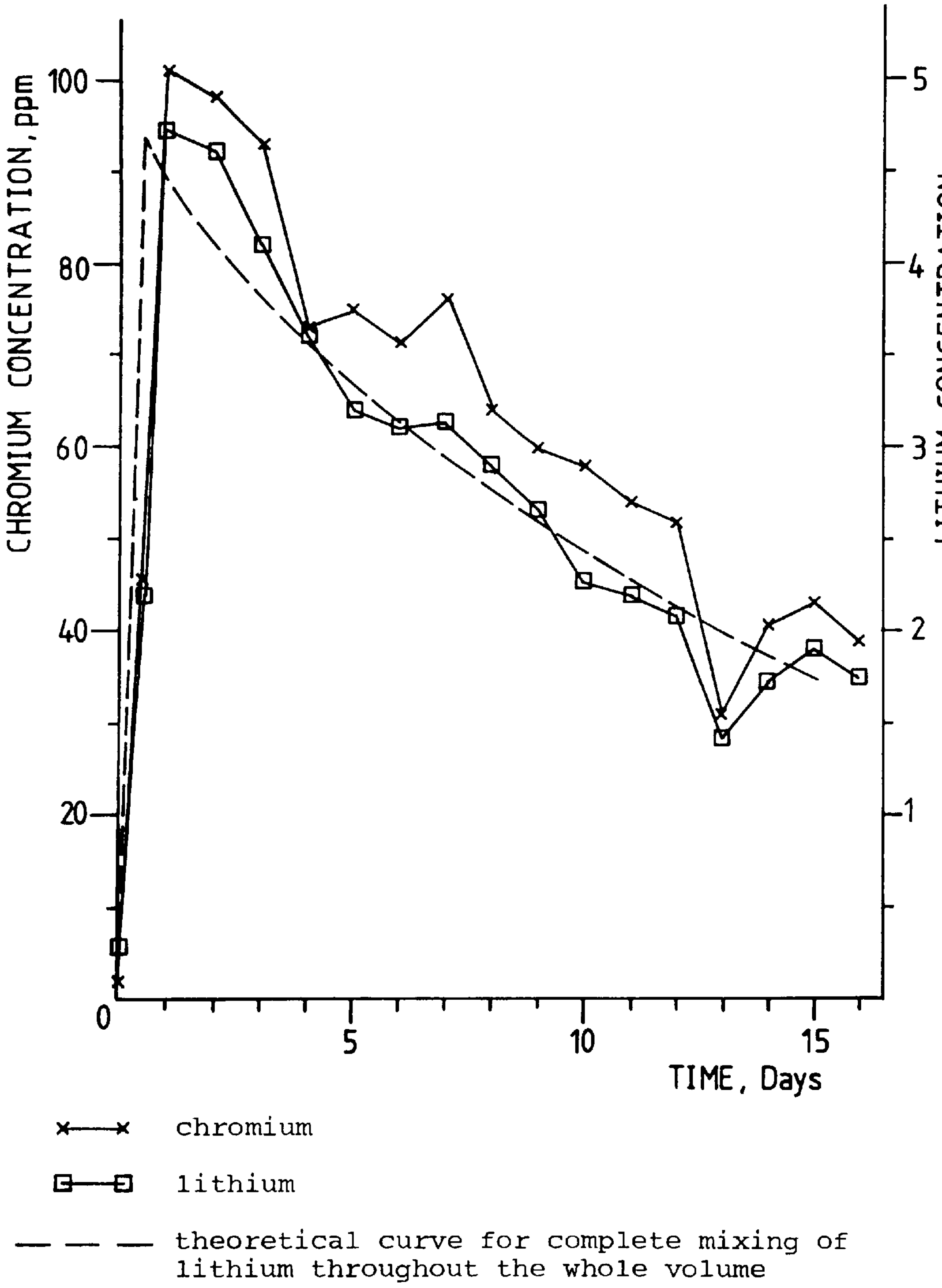


FIGURE 4.24 Tracer Experiment 2. Concentration of lithium (log scale) in effluent from Digester 2.

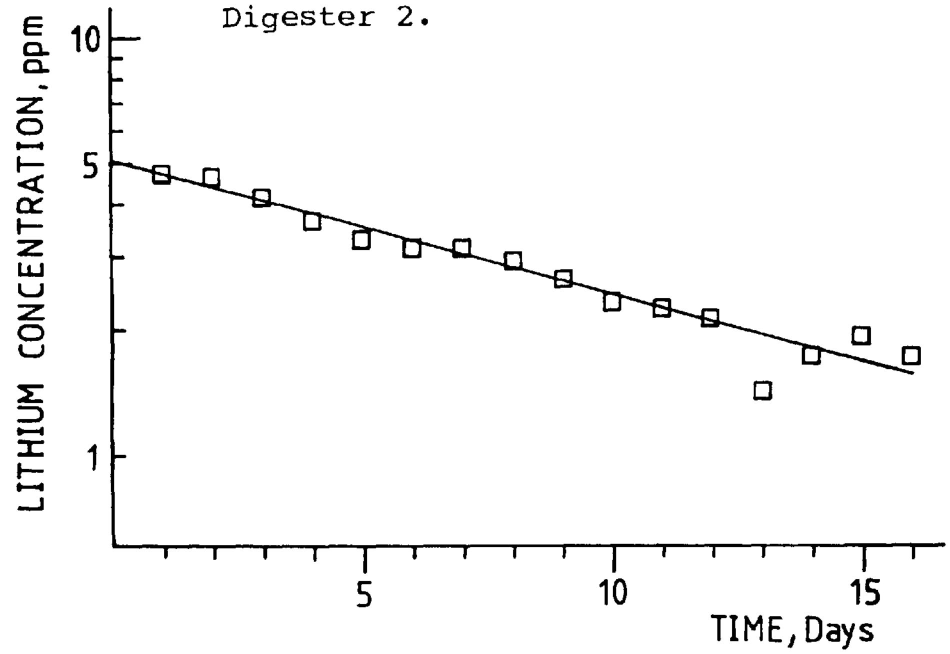
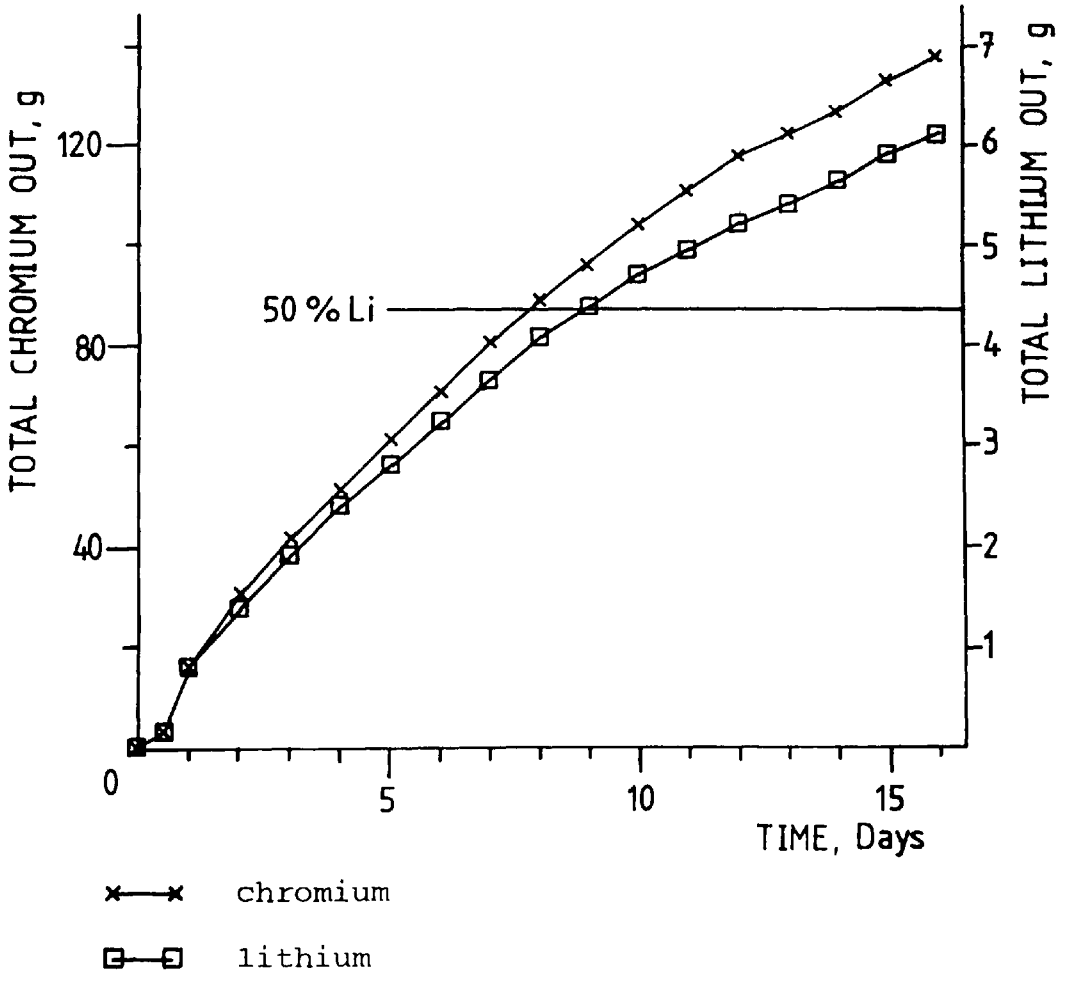


FIGURE 4.25 Tracer Experiment 2. Cumulative amount of chromium and lithium discharged from Digester 2.



the line was calculated to be :

$$C=5.02\ e^{-0.074t}$$

where C = Tracer concentration at any time t

t = Time (days) after injection of tracer

$$\text{The actual retention time} = \frac{1}{0.074} = 13.5 \text{ days}$$

Figure 4.25 shows the cumulative washout curves, 50% of the Li being recovered in 9 days, the same as for Digester 1. The total recovery of Li at 70.59% during 16 days, was again good. This is in contrast to the recovery of Cr, the total being 42.4% of the initial dose, again showing a large amount of Cr was retained in the digester, and passed out of circulation.

CHAPTER 5

DISCUSSION OF RESULTS

5.1 Discussion

The experimental programme sought to establish the effect of a range of operational and design parameters on digester performance in terms of biogas production and changes in slurry composition. The relationships investigated were those between retention time, angle of inclination, aspect ratio, temperature and type of slurry.

5.1.1 Feed Slurry Composition

During the first four experimental runs (whole slurry), the mean %TS of the feed for each run varied from 7.03-8.13%, with a maximum SD of 1.67%. Such variation is to be expected, given that the digesters were operating on a farm site and the amount of water entering the slurry, primarily as washings from the milking parlour, did vary. This is typical of the slurry produced on many large dairy units in the UK. The %TS content of the slurry leaving the cow is often in the range 12-14% (Grundey, 1980), but handling is more difficult at these higher values of %TS. Also, to ensure a daily supply of slurry at these high values of %TS is difficult, and rigorous management of washings from the parlour, collecting yards and rain water is necessary to prevent water from diluting the slurry.

In the first 4 experimental runs the volatile solids (as % of the TS) in the feed varied from 80.91-85.13%, with a maximum standard deviation of 1.98%. Large variations in the %VS of the feed would indicate a change in the proportion of the feed that was biodegradable, and result in variation in the loading rate. The %VS would not be expected to change significantly for a given livestock enterprise unless there were major changes in the animal husbandry, for example a change in bedding from sawdust to sand. The cows were bedded on sawdust during all experimental periods.

The total volatile fatty acid concentration (TVFA) in the feed showed considerable variation, the mean values for each run varying from 4809-6513 ppm, with a maximum standard deviation of 1327 ppm. These values are similar to TVFA concentrations reported by other workers e.g. 4370ppm, SD ± 1000 (whole); 4140ppm, SD ± 960 (separated) - Pain _et al_, (1984); 3650 ppm, SD ± 1462 (separated) - Hawkes _et al_, (1984). The feed TVFA were known to vary, primarily for two reasons. The inclusion of silage liquor in the feed, which is of very high VFA (>10,000ppm), did occur despite attempts to prevent this. This was a consequence of having the digesters on a large dairy unit, including 4 x 1000 tonne silage clamps.

An increase in the feed TVFA also occurred due to warming of the feed in the above ground PVC tank used for storage. The hydrolytic and fermentative bacteria increase in numbers faster than the methanogenic bacteria (Hobson & Richardson, 1985) and the VFA content of the feed is increased. If the tank were sealed, then the build-up of acids would inhibit the methanogens, and finally the hydrolytic bacteria, and cause the digestion to cease. Such a digester is commonly referred to as being "soured".

Experiment 5 used separated slurry as the feedstock. It has been shown (Pain et al, 1978) that separation with a roller press machine removes about 18% by weight of the original slurry as fibrous solids. This is seen in the feed TS value of 4.69%. Separation was found to cause a small decrease in %VS by Pain et al, 1984. During Experiment 5, the mean %VS was 77.06, SD±1.79 compared with a range of mean values per run of 80.91-85.13% for the other Experimental Runs.

The mean concentration of feed VFA during Experimental Run 5 was 4749 ppm (SD±739), compared with a range of 4809-6513 ppm during the other experimental conditions.

5.1.2 Discharge Slurry Composition

The composition of the discharge slurry is determined by the composition of the feed slurry, and the digester operating conditions during a particular run.

In Experiments 1, 2 and 4, all at 35°C and with whole slurry, the discharge %TS ranged from 4.89-5.69, with a maximum standard deviation of 1.22%. Similarly, VS% ranged from 75.73-82.55%, with a maximum standard deviation of 4.57%.

There were considerable variations in the discharge VFA concentrations from the digesters during Experiments 1, 2 and 4. The mean discharge VFA concentrations during Experimental Run 1 (20 day RT) were very similar at 673 and 667 ppm from Digesters 1 and 2 respectively. The mean discharge VFA concentrations were lower during Experimental Run 2 (15 day RT), at 578 and 469 ppm for Digesters 1 and 2 respectively. The longer RT in Experimental Run 1 would have been expected to lead to lower VFA concentrations in the discharge, than from Experimental Run 2. This is probably accounted for by the higher mean VFA concentrations in the feed for Experimental Run 1 (6034ppm), compared with 4931 ppm for Experimental Run 2.

In Experimental Run 4, the mean discharge VFA concentrations were 1367 and 2725 ppm from Digesters 1 and 2 respectively. These are very high levels of discharge VFA concentration, and suggest that considerable short-circuiting inside the digester is taking place. This was confirmed by tracer studies which were carried out (see Section 5.1.8). The two experimental runs in Period 4 were during the periods of most stable digester performance that could be identified. It must be fully recognised that any periods of digester instability occurring at this scale of operation (1 and 2m^3), and under experimental conditions, will probably occur at a magnified scale in farm scale units of this design.

Performance data for the Experiments at 35^{o}C (Experiments 1, 2 and 4) are summarised in Table 5.1.

5.1.3 The effect of temperature

Experiments 1, 2 and 4 were at 35^{o}C with whole slurry, and Experiment 3 was at 25^{o}C, again with whole slurry. The results from Experiments 1, 2 and 4 are summarised in Table 5.1, and Experiment 3 in Table 4.4.

The effect of temperature on digester performance is well documented in the literature (Van Velsen et al,1979; Hawkes, 1980). Within the mesophilic temperature range (25-40^{o}C), gas production increases with increasing temperature. At 25^{o}C, the mean gas yield (m^3/kg VS added) was 0.154, and at 35^{o}C this was 0.299 (average for both digesters).

The reduction of VS and VFA were less at 25^{o}C (20.9%, and 57.9%), compared with 29.7% (VS) and 88.9% (VFA) at 35^{o}C.

The relative importance of gas yield, and reductions in VS and VFA, depends on the purpose of the digester installation. The gas yield (m^3/kg VS added) is 90.9%

TABLE 5.1 Summary of digester performance data for Experimental Runs at 35°C and with whole slurry.

	EXPERIMENTAL RUN NUMBER						
	1	1	2	2	4	4	MEAN (SD)
Digester inclination, deg. to horizontal.	20	20	20	20	10	10	
Aspect ratio	10	20	10	20	10	20	
Digester temperature, °C	34.61	34.77	34.74	35.12	33.26	36.20	
Feed temperature, °C	10.27	10.27	17.98	17.98	13.06	15.46	
Nominal RT (days)	20	20	15	15	20	20	
GAS YIELDS m^3/m^3 reactor / day	1.001	1.042	1.279	1.253	0.907	0.801	1.144 (0.1425)
m^3/ kg TS added	0.265	0.240	0.245	0.226	0.228	0.228	0.244 (0.0161)
m^3/ kg VS added	0.318	0.288	0.308	0.282	0.282	0.286	0.299 (0.0169)
m^3/ kg TS destroyed	1.025	0.852	0.719	0.723	0.827	0.753	0.829 (0.1441)
m^3/kg VS destroyed	1.200	1.004	0.956	0.968	0.958	0.848	1.032 (0.1138)
% TS reduction	26.1	28.4	34.2	31.1	27.6	30.8	30.0 (3.49)
% VS reduction	26.8	29.1	33.1	29.6	29.4	34.3	29.7 (2.60)
% VFA reduction	88.9	88.9	87.8	89.9	78.6	56.1	88.9 (0.86)

Note : The results from Experimental Run 4 were not used in calculating the mean values.

greater at 35^{o}C, than at 25^{o}C. However, to maintain a digester at 35^{o}C compared with 25^{o}C, obviously requires more energy and the difference in net energy at these two temperatures will not be as great. The effect of the heating requirement on net energy is of particular importance at short retention times (10 days of less). In digesters where pollution control is the primary objective, operation at 25^{o}C may be acceptable, since in many instances there would be an additional treatment stage also. Where energy production is the main objective, the higher operating temperature would usually be selected. However, although the higher temperature would lead to the greatest gross energy production, the maximum <u>net</u> energy may be produced at a lower temperature. The higher operating temperature would in fact reduce the polluting power of the waste more, but extra costs would be incurred for the additional digester heating, and capital and operating costs for energy utilization equipment.

Experiment 3 was characterized by stable operation, as shown in Figures 4.7 and 4.8. The digester temperature was maintained very close to its target operating value at all times. This is important for continued performance, and stability of operation since large temperature shocks are known to cause digester failure (Peck <u>et al</u>, 1986).

5.1.4 The effect of Retention Time

During Experimental Runs 1 and 2, the digesters were operated at retention times of 20 and 15 days respectively. The configuration of the digesters, and the operating temperature was the same during both these runs.

Table 5.1 includes a summary of digester performance data from Experiments 1 and 2. Gas yields were slightly greater at 20 day RT, than at 15 day RT. This is because the longer time for digestion enables degradation of a higher proportion of the larger particles present in the feedstock (Hobson, 1983). A longer retention time implies a larger and more costly reactor, which could be justified in some circumstances.

In Experimental Run 4, tracer studies (Section 5.1.8) showed there to be large reductions in the effective working volume of each digester. The actual retention times were far less than than the nominal values. It was not possible to determine the rate at which the effective digester working volume reduced. It is possible that there were reductions in digester working volume during other Experimental Runs, but there are no results to confirm this. A consequence of the shortened retention times found in Experimental Run 4 was raised levels of digester VFA, and a reduction in the stability of the digestion process.

5.1.5 The effect of Digester Inclination

In the first three experimental runs, the digesters were inclined at 20° to the horizontal, and during the last two experiments at 10°.

In Experiment 1, digester inclination was 20°, and in Experiment 4 it was 10°. Apart from the change in digester inclination, all other aspects of digester configuration, operating temperature, retention time and feedstock were the same during both Experiments. Digester performance data for these two Experiments are included in Table 5.1.

Gas yields were greater during Experiment 1 (20°) compared with Experiment 4 (10°), but there was greater destruction of TS and VS in Experiment 4.

Figure 4.1 shows the daily gas production and feed TS content for Experiments 1-3 when the digesters were inclined at 20° to the horizontal. The same parameters are shown in Figure 4.2, when the digesters were inclined at 10° to the horizontal. From these figures it can be seen that gas production, and hence stability of the digestion process, was more uniform at 20° inclination, than 10°. It has already been commented that there was no 45 day period during Period 8 (Figure 4.2) when the performance of both digesters was relatively stable.

The greater stability of operation at 20° inclination, compared with 10°, has important consequences for larger versions of this digester design. The instability at 10^o inclination would be considered unacceptable for a full size unit of this type, operating on a similar feedstock.

Floyd (1984) used 3 inclined tubular digesters with a volume of 13 - 15 l in laboratory experiments in which the angle of inclination ranged from 16 - 20^o. It was not possible to identify any effect of angle of inclination, independently of other parameters, from that work.

5.1.6 The effect of aspect ratio

The configuration of Digester 1 was always such that the aspect ratio was 10, and for Digester 2 it was 20. The digesters were operated side by side during all experimental runs, the temperature, angle of inclination and retention time being the same for each digester.

Table 5.1 shows that the gas yield (m^3/kg VS added) was greater for an aspect ratio of 10 compared with 20, at an operating temperature of 35^oC. In experiments 1 and 2, the gas yield was greater by 10.4% and 9.2% respectively, for the digester with the aspect ratio of 10 compared with 20. This greater gas production was probably due to better mixing in the reactor where the aspect ratio was 10. The mixing is caused by the convection currents inside the reactor which arise from the heating system, and also from the evolved biogas as it moves to the uppermost end of the reactor.

The results of Floyd (1984) working with pig slurry of 5 or 10% TS, showed the opposite trend. In 3 paired runs, the mean gas yields (m^3/kg VS added) were 14, 3 and 11 % greater from a digester of aspect ratio 15, compared with a digester of aspect ratio 7. The greater gas yields were attributed to solids retention in the digester of aspect ratio 15, as indicated by the TS% near the effluent weir. However, Floyd used different digester diameters to provide different aspect ratios, and the increased yields were attributed to the greater size of the effluent apperture rather than aspect ratio per se. At the laboratory scale used by Floyd (13 - 151), and working with feedstock of 5 or 10% TS, the effect of digester dimensions is likely to be greater than at the full scale. This is shown in Floyd's work whereby the smallest difference in gas yield for the paired digesters (3%) was with a feedstock of 5% TS.

In the present study, digester diameter was the same (0.48m) for all conditions, and aspect ratio was varied by altering digester length.

In Experiment 4, the gas yields (m^3/kg VS added) were the same from digesters with aspect ratios of 10 and 20.

Similarly, in Experiment 3, Table 4.4 shows the gas yields to be the same from each digester. The digester temperature during this experiment was 25°C, and the viscosity of the digester contents would be greater than at 35°C (Kumar et al, 1972). This would reduce the mixing effect due to convection currents, and that from the evolved biogas. At this lower temperature, the gas yields (m^3/m^3 reactor day) were approximately 42% lower than at 35°C (Table 4.3).

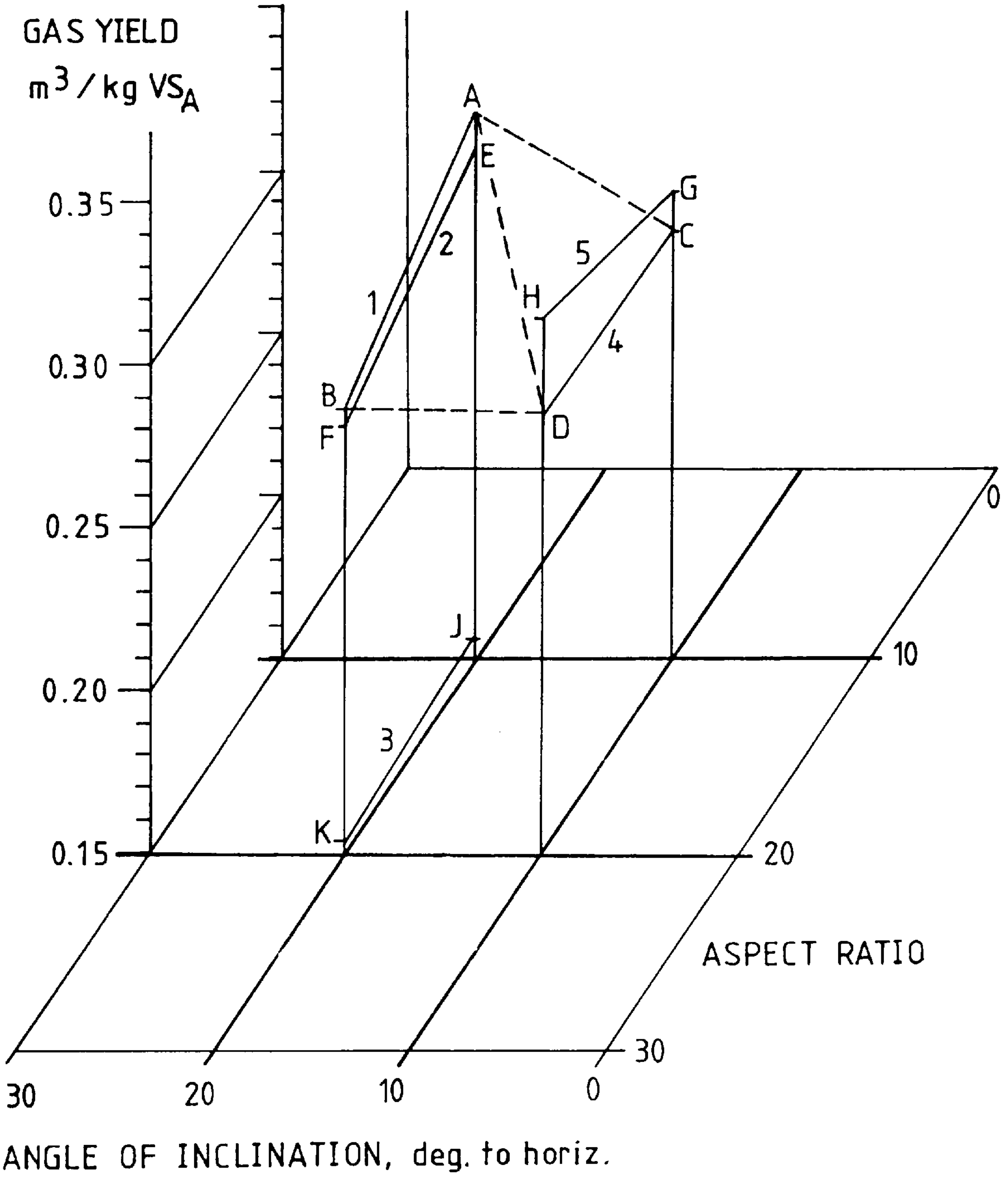
Experiment 5 was the only time when separated slurry was used as the feedstock. Gas yields were very similar from the two digesters. The gas yield (m^3/kg VS added) from Digester 2 was 0.316, which was 7% greater than Digester 1 (Aspect ratio = 10). This can perhaps be attributed to greater solids retention in the digester where aspect ratio was 20. A similar trend was observed by Floyd (1984), also working with tubular digesters.

The gas yields obtained from both digesters, during all experimental runs are presented graphically in Figure 5.1. The effect of aspect ratio and angle of inclination on gas yield is also shown in this 3 - dimensional representation.

Each point in Figure 5.1 represents the mean gas yield (m^3/kg VS added) obtained during the Experimental Run, and as presented in Tables 4.2 - 4.6. Gas yields evaluated over the same period in time are connected by solid lines, and numbered according to the Experimental Run they represent.

Considering the points ABDC, which are all for a retention time of 20 days. Line 1 indicates that for a constant digester inclination of 20° to the horizontal,

FIGURE 5.1 The effect of aspect ratio and angle of inclination on gas yield.



EXPERIMENTAL RUN	POINTS	RT (days)	TEMP (°C)	SLURRY TYPE
1	AB	20	35	whole
2	EF	15	35	whole
3	JK	15	25	whole
4	CD	20	35	whole
5	GH	15	35	sep.

gas yield increases with decreasing aspect ratio. The same trend is found with lines 2 and 3, as shown in the table below:

Experiment	Equation of line
1 (AB)	$VS_A = -3 \times 10^{-3}$ (AR) + 0.348
2 (EF)	$VS_A = -2.6 \times 10^{-3}$ (AR) + 0.334
3 (JK)	$VS_A = -3 \times 10^{-4}$ (AR) + 0.16
4 (CD)	$VS_A = 4 \times 10^{-4}$ (AR) + 0.28
5 (GH)	$VS_A = 2.3 \times 10^{-4}$ (AR) + 0.27
DB	$VS_A = 2 \times 10^{-4}$ (Angle) + 0.28
CA	$VS_A = 3.6 \times 10^{-3}$ (Angle) + 0.25
DA	$VS_A = -2.2 \times 10^{-3}$ (AR) + 0.33
DA	$VS_A = 3.2 \times 10^{-3}$ (Angle) + 0.254

where VS_A = Gas yield, m^3/kg VS added
AR = Aspect ratio
Angle = Digester inclination to horizontal (degrees)

Considering the points ABDC in terms of constant aspect ratio, gas yield increases with increasing angle of inclination. This increase is greater for an aspect ratio of 10, than 20. It has already been pointed out that in Experimental Run 4 the digesters were less stable than for other conditions, and any conclusions drawn from those results are only very tentative. However, the trends shown in the results from the 3 Experimental Runs carried out at 20° inclination, each using whole slurry, are consistent.

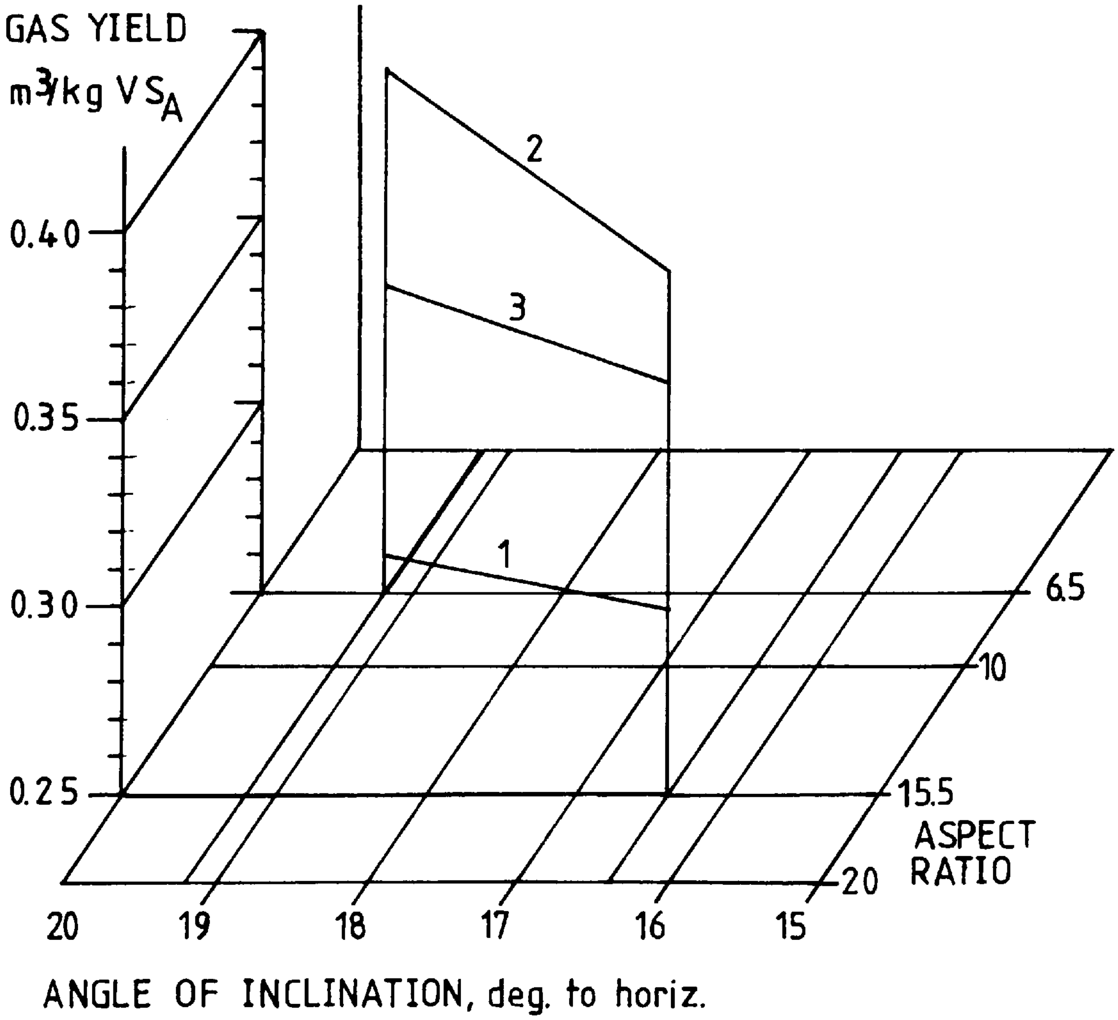
However, at the 10^{o} inclination, the opposite trend is found whereby gas yield decreases with decreasing aspect ratio as shown in line 4 (whole) and line 5 (separated slurry). However, it must be noted that line 4 represents Experimental Run 4 which was relatively unstable (Section 4.2.4), and line 5 is for separated slurry, all the other lines being for whole slurry. Therefore, the results are not directly comparable in this respect.

The lines DB and CA are both of positive slope, as given above, and it would therefore seem reasonable to regard line DA as representing the way in which gas yield can be predicted from aspect ratio or angle of inclination. The small number of points used in calculating the equations of these lines means that the conclusions drawn are somewhat tenuous.

The engineering implications of these results are significant. The greatest gas yields were obtained with low aspect ratio (10) digesters, steeply inclined (20^{o}) to the horizontal. The practical significance of these findings are further discussed in Chapter 6.

Figure 5.2 is a similar representation of the combined effects of angle of inclination and aspect ratio on gas yield from the results of Floyd (1984). Gas yields evaluated during the same period in time are connected by solid lines, and numbered according to the paired run they represent. These results are in complete

FIGURE 5.2 The effect of aspect ratio and angle of inclination on gas yield from laboratory scale inclined tubular digesters (pig slurry).



LINE	PAIRED RUNS
1	4, 8
2	6,12
3	5,9
RESULTS : Floyd, 1984	

contrast to those shown in Figure 5.1, and the equations of the lines were as follows:

Condition	Equation of line
1	$VS_A = 4.1 \times 10^{-3}\ (AR) + 0.235$
2	$VS_A = 1.34 \times 10^{-3}\ (AR) + 0.370$
3	$VS_A = 4.26 \times 10^{-3}\ (AR) + 0.294$

<u>Floyd, 1984</u>

where VS_A = Gas yield, m^3/kg VS added
AR = Aspect ratio

The lines are all of positive slope, and suggest that gas yield increases with increasing aspect ratio. However, it must be noted that these results show the <u>combined</u> effect of aspect ratio and angle of inclination obtained with digesters inclined from 16-19°, and of differing diameters and effluent weir construction.

5.1.7 Net energy production

There were shown to be wide discrepancies between the predicted and measured digester heating requirement in Section 4.3. It was also shown that at all times the gross energy production from each digester was less than the electrical energy used for digester heating.

These discrepancies are due to additional heat losses in the system which cannot be accurately quantified, and errors in predicting the wall losses and influent slurry heating requirement. Heat losses will occur in the heating system at the hot water cylinder, and in the various pipe runs between the cylinder and the digester.

The design of the heating system was such that the heating was switched on when the datalogger detected the temperature of the control thermocouple to be below a pre-set level ($25^{o}C$ or $35^{o}C$). Additionally, a thermostat in each water heating cylinder switched the immersion heater off when water temperature exceeded the pre-set value of $65^{o}C$. A water circulation pump for each heating system ran continuously. There was a separate water heating system for each digester, the only link between the two systems being a common cold water feed pipe, which was connected in to the return circuit on each digester. It is very unlikely that significant amounts of heat were transferred along this pipe.

The predicted slurry heating requirement should be close to the actual value, since the mass, temperature and specific heat capacity of the slurry were all known. The main source of error is probably in the calculation of the wall losses, and in particular calculation of the heat transfer coefficient for the digester walls, U_w.

The difficulties in accurately calculating these parameters for digester installations are seen in the work of Oliver (1984) and West (1986). Oliver (1984) calculated the term U_w for a digester installation

($125m^3$) of glass enamelled steel tanks, insulated internally with 75mm of sprayed polyurethane foam. The calculated value of the overall heat transfer coefficient was $0.507W/m^2\ ^oC$. In an experiment to validate this calculation, West (1986) constructed a cooling curve of digester temperature when no heating was applied. The experimentally determined overall heat transfer value for the digester of $140W/^oC$, corresponds to an overall heat transfer coefficient of $0.869W/m^2\ ^oC$, this being 71% greater than the calculated value. This example clearly illustrates the difficulty of calculating heat transfer coefficients for anaerobic digester installations accurately.

The term U_w is so significant because the calculated wall losses for the inclined tubular digesters were 2.1-2.8 times greater than the influent slurry heating requirement. This situation is in contrast to most CSTR installations where the influent slurry heating requirement is the greater of the two components (Oliver, 1984; Stafford et al, 1980) making up the total heat requirement. This difference occurs because wall losses are directly proportional to surface area, the surface area to volume ratio for the inclined tubular digesters of $8.5m^2/m^3$ being considerably greater than that of a CSTR installation where this value is typically $1.30m^2/m3$, for a diameter of 5m. These ratios are not dimensionless, the surface area to volume ratio for a cylindrical tank being $= \frac{4}{d} + \frac{2}{l}$

where d = tank diameter, l = tank length.

A minimum value of the surface area to volume ratio occurs where tank diameter is equal to tank length. In a survey of 39 municipal digesters, Brade and Noone (1981) found the mean aspect ratio varied from 0.40 to 2.0, the mean value being 0.79.

The surface area to volume ratio decreases with increasing diameter, so the effect of wall losses will be less significant in versions of the tubular digester design where the diameter is greater than in the present study.

In calculating the wall losses, it was assumed that the relationship between wall losses and the temperature difference between bulk and ambient temperature was linear. West (1986) suggests that wall losses are probably under-estimated at high ΔT, and over-estimated at low ΔT.

5.1.8 Tracer experiments

Two tracer experiments were conducted, the first when the feedstock was whole slurry, and the second when separated slurry was the feedstock. The fact that the digesters were fed on different feedstocks during the two experiments makes direct comparison difficult.

Additionally, chromic oxide was the only marker used in the first experiment, lithium chloride being used in the second experiment also.

Tracer Experiment 1

Tracer experiment 1, carried out at a time when the digesters were failing, showed there to be large dead volumes within both reactors, and the retention time of components marked by the chromic oxide to be approximately 2 days, compared with a theoretical value of 20 days. During the course of this experiment, the mean gas yields were 0.193 and 0.223 m^3/kg VS added for Digesters 1 and 2 respectively. These values are considerably less than the values reported for Experimental Run 4, under the same conditions, prior to this experiment when the gas yields were 0.282 and 0.286 m^3/kg VS added for Digesters 1 and 2 respectively.

The results from the tracer study question the validity of the results obtained during Experimental Run 4 , particularly for Digester 2 where the mean level of VFA in the discharge was 2725 ppm. It was not possible to determine the rate at which the working volume of the digesters reduced, from the available data. The results for Digester 1 appear acceptable, but Digester 2 had probably been operating at a much reduced retention time during Experimental Run 4, as evidenced by the high concentrations of VFA in the discharge. These results cannot be reliably considered to represent steady - state performance of the digester.

The suitability of Cr_2O_3 as a marker of digester particulate material is questionable, as shown by the low percentage recovery rates.

Tracer Experiment 2

This experiment was conducted during Experimental Run 5 when the digesters were operating under steady - state conditions. The LiCl tracer showed soluble components of the feedstock to be completely mixed throughout the digester in a period of approximately 1 day. The measured retention time of lithium marked components was 15.2 and 13.5 days, for Digesters 1 and 2 respectively. The retention times calculated from collection of the discharged slurry volume were 14.0 days for both digesters. The experiment showed Li to be a good marker of soluble components, and their movement closely resembled that of a completely mixed system.

With chromic oxide, recovery rates were again poor (<50 % in 1 retention time), and it is unclear which component of the digester contents was most closely represented by it's movement. The peak concentrations of Cr occurred 1 day after its' introduction, so some Cr obviously moved through the digester quickly. The low recovery rates of Cr were not surprising considering its' high specific gravity (5.1), and the low solids concentrations within the digester. There was no indication of material moving in true plug flow.

Lithium chloride has been successfully used as a single marker in retention time studies on plug flow digesters by Petersen (1984) and Friman (1985). In both cases the feedstock was dairy cow manure of 10 - 12 % TS.

It is always preferable to use more than 1 marker in tracer studies. If the markers move in similar ways, then the expected results would be similar. It is also preferable to include markers which follow the movement of solid particles, as well as soluble components of the feedstock. When detailed information is known about the movement of different markers relative to each other, then the use of a single marker in tracer experiments can be contemplated.

5.2 Comparison with other work

There are no reported results known of for inclined tubular digesters operating at either the pilot or full scale, on animal wastes. One full scale unit is known of (Himmel, 1984), but no performance data has been reported.

The results presented here will be compared with horizontal plug flow digesters, and continuously stirred tank reactors (CSTR).

Because of the difficulty in identifying the effect of any particular treatment (angle or aspect ratio) on digester performance, and a lack of directly comparable data, the results from this study are grouped according to feedstock, digester temperature and retention time.

Results from Experimental Runs are summarised in the following table, the values presented being the mean of all results obtained under those particular conditions. Results from Experimental Run 4 are not included.

Summary of averaged results from Experimental Runs

	Whole slurry Temperature			Separated slurry Temperature
	35°C		25°C	35°C
GAS YIELDS	Retention time (days)		RT (days)	Retention time (days)
	20	15	15	15
m^3/m^3 day	1.022	1.266	0.705	0.751
$m^3/kg\ TS_A$	0.253	0.236	0.131	0.235
$m^3/kg\ VS_A$	0.303	0.295	0.154	0.305
$m^3/kg\ TS_D$	0.939	0.721	0.676	1.012
$m^3/kg\ VS_D$	1.102	0.962	0.745	1.036

Results obtained by other workers are presented in Table 5.2 for whole slurry, and Table 5.3 for separated slurry. Referring to Table 5.2, the gas yields of 0.295-0.303 m^3/kg VS added in this study for retention times of 20 and 15 days respectively (at 35°C), are comparable with those in the literature. Higher values were reported by Jewell *et al* (1978) for a pilot scale plug flow, with a retention time of 30 days (compared to 20 and 15 days in the present study). In that study at 8.6%

TABLE 5.2 Reported results for the digestion of whole cattle slurry.

REACTOR		Temp	FEED		GAS YIELDS		REDUCTION		REFERENCE
Type	Scale	^oC	%TS	RT days	$m^3/kg\ TS_A$	$m^3/kg\ VS_A$	%TS	%VS	
PF	PILOT	35	7.61	20	0.253	0.303	27.3	28.0	Exptl. Run 1
PF	PILOT	35	8.12	15	0.236	0.295	32.7	31.4	Exptl. Run 2
PF	PILOT	25	7.66	15	0.131	0.154	19.6	20.9	Exptl. Run 3
CSTR	FULL	35	~8	20	0.230	0.288a	-	-	Oliver *et al*, 1986.
CSTR	FULL	35	~8	15	0.200	0.250a	-	-	
CSTR	FULL	25	~8	20	0.138	0.173a	-	-	.. ..
CSTR	FULL	25	~8	15	0.120	0.150a	-	-	.. ..
PF	PILOT	35	8.61	30	0.309	0.357	43.4	49.1	Jewell *et al*, 1978.
PF	PILOT	35	11.2	30	0.287	0.333	21.3	25.7	
PF	PILOT	35	11.2	15	0.209	0.236	22.3	27.1	Jewell *et al*, 1980.
PF	PILOT	35	13.2	30	0.245	0.275	25.5	27.9	
PF	FULL	35	12.9	15	0.293	0.337	29.5	34.1	Jewell *et al*, 1980.
PF	FULL	35	11.2	30	0.316	0.364	35.4	40.6	
PF	FULL	25	11.9	30	0.229	0.260	28.2	32.0	.. ..
CSTR	FULL	35	12.9	15	0.244	0.281	24.2	27.8	.. ..
PF	FULL	37	10.5	15	0.11	0.14	29.5	33.0	Friman, 1984.
PF	FULL	33	11.4	20	0.12	0.15	17.0	19.0	.. ..

a. Volatile solids assumed to be 80% of TS.

TS operation, dry fibrous material accummulated in the reactor and occupied about 40% of the reactor volume. There was no observed solids accumulation when the feed total solids was increased to the "as received" 10-12% TS. This is an important factor regarding digester design, since it is not always practical or feasible to maintain the feed total solids value at 10 - 12 %. The use of an increased feed TS was appropriate in the work of Jewell, since that is the TS at which many dairy farms in the USA handle their wastes. With the same pilot scale digester, Jewell et al (1980) reported gas yields which were lower than the present study.

In a full scale plug flow digester, Jewell et al (1980) again reported higher gas yields (m^3/kg VS added) for operation at 35°C. However, the gas yield from a full scale CSTR operated in parallel with the plug flow unit was lower, and also less than that from the present study.

Gas yields from full scale CSTR installations adjacent to the inclined tubular digesters in the present study have been reported by Oliver et al (1986), and are in all cases lower (0.250-0.288). Very much lower gas yields of 0.14-0.15 were reported by Friman (1984) for a full scale plug flow unit operating on a commercial farm. This is in contrast to all the previously quoted results, since they were obtained from digesters which were pilot and full scale units, but were operated and managed as research facilities.

The gas yield of 0.154 m^3/kg VS added obtained at 25°C is comparable with that of Oliver et al (1986), which was 0.150. Very much higher gas yields at 25°C were found by Jewell et al (1978, 1980), this can be primarily attributed to the 30 day retention time used, compared with 15 days in the present study.

Performance data for the anaerobic digestion of separated cattle slurry are presented in Table 5.3. These results were all obtained at either the laboratory sale or in farm scale CSTR installations. There is no known reported work for the digestion of separated slurry in plug flow digesters, at the farm scale.

Oliver et al (1986) obtained gas yields (m^3/kg VS added) under the same conditions of temperature and retention time, in CSTR installations, which were 11% greater than in the present study. The same trend was found by Hawkes et al (1984) who observed gas yields 3% greater than the present study, and using slurry from the same source. However, the results quoted from Hawkes et al (1984), used slurry which was separated with a belt press type separator (SCS Biotechnology Ltd.), fitted with a 1mm mesh. The separated slurry from this separator contained more finer particles (i.e. <1mm) than that of the present study, where a roller press separator (Farrow Ltd.) with 3mm screen was used.

The effect of mechanical separation on the anaerobic digestion of cattle slurry has been extensively investigated by Peck et al (1985), who suggested that

TABLE 5.3 Reported results for the digestion of separated cattle slurry.

REACTOR Scale	Temp °C	RT days	GAS YIELDS		REFERENCE
			$m^3/kg\ TS_A$	$m^3/kg\ VS_A$	
PILOT	35	15	0.235	0.305	Exptl. Run 5
FULL	35	15	0.290	0.270a	Oliver et al, 1986.
FULL	35	20	0.270	0.338a	
LAB	35	9-10.5	-	0.256	Hawkes et al, 1984.
LAB	35	15-17	-	0.314	
LAB	35	10	-	0.23	Peck et al, 1985.
LAB	35	25	-	0.33	

a. Volatile solids assumed to be 80% of TS.

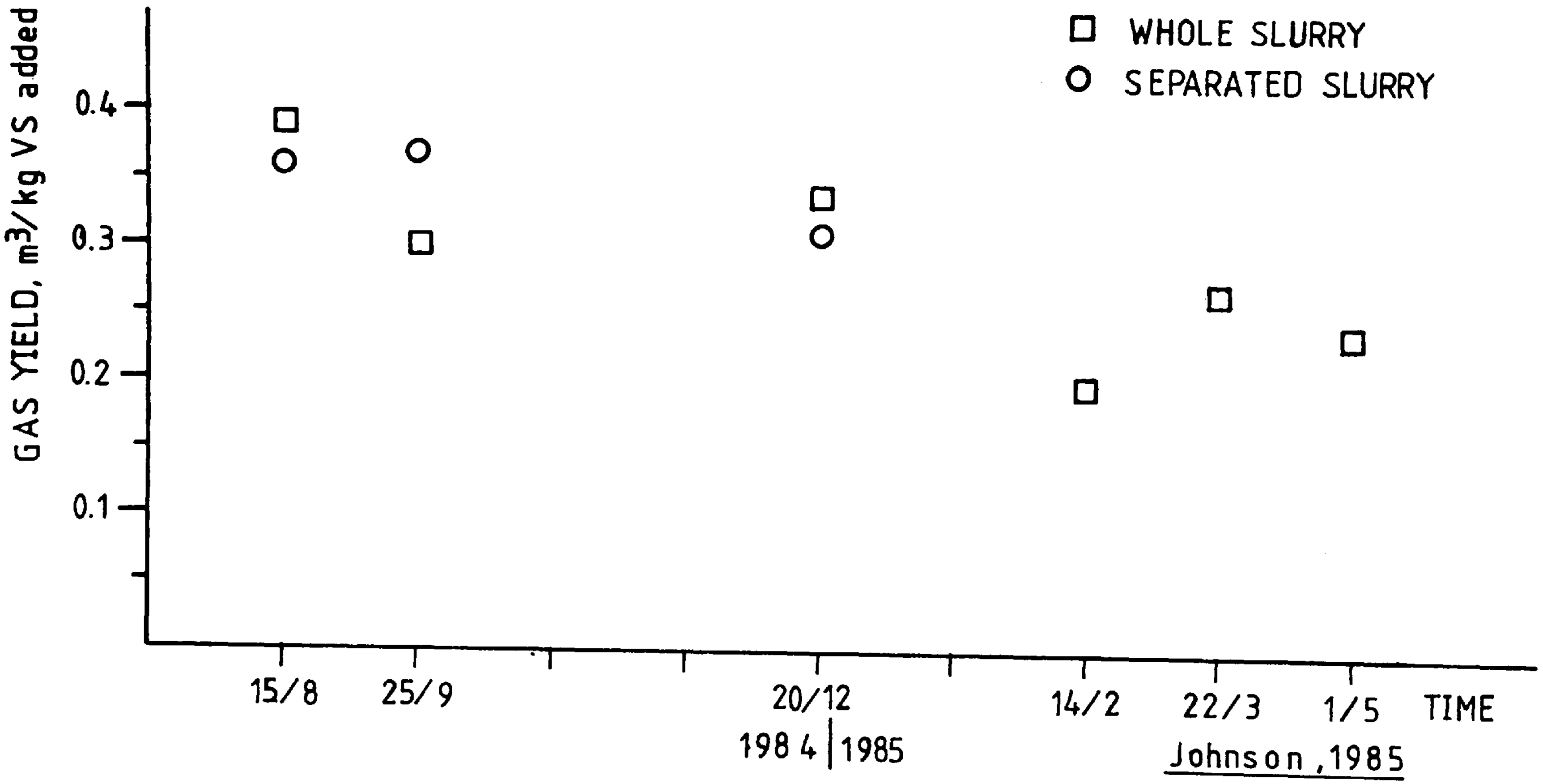
reduction in particle size may play a role in increasing the gas yield. Under laboratory conditions, a gas yield of 0.33m^3/kg VS added was obtained at a 25 day RT, which is 8% greater than this study.

From Tables 5.2 and 5.3, it can be seen that the gas yield from separated slurry is only marginally greater (3%) than that from whole slurry digested under similar conditions. Other workers have generally reported much greater differences in gas yield (m^3/kg VS added) from the digestion of separated, compared with whole slurry. For example, Oliver et al (1986) reported the difference as 26 and 35%, for retention times of 20 and 15 days. Peck et al (1985) reported a gas yield increase of 64% at a 10 day RT. Similar values have also been quoted by Liao and Lo (1985), Lo et al (1983), and Liao et al (1984) and that work particularly showed that the retention time could be reduced significantly (from 16 days to 6 days or less) for separated slurry, on account of its more rapid breakdown. In making these comparisons, it must be recognised that at the same retention time, the loading rate (kg VS/m^{3}d) is greater for whole than separated slurry (typically by a factor of 2). Higher gas yields would be expected at these lower loading rates.

In the present work, studies with whole slurry were carried out in the summer of 1984, and separated slurry in Autumn 1985. It is not possible to regard these two periods as being directly comparable, since gas yield

(m^3/kg VS added) is known to vary during the year, as shown in Figure 5.3 which is the result of batch digestion tests (30 days) from the slurry used in this study (Johnson, 1985).

Further experiments with separated slurry would have been necessary to determine gas yields and solids destruction efficiencies at different (shorter) retention times. The engineering implications of the digestion of separated slurry at retention times shorter than those used for whole slurry are significant. Separation with a roller press machine has been shown (Pain et al, 1978) to remove 18% (by weight) of the original slurry as fibrous solids. If this were done, then the amount of separated slurry entering a digester would be 18% less than the equivalent amount of whole slurry. At the same retention time, digester volume required would be approximately 18% less. Further reductions in the digester volume required could be made by reducing the retention time. These combined effects would lead to significant decreases in the digester volume required for separated slurry.



Johnson, 1985

FIGURE 5.3 Results of 30-day batch digestion experiments with cattle slurry.

CHAPTER 6

EXTRAPOLATION OF INCLINED TUBULAR DIGESTER DESIGN CONCEPT TO FULL SCALE

6.1 Limiting assumptions

Reactor vessels are usually designed to handle a particular waste stream of known volume and composition. The digester is designed for the specific circumstances pertaining to a particular site, and any constraints posed by the site are reflected in the final design. Similarly, a site may offer particular advantages and these would be incorporated in the design for a specific case.

The simplest and most sensible way of avoiding any site specific conditions is to assume a bare or greenfield site. However, it must be recognised that this assumption is not entirely satisfactory since design for a bare site means the digester will not be incorporated into the present waste handling system, and as such it will not represent optimum design for given circumstances.

The composition of the waste stream can also affect digester design. It is assumed that the digester is for a waste of 3.5 - 10% TS, containing large fibrous material. This is typical (See Section 1.4) of the cattle waste produced on many UK farms, and represents the range of TS values used with the pilot scale digesters.

Digesters are usually designed for the specific purpose of pollution reduction, odour control, energy production or a combination of these factors. If the volume and composition of the waste stream were known, then the reactor vessel could be sized. If the primary objective were pollution reduction, then a shorter retention time, and hence smaller digester volume, would be used than if energy production were the main objective. Digester volume could also be reduced if separated slurry were used, since the volume of separated slurry to be treated would be less than the original volume of whole slurry. Alternatively, for a fixed digester volume, the separated slurry could be digested for a longer period than the original volume of whole slurry. Because of these varying objectives, a reactor volume of $100m^3$ will be assumed, since this is the size appropriate to many UK dairy farms. Reactor volume is usually the last design parameter to be determined.

6.2 Process scale up

The present work has shown scale up from laboratory (Floyd, 1984) to pilot scale ($2m^3$) to be technically feasible. These two studies had geometrically similar reactors, which made comparison of performance data more valid than the situation where reactors were not geometrically similar. Since the reation vessel is three-dimensional, as the linear dimensions increase, the system volume increases as the cube of the linear dimensions. Indeed, geometric similarity was the best and probably the only criteria to use for first scale up of this design, since the simplicity of the reactor, with no forced mixing, rendered terms such as specific power input, Reynolds numbers and impeller tip speed, which are commonly used, inappropriate. This was shown by Oldshue (1966) who calculated process parameters normalized to a dimensionless value of 1 for a 20 gallon fermenter, and also for a 2500 gallon fermenter. The parameters calculated were the ratios of power to valume, impeller flow to volume, and the impeller Reynolds number. The wide discrepancies found are shown in the table below:

	Pilot scale 20 US gall	Full scale 2500 gall		
Power	1	125	3125	0.2
Power/Volume	1	1.0	25	0.0016
Impeller flow	1	42.5	125	5.0
Impeller flow/volume	1	0.34	1.0	0.04
Impeller Reynolds Number	1	8.5	25	1.0

Oldshue, 1966

The above table shows some of the difficulties in scaling up direct from laboratory to full scale. Einsele (1978) regarded geometric similarity as one of the desired pre-requisites for scale up, yet in a survey of 30 industrial plants, found that it was rarely achieved in practice.

Another parameter not yet considered is that of temperature. At the laboratory scale, digester temperatures were very uniform (Floyd, 1984). The present study has shown temperature variations to be greater at the pilot scale, the uniformity of temperature distribution decreasing with decreasing angle of inclination to the horizontal, in the range 10-20°.

6.3 Digester configuration

The reactor volume could be made up of 1 reactor of $100m^3$, or several smaller digesters. The main advantages of having one reactor lie in the simplicity of the design, and small number of components which could malfunction. With CSTR reactors over the range 30-100m^3, there are economies of scale whereby cost of digester volume increases by an exponent of 0.682, for increasing digester volume (Oliver et al, 1986).

Advantages of having several smaller digesters include:

a) One digester could be shut-down if the volume of waste decreased significantly (e.g. During the summer with over-wintered dairy cows).

b) A level of waste treatment is maintained if 1 digester had to be shut down for repair or maintenance.

Disadvantages of having more than 1 digester include an increased number of components, and therefore greater cost and likelihood of breakdown.

Other factors affecting choice of digester configuration include load bearing properties of soil which would make the cost of vertical orientation, high aspect ratio digesters, prohibitively high, due to the high cost of the foundations and support structures which would be required. Aesthetic considerations are important in some situations, and the burial or partial burial of digesters is favoured in such circumstances.

6.3.1. Aspect ratio and angle of inclination

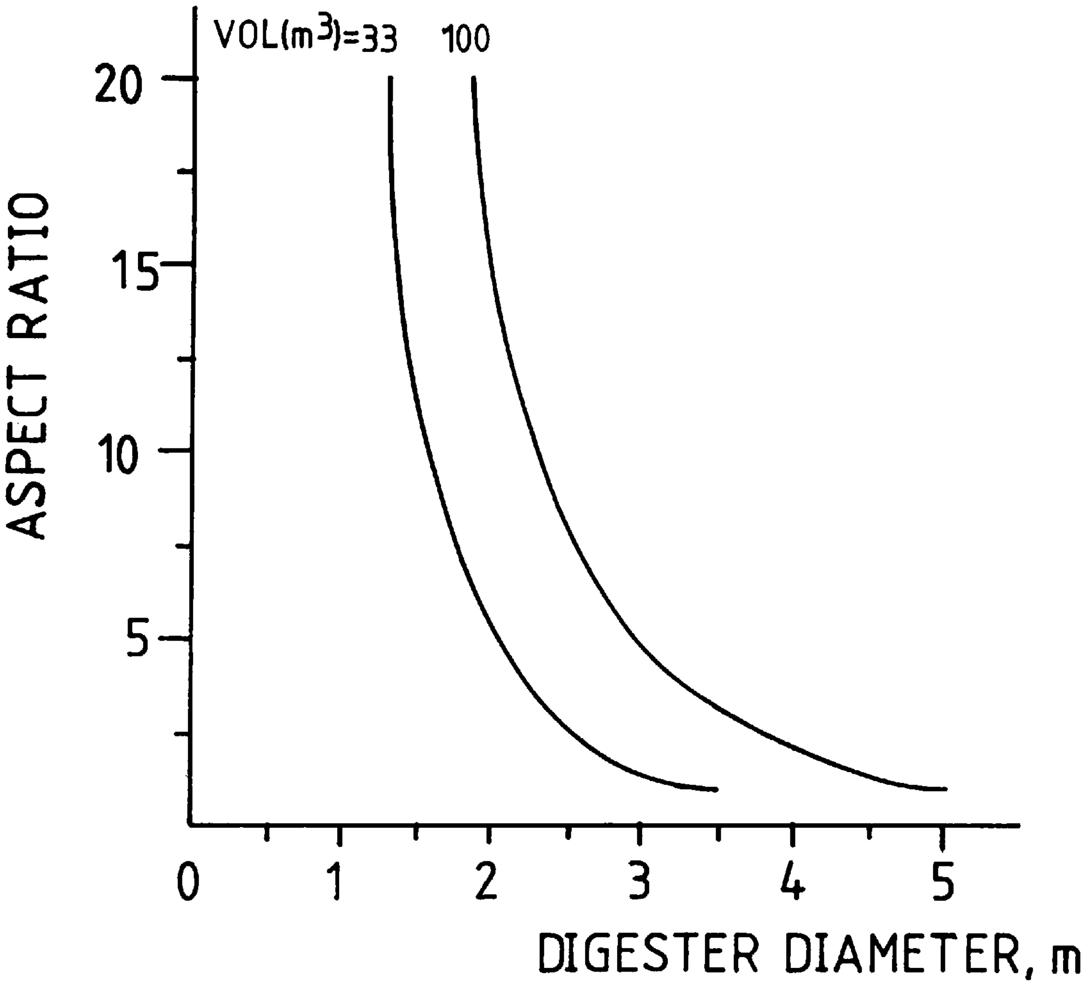
For a constant digester volume, decreasing the aspect ratio increases the tube diameter, as shown in Figure 6.1.

The combined effects of varying aspect ratio and angle of inclination for a constant digester volume can be seen in Figure 6.2, which shows the total length and total height for an inclined tubular digester. Whilst digester length for a particular situation is likely to be of little significance, the converse is true for digester height, especially where the digester is to be wholly or partly buried. The cost of excavation increases with increasing depth, and excavation to a depth greater than 4m is rarely carried out in agricultural situations.

6.3.2 Materials of construction

In selecting the materials to be used for the construction of an anaerobic digester, there are many criteria which must be considered. The most important of these are the design life, resistance to corrosion and durability. The ability to make modifications to a reaction vessel easily is also important, especially with new designs, as is the necessity to be able to stop any gas or liquid leaks. There will usually be several materials (e.g. steel, fibreglass, concrete) which will

FIGURE 6.1 Relationship between aspect ratio and digester diameter for constant digester volume.



meet the above criteria for a particular situation. The final decision is often made on grounds of cost, the cheapest solution generally being preferred, although in some municipal and industrial applications, the material thought to offer the greatest reliability will be selected (Chesshire, 1986). This is because the risk of process failure could be both expensive if alternative process capacity had to be found, and in some instances especially dangerous. The largest individual element of cost is the civil works providing the basic process volume (Brade & Noone, 1981).

The materials which have been most widely used for digester construction in the UK are reinforced concrete, and more recently vitreous enamelled steel. Calculations have shown that a reinforced concrete tank would contain at least as much steel as a similar sized vitreous enamelled steel tank (Noone and Brade, 1982).

Concrete was the favoured material for digester construction when digester design was the domain of civil engineers. However, prefabricated digestion tanks (vitreous enamelled steel, bolted, sectional tanks) have been increasingly used, due to their costing approximately 56% of a comparable reinforced concrete vessel (Noone & Brade, 1982). Additional benefits of prefabricated systems include greatly reduced construction time on site, and because they are constructed entirely above ground level, improved access to the digester for repair and maintenance.

The use of fibreglass (GRP) for digester construction has been pioneered in the UK by Farm Gas Ltd. Again, a modular prefabricated arrangement is used. The main advantages of GRP constructions are ease of construction, assembly and repair. Also GRP is not affected by the corrosive nature of slurry.

In the USA, reinforced concrete and vitreous enamelled steel constructions have been widely used. Additionally, the use of flexible liners for horizontal plug flow digesters has been pioneered by Jewell and his co-workers at Cornell University. The design concepts featured earthen basins that could be graded, insulated and lined with a hypalon rubber material. Hayes et al, (1980) found the cost of such a system to be \$35/m^3 compared with \$100/m^3 for an enclosed rigid wall tank, at the 50-100 cow scale. Even though Jewell successfully demonstrated this concept at pilot, and in several full scale units, uptake from the agricultural industry has been slow. Jewell found (Wood & Chapman, 1986) that farmers and construction companies were generally reluctant to use these unfamiliar construction methods when similar well-proven, yet more costly techniques existed. Also, because the work of Jewell was so comprehensive, commercial companies are unable to construct these low cost designs and still obtain the level of profit margin they would expect from such a project.

6.4 Capital cost estimates

Estimates of digester cost are necessary to determine the economic viability of a digester in a particular situation. In some circumstances, a digester may be judged on the value of the net energy which is produced and effectively utilized. Alternatively, a digester system may be regarded as a waste treatment process, and its economic viability judged on pollution reduction criteria such as BOD and solids reduction.

In seeking to estimate the cost of an inclined tubular digester and ultimately compare with a completely mixed reactor, the accuracy of the estimate is greatest when the number of estimated costs is the least. This can most simply be achieved by only estimating the costs of components which are different in each system. For a typical digester installation, the main elements of the cost for a greenfield site are:

Reaction vessel
Excavations/digester support
Mixing of digester contents
Pumps and associated pipework
Digester insulation
Digester heating/gas utilization
Instrumentation and control
Safety measures

Those elements likely to have different costs for inclined tubular digesters and completely mixed systems are for the reaction vessel, excavations, digester mixing and digester insulation.

The inclined tubular digester will be inclined at 20^{o} to the horizontal, and have an aspect ratio of 10-20. The completely mixed system will be an above ground tank, of aspect ratio close to unity. Provision of the basic process volume is usually the single largest item making up the cost.

It is thought likely that the inclined digester may be partly buried, so excavation costs will be evaluated. The completely mixed reactor usually has gas or mechanical mixing, the tubular digester does not have any forced mixing. Considering a bare site, it is likely that both digester systems would have similar requirements for pump feeding, and associated pipework.

The cost of digester insulation will be directly proportional to the surface area of the reaction vessel, and therefore greater for the high (10-20) aspect ratios of the tubular digester.

It is assumed that the digesters being compared are of equal volume, and that the requirements of a heating system in both types will be similar. Equipment for gas utilization (boilers, generators, combined heat and power (CHP) units) will be the same in both cases, assuming there is a similar production of energy.

Instrumentation for monitoring and control of the digesters will be of similar cost. Safety measures, such as compliance with agreed codes of practice, particularly regarding gas handling, will be the same.

6.4.1 Inclined tubular digesters

Cost estimates were obtained for the basic process volume requirement, using mild steel, helically wound glass fibre and concrete (pre-cast) as the main material of construction. Two estimates were obtained for each reactor material, for approximately $30m^3$ and $100m^3$ total volume. The costs given are ex-works, and exclude delivery to site. There were large differences in the cost/m^2 and cost/m^3, for these two different process volumes, as shown in Table 6.1. At both process volumes, there was less variation in the cost/m^2 (± 29%), than the cost/m^3 which ranged from 20-86% greater for the smaller volume. The table shows that cost/m^2 gives a more consistent cost estimate for these materials, over the range of volumes considered.

From cost considerations, the cheapest material is mild steel. This is suitable for digester construction, although corrosion protection, such as an epoxy paint, may be required. However, since the conditions within the reactor are anaerobic, corrosion is minimal, particularly when the digester contents are low in sulphur compounds (Stafford et al, 1980). A typical mild steel digester would be of factory rolled, all-welded

TABLE 6.1 Manufacturers prices for tubular tanks.

MATERIAL	VOLUME	LENGTH	DIAMETER	ASPECT RATIO	NO. OF PIECES	SURFACE AREA	COST		
							TOTAL	Area	Volume
	m^3	m	m				£	£/m^2	£/m^3
Mild steel	33	18.7	1.5	12.5	2	91.7	3000	32.7	90.9
	145	24.4	2.75	8.9	2	222.7	6000	26.9	41.4
GRP	33	18.7	1.5	12.5	2	91.7	7250	79.1	219.7
	123	25.0	2.5	10	2	201.3	12500	62.1	101.6
Concrete (pre-cast)	31	17.5	1.5	11.7	7	82.5	2590	31.4	83.6
	121	28.0	2.35	11.9	16	206.7	8400	40.6	69.4

construction and delivered to site in 2 parts. A flanged construction (bolted) would be used to join the two parts together. Mild steel is a widely used material and it is easy to work with on site and carry out basic operations such as cutting, drilling and welding.

The cost of pre-cast concrete sections (Table 6.1) for a tubular digester of 30-100m^3 would be similar to that of steel. The pipe would need laying on granular bedding material. Adjacent pipe sections have a flexible rubber sealing ring between them, and specialist equipment is used to draw pipes together when they are being laid (Anon, 1986). Particular care would need to be taken when laying precast pipes for a digester, as the large number of joints would be potential sources of leaks. The internal surface of any concrete section would need treating with a penetrative surface sealant such as magnesium silica fluoride (Belton, 1986) to minimise corrosion, particularly from sulphates. This may at first seem surprising as concrete pipes are widely used without special treatment for sewage transport. However, sewage is retained in well designed sewers for a relatively short time, self cleansing of the sewer and ventilation are incorporated to minimize damage by corrosion arising from anaerobic conditions.

The most expensive material for digester construction in Table 6.1 was GRP. The method of construction would be similar to that of a steel tank, where the tank is factory made. The structural strength is derived from helically wound glass fibre. The reactor would need making in 2 parts to facilitate road transport, and a fibreglass joint (permanent) would be made on site to join the 2 sections. The main advantage of GRP is its excellent corrosion resistance. However, it is not as versatile as steel, and any repairs or modifications would be more difficult.

6.4.1.1 Site preparation

The amount of site preparation necessary will depend mainly on the reactor material chosen, and the method of support. Above-ground structures feature good accessibility for repair and maintenance, but any installation would be unsightly. Additionally, any structure would need at least one support for each length of pipe above ground, depending on the strength of the pipes. A more aesthetically pleasing solution is provided by the burial or partial burial of the digester tank. The extent to which a digester will be buried is very much a site specific constraint. It is unlikely that ground would be excavated deeper than 4m. This is because the cost of excavation increases disproportionately with increasing depth. It is assumed here that any excavation for the digester is to be in firm earth.

For concrete pipes, it is recommended (Anon, 1986) that the overall trench width is 1.5 x Nominal size of pipe (DN). All pipes require suitable bedding material, and it is usual to over excavate by at least 200mm to allow for this. After installation of the pipe, backfilling over the exposed pipe would be carried out.

Figure 6.2 shows that for a $33m^3$ digester, inclined at 20^o to the horizontal, the overall height ranges from 5.4-9.8m, for aspect ratios ranging from 5-20. Similarly, for a $100m^3$ digester, the range is 7.8-14.4m. It will be assumed here that excavation to a depth of 4m is carried out in all cases.

For excavation to a depth of 4m, Figure 6.3 shows the geometric relationships for for a partly buried tubular digester. The general case for the volume of excavation is :

$$\text{Volume} = DN \times 1.5 \times 4m \times (d\sin\theta + 4\cot\theta) \times 0.5.$$

Where DN = Nominal size of pipe

It is assumed that the trench can be suitably graded as it is excavated, and that the ground actually under the digester will not need excavating. An additional 20% has been added to the volume needing excavating, to allow for additional excavation for feed pipes and grading. Using this formula, the following table was obtained:

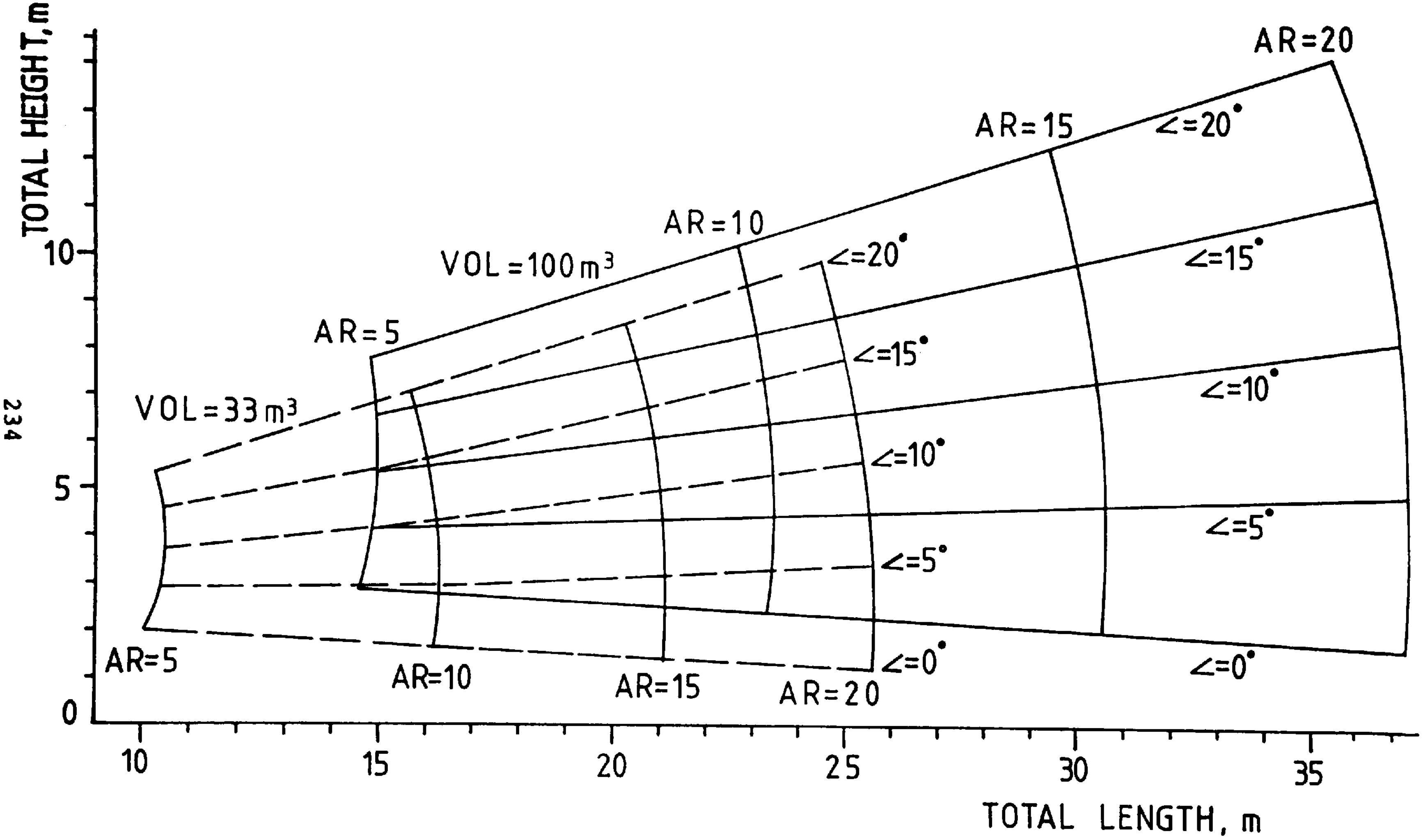
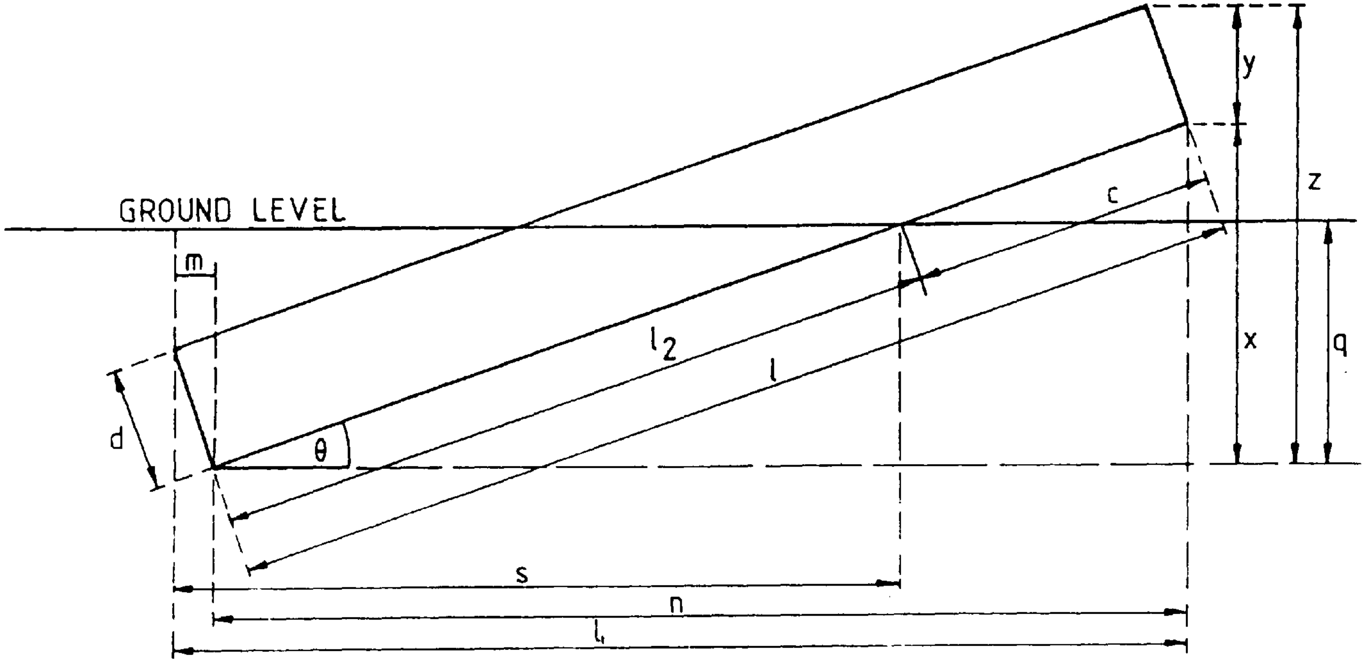


FIGURE 6.2 Inclined tubular digester size envelopes for different aspect ratios and angles of inclination.



DIGESTER DIAMETER $= d$, DIGESTER LENGTH $= l$

OVERALL DIGESTER HEIGHT $= z = x + y = l \sin\theta + d \cos\theta$

OVERALL DIGESTER LENGTH $= l_1 = n + m = l \cos\theta + d \sin\theta$

LENGTH OF EXCAVATION $= s = d \sin\theta + q \cot\theta$, EXCAVATION DEPTH $= q$

LENGTH OF DIGESTER ABOVE GROUND $= c = l - l_2 = l - q \operatorname{cosec}\theta$

FIGURE 6.3 Geometric relationships for a partly buried inclined tubular digester.

Nominal size of pipe(DN) m	Aspect ratio	Digester volume m^3	Volume of excavation m^3	Cost of excavation £	Length of pipe above ground level m
1.5	12.5	33	62	342	7.0
2.35	11.9	121	131	723	12.5
2.5	10	123	140	772	9.5
2.75	8.9	145	118	650	12.7

Assumptions: Depth of excavation = 4m

Digester is inclined at 20° to horizontal.

Cost of excavation = £5.51/m^3 (MAFF, 1986b)

The actual costs of excavation are relatively small, as presented above. This is because excavation is carried out to a constant depth of 4m, or the digester height, whichever is smallest. This means that large parts of the digester need supporting above ground level.

In the cases considered, between 7-12.7m of digester length would need supporting above ground, the corresponding horizontal distances being 6.7 and 11.8 m, repectively. This is shown in Figure 6.4. An earth ramp, suitably constrained or other support structure would add considerably to the total cost. The cost of backfilling over the reactor, and any concreting have not been included.

It has been shown that the costs of excavation are relatively small, and to minimize the length of exposed pipe above ground level, one approach would be to use a

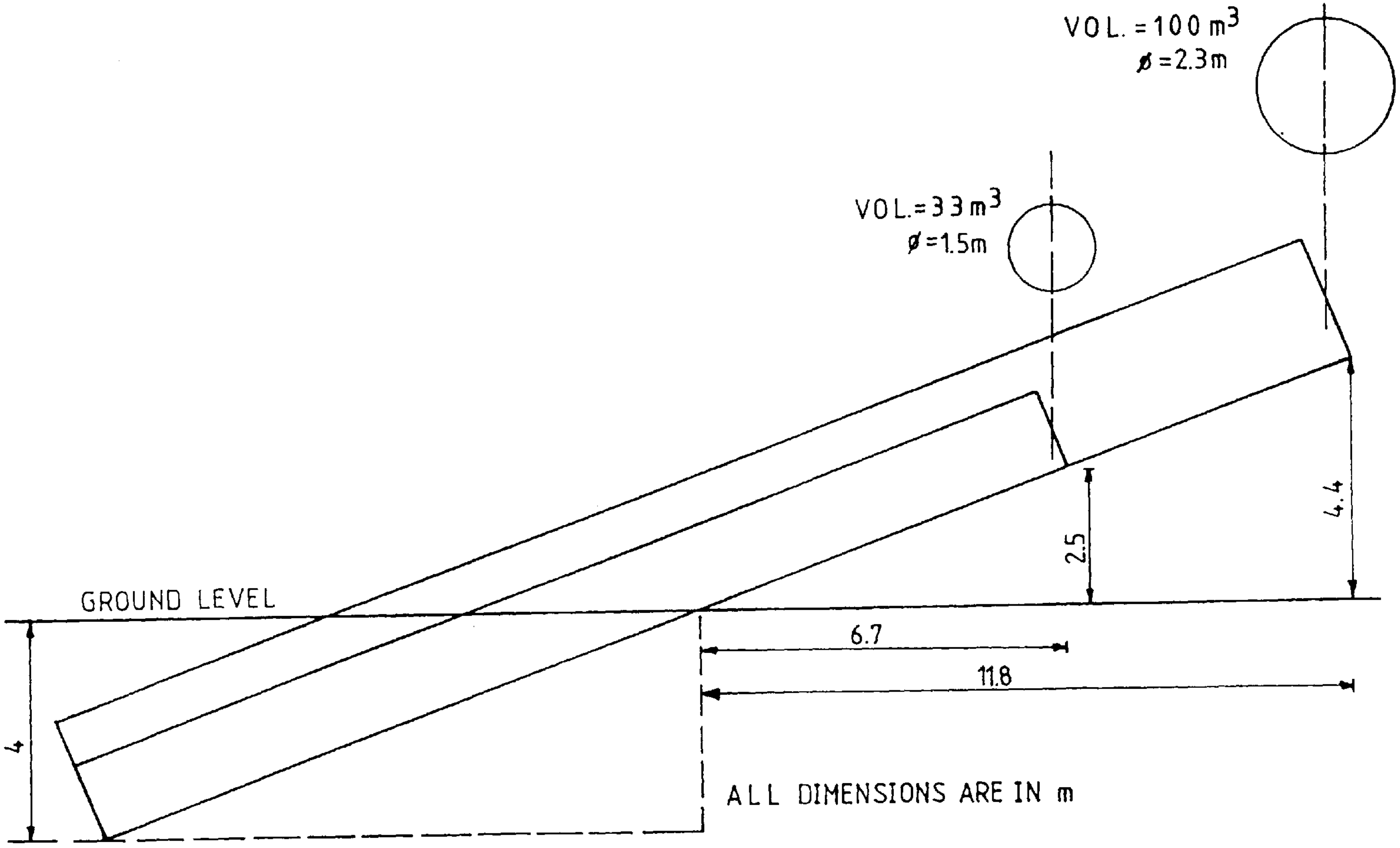


FIGURE 6.4 Scale drawing of partly buried inclined tubular digesters.

larger number of smaller digesters, with the same aspect ratio, to make up the total process volume. This is shown in Table 6.2, and diagramatically in Figure 6.5. The price of a mild steel welded tank was assumed constant at £32.7/m^2 of surface area. Table 6.2 clearly shows that this is not an economically attractive approach, due to the cost of the mild steel required for the large surface area of a multiple digester installation. This calculation does not include insulation costs which, like the cost of steel, are directly proportional to surface area. The prices quoted are for provision of basic tank volume, and do not include any provision for additional fabrication to make items such as effluent weirs.

6.4.2 Completely mixed reactor

Cost estimates were obtained for above ground, vitreous enamelled steel tanks, as are commonly used in prefabricated digester installations. The tank volumes ranged from 78-326m^3, and Figure 6.6 shows tank cost (including roof and erection, but excluding cost of base) to increase linearly in this range.

The digester size closest to 100m^3 had a volume of 97m^3, and diameter 4.6m. The maximum width which can be transported on a lorry without wide load restrictions is 2.9 m. This meant the tank would need to be prefabricated and either welded or bolted on site. The

TABLE 6.2 Excavation and tank costs for inclined tubular digesters in different configurations.

Number of digesters	DIGESTER		Total volume of excavation m^3	Total cost of excavation £	Length of exposed pipe m	Total surface area m^2	Total cost of steel. £
	Diameter m	Length m					
1	2.33	23.33	99.0	545	11.6	179.6	5871
3	1.62	16.18	201.7	1111	4.5	259.2	8474
5	1.37	13.65	281.5	1551	1.9	307.4	10,051
7	1.22	12.2	350.8	1933	0.5	343.9	11,246
9	1.12	11.22	397.0	2187	0	374.1	12,231
11	1.05	10.5	423.6	2334	0	400.0	13,079
13	0.99	9.93	447.1	2463	0	422.9	13,829
15	0.95	9.47	468.4	2580	0	443.6	14,506

Notes: Total process volume = 100 m^3 in each case.
Aspect ratio = 10
Digester inclination = 20^o to horizontal.
Digesters buried to a depth of 4m, or total digester height, whichever is smallest.
Material of construction is mild steel.

FIGURE 6.5 Excavation and tank costs for inclined tubular digesters in different configurations.

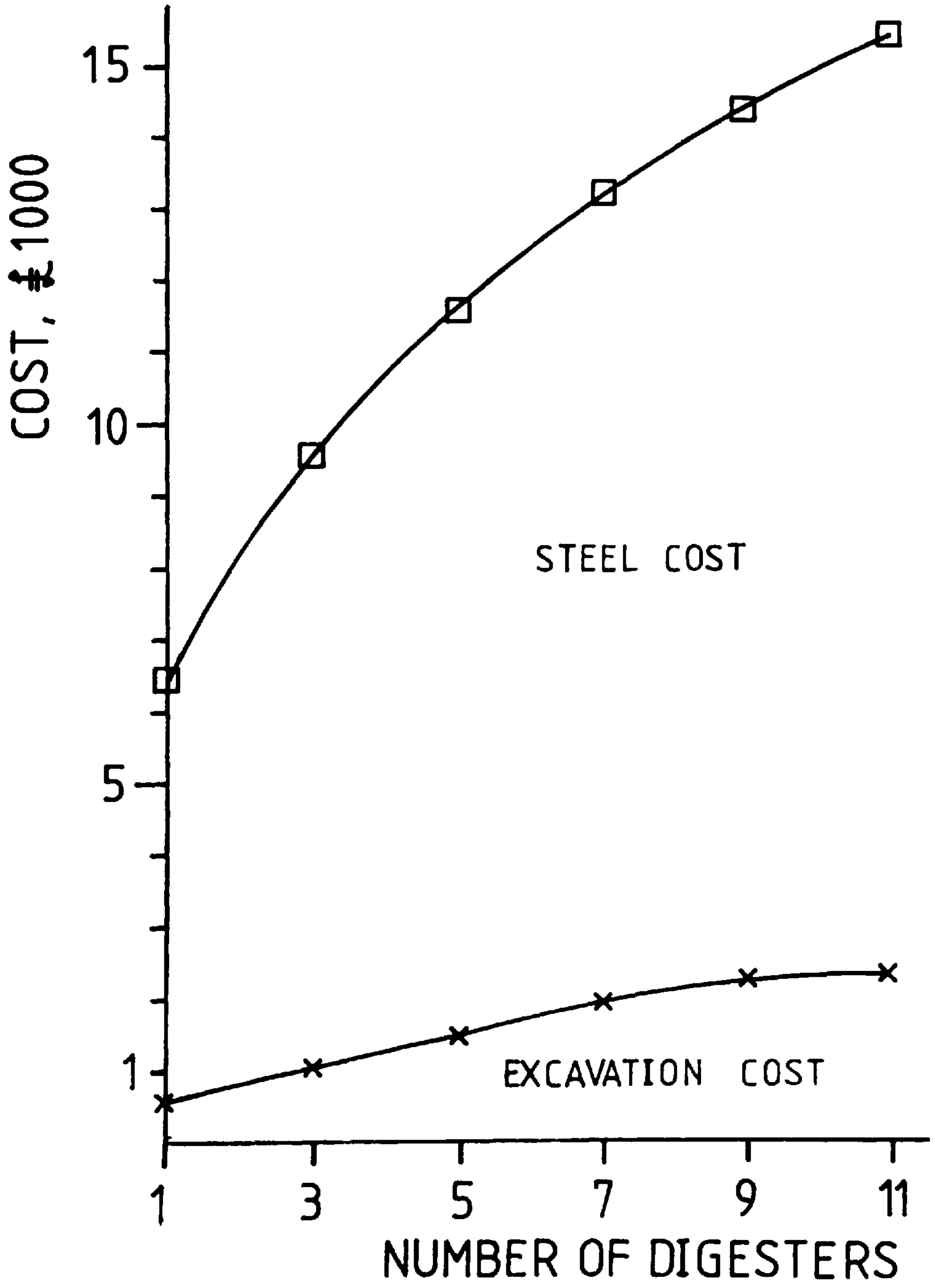
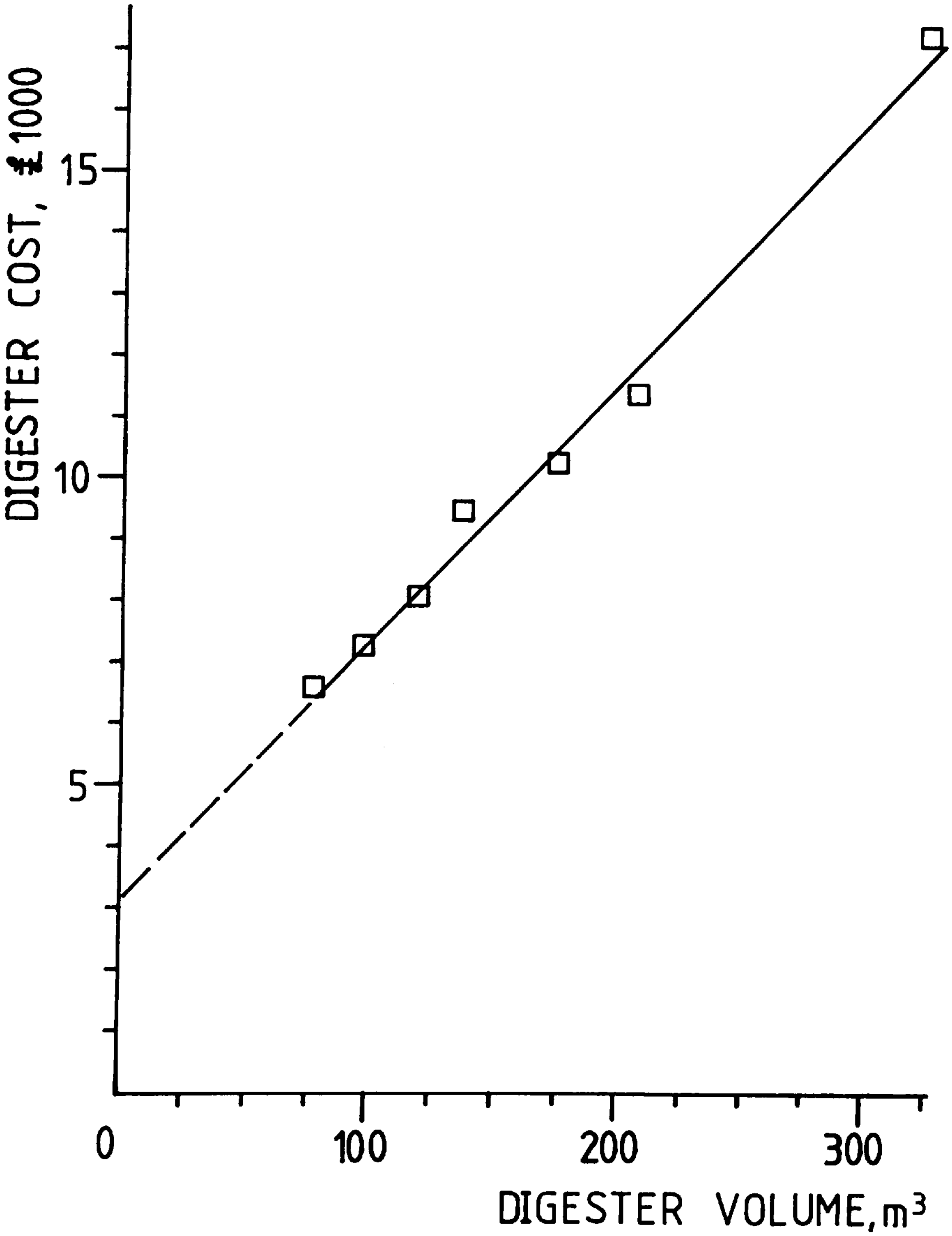


FIGURE 6.6 Relationship between volume and cost for prefabricated digestion tanks (manufacturers prices).



most common method of constructing such tanks on farms is in a bolted sectional manner. Tanks used for forage cropstores and slurry stores are constructed in this way, using essentially the same materials. The pricing of such tanks in the UK is highly competitive, there being 2 main manufacturers. Additionally, there is evidence that tanks are priced according to what the market will stand. Two specifications for tank construction are widely quoted, these being "agricultural" and "industrial" standard. On investigating this "difference", it appears that the tank specifications in both cases are identical, apart from the sealant used between sheets. Industrial specification tanks generally incorporate ladders (external) as standard, whereas agricultural tanks do not.

The linear regression line relating prefabricated digester tank cost and volume was calculated to be:

Total tank cost (£) = 42.2 x (Tank Volume, m^3) + 3183

For a tank of $100m^3$, Total tank cost = £7403, or £$74/m^3$

Because the tanks are constructed from prefabricated sheets, tank volume is increased by either increasing the tank circumference, or the tank height. In each case whole numbers of sheets must be used.

The cost estimate for the $97m^3$ tank had a diameter of 4.6m, and height 5.98m. The quoted cost (erected) was £7314, excluding VAT.

The price of a 33m^3 tank using the regression equation was estimated to be £4575, or £138/m^3. It must be recognised that this is an extrapolation of the regression line which was calculated from tank prices for volumes ranging from 78-326m^3.

6.4.2.1 Site preparation

Prefabricated steel tanks are erected on prepared concrete bases. A flat or slightly dished concrete floor is used. Manufacturer's recommendations for preparing the base for a tank 4.6mØ, 5.98m high are as follows:

	£/m^3	TOTAL FOR TANK
1) Excavate trench for footings, 500mm depth, 900mm wide	5.07	16.41
2) Excavate whole area to a depth of at least 250mm	1.92	7.91
3) Pour footings or "ring beam", 300mm depth	42.38	137.17
	£/m^2	
4) Lay insulation in base, 50mm thick	5.25	86.49
5) Lay floor mesh, 6.16kg/m^2	2.94	48.44
6) Pour base, 200mm depth	11.30	186.17
	TOTAL	£ 482.59

Note: Excavations + concrete are to the tank manufacturers specification. All prices quoted are estimates taken from the MAFF "Cost of Buildings Handbook" (MAFF, 1986b).

The total cost for the base is £482.59, or £29.29/m^2 of concreted digester base. Assuming the same unit cost for a 33m^3 digester, aspect ratio = 1, diameter = 3.47m, the base cost would be £277.

There is considerable experience in building prefabricated steel tanks on correctly prepared bases. The main restrictions regarding siting of such tanks are that the load bearing capacity of the ground (must be >50 kN/m^2), and the site must be well drained. The safe bearing pressure required for tanks with more than 4 rings is higher (100 kN/m^2), and it is necessary to use a stronger concrete base (more floor mesh or different specification concrete).

6.4.3 Cost comparison of inclined tubular and completely mixed reactors

The manufacturers price estimates for the 33m^3 digester tanks were:

	Tank cost (£)	Unit Cost £/m^3
Inclined digesters		
Mild steel	3000	90.9
Concrete (pre-cast)	2590	78.5
GRP	7250	219.7
Completely mixed		
Prefabricated, glass enamelled steel tank. (estimated) Aspect ratio = 1	4575	138.6

Notes

1. For the inclined digesters, there were excavation charges of £342, and 7.0m of pipe above ground level, the support of which has not been costed.

2. For the completely mixed reactor, there was a base preparation cost of £277.

It is not possible to directly compare the actual prices for the inclined digesters, with the estimated cost for a completely mixed reactor, due to the different materials of construction. However, the prefabricated completely mixed reactor is a superior system, on account of the vitreous enamelled finish to the steel. Although the tank prices for mild steel concrete are cheaper, they do not include any cost for installation, and more significantly, support of the 7m length of digester above ground. The cost of installing a 7-piece concrete digester would add considerably to the digester price.

The price of the inclined tubular GRP digester was greater than that for the completely mixed reactor, and it is very unlikely that GRP would be selected for the tank material.

For reactors of approximately $100m^3$, the following manufacturers prices were obtained:

Inclined digesters	Volume m^3	Tank Cost £	Unit Cost £/m^3
Mild steel	145	6000	41.4
Concrete (pre-cast)	121	8400	69.4
GRP	123	12500	101.6
Completely mixed			
Prefabricated, glass enamelled steel tank	97	7314	75.4

Notes

1. For the inclined digesters, there were excavation charges of £650-£772. The length of pipe above ground level was 9.5-12.7m. No costs for supporting that pipe have been included.

2. For the completely mixed reactor, there was a base preparation cost of £482.

For tanks of approximately $100m^3$, construction costs per m^3 are about 50% of the unit cost for $33m^3$ tanks, showing economies of scale, with the exception of concrete as a material of construction. Concrete was 12% cheaper at the $100m^3$ scale.

The cheapest material was again mild steel. The tank had an aspect ratio of 8.9, and 12.7m of the tank needed supporting above ground level. When the cost of this additional support is included, it is likely that the total cost of the mild steel system would be greater than the completely mixed, prefabricated glass enamelled steel tank. The mild steel digester would need transporting as 2 loads because of its size. Again, the prefabricated system consists of coated mild steel which is known to be resistant to corrosion. It is unlikely that the mild steel tubular digester could receive a similar quality surface treatment, due to logistical problems.

The inclined tubular concrete system had 16 sections. There would be large extra costs for assembly, and support of the pipe above ground level. This support would need to be continuous, or, there would have to be at least one support per length of pipe.

The GRP system is financially unattractive, the basic tank costing 35% more than an erected, prefabricated glass enamelled steel tank.

The completely mixed reactor system would need a mixing system which has not been costed. The inclined tubular digester has no forced mixing, and would not require a mixing system. This is a cost benefit to the inclined tubular design, both as a capital cost, and also as an operating cost in terms of power consumed, repairs and maintenance.

6.5 Discussion

The feasibility of scaling up the inclined tubular digester to farm scale has been considered, the major limitations being those associated with the supporting of large (>$100m^3$) reactors. Support structures for this purpose could be made, but the cost would be prohibitively high compared with the expected benefits. The construction of completely mixed reactors using prefabricated tanks is well established technology, and even if inclined tubular digesters could be built at the 100 m^3 scale for a similar cost, process efficiency would probably need to be greater by a factor of 1.5 - 2, before widespread adoption and departure from conventional structures could be envisaged.

An aspect ratio for the tubular digesters of approximately 10 was chosen for the cost comparisons since the the experimental results showed digester performance to be greatest at this value. It is possible that the same or increased gas yields would be obtained with digesters of aspect ratio < 10. This would have the advantage of reducing the surface area, and hence heat losses, in relation to digester volume, and for a particular volume, digester length would be reduced. This is important because of the need for extensive support structures which are required for inclined tubular digesters at the farm scale. The constructional problems, and hence cost, associated with high (20) aspect ratio digesters would be greater.

Digester performance at 10° inclination was less stable with whole slurry, compared to 20° inclination, and was considered to be too unreliable for a full scale unit. Implications regarding reliability are important at the full scale since process failure would be time consuming and expensive to rectify. The cost of any alternative treatment process during a period of digester failure would have to be ascribed to the digestion system.

CHAPTER 7

CONCLUSIONS

Two pilot scale inclined tubular digesters were designed and constructed adjacent to a large dairy unit. The digesters and associated system of instrumentation fulfilled the deign criteria, and functioned reliably over a period of 20 months.

The maximum gas yield was 0.318 m^3/kg VS added, this value being obtained under the following conditions : Digester inclination of 20^o to the horizontal, an aspect ratio of 10, feedstock of whole slurry, operating temperature of 35^oC and a retention time of 20 days. Digester performance was similar to that normally reported for CSTR type designs.

At a digester inclination of 20^o, gas yields (m^3/kg VS added) from whole slurry were higher at an aspect ratio of 10 than at 20 for both 25 and 35^oC operating temperatures. At 10^o inclination, gas yields were marginally lower at an aspect ratio of 10 for both whole and separated slurry.

Stability of the digestion process was greatest at 20° inclination. This was attributed to the more uniform temperature distribution throughout the reactor due to a greater degree of mixing caused by movement of the evolved biogas and convection currents.

There was no evidence of scum and crust formation inside the digesters. Minor problems with crust formation in the effluent box <u>external</u> to each digester were experienced, but were overcome by re-designing the effluent weir.

At an inclination of 10° , digester performance on whole slurry was particularly unstable, and there was a progressive decrease in gas production. A tracer study showed there to be considerable accumulation of solid material in both digesters.

At 10° inclination with separated cattle slurry, digester performance was stable. A tracer study showed the digester contents were well mixed, and the flow of soluble feed components through the digesters was similar to a completely mixed system.

The measured digester heating requirement exceeded gross energy production due to the experimental nature of the plant. The largest component of the heating requirement was to overcome wall losses. Calculations showed that this situation would be improved in farm scale versions of this design, where the digester surface area to volume ratio would be lower.

A technical and economic appraisal established that the design could be scaled up to farm scale ($>100m^3$), but that major problems would arise from the need to support a reactor of this size at the optimum inclination of 20^o, even if the digester was partly buried. This was reflected in estimates of capital cost. It was estimated that it would be more expensive to use a number of smaller digesters to make up the total process volume, and that there would still be extensive requirements for support structures. However, a number of smaller digesters would afford greater flexibility, and enable total process volume to be adjusted should there be variations in the volume of waste to be treated.

The inclined tubular digester is a simple design requiring a minimal amount of daily labour input. There is no requirement for forced mixing. The main applications of this design are likely to be for treating particulate wastes of <8% TS concentration, where some settling will occur, leading to increased retention of solid components in the feedstock.

REFERENCES

ANON (1963)
Methane Production and Effluent Disposal on the Farm.
Wright Rain Technical Bulletin B232.

ANON (1972)
Analysis of Raw, Potable and Waste Waters.
London : H.M.S.O.

ANON (1986)
Concrete Pipes for Drainage and Sewerage.
Publication 73.357-0186
Leicester: Concrete Pipe Association.

BABA (1982)
A code of practice on safety in and around anaerobic digesters.
Edition 1. British Anaerobic and Biomass Association Ltd., Marlborough, Wilts. 20 pages.

BARRY, M.; COLLERAN, E. and WILKIE, A. (1982)
Two-stage Anaerobic Digestion of Organic Residues and Energy Crops.
pp. 75-87 In Vogt, F. (ed) Energy Conservation and use of Renewable Energies in the Bio-Industries 2.
Oxford: Pergamon Press.

BELTON, N.B.G. (1986)
Personal communication.
ADAS Farm Buildings Group,
MAFF, Coley Park, Reading.

BRADE, C.E. and NOONE, G.P. (1981)
Anaerobic Sludge Digestion - Need it be Expensive?
Making More of Existing Resources.
Water Pollution Control, 80, (1), 70-94.

BRAUN, R. and HUSS, S. (1982)
Anaerobic Digestion of Distillery Effluents.
Process Biochemistry 17, (4), 25-27.

BRUCE, A.M. & OLIVER, B. (1986)
Heating and Cooling of Sewage Sludges - Some recent Developments.
Institute of Water Pollution Control. In Press.

BRYANT, M.P. (1979)
Microbial methane production - Theoretical aspects.
J. of Animal Science, 48, (1), 193-201.

BUHLERT, J.E.; YORK, G.K. and LEWIS, M.J. (1981)
Demonstration of the performance of an inclined anerobic digester in wine stillage and pea blancher wastewater treatment.
J. Food Science, 46, 1747-1750.

BUSWELL, A.M. and BORUFF, C.S. (1933)
Mechanical Equipment for Continuous Fermentation of Fibrous Materials.
Industrial and Engineering Chemistry, 25, (2), 147-149.

CALLANDER, I.J. and BARFORD, J.P. (1983)
Recent Advances in Anaerobic Digestion Technology.
Process Biochemistry, 18, (4), 24-30, 37.

CASEY, T.J. and POWER, J. (1984)
Development of a plug-flow digester for sewage sludge.
pp. 454-456 In Ferrero, G.L.; Ferranti, M.P. and Naveau, H. (eds.) Anaerobic Digestion and Carbohydrate Hydrolysis of Waste.
Barking: Elsevier Applied Science Publishers. 517 pages.

CASEY, T.J. (1986)
Requirements and Methods for Mixing in Anaerobic Digesters.
pp. 90-103, In Bruce, A.M.; Kouzeli-Katsiri, A. and Newman, P.J. (eds.) Anaerobic Digestion of Sewage Sludge and Organic Agricultural Wastes. Barking: Elsevier Applied Science Publishers.

CHEN, J.S.; FREY, B.C. and BRODIE, H.L. (1983)
Solids concentration effect of scum formation in plug-flow digester.
American Society of Agricultural Engineers,
Paper No. 83-4055.

CHEN, J.S.; FREY, B.C. and BRODIE , H.L. (1984)
Scum Accumulation in Plug-Flow Digesters - A Problem with Flushed Dairy Manure.
Transactions of the ASAE., 26,(6), 1894-1896.

CHESSHIRE, M. (1986)
A Comparison of the Design and Operational Requirements for the Anaerobic Digestion of Animal Slurries and of Sewage Sludge.
pp. 33-54 In Bruce, A.M.; Kouzeli-Katsiri, A. and Newman, P.J. (eds) Anaerobic Digestion of Sewage Sludge and Organic Agricultural Wastes.
Barking: Elsevier Applied Science Publishers.

COHEN, A. (1983)
Two-phase digestion of liquid and solids wastes.
pp. 123-138 In Anaerobic Digestion 1983. Proc. 3rd Intnl. Symp. on Anaerobic Digestion, Boston, USA.
Cambridge, Mass. : The 3rd Intnl. Symp. on Anaerobic Digestion.

COURNOYER, M.S.; DELISLE, U. and FERLAND, D. (1984)
A mixed plug flow anaerobic digester for dairy manure.
American Society of Agricultural Engineers,
Paper No. 84-4562.

CROCKER, S.A. (1985)
Integrated farm waste management systems.
pp. 97 - 114 In Pain, B.F. and Hepherd, R.Q. (eds.) Anaerobic Digestion of Farm Waste. Technical Bulletin 7.
Reading: The National Institute for Research in Dairying.

DEMUYNCK, M.G.; NAVEAU, H.P. and NYNS, E.J. (1982)
Assessment Study on Biogas Plants in Europe.
pp. 602-605 In Strub, A.; Chartier, P. and Schlesser, G. (eds.) Energy from Biomass. Proc. 2nd E.C. Conference, Berlin. London: Applied Science Publishers.

DEMUYNCK, M.; NYNS, E.J. and PALZ, W. (1984)
Biogas Plants in Europe.
Solar Energy R & D in the European Community.
Series E. Volume 6. Energy from Biomass.
Dordrecht: D. Reidel Publishing Company. 339 pages.

DODSON, C.E. (1981)
Anaerobic Treatment of Food Processing Wastes and Agricultural Effluents.
pp. 85-91 In Herzka, A. and Booth, R.G. (eds.) Food Industry Wastes: Disposal and Recovery.
Barking: Applied Science Publishers.

EINSELE, A. (1978)
Scaling Up Bioreactors.
Process Biochemistry, 13, (7), 13-14.

ELLIS, W.C.; LASCANO, C.; TEETER, R. and OWENS, F.N. (1982)
Solute and Particulate Flow Markers.
pp. 37-56 In Owens, F.N.(ed) Protein Requirements for Cattle. Proc. of an International Symp., Publication MP109, Division of Agriculture, Oklahoma State University. 363 pages.

FANFONI, K.J.; HAYES, T.D. and JEWELL, W.J. (1978)
Simplified Fermentor Development.
pp. 73 - 161 In JEWELL, W.J.; CAPENER, H.R.; DELL'ORTO.S; FANFONI, K.J.; HAYES, T.D.; LEUSCHNER, A.P.; MILLER, T.L.; SHERMAN, D.F.; VAN SOEST, P.J.; WOLIN, M.J. and WUJCIK, W.J. (1978) Anaerobic Fermentation of Agricultural Residue : Potential for Improvement and Implementation. Final Report. U.S. Dept. of Energy Report No. HCPT/2981-07. 427 pages.

FERRARA, R.; BARBERIS, R.; JODICE, R.; VICENZINO, E. and VANNI, A. (1984)
Influence of Kinetics of Hydrolysis, Acidogenesis and Methanogenesis on Enhancement of the Production of Biogas from Animal Excreta.
Agricultural Wastes, 11, 79-90.

FLOYD, J.R.S. (1984)
Operation of a tubular anaerobic digester on piggery waste.
Unpublished PhD. The Polytechnic of Wales.

FLOYD, J.R.S. and HAWKES, F.R. (1986)
Operation of a laboratory-scale tubular digester on piggery waste.
Agricultural Wastes,18,(1),39-60.

FRIMAN, R. (1984)
Monitoring Anaerobic Digesters on Farms.
J. agric. Engng. Res., 29, 357-365.

FRIMAN, R.M. (1985a)
The ADAS Mobile Monitoring Laboratory.
pp. 141-148, In Pain, B.F. and Hepherd, R.Q. (eds.) Anaerobic Digestion of Farm Waste. Technical Bulletin 7. Reading: The National Institute for Research in Dairying.

FRIMAN, R.M. (1985b)
Personal communication.
ADAS Farm Waste Unit, MAFF, Coley Park, Reading.

FRIMAN, R.M. (1986)
Anaerobic Digestion on Farms in the United Kingdom.
pp. 135-144 In Bruce, A.M.; Kouzeli-Katsiri, A. and Newman, P.J. (eds.) Anaerobic Digestion of Sewage Sludge and Organic Agricultural Wastes.
Barking: Elsevier Applied Science Publishers.

FRY, L.J. (1974)
Practical Building of Methane Power Plants for Rural Energy Independence.
Santa Barbara, California: Standard Printing. 96 pages.

GHOSH, S. and KLASS, D.L. (1978)
Two-phase anaerobic digestion.
Process Biochemistry, 13, (4), 15-24.

GILMAN, F.E. and BENNETT, S. (1984)
Northern New England's Dairy Manure Digesters.
American Society of Agricultural Engineers,
Paper No. 84-4558.

GRUNDEY, K. (1980)
Tackling Farm Waste.
Ipswich: Farming Press Ltd. 245 pages.

HAMAD, M.A.; ABDEL DAYEM, A.M. and EL HALWAGI, M.M. (1983)
Evaluation of the performance of two rural biogas units of Indian and Chinese design.
Energy in Agriculture, 1, (3), 235-250.

HAO, P.L.C.; TANG, S.H.; HSU, W.W. (1979)
PVC Red-Mud Compositions.
U.S. Patent 4,161,465

HAWKES, D.L. (1980)
Factors Affecting Net Energy Production from Mesophilic Anaerobic Digestion.
pp. 131 - 150 In Stafford, D.A.; Wheatley, B. and Hughes, D.E. (eds.) Anaerobic Digestion. Proc. 1st International Symposium on Anaerobic Digestion, Cardiff, UK.
London : Applied Science Publishers.

HAWKES, F.R.; FLOYD, J.R.S. and HAWKES, D.L. (1981)
Operation of a laboratory-scale plug flow type of digester on pig manure.
pp. 398-405, In Palz, W.; Chartier, P. and Hall, D.O. (eds) Proc. 1st E.C. Conference on Energy from Biomass.
London: Applied Science.

HAWKES, F.R.; ROSSER, B.L.; HAWKES, D.L. and STATHAM, M. (1984)
Mesophilic Anaerobic Digestion of Cattle Slurry After Passage Through a Mechanical Separator : Factors Affecting Gas Yield.
Agricultural Wastes, 10, 241-256.

HAWKES, D.L. (1985)
Digester design, operation and performance.
pp. 29 - 40 In Pain, B.F. and Hepherd, R.Q. (eds.) Anaerobic Digestion of Farm Waste. Technical Bulletin 7.
Reading: The National Institute for Research in Dairying.

HAYES, T.D.; JEWELL, W.J.; DELL'ORTO, S.; FANFONI, K.J.; LEUSCHNER, A.P. and SHERMAN, D.P. (1980)
Anaerobic digestion of cattle manure.
pp. 255-288 In Stafford, D.A.; Wheatley, B. and Hughes, D.E. (eds.) Anaerobic Digestion. Proc. 1st International Symposium on Anaerobic Digestion, Cardiff, UK.
London : Applied Science Publishers.

HERMANSON, R.E. (1985)
Flush Cleaning Dairy Barns - Case Studies.
pp. 590-597 In Agricultural Waste Utilization and Management. Proc. 5th Intnl. Symposium on Agricultural Wastes, Chicago, USA.
St. Joseph, Michigan: American Society of Agricultural Engineers.

HILL, W.D.; ROTHFUS, R.R. and LI, K. (1977)
Boundary - Enhanced Sedimentation due to Settling Convection.
Intnl. J. of Multiphase Flow, 3, 561-583.

HILLS, D.J. (1983)
Intermittently Mixed and Plug Flow Digestion of Beef Feedlot Manure at Various Water Contents.
Transactions of the ASAE., 26 (3), 884-889.

HIMMEL, W. (1984)
Personal Communication.
Technische Universitat, Graz, Austria.

HOBSON, P.N. (1983)
The Kinetics of Anaerobic Digestion of Farm Wastes.
J. Chem. Tech. Biotechnol., 33B, 1-20.

HOBSON, P.N.; BOUSFIELD, S. and SUMMERS, R. (1974)
The anaerobic digestion of organic matter.
Critical Reviews in Environmental Control, 4, (2), 131-191.

HOBSON, P.N.; BOUSFIELD, S. and SUMMERS, R. (1981)
Methane Production from Agricultural and Domestic Wastes.
Barking : Applied Science Publishers. 259 pages.

HOBSON, P.N. and RICHARDSON, A.J. (1985)
The Microbiology of Anaerobic Digestion.
pp. 15-28, In Pain, B.F. Heperd, R.Q.(eds.) Anaerobic Digestion of Farm Waste. Technical Bulletin 7. Reading : The National Institute for Research in Dairying.

HORTON, R. (1980)
The Implications of Engineering Design on Anaerobic Digester Systems.
pp. 321-343 In Staford, D.A.; Wheatley, B.I. and Hughes, D.E. (eds.) Anaerobic Digestion.
Proc. 1st International Symposium on Anaerobic Digestion, Cardiff, UK. London: Applied Science Publishers.

HULL-BANG DAO (1974)
Production et utilizations du Gay de fumier. Methane Biologique,
CNEEMA Biu°200.

JEWELL, W.J.; DAVIS, H.R.; GUNKEL, W.W,; LATHWELL, D.J.; MARTIN, J.H.; McCARTY, G.R.; MORRIS, G.R.; PRICE, D.R. and WILLIAMS, D.W. (1976)
Bioconversion of Agricultural Wastes for Pollution Control and Energy Conservation. U.S. Energy Research and Development Administration, Report No. T1B 27164.
Available from National Technical Information Service (NTIS) , Springfield, Va.

JEWELL, W.J.; CAPENER, H.R.; DELL'ORTO.S; FANFONI, K.J.; HAYES, T.D.; LEUSCHNER, A.P.; MILLER, T.L.; SHERMAN, D.F.; VAN SOEST, P.J.; WOLIN, M.J. and WUJCIK, W.J. (1978)
Anaerobic Fermentation of Agricultural Residue : Potential for Improvement and Implementation. Final Report. U.S. Dept. of Energy Report No. HCPT/2981-07. Available from NTIS. 427 pages.

JEWELL, W.J. (1980)
Future Trends in Digester Design.
pp. 467-491 In Stafford, D.A.; Wheatley, B.I. and Hughes, D.E. (eds.) Anaerobic Digestion.
Proc. 1st International Symposium on Anaerobic Digestion, Cardiff, UK.
London: Applied Science Publishers.

JEWELL, W.J.; DELL'ORTO, S.; FANFONI, K.J.; HAYES, T.D.; LEUSCHNER, A.P. SHERMAN, D.F. (1980)
Anaerobic fermentation of agricultural residue: Potential for improvement and implementation.
Volume II. U.S. Dept. of Energy Final Report, Project No. DE-AC02-76ET20051. Available from NTIS. 599 pages.

JEWELL, W.J.; CUMMINGS, R.J.; KABRICK, R.M. METZGER, J.A. (1981a)
Long-term operational comparison of two full scale dairy manure digesters.
U.S. Dept. of Energy Report XB-0-9038-1-8. Available from NTIS. 107 pages.

JEWELL, W.J.; KABRICK, R.M.; DELL'ORTO, S.; FANFONI, K.J. and CUMMINGS, R.J. (1981b)
Earthen-Supported Plug Flow Reactor for Dairy Applications.
pp. 1-24 In Proc. Methane Technology for Agriculture Conference, Ithaca, New York. Ithaca: Cornell University.

JEWELL, W.J.; ADAMS, B.A.; ECKSTROM, B.P.; FANFONI, K.J.; KABRICK, R.M. and SHERMAN, D.F. (1982)
The Feasibility of Biogas Production on Farms.
Solar Energy Research Institute Report No. XB-0-9038-1-10. Available from NTIS. 163 pages.

JOHNSON, A.P. (1985)
Farm-scale and laboratory studies on anaerobic digestion of dairy cow slurry.
Industrial Training Report.(unpublished).
The Animal and Grassland Research Institute, Shinfield.

KINOSITA, K. (1949)
Sedimentation in tilted vessels (1).
J. Colloid Interface Sci., 4, 525-536.

KUMAR, M.; BARTLETT, H.D. and MOHSENIN, N.N. (1972)
Flow Properties of Animal Waste Slurries.
Trans. ASAE., 15, (4), 718 - 722.

LARKIN, S.B.C.; MORRIS, R.M.; NOBLE, D.H. and RADLEY, R.W. (1981a)
Resource Mapping of Agricultural Wastes and Residues.
Report on a study conducted on behalf of the Energy Technology Support Unit.
National College of Agricultural Engineering, Silsoe.

LARKIN, S.B.C.; MORRIS, R.M.; NOBLE, D.H. and RADLEY, R.W. (1981b)
Production and Distribution of Agricultural Wastes in the United Kingdom and their Potential for use as an Energy Source.
pp. 335 - 351In Vogt, F. (ed.) Energy Conservation and the use of Solar and other Renewable Energies in Agriculture. Oxford: Pergamon Press.

LETTINGA, G.; VAN VELSEN, L.; DE ZEEUW, W. and HOBMA, S.W. (1980)
The Application of Anaerobic Digestion to Industrial Pollution Treatment.
pp. 167-186 In Stafford, D.A.; Wheatley, B. and Hughes, D.E. (eds.) Anaerobic Digestion. Proc. 1st International Symposium on Anaerobic Digestion, Cardiff, UK.
London: Applied Science Publishers.

LETTINA, G. (1984)
The prospects of anaerobic waste water treatment.
pp. 262-273 In Ferrero, G.L.; Ferranti, M.P. and Naveau, H. (eds.) Anaerobic digestion and carbohydrate hydrolysis of waste.
Barking: Elsevier Applied Science Publishers. 517 pages.

LIAO, P.H. and LO, K.V. (1985)
Methane Production using Whole and Screened Dairy Manure in Conventional and Fixed-Film Reactors.
Biotechnology and Bioengineering, 27, 266-272.

LIAO, P.H.; LO, K.V. and CHIENG, S.T. (1984)
Effect of Liquid-Solids Separation on Biogas Production from Dairy Manure.
Energy in Agriculture, 3, 61-69.

LO, K.V.; BULLEY, N.R.; LIAO, P.H. and WHITEHEAD, A.J. (1983)
The Effect of Solids-Separation Pretreatment on Biogas Production from Dairy Manure.
Agricultural Wastes, 8, 155-165.

MAFF (1981)
Energy in Agriculture.
Report of the ADAS/NFU Working Party on Energy in Agriculture.

MAFF (1983)
General Information. Farm Waste Management.
Booklet 2077.

MAFF (1985)
Farm environment and energy-saving grants.
Agriculture Improvement Scheme.
Booklet AIS 4.

MAFF (1986a)
Frequency distribution tables nos. 434, 443, 446 and 449 for the years 1981, 1983 and 1985.
Agricultural Census Branch, MAFF, Guildford.

MAFF (1986b)
Cost of Buildings Handbook.
Alnwick: Ministry of Agriculture, Fisheries and Food.

McINERNEY, M.J.; BRYANT, M.P. and STAFFORD, D.A. (1980)
Metabolic Stages and Energetics of Micrbial Anaerobic Digestion.
pp. 91-98 In Stafford, D.A.; Wheatley, B. and Hughes, D.E. (eds.) Anaerobic Digestion. Proc. 1st International Symposium on Anaerobic Digestion. London : Applied Science Publishers.

MARTIN, J.H. and DALE, G. (1978)
Energy recovery from manure using plug-flow digesters.
American Society of Agricultural Engineers,
Paper No. 78-4568.

MARTIN, J.H. and LICHTENBERGER, P.L. (1981)
Operation of a commercial farm-scale plug-flow manure digester plant.
pp. 439-462 In Symposium Papers, Energy from Biomass and Wastes V. Chicago: Institute of Gas Technology.

MEYNELL, P.J. (1976)
Methane : Planning a Digester.
Dorchester : Prism Press. 150 pages.

MEYNELL, P.J. (1985)
Safety in and around Anaerobic Digesters.
pp. 82-86, In Pain, B.F. and Hepherd, R.Q. (eds.) Anaerobic Digestion of Farm Waste. Technical Bulletin 7. Reading : The National Institute for Research in Dairying.

MONTEITH, H.D. and STEPHENSON, J.P. (1981)
Mixing efficiencies in full-scale anaerobic digesters by tracer methods.
J. Water Pollution Control Federation 53, 78-84.

MONO PUMPS LTD. (1982)
Curve 1569-M, MD/MH60 pump.
Mono Pumps Ltd., Audenshaw, Manchester.

MOORE, J.A. and MINER, J.R. (1985)
Wastewater Characteristics of a Flushing Dairy.
pp. 606-612 In Agricultural Waste Utilization and Management. Proc. 5th International Symposium on Agricultural Wastes, Chicago, USA. St. Joseph, Michigan: American Society of Agricultural Engineers.

MOSEY, F.E. (1981)
Methane fermentation of organic wastes.
Trib. Cebedeau, Nos. 453-454, pp. 389-400.

NIELSEN, V.C. (1985)
Installation and operation of digesters on farms.
pp. 87-95 In Pain, B.F. and Hepherd, R.Q. (eds.)
Anaerobic Digestion of Farm Waste.Technical Bulletin 7. Reading: The National Institute for Research in Dairying.

NOONE, G.P. and BRADE, C.E. (1982)
Low-Cost Provision of Anaerobic Digestion : II. High-Rate and Prefabricated Systems.
Water Pollution Control, 81, 479-510.

O'CALLAGHAN, J.R.; DODD, V.A. and POLLOCK, K.A. (1973)
The Long Term Management of Animal Manures.
J. agric. Engng. Res., 18, 1-12.

OLDSHUE, J.Y. (1966)
Fermentation Mixing Scale-Up Techniques.
Biotechnology and Bioengineering, 8, 3-24.

OLIVER, B.O. (1984)
Heat recovery to reduce the energy requirements of anaerobic digestion. Part 1 : Predictions. Div. Note DN/1232, Natn. Inst. agric. Engng., Silsoe. (unpubl.)

OLIVER, B.; PAIN, B.F. and PHILLIPS, V.R. (1986)
Mesophilic anaerobic digestion of dairy cow slurry on a farmscale : Economic considerations.
J. Agric. Engng. Res.,34, 229-243

ORTH, H.W. (1981)
Determination of heat transfer data for slurry. (Bestimmung von Kennzahlen zur Warmeubertragung bei Flussigmist). Grundlagen der Landtechnik, 1981, 31, (2), 47-50. Translation T.477, National Institute of Agricultural Engineering, Silsoe. 9 pages.

PAIN, B.F.; HEPHERD, R.Q. and PITTMAN, R.J. (1978)
Factors affecting the performance of four slurry separating machines.
J. agric. Engng. Res., 23, (3), 231-242.

PAIN, B.F.; WEST, R.; OLIVER, B. and HAWKES, D.L. (1984)
Mesophilic Anaerobic Digestion of Dairy Cow Slurry on a Farm Scale : First Comparisons between Digestion Before and After Solids Separation.
J. agric. Engng Res., 29, 249-256.

PAIN, B.F. (1986)
Personal communication.
The Animal and Grassland Research Institute,
Hurley, Maidenhead.

PECK, M.W.; HAWKES, F.R. and HAWKES, D.L. (1985)
Effect of mechanical separation on the anaerobic digestion of cattle slurry.
pp. 34-42, In Advances in Fermentation II. Rickmansworth : Turret Wheatland.

PECK, M.W.; SKILTON, J.M.; HAWKES, F.R. and HAWKES, D.L. (1986)
Effects of temperature shock treatments on the stability of anaerobic digesters operated on separated cattle slurry.
Wat. Res., 20, (4), 453-462.

PERRY, J.H. (1963)
(ed.) Chemical Engineers' Handbook.
Fourth Edition.
New York : McGraw-Hill Book Company.

PERSSON, S.P.E.; BARTLETT, H.D.; BRANDING, A.E. and REGAN, R.W. (1979)
Agricultural Anaerobic Digesters. Design and Operation.
Bulletin 827, Agricultural Experiment Station : The Pennsylvania State University College of Agriculture.50 pages.

PETERSEN, G. (1984)
Tracer study on a full-scale plug flow biogas plant using Li as a tracer.
pp. 406-408, In Egneus, H. and Ellegard, A. (eds.) Bioenergy 84. Vol III. Biomass conversion.
Barking: Elsevier Applied Science Publishers.

RIPLEY, L.E. and BOYLE, W.C. (1983)
Anaerobic Digestion Models : Implications for the Design Engineer.
pp. 451-463 In Anaerobic Digestion 1983. Proc. 3rd Intnl. Symp. on Anaerobic Digestion, Boston, USA.
Cambridge, Mass. : The 3rd Intnl. Symp. on Anaerobic Digestion.

ROCKWOOL LTD. (1981)
Rockwool lamella mats.
Data sheet 16.
Rockwool Ltd., Pencoed, Bridgend, Mid Glamorgan.

ROZZI, A. and PASSINO, R. (1985)
State of the art of anaerobic digesters in Europe.
pp. 115-124 In Pain, B.F. and Hepherd, R.Q. (eds)
Anaerobic Digestion of Farm Waste. Technical Bulletin 7.
Reading: The National Institute for Research in Dairying.

SCHELLER, W.A. (1982)
Commercial experience with a plug flow anaerobic digester for the production of biogas from agricultural and food processing wastes.
pp. 492-496 In Strub, A.; Chartier, P. and Schlesser, G. (eds) Energy from Biomass. Proc. 2nd E.C. Conference, Berlin. London : Applied Science Publishers.

STAFFORD, D.A. and ETHERIDGE, S.P. (1982)
Farm wastes, energy production and the economics of farm anaerobic digesters.
pp. 255-268 In Hughes, D.E.; Stafford, D.A.; Wheatley, B.I.; Baader, W.; Lettinga, G.; Nyns, E.J.; Verstraete, W. and Wentworth, R.L. (eds.) Anaerobic Digestion 1981. Proc. 2nd International Symposium on Anaerobic Digestion, Travemunde, FRG. Amsterdam : Elsevier Biomedical Press.

STAFFORD, D.A.; HAWKES, D.L. and HORTON, R. (1980)
Methane production from waste organic matter.
Florida : CRC Press. 285 pages.

STAFFORD, D,A. and SPENSLEY, R.A. (1986)
The Control and Optimisation of Anaerobic and Aerobic Waste Treatment Systems.
pp. 151-161 InEffluent Treatment and Disposal. I. Chem. E Symp. Series No. 96.

STEINSBERGER, S.C. and SHIH, J.C.H. (1984)
The Construction and Operation of a Low-Cost Poultry Waste Digester.
Biotechnology & Bioengineering, 26, (5) 537-543.

STEVENSON, A.E. and DE LANGEN, H. (1960)
Measurement of feed intake by grazing cattle and sheep. VII. Modified wet digestion method for determination of chromic oxide in faeces.
New Zealand J. agric. Res., 3, 314-319.

STRONACH, S.M.; RUDD, T. and LESTER, J.N. (1986)
Anaerobic digestion processes in Industrial Wastewater Treatment.
In Aiba, S.; Fan, L.T.; Fiechter, A. and Schurgel, K. (eds.) Biotechnology Monographs. Volume 2.
Berlin : Springer-Verlag.

STUCKEY, D.C. (1983a)
Biogas in developing countries : A critical appraisal. pp. 253-269 In Anaerobic Digestion 1983. Proc. 3rd Intnl. Symp. on Anaerobic Digestion, Boston, USA.
Cambridge, Mass. : The 3rd Intnl. Symp. on Anaerobic Digestion.

STUCKEY, D.C. (1983b)
Technology asessment study of biogas in developing countries.
International Reference Centre for Waste Disposal, Dubendorf, Switzerland, 1-137.

TENNEY, M.W. and BUDZIN, G.J. (1972)
How good is your mixing?
Water and Wastes Engineering, 9, 57-59.

THORN EMI, (1983)
Product Test Laboratory Report T.L.R. 706. March 1983.
Thorn EMI Flow Measurement, Stretford, Manchester.

TIETJEN, C. (1975)
From Biodung to Biogas - Historical Review of European Experience.
pp. 247-259, In Jewell, W.J. (ed.) Energy, Agriculture and Waste Management. Proc. of the 35th Cornell Agricultural Waste Management Conference. Michigan : Ann Arbor Science.

UDEN, P.; COLUCCI, P.E. and VAN SOEST, P.J. (1980)
Investigation of Chromium, Cerium and Cobalt as Markers in Digesta. Rate of Passage Studies.
J. Sci. Food Agric., 31, 625-632.

VAN DEN BERG, L. and KENNEDY, K.J. (1981)
Support materials for stationary fixed film reactors for high rate methanogenic fermentations.
Biotechnology Letters, 3, (4), 165-170.

VAN VELSEN, A.F.M.; LETTINGA, G. and DEN OTTELANDER, D. (1979)
Anaerobic digestion of piggery waste. 3. Influence of temperature.
Neth. J. agric. Sci. 27, 255-267

VOERMANS, J.A.M. (1985)
Biogas Production on Livestock Farms.
pp. 132 -137 In Agricultural Waste Utilization and Management. Proc. 5th International Symposium on Agricultural Wastes, Chicago, USA.
St. Joseph, Michigan: American Society of Agricultural Engineers.

WELLINGER, A. (1985)
pp. 125 - 134 In Pain, B.F. and Hepherd, R.Q. (eds) Anaerobic Digestion of Farm Waste. Technical Bulletin 7. Reading: The National Institute for Research in Dairying.

WENTWORTH, R.L. (1984)
Anaerobic Digestion in North America.
pp. 348-358 In Ferrero, G.L.; Ferranti, M.P. and Naveau, H. (eds.) Anaerobic Digestion and Carbohydrate Hydrolysis of Waste. Barking: Elsevier Applied Science Publishers.

WEST, R. (1984)
Personal communication.
National Institute for Research in Dairying, Shinfield, Reading.

WEST, R. (1986)
Personal communication.
The Animal and Grassland Research Institute, Shinfield, Reading.

WHEATLEY, B.I. (1978)
Anaerobic Digestion on Farms - The European Scene.
pp. 94-102 In Anaerobic Digestion of Farm Wastes.
Report of a Seminar organised by ADAS Farm Waste Unit at Cardington, Bedford.
Reading : ADAS Farm Waste Unit, MAFF.

WILLIAMS, A.G.; SHAW, M. and ADAMS, S.J. (1984)
The Biological Stability of Aerobically-treated Piggery Slurry during Storage.
J. agric. Engng Res., 29, 231-239.

WILTON, B. (1985)
Personal communication.
Nottingham University School of Agriculture,
Sutton Bonnington, Loughborough, Leics.

WOOD, P.R. and CHAPMAN, J.M. (1986)
Report of a visit to Cornell University and ISAW 85, USA.
Visit Report Overseas, URO 206, National Institute of Agricultural Engineering, Silsoe. (unpublished). 26 pp.

YOUNG, J.C. and McCARTHY, P.L. (1969)
The anaerobic filter for waste water treatment.
J. Wat. Pollut. Control Fed., 41, R160-R173.

ZEIKUS, J.G. (1980)
Microbial populations in digesters.
pp. 61-89 In Staford, D.A.; Wheatley, B.I. and Hughes, D.E. (eds.) Anaerobic Digestion.
Proc. 1st International Symposium on Anaerobic Digestion, Cardiff, UK. London: Applied Science Publishers.

ZOLTEK, J. and GRAM, A.L. (1975)
High-rate digester mixing study using radioisotope tracer.
J. Water Pollution Control Federation, 47, (1), 79-84.

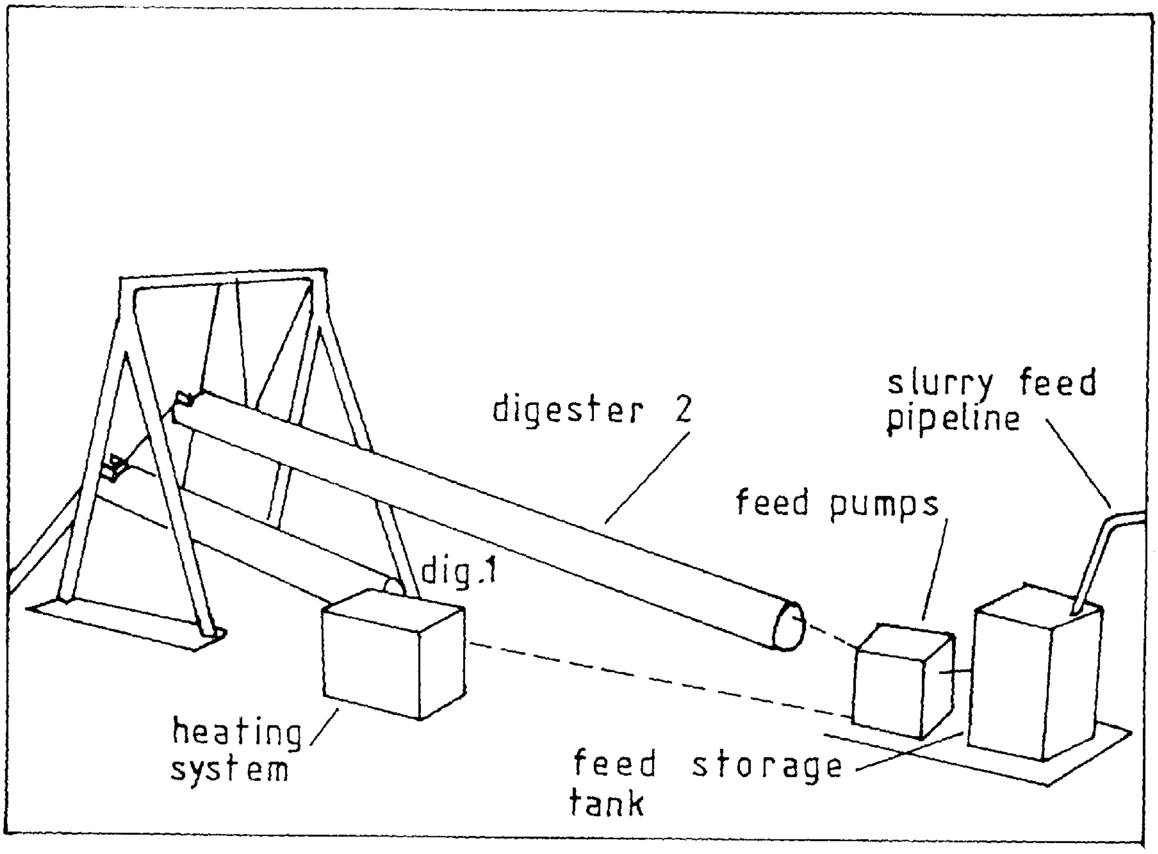


PLATE 1–KEY

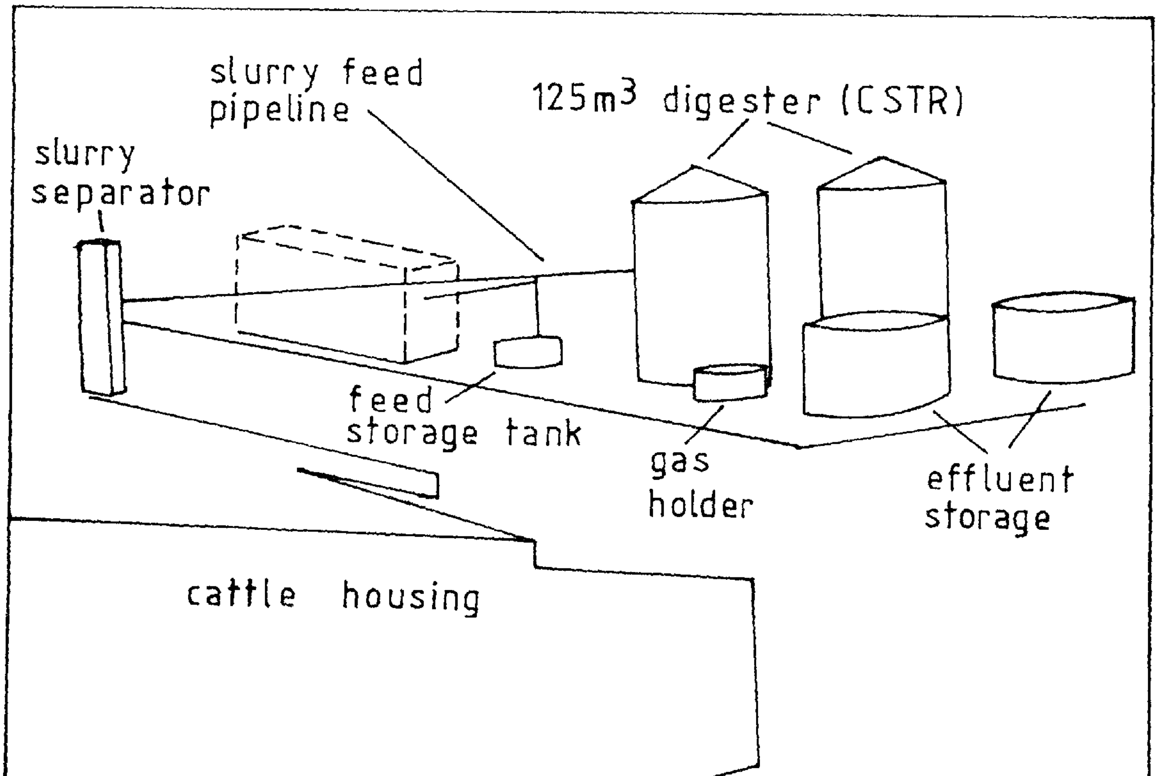


PLATE 2–KEY