THE STOICHIOMETRY AND KINETICS OF BIOLOGICAL HEAT GENERATION IN THE AEROBIC STAGE OF DUAL DIGESTION.

by

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A thesis submitted in fulfillment of the
requirements for the degree of Doctor of
Philosophy at the University of Cape Town.

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March 1991

DECLARATION BY CANDIDATE

I, John Robert Messenger, hereby declare
that this thesis is my own work and that it has
not been submitted for a degree at another University.

Signed by candidate

Signature Removed

FRONTISPIECE

A scene at the Potsdam Waste Water Treatment Works showing the dual digestion system. The lagged 45 m^3 fibre glass aerobic reactor and the sludge thickener are situated to the left of the foreground of the photograph. The huts to the right of the reactor and to the right of the thickener respectively housed the pumping and electrical control equipment. The digestor on the right served as the anaerobic stage in the research while the digestor directly behind the reactor and thickener was employed as a sludge storage vessel for the Zimpro sludge treatment facility situated to the left of the photograph. The liquid oxygen storage vessel and vaporizer can be seen in the foreground.

SYNOPSIS

RESEARCH INTO DUAL DIGESTION

Over the past 10 years the Water Research Commission (WRC) has recognized the potential benefits of the dual digestion system for South Africa where anaerobic digestion is the preferred method of sludge stabilization. Accordingly, from 1981 to 1984 under WRC sponsorship, the Johannesburg City Council (JCC) undertook a pilot plant study into the performance of the autothermal thermophilic aerobic reactor oxygenated with pure oxygen. As a consequence of the encouraging results obtained in this study (Trim and McGlashan, 1984; Trim, 1984), in 1987 the WRC and Milnerton Municipality supported a full scale evaluation of dual digestion (45 m^3 pure oxygen oxygenated aerobic reactor and 600 m^3 anaerobic digestor) at Milnerton's Potsdam Wastewater Treatment Works, with scientific and academic input by the Division of Water Technology of the CSIR and the Department of Civil Engineering, of UCT. In this evaluation a number of important considerations needed to be addressed such as

(1) oxygen requirements and oxygen utilization efficiency with pure oxygen and air oxygenation,

(2) minimum retention time that can be achieved in the aerobic reactor and in the anaerobic digestor without compromising the attainment of thermophilic temperature in the reactor and VS removal, gas production and sludge stability in the digestor,

(3) temperature control of the aerobic reactor and anaerobic digestor,

(4) efficacy of pathogen inactivation and mode of operation to prevent recontamination,

(5) dewaterability of the anaerobically digested sludge,

(6) operation and economic evaluation of the system.

In this thesis, the results obtained pertaining to operation and performance of the Milnerton aerobic reactor are presented in detail. Also, the observed results are generalized and, based on the observed general principles, a design procedure and a simulation algorithm and computer programme for pure oxygen and air oxygenated reactors are developed, in accordance with objectives (1) to (3) above.

KINETICS OF BIOLOGICAL HEAT GENERATION

In order to meet objectives (1) to (3) above for the aerobic reactor, it was necessary to examine the kinetics of biological heat generation. In this regard it was realized during the investigation that a distinction needs to be made between the objectives of the thermophilic aerobic reactor in dual digestion and thermophilic aerobic digestion as a stand alone process. In the latter, the objective is to reduce the sludge energy as measured by VS removal; accordingly the process kinetics are defined in terms of a VS degradation rate, and from it a retention time to achieve a specified VS removal can be calculated. Also, the rates of biological heat generation and oxygen utilization are both related proportionally to the VS degradation rate. For the kinetics to be controlled by the VS degradation rate, the process needs to be operated under oxygen sufficient conditions. This approach is unproductive for modelling the aerobic reactor in dual digestion principally because it was found in this investigation that (1) the reactor was not operated under oxygen sufficiency conditions but under oxygen limiting conditions and (2) very little VS was removed in the reactor even though large quantities of biological heat were generated. Accordingly, because (1) biological heat generation theoretically can be shown to be stoichiometrically related to oxygen consumption rate and (2) the reactor is operated under oxygen limiting conditions, it was proposed that the oxygen supply rate and sludge oxygen consumption rate are the most useful parameters for operation, control, design and simulation of the aerobic reactor. Therefore the principal objective of the research became to substantiate this proposal by (1) verifying direct proportionality between the rates of biological heat generation and oxygen consumption via the specific heat yield (Y_H, MJ/kgO) and (2) demonstrating that with the oxygen supply rate, instantaneous and complete control of the reactor sludge temperature could be obtained.

EVALUATION OF AEROBIC REACTOR

In order to measure the specific heat yield (Y_H), accurate heat and oxygen mass balances needed to be made over the aerobic reactor. Because the aerobic reactor was batch fed on a draw and fill basis, (to avoid contamination of the effluent sludge with influent sludge; this incidentally, also provided an accurate means of sludge flow measurement), the reactor temperature fluctuated over a batch cycle resembling a saw-tooth pattern with 3 phases i.e. (1) a sludge transfer phase (draw) of 4 min, (2) a feeding phase (fill) of 3 min and (3) a heating phase of 2 to 6h depending on the reactor retention time. By measuring during the heating phase, (1) the sludge temperature increase rate, which is proportional to the sludge enthalpy increase rate (H_{net}), (2) the oxygen injection recirculation pump mechanical heat input rate (H_{mi}), (3) the wall heat loss rate (H_{we}) and (4) the vent gas water vapour heat loss rate (H_{ve}), the biological heating rate (H_{bi}) could be determined. Then by dividing H_{bi} by the mass oxygen consumption rate (O_c) obtained from the oxygen mass balance over the same heating period, the specific heat yield (Y_H) was calculated.

To measure the oxygen consumption rate required careful monitoring of the influent and vent gas volumetric flow rates. With regard to the vent gas, the total vent gas flow was collected, dehumidified to dry it and measured in a town's gas meter. The water vapour removed from the vent gas in condensers was collected and from this the water vapour heat loss rate was calculated. The oxygen and carbon dioxide mass and molar flow rates in the vent gas were obtained by measuring the oxygen content of the dried vent gas at the gas meter.

Apart from the heat and oxygen mass balances, influent sludge VS and COD concentrations also were measured to check whether or not the rates of biological heat generation or oxygen consumption correlated with VS or COD removal rates.

AEROBIC REACTOR RESULTS

Over a period of 8 months, 116 heat and oxygen mass balances were done over the aerobic reactor. During this time the operating conditions differed widely viz. reactor sludge temperature 54 to 69°C, retention time 1.2 to 3d, average ambient temperature 8 to 30°C and oxygen supply rates (OSR) from 0.13 to 0.44 kgO/(m^3.h) giving oxygen transfer rates (OTR, or equivalently sludge oxygen consumption rate, O_C) from 0.13 to 0.39 kgO/(m^3.h) and oxygen transfer efficiencies (OTE = OTR/OSR) from 1.00 to 0.80 respectively. From these tests the following results were obtained:

(1) Biological heat generation rate was directly proportional to the oxygen

transfer rate (OTR), or equivalently the sludge oxygen consumption rate (O_c). The constant of proportionality is the specific heat yield (Y_H) and was measured to be 12.77 ± 0.58 MJ/kgO. This value conforms closely to thermodynamically and bioenergetically calculated values. The value was found to be independent of reactor sludge temperature and retention time but was slightly dependent on oxygen limitation (defined as 1–OTR/OUR), increasing as oxygen limitation increased.

(2) Reactor sludge temperature increases could be completely, and instantaneously, controlled by means of the oxygen supply rate (OSR) *for as long as the reactor was oxygen limited* i.e. while the oxygen transfer rate (OTR) was less than the maximum biological oxygen utilization rate (OUR). For the Milnerton sludge, which was a mixture of primary and humus (biofilter) sludge, OUR was measured to be about 0.38 kgO/(m^3.h) at an average concentration of 30 kgVS/m^3. Step increases in OSR as high as 330% [from 0.13 to 0.43 kgO/(m^3.h)] which increased OTR and hence the biological oxygen consumption rate by 300% [from 0.13 to 0.39 kgO/(m^3.h)] caused an immediate (<2h) increase in the biological heat generation rate (H_{bi}) and hence an increasing reactor sludge temperature. In the same way, decreasing the OSR caused an immediate decrease in H_{bi} and a decreasing reactor sludge temperature. Increases in OSR did not cause an equivalent increase in OTR because, as OTR increased so OTE (= OTR/OSR) decreased. This was the case with the Milnerton Vitox pumped recirculation pure oxygen injection oxygenation system, and probably also will be the case with other pure oxygen and air oxygenation systems. This aspect is of considerable significance in design.

(3) Respiration quotient i.e. mole CO_2 generated per mole O_2 utilized was 0.66 instead of 1.0 often assumed. From this it appeared that the oxidation reactions in the aerobic reactor were not simply those of VS degradation.

(4) Vent gas was saturated with water vapour at all vent gas flow rates.

(5) COD and/or Volatile Solids (VS) removal rates were poor parameters for (i) quantifying the biological heat generation rate and (ii) controlling the reactor sludge temperature because the tests (i) are prone to significant variability when dealing with sewage sludges and (ii) take too long to give a

result.

The close correlation, and rapidity of response, between the biological heat generation rate and the oxygen transfer rate make the OTR a pivotal parameter in design and simulation of autothermal thermophilic aerobic reactors in dual digestion.

DESIGN AND SIMULATION OF THE AEROBIC REACTOR

Aerobic reactor design and simulation procedures were derived from the results of the Milnerton aerobic reactor performance. These are founded on the basic heat and mass balance principles and accordingly are suitable for general application to aerobic reactor design. Therefore, whereas pure oxygen oxygenation was used on the Milnerton reactor, the derived design and simulation procedures are general and apply to reactors oxygenated also with air or oxygen enriched air.

The design procedure is based on the solution of the steady state heat balance across the reactor. Such a heat balance yields a constant temperature for the reactor sludge and is applicable only to reactors that are continuously fed. However, in practice, the reactor is batch fed to avoid recontamination of pasteurized sludge. This causes the reactor sludge temperature to fluctuate between 2 and 4°C per batch cycle. Nevertheless, despite batch feeding, the steady state approach is adopted for design because it greatly simplifies the design procedure.

Two objectives need to be met by the aerobic reactor in dual digestion, viz. (1) pasteurization by exposure of the sludge to a minimum temperature for a minimum length of time generally above 60°C for 2h or above 70°C for 30 min and (2) pretreatment through oxygen limitation for enhanced performance of the anaerobic digestor. The literature does not define the temperature and the degree of oxygen limitation at which sludge pretreatment is best accomplished, nor could this be established at Milnerton. Thus the approach to design is to ensure that (1) the specified pasteurization temperature and times are maintained in the reactor, and (2) the sludge is oxygen limited, i.e. the OTR [kgO/(m^3.h)] to the sludge by the oxygenation system is controlled at a lower value than the sludge's maximum OUR [kgO/(m^3.h)]. Not only is oxygen limitation important for sludge pretreatment but, as mentioned above, also for reactor sludge temperature control. Sludge stabilization (i.e. VS removal) is *not* an objective of the aerobic reactor; this is accomplished in the anaerobic digestor. Therefore the oxygen supply rate is governed solely by the requirement of generating sufficient heat biologically to achieve the pasteurization specifications.

Accepting that design of the aerobic reactor centres on minimizing the retention time through maximizing the heat sources (i.e. biological heat generation and mechanical heat input) and minimizing the heat losses (i.e. wall heat loss and vent gas water vapour and sensible heat losses) it is demonstrated with the aid of the steady state heat balance that 3 parameters are of crucial importance, viz:

(1) The sludge oxygen utilization rate (OUR), which fixes the maximum biological heat generation rate through the specific heat yield (Y_H).

(2) The oxygen transfer rate (OTR) of the oxygenation system. The actual biological heat generation rate is directly proportional to the OTR through the specific heat yield (Y_H). The OTR should be less than the OUR to ensure (1) sludge pretreatment through oxygen limitation and (2) reactor sludge temperature control through control of the oxygen supply rate (OSR).

(3) The oxygen transfer efficiency (OTE) of the oxygenation system. For a particular pure oxygen or air oxygenation system at a given OTR, the OTE controls the vent gas volumetric flow rate; the lower the OTE the greater the vent gas flow rate and hence the greater the vent gas heat losses (via water vapour and sensible heats).

At a certain OTR and mechanical heat input, the heat sources are fixed. If OTE is high (>0.80) and oxygenation is with pure oxygen, the vent gas heat losses are small with the result that most of the heat generated can be lost via hot effluent sludge thereby allowing short retention times (~ 1.25 to 2d); if OTE is low (0.10–0.20) and oxygenation is with air, then vent gas heat losses are high with the result that much less heat can be lost via the hot effluent sludge thereby forcing long retention times (4–6d). At Milnerton, where the sludge OUR was observed to be around 0.38 kgO/(m^3.h) and high OTR's and OTE's of 0.37 kgO/(m^3.h) and > 0.80 respectively were achieved with the use of pure oxygen, the vent gas heat losses were very low allowing 1.25d retention time operation and reactor temperatures above 60°C. In contrast, with air oxygenation, OTR's and OTE's tend to be considerably lower [e.g. at Athlone, 0.14 kgO/(m^3.h) and 0.12 respectively, Pitt, 1990] and the major part of the heat generated is lost via the vent gas with the result that only a minor part of the heat can be lost via hot effluent sludge thereby forcing long retention time operation to maintain temperatures above 60°C.

Biological heat generation has the drawback in that increasing its rate requires increases in OTR. Increases in OTR require increases in oxygen or air supply rates (OSR) which, together with the reduction in OTE usually caused by increases in OTR, causes larger vent gas heat losses. In contrast to biological heat generation, supplementary heat sources such as (1) heat exchange between reactor effluent and feed sludge (2) increased mechanical heat input, or (3) anaerobic digestor gas combustion, have the effect of increasing the heat sources without increasing the heat losses, and therefore allow a *pro rata* reduction in retention time.

The design procedure described above is based on the steady state heat balance which accepts that the reactor sludge temperature is constant and therefore applies only to a continuously fed reactor. Pasteurization requires the reactor to be batch fed, resulting in a continuously changing reactor sludge temperature over about 3°C in a saw-tooth pattern, decreasing at times of feeding and increasing between the feeding times, with the result that the reactor is not at steady state. The error arising from assuming steady state conditions is small and well justified by the simplifications in the heat balance it affords. The saw-tooth temperature profile can only be determined by solving the unsteady state heat balance by a forward integration method. This is dealt with in this thesis and an algorithm and computer programme are presented which solve the unsteady heat balance in order to simulate the temperature profile of the batch fed aerobic reactor. For specified input conditions identical to those observed in the Milnerton reactor, the simulation programme is shown to simulate the reactor sludge temperature profile to within 0.5°C. Whereas the steady state design allows only the required pasteurization temperature to be specified, the simulation programme allows both the sludge pasteurization temperature and the undisturbed sludge detention period to be defined, the latter in terms of the interval between consecutive batch feeds.

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ACKNOWLEDGEMENTS

I wish to express my sincere thanks and appreciation to:

- Associate Professor G A Ekama, my supervisor, for his guidance and assistance throughout the research.
- Professor G v R Marais for his substantive support and advice
- Mr H A de Villiers of the Department of Water Technology of the CSIR for countless hours of technical assistance at the Potsdam research site.
- Mr K Kenmuir and staff at the Potsdam Waste Water Treatment Works, both for their assistance and for their patient tolerance during the research program.
- Afrox Limited for their financial support during the research.
- The Water Research Commission and the Foundation for Research and Development for providing financial support.
- Heather Bain who typed this thesis; her cheerful good nature never flagged in spite of receiving a seemingly unending stream of drafts, redrafts and alterations.
- My fellow colleagues and friends for their constant moral support

A very special word of thanks to my wife, Melissa, whose moral support and encouragement made the onerous task of completing this thesis that much more bearable.

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Symbols and abbreviations which are employed in this thesis are listed below. To facilitate the use of this list an illustration of the reactor is given on Page xxxii showing the positions on the reactor where the parameters denoted by some of these symbols are likely to be encountered.

A = the surface area of the aerobic reactor and Vitox oxygenation system pipework (m^2)

ATAD = autothermal thermophilic aerobic digestion

[BVS] = biodegradable volatile solids concentration (kg/m^2)
$[BVS]_t$ = [BVS] at time t (kg/m^2)
$[BVS]_i$ = initial [BVS] i.e. prior to sludge digestion (kg/m^2)
$[BVS]_f$ = final [BVS] i.e. after sludge digestor or [BVS] in the mixed liquor of a completely mixed digestor (kg/m^2)

COD = chemical oxygen demand [(g/ℓ) or (kg/m^3)]
Cos ϕ = power factor of Vitox recirculation pump electric motor
C_p = specific heat of sludge (MJ/(ton.°C)
C_{pg} = composite specific heat of the aerobic reactor vent gas (MJ/(ton.°C)
CSIR = Council for Scientific and Industrial Research.

dT_{se}/dt = the rate of change of aerobic reactor sludge temperature (°C/h)
DWT = Division of Water Technology.

F = the rate of change of the aerobic reactor operating volume during one of the 3 phases of a complete batch cycle.
f_b = the batch fraction ie the ratio of the volume of a single batch of feed sludge to the full operating volume of the aerobic reactor.

f_{mech} = the fraction of electrical power consumed by the Vitox recirculation pump motor which was converted to heat

f_{mo_2} = the mass fraction of oxygen in the influent gas to the aerobic reactor.

f_{O_2} = the volume (molar) fraction of oxygen in the dry vent gas from the aerobic reactor

F_s = the flow rate of sludge passing through the aerobic reactor [(m^3/h) or (m^3/d)]

G_{in} = the free energy of the feed sludge to a digestor

G_{out} = the free energy of the effluent sludge from a digestor

H_{bi} = the rate of biological heat generation (MJ/h)

H_{ex} = the rate of heat available for heat exchange (MJ/h)

h_{fg} = the latent heat of vaporization of water (MJ/kg)

H_{gi} = the rate of heat gain to the aerobic reactor due to the sensible heat of the feed gas (MJ/h)

H_{ge} = the rate of heat loss from the aerobic reactor due to the sensible heat of the aerobic reactor vent gas (MJ/h)

H'_{ge} = the net rate of heat loss from the aerobic reactor due to the sensible heats of the reactor feed and vent gases (MJ/h)

H_{losses} = the sum of the rates of all the heat losses from the aerobic reactor in various forms excluding only the net rate of heat loss via the hot effluent sludge, H'_{se} (MJ/h)

H_{mi} = the rate of heat generation in the aerobic reactor sludge due to the mechanical action of pumps and stirrers (MJ/h)

H_{net} = the rate of aerobic reactor sludge enthalpy increase during the specific heat yield test (MJ/h)

H_{se} = the rate of heat loss from the aerobic reactor due to the sensible heat of the reactor effluent sludge (MJ/h)

H'_{se} = the net rate of heat loss from the aerobic reactor due to the sensible heat of the reactor feed and effluent sludges (MJ/h)

= the rate of heating required to change the temperature of the sludge from the feed to the effluent value (MJ/h)

H_{si} = the rate of heat gain to the aerobic reactor due to the sensible heat of the reactor feed sludge (MJ/h)

H_{sinks} = the sum of the rates of all the heat losses from the aerobic reactor in all forms including the net rate of heat loss via the hot effluent sludge, H'_{se} (MJ/h)

$H_{sources}$ = the rate of heat supply to a digestor/reactor needed to maintain a desired thermophilic temperature in the digestor/reactor sludge (MJ/h)

H_{ve} = the rate of heat loss from the aerobic reactor in the latent heat of the reactor vent gas (MJ/h)

H_{we} = the rate of heat loss from the surfaces of the aerobic reactor and the Vitox oxygenation system (MJ/h)

I = the phase current drawn by the Vitox recirculation pump (amps)

k = VS reduction rate coefficient (h^{-1})

m_{dg} = the molar flow rate of dry vent gas leaving the aerobic reactor (kmol/h)

M_{wc} = the mass flow rate of condensate accumulation in the water traps of the aerobic reactor vent gas metering system during the specific heat yield test (kg/h)

M_{we} = the mass flow rate of water vapour leaving the aerobic reactor in the vent gas (kg/h)

M_{wp} = the predicted mass flow rate of water vapour leaving the aerobic reactor in the vent gas (kg/h)

M_{wr} = the mass flow rate of water vapour in the partially dried aerobic reactor vent gas leaving the condenser of the vent gas metering system

O_c = the mass rate at which the aerobic reactor sludge biomass consumes oxygen (kgO/h)

O_{cmax} = the maximum mass rate at which the aerobic reactor sludge biomass consumes oxygen ie when the reactor is operated under oxygen sufficient conditions (kgO/h)

O_{eff} = the fraction of oxygen supplied to the aerobic reactor which is consumed by the reactor sludge NB this symbol is used in the specific heat yield test and is equivalent to OTE, the symbol preferred in design procedures.

O_s = the mass rate at which oxygen is supplied to the aerobic reactor sludge biomass (kgO/h)

OSR = the oxygen supply rate to the aerobic reactor [kgO/(m^3.h)]

OTE = oxygen transfer efficiency i.e. the fraction of the oxygen supply which the oxygenation system is able to transfer to the mixed liquor NB this symbol is used in design procedures and is equivalent to O_{eff}, the symbol preferred in the specific heat yield test.

OTR = the oxygen transfer rate i.e. the rate at which oxygen is transferred to the mixed liquor [kgO/(m^3.h)]

OUR = the biological oxygen utilization rate i.e. the volume specific rate at which sludge consumes oxygen under oxygen sufficient conditions [kgO/(m^3.h)]

P = the substitute energy parameter

p_{dg} = the partial pressure of the dry vent gas (mmHg)

PST = primary sedimentation tank

ρ_s = the sludge density (ton/m^3)

P_T = the total absolute pressure of the vent gas (mmHg)

p_w = the partial pressure of water vapour in the vent gas (mmHg)

Q_{bi} = a quantity of biologically generated heat

R_H = the hydraulic retention time of the sludge in the aerobic reactor

t = time (h)

T_{amb} = the ambient temperature (°C)

t_c = the cycle time ie the duration of the 3 phases of a complete batch cycle (ie $t_c = t_D + t_{fpr} + t_{tpr}$) [h]

t_D = the duration of phase 3 of a batch cycle which is equivalent to the period of exposure of the aerobic reactor sludge to disinfection conditions (h)

t_{fpr} = the duration of phase 2 of a batch cycle (h)

T_{ge} = the temperature of the vent gas at the point of exit from the aerobic reactor (°C)

T_{gi} = the temperature of the oxygen/air entering the aerobic reactor (°C)

T_{gm} = the temperature of the aerobic reactor vent gas passing through the vent gas meter during a specific heat yield test (°C)

T_{min} = the minimum temperature at which the enthalpy of the sludge transferred from the aerobic reactor to the anaerobic digestor is sufficient to maintain mesophilic temperatures in the digestor (°C)

T_{si} = the temperature of the feed sludge to the aerobic reactor (°C)

T_{se} = the temperature of the sludge in the aerobic reactor *or* the temperature of the effluent sludge from the aerobic reactor (°C)

TS = total solids concentration [(g/ℓ) or (kg/m³)]

t_{tpr} = the duration of phase 1 of a batch cycle (h)

U = the overall heat transfer coefficient for the aerobic reactor and oxygenation system pipework (MJ/(m².h.°C)

UCT = University of Cape Town.

V = the phase voltage of the Vitox recirculation pump electric motor (volts)

V_b = the batch volume ie the volume of a batch of sludge fed to the aerobic reactor.

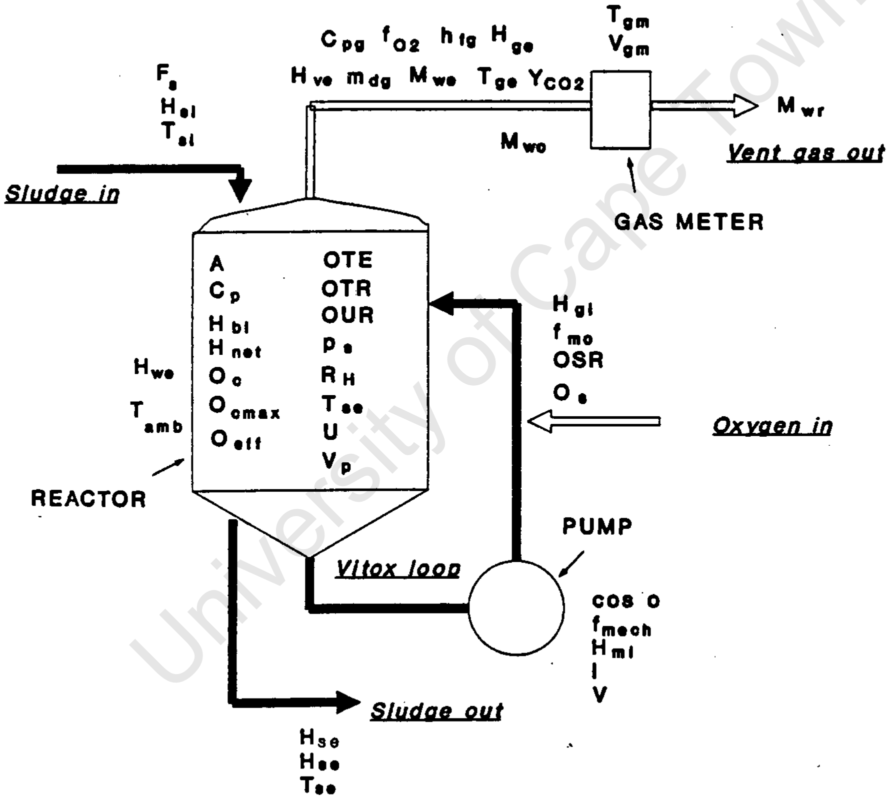
V_{gm} = the volumetric flow rate of vent gas measured by the vent gas meter during the specific heat yield test (m³/h)

V_p = the operating volume of the aerobic reactor (m³)

VS = the volatile solids concentration [(g/ℓ) or (kg/m³)]

Y_H = the specific heat yield i.e. the quantity of biological heat generated per unit change in substitute energy parameter P.

SOME OF THE SYMBOLS USED IN THE TEXT



CHAPTER 1

INTRODUCTION

1.1 THE NATURE OF SEWAGE SLUDGE

At municipal wastewater treatment plants, there are usually two main stages of treatment, i.e. (1) wastewater treatment and (2) sludge treatment and disposal. In the wastewater treatment stage, a number of physical and biological unit operations follow sequentially whereby particulate and dissolved organic pollutants at a low concentration in the wastewater are transformed and removed from the wastewater as a concentrated low volume sludge to produce a clear water stream as effluent. Having treated the wastewater in the first stage, it remains in the second stage to treat and dispose of the produced sludge in such a way as to prevent pollution of the environment and to safeguard the health of man and animals.

The sludge produced at wastewater treatment plants is a putrefactive concentrated aqueous suspension of particulate organic material containing mainly biodegradable but also inert substances. Its organic putrefactive nature makes it potentially both a public nuisance and a health hazard. It is very active biologically and its uncontrolled biodegradation very rapidly generates unpleasant odours and provides a site for insect and fly breeding. Its indiscriminate disposal can pollute surface and ground waters and contaminate land with human pathogens and heavy metals.

The need to treat the sludge before disposal was recognized early on and sludge treatment technology evolved as wastewater treatment technology evolved. Generally the treatment of sludge involves the controlled degradation of the putrefactive organic material, called *stabilization* and further concentrating the sludge called *dewatering*. The stabilized and concentrated sludge can be disposed of in a number of ways such as incineration, landfill or used in beneficial ways such as a soil conditioner and, more recently, brick making.

By far the most widely employed beneficial use of sludge in many countries, particularly in Europe, is as a soil conditioner. The sludge not only serves as a nutrient source of N and P but also lends a number of other beneficial properties to the soil such as improving moisture retention and soil structure. However, there are attendant health risks involved in the agricultural use of sludge. In the stabilization and dewatering processes the sludge is not disinfected with the result that it still contains high levels of pathogens such as *Salmonella, hepatitis A and B*, enteric coliforms and *Helminths*, hook worm, round worm and *Ascaris Lumbricoides.* In the Western Cape of South Africa, the incidence of *Ascaris* is very high and poses a considerable health hazard particularly to children. Depending on the industrial contribution to the municipal wastewater, the sludge produced in its treatment will contain heavy metals such as Cadmium, Cobalt, Chromium, Copper, Mercury, Molybdenum, Nickel, Lead, Zinc, which precipitate out in the treatment process and become part of the sludge. These heavy metals are difficult to remove from the sludge and the best way to keep them out of the sludge is to prevent them from entering the sewerage system by having good regulation of industrial wastewater discharges to the sewers. Nevertheless most sludges invariably contain some heavy metals and depending on the level, if the sludge is used as a soil conditioner, the metals can enter the animal and human food chain and pose a health risk.

1.2 REGULATIONS REGARDING USAGE OF SLUDGE

In order to control the disposal and use of sewage sludge to reduce its public health risk, health authorities in many countries have set down guidelines or regulations stating what may be done with certain classes of sludge, where the class of sludge depends on the treatment it has received, and the level of heavy metals in it. As an example of such guidelines, those recommended in South Africa by the Department of Health and Population Development are summarized in Table 1.1 (reproduced from Vivier, *et al.*, 1988). An overview of USA and German regulations with respect to stability and pathogens for the agricultural use of sludge is given by Heidman (1989).

The regulations and guidelines stipulate heavy metal limits and appropriate sludge treatment before application to land in order to stabilize and pasteurize the sludge to a degree where it no longer attracts disease vectors such as insects and rodents and to reduce pathogens. With regard to stabilization and pathogen removal, two

TABLE 1.1: CLASSIFICATION OF SEWAGE SLUDGE TO BE USED OR DISPOSED OF ON LAND
(source: Vivier et al (1991))

TYPE OF SEWAGE SLUDGE	ORIGIN/TREATMENT (EXAMPLES)	CHARACTERISTICS/QUALITY STANDARD
Type A sludge	Raw sludge Cold digested sludge Septic tank sludge Oxidation pond sludge (night soil)	* Usually unstabilized and can cause odour nuisances and fly breeding * Contain pathogenic organisms * Variable metal content
Type B sludge	Anaerobic digested sludge (heated digestor) Surplus activated sludge Humus tank sludge	* Fully or partially stabilized - should not cause significant odour nuisances or fly breeding * Contain pathogenic organisms * Variable metal content
Type C sludge	Pasteurized sludge Heat treated sludge Lime stabilized sludge Composted sludge Irradiated sludge Fumigated sludge	* Certified to comply with the following quality requirement: # Stabilized - should not cause odour nuisances or fly breeding # Contain no viable Ascaris Ova per 10g dry sludge # Maximum 1 Salmonella organism per 100g dry sludge # Maximum 10 faecal coliforms per 100g dry sludge immediately after treatment (disinfection/sterilization) NB If not certified this sludge is considered a Type B sludge * Variable metal content
Type D sludge	Pasteurized sludge Heat treated sludge Lime stabilized sludge Composted sludge Irradiated sludge Fumigated sludge Produced for unlimited use on land with or without addition of plant nutrients or other materials	* Certified to comply with the following quality requirement: # Stabilized - should not cause odour nuisances or fly breeding # Contain no viable Ascaris Ova per 10g dry sludge # Maximum 1 Salmonella organism per 100g dry sludge # Maximum 10 faecal coliforms per 100g dry sludge immediately after treatment (disinfection/sterilization) # Maximum metal content in mg/kg dry sludge: Cadmium 20 Cobalt 100 Chromium 2750 Copper 750 Mercury 10 Molybdenum 25 Nickel 200 Lead 250 Zinc 2750 * User must be informed about the moisture and N P K content * User must be warned that not more than 8 t/ha/yr (or kg/10 sq.m) may be applied to soil and that the pH of the soil should preferably be higher than 6

categories of sludge treatment technology are defined in the guidelines or regulations depending on the level of pathogen reduction that is achieved, i.e.

1) Processes to Significantly Reduce Pathogens (PSRP) which include ambient aerobic digestion and mesophilic anaerobic digestion, and

2) Processes to Further Reduce Pathogens (PFRP) which include wet air oxidation (Zimpro), composting under certain conditions, thermophilic aerobic digestion, β or γ ray radiation and pasteurization (70° C for 30 min).

The anticipated disinfection performance achieved by the two categories of processes is summarized by Heidman (1989).

1.3 SLUDGE STABILIZATION AND DISINFECTION PROCESSES

In South Africa and the UK, anaerobic digestion of sewage sludge is the preferred method of stabilization, whereas in the European countries aerobic digestion is a popular stabilization method. With the growing concern about the health risks of using sludge for agricultural purposes, disinfection or pasteurization are now also required or recommended. For example, already in the late 1960's Switzerland promulgated legislation that required all sludge used in agriculture to be pasteurized. This legislation was repealed in 1977 when it was found that the post stabilization–pasteurization approach which had been adopted was ineffective due to regrowth of pathogens in the treated sludge. This problem of reinfection of pasteurized sludge was researched in depth in Europe (Switzerland, Germany and Holland) and because major modifications to existing pasteurization plants needed to be made, new legislation was promulgated in 1981 which stipulated pasteurization of sludge by 1990.

Processes other than conventional pasteurization (70° C for 30 min) are also capable of disinfection. The process of thermophilic aerobic digestion allows simultaneous stabilization and pasteurization of sludge. In this process, sludge temperatures in excess of 60° C can be maintained autothermally at 6 to 8 days retention time thereby obviating the need for costly external heating of the sludge. The autothermal heat is generated from the biological oxidation reactions which lead to stabilization of the sludge. However, in South Africa where mesophilic anaerobic digestion of sludge for stabilization is popular, to change to thermophilic aerobic digestion would change the anaerobic digestors from an energy producing (through methane generation) stabilization system, to an energy consuming (through aeration) stabilization system. The process known as dual digestion overcomes this problem. It combines the advantage of autothermal aerobic digestion by providing a certain degree of pasteurization and the advantage of anaerobic digestion by providing energy efficient stabilization.

1.4 THE DUAL DIGESTION PROCESS

The dual digestion process comprises an autothermal thermophilic aerobic first stage and a mesophilic anaerobic second stage. The aerobic stage is based on the principle that if the sludge mass is maintained in the thermophilic temperature range (50–70°C) under aerobic conditions by supply of air or pure oxygen, a population of thermophilic organisms develop which oxidize some of the sludge organic material. Due to the high rate of metabolism at the thermophilic temperatures, sufficient heat is generated biologically to sustain very high sludge loading rates and typically retention times of 1–2 days are possible. At this retention time only a small portion of the organic matter in the sludge is oxidized so that only partial stabilization is achieved. In the anaerobic stage the stabilization of the sludge is completed and, because very little sludge stabilization takes place in the aerobic reactor, the gas production is not significantly reduced. Indeed, it is claimed that due to the pretreatment of the sludge in the aerobic stage, the anaerobic digestor apparently can be operated at significantly shorter retention times (~ 10 days) than conventional sludge stabilization anaerobic digestors. In combining a short retention time autothermal thermophilic aerobic reactor and a mesophilic anaerobic digestor, a number of advantages are apparently obtained. The advantages that are claimed for the dual digestion process are that in the aerobic reactor;

(1) the thermophilic temperatures pasteurize the sludge making it safer for disposal,

(2) the sludge is 'pretreated' through partial solubilization of particulate organic matter allowing short retention times (10 days) in the anaerobic stage,

(3) solubilization produces alkalinity through ammonification of proteins lending greater pH stability to the anaerobic stage,

(4) very little sludge stabilization takes place in the aerobic stage – only to the degree that the heat generated biologically maintains thermophilic temperatures; final and full stabilization takes place in the anaerobic stage – methane generation is reduced only marginally by the aerobic pretreatment stage,

(5) the major portion of the heat required to achieve thermophilic temperatures is biologically generated.

(6) In the anaerobic stage, the heat required to maintain mesophilic temperatures is derived solely from the hot aerobic sludge feed with the result that the methane generated can be used for purposes other than digestor heating.

From a practical point of view, claim (2) is of particular importance; if substantiated, then dual digestion would be a viable system for upgrading existing fully loaded anaerobic digestors. This single feature would justify implementation of dual digestion with sludge pasteurization being an additional benefit. The dual digestion system is therefore particularly appropriate to South Africa where anaerobic digestion is the preferred method for sludge stabilization; where pasteurization of sludge is required or where existing digestors are overloaded, dual digestion could be implemented to meet both pasteurization and increased capacity requirements. With dual digestion, if the claims can be substantiated, anaerobic digestor sludge loading rates can be increased by a factor of 2 as well as eliminating the need to replace or renovate ageing heat transfer equipment. On the negative side it appears that the system as conventionally designed, qualifies only as a process to significantly reduce pathogens (PSRP) and not as a process to further reduce pathogens (PFRP) in terms of the current USA sludge regulations (see Heidman, 1989). However, being a pre (stabilization) pasteurizing process, it probably will overcome the problem of sludge recontamination which the Swiss experienced with post (stabilization) pasteurization mentioned above.

1.5 DUAL DIGESTION RESEARCH REQUIREMENTS.

Early in the development of the dual digestion process in the USA, the potential benefits of the dual digestion process were recognized by the Water Research Commission (WRC) in South Africa and in 1979 a research programme between the WRC and the Johannesburg City Council (JCC) was initiated. This research focused on the performance of the autothermal thermophilic aerobic reactor with regard to its effectiveness of pathogen destruction/inactivation, to evaluate operating parameters such as sludge loading rate, retention time, oxygen consumption, volatile solids (VS) removal and to assess the prospects of combining it with second stage anaerobic digestion with a view to pathogen reduction combined with cost-effective sludge stabilization.

The results of the research were very encouraging (Trim, 1984a, 1984b; Trim and McGlashan, 1985). The pilot plant aerobic reactor was found to operate reliably and stably. It could be operated at short retention times and efficient utilization of the pure oxygen was achieved. Disinfection of sludge was attained (zero viable *Ascaris* ova) provided sufficiently high temperatures were maintained (> 60°C). The sludge produced by the pilot plant was capable of undergoing anaerobic digestion but it exhibited very poor dewatering properties. It was estimated that despite very efficient utilization of oxygen the principal cost of the reactor lay in the pure oxygen supply. From the study it was concluded that the aerobic reactor could serve as a first step in dual digestion but that to make the process as economically attractive as possible, it was necessary to investigate using a combination of air and pure oxygen for the aerobic reactor.

The JCC/WRC research demonstrated that the dual digestion process holds promise to produce a sludge which is stable and disinfected. However, before the process could be adopted a number of aspects needed careful study in order to ensure successful application in South Africa, viz.

1) Operation at a sufficiently large scale to develop design criteria and undertake an economic evaluation.

2) Oxygen requirements and oxygen utilization efficiency, with pure oxygen and air oxygenation and the influence of the foam layer resulting from air oxygenation.

3) Minimum retention time of the aerobic reactor without compromising attainment of thermophilic temperatures.

4) Temperature control of the aerobic reactor and anaerobic digestor.

5) Efficacy of pathogen inactivation in the aerobic reactor and mode of operation to prevent recontamination.

6) Minimum retention time of the anaerobic digestor without compromising digestor performance and sludge stability.

7) Performance of anaerobic digestor with respect to VS removal, gas production, sludge stability and dewaterability.

In 1987 under the sponsorship of the Water Research Commission and Milnerton Municipality, a full scale (45 m^3 aerobic reactor and 600 m^3 anaerobic digestor) research project into dual digestion was initiated at the Potsdam Wastewater Treatment Works (Milnerton, Western Cape) to investigate the above aspects of the dual digestion system. Contributing organizations to this project were the Water Research Commission, the Milnerton Municipality, the Division of Water Technology of the CSIR and the Department of Civil Engineering at the University of Cape Town, the latter two for scientific and academic input.

The research presented in this thesis focuses on the performance, control and design of the autothermal thermophilic aerobic reactor in dual digestion based on the results obtained on the Milnerton aerobic reactor.

Already at the design stage of the Milnerton aerobic reactor, it became evident that biological heat generation and oxygen requirements were poorly understood: From past research a number of parameters, largely empirical, were available for describing autothermal thermophilic aerobic sludge stabilization, but these were not useful because firstly, they were vague and produced conflicting results and secondly, they were oriented towards describing sludge degradation in sludge stabilization; in the aerobic stage of dual digestion, sludge stabilization is *not* an objective, but sludge pretreatment and pasteurization are. Accordingly it was necessary to

1) review and investigate the stoichiometry and kinetics of aerobic biological heat generation,

2) identify the most suitable parameter in terms of which to quantify the biological heat generation,

3) quantify oxygen requirements for design purposes,

4) develop a mathematical model to simulate the aerobic reactor temperature for design and control purposes.

The results obtained in the Milnerton dual digestion project for the aerobic reactor insofar as the above four tasks are concerned are presented in this thesis.[1]

In Chapter 2 the kinetics of biological heat generation in thermophilic aerobic digestion is reviewed and found to be deficient for application to the aerobic reactor in dual digestion because (1) VS removal is not an objective in the aerobic reactor and (2) the reactor is likely to be operated under oxygen limiting conditions. Oxygen consumption is proposed as a means for evaluating biological heat generation.

In Chapter 3 the layout, equipment and operating procedures of the Potsdam dual digestion system, are described as well as the equipment, procedures and calculations employed to construct and solve simultaneous oxygen mass and unsteady state heat balances, and the VS, TS and COD balances, across the aerobic stage of the dual digestion system. A sample solution of a simultaneous

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[1] Details of the operation and performance of the anaerobic digestor are not presented in this thesis; these are reported by de Villiers *et al.* (1991). Aspects of the anaerobic digestor operation and performance that are presented are those required for clear exposition of the aerobic reactor operation and performance as part of the dual digestion system and can be found in the following sections:

- Dual digestion plant layout and operation – Chapter 3, Section 3.2
- Operation of anaerobic digestor – Appendix 5B, Section 5B.2

A list of all the publications to date emerging from the Milnerton dual digestion project is given at the end of this chapter after the references.

oxygen mass and an unsteady state heat balance in order to calculate a specific heat yield (MJ generated/kgO consumed), and the procedures and calculations whereby monitoring equipment was calibrated to ensure accuracy in measured data, are given in this chapter.

In Chapter 4 the results of 116 specific heat yield tests, as well as those of the TS, VS and COD concentration measurements, are described. From these results it is shown *inter alia* that the biological heating rate was directly proportional to the biological oxygen consumption rate with the constant of proportionality, called the specific heat yield (Y_H), measured to be 12,77 ± 0,58 MJ/kgO. The reactor temperature could be completely and instantaneously controlled by means of the oxygen supply rate for as long as the reactor was oxygen limited. It is demonstrated that the biological oxygen consumption rate and the oxygen supply rate, are two pivotal parameters in the modelling, design and control of autothermal thermophilic aerobic reactors in dual digestion.

In Chapter 5 design considerations for pure oxygen and air oxygenated aerobic reactors are outlined based on the steady state heat balance. Accepting that the objectives of the aerobic reactor are sludge pretreatment through oxygen limitation and pasteurization, it is demonstrated that three parameters are of crucial importance to design which centres around minimizing the reactor retention time: (1) the oxygen utilization rate (OUR) of the sludge which fixes the maximum biological heat generation rate of the sludge, (2) the oxygen transfer rate (OTR) of the oxygenation system which should be somewhat less than OUR to ensure oxygen limitation, and (3) the oxygen transfer efficiency, which determines the vent gas volumetric flow rate.

In Chapter 6 procedures are developed which enable the temperature profile of the reactor sludge temperature versus time in a batch fed aerobic stage of a pure oxygen or air oxygenated dual digestion system to be simulated. With the aid of an unsteady state heat balance across the reactor over a small discrete time interval, Δt, the increment or decrement in reactor sludge temperature may be calculated over this time interval. Repetitive computation of the unsteady state heat balance over a large number of time intervals allows the profile of the reactor sludge temperature to be plotted against time. From this procedure an interactive computer model is developed which calculates and plots the sludge temperature versus time in a batch fed reactor which is oxygenated either by air or by pure oxygen. The model is shown to accurately predict the temperature–time profiles in the aerobic reactor of the Milnerton dual digestion system for a range of different operating conditions.

In Chapter 7 the conclusions emerging from the research on the aerobic reactor are presented.

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CHAPTER 2

REVIEW OF THEORY AND LITERATURE PERTAINING TO BIOLOGICAL HEAT GENERATION IN THE AEROBIC STAGE OF DUAL DIGESTION

ABSTRACT

In this chapter the kinetics of biological heat generation in thermophilic aerobic digestion is reviewed and found to be deficient for application to the aerobic reactor in dual digestion because (1) VS removal is not an objective in the aerobic reactor and (2) the reactor is likely to be operated under oxygen limiting conditions. Oxygen consumption is proposed as a means for evaluating biological heat generation. This approach finds support in the literature of theoretical bioenergetics from which a specific heat yield Y_H (MJ heat generated per kg oxygen consumed) of 13 MJ/kgO is estimated. Practical measurements of the specific heat yield in the fields of human physiology, microbiology and operating thermophilic aerobic digestors support this approach, but widely differing values of Y_H are reported, particularly in the latter 2 fields.

2.1 INTRODUCTION – SLUDGE DIGESTION

The principal objective of a sludge digestion process is to produce a stable, dewaterable sludge in the shortest possible process retention time. Operators of a sewage treatment works will generally judge a sludge to be stable when it neither generates unpleasant odours nor attracts flies. Scientists and sanitary engineers define a stable sludge as one in which the biodegradable substrate energy has been reduced to a level which does not support excessive microbial activity.

Practical measures of the increase in sludge stability are:

1. The decrease in sludge COD across the digestion process.

2. The decrease in volatile organic content (VS) of the sludge.

3. A reduction in biological oxygen demand in the sludge.

The second measure (VS) has traditionally found wide use in defining sludge stability and has the additional benefit of determining how much solid matter has been destroyed by digestion and, accordingly, how much dry mass remains behind for the treatment works to dispose of.

The above definitions of sludge stability cannot be used in a fixed measure to quantify the degree of sludge stabilization. It is the removal of *biodegradable* energy that results in stability. Hence a sludge in which only 40% of the volatile organic matter is biodegradable will be fully stabilized by a 40% VS removal. Nevertheless guidelines and regulations governing sludge disposal generally do specify the fraction of VS that a treatment process should remove from sludge in order to stabilize it, for example, a 38% reduction in volatile suspended solids (VSS) - this criterion varies from country to country.

The stabilization of sludge at a wastewater treatment facility is typically accomplished by either aerobic or anaerobic digestion. Dual digestion combines both of these processes into a single system, in which both sludge stabilization and pasteurization are achieved. Because this thesis describes an investigation into biological heat generation in the *aerobic stage* of a dual digestion system, anaerobic digestion will not be discussed, other than where it serves to clarify details of the research plant at Milnerton and where operational experience with the research plant is reported (see Appendix 5B).

A wealth of experience and information has been gained regarding the theory and practice of aerobic digestion at both ambient and thermophilic temperatures. It is tempting, for lack of a better alternative, to use this experience and information in quantifying biological heat generation in the design and modelling of the aerobic stage of dual digestion. This temptation is further encouraged by the apparent similarities between the processes. However, focusing on these similarities and on the experience gained in aerobic digestion, may be misleading and inappropriate to the requirements of the aerobic stage of dual digestion. Nevertheless, before proceeding with the task of evaluating procedures which may be used to describe biological heat generation in the aerobic stage of dual digestion, aspects of both conventional and thermophilic aerobic digestion merit attention, so that their design objectives and requirements may be more clearly distinguished from those of the aerobic stage in dual digestion.

2.2 THE PROCESS DESIGN OF CONVENTIONAL AEROBIC DIGESTION

The hydraulic retention time and the oxygen demand required by aerobic digestion to effect an adequate degree of sludge stabilization, largely determine the capital and operating costs of the process and are of primary importance in the process design of aerobic digestion. The hydraulic retention time is controlled by the kinetic parameter dVS/dt – the rate of volatile solids reduction in the aerobic digestor. It is desirable for this rate to be as high as possible in order to minimize the sludge retention time required to effect a given degree of VS reduction in the system. Oxygen demand is determined from a stoichiometric ratio of the mass of oxygen required per unit mass of VS removed. Because oxygen and COD are equivalent the COD/VS ratio of the sludge is used for this purpose where COD removed is equivalent to oxygen required.

Sanitary engineers generally define the rate of VS reduction under oxygen sufficient conditions in terms of a first order degradation process:

$$\frac{dBVS}{dt} = k \times [BVS]_t \quad (kg/h) \tag{2.1}$$

where

BVS = biodegradable volatile solids concentration (kg/m³)

k = VS reduction rate coefficient (h^{-1})

t = time

and subscript t denotes time t.

For a completely mixed continuous flow aerobic digestor the above equation can be solved to determine the hydraulic retention time, R_H:

$$[BVS]_f = \frac{[BVS]_t}{(1 + k.R_h)} \quad (kg/m^3) \tag{2.2a}$$

$$R_h = \frac{1}{k}\left(\frac{[BVS]_i}{[BVS]_f} - 1\right) \quad (h) \tag{2.2b}$$

where

R_h = hydraulic retention time (h)
and subscripts i and f denote initial and final states respectively.

The carbonaceous oxygen utilization rate in the aerobic digestor is found from the product of the rate of VS reduction and the COD/VS ratio of the sludge i.e.

$$OUR = k \times [BVS]_f \times \frac{COD}{VS} \qquad [kgO/(m^3.h)] \qquad (2.3)$$

where
OUR = carbonaceous oxygen utilization rate $[kgO/(m^3.h)]$

In aerobic digestion at ambient temperatures, oxygen may also be consumed by nitrification in the sludge but this oxygen demand does not contribute to VS removal and therefore will be omitted in this discussion.

Thus, in terms of Eqs 2.2 and 2.3, the rate coefficient, k, is the single most important variable in determining both the retention time and the rate at which oxygen is consumed in order to effect a specified degree of sludge stabilization. Its value increases with temperature and this dependency is frequently expressed using an Arrhenius equation:

$$k_T = k_{20}\, \theta^{(T - 20)} \qquad (h^{-1}) \qquad (2.4)$$

The increase in k with temperature provides the rationale for operating aerobic digestion at thermophilic temperatures.

2.3 THERMOPHILIC AEROBIC DIGESTION

2.3.1 The justification for thermophilic operation

The rate of VS removal in aerobic sludge digestion occurs at a much greater rate at thermophilic temperatures than at conventional ambient temperatures. Hence a specified degree of VS reduction can be achieved at shorter retention times in a thermophilic aerobic digestor than in a conventional aerobic digestor. The resulting reduction of capital cost with no increase in oxygen demand being required (the oxygen requirement for unit VS oxidation is assumed to remain constant with temperature) is the direct benefit of such operation. The benefits of increasing temperature are illustrated in Fig 2.1 which shows the % removal of

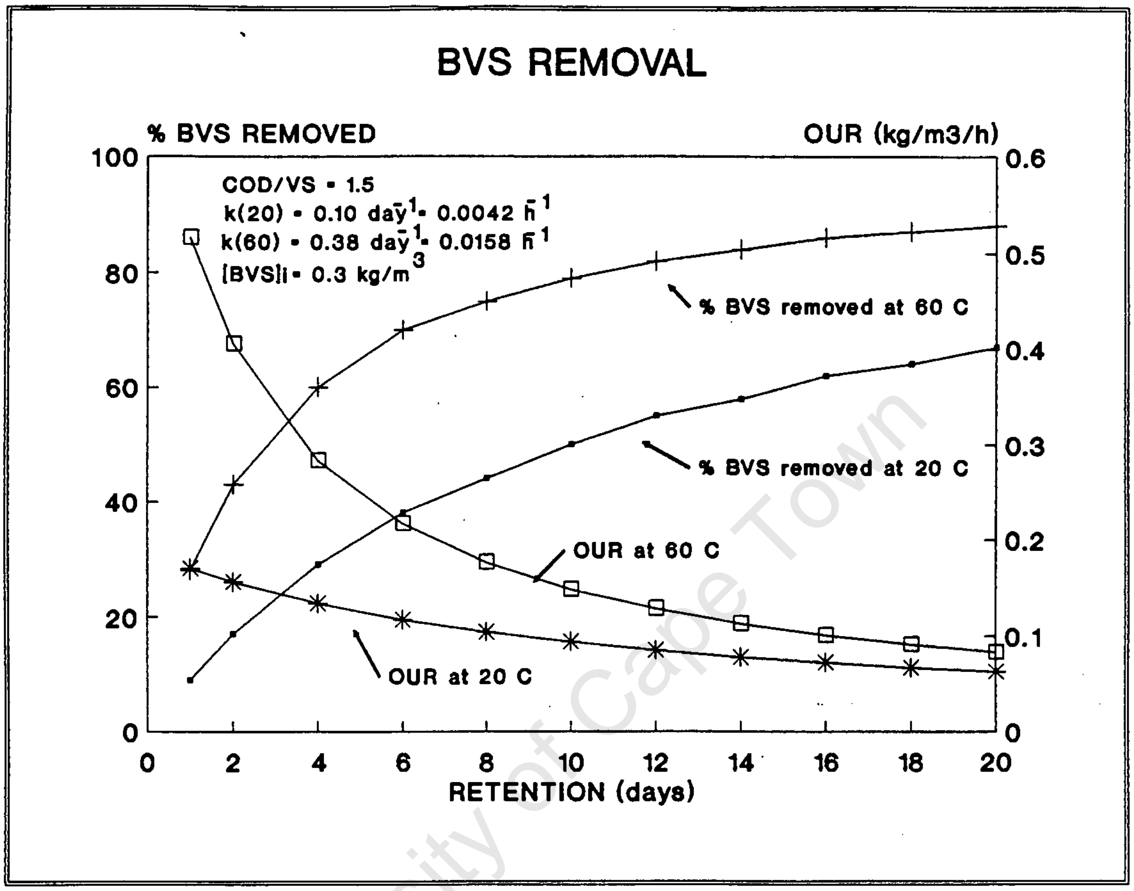

Fig 2.1: The % removal of biodegradable volatile solids (BVS) and the sludge oxygen utilization rate (OUR) at 20°C and 60°C as functions of retention time (R_H), in an aerobic digestor. The plots are derived from calculations based on 1st order BVS removal kinetics [Eq (2.2b) and Eq (2.3)].

BVS [from Eq (2.2a)] and oxygen utilization rate based on a COD/VS ratio of 1,5 [Eq (2.3)] against the retention time, R_H, for values of k at 20°C and 60°C; to effect a removal of 50% of the influent VS requires 10 and 2.6 day sludge hydraulic retention times at 20°C and 60°C respectively. While the OUR needed to effect the reduction of sludge BVS increases with the increased temperature from 0.094 to 0.358 kgO/(m³.h), the daily mass of oxygen (excluding nitrification oxygen demand) required to treat a given volume of sludge remains the same because the %BVS removal is the same (note that nitrification is suppressed at thermophilic temperatures). An additional benefit of the thermophilic operation is that the sludge is pasteurized by the high temperature, resulting in the reduction of pathogens and rendering a safer sludge for use as a soil conditioner on land.

In order to heat the digestor feed sludge to thermophilic temperatures and to maintain these temperatures in spite of heat losses to the ambient surroundings, requires heat to be supplied to the process at a defined rate. Hence an important parameter in thermophilic digestion is the rate of heat supply required to maintain a desired operating temperature.

2.3.2 Heat generation in thermophilic aerobic digestion

In thermophilic aerobic digestion, feed sludge enters the digestor at a temperature T_{si} and effluent sludge is discharged from the digestor at a higher temperature T_{se}. The rate at which the sludge enthalpy increases in order to effect this temperature increase from T_{si} to T_{se} may be calculated with the aid of the following equation:

$$H'_{se} = F_s \times \rho_s \times Cp \times (T_{se} - T_{si})$$

$$= \frac{V_p}{R_H} \times \rho_s \times Cp \times (T_{se} - T_{si}) \qquad (MJ/h) \qquad (2.5)$$

where

H'_{se}	= the rate of sludge enthalpy increase	(MJ/h)
F_s	= the flow rate of sludge through the digestor	(m³/h)
R_H	= the retention time of the sludge in the digestor	(h)
V_p	= the operating volume of the digestor	(m³)

ρ_s = the density of the sludge (ton/m³)
C_p = the specific heat of the sludge [MJ/(ton.°C)]
T_{se} = the temperature of the effluent digestor sludge (°C)
T_{si} = the temperature of the feed sludge to be digested (°C)

The rate of heat supply required to maintain the digestor at a specified temperature, T_{se}, must be greater than the rate of increase in sludge enthalpy, H'_{se}, in order to compensate for heat which is lost from the digestor to various heat sinks. The losses are typically due to heat transfer through the digestor walls and due to the vaporization of water from the hot mixed liquor via vent gases. Both of these heat losses can be kept to a minimum: The former by lagging the digestor walls, and the latter by ensuring that the vent gas flow rate is as small as possible. Although the water vaporization heat losses in the vent gas may be insignificant when pure oxygen is used to oxygenate the digestor sludge, in air oxygenated thermophilic digestors the large volumes of inert nitrogen passing through the sludge can cause substantial water vaporization heat losses which can be critical in determining whether or not the desired operating temperature will be achieved. Much discussion of the thermophilic aerobic digestion process deals with the relative viability of using air as opposed to pure oxygen (Andrews and Kambhu, 1971; Gould and Drnevich, 1978; Wolinski, 1985) to maintain thermophilic temperatures in the sludge.

The rate at which heat must be supplied from various heat sources in order to maintain a digestor sludge temperature T_{se} is calculated with the aid of the following equation:

$$H_{sources} = H'_{se} + H_{losses} \qquad \text{(MJ/h)} \tag{2.6}$$

where,

$H_{sources}$ = the rate at which heat must be supplied to a digestor/reactor in order to maintain a desired thermophilic temperature, T_{se}, in the sludge (MJ/h)
H_{losses} = the rate at which heat is lost from the digestor (MJ/h)

There are a number of sources from which, $H_{sources}$, can be derived:

(1) The mechanical action of a pump or a mixer.

(2) Submerged combustion of a fuel gas such as anaerobically produced methane.
(3) The injection of steam.
(4) Autothermally ie the generation of biological heat within the digestion process through oxygenation of the sludge with pure oxygen or air as in the autothermal thermophilic aerobic digestion process (ATAD).

The kinetics of the former three heating options above are easy to determine from standard thermodynamic engineering techniques but biological heat generation is less clearly described in the literature. To determine this is the major objective of this thesis.

In the literature, biological heat generation is generally defined stoichiometrically ie the quantity of heat generated for each unit change in a designated parameter. The stoichiometric definition is then given suitable kinetic expression. This approach is used in the following discussion of biological heat generation.

2.4 THE STOICHIOMETRY OF BIOLOGICAL HEAT GENERATION

The manner in which the stoichiometry of biological heat generation is defined is largely influenced by both the system in which it occurs and the priorities of those who wish to quantify it. Thus physiologists define biological heat production in the human body in terms of oxygen consumed during respiration (Consolazio, 1963). Sanitary engineers investigating autothermal thermophilic aerobic digestion (ATAD) have related it to the reduction of sludge volatile solids concentration (Andrews and Kambhu, 1969).

2.4.1 Definition of biological heat generation in aerobic digestion

In a biological system, the degradation of organic matter and the growth of biomass are accompanied by the release of heat. Biological heat generation is the waste heat arising from the inefficient capture and utilization of energy by micro-organisms. In a biological culture such as that found in aerobic sludge digestion, the stepwise transformations of substrate energy by micro-organisms are each subject to a degree of inefficiency which manifests itself in the generation of heat.

It is reasonable to postulate that the energy released from a sludge by an aerobic biological culture is liberated principally via oxidation reactions and that this energy has only two ultimate destinations:

(1) Capture in synthesized chemical bonds eg. growth of new micro-organisms.
(2) Waste heat.

It is the waste heat that is available as a source of biological heat in a process such as thermophilic aerobic digestion. The quantity of heat that is available may be determined from an energy balance.

2.4.2 Steady state energy balance

In accordance with the postulation stated in the 2 points above, the aerobic digestion process will change the energy stored in the chemical structure of a sludge. By measuring the change in sludge energy across a digestor with the aid of a steady state energy balance, it is possible to determine the amount of heat which has been released in the digestor. For the purpose of this determination, the energy balance is subject to the following specific constraints:

(i) the system is at constant temperature and pressure
(ii) no phase changes occur in the digestor i.e. no vaporization of water is considered and all heat generated is abstracted in order to maintain constant temperature.
(iii) potential and kinetic energies do not change through the system.

The energy balance can be stated as follows: 'The change in heat released by the biological activity within the digestor is equal to the decrease in free energy within the sludge flowing through the system'

$$Q_{bi} = G_{in} - G_{out} \tag{2.7}$$

where,

Q_{bi} = a quantity of biologically generated heat
G_{in} = the free energy of feed sludge to a digestor
G_{out} = the free energy of effluent sludge from a digestor.

2.4.3 The specific heat yield

Absolute measures of G_{in} and G_{out} are not possible and only a measure of their change is necessary to enable the determination of the heat generation. If it can be shown that the free energy content of the sludge is proportional to some substitute measurable parameter, P, then the stoichiometry of the heat generation when the bioreactor is at steady state can be presented as:

$$Q_{bi} = Y_H (P_{in} - P_{out}) \qquad (2.8)$$

where,

Y_H = the quantity of biological heat generated per unit change in substitute energy parameter, P.

P = the substitute sludge energy parameter.

In this thesis the constant of proportionality, Y_H, will be referred to as the *specific heat yield.*

2.4.4 Substitute measures of energy used in quantifying biological heat generation

In literature relating to aerobic sludge digestion systems, three substitute measures of sludge energy are commonly used to predict heat generation:

(i) sludge heat of combustion
(ii) chemical oxygen demand (COD)
(iii) sludge VS content

All three parameters have drawbacks when applied to design, control and simulation of aerobic sludge digestion systems. These drawbacks arise from the practical use of the three parameters; the drawbacks are discussed below after evaluations of the theoretical soundness of each parameter.

2.4.4.1 *Heat of combustion*

Theoretical support for sludge heat of combustion can be found in the work of Linton and Stephenson (1978) who related the substrate heat of combustion to the maximum biomass yield coefficient, both based on a 'unit weight of substrate carbon'. Hamer and Bryers (1985) suggested assessing the heat generation in an aerobic bioreactor on the basis of the difference between the heats of combustion of the products and reactants. However while the heat of combustion may be a theoretically sound substitute measure of the change in sludge energy, the disadvantage of this method is that the average wastewater laboratory is not equipped to measure heats of combustion. Hence it is inappropriate for routine use in the operation of ATAD.

2.4.4.2 *Chemical oxygen demand*

Much theoretical and empirical support can be found for using chemical oxygen demand as a substitute energy parameter in predicting heat generation. Servizi and Bogan (1963) showed that the biomass synthesized is proportional to the amount of substrate COD utilized and hence heat generation in a bioreactor should likewise be proportional to that value. This is in accordance with the earlier definition in Section 2.4.1 that the energy released during biological oxidation has only 2 destinations – the growth of biomass and waste heat.

McCarty (1972) showed that with oxygen as an electron acceptor, and with products and reactants at unit activity, the energy released by different organic substrates per mole of electrons transferred falls within the range –24.4 to –30.2 kcal/e^- mole. These extremes lie within 11% of their arithmetic mean (27.3 kcal/e^- mol) and it is very likely that the value of this parameter for a mixed substrate like sludge, lies close to the mean of 27.3 kcal/e^- mole. Expressed in oxygen mass equivalents the average of 27.3 kcal/e^- mole is equivalent to 14.28 MJ/kgO.

Kambhu (1971) measured the heats of combustion of thermophilic organisms, relating them to the organism COD with an average of 5860 BTU/lb COD (13.6 MJ/kgCOD) being measured. Wright (1975), as quoted in Jewell and Kabrick (1980), observed experimentally that the range of energy of organic materials lay between 3 to 4 kcal/gCOD (12.6 – 16.7 MJ/kgCOD).

Thus COD is a suitable substitute energy parameter in Eq (2.8) with YH defined as MJ/(kgCOD removed). Although sensitive to sampling and analytical error, it is a routine test conducted at wastewater plants. Furthermore a change in COD is equivalent, by definition, to oxygen demand and thus heat generation and process oxygen requirements can be directly related.

2.4.4.3 *Volatile solids*

The volatile solids test finds use as a substitute energy parameter in sludge treatment because, whether appropriate or not, guidelines often define sludge stability in terms of a certain % VS reduction. Hence, by convenience VS has become the substitute energy parameter most frequently used in defining biological heat generation in ATAD. In models of ATAD, Andrews and Kambhu (1969) and Vismara (1985) expressed the stoichiometry of biological heat generation in terms of the VS removal across the bioreactor, where each kg of VS removed generated a fixed amount of heat. The value of Y_H used in these models was the average heat of combustion of the sludge expressed in VS mass equivalents. Values of Y_H are reported in the work of Fair and Moore (1932) and can range from 20 to 28 MJ/kgVS removed.

Provided the COD/VS ratio of the sludge remains constant during treatment, the VS parameter is as theoretically sound as the COD parameter for use as a substitute measure of energy.

2.4.4.4 *Practical drawbacks of the three parameters*

All three of the above stoichiometric parameters share common disadvantages. These are:

(1) Their accuracy depends on the correct and representative sampling of sludges. When dealing with sludges, even the most vigilant sampling procedures are prone to error.

(2) They are all, from an operational point of view, historical data and the information derived from them is at least 4 hours old. Although this is of no consequence when the data is to be used in process design, it is a disadvantage where it is required for process temperature control where much more immediate information is required.

2.5 THE KINETICS OF BIOLOGICAL HEAT GENERATION

The rate at which heat must be supplied to an aerobic thermophilic digestor in order to maintain a specified thermophilic temperature in the sludge, T_{se}, is defined by Eq (2.6). If this heat is to be generated biologically then the kinetics of biological heat generation must be understood if a desired rate of heating is to be achieved.

In the literature on ATAD, biological heat generation stoichiometry has been related to either VS (Andrews and Kambhu, 1969; Jewell and Kabrick, 1980) or COD (Andrews and Kambhu, 1971; Gould and Drnevich, 1978) removal. In contrast, biological heat generation kinetics has generally been presented as a function only of BVS removal kinetics (with the exception of a model proposed by Vismara (1985) which uses an empirical relationship based on the work of Koers and Mavinic (1977). Equation (2.1) is a description of the BVS removal kinetics and from this the biological heat generation kinetics in terms of VS or COD may be presented as:

When the stoichiometry is related to VS:

$$H_{bi} = Y_H \times \frac{d[BVS]}{dt} = Y_H \times k \times [BVS]_f \qquad (MJ/h) \qquad (2.9a)$$

When the stoichiometry is related to COD:

$$H_{bi} = Y_H \times k \times [BVS]_f \times \frac{COD}{VS} \qquad (MJ/h) \qquad (2.9b)$$

where,

H_{bi} = the rate of biological heat generation (MJ/h).

The oxygen demand exerted by thermophilic aerobic digestion is only to meet carbonaceous requirements as nitrification does not occur at thermophilic temperatures:

$$OUR = k \times [BVS]_f \times \frac{COD}{VS} \qquad [kgO/(m^3.h)]) \qquad (2.10)$$

These kinetics mean that the BVS removal rate coefficient, k, not only controls the retention time required to effect a desired VS removal (Eq 2.2b), and the oxygen utilization rate in the digestor, but it also controls the biological heat generation rate. Andrews (1971) has presented a number of values for k which were measured for different sludges at different temperatures.

Figure 2.2 illustrates the use of Eqs (2.6) and (2.9b) with a COD/VS ratio of 1.5 kgCOD/kgVS and a specific heat yield Y_H of 14 MJ/kgCOD removed. In Eq (2.6) the heat losses through the reactor walls and due to water vaporization are assumed to be 50% of $H_{sources}$. At a hydraulic retention time of 3.5 days the

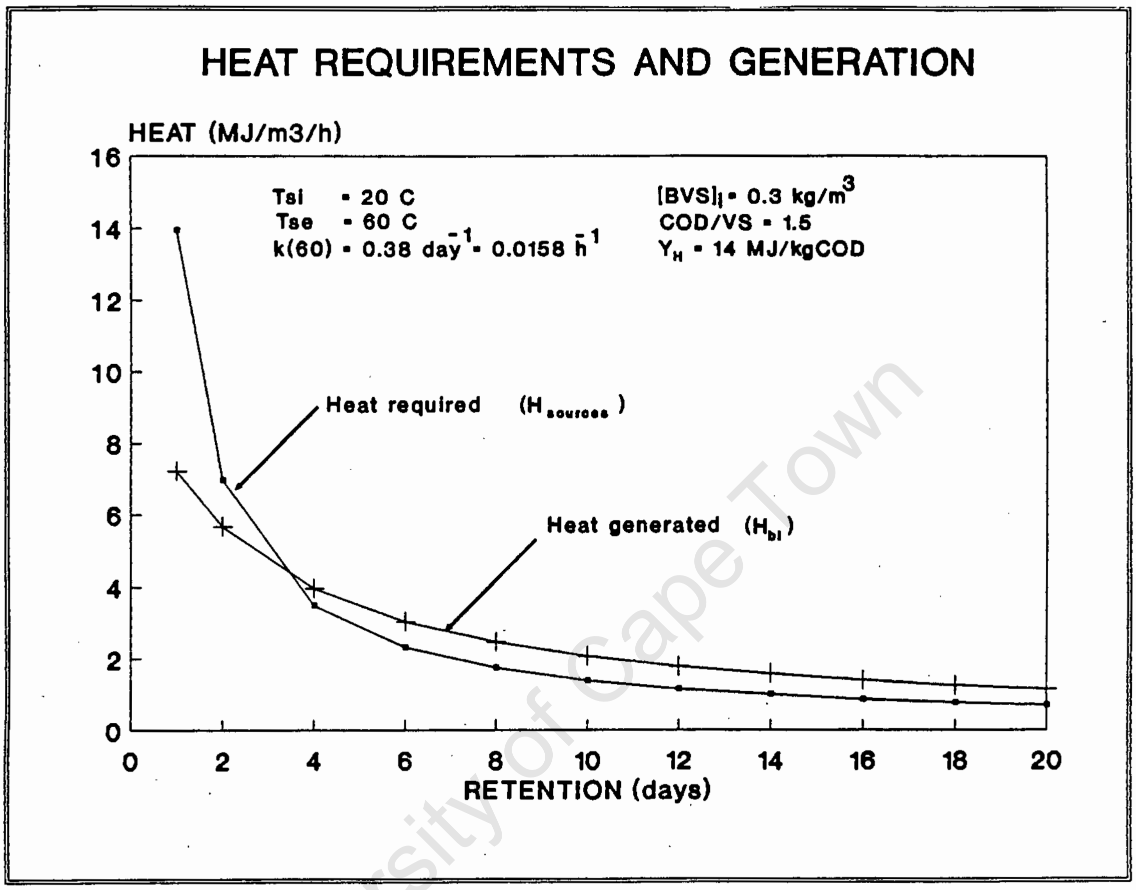

Fig 2.2: The rate of heating required ($H_{sources}$) to maintain a digestor sludge temperature (T_{se}) of 60° C, and the rate of biological heat generation (H_{bi}) as functions of hydraulic retention time (R_H) in a thermophilic aerobic digestor. The rate of biological heat generation is calculated on the basis of 1st order BVS removal kinetics [Eq (2.9b)]. At a retention time of 3.4 days the required rate of heating and the biological heat generation rate are equal and the digestor will operate at a temperature of 60° C and will remove 53.4% of the feed sludge BVS (see Fig 2.1). At retention times of less than 3.4 days, less than 53.4% of the feed sludge BVS will be removed and the digestor sludge temperature will be less than 60° C.

rate of biological heat generation is equal to the required heat generation rate. *Hence the rate of biological heat generation determined the minimum hydraulic retention time required to operate the aerobic digestor at the desired temperature of 60 °C.* At a shorter retention time the biological rate will be insufficient and sludge temperatures will drop below 60°C and, conversely longer retention will result in temperatures greater than 60°C. Figure 2.1 gives the % BVS removal and the OUR at the 3.5 day retention to be 53.4% and 0.31 kg/(m^3.h). If a larger BVS removal is required, then the digestor sludge temperature will increase above 60°C.

However this approach to the kinetics of biological heat generation has a number of flaws:

(1) The fraction of biodegradable volatile solids (BVS) in sludge cannot be measured. The reduction of the volatile solids content of a sludge is accompanied by the generation of inert endogenous residue. This residue is measured as VS and cannot be distinguished from the biodegradable VS. Hence in batch digestion of sludge, biodegradable VS is reduced while inert VS is generated by endogenous respiration: For 100 units of BVS reduction, 20 units of inert VS are generated and 80 units are oxidized to CO_2 and H_2O (Marais and Ekama, 1976). Equation (2.1) does not account for inert VS generation.

(2) The formulation of Eq (2.1) for the BVS removal kinetics applies to a specific condition where endogenous respiration is the dominant biological activity. This occurs, for example, when waste activated sludge is treated by conventional aerobic digestion. In this instance, the loss of VS is because a fraction of the organism mass disappears to provide energy for the maintenance of the mass remaining i.e. the substrate is internally generated (Dold, Ekama and Marais, 1980). However, when a primary sludge serves as substrate in a digestor, biodegradable substrate external to the organism is available, and this substrate is available for synthesis of new organism mass after passing through the phases of adsorption, storage and hydrolysis. These biological activities are not described by the endogenous respiration type formulation of Eq (2.1). Similarly, in thermophilic aerobic digestion, the feed sludge (primary or waste activated sludge) which serves as substrate to the digestor, undergoes a temperature transition from ambient

to thermophilic temperature. None of the live biomass in the feed sludge are likely to survive this transition to a thermophilic temperature. The biodegradable component of this live biomass will serve as substrate to a different and acclimatized culture of thermophilic organisms in the thermophilic aerobic digestor. Thus kinetic modelling on the basis of endogenous respiration is inappropriate.

(3) The reliability of predictive equations based on the VS removal rate coefficient, k, is doubtful due to the wide variation shown by measured values of k. The wide range of values, even where measured at the same temperature, suggests that individual determinations are required in digestor designs for different sludges. It is possible that this variation is a consequence of the incorrect determinations of k due to point (1) above. In detailing a method of measuring the endogenous respiration constant[1] for the activated sludge model, Marais and Ekama (1976) have concluded that measuring the change in VS concentration with time in a batch digestion is not a good procedure for determining this constant. They propose that this is best achieved by measuring the change in oxygen utilization rate with time in a batch digestion as well as incorporating the generation of inert endogenous residue.

(4) As will be evident in the results of the research reported in this thesis, biological heat generation and COD removal can occur without a measurable VS removal ie the COD/VS ratio of the sludge can change during digestion. This demonstrates the unreliability of VS as a substitute energy measure in sludge digestion.

(5) From (4) above, biological heat generation stoichiometry appears to be more reliably expressed in COD equivalents than in VS equivalents. However, any advantage gained by adopting this approach is nullified by linking the COD based stoichiometry to VS removal kinetics via a COD/VS ratio [as in Eq (2.9b)] as this implies proportionality between COD and VS.

(6) The equations describing VS degradation kinetics imply that oxygen

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[1] The endogenous respiration constant is conceptually equivalent to the VS removal rate coefficient, k.

supplied is in excess of biological oxygen demand. When this condition is not met, the above equations no longer apply. The rate of oxygen dissolution in the sludge will then control the rates of VS removal and biological heat generation. Consequently k values can be suppressed under oxygen limited conditions.

In spite of the shortcomings cited above, for lack of a better alternative, VS removal kinetics have found wide use in describing biological heat generation (and the rate of sludge stabilization) in thermophilic aerobic digestion.

When considering biological heat generation in the aerobic stage of the dual digestion process, it may be thought appropriate, for lack of a better approach, to apply the above mentioned kinetics of biological heat generation directly to the aerobic stage of dual digestion. Such a step is encouraged by the apparent similarity between ATAD and the aerobic stage of dual digestion. However, taking cognizance of the above cited deficiencies in the ATAD heat generation/VS removal kinetics, this approach to evaluating biological heat generation in the first stage of dual digestion may not be appropriate for the following reasons:

(1) VS removal is not an objective in the aerobic stage of dual digestion.
(2) The stage is likely to be oxygen limited.

With regard to the system being oxygen limited, it may appear from Fig 2.2 that this cannot be so at short retention times: The heat required to maintain a thermophilic temperature of 60°C is greater than the biological heat generated through VS destruction. To enable the sludge to generate the heat through VS destruction at its maximum biological rate, sufficient oxygen needs to be supplied. Hence in the aerobic reactor of dual digestion at short retention times, it would appear that oxygen sufficient conditions are required to generate heat at maximum rate so that oxygen limiting conditions cannot be imposed on the aerobic reactor at short retention times. However, in this investigation, it was possible to operate the aerobic reactor at short retention time (1,25d), and generate sufficient heat for pasteurization (60°C) under oxygen limiting conditions and with negligible VS destruction (<6%). This clearly demonstrated that heat generation is not necessarily directly linked to VS destruction at short retention times and underlines the deficiency of the aerobic digestion VS destruction approach to model the aerobic reactor in dual digestion.

2.6 THE AEROBIC STAGE OF THE DUAL DIGESTION PROCESS

2.6.1 Process objectives

The layout and objectives of the dual digestion process have been discussed in the previous chapter. As a system, the aerobic stage of dual digestion and an ATAD digestor are similar. However they differ in their objectives and operation. In dual digestion the aerobic stage is not required to stabilize sludge – this function is fulfilled by the anaerobic stage – hence achieving VS removal is not important. The role of the aerobic stage is to expose the sludge to thermophilic temperatures in order to *disinfect* it and to *pretreat* it to enhance subsequent anaerobic digestion.

2.6.1.1 *Sludge disinfection*

Disinfecting of sludge by means of pasteurization requires the sludge to be exposed to a temperature, which exceeds a defined minimum, for a minimum length of time. These minima vary and standards for disinfection may differ from country to country but generally are of the order of not less than 55°C for not less than 2 hours. In a biological process, such as the aerobic stage of dual digestion, the pasteurization temperature therefore will fall in the range 55 to 70°C.

2.6.1.2 *Sludge pretreatment*

The sludge pretreatment achieved in the aerobic stage of dual digestion is said to enable the stable operation of the subsequent anaerobic stage at shorter retention times than would be possible in conventional mesophilic anaerobic digestion. This pretreatment may be due to both the biological oxidation reactions occurring in the sludge and to the thermal effects caused by the thermophilic temperatures[2]. Nevertheless, research by Mason (1986) suggests that oxygen limitation in the aerobic stage enhances its pretreatment ability. It increases production of extra-cellular enzymes and hence more solubilization of particulate material occurs, which is favourable for the subsequent anaerobic stage – according to Eastman and Ferguson (1981), the hydrolysis of particulate material to soluble substrates is the rate limiting step in the anaerobic digestion of sewage sludge.

Therefore the principal objective of the aerobic stage of dual digestion is to ensure

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[2] Thermal pretreatment of sludge at temperatures ranging from 135 to 180°C has been shown to enhance the anaerobic treatment of certain sludges (Pinnekamp (1989)). However the relative benefits of thermal pretreatment and biological pretreatment by oxygenation are not clearly understood.

thermophilic conditions, preferably under conditions of oxygen limitation. Because VS removal is not important, the process may be operated at the shortest possible retention time - subject, naturally, to the time constraints imposed by disinfection. As for the ATAD process, the rate of biological heat generation is the most important parameter in defining the minimum hydraulic retention time in the first stage of dual digestion.

2.6.2 Biological heat generation in the aerobic stage of dual digestion

The required heat for this aerobic stage may be derived from combinations of the same heat sources available to thermophilic aerobic digestion. These are:

(1) The mechanical action of a pump or a mixer.
(2) Submerged combustion of a fuel gas such as anaerobically produced methane.
(3) The injection of steam.
(4) Autothermally ie the generation of biological heat within the digestion process through oxygenation with pure oxygen or air.

Although heating the aerobic stage in dual digestion can be achieved through (2) and (3) above because methane is available from the anaerobic digestor, biologically generated heat is considered the principal heat source in this investigation.

As described earlier, the rate of biological heat generation in ATAD has been described via the rate of VS removal. While the use of VS removal rate is subject to the deficiencies listed above (in Section 2.5), it has the advantage of describing both biological heat generation and the rate of sludge stabilization, the principal objective of the ATAD process. However this advantage does not apply to the aerobic stage of dual digestion where sludge stabilization is not important. Furthermore, the possibility of oxygen limitation, and heat generation in the absence of VS removal, in the aerobic stage of dual digestion is sufficient reason for not using VS removal kinetics in the design and modelling of this stage. Hence an alternative to VS removal kinetics needs to be found.

Such an alternative might be COD removal kinetics - certainly the theoretical and empirical data presented earlier (in Section 2.4.4.2.) suggests that biological heat generation stoichiometry in the aerobic stage of dual digestion would be adequately described in terms of the removal of COD from the sludge. However, there are practical considerations which mitigate against this approach:

(1) The COD test is imprecise and the small COD removals likely to occur in the aerobic stage will be subject to large errors – this will be clearly demonstrated in this investigation.

(2) Oxygen limitation in the aerobic stage will cause biological heat generation and COD removal rates to be controlled by oxygen supply (and dissolution) rates rather than by biological kinetic rates.

A better approach might be to describe the kinetics of biological heat generation in terms of oxygen utilization. Under aerobic conditions COD removal and oxygen utilization are equivalent and if the stoichiometry of biological heat generation is expressed in terms of oxygen utilization then the kinetics of biological heat generation can be defined by the oxygen utilization rate. This approach retains the theoretical advantages of the COD measurement and it overcomes the above disadvantages because:

(1) Oxygen utilization rates can be more precisely and more quickly measured than COD removal rates.

(2) The kinetics of biological heat generation are controlled by oxygen utilization rates under both oxygen sufficient and oxygen limiting conditions.

The oxygen utilization approach to biological heat generation not only has the above practical operational advantages but also, from a design point of view, it provides a direct measure of the oxygen supply rate which will enable biological heat to be generated at the rate required to achieve thermophilic temperatures.

2.7 BIOLOGICAL HEAT GENERATION STOICHIOMETRY AND KINETICS AS A FUNCTION OF OXYGEN UTILIZATION

A theoretical relationship between biological heat generation and oxygen utilization can be derived from fundamental bio-energetics. Also experimental measurements relating biological heat generation to oxygen utilization have been made in the fields of human physiology, microbiology and operating thermophilic aerobic digestors. These sources of information on the link between biological heat generation and oxygen utilization are discussed below.

2.7.1 Predicting biological heat generation using bio–energetics

By considering the ATP/ADP energy capture and transport system, it is possible to derive equations which quantify biological heat generation in terms of oxygen utilization in an aerobic microbial culture. The postulate which is fundamental to this derivation has been stated earlier in Section 2.4.1. viz that the energy released by biological oxidation in an aerobic culture has only 2 destinations:

(1) Capture in synthesized chemical bonds e.g. growth of new biomass.
(2) Waste heat.

By calculating the energy released from a biological oxidation reaction and by determining the amount of energy captured in (1), it is possible to calculate the amount of waste heat liberated in (2) by difference and to express this waste heat in oxygen equivalents ie as the specific heat yield (MJ/kgO). This derivation is subject to certain assumptions which will be defined at the end of this section.

2.7.1.1 *Discussion*

In an aerobic heterotrophic culture energy is released for use by micro–organisms by the oxidation of organic substrates and the transfer of electrons to oxygen. The micro–organisms are not able to directly utilize this energy and therefore require the use of an energy capture and transport system. The system used comprises 2 principal chemical agents, adenosine diphosphate (ADP) and adenosine triphosphate (ATP).

Energy capture is effected by ADP and energy transport by ATP. A portion of the energy released by the oxidation of substrate is employed to attach a phosphoryl group to a molecule of ADP which then forms a molecule of ATP. The energy not captured into the molecule of ATP due to inefficient energy capture is lost as waste heat. When energy is required by a cell function, the ATP is converted back to ADP, and the captured energy is released when the phosphoryl bond is broken. A portion of this energy is consumed by the cell function while the remainder is lost as waste heat due to inefficient energy utilization. It is possible to determine the amount of waste heat lost in these two energy transactions, and thereby also the specific heat yield, by (1) defining the amount of energy consumed by a cell function, (2) calculating the number of phosphoryl bonds which are broken to provide this amount of energy and (3) calculating the amount of substrate which must be oxidized in order to provide this number of phosphoryl bonds.

2.7.1.2 *Derivation of a specific heat yield equation*

The specific heat yield equation is derived by considering that an amount of energy, A, is required to effect a cell function. To accomplish the derivation 6 variables are defined:

(1) A (MJ) – the amount of energy required by a cell function. This function may include modifying organic molecules for inclusion into protoplasm, motility and the maintenance of membrane potentials and cellular integrity. Note that if A MJ of energy is consumed for purposes of biomass synthesis, then it will be incorporated into the biomass structure as chemical bond energy. However, when it is consumed by a function such as motility, then it will ultimately revert to waste heat.

(2) E – the efficiency with which the amount of energy, A, is provided from the conversion of ATP to ADP. Thus where A MJ of energy is consumed, A/E MJ of energy must be provided by converting ATP to ADP. This efficiency, E, can only be estimated because it varies in biological systems.

(3) B (MJ/kmol) – the phosphoryl bond energy. This is the amount of energy released by the conversion of 1 kmol of ATP to 1 kmol of ADP or captured in the conversion of 1 kmol of ADP to 1 kmol of ATP. A typical *in vitro* value for B of 50 MJ/kmol will be used in this derivation.

(4) G_{ox} (MJ/kmol) – the amount of energy released by the oxidation of 1 kmol of substrate.

(5) z (kg O_2/kmol substrate oxidized) – the mass of oxygen required to oxidize 1 kmol of substrate. Values of z for a number of organic compounds are given in Table 2.1.

(6) N_{ATP} – the kmol of ATP generated by the oxidation of 1 kmol of substrate. Values of N_{ATP} for a number of organic compounds are given in Table 2.1.

Note from the above definitions that whereas the efficiency of energy utilization is defined by the parameter E, the efficiency of energy capture is defined by the parameters B and N_{ATP}. For example, the oxidation of 1 kmol of glucose releases 2874 MJ and converts 38 kmol of ADP to ATP. Given that the conversion of 1 kmol of ADP to ATP consumes 50 MJ of energy, the 38 kmol of ATP formed by the oxidation of 1 kmol of substrate represents 1900 MJ of energy ie a 66% energy capture.

With the aid of the above definitions, the specific heat yield equation is derived as follows:

If the amount of energy consumed by a cell function is equal to A MJ, then, due to inefficient utilization of energy, the amount of energy that must be provided to the cell is equal to:

$$\frac{A}{E} \qquad \text{(MJ)}$$

The kmol of phosphoryl bonds that must be broken to yield this energy is equal to:

$$\frac{A}{E \times B} \qquad \text{(kmol ATP)}$$

The amount of substrate oxidized to produce A/(E.B) kmol of phosphoryl bonds is equal to:

$$\frac{A}{E \times B \times N_{ATP}} \text{ (kmol substrate)}$$

The total amount of energy released during this oxidation is equal to:

$$\frac{A \times G_{ox}}{E \times B \times N_{ATP}} \qquad \text{(MJ)}$$

and the stoichiometric oxygen requirement is:

$$\frac{A \times z}{E \times B \times N_{ATP}} \quad (\text{kg } O_2)$$

Of the total amount of energy released only A MJ is used. If it is captured as cellular matter then the heat released in effecting the transaction for which A MJ was required is equal to:

$$\frac{A \times G_{ox}}{E \times B \times N_{ATP}} - A \quad (\text{MJ})$$

Defining the specific heat yield Y_H as the heat lost in the energy transaction divided by the oxygen consumed, then,

$$Y_H = \frac{\left[\dfrac{A \times G_{ox}}{E \times B \times N_{ATP}} - A\right]}{\left[\dfrac{A \times z}{E \times B \times N_{ATP}}\right]}$$

$$= \frac{1}{z} \times (G_{ox} - E \times B \times N_{ATP}) \quad (\text{MJ/kg } O_2) \qquad (2.11)$$

For Eq (2.11) to apply to a biological system such as the aerobic stage of a dual digestion system, specific conditions have to apply. These are:

(i) In the derivation of Eq (2.11) only the ADP/ATP energy transport system was considered. Hence, if Eq (2.11) is to calculate Y_H accurately, energy transactions in aerobic digestion other than those mediated by this ADP/ATP system must be negligible.

(ii) The waste heat released by biological activity arises from the inefficiencies in energy capture and utilization in the manner outlined above and is directly linked to the transfer of electrons to oxygen. Heat generation without the transfer of electrons to oxygen may occur when organic substrate is modified for inclusion into protoplasm. The modification of substrate for inclusion into protoplasm involves the formation of an intermediary or precursor and energy may be either released or consumed in the conversion of the substrate to the intermediary. This depends on whether an energy excess or energy deficit substrate is involved (Babel and Müller, 1985). In the case of an energy excess substrate, heat would be released which was not linked to the transfer of electrons to oxygen. Sewage sludge will probably be an energy balanced substrate in aerobic digestion processes especially where cryptic growth occurs (Hamer and Zwiefelhofer, 1986). Hence it is reasonable to assume that during aerobic digestion, not much release of heat takes place without the transfer of electrons to oxygen.

TABLE 2.1: PHYSICAL AND THERMODYNAMIC PROPERTIES OF SOME ORGANIC COMPOUNDS.

SUBSTRATE	G_{ox} kcal/kg	G_{ox} MJ/kmol	N_{ATP}	z	Y_H MJ/kg E=0.15	Y_H MJ/kg E=0
GLUCOSE	-3.81	-2874	38	192	13.5	15.0
FUMARIC ACID	-2.87	-1393	18	96	13.1	14.5
PYRUVIC ACID	-3.20	-1180	15	80	13.3	14.7
BUTYRIC ACID	-5.84	-2151	28	160	12.1	13.4
GLUTAMIC ACID	-3.22	-1987	26	144	12.4	13.8
LACTIC ACID	-3.78	-1423	18	96	13.4	14.8
SUCCINIC ACID	-3.13	-1544	20	112	12.4	13.8
LACTOSE	-4.03	-5778	76	384	13.5	15.0
GLYCEROL	-4.42	-1703	22	112	13.7	15.2

2.7.1.3 *Predicted values of Y_H*

With the aid of Eq (2.11), the specific heat yield Y_H was calculated for 9 widely differing organic compounds and listed in Table 2.1. In Table 2.1, 2 sets of values for Y_H, in units of MJ/kg O_2, are listed, for different energy efficiencies, i.e. 0.15 and 0.0. This efficiency, E, should not be seen only as a fraction of energy which is incorporated in biomass, but rather as a net efficiency of utilization which includes losses due to microbial maintenance and motility. At high temperatures it is probable that maintenance requirements are large (Marr *et al.*, 1963) especially where high concentrations of extracellular enzymes are required to effect the degradation of particulate substrate, and where the organism itself might be subject to the action of some of these enzymes.

The maximum specific heat yield predicted by Eq (2.11) occurs when E = 0 and is simply the oxidation or combustion energy of the substrate expressed in stoichiometric oxygen equivalents. Equation (2.11) will underestimate Y_H if exothermic reactions occur in the sludge without the transfer of electrons to oxygen. In an aerobic digestion process Y_H should always be less than the maximum at E = 0 because E = 0 implies that no energy is conserved in the growth of biomass. Accepting the efficiency E is 0.15 i.e. 85% of the oxidation energy in the substrate is lost as heat and 15% is captured and retained as organism mass, gives specific heat yield values for the 9 substrates ranging from 12.1 to 13.7 MJ/kgO.

2.7.2 Measurements of biological heat generation and oxygen consumption in human physiology

The literature contains a number of measurements of Y_H. The field of human physiology provides some of the earliest of these measurements by way of direct calorimetry where energy expenditure by the human body was measured as heat and related to oxygen consumption and carbon dioxide production. Measurements were initially performed by Atwater and Benedict (1903) and Benedict and Milner (1907). In indirect calorimetry, the known proportionality between oxygen consumption, carbon dioxide production and energy generation was used to measure energy expenditure. Agreement between direct and indirect calorimetry was shown by Gephart and Du Bois (1915). Typical physiological values of Y_H from these direct and indirect measurements range from 13.8 to 16.1 MJ/kgO consumed. Consolazio *et al.* (1963) recommend an average value of 14.6 MJ/kgO for general human physiological energy expenditure calculations.

2.7.3 Measurements of biological heat generation and oxygen consumption in microbiology

After showing that heat output from slipe wool was the result of microbial activity (Rothbaum, 1961), in 1961 Rothbaum and Stone presented measured heat outputs of, and the concomitant consumption of oxygen by, *Escherichia Coli* using 10 different concentrations of a mixed substrate. Their results gave specific heat yields ranging from 16.3 to 26.7 MJ/kgO. Cooney, Wang and Mateles (1968) measured the heat output, as functions of oxygen consumption and carbon dioxide production, of yeast, bacteria and a mold, using glucose, molasses and soy bean meal as substrate. Eight fermentations with combinations of these substrate and biomass were performed and results were reported as individual tests during each fermentation and as 8 cumulative values, one for each fermentation. The cumulative values give specific heat yields ranging from 11.4 to 18.6 MJ/kgO, with an average of 14.4 ± 1.31 MJ/kgO.

The high values obtained by Rothbaum and Stone are difficult to explain as they exceed the bioenergetic maxima predicted by Eq. 2.11. Similarly, of Cooney's data, only the slope of total heat generation against oxygen consumption approximates the predicted values in Table 2.1. The wide scatter in the measured values data suggests that experimental error might have been significant because from Table 2.1 it is reasonable to expect the Y_H values to fall within a fairly narrow range. Given the range of theoretical Y_H values in Table 2.1 it is unlikely that energy excess substrates could have caused such high values of Y_H as measured by Rothbaum & Stone.

2.7.4 Measurements of biological heat generation and oxygen consumption in thermophilic aerobic digestion

With regard to thermophilic aerobic digestion, very little experimental information on Y_H is available. The heat balance data of Wolinski (1985) suggests a value of 21 MJ/kgO. Calculations from the work of Trim (1984) gives values of 9.53 (3.2 day retention), 5.3 (3 day retention), 6.1 (1.8 day retention), and 11.7 MJ/kgO (1 day retention). In neither of the two investigations was the objective specifically aimed at determining Y_H and consequently the values of Y_H quoted above were calculated from their average data. The large variation in Y_H values between these two investigations seems to suggest that large errors can readily be made in measuring the Y_H value from heat and oxygen mass balances around the aerobic reactor. Clearly, if Y_H is to be accurately measured on the aerobic reactor of a dual digestion system, great caution and care needs to be taken in setting up and measuring the heat and oxygen mass balances; if this is not done, then from the experience of the two investigations cited above, large variations and uncertainty in the measured Y_H value will arise.

2.8 CONCLUSIONS

The biological heat generation kinetics defines the minimum retention time at which the aerobic stage of a dual digestion system can be operated if a specified thermophilic temperature, T_{se}, is to be maintained in this stage. The stoichiometry and kinetics of biological heat generation in autothermal thermophilic aerobic digestion, a process superficially similar to the aerobic stage of dual digestion, have been related to the rate of VS removal in the digestor sludge, in spite of shortcomings in this approach. However, relating biological heat generation and VS removal in the aerobic stage of dual digestion is inappropriate, primarily because of possible oxygen limitation in this stage and because the stage is not required to remove volatile solids.

Theoretical and empirical evidence suggests that COD removal from an aerobic sludge is stoichiometrically related to biological heat generation in the sludge. Thus for the aerobic stage of dual digestion it is proposed that this stoichiometry may be defined in terms of the amount of oxygen consumed by a sludge, and that the kinetics of biological heat generation will be defined by the oxygen utilization rate of the sludge.

Accordingly, the objectives of the investigation reported in this thesis are:

(1) To determine the stoichiometric relationships between biological heat generation and oxygen consumption in the aerobic stage of a full scale dual digestion system treating sewage sludge.

(2) To show that the rate of biological heat generation in this aerobic stage is determined by the oxygen utilization rate (OUR) of the sludge and that under oxygen sufficient conditions, this OUR defines the minimum stage hydraulic retention time.

(3) To examine VS and COD removal in the aerobic stage of the dual digestion system and to determine whether these 2 parameters are related to biological heat generation in the stage.

(4) To develop procedures which may be used both in the design of the aerobic stage of dual digestion and in the dynamic simulation of the change in sludge temperature with time in this stage.

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CHAPTER 3

OPERATION AND MEASUREMENT METHODS FOR EVALUATION OF AEROBIC REACTOR PERFORMANCE

ABSTRACT

In this chapter the following are described:

(1) The layout, equipment and operating procedures of the Potsdam dual digestion system.
(2) The equipment, procedures and calculations employed to construct and solve simultaneous oxygen mass and unsteady state heat balances, and VS and COD balances, across the aerobic stage of the dual digestion system.
(3) A sample solution of a simultaneous oxygen mass and unsteady state heat balance in order to calculate a specific heat yield value.
(4) The procedures and calculations whereby equipment in (2) was calibrated to ensure accuracy in data measurement.

3.1 THE POTSDAM WASTEWATER TREATMENT WORKS

The Potsdam Wastewater Treatment Works falls within the greater Cape Town area and is operated by the Municipality of Milnerton. The works serves a population of 80,000 and treats an average dry weather flow (ADWF) of 18 megalitres per day. Sludge production is 150 m^3/d at about 3.5% total settleable solids concentration. Treatment of sewage is currently achieved by screening, grit removal, primary sedimentation, biofiltration with humus tanks and maturation. A layout of the Potsdam works is given in Fig 3.1.

Eight biofilters discharge to 8 humus tanks, the underflow from these tanks returning to the head of works. Both the humus and the primary sludges are removed in 7 primary sedimentation tanks (PST). Desludging of the PST's is manually controlled: At approximately 5h intervals the accumulated sludge is discharged to a central sludge collection sump; a desludging operation is stopped when, in the estimation of the operator, the underflow concentration has decreased perceptibly. The contents of the central sludge collection sump are pumped to the

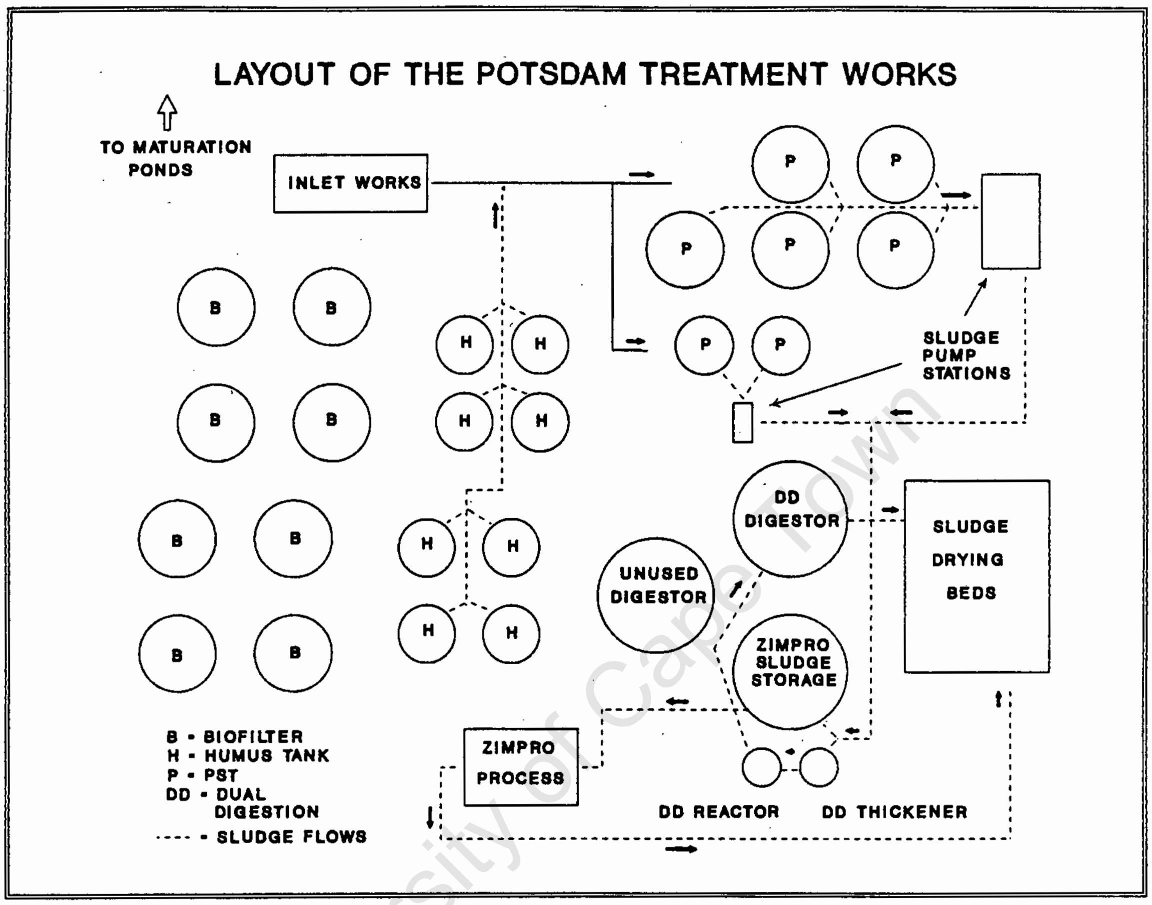

Fig 3.1: A schematic layout of the Potsdam waste water treatment works showing the integration of the dual digestion research plant with the treatment facility. Note that only sludge flows are depicted in the illustration - flows of sewage are not shown.

sludge treatment facilities about 4 times daily.

At Milnerton, sludge originally was treated by cold anaerobic digestion in three 1000 m^3 concrete digestors. In 1976 these digestors were decommissioned and a Zimpro treatment facility was installed to treat the sludge, a facility still in operation today (1991). With the Zimpro process, sludge is pumped from the central collection sump to one of the decommissioned anaerobic digestors for sludge storage and thickening before discharge to the Zimpro process.

The Zimpro process treated all the sludge produced until 1987, when the experimental dual digestion plant was put into operation. To give an impression of the scale of the experimental dual digestion plant, it treated about 1/4 of the daily sludge production, i.e. about 36 m^3/d or 1250 kg dry solids/d.

3.2 DUAL DIGESTION PLANT LAYOUT AND OPERATION

A schematic layout of the Milnerton dual digestion plant is shown in Fig 3.2. The plant comprises 3 tanks in series, i.e. a sludge thickener, an aerobic stage and an anaerobic stage. Throughout this thesis, the aerobic stage of the Milnerton dual digestion system will be termed 'the reactor' and the anaerobic stage, 'the digestor'.

The thickener and reactor are identical 50 m^3 fibre glass tanks which differ only in regard to their piping configurations. The thickener has 3 draw-off points for the draining of supernatant during gravity thickening of the sludge. These are located at 2.1, 3.1 and 4.1 metres from the bottom point of the thickener cone. Sludge feed for the thickener was pumped from the central collection sump. Supernatant from the thickener was returned to the head of works.

The resin used in the construction of the thickener and reactor is sensitive to temperature and the manufacturer's guarantee is invalidated if sludge temperatures rise above 68°C in the aerobic reactor. Accordingly 68°C was set as the maximum allowable reactor sludge temperature. To reduce heat losses, the reactor was lagged with 50mm thick polyurethane foam. The operating volume of the reactor was set at 45 m^3; mixing and oxygenation were by means of a Vitox pure oxygen injection system.

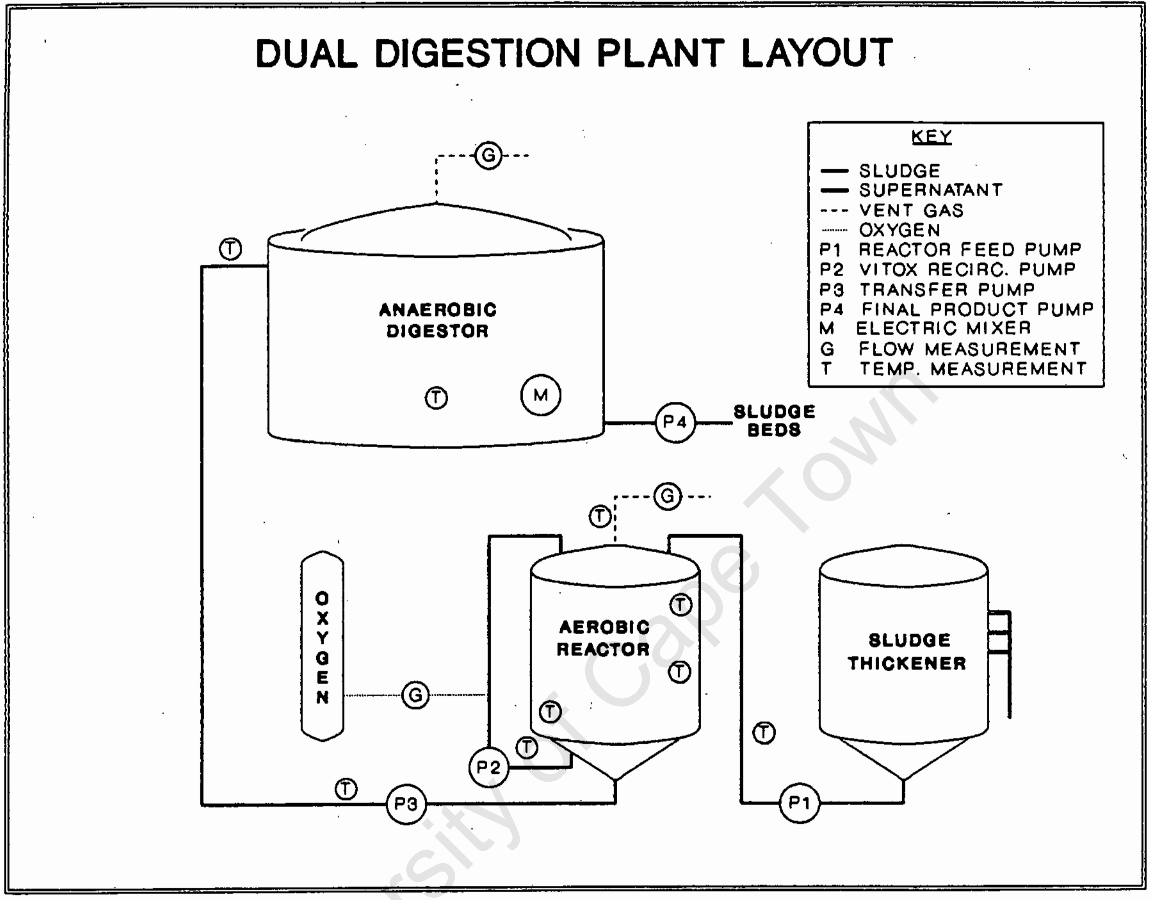

Fig 3.2: A schematic layout of the dual digestion plant at Potsdam showing the sludge thickener, the aerobic reactor and the anaerobic digestor, pumps, and positions of temperature and gas flow measurement.

The reactor was operated on a draw and fill basis to avoid contamination of the effluent pasteurized sludge with feed sludge. Draw and fill operation also allowed the sludge flow rate to be monitored accurately. A centrifugal pump first transferred a fixed batch volume of hot, pasteurized sludge from the aerobic reactor to the anaerobic digestor and then a macerator feed pump refilled the reactor to its operating volume with thickened (to ≈ 3.5%) feed sludge drawn from the base of the sludge thickener. The transfer and feeding operations were completed in about 3 and 4 min respectively. The feed to the sludge thickener was a mixture of primary and humus sludge drawn from the central sludge collection sump.

The draw and fill batch volume was fixed at 1/12 of the reactor volume i.e. 3.75 m^3 and the time interval between feeding was varied between 2 and 6h to achieve retention times varying from 1.2 to 3 days. The minimum batch cycle time interval of 2h was set by the time constraints imposed by sludge disinfection[1], and this gave the minimum hydraulic retention time of 1.0 day. Vent gas temperature was measured as it emerged from the reactor, and specialized equipment was constructed to measure the vent gas composition, water vapour content and volumetric flow rate. This comprehensive monitoring of the vent gas allowed calculation of the oxygen consumption rate, the oxygenation efficiency, the degree of vent gas water vapour saturation and the respiration quotient i.e. moles carbon dioxide produced per mole of oxygen consumed.

The anaerobic digestor was one of the decommissioned cold anaerobic digestors. It is a domed concrete structure of 1000 m^3. As the anaerobic stage of the dual digestion plant, the operating volume of the digestor ranged between 400 and 600 m^3 depending on the anaerobic retention time required. Because of the batch operation of the reactor, the digestor also was batch fed, receiving 3.75 m^3 of hot transfer sludge from the reactor every 2 to 6h, depending on the reactor retention time. No external heating of the digestor was provided. The hot (±60°C) feed sludge from the reactor was the only heat source to maintain the digestor at mesophilic temperatures (±37°C). In winter the heat loss from the digestor was sufficient to prevent the digestor from overheating but during the hot summer

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[1] The time and temperature constraints imposed by disinfection have been discussed in Chapter 2, Section 2.6.1.1. In this investigation it was accepted that pasteurization would be achieved by exposing the sludge to a temperature greater than 55°C for a minimum of 2h.

months, the digestor heat loss was insufficient and overheating (44°C) occurred under normal operation. Because no feed sludge cooler or heat exchanger was available, the digestor temperature was controlled by diverting batches of hot sludge from the reactor to the Zimpro feed sludge storage tank.

To maintain the required digestor operating volume, sludge was discharged once daily from the digestor in a single batch. The volume discharged was monitored via a 6m high external sight-glass fitted to the digestor for sludge liquid level control. Gas production was collected, cooled to condense moisture, reheated to reduce the relative humidity of the gas and passed through a Parkinson Cowan positive displacement gas meter to measure its volume. Reheating was a necessary precaution to prevent moisture condensing in the gas meter.

The above discussion briefly describes the operation of the thickener, aerobic reactor and anaerobic digestor comprising the dual digestion plant. This investigation focused on the performance of the aerobic reactor, and, consequently, aside from a brief description of the performance of the anaerobic digestor in Appendix 5B, no further details on digestor operation and performance will be given. A more detailed description of the reactor operation and monitoring is given below.

3.3 AEROBIC REACTOR OPERATION AND MONITORING

3.3.1 Sludge collection and thickening

Sludge for the dual digestion system was pumped from the central sludge collection sump and discharged to the dual digestion sludge thickener. At regular intervals, dual digestion operators decanted clear supernatant through one of the 3 drain ports on the thickener. At times dilute sludge was wasted in the same manner in an attempt to ensure the most concentrated feed sludge possible for the reactor. Long retention times in the Milnerton sewerage reticulation system resulted in some sludge septicity, and the degree of consolidation achieved in the thickener at times appeared to be compromised by the generation of anaerobic gases by the sludge. On a few occasions during very hot weather, the sludge was buoyed to the surface resulting in complete sludge/supernatant inversion which necessitated the withdrawal of supernatant from the base of the thickener to achieve thickening. The concentration of a composite sample of thickened sludge was monitored daily. The total feed solids concentration achieved during the period covered by this investigation (November 1987 – November 1988) ranged from 27.8 to 57.9 g/ℓ with an average of 37.5 ± 4.4 g/ℓ.

3.3.2 Sludge feeding and level control

Accurate sludge level monitoring inside the reactor was essential to control sludge levels during the batch draw and fill operation. The level control system is illustrated in Fig 3.3.

It comprised a stainless steel float, which rested on the sludge liquid surface, and to which was connected a thin rigid rod with a plastic bottle containing a magnet fastened to its top end. The magnet would activate reed switches fixed to a rule at measured points. The reed switch at the high level corresponded to the 45 m^3 operating volume and at the low level to 41.25 m^3, the difference between the two levels being the 3.75 m^3 batch volume.

At intervals of not less than 2h, a quartz pulse timer activated the transfer pump and sludge was pumped from the reactor to the digestor causing the stainless steel float to drop until the low level reed switch was activated by the magnet. This terminated the transfer operation with exactly 3.75 m^3 having been transferred. Immediately after termination of sludge transfer to the digestor, sludge feeding from the thickener would commence. This feeding was terminated by the high level reed switch when the operating volume was once again 45 m^3. The pulse timer merely initiated this transfer/feeding cycle and after initiation, the level control system controlled the sludge transfer and feed operations. Additional reed switches were attached to the rule above and below the high and low levels. These were alarm switches which would be activated in the event of a failure of the high or low level switches.

A 1.5mm stainless steel cable was connected to the plastic bottle containing the magnet at the top of the vertical stainless steel rod. The cable was looped over a pulley wheel fitted onto the shaft of a potentiometer, and was connected to a counterweight on its other end. Vertical movement of the stainless steel rod rotated the pulley wheel and the potentiometer shaft. The rotation of the potentiometer changed the flow of electrical current in a circuit of which the potentiometer was a component. The change in electrical current was displayed on an ammeter, which was located in the control hut. This arrangement provided

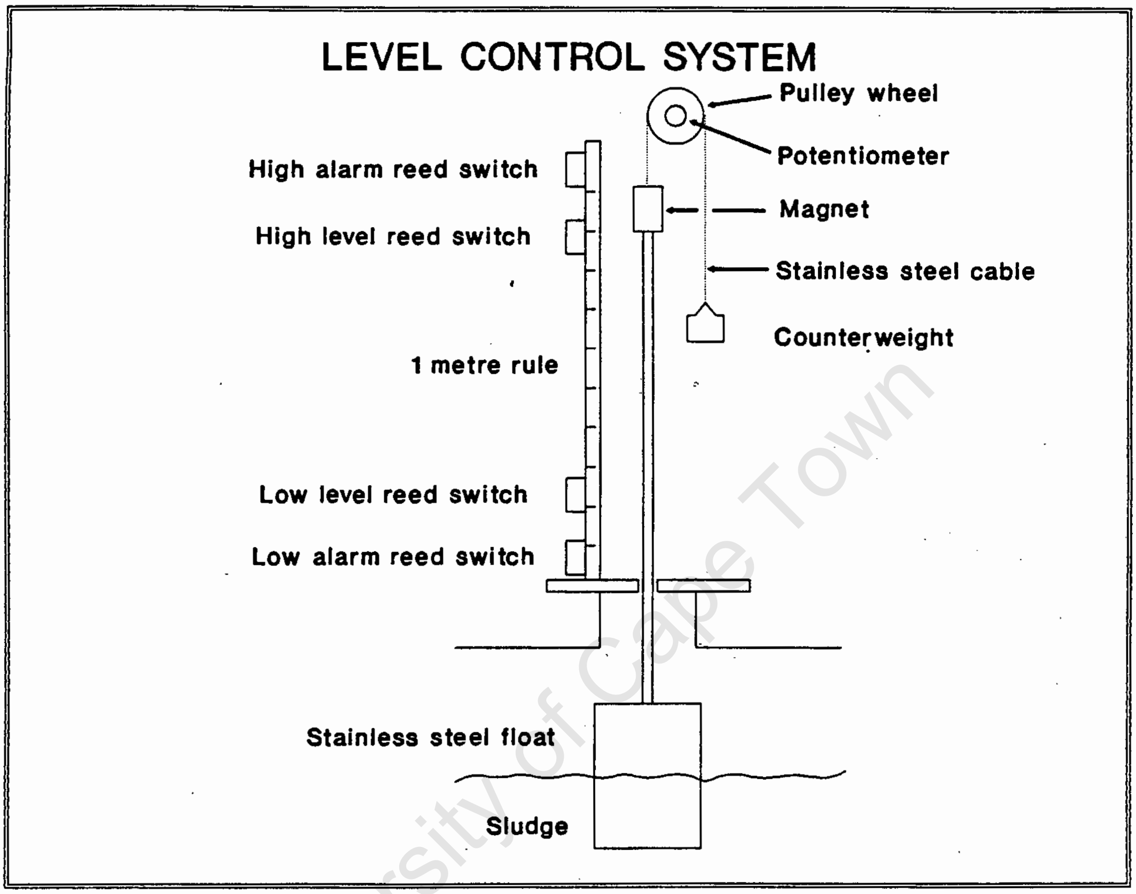

Fig 3.3: The aerobic reactor level control system which maintained a constant sludge level of $45m^3$, and which controlled the batch draw and fill system to ensure that $3.75m^3$ batches of sludge were transferred from and fed to the reactor.

a visual indication of the sludge level in the reactor.

The level control system proved to be reliable, robust, and not subject to fouling and foaming caused by material in the sludge. It operated much more reliably than more sophisticated (and more costly) systems which, early in the investigation, had proved inadequate and unreliable. More discussion of level control is given in Appendix 5B.

3.3.3 Oxygenation system

A Vitox[2] system oxygenated and mixed the aerobic reactor. The Vitox system comprised a recirculation loop through which sludge was pumped continuously. Oxygen was injected into the sludge via a venturi located in the loop. Sludge was drawn from a point on the conical base of the reactor, passed through the recirculation loop by a torque flow centrifugal pump (see Appendix 5B for more detail on this pump) and returned to the reactor through a flange located on the top of the reactor. The recirculated sludge discharged into the reactor, below the sludge surface, through a single nozzle[3] offset axially and radially in the reactor to achieve effective mixing. The recirculation flow was about 150 m^3/h giving a reactor contents turnover time of about 20 min.

Although the efficiency of the torque flow centrifugal pump was only 40-50%, it was not prone to blockage by rags and similar materials present in the sludge. The relatively high power consumption by the pump (±20 kW) was not wasted because in the heat balance tests it was found that about 90% of this power was converted to heat and contributed directly to heating the sludge. This pump is discussed in more detail in Appendix 5B.

Oxygen at 400 kPa was injected into the venturi located in the recirculation loop downstream of the pump. The oxygen partially dissolved in a tortuous path (four 180°C bends) in the recirculation loop prior to discharge into the reactor where dissolution was complete. The oxygen supply was stored as liquid in a 6000ℓ vacuum insulated vessel. The oxygen supply to the reactor was vaporized to ambient temperature and the rate of supply to the reactor was controlled with two calibrated rotameters, each fitted with a needle valve. Either the low flow

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[2] Patented by the British Oxygen Company (BOC).

[3] Patented by African Oxygen Limited.

rotameter (6 – 20 kg/h) or the high flow rotameter (10 – 55 kg/h) was used depending on the required oxygen supply rate.

Both the recirculation pump and oxygen supply operated continuously, with the oxygen supply rate fixed for any specific 'steady state' condition. Control on the basis of dissolved oxygen concentration in the reactor was not employed because in previous research (Trim, 1984) continuous oxygen supply at a certain specified rate was found more reliable than dissolved oxygen control.

3.3.4 Temperature measurement

3.3.4.1 *Equipment and monitoring points*

Ten PT100 resistance thermometers were positioned around the dual digestion system to measure temperatures. The locations of the monitoring points are given in Fig 3.2 i.e.

3 in the reactor
1 in the recirculation loop
1 in the feed sludge
1 in the reactor vent gas
1 in the digestor feed sludge immediately after transfer from the reactor
1 in the digestor feed sludge immediately before discharge to the digestor
1 in the digestor
1 in the ambient air

Temperatures were displayed to 1 decimal place and recorded on chart paper by a 12-channel Honeywell recorder. The respective temperature traces from the 10 thermometers were identified on the chart by labels and by the use of 6 different pen colours and 2 printing styles.

The recorder was equipped with 2 alarm contacts. One of these was set to prevent the reactor temperature exceeding the limit of 68°C. When the temperature in the reactor exceeded 68°C the alarm actuated a control that switched the oxygen supply from the high to the low flow oxygen supply rotameter.

3.3.4.2 *Batch cycle temperature profile*

The draw and fill batch feed operation produced a distinctive saw tooth profile for each of the 4 measured reactor sludge temperatures (i.e. 3 in the reactor and one in the recirculation loop – see Fig 3.2). A typical profile for a single temperature probe is illustrated in Fig 3.4. There are 3 phases in a batch cycle viz. (1) the transfer phase, (2) the feeding phase and (3) the heating phase. From the temperature measurements, sludge transfer from the reactor (phase 1) had no discernible effect on the temperature profile and the heating rate during this phase appeared to be the same as that during the previous phase[4] i.e. the heating phase 3. Feeding cold sludge (phase 2) caused a sharp drop of ±3° C in temperature as it mixed in with the hot reactor contents. After the cessation of feeding, phase 3 commenced and during this phase the temperature rose linearly at constant oxygen supply rate until the new batch cycle commenced.

3.3.4.3 *Sludge temperatures and reactor mixing*

As mentioned above, the reactor sludge temperature was monitored at 4 different points (Fig 3.2). The rapidity, and similarity, of the 4 individual thermometer responses[5] to the sudden temperature changes that took place during feeding (phase 2, Fig 3.4), were used as indicators of the quality of reactor mixing. When the cold feed sludge was rapidly distributed into the hot reactor contents then the mixing was good and the temperature probes at different reactor positions showed an equally rapid and similar response to the sudden temperature decrease during feeding. On the other hand, when the mixing was poor, then the cold feed sludge was not mixed into the hot sludge at a rapid and uniform rate and the temperature probes showed different rates of response and gave dissimilar temperature readings. Temperature profiles in Fig 3.5 illustrate ideal and non-ideal mixing. The conditions for good mixing in the reactor were found by trial and error positioning of the recirculation flow discharge nozzle in the reactor contents. Reactor mixing is discussed more thoroughly in Appendix 5B.

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[4] In order to illustrate the 3 phases in Fig.3.4 more clearly, the durations of phases 1 and 2 are shown to be longer than they actually were at Milnerton. In reality these phases were short (about 3 min each) compared to phase 3 (2 to 6h).

[5] The response characteristic of a single probe was measured and was found to give reliable readings if the rate of temperature change was less than 2° C/min. The rate of temperature change at Milnerton was always less than 1° C/min.

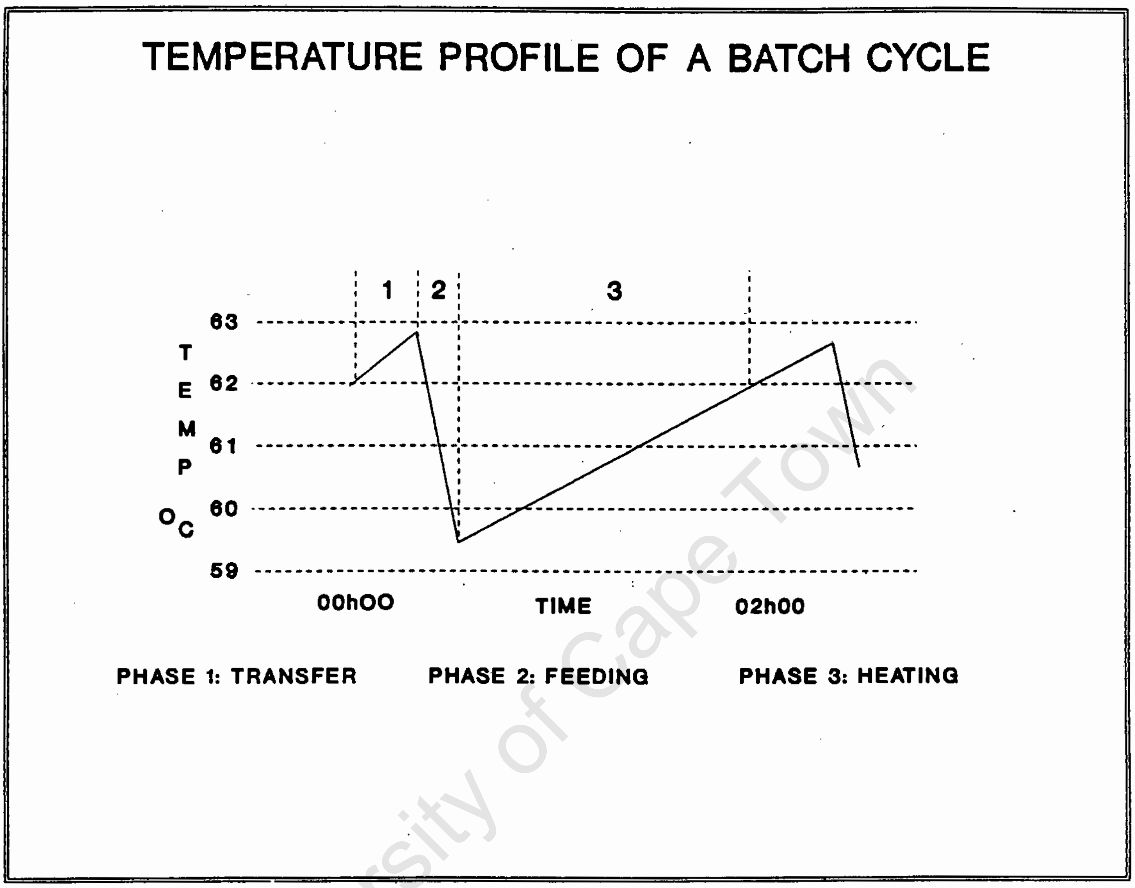
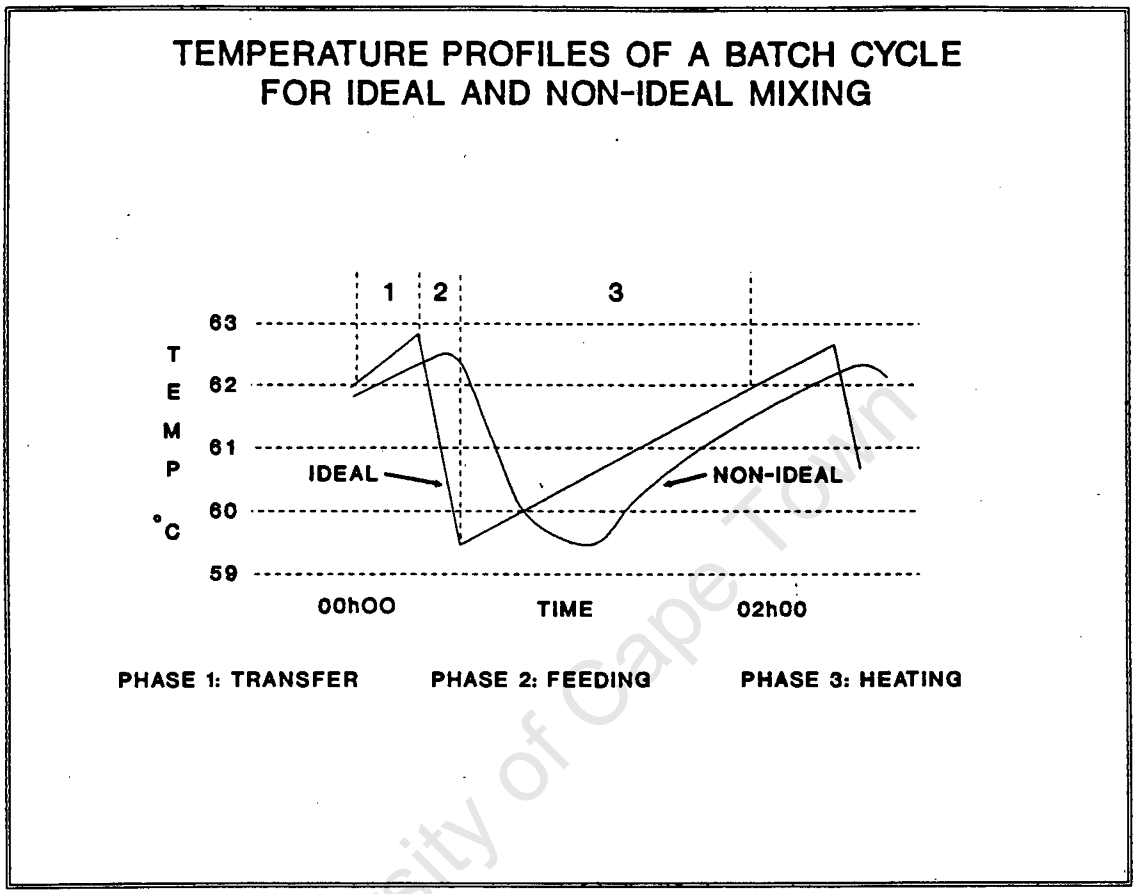


Fig 3.4: A typical profile of the reactor sludge temperature (T_{se}) against time for the 3 phases of a complete batch cycle.



<u>Fig 3.5</u>: The profiles of reactor sludge temperature (T_{se}) against time, illustrating ideal and non-ideal mixing via the responses of 2 temperature probes. For ideal mixing the response of the probe to change in sludge temperature is rapid and the points of change are clearly defined. For non-ideal mixing the probe response lags behind the changes in sludge temperature with poorly defined points of change.

3.4 OXYGEN MASS AND HEAT BALANCES OVER THE REACTOR

Because the primary objective of this thesis was to determine whether biological heat generation in the aerobic stage was proportional to the oxygen consumption, simultaneous oxygen mass and heat balances around the reactor, from which the *specific heat yield* (MJ/kg O_2) could be calculated, needed to be performed. The time interval chosen for performing these balances, was the 2 to 6h period of linear heating from feed termination to transfer commencement (i.e. phase 3) shown in Fig 3.4. The reasons for selecting this period were:

(1) During this period the sludge volume remains constant and so the vent gas volumetric flow rate and composition are not disturbed by a changing sludge volume. In contrast, during the sludge transfer phase, the sludge volume in the reactor decreases and this causes air to be sucked into the reactor head space, and during feeding, this air, mixed with the vent gas, is again expelled from the reactor. Consequently during the transfer and feeding phase, the vent gas volumetric flow rate and composition is disturbed by the changing sludge volume.

(2) The transfer and feeding phases were both of short duration (about 3 and 4 min respectively) compared to the heating phase (2 to 6h). At least 1h was required to perform the heat and mass balance measurements.

(3) At constant sludge volume, the heat balance is simplified in that the heat gained by the sludge is directly proportional to the sludge temperature increase.

All the specific heat yield tests reported in this thesis were conducted during the linear heating phase (phase 3) and as a result this phase is referred to as the specific heat yield test period.

A secondary objective of this thesis was to measure the change in COD and VS concentrations across the reactor and to determine whether these also could be related to the biological heat generation.

3.4.1 The oxygen mass balance

The oxygen mass balance enabled the oxygen consumption by the sludge in the reactor to be measured over the same time interval in which the heat balance was conducted. Influent oxygen mass flow was measured with the oxygen feed rotameters. Effluent oxygen mass flow was determined by measuring the volumetric flow rate of, and oxygen concentration in, the reactor vent gas.

In many previous research investigations into autothermal thermophilic aerobic digestion (ATAD), the oxygen consumed was estimated by assuming unity for the respiration quotient i.e. 1 mole carbon dioxide generated per mole oxygen consumed. Simple stoichiometry suggests that this is permissible e.g. $C_5H_7O_2N + 5O_2 \rightarrow 5CO_2 + 2H_2O + NH_3$. Assuming unity for the respiration quotient obviates having to measure the total dry vent gas molar flow rate - this would be equal to the influent dry gas molar flow rate. Only the oxygen fraction of the dry vent gas needs to be measured and from this the mass of oxygen consumed can be calculated. However, should the respiration quotient be less than unity, then the calculated oxygen consumed by this approach will be less than the actual oxygen consumed. In this investigation assuming unity for the respiration quotient was not considered sufficiently reliable. Consequently, because the biological reactions in the reactor were not believed to be solely sludge digestion (as the $C_5H_7O_2N$ reaction above suggests, see Chapter 2, Section 2.5) it was decided to measure the total dry vent gas flow rates. In hindsight much of the accuracy of the specific heat yield data rested on this decision because the Milnerton results showed the respiration quotient in the reactor to be 0.66, not unity (see Chapter 4, Section 4.6).

3.4.1.1 *Influent oxygen measurement*

The pure oxygen supply was stored as liquid in a vacuum insulated vessel. It was vaporized and piped to the rotameters in the Vitox oxygenation control panel, at which stage its temperature had equilibrated with the ambient surroundings.

Oxygen supply flow to the reactor was continuous and controlled with the two rotameters mentioned earlier i.e. a high flow rotameter (10 to 55 kgO/h) and a low flow rotameter (6 to 20 kgO/h). Apart from the automatic alarm control switch from the high to the low flow rotameter in the event of the reactor sludge temperature, T_{sc}, exceeding 68°C mentioned earlier (see Section 3.2), the oxygen supply rate was manually set with the aid of rotameters. For a given reactor retention time and temperature, the oxygen supply rate was set at a specified value and remained unaltered, sometimes for weeks at a time.

The oxygen supply rotameters were calibrated by timing the flow of measured masses of pure oxygen through them at a number of different supply rate settings. The calibration tests are described in Appendix 3A.

3.4.1.2 *Effluent oxygen measurement*

The decision, in this investigation, to measure the total vent gas volumetric (molar) flow rate and oxygen composition as part of the heat and oxygen mass balances around the reactor, required the total vent gas flow to be dehumidified. The high humidity of the vent gas complicates flow rate measurement because water vapour condenses in the gas meter and interferes with its operation and accuracy. To avoid water vapour condensation in the gas meter, the vent gas was partially dried in a water-cooled condenser before passing through the gas meter. The temperature and oxygen content of the partially dried vent gas were measured at the gas meter.

Not only was the water-cooled condenser useful to protect the vent gas meter, it also provided a means for measuring the water vapour condensate accumulation rate. This rate was required to calculate the vent gas water vapour heat loss rate in the heat balance. This vent gas metering system is illustrated in Fig 3.6 and its operation is described below.

The temperature of the vent gas was measured in the vent gas carrier line very close to the point of exit of the vent gas from the head space of the reactor. The temperature measurement point was positioned at the 'watershed' point in the vent gas carrier line, i.e. at the point where water condensate produced prior to the temperature measurement point would flow back into the reactor head space and that produced after the temperature measurement point would flow down stream along the vent gas carrier line and collect in the primary water trap (Fig 3.6). This procedure was necessary to obtain the correct measurements of the water vapour condensate accumulation rate to accurately determine the water vapour heat loss rate (H_{ve}). At the watershed point the vent gas carrier line bent downwards (through an angle of about 100° from the vertical) so that from the bend, any water vapour condensate formed after the bend would flow along the line of flow into the primary water trap. The temperature of the vent gas was measured at the watershed point in the carrier line and this was accepted to be the average temperature of the vent gas in the head space.

After temperature measurement at the watershed point in the gas carrier line, the vent gas passed through the primary water trap. From the primary trap, the vent gas entered a counter-current condenser which was cooled with water from one of the final maturation ponds of the treatment works. The condenser consisted of 25mm diameter thin walled copper tubing which conveyed the condensate. The thin walled copper tubing was enclosed in a jacket of 50mm diameter plastic pipe which conveyed the coolant. The cooled surface extended for a length of ±20m. A second water trap collected the condensate from the condenser. The volume of water collected in both water traps during a specific heat yield test period was measured. The gas leaving the condenser was not completely dry as it was still saturated with water vapour at the temperature of the cooling water (±20°C). The remaining water vapour in the cooled vent gas was calculated with the aid of thermodynamic equations (Equation 3.13 given below) and was added to the collected water vapour to give the total vent gas water vapour flow rate. Knowing the total vent gas water vapour flow rate allowed calculation of (1) the heat loss via water vaporization and (2) the degree of water vapour saturation of the vent gas.

To decrease the gas relative humidity and prevent any further water condensation after cooling to 20°C, the gas was reheated using a length of wire element wrapped around the copper vent gas pipe after the second water trap. The temperature, pressure, composition and volumetric flow rate of the reheated vent gas was measured.

The vent gas volumetric flow rate was measured with a totalising Parkinson Cowan positive displacement towns gas meter by dividing the volume of vent gas generated during the specific heat yield test, by the time interval of the specific heat yield test period. The accuracy of the meter was verified by checking its measured volumetric flow rate against calculated volumetric flow rates from measured masses of nitrogen gas that were passed through the meter over measured time intervals (see Appendix 3A for details). The vent gas composition was measured with an Orsat apparatus which absorbs the carbon dioxide from the gas. The remaining component of the vent gas was accepted to be all oxygen. That this remaining gas was indeed all oxygen was verified by occasionally absorbing it with pyrogallol – in a number of such tests no gas was left over after carbon dioxide and oxygen absorption. The ORSAT analysis of the vent gas

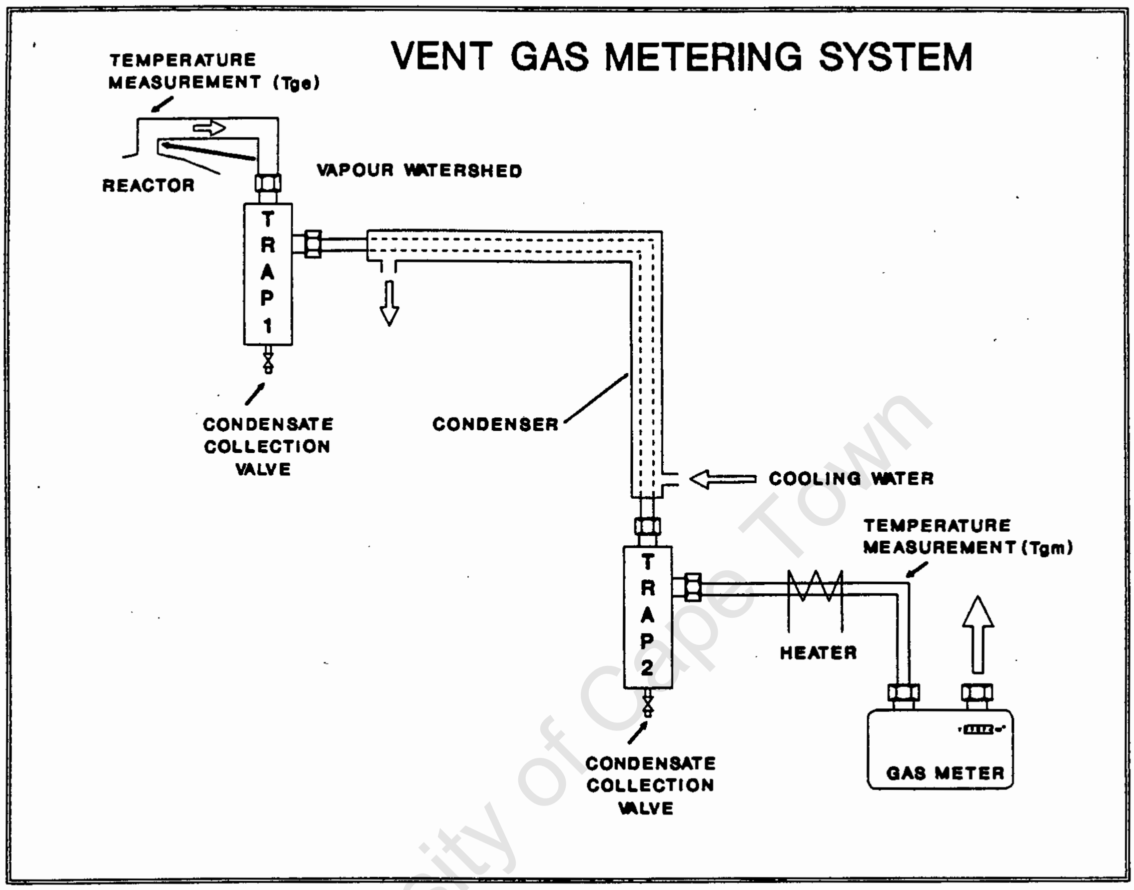


Fig 3.6: The layout of the vent gas metering system. The vent gas metering system (1) cooled the hot, moisture laden vent gas to 20°C, (2) enabled the condensate from the cooled and partially dried vent gas to be collected and (3) reheated this cooled and partially dried vent gas prior to the measurement of its volume in the gas meter.

carbon dioxide concentration was conducted at the end of the specific heat yield test period i.e. at the end of the heating phase 3. This time was selected to ensure that the vent gas sample was representative and was free of contamination by air. As has been mentioned earlier (Section 3.4 above), the transfer of sludge from the reactor during phase 1 caused air to be drawn into the reactor head space where it mixed with the vent gas produced by the sludge. Although the feeding of sludge into the reactor, during phase 2, expelled much of this air, some remained in the head space, contaminating the vent gas. By the end of phase 3, vent gas generation caused 4 to 5 displacements of the head space volume, sufficient to ensure that the contaminated vent gas had been purged. Therefore in so far as composition is concerned, a vent gas sample taken at the end of the heating phase (3) would be representative of the vent gas generated.

With the vent gas measuring equipment described above, the following data were measured during a specific heat yield test:

1. the temperature of the vent gas at the point of exit from the reactor, T_{ge}

2. the water vapour condensate accumulation rate, M_{wc}

3. the flow rate of partially dried vent gas, V_{gm}, at a temperature of T_{gm}

4. the fraction (by volume - f_{O_2}) of oxygen in the vent gas.

With the aid of this data it was possible to calculate (1) the mass of oxygen consumed by the reactor, O_c (kgO/h), (at a given oxygen feed supply rate, O_s (kgO/h), (2) the oxygenation efficiency of the Vitox system, O_{eff} (O_c/O_s), (3) the respiration quotient, Y_{CO_2} (mole CO_2 generated/mole O_2 consumed), and (4) the degree of water vapour saturation in the vent gas. The equations required for these calculations are derived in Appendix 3B and are listed below in Equations (3.1) to (3.3) for the first 3 parameters. With regard to the 4th parameter (the degree of water vapour saturation in the vent gas), this could not be calculated simply by a derived equation and required a more complex approach. The calculations for this are briefly set out in Section 3.5 of this chapter, and the approach and derivation of the equations are given in Appendix 3B.

$$O_c = O_s - \frac{380.7 \text{ x } f_{O_2} \text{ x } V_{gm}}{273 + T_{gm}} \qquad (kgO/h) \tag{3.1}$$

$$O_{eff} = \frac{O_c}{O_s} \text{ x } 100 \tag{3.2}$$

$$Y_{CO_2} = \frac{380.7 \text{ x } (1-f_{O_2}) \text{ x } V_{gm}}{O_c \text{ x } (273 + T_{gm})} \tag{3.3}$$

where

O_c = the mass oxygen consumption rate in the reactor (kgO/h)

O_s = the mass oxygen supply rate to the reactor (kgO/h)

f_{O2} = the volume fraction of oxygen measured in the vent gas passing through the Parkinson Cowan gas meter

O_{eff} = the fractional oxygenation transfer/utilization efficiency

V_{gm} = the volumetric flow rate of vent gas measured by the Parkinson Cowan gas meter (m^3/h)

T_{gm} = the temperature of the vent gas passing through the Parkinson Cowan gas meter (°C)

3.4.2 Heat balance across the reactor

Two types of heat balance can be made across the reactor, a steady state one and an unsteady state one. The steady state heat balance is generally applied to a reactor which is continuously fed – such a reactor is judged to be at steady state when the heat balance parameters, such as reactor sludge temperature, do not change over time. Under this condition, at any given moment in time, the sum of heat sources to the reactor is equal to the sum of heat sinks from the reactor. Typical heat sources are: The sensible heats of influent gas and sludge, heat generated by the mechanical action of mixing and oxygenation systems, and heat generated from biological oxidation. Typical heat sinks are: The sensible heats of effluent gas and sludge, the heat of vaporization of water, and heat conducted through the reactor walls to the ambient surroundings. When the oxygenation of the reactor sludge is accomplished with pure oxygen, then the contributions of influent and effluent gas sensible heats will be very small (typically $< 0.3\%$ of the sources and < 0.4% of the sinks) and therefore may be neglected. Therefore a typical steady state heat balance across a continuously fed reactor is given as:

$$H_{bi} + H_{mi} + H_{si} = H_{se} + H_{ve} + H_{we} \qquad (3.4)$$

where

H_{bi} = the rate of biological heat generation (MJ/h)

H_{mi} = the rate of heat generation in the reactor due to the mechanical action of mixing (MJ/h)

H_{si} = the rate of heat generation due to the sensible heat of the influent feed sludge (MJ/h)

H_{se} = the rate of heat loss due to the sensible heat of the reactor effluent sludge (MJ/h)

H_{we} = the rate of heat loss from the surfaces of the reactor and the Vitox oxygenation system (MJ/h)

H_{ve} = the rate of heat loss in the vent gas water vapour (MJ/h)

However, at any given moment in time, a batch fed reactor is never at steady state. The heat sources are either greater or smaller than the heat sinks, causing the sludge temperature to be continually increasing or decreasing. For example, between feed termination and transfer commencement (phase 3, Fig 3.4), the heat sources are greater than the sinks, resulting in a steadily increasing sludge temperature. During feeding (phase 2) the opposite is the case and reactor sludge temperature decreases rapidly. A 'steady state' condition in a batch fed reactor can only be identified over a complete batch cycle [i.e. between corresponding times encompassing all 3 phases together (Fig 3.4)] and should comply with the following criteria:

(1) the reactor sludge temperature at corresponding points of successive batch cycles must be equal

(2) reactor operating parameters must not have been changed for 3 sludge ages.

From the above, if time intervals other than those between corresponding times in successive batch cycles are taken, an unsteady state heat balance must be applied to a batch fed reactor. In this research, the unsteady state heat balance across the Milnerton reactor was constructed over the linear heating phase (phase 3) and formed the basis for determining the rate of biological heat generation, H_{bi}. During this phase, the sludge volume in the reactor remained constant (due to an absence of sludge transfer and feeding). Also, because of the low gas flows through the reactor resulting from the use of pure oxygen, the mass of water vaporized from the reactor during the heating phase was less than 0,007% of the 45 m³ volume i.e. negligible. This constancy of volume during the heating phase has, from a measurement point of view, two advantages which were mentioned earlier viz. (1) the vent gas volumetric flow rate and composition were not disturbed and (2) the heat gained by the sludge is directly proportional to the sludge temperature increase. The latter advantage allows the unsteady state heat balance over the linear heating phase (3) to be written as:

The change in reactor sludge enthalpy = Σ heat sources – Σ heat sinks

Accepting, as mentioned above, that for the Milnerton pure oxygen oxygenation system, the gas sensible heats are negligible, the unsteady state heat balance can be expressed mathematically as:

$$H_{net} = H_{bi} + H_{mi} - H_{ve} - H_{we} \quad \text{(MJ/h)} \tag{3.5}$$

and

$$H_{net} = V_p \times \rho_s \times Cp \times \frac{dT_{se}}{dt} \quad \text{(MJ/h)} \tag{3.6}$$

where,

H_{net} = the rate of reactor sludge enthalpy increase during the specific heat yield test (MJ/h)
V_p = the operating volume of the reactor (m³)
C_p = the specific heat of the sludge [MJ/(ton.°C)]
ρ_s = the density of the sludge (ton/m³)
dT_{se}/dt = the rate of reactor sludge temperature change (°C/h)

For sewage sludge, the specific heat C_p and the density ρ_s were accepted to be equal to those of water, i.e. $C_p = 4.184$ MJ/(ton.°C) and $\rho_s = 1.000$ ton/m³. The presence of up to 4% dry mass solids of which 80% is organic was considered to contribute negligibly to the specific heat because its contribution to the volume is included in the sludge volume measurement and the product of the specific heat and density of the organic (80%) and inorganic (20%) dry solids is less than half of that for water, e.g. for wood as representative of organic material C_p = 2.7 MJ/(ton.°C) and ρ = 0,42 ton/m³ which gives $C_p\rho_s \approx$ 1,0 MJ/(m³.°C) and for marble, as representative of the inorganic material C_p = 0.8 MJ/(ton.°C) and ρ_s = 2.6 ton/m³ giving a $C_p\rho_s$ = 2.1 MJ/(m³.°C). Hence, accepting the sludge volumetric flow to be pure water overestimates the product of the specific heat and density of the sludge liquor but clearly negligibly so.

All the terms in Eqs (3.5) and (3.6), with the exception of the biological heat generation rate, H_{bi}, can be calculated from measurements; H_{bi} therefore can be found by difference. The methods employed for determining dT_{se}/dt, H_{mi}, H_{we} and H_{ve} are described below.

3.4.3 Measuring the heat balance terms

3.4.3.1 *Rate of sludge enthalpy increase (H_{net})*

The operating volume, V_p, was set at 45 m³ by the level control system (Section 3.3.2). As mentioned above (Section 3.4.2) the sludge density, ρ_s, and specific heat, Cp, were taken to be that of water i.e. 1.00 ton/m³ and 4.184 MJ/(ton.°C) respectively. The rate of reactor sludge temperature increase dT_{se}/dt was always linear throughout a specific heat yield test provided the oxygen supply rate was kept constant. During the specific heat yield test the reactor sludge temperatures from 2 of the 4 reactor temperature monitoring points were recorded at 10 min intervals for periods ranging from ± 1.5 h (at short retention times and high heating rates) to 5.5 h (at long retentions and low heating rates). Hence 2 sets of temperature and time data were available to accurately determine the value of dT_{se}/dt. Determination of dT_{se}/dt could be obtained either from the slope of a linear regression of T_{se} on time or by dividing the difference between the last and the first temperature reading by the test time interval ($\Delta T_{se}/\Delta t$). The latter method was preferred as on a number of occasions it yielded identical results for both of the two temperature data sets, while the former gave different values for the two data sets. The measured value of dT_{se}/dt was substituted into Eq (3.6) to calculate H_{net}.

3.4.3.2 *Mechanical heat input rate (H_{mi})*

The heat source due to mechanical agitation resulted from the action of the Vitox recirculation pump. The rate of this mechanical heat input was a function of the power consumed by the pump's electric motor. Of the power consumed, some was lost due to electric motor inefficiencies and due to transmission losses in bearings and couplings. These losses did not generally amount to more than 10-15% of the total power consumption with the remaining power all converted to heat in the pumped sludge. The mechanical heat input (H_{mi}) therefore was determined from the following equation:

$$H_{mi} = f_{mech} \times V \times I \times \sqrt{3} \times \cos\phi \times \frac{3.6}{1000} \quad \text{(MJ/h)} \qquad (3.7)$$

where

f_{mech} = the fraction of electric power consumed by the pump electric motor which was converted to heat in the reactor sludge
V = phase voltage (volts)
I = phase current (amps)
$\cos\phi$ = power factor
3.6/1000 = conversion from watts to MJ/h

The phase voltage (V) at Milnerton was 380 volts and the phase current (I) was measured on the pump ammeter during a specific heat yield test. From engineering data the power factor ($\cos\phi$) was estimated at 0.82 for the operational conditions of the motor. The fraction of power converted to heat, f_{mech}, was measured experimentally on the reactor and found to be 0.85. Details of the measurement of f_{mech} are given in Appendix 3C.

The factor f_{mech} was measured by filling the reactor with sludge to its operating volume of 45 m³ at ambient temperature (i.e. no wall heat loss), running the recirculation pump over a 5 to 6h period with no oxygen input (no biological heat generation and no water vapour heat loss) and noting the temperature increase in the sludge as a result of the pump energy input.

Substituting the appropriate values into Eq (3.7) yields:

$$H_{mi} = 1.65 \times I \quad (MJ/h) \tag{3.8}$$

Equation (3.8) was used to determine H_{mi} in each of the specific heat yield tests from the observed current drawn (I).

3.4.3.3 *Wall heat loss rate (H_{we})*

The heat loss through the walls and pipework of the aerobic reactor is due to a combination of conduction and convection. It can be quantified with the aid of an overall heat transfer coefficient (U) as in Eq (3.9) i.e.

$$H_{we} = U \times A \times (T_{se} - T_{amb}) \quad (MJ/h) \tag{3.9}$$

where,

A = surface area of the reactor and the Vitox oxygenation system (m^2)
U = the overall heat transfer coefficient [$MJ/(m^2.^\circ C.h)$]
T_{se} = the temperature of the reactor sludge (°C)
T_{amb} = the ambient temperature (°C)

The value of A for the Milnerton system was calculated to be 77 m^2. The value of the overall heat transfer coefficient (U) was measured experimentally on the aerobic reactor and found to be 0.00733 $MJ/(m^2.^\circ C.h)$. The U value was measured by operating the reactor with sludge at pasteurization temperature (~ 60°C) (i.e. with wall heat loss), running the recirculation pump with no oxygen input (i.e. no biological heat generation and no water vapour heat loss), and noting the sludge temperature change. Knowing the pump heat input rate (H_{mi}) from Eq(3.8) and the rate of sludge enthalpy increase (H_{net}) from Eq(3.6) allowed H_{we} to be calculated with the aid of Eq(3.5). Hence U could be calculated via Eq(3.9). Details of the measurement of U are given in Appendix 3D.

Substituting the measured values of U and A into Eq (3.9) yields

$$H_{we} = 0.564 \times (T_{se} - T_{amb}) \quad (MJ/h) \tag{3.10}$$

The value of T_{se} increased during the specific heat yield test and hence the average value for the test was substituted for T_{se} in Eq (3.10). Equation (3.10) was used to determine H_{we} from measured values of T_{se} and T_{amb} for each of the specific heat yield tests.

3.4.3.4 *Vent gas water vapour heat loss rate, (H_{ve})*

The flow of dry oxygen through the hot sludge in the reactor caused water to be vaporized. The amount of water vaporized is a function of the temperature and flow rate of the vent gas. The amount of heat lost due to water vaporization is represented by the following equation:

$$H_{ve} = h_{fg} \times M_{we} \quad (\mathrm{MJ/h}) \tag{3.11}$$

where

h_{fg} = the latent heat of vaporization of water (MJ/kg)

M_{we} = the mass flow rate of water vapour leaving the reactor in the vent gas (kg/h)

The latent heat of vaporization, h_{fg}, was obtained from steam tables [Mayhew and Rogers (1977)] and the following function was derived:

$$h_{fg} = 2.501 - 2.386 \times 10^{-3} \times T_{ge} \quad (\mathrm{MJ/kg}) \tag{3.12}$$

where

T_{ge} = the temperature of the vent gas at the point of exit from the reactor (°C).

The vent gas temperature from the reactor was recorded during each of the specific heat yield tests. Over a specific heat yield test time interval it changed linearly in the same fashion as the reactor sludge temperature, i.e. increased ± 3°C, but was 2 to 5°C below the corresponding sludge temperatures, depending on the vent gas flow rate (2°C at high flow rates and 5°C at low flow rates). The change in h_{fg} with the ± 3°C temperature change during an individual specific heat yield test is so slight that using an average value for the range of T_{ge} recorded during an individual test was found to be sufficiently accurate for calculating the vent gas vapour heat loss (H_{ve}).

The temperature difference between the reactor sludge (T_{se}) and the vent gas at the point of exit from the reactor (T_{ge}) arises as a result of wall heat losses through the reactor head space. In effect, this wall heat loss is not fully taken account of in the measurement of the overall heat transfer coefficient because, in this test no oxygenation took place and therefore no vent gas was produced. However, this heat loss is negligibly small. It is equal to the heat of vaporization of the vent gas flow in cooling from T_{se} to T_{ge}. In cooling the vent gas from T_{ge} to the water vapour condenser temperature (20°C), the heat loss was less than 10% of the total heat losses. Therefore in the vent gas cooling from T_{se} to T_{ge} (i.e. 2 to 5°C) the heat loss will be even less and therefore small enough to be ignored.

To measure the mass of water vaporized, M_{we}, the reactor vent gas was cooled in a condenser and the condensate was collected as described above in Section 3.4.1.2 (see Fig 3.6). Although the condenser removed most of the water vapour from the vent gas in cooling it, the vent gas nevertheless remained saturated with water vapour at the temperature of the condenser coolant (± 20°C). The flow rate of this remaining water vapour was calculated from the following equation [for derivation of Eq (3.13) see Appendix 3B]:

$$M_{wr} = \frac{5.22 \text{ x } V_{gm}}{2.73 + T_{gm}} \quad \text{(kg/h)} \qquad (3.13)$$

where

M_{wr} = the flow rate of the remaining water vapour in the partially dried vent gas (kg/h)

V_{gm} = the vent gas flow rate measured at the vent gas meter (m³/h)

T_{gm} = the temperature of the vent gas in the vent gas meter (°C)

The total vent gas water vapour flow rate (M_{we}) was taken to be the sum of the measured rate of condensate accumulation in the two water traps during a specific heat yield test (M_{wc}) and the calculated remaining vent gas water vapour flow rate (M_{wr}). Knowing the total vent gas water vapour mass flow rate (M_{we}) and the latent heat of vaporization (h_{fg}) at the average vent gas temperature during a specific heat yield test (T_{ge}), allowed calculation of the rate of heat loss via water vaporization (H_{ve}) with the aid of Eq (3.11).

3.5 SAMPLE SPECIFIC HEAT YIELD CALCULATION

The measurement and calculation procedures for determining the mass biological oxygen consumption rate (O_c) and the 4 measurable terms in the unsteady state heat balance viz. the rate of increase in reactor sludge enthalpy (H_{net}), the mechanical heat input rate (H_{mi}), the wall heat loss rate (H_{we}) and the vent gas vapour heat loss rate (H_{ve}), allowed calculation of the biological heat generation rate (H_{bi}) and the specific heat yield (Y_H) i.e. the biological heat generated per kg of oxygen consumed. In addition, the oxygen utilization efficiency (O_{eff}), the respiration quotient (Y_{CO_2}), and the degree of water vapour saturation in the vent gas could be calculated from the specific heat yield test data.

To demonstrate the calculation procedure of the biological heating rate and the specific heat yield, a sample calculation employing the results of an actual test is given below. The test data are presented in Table 3.1 in the format in which they were measured. In the calculations, the equations required are referred to by number and are written with the appropriate values substituted for the equation variables.

3.5.1 The rate of increase in reactor sludge enthalpy, H_{net}

Taking the first and last reactor sludge temperature readings (this gave less variable results for the net heating rate than a linear regression on the temperature time data - see section 3.4.3.1) for each of the 2 probes, gives the heating rates to be:

$$\text{Probe 1: } \Delta T_{se}/\Delta t = 2.9/2.167 = 1.338\ ^\circ C/h$$
$$\text{Probe 2: } \Delta T_{se}/\Delta t = 2.9/2.167 = 1.388\ ^\circ C/h$$

Hence, via Eq (3.6), the rate of change of sludge enthalpy is:

$$\begin{aligned} H_{net} &= V_p * \rho_s * Cp * dT/dt \\ &= 45 * 1 * 4.184 * 1.338 \\ &= 251.9 \text{ MJ/h.} \end{aligned}$$

3.5.2 Mechanical heat input rate, H_{mi}

The current drawn by the Vitox recirculation pump was 45 amps. Hence from Eq (3.8):

$$\begin{aligned} H_{mi} &= 1.65 * 45 \\ &= 74.3 \text{ MJ/h} \end{aligned}$$

3.5.3 Wall heat loss rate, H_{we}

The ambient temperature during the specific heat yield test was 13.5° C and the average reactor sludge temperature is calculated as 55.5° C. Hence from Eq (3.10):

$$\begin{aligned} H_{we} &= 0.564 * (55.5 - 13.5) \\ &= 23.7 \text{ MJ/h} \end{aligned}$$

3.5.4 Vent gas vapour heat loss rate, H_{ve}

The average vent gas temperature was 52.9° C during the 2h period of condenser operation in the specific heat yield test. The latent heat of vaporization at 52.9° C is calculated from Eq (3.12):

$$\begin{aligned} h_{fg} &= 2.501 - 2.386 * 10^{-3} * 52.9 \\ &= 2.375 \text{ MJ/kg} \end{aligned}$$

The vent gas condensate accumulated at a rate of 1.74ℓ in 2h from which M_{wc} = 0.87 kg/h. To this is added the remaining uncondensed vent gas water vapour M_{wr} calculated from the measured vent gas flow rate and vent gas temperature via Eq (3.13) i.e.

$$\begin{aligned} M_{we} &= M_{wc} + M_{wr} \\ &= 0.87 + 5.22 * 10.2/(273 + 27) \\ &= 1.05 \text{ kg/h} \end{aligned}$$

Hence from Eq (3.11) the heat loss via water vapour in the vent gas is:

$$\begin{aligned} H_{we} &= h_{fg} * M_{we} \\ &= 2.375 * 1.05 \\ &= 2.5 \text{ MJ/h} \end{aligned}$$

TABLE 3.1: Sample specific heat yield test data

Date: 12 September 1988		Weather: Sunny & clear	
Ambient 13.5°C	Feed temp 18.3°C	Amps 45	Oxygen 18.5 kg/h

Time	Reactor sludge T_{se} Temp 1	Reactor sludge T_{se} Temp 2	Vent Temp T_{ge}	Gas Meter Temp	Gas Meter Reading
9h20	53.8	54.2	51.6	25.5	3618.5
30	54.0	54.4	51.8	26.0	3620.2
40	54.3	54.6	52.0	26.5	3621.8
50	54.5	54.8	52.2	26.5	3623.4
10h00	54.8	55.0	52.4	27.0	3625.0
10	55.0	55.3	52.6	27.0	3626.7
20	55.2	55.5	52.8	27.0	3628.5
30	55.3	55.8	53.1	27.0	3630.2
40	55.6	56.0	53.3	27.5	3631.9
50	55.9	56.2	53.5	27.5	3633.7
11h00	56.0	56.4	53.7	27.0	3636.4
10	56.3	56.6	53.9	28.0	3637.3
20	56.5	56.8	54.2	27.5	
30	56.7	57.1	54.4		
ΔT_{se}	2.9	2.9			

Gas meter @ 11h15 3638.0
Gas meter @ 9h15 3617.6
(m^3) 20.4

Condensate collected from
9h20 - 11h20 = 1.74 litre

Orsat analysis
@ 11h10 : 23% O_2
@ 11h18 : 23% O_2

Data summary

V_{gm} : 10.2 m^3/h
T_{gm} : 27 °C
f_{O2} : 0.23
O_s : 18.5 kgO/h
M_{wc} : 0.87 kg/h
T_{ge} (avg) : 53 °C
$\Delta T_{se}/\Delta t$: 1.338 °C/h
I : 45 amps
T_{amb} : 13.5 °C

3.5.5 Biological heat generation rate, H_{bi}

Knowing values for all the measurable terms of the unsteady state heat balance, allows the calculation of H_{bi} by difference, i.e. from Eq (3.5)

$$\begin{aligned} H_{bi} &= H_{net} - H_{mi} + H_{we} + H_{ve} \\ &= 251.9 - 74.3 + 23.7 + 2.5 \\ &= 203.8 \text{ MJ/h} \end{aligned}$$

3.5.6 Mass oxygen consumption rate, O_c

This is obtained with the aid of Eq (3.1) from the difference in the mass oxygen supply rate (O_s) and the mass flow rate of oxygen in the vent gas. The former was set at 18.5 kgO/h. The vent gas flow rate, temperature and oxygen fraction were 10.2 m³/h, 27°C and 0.23 respectively; hence from Eq (3.1), the biological oxygen consumption rate (O_c) is

$$\begin{aligned} O_c &= 18.5 - (380.8 * 0.23 * 10.2)/(273 + 27) \\ &= 15.5 \text{ kgO/h} \end{aligned}$$

3.5.7 The specific heat yield, Y_H

The specific heat yield is the ratio of the biological heat generation rate (H_{bi}) and the biological mass oxygen consumption rate (O_c) i.e.

$$\begin{aligned} Y_H &= H_{bi}/O_c \\ &= 203.8/15.5 \\ &= 13.1 \text{ MJ/(kg } O_2) \end{aligned}$$

3.6 VENT GAS ANALYSIS CALCULATIONS

From the data that were required to be measured on the vent gas for the specific heat yield calculation, there are 3 additional parameters that were calculated and which are not directly associated with the specific heat yield. These are:

(1) the oxygenation efficiency (O_{eff})
(2) the respiration quotient (Y_{CO_2})
(3) the degree of water vapour saturation of the vent gas.

The first 2 of these parameters were discussed earlier in Section 3.4.1.2 of this chapter and Eqs (3.2) and (3.3) were presented for their calculation (the derivation of these equations is outlined in Appendix 3B). The last parameter, that is, the degree of water vapour saturation, was more difficult to calculate and required a different approach. The details of this approach are outlined in Appendix 3B and an example of the calculations is given below for the specific heat yield test data given in Table 3.1.

3.6.1 The oxygenation efficiency, O_{eff}

This is simply the biological oxygen consumption rate as a percentage of the oxygen supply rate, i.e. from Eq (3.2) we get:

$$\begin{aligned} O_{eff} &= O_c/O_s * 100 \\ &= 15.5/18.5 * 100 \\ &= 83.8\% \end{aligned}$$

3.6.2 The respiration quotient, Y_{CO_2}

The respiration quotient, that is, the moles of carbon dioxide produced per mole of oxygen consumed, is calculated from the measured vent gas flow rate (V_{gm}), the temperature of the vent gas at the gas meter (T_{gm}) and the fractional volume of oxygen in the dried vent gas (f_{O_2}). From Eq (3.3):

$$\begin{aligned} Y_{CO_2} &= 380.7 * (1-0.23) * 10.2/[15.5 * (273 + 27)] \\ &= 0.64 \end{aligned}$$

3.6.3 The degree of water vapour saturation in the vent gas

The degree of water vapour saturation in the vent gas is assessed by comparing the actual mass flow rate of water vapour measured during the specific heat yield test, M_{we}, with the predicted mass flow rate, M_{wp}. Procedures employed for calculating the predicted mass flow rate, M_{wp}, assume that the vent gas was saturated with water vapour. Details of this procedure are briefly set out below but are detailed in Appendix 3B.

$$M_{wp} = 18 * m_{dg} * p_w/p_{dg} \quad (kg/h)$$

where

18 $=$ molar mass of water (kg/kmol)

m_{dg} $=$ molar flow rate of dry vent gas

$= 11.9 * V_{gm}/(273 + T_{gm})$

$= 11.9 * 10.2/(273 + 27)$

$= 0.4046$ kmol/h

p_w $=$ partial pressure of saturated water vapour at vent gas temperature at the point of exit from reactor T_{ge}

$= 10^{[8.896 - 2238/(Tge + 273)]}$

$= 10^{[8.896 - 2238/(53 + 273)]}$

$= 107$ mmHg

p_{dg} $=$ dry vent gas partial pressure (mmHg)

$=$ atmospheric pressure $- p_w$

$= 760 - p_w$

$= 760 - 107$

$= 653$ mmHg

Therefore,

$$M_{wp} = 18 * 0.4046 * 107/653$$
$$= 1.19 \text{ kg/h}$$

The measured mass flow rate of vent gas water vapour, M_{we}, was 1.05 kg/h (see Section 3.5.4 above), 12% lower than the predicted value of 1.19 kg/h. Hence for this particular specific heat yield test the vent gas from the reactor was 88% saturated with water vapour.

3.7 FLOW RATE AND CHEMICAL ANALYSES OF THE SLUDGE

In order to determine the mass of sludge treated per day, and the mass of volatile solids (VS), total solids (TS) and chemical oxygen demand (COD) removed in the aerobic reactor, it was necessary to do TS, VS and COD mass balances across the aerobic reactor. This required the flow rate and the TS, VS and COD concentration measurements to be conducted on the sludge. Other measurements of interest conducted on the sludge were pH and conductivity.

From time to time the results obtained from the chemical tests were cross checked against results obtained on identical samples by the Division of Water Technology

(DWT). Generally a favourable comparison was obtained – details of the comparison are given in Appendix 4D.

The DWT also undertook a number of additional tests on the sludge, not only on the aerobic reactor but also on the anaerobic digestor. A list of these tests is given in Appendix 5B. Because these results are not of direct interest to this thesis, they are omitted. These results are given in other reports on the dual digestion project [De Villiers *et al* (1987 to 1990)].

3.7.1 VS and COD balance

The masses of VS and COD removed from the sludge by the aerobic reactor are found by the product of the sludge flow rate and the changes in concentration of these parameters across the reactor. The accuracy of the former depended on the level control system and of the latter, on the sludge sampling and analytical procedures.

3.7.2 Sludge flow rate measurement

The daily volume of sludge passing through the reactor was determined by the product of the number of batch feeds and the batch volume. As described in Section 3.3.2, the level control system ensured that each batch of sludge fed to the reactor had a volume of 3.75 m^3. From the chart recordings of the reactor temperature trace, the saw-tooth profile clearly indicated when feeding took place and the number of feeding periods per day could be easily counted.

3.7.3 Sludge sampling and analysis

The nature of sludge makes accurate analysis vulnerable to errors in the collection of samples. Sludge concentrations in the central sludge collection sump of the Milnerton treatment works vary over the day due to: (a) manual desludging by different operators who stopped the PST underflow to the sump when, in their opinion, the sludge appeared to become dilute, (b) scum collection systems on two of the PST's which dumped varying quantities of water/scum into this sump.

The fibreglass prethickener partially smoothed out concentration variance but some stratification did occur, especially in summer when high temperatures led to gas generation in the vessel.

A system of composite sampling was devised to ensure obtaining representative samples. Samples were taken for each of those batch feeds that occurred between 08h00 and 24h00, a period when the shift operators were on duty.

Samples were drawn from the discharge pipework of the transfer and feed pumps. Stratification and the deliberate absence of mixing of the sludge in the sludge thickener could have resulted in variations in the concentration of feed sludge over the pumping cycle. To minimize the effect of these variations on the feed sample, the sludge was tapped from the pump discharge line in short, regular bursts over the entire duration of the feed cycle. The same procedure was followed for the transfer sludge sample, although, given the degree of mixing in the aerobic reactor (400 W/m^3), the transfer sludge concentration was likely to be much more uniform than the feed sludge.

The transfer pump operated for about 4 min during which time the operator bled a sample in short bursts from the transfer line into a 25ℓ bucket. 1.5ℓ of this sample was deposited into another bucket after which the sampling apparatus was cleaned. The same procedure was followed for the feed sludge sample. The following morning at the start of the first shift, the composite buckets containing the feed and transfer samples were stirred thoroughly and a 1.5ℓ sample was taken from each for laboratory analysis.

3.7.4 Solids concentration measurement

Initially, total and volatile *suspended* solids tests were performed, but the samples of aerobically treated sludge (i.e. the transfer sample) contained a considerable proportion of suspended colloidal material which resisted separation by both filtration and centrifugation. It was therefore decided to do total and volatile solids tests by drying the whole sample volume without prior liquid/solid separation.

Feed and transfer sludge samples of 50 mℓ were poured into dry weighed crucibles, dried for 24h at 103° C and cooled in a desiccator. After weighing to determine total solids, the samples were incinerated at 600° C for 1h and, following cooling in a desiccator again, were reweighed to determine the volatile solids.

3.7.5 COD concentration measurement

Feed and transfer sludge samples of 100 mℓ were macerated and diluted 100 times. A volume of 10 mℓ of the diluent was taken for the COD test which was carried out in accordance with the method outlined in Standard Methods for the Examination of Water and Wastewater (1985). The COD test procedure was occasionally checked against standards and was found to be accurate.

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CHAPTER 4

DESCRIPTION AND DISCUSSION OF RESULTS MEASURED ON AEROBIC REACTOR

ABSTRACT

In this Chapter the results of 116 specific heat yield tests, as well as those of the TS, VS and COD concentration measurements, are described. From these results it was found that: the biological heating rate was directly proportional to the biological oxygen consumption rate. The constant of proportionality is the specific heat yield (Y_H) which was measured to be 12,77 ± 0,58 MJ/kgO, a value which conforms closely to bioenergetically calculated and microbiologically measured values. (2) The respiration quotient (Y_{CO_2}) i.e. mole CO_2 generated per mole O_2 utilized) was 0,66 instead of 1,0 usually assumed. From this it appears that the oxidation reactions in the sludge are not only those of VS degradation, as is commonly accepted. (3) The vent gas was saturated with water vapour even at the high vent gas flow rates. (4) The COD and/or VS removal rates are poor parameters for (i) quantifying the biological heat generation rate and (ii) controlling the reactor temperature, because the tests (a) are prone to significant variability when dealing with sewage sludges and (b) take too long to give a result. (5) The reactor temperature increases could be completely and instantaneously controlled by means of the oxygen supply rate for as long as the reactor was oxygen limited, i.e. while the oxygen transfer rate (OTR) was less than the maximum biological oxygen utilization rate (OUR), which was 380 mgO/(m^3.h) at an average VS concentration of 30 kgVS/m^3. (6) The close correlation and rapidity of response between biological heat generation rate and the biological oxygen consumption rate and oxygen supply rate, make these two parameters pivotal ones in the modelling, design and control of autothermal thermophilic aerobic reactors in dual digestion.

4.1 INTRODUCTION

In Chapter 3, the design and operation of the aerobic reactor in the Milnerton dual digestion system were described. Also the oxygen mass and unsteady state heat balance formulations were derived, and the procedures outlined whereby the various parameters in these balances were measured. In this chapter, the results obtained from the oxygen mass and heat balance measurements are presented and discussed.

From Chapter 3, the unsteady state heat and oxygen mass balances are:

$$H_{net} = V_p * C_p * \rho_s * \frac{dT_{se}}{dt}$$

$$= H_{bi} + H_{mi} - H_{we} - H_{ve} \qquad \text{(MJ/h)} \qquad (3.6)$$

$$O_c = O_s - \frac{380.8 \text{ x } f_{O_2} * V_{gm}}{273 + T_{gm}} \qquad \text{(kg/h)} \qquad (3.1)$$

where,

H_{net} = rate of increase of reactor sludge enthalpy {MJ/h}
V_p = the sludge process volume in the reactor {45m³}
ρ_s* = density of sludge liquor {1000 kg/m³}
C_p* = the specific heat of the sludge liquor {4.184 MJ/ton/°C}
dT_{se}/dt = rate of change of reactor sludge temperature {°C/h}
H_{bi} = biological heat generation rate {MJ/h}
H_{mi} = mechanical heat generation rate {MJ/h}
H_{we} = wall heat loss rate {MJ/h}
H_{ve} = water vapour heat loss rate {MJ/h}
O_c = biological oxygen consumption rate {kg/h}
O_s = mass oxygen supply rate {kg/h}
f_{O_2} = volume fraction of oxygen in the dry vent gas
V_{gm} = the volumetric flow rate of vent gas passing through the vent gas meter {m³/h}
T_{gm} = the temperature of the gas passing through the vent gas meter {°C}

* Values accepted to be equal to those of pure water.

The simultaneous execution of an unsteady state heat balance and an oxygen mass balance constituted a *specific heat yield test* in which 6 parameters, viz:

(1) the biological heating rate (H_{bi})
(2) the oxygen consumption rate (O_c)

(3) the specific heat yield (Y_H)
(4) the oxygenation efficiency (O_{eff})
(5) the respiration quotient (Y_{CO_2})
(6) the degree of water vapour saturation in the vent gas

were calculated from 8 measurements, viz:

(1) rate of increase of reactor sludge temperature ($\Delta T_{se}/\Delta t$)
(2) recirculation pump electric current drawn (I)
(3) ambient temperature (T_{amb})
(4) oxygen mass supply rate (O_s)
(5) temperature and volumetric flow rate of the partially dried vent gas passing through the vent gas meter (T_{gm}, V_{gm})
(6) volume fraction of oxygen in the dry vent gas (f_{O_2})
(7) water vapour condensate accumulation rate (M_{we})
(8) temperature of the vent gas at the exit point of the reactor (T_{ge})

In addition to their use in calculating the 6 parameters listed above, on 3 occasions certain of the specific heat yield tests were employed to assess the degree of oxygen limitation in the reactor sludge. This assessment of oxygen limitation was of a semi-quantitative nature and entailed comparing the oxygen consumption rates (O_c) measured in specific heat yield tests which were performed immediately before and after a step increase in the oxygen supply rate (O_s).

Altogether over a period of 8 months of operation, 116 specific heat yield tests (i.e. oxygen mass and unsteady state heat balances) were conducted, covering a range of mass oxygen supply rates (O_s) from 5.9 to 24.3 kgO/h, retention times (R_H) from 1.2 to 3 days and reactor sludge temperature (T_{se}) from 54 to 69°C. The number of tests conducted at each reactor sludge temperature, mass oxygen supply rate range and retention time are set out in Table 4.1.

In this chapter the results of the 116 specific heat yield tests are discussed, covering aspects such as (1) the factors that influence the magnitudes of the terms in the heat and oxygen balances [Eqs (3.5) and (3.1)], and (2) the results of the 6 calculated parameters listed above and the 3 assessments of oxygen limitation. Also the COD and VS removal achieved in the reactor are discussed.

TABLE 4.1: RANGES OF AEROBIC REACTOR OPERATING CONDITIONS.

T_{se} (°C)	54	55	56	57	58	59	60	61
TESTS	4	2	9	6	7	17	9	13
T_{se} (°C)	62	63	64	65	66	67	68	69
TESTS	12	9	7	6	1	5	5	4
O_s (kg/h)	5.9 - 8.1		14.5		16.0 - 16.5		18.5	
TESTS	13		7		29		15	
O_s (kg/h)	19.4		22		24.3			
TESTS	13		28		11			
R_h (days)	1.2		1.25		1.5		3	
TESTS	25		47		31		13	

4.2 THE BIOLOGICAL HEATING RATE H_{bi}

From the unsteady state heat balance [Eq (3.5)], the biological heating rate H_{bi} could be calculated from the experimentally determined values of H_{net}, H_{mi}, H_{ve} and H_{we}. In Fig 4.1 the results obtained for these measured terms of the heat balance as well as the calculated H_{bi} for the 116 tests, are shown plotted with the data ranked in order of increasing H_{net}. A list of the data set plotted in Fig 4.1 is given in Appendix 4A. Factors influencing the magnitude of the 4 measured terms of the heat balance are discussed below.

4.2.1 The rate of increase of reactor sludge enthalpy, H_{net}

The rate of increase of reactor sludge enthalpy was calculated from measurements of the rate of change of sludge temperature dT_{se}/dt. As mentioned in Chapter 3, Section 3.4.3.1, dT_{se}/dt was obtained from reactor sludge temperature T_{se} readings taken at 10 min intervals for the duration of the specific heat yield test from 2 of the 4 PT100's that monitored reactor sludge temperature. In this way 2 sets of temperature and time readings were available for the calculation of H_{net} in each test. The changes in T_{se} with time were clearly linear for both sets

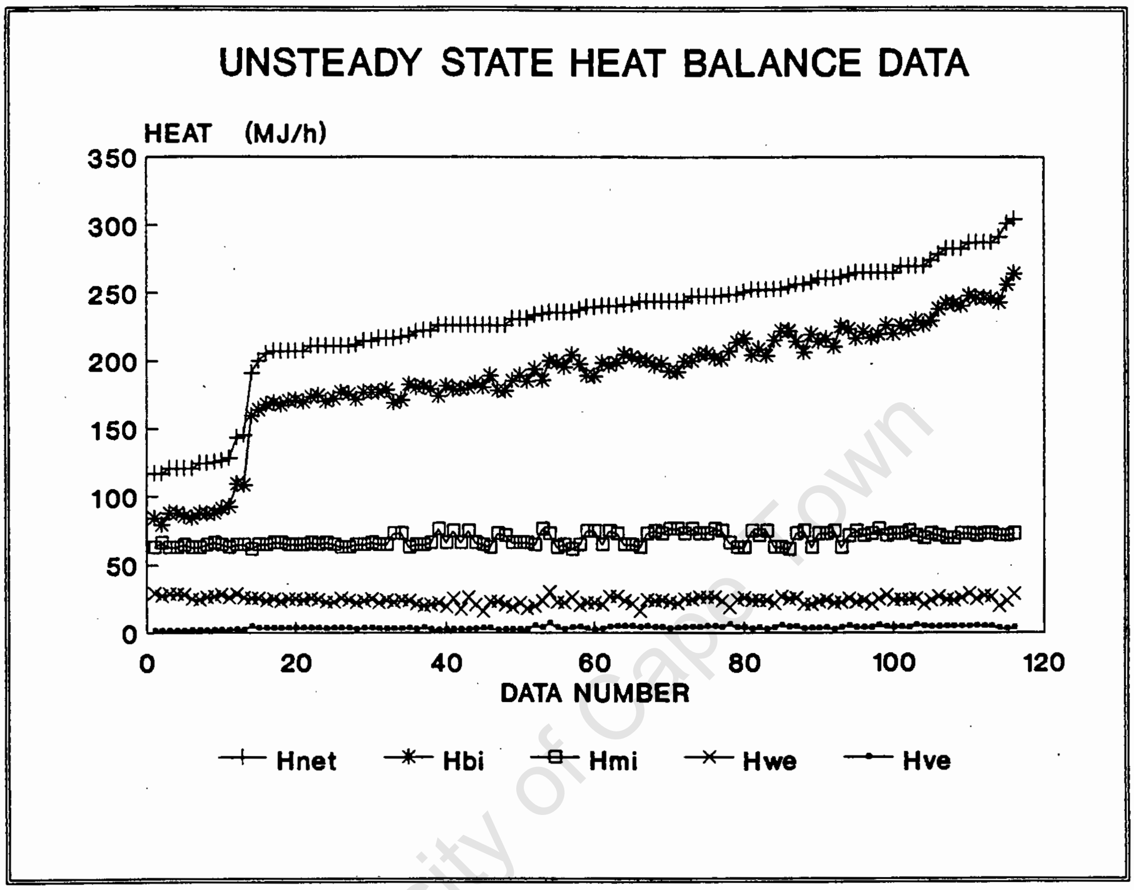

Fig 4.1: The data for the 5 components of the unsteady state heat balance, H_{net}, H_{bi}, H_{mi}, H_{we} and H_{ve}, measured during the 116 specific heat yield tests and ordered in terms of increasing H_{net}.

of data, and this linearity was verified by either graphical plot or linear regression. For example, in 106 regressions that were performed out of the 232 sets of time and temperature data (i.e. 2 in each of the 116 specific heat yield tests), the correlation coefficient (R) was greater than 0.99 in each case. In more than 60% of the 106 cases, the correlation coefficient was greater than 0.999. This linearity of the rate of increase in sludge temperature, dT_{se}/dt allowed this rate to be calculated by dividing the difference between the first and the last reactor sludge temperature values in the specific heat yield test by the test time interval i.e. $\Delta T_{se}/\Delta t$. While dT_{se}/dt could also have been calculated from the slope of the linear regression on the temperature (T_{se}) and time data, the values of dT_{se}/dt so obtained were not as reproducible and consistent between the two temperature probes as those calculated from $\Delta T_{se}/\Delta t$. Knowing the dT_{se}/dt from $\Delta T_{se}/\Delta t$, H_{net} was calculated with the aid of Eq (3.6). For the 116 specific heat yield tests, the values of H_{net} ranged from 116.7 MJ/h at a 3 day reactor retention time (R_H) to 304.1 MJ/h at a 1.25 day reactor retention time (R_H) (see Fig 4.1).

4.2.2 The mechanical heat input rate, H_{mi}

The mechanical heat input rate, H_{mi}, was calculated from the current drawn (I) by the Vitox recirculation pump during the specific heat yield test with Eq (3.8). The value of I, and hence H_{mi}, did not vary during a specific heat yield test because the pumped head remained constant and pumping was continuous. However, closer examination of the H_{mi} data over the 8 month period during which the specific heat yield tests were conducted reveals an interesting phenomenon.

The Vitox recirculation pump was subject to considerable wear due to the grit content of the sludge (for details of this pump performance see Appendix 5B). At about 6 month intervals it was removed and replaced with a similar standby pump while its casing was rebuilt to the original dimensions. Replacing the worn pump casing with the standby one resulted in an immediate decrease in the pump heat input. This is illustrated in Fig 4.2 where the mechanical (pump) heat input rate, H_{mi}, is plotted against the specific heat yield test number, 1 to 116, where the test number increases in chronological order over the 8 month period. In Fig 4.2, it can be seen that as the size of the pump volute increased due to wear with time, so did the pumping rate, the current drawn and hence the value of H_{mi}.

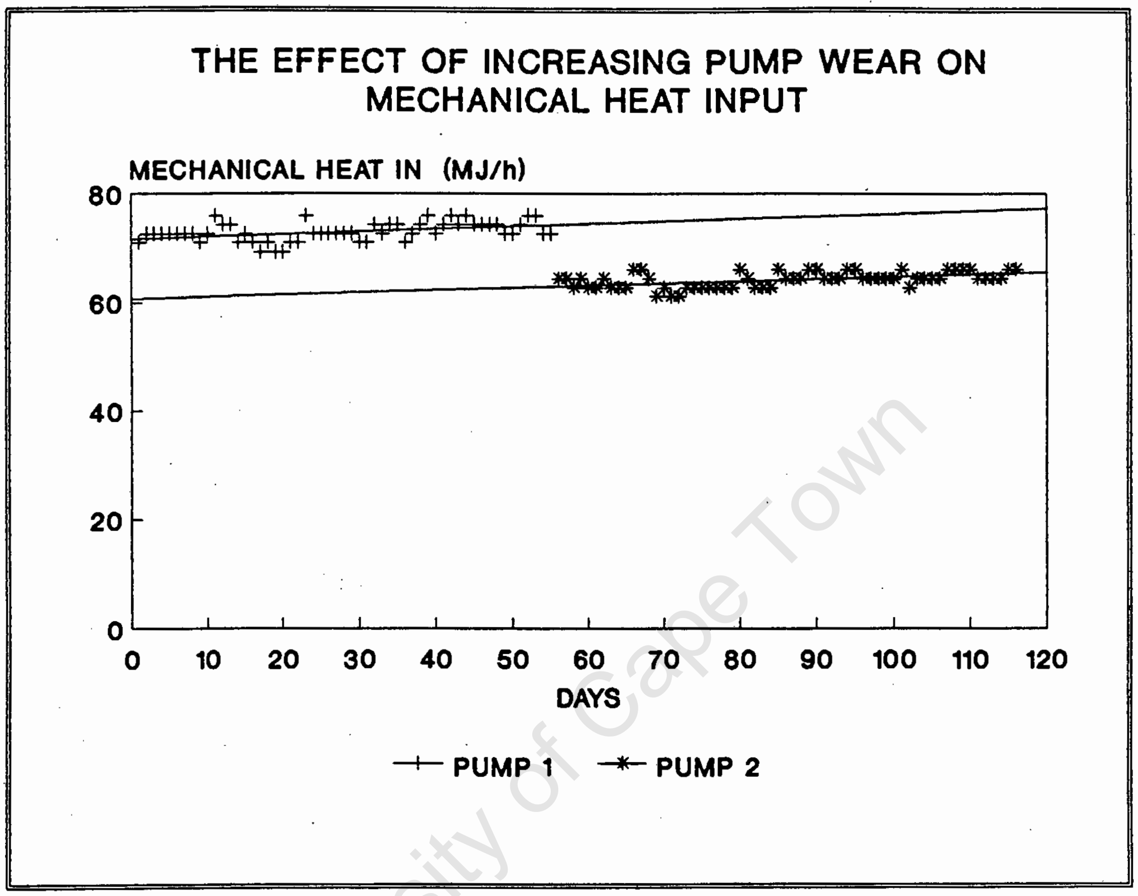

Fig 4.2: The 116 measured values of the rate of mechanical heat input from the Vitox recirculation pump, H_{mi}. The data is arranged in chronological order to show how H_{mi} increased over time due to the gradual increase in the volume of the pump volute caused by abrasive wear inside the cast iron pump casing. As the volute volume increased with time, so did the pumping rate, the current drawn and hence the value of H_{mi} calculated from Eq (3.8).

4.2.3 The wall heat loss rate, H_{we}

The values of the wall heat loss rate, H_{we}, for the 116 tests were calculated from the difference in the measured reactor sludge temperature T_{se} and the ambient temperature T_{amb} with Eq (3.10). It was mentioned in the calibration of Eq (3.10) (Appendix 3D) that the wall heat losses are probably not very accurately estimated by this equation for 2 reasons. The first is because the reactor was subject to cooling by winds of different velocity and was sometimes partly in shade and partly in sunshine. The ambient temperature for Eq (3.10), T_{amb}, was measured in the shade and this value on its own could not possibly fully account for all the environmental factors influencing the wall heat loss. The second is a consequence of the sludge not being oxygenated during the calibration test. In the absence of oxygenation the heat losses from the reactor headspace walls were slightly smaller than those that occurred during normal reactor operation. The vent gases arising from oxygenation convey vapour to these headspace walls and a small portion of this vapour condenses on the walls, cooling the vent gas by about 2° to 5° C, depending on the vent gas flow rate, and releasing heat, most of which is conducted out of the reactor. The smaller headspace heat losses during the calibration test caused the overall heat transfer coefficient U to be slightly underestimated. However, this underestimation is clearly insignificant when considering that the heat lost in the headspace, which is equivalent to the heat of vaporization of the vent gas flow in cooling down 2° to 5° C from the reactor sludge temperature, is a very small fraction of the total water vapour heat loss H_{ve} which was found to be less than 1% of the H_{net} (see Section 4.2.4 below).

From Fig 4.1 it can be seen that H_{we} is relatively small, about 30 MJ/h, comprising 6.4 to 24.7% of H_{net}[1] for the 116 data, the lower percentages at the short retention times (1,2-1,5d) and the higher percentages at the long retention times (3d). For the 103 specific heat yield tests at short retention times (see Table 4.1), H_{we} is less than 12% of the corresponding H_{net}. Consequently the error in H_{we} is considered small and random, with the result that it would not significantly affect the calculation of the biological heat generation rate, H_{bi}, with Eq (3.5).

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[1] Although H_{we} was effectively constant over time, it did not form a constant proportion of H_{net} because H_{net} increased with decreasing retention time R_H.

4.2.4 Vent gas vapour heat loss rate, H_{ve}

The values of the vent gas vapour heat loss rates, H_{ve}, were calculated from the measured vent gas water vapour mass flow rate M_{we} (kg/h), with Eqs (3.11 to 3.13). For the 116 specific heat yield tests, the water vapour mass flow rate ranged from 0.63 to 3.37 kg/h and the resulting heat loss, H_{ve} from 1,5 to 5,8 MJ/h. This heat loss rate formed a very small portion of the heat balance, typically less than 1% of H_{net}. Closer examination of the 116 data showed that H_{ve}, and therefore the mass flow rate of water vapour in the vent gas, M_{we}, increased with increasing oxygen supply, O_s, and vent gas temperature, T_{ge}. This behaviour is illustrated in Fig 4.3 where the vent gas vapour heat loss rate, H_{ve}, is plotted against the vent gas temperature at the point of exit from the reactor T_{ge}, for different mass oxygen supply rates, O_s. This behaviour is consistent with water vapour generation rates predicted from published steam table data, i.e. the greater the vent gas flow rate and temperature the greater the water vapour mass flow rate, M_{we}, and therefore H_{ve}. In Section 4.7 below, such predicted water vapour generation rates will be used to show that the vent gas was saturated with water vapour in all of the 116 specific heat yield test conditions. The low water vapour mass flow rates of 0.63 to 3.37 kg/h verifies that the reactor operating volume remained constant at 45 m³ during the linear heating phase – 3,37 kg/h water loss represents a volume change of 0,007% per hour.

As mentioned in Chapter 3, Section 3.4.1.2, the vent gas temperature at the point of exit from the reactor (T_{ge}) was measured at the 'watershed' point on the vent gas carrier line, i.e. at the point where condensate produced prior to the temperature measurement point would flow back into the reactor and that produced after the temperature measurement point would flow down the vent gas carrier pipe and collect in the primary water trap (Fig 3.6). The value of T_{ge} was monitored during the specific heat yield test and was found to increase over the test period at the same linear rate as the reactor sludge, T_{se}, but was always a constant 2 to 5°C cooler than T_{se}. The difference between T_{se} and T_{ge} was inversely related to the oxygen supply rate, O_s, which is illustrated in Fig 4.4. This inverse relationship is probably a result of increased vent gas retention time in the 5 m³ reactor headspace at low mass oxygen supply rates. A longer headspace retention time exposes the gas to a longer period of heat loss through the reactor head space walls which would allow the vent gas to cool more prior to leaving the reactor – see section 4.2.3 above.

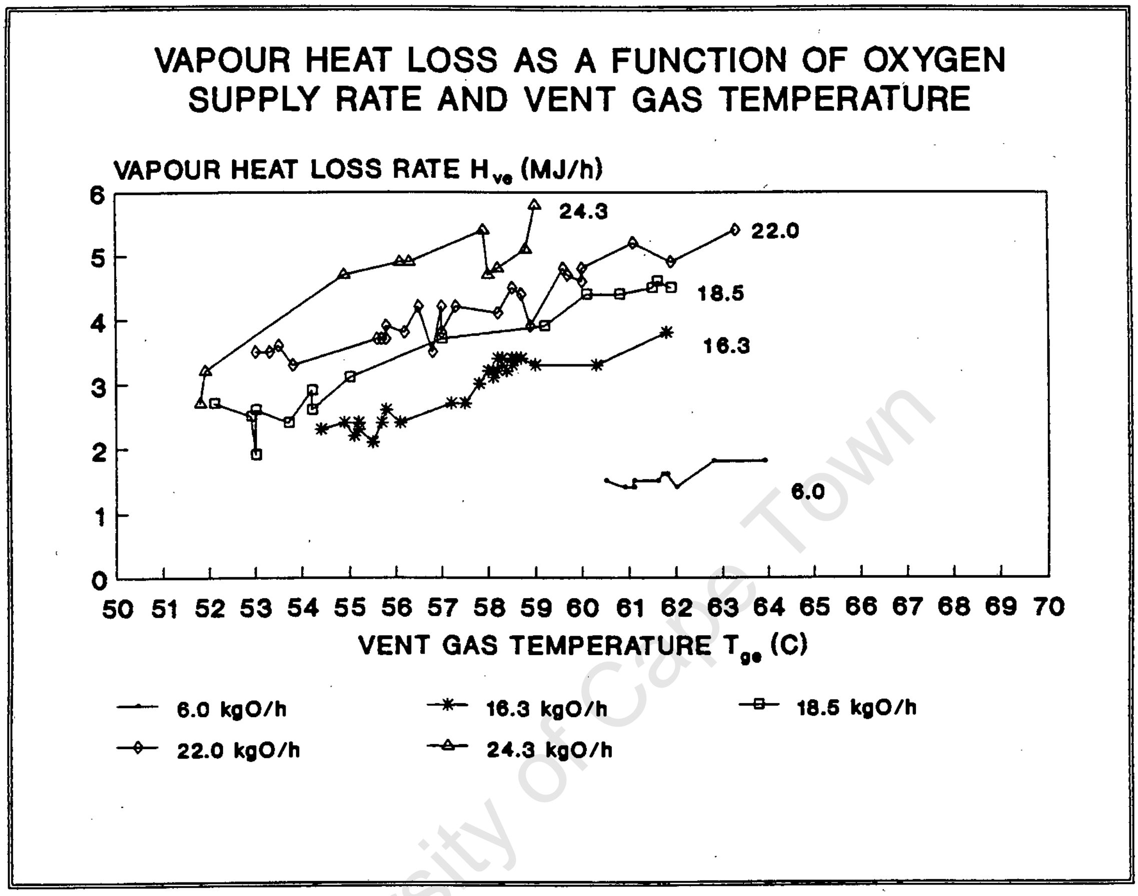


Fig 4.3: 97 of the measured values of the rate of heat loss in the vent gas vapour (H_{ve}) plotted against the vent gas temperature (T_{gr}) for 5 oxygen supply rates (O_s). 19 values of H_{ve} measured at supply rates of 14.5 and 19.4 kgO/h have been neglected in this plot in order to present the relationship between H_{ve}, T_{ge} and O_s more clearly. The plots show H_{ve} to be proportional to T_{ge} and O_s in accordance with accepted thermodynamic theory.

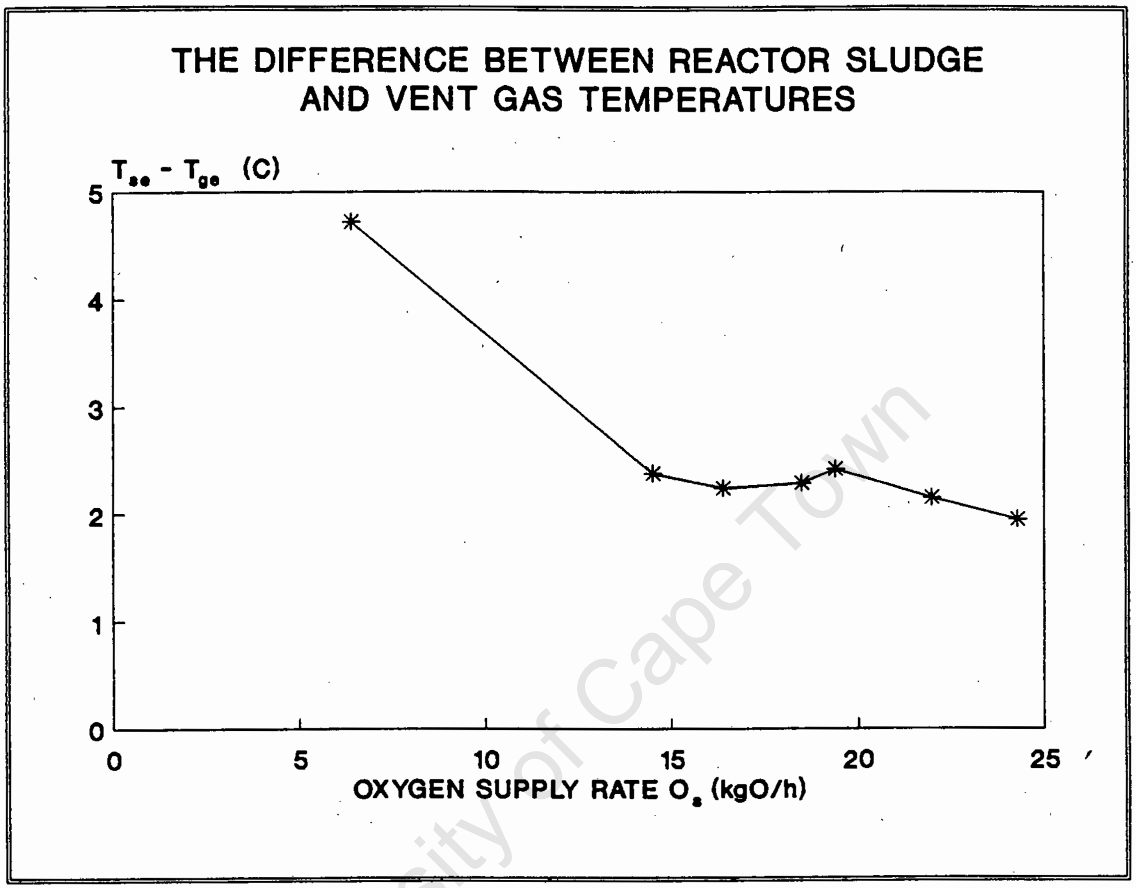


Fig 4.4: The difference between the reactor sludge temperature (T_{se}) and the vent gas temperature (T_{ge}) as a function of the oxygen supply rate, O_s. Each of the 7 plotted values of T_{ge} is the average of the T_{ge} values measured at one of the 7 different oxygen supply rates.

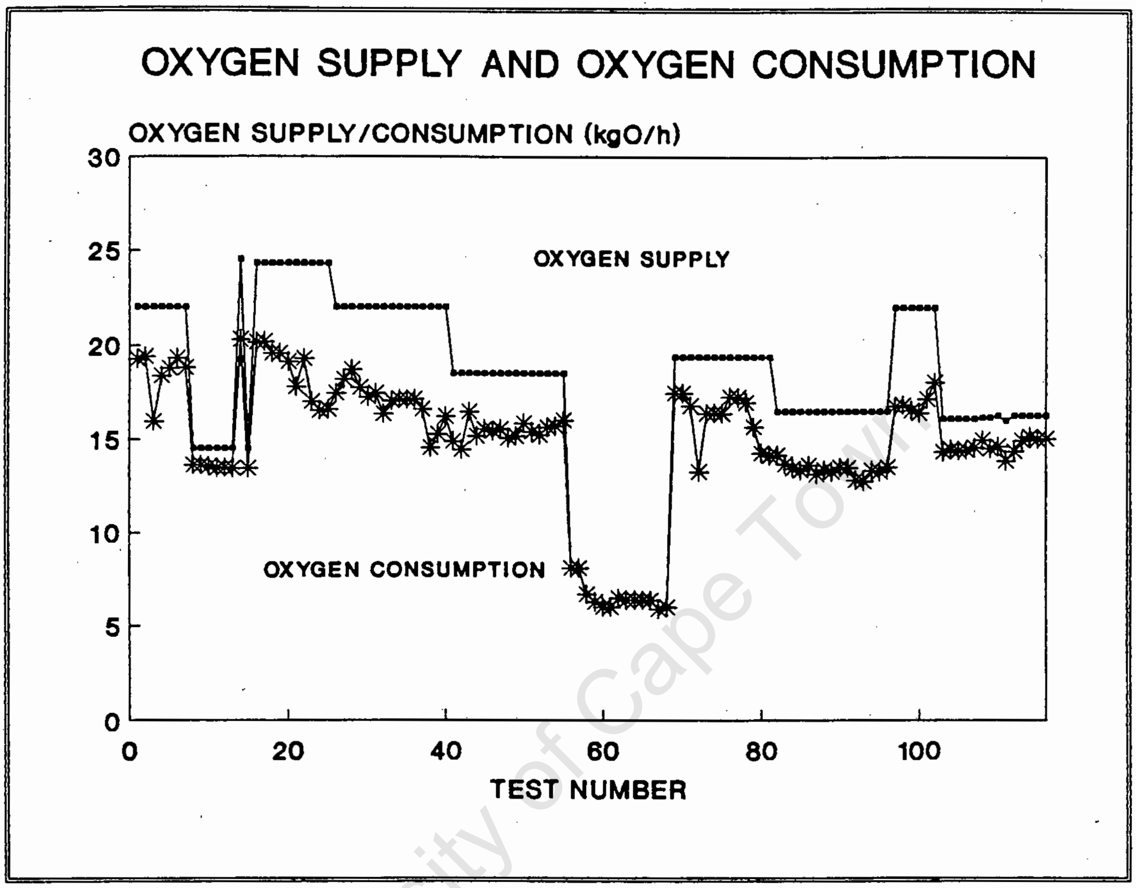
4.2.5 The biological heating rate, H_{bi}

The biological heating rate, H_{bi}, was calculated from the experimentally determined H_{net}, H_{mi}, H_{we} and H_{ve} with the aid of Eq (3.5) and is shown plotted in Fig 4.1 for increasing H_{net} value. It ranged from 85 MJ/h (at 3 day retention time) to 263 MJ/h (at 1,25 day retention time) and follows the same trend as H_{net}. The reason for the similarity in trend with H_{net} is that the mechanical heat input rate heat gain H_{mi} and wall and vent gas heat losses (H_{we} and H_{ve}) remain relatively unchanged at different retention times so that any change in H_{net} has to be accommodated by a change in H_{bi} (Fig 4.1).

4.3 THE BIOLOGICAL OXYGEN CONSUMPTION RATE, O_c

The biological oxygen consumption rate, O_c, was calculated from Eq (3.1) as the difference between the oxygen supply rate, O_s, and the oxygen mass flow rate in the vent gas. The lowest O_c measured was 5.9 kgO/h (at 3 day retention time) and the highest was 20.3 kgO/h (at 1.25 day retention time). It was possible to control O_c through the oxygen supply rate O_s, and this is illustrated in Fig 4.5 where the O_c and O_s values measured in each specific heat yield test are plotted against test number arranged in chronological order. This control was limited however, and a maximum O_c would be reached above which a further increase in oxygen supply would not increase oxygen consumption. This maximum O_c was a steady state value of 17 kgO/h achieved at a 1.25 day retention. The above mentioned highest measured O_c of 20.3 kgO/h was not a steady state value, but was transient and could not be sustained by the sludge for more than 3 days before decreasing gradually to 17 kgO/h (see Section 4.8 below on oxygen limitation).

The oxygen consumption rate, O_c, was constrained to a maximum of 17 kgO/h by biological limitations rather than mechanical transfer limitations. It was measured under reactor operating conditions of (1) a reactor sludge temperature of ±60°C (2) a reactor sludge VS concentration of ±30 g/ℓ and (3) a retention time of 1.25 days. While it was not possible to examine the relationship between the maximum O_c and each of these 3 operating conditions, this maximum O_c did appear to be lower at a 3 day retention time, where a value of 14 kgO/h was deduced (see Section 4.8 below). To distinguish the maximum O_c from O_c values which have been constrained by mechanical oxygen transfer limitations (such as by constraining the oxygen supply), it has been expressed as a volume specific



<u>Fig 4.5</u>: The values of oxygen supply rate (O_s) and oxygen consumption rate (O_c) for the 116 specific heat yield tests versus test number, arranged in chronological order. The plots show O_c to be directly proportional to O_s.

rate termed the biological oxygen utilization rate (OUR). Thus the OUR of the reactor sludge was 0.38 kgO/(m³.h) (≡ 17 kgO/h) at a 1.25 day retention time and 0.31 kg/(m³.h) (≡ 14 kgO/h) at a 3 day retention time.

The transient O_c of 20.3 kgO/h indicated that the Vitox oxygenation system was capable of transferring at least 20.3 kgO/h of oxygen to the sludge mixed liquor. In all likelihood the Vitox system was capable of transferring oxygen at a higher rate, but this could not be ascertained due to the biological limitation to oxygen consumption – i.e. in no specific heat yield test did the reactor sludge consume oxygen at a rate greater than 20.3 kgO/h.

Because 17 kgO/h was a sustainable maximum O_c, it was accepted that at oxygen consumption rates (O_c) of less than 17 kgO/h the reactor was oxygen limited. Oxygen limitation in the reactor is discussed in detail in Section 4.8 below.

4.4 THE SPECIFIC HEAT YIELD, Y_H

In Fig 4.6, the biological heating rate H_{bi} is plotted against oxygen consumption rate O_c and a very well defined proportionality between H_{bi} and O_c is apparent with very little scatter about the best fit line drawn through the plotted points. The constant of proportionality between H_{bi} and O_c is the specific heat yield Y_H, i.e. MJ of heat generated biologically per kg oxygen consumed. For the 116 specific heat yield tests, Y_H ranges from 11,4 to 14,6 MJ/kgO with an average of 12,77 ± 0,58 MJ/kgO. The slope of the best fit line in Fig 4.6 also is the specific heat yield Y_H. This value is 11.81 which is different to the mean of 12,77 MJ/kgO.

4.4.1 Comparison between Y_H and similar values reported in literature

The average Y_H value of 12.77±0.58 MJ/kgO measured at Milnerton is generally lower than average values reported in or deduced from the literature. Seven sets of values from the literature were discussed in Chapter 2, Sections 2.4.4.2, 2.7.3 and 2.7.4, and fall into 4 categories:

(1) thermodynamic heats of oxidation or organics
(2) measured ratios of heats of combustion to COD
(3) measured Y_H values in microbiology
(4) Y_H values deduced from data reported in investigations into autothermal

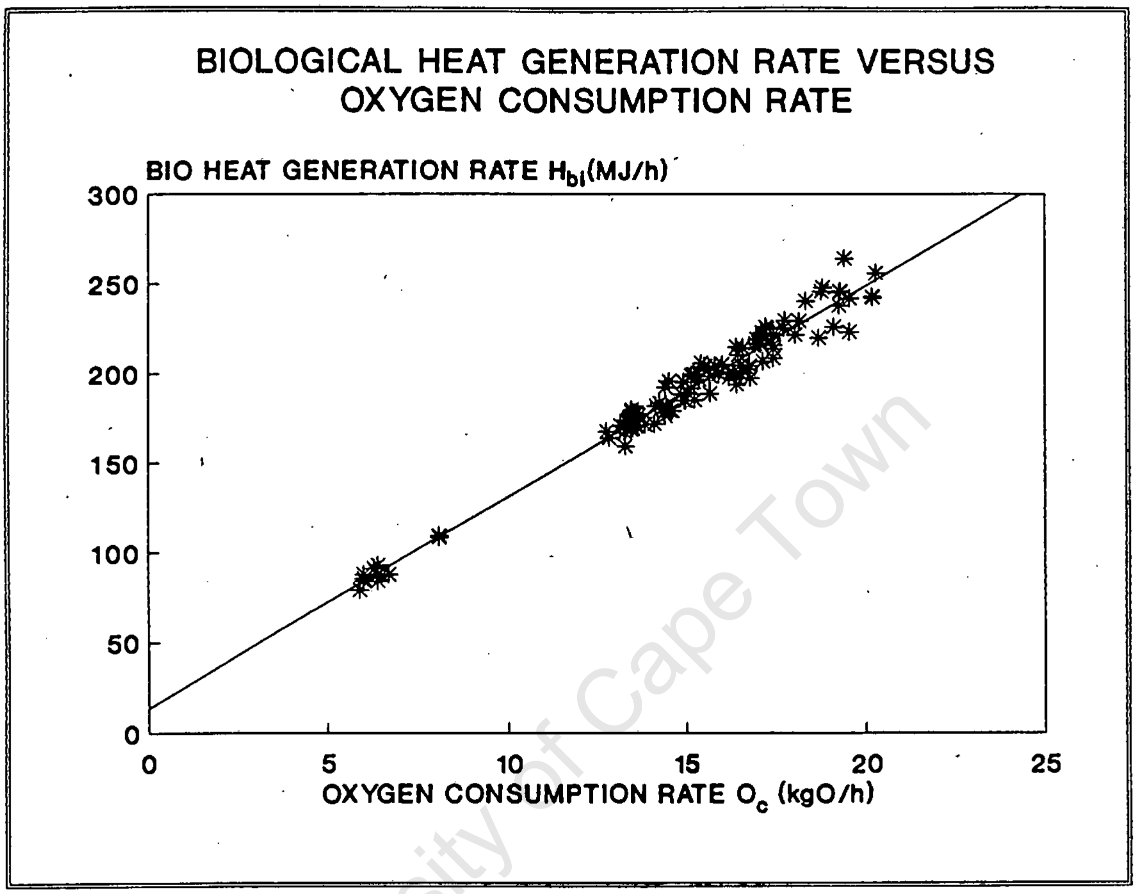


Fig 4.6: Plot of the biological heat generation rate (H_{bi}) versus the oxygen consumption rate (O_c) from the values measured during the 116 specific heat yield tests. Direct proportionality between H_{bi} and O_c is shown with the constant of proportionality being the specific heat yield Y_H (MJ/kgO).

thermophilic aerobic digestion (ATAD).

From thermodynamic data reported by McCarty (1972) an average heat of oxidation of 14.4 MJ/kgO can be deduced for a wide range of organics. Kambhu (1971) determined the mean ratio of the heats of combustion and COD's of thermophilic organisms grown on glucose to be 5860 BTU/lb (13.6 MJ/kg COD) while Wright (1975) observed that this ratio for organic materials ranged from 12.6 to 16.7 MJ/kg COD. The Milnerton average value compares favourably with these 3 sets of data and the fact that they are slightly higher is consistent with the hypothesis in Chapter 2, Section 2.7.1.3 – namely that the Y_H value measured in a biomass is unlikely, subject to certain assumptions, to be higher than the specific energy (MJ/kgO) released during the combustion (or oxidation) of the biomass substrate, as a portion of this energy will be retained in synthesized organism mass.

With the exception of an average value of 14.4 MJ/kgO reported by Cooney *et al.* (1968), less favourable agreement is found between the Milnerton data and other values which have been measured experimentally or deduced from research data. Measurements of Y_H in the field of microbiology show wide variation between individual values. Cumulative values reported for yeast by Cooney *et al.* range from 11.4 to 18.6 with an average of 14.4 MJ/kgO and values for *E.Coli* reported by Rothbaum and Stone (1961) range from 16.3 to 26.7 MJ/kgO. Variations in these measured values are probably the consequence of inadvertent errors in the experimental procedures whereby Y_H was determined. From work published on ATAD by Trim (1984) and Wolinski (1985) both lower and higher values can be deduced; calculations on the reported data of Trim give specific heat yield values ranging from 5.3 to 11.7 MJ/kgO and a heat and oxygen mass balance presented by Wolinski suggests a value of 21 MJ/kgO. In these ATAD investigations, the research was not specifically directed at measuring the specific heat yield. Consequently, although heat and oxygen mass balances were reported, they did not appear to have been constructed with the vigilance and detail which is required to accurately measure Y_H.

The above discussion indicates that not much store can be placed on previously experimentally measured values due to their unreliability. As a consequence, the accuracy of the 12,77 MJ/kgO measured value at Milnerton cannot be tested against these previously measured experimental values. Confidence in the accuracy of the 12,77 MJ/kgO value comes not from comparison with other measured Y_H values, but from 2 other sources, i.e.

(1) the favourable comparison with McCarty's calculated values and Kambhu's (1971) and Wright's (1975) heat of combustion values, and

(2) the thoroughness and uncompromising vigilance and attention to every detail, to the degree that resources and understanding made possible, in the experimental procedure employed to measure the Y_H value at Milnerton. (Details of the experimental method are given in Chapter 3).

In the absence of reliable experimental data on Y_H with which to compare the 12,77 MJ/kgO measured value, the measured Y_H data need to be examined very carefully before these are accepted. This examination took 3 forms:

(1) a statistical evaluation of the Y_H data
(2) the effect of reactor operating conditions such as reactor temperature, retention time and oxygen consumption rate, on the Y_H data, and
(3) the sensitivity of the Y_H data to experimental measurement errors.

Each of these 3 aspects are discussed below.

4.4.2 Statistical evaluation of the Y_H data

A bar histogram showing the frequency of occurrence of a particular measured Y_H value for the 116 Y_H data is given in Fig 4.7. The chart indicates that the Y_H data are not normally distributed, which suggests that the sample may contain more than 1 population distribution. A probability plot of Y_H values (Fig 4.8) confirms this. In order to find an assignable cause for the non-normal distribution of the Y_H data, these were examined to check whether there was a dependence of Y_H on plant operating parameters that might have led to distinct population groupings in the data. Reactor sludge temperature, T_{se}, retention time, R_H, and

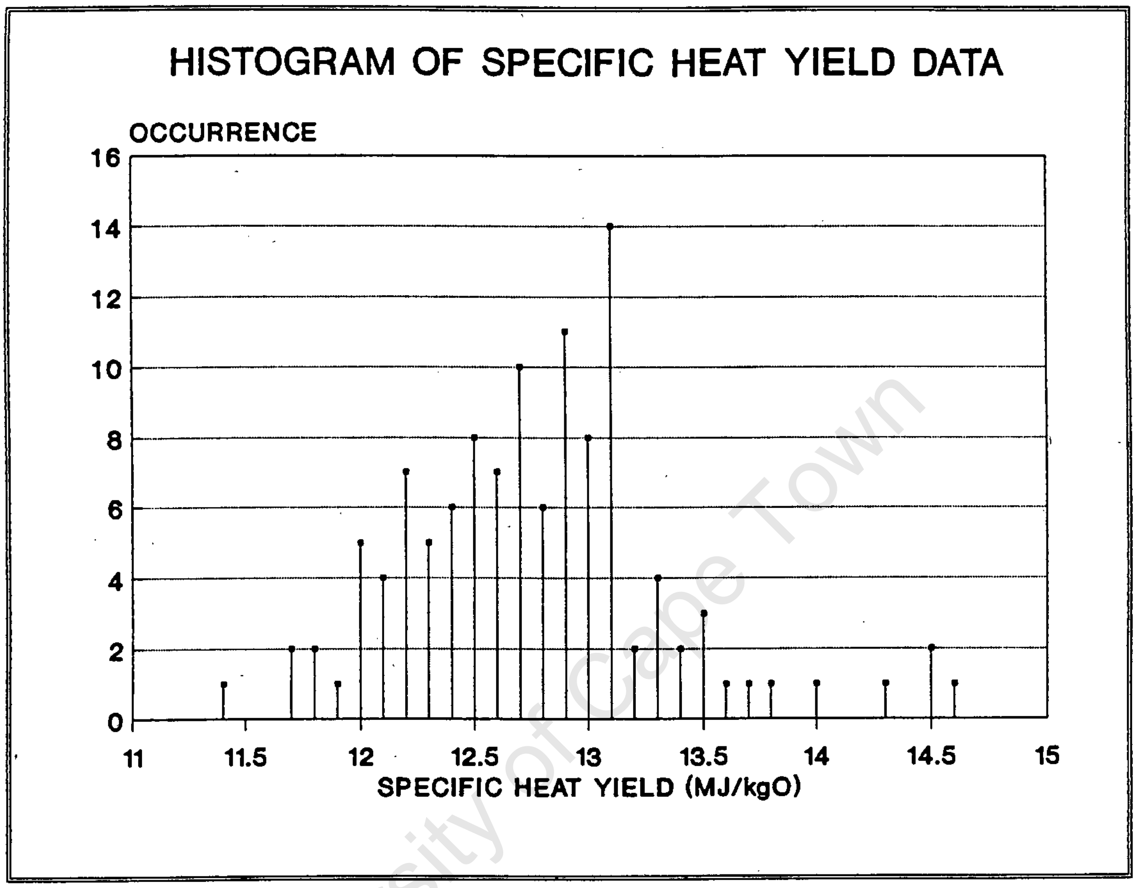


Fig 4.7: A histogram of the 116 specific heat yields (Y_H) which suggests that the data is not normally distributed. The greatest number of Y_H values were recorded at a value of 13.1 MJ/kgO although the arithmetic average of the 116 data is calculated as 12.77 MJ/kgO.

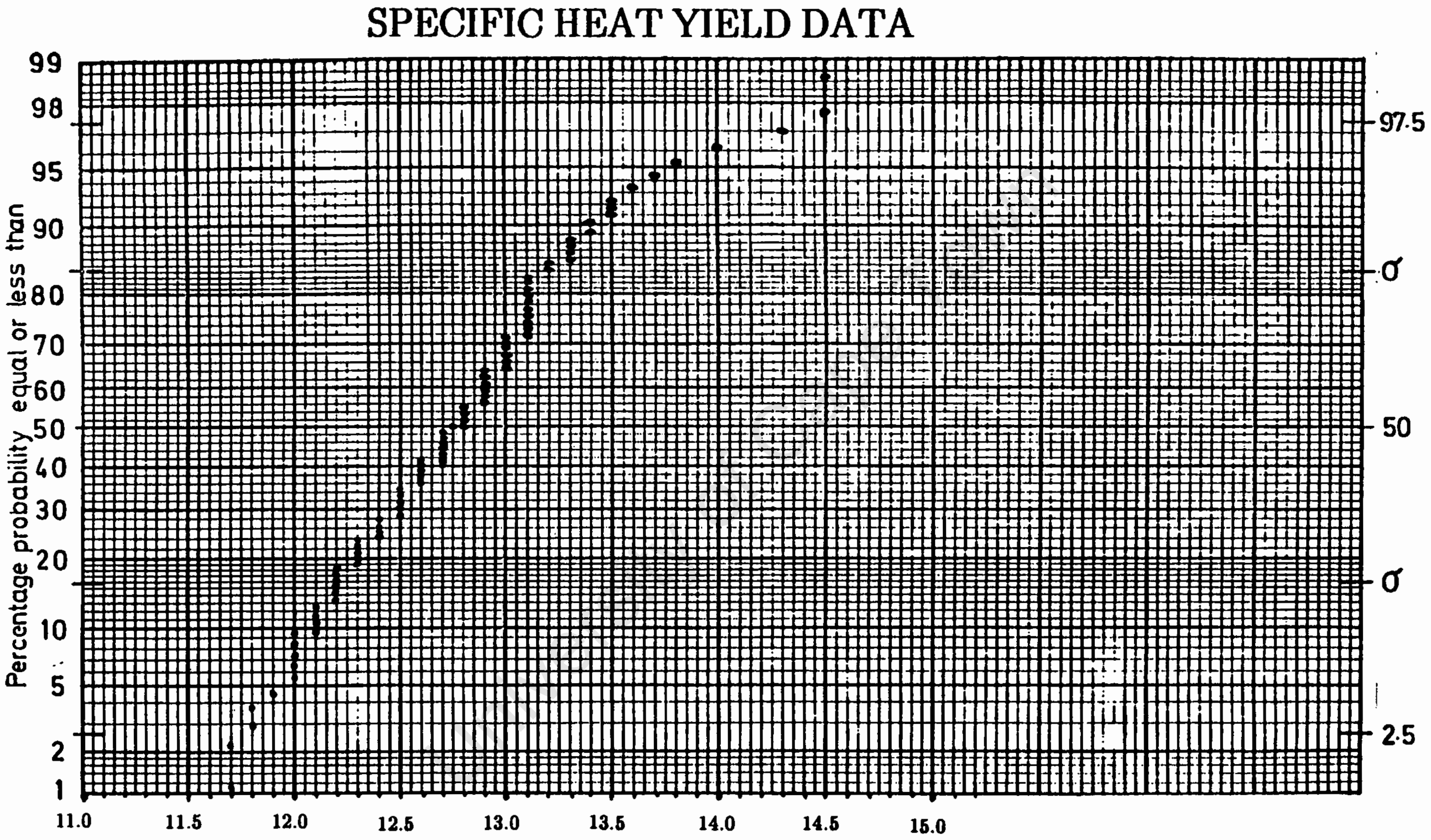


Fig 4.8: A probability plot of the 116 specific heat yield values (Y_H). The data does not lie on a straight line and hence it is concluded that it does not conform to a normal distribution.

oxygen consumption rate, O_c,[2] were the only controlled parameters whose effects on the plant were non-random, and these 3 parameters were accepted as possibly having an effect on the value of Y_H.

4.4.3 Effect of reactor operating conditions on Y_H

4.4.3.1 *Effect of reactor sludge temperature, T_{se} on Y_H*

When ordering the Y_H data with increasing T_{se}, no obvious relationship between the two can be discerned - see Fig 4.9 where the average Y_H for 5 ranges of T_{se} is plotted against the average T_{se} in each group. Also if the 116 Y_H data are sorted into 3 groups i.e. 13 values at high Y_H (average 13,8 MJ/kgO), 4 values at medium Y_H (average 13,15 MJ/kgO) and 99 values at low Y_H (average 12,62 MJ/kgO), as in Table 4.2, then the mean value of the average reactor sludge temperature during the specific heat yield test for the respective groups is high (66,6 °C) at high Y_H, low (53,2°C) at medium Y_H and high (61°C) at low Y_H. Clearly Y_H changes inconsistently with increasing reactor sludge temperature T_{se}, and T_{se} was rejected as a possible cause for the non-normality in the Y_H data.

4.4.3.2 *Effect of retention time R_H on Y_H*

To comply with the temperature specifications for pasteurization, the value of T_{se} was maintained in the range 55 – 65°C. In order to maintain T_{se} in this temperature range, the biological heat generation rate H_{bi} had to be increased (decreased) whenever the retention time in the reactor was decreased (increased). This was accomplished by increasing (decreasing) the oxygen supply rate which increased (decreased) the oxygen consumption rate O_c. Therefore a decrease (increase) in retention time was always accompanied by an increase (decrease) in O_c and the separate influences of O_c and R_H on Y_H could not be determined from the available data. Consequently R_H was not considered as an independent assignable cause for the non-normality in the data, and attention was directed at the oxygen consumption rate O_c as a possible cause.

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[2] From the discussion in Section 4.3 above and from Fig 4.5, it can be seen that O_c was a controlled parameter – control of O_c was effected through control of the oxygen supply rate, O_s.

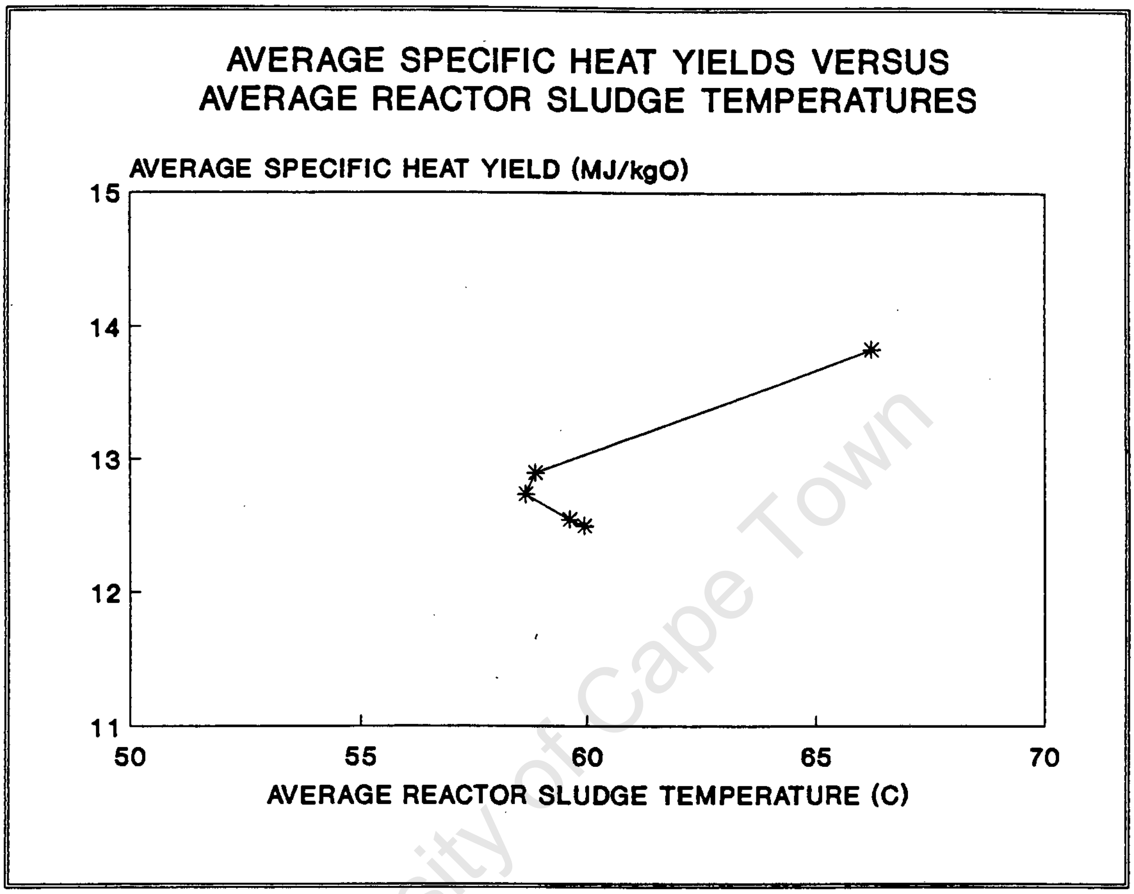

Fig 4.9: The average specific heat yields, Y_H, calculated for 5 ranges of reactor sludge temperature, T_{se}, plotted against the average temperature for each range. The plot suggests that Y_H and T_{se} are not related in any obvious manner.

TABLE 4.2: Specific heat yield, reactor sludge temperature and oxygen consumption.

AVERAGE Y_H (MJ/kgO)	AVERAGE T_{se} (°C)	AVERAGE O_C (kg/h)	NUMBER OF TESTS
13.80	66.6	6.6	13
13.15	53.2	12.5	4
12.62	61.0	15.91	99

TABLE 4.3: The average Specific Heat Yield (Y_H) values for 3 groups of Oxygen Consumption (O_C).

O_C (kg/h)	GROUP 1 5.9 - 8.1	GROUP 2 12.7 - 15.9	GROUP 3 16.0 - 20.3
DATA PTS	13	56	47
MEAN Y_H	13.80 ±0.52	12.75 ±0.36	12.50 ±0.48

4.4.3.3 *Effect of oxygen consumption rate O_c on Y_H*

While it is clear from Table 4.2 that there is no obvious relationship between Y_H and T_{se}, the average Y_H of each group is found to be consistently related to the average oxygen consumption rate O_c of each group; as O_c decreases so Y_H increases. For the purposes of statistical analysis the 116 Y_H data was sorted into 3 groups (Table 4.3) different to the 3 groups in Table 4.2, and once again the trend of increasing Y_H with decreasing O_c is evident. Probability plots were made of each group of Y_H values in Table 4.3 in order to check the nature of their distributions. These plots (see Appendix 4D) indicate that the Y_H values of each group are normally distributed. Statistical F and t-tests (given in Appendix 4D) confirm that at the 95% confidence interval, the means of each group belong to different populations, supporting the observation from Table 4.3 that Y_H showed an inverse dependence on O_c. If groups 2 and 3 are further subdivided to yield a total of 5 groups, then the average Y_H values for each group confirm the inverse relationship, although now F and t-tests show that the means are no longer from independent populations. This probably happens because now the number of data in each of the 5 groups are too few to allow distinction between the means of the groups at the 95% confidence interval. Nevertheless, plotting the average Y_H value versus the average oxygen consumption rate for each of the 5 groups, demonstrates not only the inverse relationship between Y_H and O_c, but also that this relationship is linear (Fig 4.10). This implies that decreasing oxygen consumption rate O_c causes increasing amounts of biological heat to be generated per kgO consumed by the sludge. If O_c is converted to a volume specific value and called the oxygen transfer rate[3] i.e. OTR = O_c/V_p[kgO/(m³.h)] the same conclusion would emerge, i.e. a reduction in OTR causes an increase in Y_H. As O_c (or OTR) decreases, which is imposed on the sludge mass by reducing the oxygen supply rate, the sludge becomes more oxygen limited. Consequently it would appear that the specific heat yield Y_H increases linearly as the sludge becomes more oxygen limited. There are 2 possible causes for this observation:

(1) systematic experimental error in the measurements of the parameters from which Y_H is calculated,

(2) a biological phenomenon.

In examining the sources of experimental error in the parameters from which Y_H is calculated, the oxygen supply rate O_s, from which the oxygen consumption rate was determined, was identified as the most likely source of error. The effect of oxygen supply rotameter calibration on Y_H is discussed below.

4.4.3.3.1 *Effect of oxygen supply rotameter calibration on Y_H*

The group of 13 high Y_H values were obtained at low oxygen consumption rates for which the oxygen supply rate necessarily also needed to be low. The low flow oxygen feed rotameter was used to deliver these low oxygen supply rates. In contrast, the 103 lower Y_H data at higher oxygen consumption rates (> 12 kgO/h) were obtained with the high flow oxygen rotameter. Not only are the Y_H

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[3] As a consequence of the oxygen limitation the oxygen consumed by the sludge is equal to the oxygen transferred to the bulk liquid by the oxygenation system.

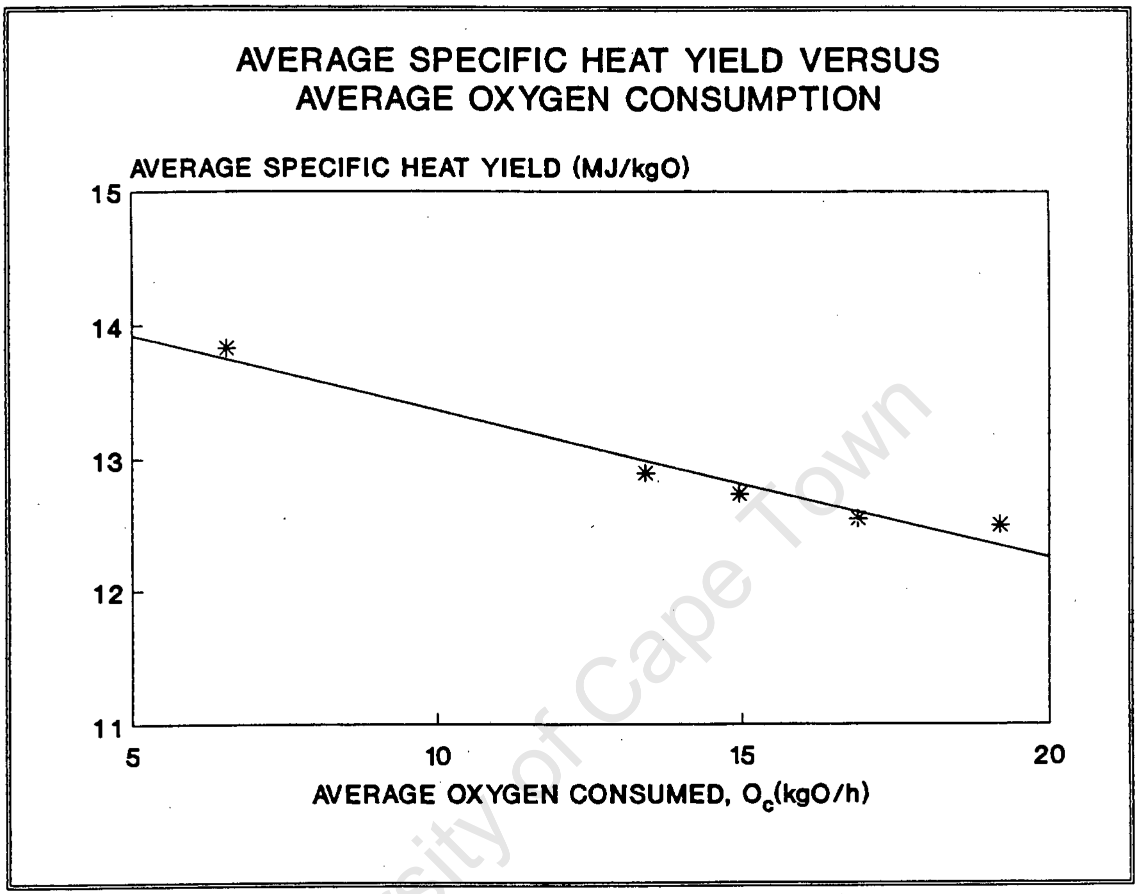

Fig 4.10: The average specific heat yields, Y_H, calculated for 5 ranges of oxygen consumption rate, O_c, plotted against the average oxygen consumption rate for each range. The plot clearly illustrates an inverse proportionality between Y_H and O_c.

values for these 13 specific heat yield tests higher than the others but also the respiration quotient Y_{CO_2} i.e. moles CO_2 generated per mole oxygen consumed (see Section 4.6 below) is higher than the other 103. If the actual oxygen mass flow through the low flow rotameter was 10% more than measured because of a calibration error, then the average Y_H value for the group of 13 data at the low oxygen flow would decrease from 13,8 (see Table 4.3) to 12,6 MJ/kgO. Also the average Y_{CO_2} would decrease from an above average 0,78 value to the average value of 0,66 (see Section 4.6 below). Consequently from the higher values of Y_H and Y_{CO_2} for these 13 data, it could be asserted that the low flow oxygen rotameter was not accurately calibrated. However, two factors lead to rejection of this assertion –

(1) Even if the 13 tests in question are omitted from the data set, the trend of increasing Y_H with decreasing O_c is still evident in the remaining 103 tests for which the oxygen supply was measured with the high flow rotameter only.

(2) A high degree of precision could be achieved in the calibration of the low flow rotameter because at low flows, a much longer time is required to dispense the fixed masses of oxygen through the rotameter (see Appendix 3A for calibration of rotameters).

By rejecting the assertion that the high Y_H values were obtained as a result of poor calibration of the low flow oxygen supply rotameter, it is accepted that the increase in Y_H with decreasing O_c may be attributed to the 2nd of the 2 possible causes listed above, viz. that it is a biological phenomenon caused by increasing oxygen limitation in the reactor.

From Chapter 2, Section 2.7.1.3, the possibility of Y_H decreasing with increasing oxygen limitation is apparent; under the stressful condition of oxygen starvation, the organisms appear to expend a greater proportion of the energy available in the substrate, on organism maintenance requirements, which ultimately leads to heat generation, rather than organism mass synthesis, which remains as organism mass and therefore does not generate heat.

4.4.3.3.2 *Effect of oxygen limitation on Y_H*

As mentioned earlier in Section 4.3 above, a sustainable maximum oxygen consumption rate $O_{c\ max}$ of 17 kgO/h was observed. This gives a biological oxygen utilization rate (OUR) of 0.38 kg/(m³.h) which was achieved at an average reactor VS concentration of about 30 kgVS/m³.[4] At oxygen consumption rates O_c less than 17 kgO/h, the sludge mass would be oxygen limited, with the severest oxygen limitation occurring at the lowest oxygen consumption rate of around 6 kgO/h. If the degree of oxygen limitation O_L is defined as the ratio of the actual sludge volume specific oxygen consumption rate OTR and the OUR i.e. O_L = OTR/OUR, then because of the linear relationship between Y_H and O_c (Fig 4.11), the specific heat yield Y_H can be described as

$$Y_H = Y_{HM} + (Y_{HO} - Y_{HM})\, O_L$$

where

O_L = (1 - OTR/OUR) = oxygen limitation

Y_{HM} = Y_H at OUR [MJ/(kgO.m³)]

Y_{HO} = hypothetical Y_H value at zero OTR

4.4.4 The sensitivity of the specific heat yield, Y_H, to experimental error

In order to gain confidence in the specific heat yield values measured on the aerobic reactor, an analysis was undertaken of the sensitivity of the calculated Y_H values to experimental error. Eight parameters were measured in the heat and oxygen mass balances in order to calculate Y_H (see Section 4.1 above). Of these 8 parameters the 6 listed below, as well as the reactor operating volume, V_p, were examined individually to determine how the Y_H value changes with errors in their measurement. The 7 parameters are:

(1) The flow rate of partially dried vent gas passing through the gas meter, V_{gm}, the first of the 3 parameters from which *inter alia*, the oxygen consumption rate O_c was calculated.

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[4] The temptation of converting the OUR to a VS concentration specific rate should be resisted because the effect of VS concentration on OUR is uncertain. It may be thought that doubling the reactor VS concentration would double OUR, but this may not necessarily be so.

(2) The fraction of oxygen in the dry vent gas (f_{O_2}) – the second measurement required for determining O_c.

(3) The oxygen supply rate (O_s) – the third measurement required for determining O_c.

(4) The reactor sludge temperature increase rate ($\Delta T_{se}/\Delta t$) – one of two measurements from which the rate of reactor sludge enthalpy H_{net} was determined.

(5) The operating volume of the reactor V_p – the second measurement required for determining H_{net}.

(6) The ambient temperature, T_{amb} – from which the wall heat losses, H_{we}, were calculated.

(7) The current drawn by the Vitox recirculation pump (I) from which the mechanical heat input rate H_{mi} was determined.

The measurements for the determination of the vent gas water vapour heat loss rate H_{ve} were neglected in this sensitivity analysis because H_{ve} makes only a very small contribution to the heat sinks (typically < 10%) with the result that errors in these measurements would have a negligible influence on the specific heat yield determination.

The procedure employed in the sensitivity analysis is summarized as follows (details of the results are given in Table 4.4): Taking one of the 7 measurements listed above, a step increment (and decrement) was made to its actual measured value in one of the 116 specific heat yield tests. The magnitude of the increment (and decrement) was selected on the basis of the magnitude of the error that could be made in its measurement; for example with the reactor sludge temperature, this was recorded to one decimal place of a degree, so that the increment given to this measurement was 0.1°C. Incrementing only the particular measurement in question, and leaving all the others at the recorded value, the specific heat yield was recalculated. By comparing the recalculated Y_H value with the original value, the sensitivity, or gearing, of experimental measurement error on the Y_H

value could be established. This procedure was applied to all 7 of the listed measured parameters, both by incrementing and by decrementing the measured value by the error that could be made in its measurement.

The results of the sensitivity analysis show that of the 7 parameters, errors in the measurement of 4 of these, i.e. the ambient temperature (T_{amb}), % oxygen in the dry vent gas (f_{O_2}), vent gas flow rate (V_{gm}) and recirculation pump current drawn (I) have a negligible effect on the Y_H value whereas errors in the measurement of the remaining 3, i.e. the oxygen supply rate O_s, the reactor sludge heating rate dT_{se}/dt and operating volume of the reactor V_p have a significant effect on Y_H. For example, a 34% error in T_{amb}, which causes an incorrect prediction of the wall heat loss H_{we} by Eq (3.10) results in only a 1,6% error in Y_H; it was for this reason, as mentioned in Section 4.2.3 above (see also Appendix 3D) that a more complex equation for estimating H_{we} was not required. In contrast, if O_s, $\Delta T_{se}/dt$ or V_p is over- or underestimated by as little as 2,7%, an error of more than 3% would result in Y_H. It was for this reason that great care was taken in determining these three parameters; the first through careful calibration of the oxygen supply rotameters, the second through calculating as accurately as possible the sludge heating rate by different methods from not only 1, but 2 temperature probes, and the third through careful calculation from the dimensions on the drawings of the mould on which the reactor was cast and checking by measuring the internal dimensions of the reactor itself.

From the sensitivity analysis it is clear that only errors in O_s, dT_{se}/dt and V_p would have a significant influence on the calculated value of Y_H so that if the assignable cause of the non-normality of the Y_H data does not lie in oxygen limitation but in experimental error, then it is in the errors in these three parameters that this assignable cause is likely to be found. Accepting that both oxygen supply rotameters were accurately calibrated (the reason for this was already discussed in Section 4.4.3.3.1 above and Appendix 3A) then errors in O_s would be random and these would not affect the averages of the Y_H values of the 3 groups of Y_H data in Tables 4.2 and 4.3. Therefore even random errors in O_s could not cause an increasing Y_H with decreasing O_c. If the Honeywell temperature recorder did not give the correct absolute temperature of the sludge, this would not significantly affect the sludge heating rate dT_{se}/dt, where this rate is measured over a fairly narrow temperature range of 3°C. Therefore the

TABLE 4.4: Sensitivity of Y_H to experimental error.

Data	Data Value	Specific Heat Yield	% Error in Data	%Error in Y_H	Ratio of % errors
T_{amb}	10.0	13.1	-34.6	1.6	0.04
	15.3	12.9	----		
	20.0	12.7	30.7	-1.6	0.05
m^3/h vent	10.00	12.8	-4.9	-0.8	0.16
	10.52	12.9	----		
	11.00	13.0	4.6	0.8	0.17
f_{O2}	17	12.6	-15.0	-2.3	0.16
	20	12.9	----		
	23	13.2	15.0	2.3	0.16
O_s	18.0	13.3	-2.7	3.1	1.15
	18.5	12.9	----		
	19.0	12.5	2.7	-3.1	1.15
dT_{se}/dt	1.2750	12.5	-2.9	-3.1	1.09
	1.3125	12.9	----		
	1.3500	13.4	2.9	3.9	1.36
V_p	43	12.2	-4.4	-5.4	1.23
	45	12.9	----		
	47	13.6	4.4	5.4	1.23
Amps	40	13.3	-9.1	3.1	0.34
	44	12.9	----		
	48	12.5	9.1	-3.1	0.34

non-normality of the Y_H data is not likely to lie in incorrect reactor sludge temperature measurement. Regarding the accuracy of the reactor operating volume V_p, this is completely dependent on the measured dimensions of the reactor, and a constant 45 m³ was accepted for all 116 specific heat yield tests. If this volume was incorrect, the error in Y_H would be systematic, resulting in all the Y_H values being either too high or too low relative to the true value.

4.4.5 Conclusion regarding the specific heat yield Y_H

From the above discussion on the possible errors in the experimentally determined specific heat yield Y_H values, it can be concluded that

- the individual Y_H values determined from the 116 specific heat yield tests are reliable and accurate owing to the precision and care with which the experimental measurements were taken,

- the linear increase in Y_H with increasing oxygen limitation is most likely real and not an artifact arising out of errors in experimental measurement.

4.5 OXYGEN UTILIZATION EFFICIENCY (O_{EFF})

The oxygen utilization efficiency (O_{eff}) is given by the oxygen consumption rate (O_c) as a ratio of the oxygen supply rate (O_s) i.e. $O_{eff} = O_c/O_s$. For the 116 specific heat yield tests, O_{eff} ranged from 66 to 100%. Examination of the data shows that O_{eff} decreased as O_s increased. This is illustrated in Fig 4.11 where the average values of O_c and O_{eff} are plotted against O_s for the 7 different O_s values ranging from 6 to 24,3 kgO/h, at which the aerobic reactor was operated during the investigation. The inverse proportionality between O_{eff} and O_s shown in Fig 4.11 probably results from the Vitox oxygenation system transferring oxygen less efficiently as O_s increases. However, at the high oxygen supply rates of 24 kgO/h, the efficiency of oxygen utilization in the reactor was more likely limited by the biological oxygen utilization rate (OUR) of the sludge than by the oxygen transfer ability of the Vitox system.

4.6 THE RESPIRATION QUOTIENT (Y_{CO_2})

The respiration quotient, Y_{CO_2}, is the number of moles of carbon dioxide produced per mole of oxygen consumed by the reactor sludge. The results from the 116 specific heat yield tests performed in this investigation showed that Y_{CO_2} varied from 0.53 to 0.85 with an average of 0.66 ± 0.04. Examination of the data indicates that at low oxygen consumption rates O_c of 6 to 8 kgO/h, the values of Y_{CO_2} are higher than the values measured at higher O_c; the average for the 13 data at low O_c is 0.72 ±0.02. However after scrutinizing the data, neither oxygen consumption rates through the oxygen supply rate (see Section 4.4.3.3.1 above) nor any other parameter could be shown to systematically influence the values of Y_{CO_2} and accordingly, the data was examined to see if it conformed to a normal distribution.

In analyzing the distribution of the data, 6 of the 116 values were discarded, 4 at the bottom end and 2 at the top end, i.e. 0.53, 0.57, 0.58, 0.59, 0.81, 0.85. The 4 at the bottom end were discarded on experimental grounds, and the 2 at the top end on statistical, grounds. The reason for rejecting the 4 bottom end values was that during the investigation, 9 step changes in oxygen supply rate were made, 6 downward and 3 upward - see Fig 4.5 in which O_s and O_c are plotted in chronological order for the 116 specific heat yield tests. (Data listed in Appendix 4A). In 4 of these instances (3 down, 1 up) the value of Y_{CO_2} dropped by more than 0.05 to below 0.6 immediately following the change, recovering to higher values the next day. No explanation can be advanced as to why these low Y_{CO_2} values should have occurred. With regard to the two high Y_{CO_2} values, 0.81 and 0.85, these did not occur at any of the 9 changes. Accepting the Y_{CO_2} data, excluding the 4 low values, are normally distributed, using statistical methods outlined by Laubscher the two high values were found to be outliers in the data set. The remaining 112 data points are plotted in a bar chart of Y_{CO_2} values against frequency of occurrence in Fig 4.12. A probability plot of the 112 data in Appendix 4D shows the distribution to be normal.

Theoretical values of Y_{CO_2} may be calculated by considering typical formulations of sludge elemental composition reported in the literature, viz. $C_5H_7O_2N$, $C_4H_9O_3N$, $C_7H_{10}O_3N$ and $C_5H_8O_2N$. The complete oxidation of these compounds to CO_2, H_2O and NH_3 by stoichiometric amounts of oxygen, yields respiration quotients ranging from 0.95 to 1. There are a number of possible

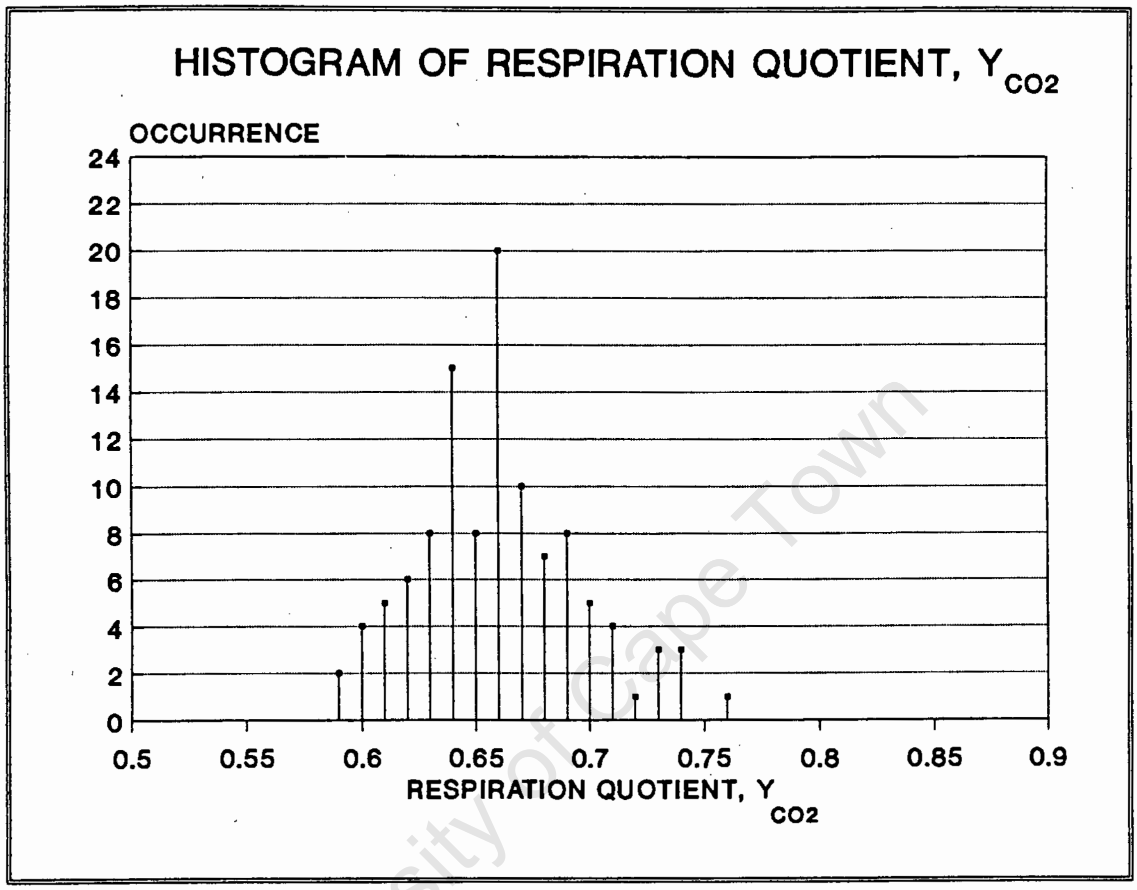

Fig 4.12: A histogram of the 116 measured values of the respiration quotient, Y_{CO_2}.

reasons why the values of Y_{CO_2} measured at Milnerton are smaller than these theoretical values:

(1) The composition of oxidized substrate differs from those formulations listed above.

(2) Some of the CO_2 produced remained in the sludge liquid in dissolved form as a result of the increase in $H_2CO_3^*$ alk in the reactor through the ammonification of organic nitogen in the sludge and the high partial pressure of CO_2 (0.8 atm) in the reactor headspace.

The probability of (2) being the cause of this observation is examined in Appendix 4E where it is shown that only a relatively small amount of CO_2 produced by sludge respiration remains dissolved in the sludge liquor. Dissolution of this small amount of CO_2 occurs rapidly, at the start of a batch cycle, immediately following a batch feed. Thus by the time vent gas volume measurement commences, dissolution will be complete and the measured flow rate of CO_2 in the vent gas will be equal to the rate of CO_2 generation by the sludge. Thus the dissolution of CO_2 in the sludge liquor will not have been the cause of the measured Y_{CO_2} being lower than expected stoichiometric theoretical values.

4.7 DEGREE OF WATER VAPOUR SATURATION

The observed proportionality between vent gas vapour heat losses, H_{ve}, and vent gas temperature, T_{ge}, (Fig 4.3) conforms to theoretical predictions that the mass of water vapourised by a gas increases as the gas temperature increases. It is the elevation of water vapour pressure with temperature that causes more water to be vapourised at higher values of T_{ge}. The relationship between water vapour pressure and temperature can be described using the Antoine equation.

In Appendix 3B equations have been derived which predict the water vapour mass flow rate (M_{wp}) in the vent gas if it were saturated with water vapour at the vent gas temperature, T_{ge}. For the 116 specific heat yield tests the predicted saturated water vapour mass flow rate M_{wp} was calculated with the aid of these equations and plotted against the measured total water vapour mass flow rate (M_{we}) in Fig 4.13. The predicted values show good agreement but are slightly higher than the measured ones, indicating that the vent gas is close to saturation. The small overprediction could be a consequence of the vent gas not being

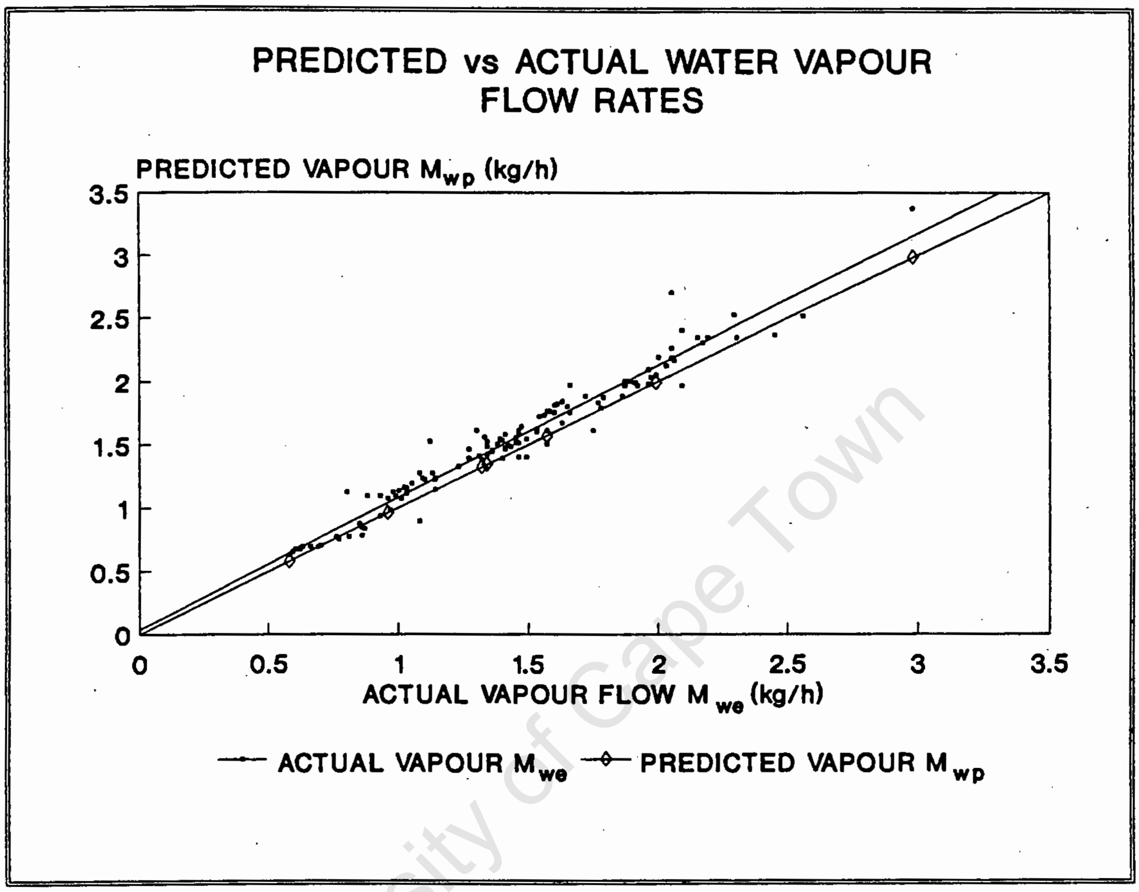

Fig 4.13: The values of the mass flow rate of water vapour in the aerobic reactor vent gas, M_{we}, measured during the 116 specific heat yield tests, plotted against the vent gas water vapour mass flow rates predicted for each test, M_{wp}. The plot shows good agreement between measured and predicted vapour flow rates although the former are slightly lower than the latter.

saturated with water vapour, but this is unlikely due to the relatively low gas flows and the intimate mixing between gas and sludge achieved in the reactor. A more probable cause for the slight undersaturation is the presence of dissolved salts in the sludge which would depress the vapour pressure in accordance with Raoult's law. Applying an Antoine equation for pure water to a water/sludge/salt system will overestimate the vapour pressures occurring in the system. Taking this phenomenon into consideration it is reasonable to conclude from the data that the vent gas was saturated with water vapour at the vent gas temperature, T_{ge}.

4.8. EVALUATION OF OXYGEN LIMITATION IN THE AEROBIC REACTOR

In the course of the investigation it was found that while the reactor was oxygen limited, controlling oxygen supply rate allowed direct control over the reactor sludge temperature. In general, the oxygen supply rate (O_s) was selected at different retention times to ensure that reactor sludge temperature remained at around 60°C to comply with pasteurization specifications and not on the basis of a planned experimental program. However, in three instances, the oxygen supply rate, O_s, was deliberately increased in order to test the effect of a step increase in O_s on O_c, (see Fig 4.14a, b and c). These step increases were useful in examining the extent of oxygen limitation in the aerobic reactor and in testing the rapidity of changes in O_c to changes in O_s. Should the response be sufficiently rapid, then, while the reactor is oxygen limited, control of the oxygen supply rate (O_s) would be an extremely convenient means for controlling the reactor sludge temperature. because the biological heating rate H_{bi} is directly related to the oxygen consumption rate (O_c) through the specific heat yield (Y_H). Step decreases were not examined because, owing to oxygen limitation, this would have an instantaneous effect on O_c and hence reactor sludge temperature. The step increases in O_s were made during the course of conducting the specific heat yield tests. Hence each measurement of O_c mentioned in this discussion on oxygen limitation was made during one of the 116 specific heat yield tests.

The three cases in which the oxygen supply rate (O_s) was suddenly increased are (see Fig 4.14):

(1) from 14.5 to 24.3 kgO/h. The retention was 1.25 days at both rates.

(2) from 6.5 to 19.4 kgO/h. The retention was simultaneously reduced from 3 to 1.2 days.

(3) from 16.5 to 22.0 kgO/h. The retention was 1.25 days at both rates.

The O_s and O_c measured over 1 to 2 days before and 6 to 20 days after the step increase in O_s for the 3 tests are shown plotted in Figs 4.14a, b and c respectively. To test the rapidity with which O_c responded to the step increase in O_s, O_c and O_s were measured in specific heat yield tests in the batch cycles immediately preceding and immediately succeeding the step change: The step increase in O_s was made at the start of phase 3 of the batch cycle, i.e. immediately after sludge feeding to the reactor was completed. The oxygen consumption rate O_c was measured at the end of phase 3 in the batch cycles immediately preceding and succeeding the O_s step change. The reason that the earliest O_c after the step change in O_s could only be measured at the end of phase 3 was that for representative oxygen fraction in the vent gas (f_{O_2}) measurements, a recovery period is required. This is because during the drawing (transfer phase 1) phase of the batch cycle, air is sucked into the reactor, and during filling (feeding phase 2), this air, mixed with gas generated, is expelled from the reactor (see Chapter 3, Section 3.4.1.2). The length of the recovery period depends on the vent gas flow rate and hence on the oxygen supply rate. This constraint on the O_c measurement means that the earliest a response in O_c to O_s could be detected was 2h after the step increase in O_s. However, because of the direct link between the rate of sludge temperature increase dT_{se}/dt and O_c [through H_{bi} and Y_H; see Eqs (3.5) and (3.6)], if dT_{se}/dt is linear throughout the heating period (phase 3) after the step increase in O_s, then O_c is likely to have been constant throughout the heating period at the value measured at the end of the heating period.

For each of the three cases, the earliest O_c measurement after the step increase in O_s, showed that O_c had increased in response to O_s. In test 3 (Fig 4.14c), where O_s was increased from 16,5 to 22,0 kgO/h (33% increase), the new value of O_c was sustained over the 6 days that followed. In test 1 (Fig 4.14a) where O_s was increased from 14,5 to 24,5 kgO/h (69% increase), O_c immediately increased from 14.2 to 20.3 kgO/h but thereafter progressively decreased to 17,0 kgO/h over the subsequent 11 days. Apparently the O_c of 20.3 kgO/h was a transient value and

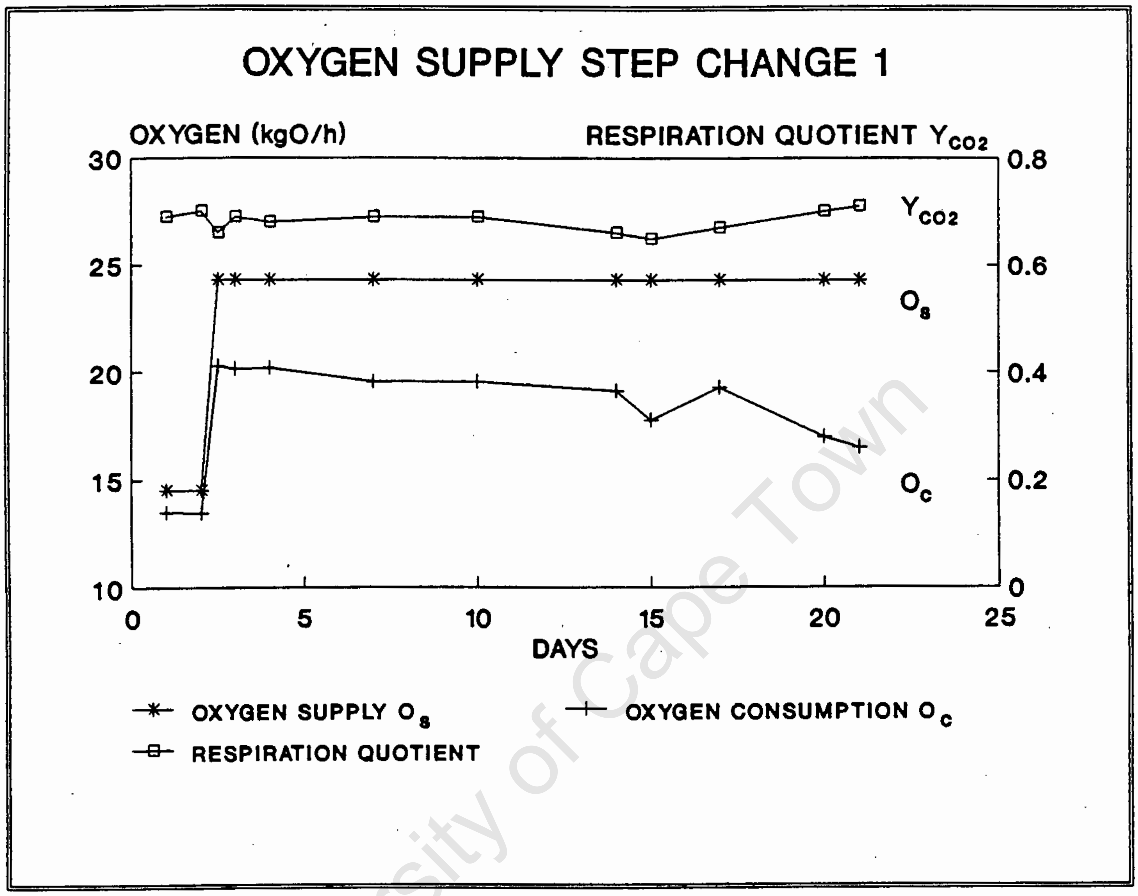

Fig 4.14(a): The change, with time, in the oxygen consumption rate, O_c, and the respiration quotient, Y_{CO_2}, in the aerobic reactor following a step increase in oxygen supply rate, O_s, from 14.5 to 24.3 kgO/h at a constant hydraulic retention time, R_H.

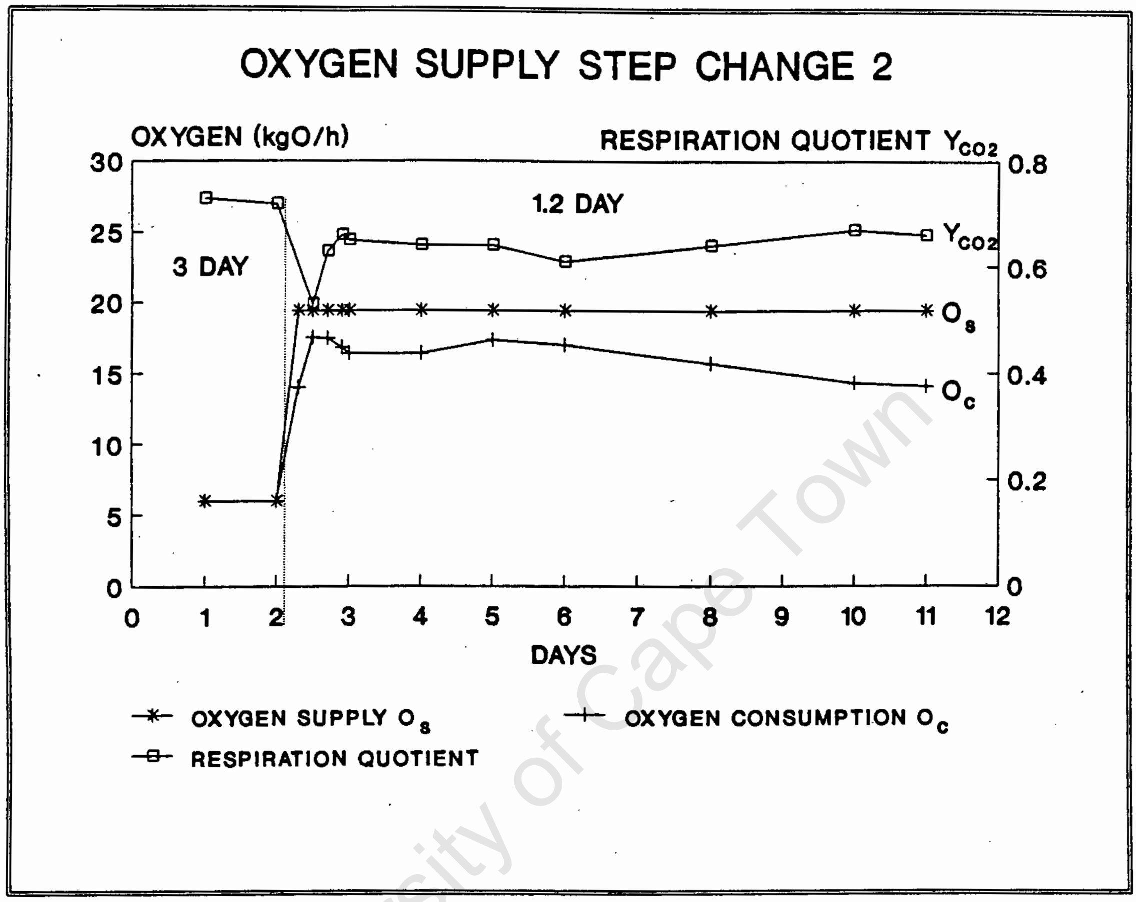

Fig 4.14(b): The change, with time, in the oxygen consumption rate, O_c, and the respiration quotient, Y_{CO_2}, in the aerobic reactor following a step increase in oxygen supply rate, O_s, from 6.0 to 19.4 kgO/h with a simultaneous decrease in hydraulic retention time, R_H, from 3.0 to 1.2 days.

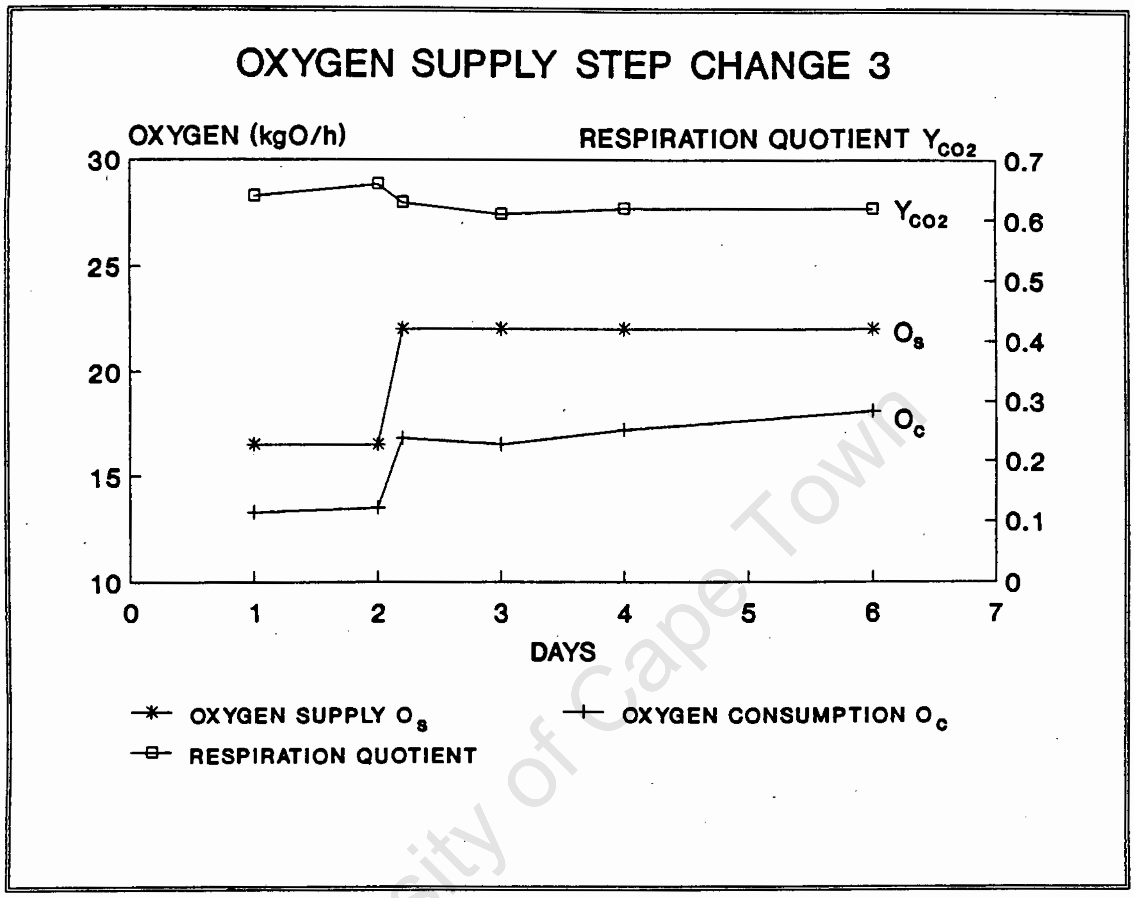

Fig 4.14(c): The change, with time, in the oxygen consumption rate, O_c, and the respiration quotient, Y_{CO_2}, in the aerobic reactor following a step increase in oxygen supply rate, O_s, from 16.5 to 22.0 kgO/h at a constant hydraulic retention time, R_H.

the sludge was unable to maintain this high oxygen consumption rate. In test 2, which involved the largest step increase in O_s from 6.5 to 19.4 kgO/h (198%), with a simultaneous decrease in retention time from 3 to 1.25 days, O_c increased initially to about 14 kgO/h[5], 2h after, and then to 17.4 kgO/h, 4h after, the step increase in O_s. The second O_c value of 17 kgO/h was maintained for about 4 days before decreasing slowly to 15 kgO/h after 9 days. Therefore, whereas the step increase in O_s for tests 1 and 3 caused O_c to increase immediately (i.e. within 2h) to the highest new level value, in test 2, O_c increased to the highest new level value only after 4h. This second test is also the test in which the largest step increase in O_s was made. Because of the large step increase in O_s it is possible that this slower response of O_c was caused by the need for growth of new biomass in response to the sudden reduction in retention time and increased availability of oxygen; the active biomass concentration in the sludge would probably have been significantly lower at 3 days retention time operation with O_s at 6 kgO/h than at 1.2 days retention operation with O_s at 17 kgO/h.

From the 3 tests discussed above it would appear that

(1) The maximum biological oxygen consumption rate of the sludge was approximately 17.0 kgO/h, i.e. the reactor sludge OUR was 0.377 kgO/(m^3.h). At O_c values less than this the sludge would be oxygen limited. In terms of this definition of oxygen limitation, 74% of the specific heat yield tests were conducted under oxygen limiting conditions; 86 of the 116 tests gave O_c values less than 17 kgO/h. This definition was applied to qualitatively describe the extent of oxygen limitation in the reactor. This aspect was described in Section 4.4.3.3.2 above wherein it was shown that the specific heat yield Y_H increases linearly with increasing oxygen limitation. However, it should not be overlooked that it is possible that the

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[5] Note that the value of 14 kgO/h had to be deduced from the rate of sludge temperature increase, dT_{se}/dt, as sludge foaming prevented it from being measured. The reactor sludge did not immediately acclimatize to the step increase in O_s from 6 to 19.4 kgO/h and foamed slightly for about 1h following the increase. The foam prevented vent gas measurement as it would have fouled the vent gas meter. Foaming in the reactor was not a usual occurrence and was only observed at reactor start up, and on this particular O_s step increase test, in which the retention time was simultaneously reduced with the increase in O_s. Foaming is discussed further in Appendix 5B.

maximum biological oxygen consumption rate of the sludge, i.e. the value to which it would immediately increase if sufficient oxygen were available, is lower at the 3 day retention time than at the 1.25 day retention time because the sludge substrate loading rate (kgVS/h) at the former retention time was lower than at the latter retention time. It was not possible to experimentally determine the biological oxygen consumption rate for the 3 day retention but it is estimated to be about 14 kgO/h from the results of the second O_s step increase test when O_s was increased from 6.5 to 19.4 kgO/h - the O_c of 14 kgO/h could not be measured directly in the 2h heating phase immediately after the step increase due to foaming, but was estimated from the sludge heating rate dT_{se}/dt.

(2) The rapidity of response of O_c, and hence of biological heat generation rate H_{bi} through Y_H, to increases in O_s, makes O_s a very convenient and virtually instantaneous control parameter for the reactor sludge temperature, T_{se}. Step decreases in O_s would naturally through oxygen limitation cause an instantaneous reduction in O_c.

4.9 THE STEADY STATE HEAT BALANCE ACROSS THE AEROBIC REACTOR

At any given moment in time, the aerobic reactor is not at steady state with regard to its heat sources and sinks due to the system of batch feeding. The moment by moment imbalance in the heat sources and sinks results in the saw tooth reactor sludge temperature profile (Fig 3.4). From this unsteady state condition, an unsteady state heat balance was set up from which the specific heat yield Y_H, that is, the MJ biological heat generated per kgO consumed, was determined. With the specific heat yield known, the unsteady state heat balance can form the basis of a reactor sludge temperature simulation model. This aspect is presented in Chapter 6. However, for design of the aerobic reactor, the unsteady heat balance is unnecessarily cumbersome; a steady state heat balance is much simpler and sufficient. In this section, the steady state heat balance over the aerobic reactor is discussed: Its application to design of aerobic reactors in dual digestion in general, which are oxygenated with pure oxygen or air, is presented in Chapter 5.

For the batch fed reactor, a steady state operation can only be determined over a complete batch cycle. The aerobic reactor was judged to be at steady state when

(i) the reactor sludge temperature at the start of a number of consecutive batch cycles (phase 1 of Fig 3.4) was the same, and

(ii) the plant operating parameters had not been changed for at least 3 retention times (sludge ages).

Of the 116 unsteady state heat balances that were conducted, 65 were done while the reactor was at a steady state as defined by the 2 conditions above.

For a steady state heat balance to be constructed across a reactor, all of the heat balance parameters, such as the reactor sludge temperature, T_{se}, must remain constant over a defined time interval. While this condition may exist in a continuously fed reactor, it clearly cannot exist in a batch fed reactor where T_{se}, and those heat losses which are functions of T_{se}, change continuously. However a steady state heat balance across a batch fed reactor which is at a steady state as defined above, can be approximated by assuming continuous feed and by accepting that the steady state value of T_{se} is equal to the value of T_{se} in the effluent reactor sludge (i.e. the sludge transferred from the reactor during phase 1), which is the highest reactor sludge temperature attained during the batch cycle. The steady state heat balance is given by equating the heat sources with the heat sinks, i.e.

$$\text{Heat sources} = \text{Heat sinks}$$

$$H_{bi} + H_{mi} + H_{si} = H_{se} + H_{ve} + H_{we} \quad \text{(MJ/d)} \tag{3.4}$$

where

H_{bi} = biological heating rate (MJ/d)

H_{mi} = mechanical heat input rate (MJ/d)

H_{si} = feed sludge heat input rate (MJ/d)

H_{se} = effluent reactor sludge heat loss rate (MJ/d)

H_{ve} = vent gas water vapour heat loss rate (MJ/d)

H_{we} = wall heat loss rate (MJ/d)

Now for the aerobic reactor oxygenated with pure oxygen, it is reasonable to accept that the influent and effluent sludge mixed liquor specific heats and volumetric flow rates are the same. This is because (1) the volatile or total solids suspended in the water contribute negligibly to the specific heat of the mixed liquor so that whatever VS reduction takes place does not affect the specific heat (C_p) of the mixed liquor, and (2) with pure oxygen, only a negligibly small proportion of water is vaporized in the reactor [< 0,007%, see Section (4.2.4) above]. Accepting the influent and effluent sludge mixed liquor (i.e. water and suspended solids together) specific heats (C_p) and volumetric flow rates to be the same, the net effluent reactor sludge heat loss rate is related to the temperature difference between the feed (T_{si}) and transfer (T_{se}) sludges as follows:

$$H'_{se} = H_{se} - H_{si} = F_s \rho_s C_p (T_{se} - T_{si}) \quad \text{(MJ/d)} \tag{4.1}$$

H'_{se} = net rate of heat loss in the reactor effluent sludge with respect to the influent sludge temperature T_{si} (MJ/h)

ρ_s = density of the sludge liquor (1000 kg/m³)

C_p = specific heat [4.18 MJ/(ton.°C)]

F_s = sludge volumetric flow rate (m³/d)

= V_p/R_H

where V_p = reactor operational volume (45 m³)

and R_H = retention time in days.

The rate H'_{se} calculated with Eq (4.1) is not only the rate of heat loss in the reactor effluent sludge, but compared to the unsteady state heat balance (Eq 3.5) is also equal to the rate of increase in sludge heat content, H_{net}, from feed sludge temperature T_{si} to reactor sludge temperature T_{se}, i.e. in the unsteady heat balance the 45 m³ sludge volume, which was cooled about 3°C due to the addition of cold feed sludge, is reheated to the effluent sludge temperature, whereas in the steady state heat balance, the reactor sludge temperature is accepted to remain constant at the effluent sludge temperature and only the feed sludge volume is heated from the feed temperature to the effluent sludge temperature.

By substituting Eq (4.1) into Eq (3.4) the steady state heat balance becomes

$$H_{bi} + H_{mi} = \frac{V_p}{R_h} \rho_s C_p (T_{se} - T_{si}) + H_{we} + H_{ve} \quad \text{(MJ/d)} \tag{4.2}$$

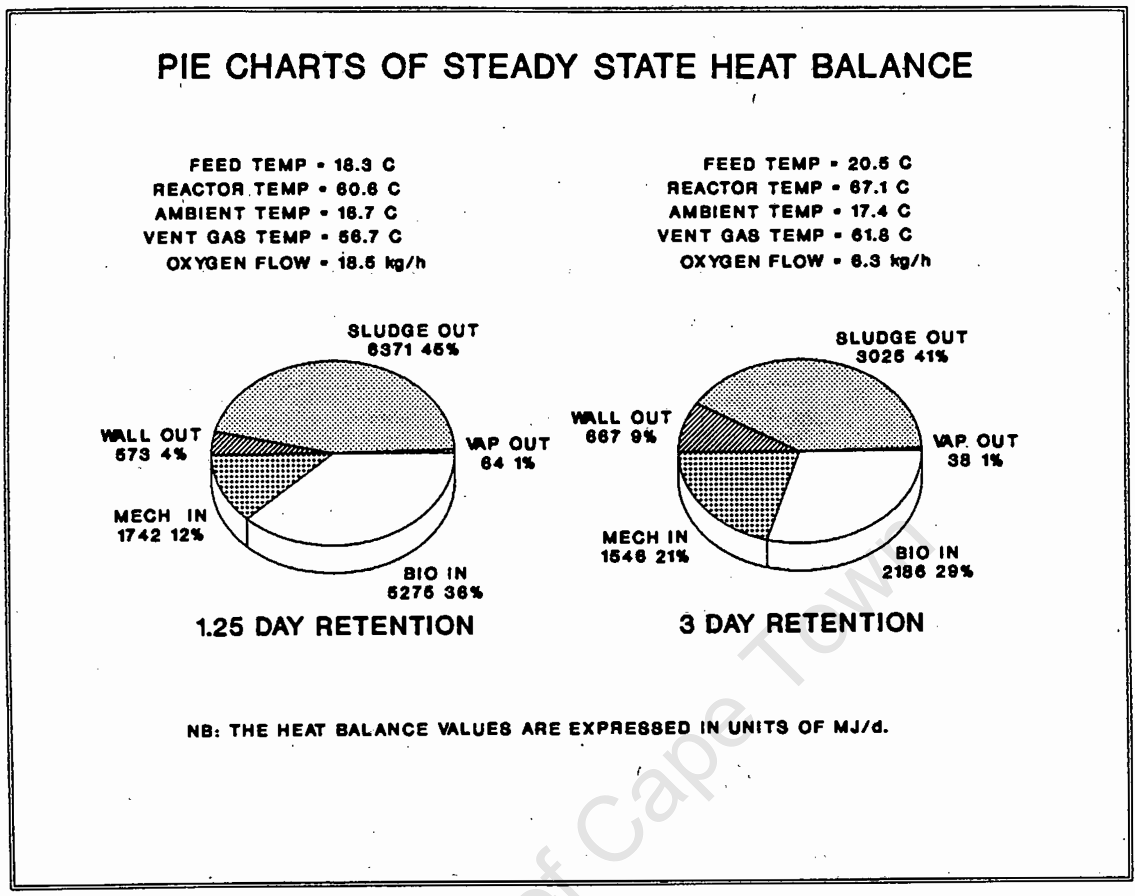
Taking one day as the unit of time, the steady state heat balance given by Eq (4.2) is illustrated in Fig 4.15 in the form of pie charts for 1,25 and 3 days retention time conditions based on the particular conditions of one of the specific heat yield tests at these retention times. The bottom half of each of the pie charts constitute the heat sources (LHS of Eq 4.2 above) and the top half the heat sinks (RHS of Eq 4.2 above). The biological heat generation rate H_{bi} and the effluent reactor sludge heat loss rate H'_{se} constitute the largest source and sink respectively, both of which decrease with increasing retention time (Fig 4.15).

From Fig 4.15, it can be seen that the values of H_{bi}, H_{se} and H_{ve} (MJ/d) more than doubled with reduction in retention time from the 3 to the 1.25 days; in the case of H_{bi} and H_{ve}, this was due to increased oxygen supply rate, and in the case of H_{se} due to increased sludge flow rate (F_s). In contrast H_{we} decreased in value when the retention time was reduced from 3.0 to 1.25 days. This reduction, which is only 16%, would not have taken place had the reactor sludge temperature been the same at 3 and 1,25 days retention time; although it was attempted to maintain reactor temperatures at about 60°C, high temperatures at the 3 day retention time were a consequence of the low flow oxygen supply rotameter being unable to accurately dispense less than 6 kgO/h, an amount which generated more biological heat than required for a reactor sludge temperature of about 60°C.

For the 65 specific heat yield tests that were done while the reactor was at steady state (as defined by the 2 conditions above), the calculated values of H_{bi}, H_{mi}, H_{we} and H_{ve} are plotted versus increasing H'_{se} in Fig 4.16. Because in the specific heat yield tests, the temperature difference between the transfer and feed sludges (T_{se}-T_{si}) did not change very much (min. 35,6; max. 48,6 °C), the change in retention time R_H represents the largest influence on H'_{se}, which decreased as R_H increased (see Eq 4.1). Consequently in Fig 4.16, increasing H'_{se} is related approximately to decreasing R_H. The inter-relationship between H'_{se} and the 4 parameters H_{bi}, H_{mi}, H_{we} and H_{ve} is important in the steady state design of the aerobic reactor and each of the 4 parameters are discussed below with reference to Fig 4.16.

4.9.1 <u>The mechanical heat input rate, H_{mi}, and wall heat loss rate, H_{we}</u>

Both H_{mi} and H_{we} are independent of H'_{se} and retention time R_H. In the case of H_{mi}, this is because the heat input from the Vitox recirculation pump was



<u>Fig 4.15</u>: Two pie charts illustrating typical daily heat balance data for the aerobic reactor measured under 'steady state' conditions at a 1.25 and a 3 day hydraulic retention time.

effectively constant – as discussed in Section 4.2.2 above; its value increased only very slightly with time as the size of the pump volute increased due to wear caused by the abrasive action of grit contained in the sludge (see Section 4.2.2). In the case of the wall heat rate loss, H_{we}, this is a function only of the reactor sludge temperature, T_{se}, and the ambient temperature, T_{amb}, as Eq (3.9) shows, and if the difference between T_{se} and T_{amb} remains approximately constant, the wall heat loss will remain approximately constant at different retention times.

4.9.2 Vent gas water vapour heat loss rate, H_{ve}

The steady heat balance data shown in Figs 4.15 and 4.16 show that the vent gas water vapour heat loss rate H_{ve} contributes less than 1% to the heat sinks; which really is too small to warrant serious evaluation. This low H_{ve} value is the result of using a pure oxygen oxygenation system which has the result that gas flows through the sludge are kept at a minimum. Even though not of much relevance to pure oxygen aerobic reactors an evaluation of this form of heat loss is made below to seek the principles involved for possible application to air oxygenated aerobic reactors (see Chapter 5, Section 5.4.5).

A plot of H_{ve} versus H_{bi} for the 65 specific heat yield tests conforming to steady state conditions (i.e. the H_{ve} data from Fig 4.16) is given at an expanded scale in Fig 4.17. This plot shows that H_{ve} is related to H_{bi}, increasing as H_{bi} increases. This increase in H_{bi} is brought about through an increase in oxygen supply rate O_s which increases the vent gas flow rate. Because, as shown earlier (see Section 4.7), the vent gas is saturated with water vapour even at the highest vent gas flow rates, H_{ve} increases as the vent gas flow rate increases.

4.9.3 The biological heat generation rate, H_{bi}

From the above discussion, it is clear that H_{mi}, H_{we} and H_{ve} either are approximately constant or negligible in comparison with net effluent sludge heat loss rate H'_{se}. From this it follows that the biological heat generation rate H_{bi} is closely proportional to the required sludge heating rate H'_{se}. This proportionality is clearly shown in Fig 4.16. From this proportionality and the direct link between H_{bi} and the oxygen consumption rate O_c, which in turn can be controlled by the oxygen supply rate under conditions of oxygen limitation, it is clear that control of H_{bi} through control of oxygen supply rate O_s, is a very convenient means for controlling the reactor sludge temperature T_{se}; H_{bi} needs to be

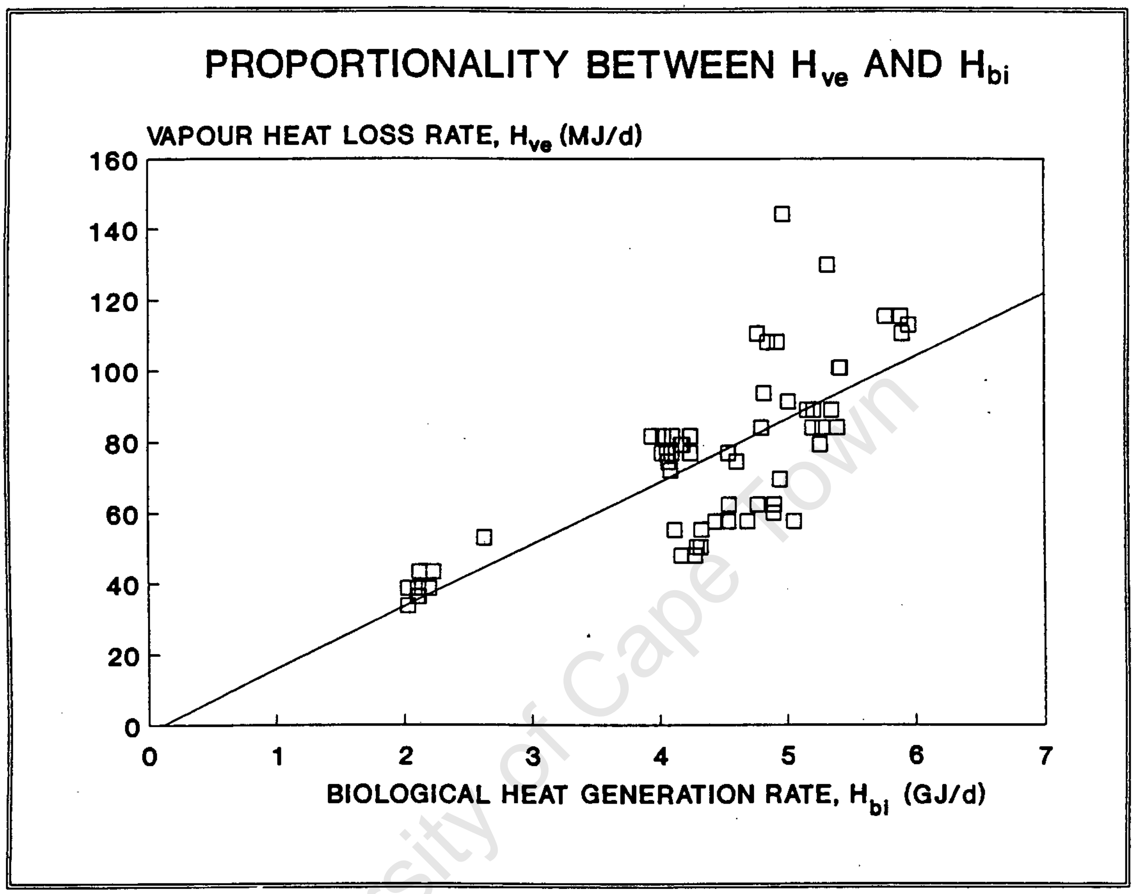


Fig 4.17: The daily rate of water vapour heat loss in the vent gas, H_{ve}, against the daily biological heat generation rate, H_{bi}. The plot is derived from 61 of the 116 unsteady state heat balances which were conducted while the reactor was at 'steady state' and shows that the 2 sets of data are directly proportional.

increased whenever the feed and transfer sludge temperature difference increases or the retention decreases (see Eq 4.2), and this is brought about by increasing O_s. However this control of H_{bi} and hence T_{se} by means of O_s is only possible while the reactor is oxygen limited. Under conditions of oxygen sufficiency, the oxygen consumption rate in the reactor is at a maximum and a further increase in O_s not increase the biological oxygen consumption rate and therefore also not the biological heat generation rate of the sludge. Thus, for specified feed and reactor sludge temperatures, the minimum retention time that can be achieved in the aerobic reactor is defined by the maximum biological heat generation rate $H_{bi\ max}$ which is directly proportional to the biological oxygen utilization rate of the sludge (OUR) through the specific heat yield (Y_H).

For the Milnerton plant, at which $O_{c\ max}$ was 17 KgO/h, $H_{bi\ max}$ would be about $17.0 \cdot 12.8 \cdot 24 = 5.22$ GJ/d, which gives a minimum retention time of 1.25 days and 1.0 days in winter and summer respectively. However, this minimum retention time can vary from plant to plant depending on the OUR of the sludge. The OUR of the sludge is likely to be affected by the type of sludge under treatment, i.e. primary, humus tank, activated or mixtures of these and possibly the concentration of sludge in the aerobic reactor.

If the OUR of a particular sludge is known then for a particular plant set up, i.e. mechanical heat input H_{mi}, wall heat loss H_{we}, vent gas vapour heat loss H_{ve} and feed sludge temperature increase to reach pasteurization temperature, the minimum aerobic reactor retention can be directly calculated with the aid of the specific heat yield Y_H.

4.10 VS AND COD REDUCTION

The advantages and disadvantages of volatile solids (VS) and COD removal as substitute energy parameters in stoichiometrically quantifying biological heat generation were discussed in Chapter 2, Section 2.4.4. During the investigation, VS and COD tests were performed on the aerobic reactor influent (feed) and effluent (transfer) sludge flows with the objective of evaluating these as substitute energy parameters i.e. can the biological heat generated be consistently and reliably related to the VS or COD removal? Additionally, the measurements of VS and COD in the influent and effluent sludge allowed calculation of the COD/VS ratio of the influent and effluent sludge and, with the measured biological oxygen consumption rate O_c, the kgO consumed per kgCOD or kgVS removed.

During the investigation 180 influent and effluent sample pairs were tested for total solids (TS) and VS, and 75 for COD. The complete test data is listed in Appendix 4B and is summarized in Table 4.5. The TS, VS and COD results obtained were periodically cross checked by 31 parallel analyses at the Bellville laboratory of the Division of Water Technology (DWT) of the Council for Scientific and Industrial Research (CSIR). The results of these analyses are presented in Appendix 4C and are summarized in Table 4.6. Of the 31 parallel analyses, the samples on which 12 were performed were duplicates of samples analysed at Milnerton. The 12 Milnerton and DWT analyses are listed in Table 4.7.

4.10.1 VS removal

From Table 4.5, comparing the average influent and effluent VS concentrations, it can be seen that, on average over the investigation, very little VS was removed in the aerobic reactor i.e. 0.39 g/ℓ. A probability density plot of the 180 paired influent and effluent VS data is given in Fig 4.18. The plot suggests that the data is normally distributed and shows that the difference between influent and effluent VS concentrations is very small i.e. the difference between their mean values is less than 13% of 1 standard deviation of the influent VS data. A statistical t-test performed on the influent and effluent VS concentrations shows that there is no significant difference between their mean values at the 95% confidence interval (see Appendix 4D).

At a retention time of 1.25d, at which most of the tests were undertaken, the sludge flow rate was 36 m^3/d at 30.28 gVS/ℓ giving a daily sludge mass treated of 1090 kgVS/d. The VS removal therefore is only 0.39x36 = 14 kgVS/d which is only 1.3% of the sludge mass treated per day.

Examining the difference between the paired influent and effluent VS concentrations, i.e. the VS removal, in g/ℓ, it can be seen from the shape of the probability density plot shown in Fig 4.19, that this is also approximately normally distributed. However, for 80 of the 180 samples (44%), the effluent VS concentration was higher than that of the influent. This is illustrated by the

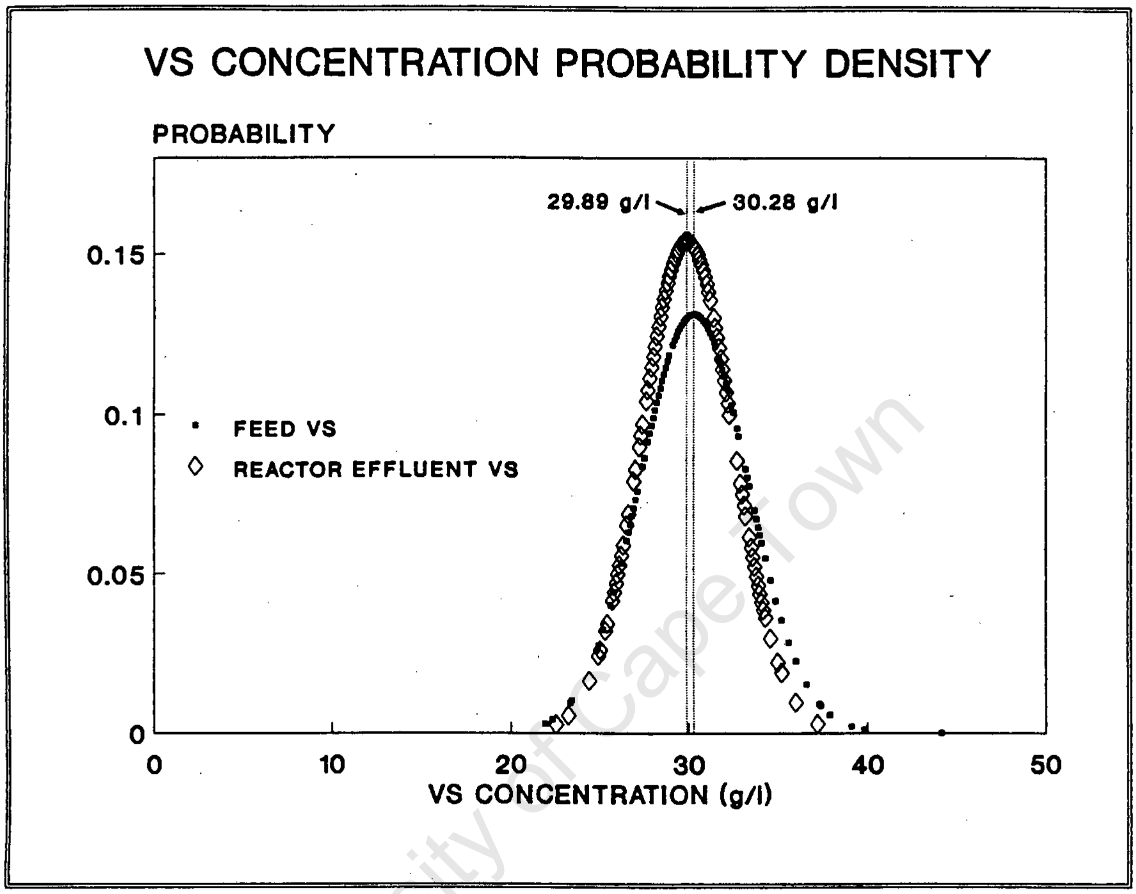
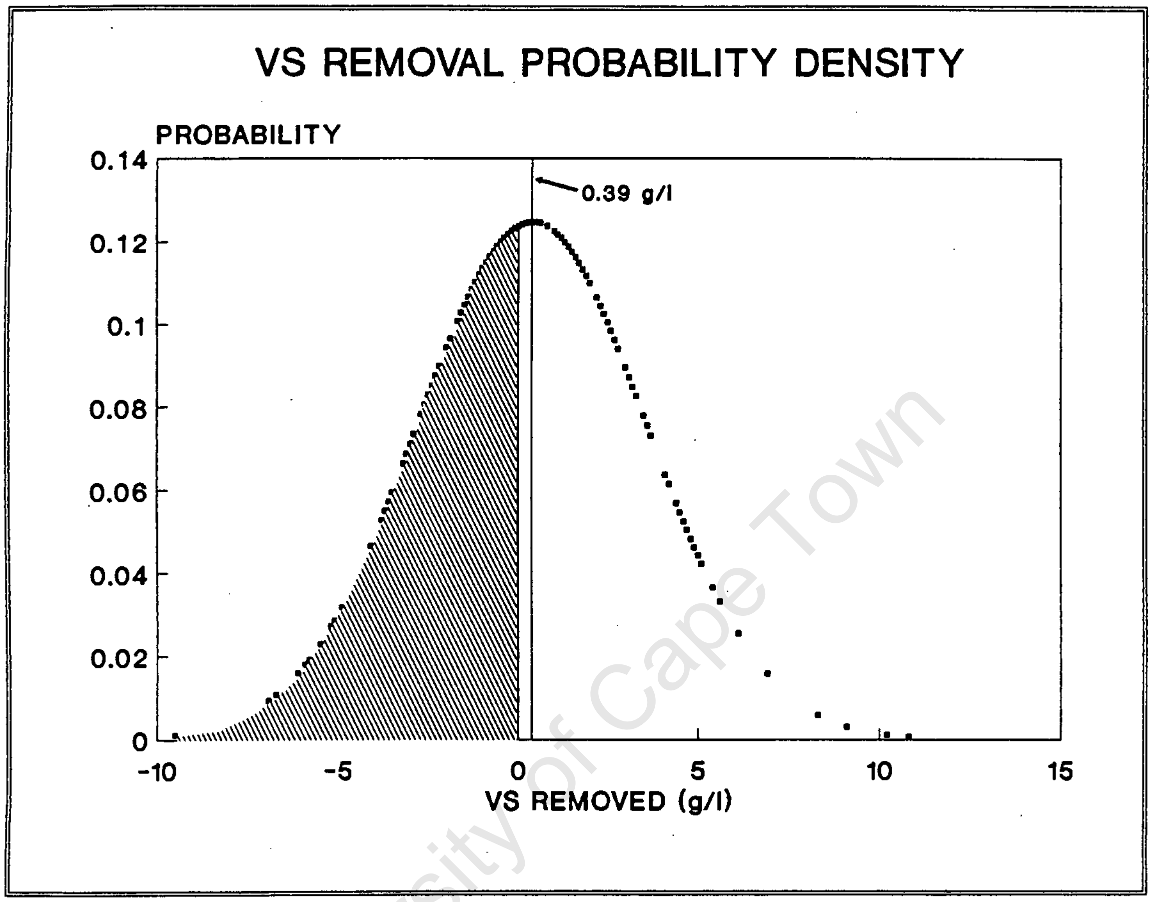


Fig 4.18: A probability density plot of feed and effluent VS concentrations measured in the aerobic reactor sludge. The plot shows that, on average, there is very little difference between the concentrations of VS in the feed and effluent sludges.



<u>Fig 4.19</u>: A probability density plot of VS removal (g/ℓ) in the aerobic reactor sludge. The shaded area represents the data where VS removal was negative ie the VS concentration in the effluent sludge was higher than the VS concentration in the feed sludge.

TABLE 4.5. Average influent and effluent Total Solids (TS), Volatile Solids (VS) and COD values measured on the aerobic reactor.

AVERAGE VALUES	TS g/ℓ	VS g/ℓ	COD g/ℓ	COD/VS
INFLUENT	37.54	30.28	52.99	1.721
STD DEV	4.35	3.04	5.95	0.097
DATA PTS	180	180	75	75
EFFLUENT	37.31	29.89	46.69	1.548
STD DEV	3.84	2.59	5.56	0.098
DATA PTS	180	180	75	75

TABLE 4.6. Average influent and effluent Total Solids (TS), Volatile Solids (VS) and COD values measured on the aerobic reactor by the CSIR.

AVERAGE VALUES	TS g/ℓ	VS g/ℓ	COD g/ℓ	COD/VS
INFLUENT	37.07	29.86	50.34	1.686
STD DEV	4.39	3.10	7.42	0.177
DATA PTS	31	31	31	31
EFFLUENT	34.54	27.59	43.15	1.561
STD DEV	4.28	2.67	6.87	0.156
DATA PTS	31	31	31	31

shaded area in Fig 4.19.

Now if the biological action in the reactor is aerobic VS degradation then, accepting for the sake of illustration that the VS has a stoichiometric formula of $C_5H_7O_2N$, every kgVS oxidized to carbon dioxide and water would generate 0,53 kgC in carbon dioxide. So for a VS removal of 14,0 kgVS/d, the mass flow rate of C in carbon dioxide generated would be 7,5 kgC/d. From the vent gas measurements the mass flow of C in the vent gas could be calculated and at 1,25d retention time was found to be about 100 kgC/d. Based on the stoichiometry above the 100 kgC/d in the vent gas should have resulted in a VS reduction of about 190 kgVS/d i.e. a 17% reduction. Yet only a 1.3% reduction was measured. This raises the question of whether the absence of measured VS removal is a valid observation or the consequence of analytical or sampling error. The accuracy of the analytical and sampling procedures is reviewed below before coming to a decision regarding the nature of the biological reactions in the reactor.

4.10.1.1 *Analytical error*

Three important factors point to the acceptance of the VS results as measured:

1) The VS test is simple and straightforward and not prone to large error if the correct procedures are observed and carefully followed.

2) Duplicate tests were performed on each sample and the two results of a pair seldom differed by more than 1%.

3) The 31 parallel analyses performed by the DWT agree closely with the analyses performed at Milnerton (see the summary in Table 4.6), although the former do indicate, on average, higher VS removal than the latter (7.6% as opposed to 1.3%). Although statistical f and t tests performed in Section 4D.3 of Appendix 4D confirm, on the basis of the DWT data, that there was significant VS removal in the aerobic reactor, this VS removal is nevertheless too slight to suggest experimental error in the Milnerton data.

4) 12 of the 180 pairs of samples analyzed at Milnerton by the writer were cross-checked in duplicate analyses performed by the Division of Water Technology (DWT) laboratory at the CSIR (see Table 4.7). The means and a statistical comparison of the Milnerton and CSIR results for TS_{in}, TS_{out}, VS_{in}, VS_{out}, COD_{in} and COD_{out} are given in Appendix 4D, Table 4D.4. From this Table it can be seen that the percentage difference between the VS values of the Milnerton and DWT laboratories are very small – less than 1,6% and 0,6% on the influent and effluent samples. Also, the statistical t-tests given in Table 4D.4 shows that there is no significant difference between the results of the two laboratories' results at the 95% confidence interval.

TABLE 4.7: DUPLICATE ANALYSES OF VS, TS AND COD BY THE CSIR AND UCT.

	CSIR TS IN g/ℓ	MILN TS IN g/ℓ	CSIR TS OUT g/ℓ	MILN TS OUT g/ℓ	CSIR VS IN g/ℓ	MILN VS IN g/ℓ	CSIR VS OUT g/ℓ	MILN VS OUT g/ℓ	CSIR COD IN g/ℓ	MILN COD IN g/ℓ	CSIR COD OUT g/ℓ	MILN COD OUT g/ℓ
	36.0	36.3	34.0	34.2	30.4	30.1	28.1	27.9	53.0	47.9	42.0	44.2
	36.1	36.0	31.6	30.7	29.8	29.6	25.8	24.9	48.0	50.9	38.0	36.6
	43.5	42.2	43.6	43.4	31.7	30.5	29.7	29.1	48.4	50.6	42.3	43.3
	43.8	43.2	42.0	42.5	34.8	34.5	33.0	34.0	56.5	57.7	48.5	49.4
	35.6	35.7	38.4	38.3	29.2	28.8	31.2	30.7	47.7	48.5	45.6	47.3
	39.2	38.5	35.9	37.0	32.3	31.9	29.6	30.5	52.2	55.6	50.2	45.0
	34.6	34.8	35.2	36.3	26.5	26.9	27.6	28.2	49.2	44.1	39.5	42.4
	35.7	34.1	33.8	32.3	27.9	26.9	26.2	25.7	45.9	45.1	42.0	39.2
	41.4	40.7	43.9	43.1	31.8	31.4	32.7	32.0	56.0	54.7	49.2	52.7
	41.2	40.2	40.3	40.1	33.1	31.7	31.2	30.7	72.7	55.2	49.8	50.0
	42.9	42.9	42.8	42.8	33.6	33.9	33.5	33.6	57.6	56.5	68.8	52.0
	38.1	37.7	36.9	35.9	30.5	29.8	29.5	28.3	49.8	59.5	49.4	44.7
AVERAGE	39.01	38.53	38.20	38.05	30.97	30.50	39.01	38.53	38.20	38.05	30.97	30.50
VARIANCE	11.74	10.50	18.02	19.22	5.78	5.64	11.74	10.50	18.02	19.22	5.78	5.64
STD DEV	3.43	3.24	4.25	4.38	2.40	2.37	3.43	3.24	4.25	4.38	2.40	2.37

It is clear from the above 4 points that the analytical results can be accepted with confidence.

4.10.1.2 *Sampling error*

The sampling procedures, which were described in Chapter 3, Section 3.7.2 were devised specifically to obtain as representative samples as possible so as to minimize errors. However the nature of primary sludge is such that accurate sampling is difficult, particularly in the case of the Milnerton feed sludge. Stratification and the absence of mixing in the feed sludge thickener may have led to feed sludge of varying concentration entering the reactor during the 3 min feeding cycle. The precautions taken to minimize the effects of this have been described in Chapter 3, Section 3.7.2. Even though during each day 1/3 of the batches of sludge fed to and transferred from the aerobic reactor were not sampled as no operator staff were available during the 0h00 – 08h00 shift, it is unlikely that the sampling procedure could have led to a consistently small difference between influent and effluent sludge concentrations if, in reality, the difference actually was large. It can be accepted that with the feed sludge in particular, deviations between the sample and sludge stream can occur but these would be expected to be (1) random and (2) not consistently low by as much as 17%, the quantity expected from the vent gas carbon flow if the biological reaction is simply that of VS degradation (see above). With regard to the effluent sludge, the intense and thorough mixing in the reactor makes this more homogeneous compared to the feed sludge with the result that the precision of the effluent sludge samples was greater. This is reflected in the data, in that the standard deviation of the effluent VS concentration is smaller than that of the feed sludge. The reproducibility of the effluent sludge VS concentration sampling and analysis method was checked on one occasion when 5 samples were drawn from the reactor at 20 min intervals and the TS concentration of each determined. The results ranged from 30.8 to 32.0 g/ℓ, with a standard deviation of 0.47 and a coefficient of variation of 1.5%. They varied randomly and showed no consistent reduction with time as sludge heating progressed.

Noting that the sampling procedures were devised specifically to obtain samples as representative as possible, bias in the VS (and TS and COD) concentration measurements is unlikely. Errors in these measurements therefore would be random rather than systematic, and random errors do not affect the mean value of a large number of samples. The randomness of the errors was checked in 2 ways (i) by plotting the data on a probability plot and (ii) by ordering the 180 VS data set chronologically, dividing the set into a number of groups and comparing the mean influent and effluent VS concentrations of each group. The probability plots indicated that both the influent and effluent data set plotted as straight lines confirming that the data are normally distributed. Because the data are normally distributed the influences that introduce error on the data are, numerous, random and independent. Dividing the chronologically ordered VS data into 3 groups of 60 each for both the influent and effluent, the respective influent and effluent means of each group, in g/ℓ are: 29.38 and 29.59; 30.30 and 29.35; 31.18 and 30.73. The small differences between the values of each influent and effluent pair and their similarity to the overall mean of all 180 data, support the conclusion that the influent and effluent VS measurements are normally distributed and consistent over the entire test period in that the time in the test period when the samples were taken did not influence the concentration.

4.10.1.3 Conclusion

It is clear from the above that considerable effort went into the collection and analysis of influent and effluent sludge samples. Because the sampling technique provided a means of obtaining representative daily composite samples, the mean of the 180 influent and effluent values is likely to be accurate, i.e. close to the true mean and the random errors to which the data were subject do not affect the accuracy, only the precision (i.e. large or small standard deviation). Because the VS data can be accepted to be accurate, it can be concluded that the low VS removal measured in the aerobic reactor is not an artifact of sampling and analysis, but is real. Accepting the VS removal to be the low 1.3% measured, implies that the biological heat generating reactions on the reactor are not simply those of aerobic VS ($C_5H_7O_2N$) degradation because the mass of C generated in vent gas carbon dioxide is 13,3 times greater than expected from VS degradation stoichiometry.

4.10.2 COD removal.

As mentioned earlier, on 75 of the 180 influent and effluent sludge samples, COD concentrations were measured. For each influent and effluent sample, 3 COD tests were done and the mean of the 3 was accepted to be the COD value of the sample (see Chapter 3, Section 3.7.2). After checking that the influent and effluent COD data sets were normally distributed by plotting on probability paper (see Appendix 4D), the mean and standard deviation of the 75 influent and effluent COD data were calculated and listed in Table 4.5. Knowing the mean and standard deviation, the probability density plots for the influent and effluent COD concentrations were constructed and are shown in Fig 4.20. From Fig 4.20 it can be seen that in contrast to the VS concentration, with the COD the difference between the mean influent and effluent COD concentration is greater and there is less overlap in the probability density diagrams of the influent and effluent COD. Statistical F and t-tests (see Appendix 4D) confirm that the means of the influent and effluent COD concentrations are significantly different at the 95% confidence interval.

From Table 4.5 the mean influent and effluent COD concentrations are 52.99 and 46.69 g/ℓ respectively, giving an average COD removal of 6.3 g/ℓ. At 1.25 day retention time (at which most of the COD tests were done) the sludge flow was 36 m^3/d giving a mass COD removal of $6.3 \cdot 36 = 227$ kgCOD/d which is 11.9% of the influent COD mass flow. Considering the individual COD removal data (obtained from the difference of the 75 influent and effluent COD concentration pairs), it can be seen from the shape of the probability plot that these also are approximately normally distributed, (Fig 4.21). From Fig 4.21 it would appear that COD removal is a better parameter than VS removal, because with the COD, only 8 out of the 75 (compared to 80 out of the 180 VS data) have effluent COD concentrations greater than the influent (shaded area in Fig 4.21). However, this is not so because the coefficient of variation (standard deviation divided by the mean) for the COD data is no better than for the VS data. Consequently, even though more COD was removed than VS, the large variation in COD removal (from –14 to +28 g/ℓ), makes this a very poor parameter on which to base biological heat generation for temperature control.

4.10.2.1 *Sampling error*

Because COD was tested on the same sample as the VS, the conclusion that representative samples were obtained for the VS applies also to the COD. The only difference is that the VS test was conducted on an undiluted sample while the COD sample was diluted by a factor of 100. This dilution would have the effect of a 100 fold increase in the event of a sampling error.

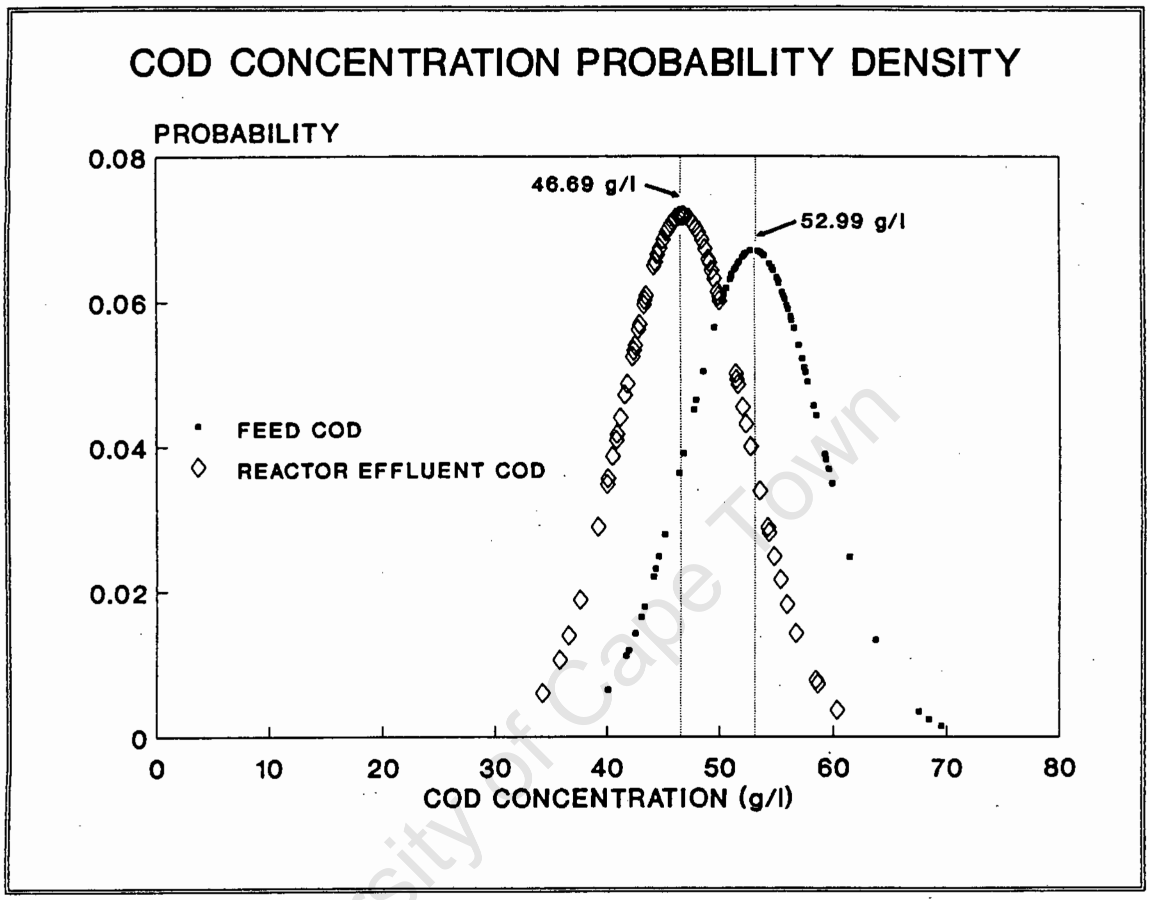


Fig 4.20: A probability density plot of feed and effluent COD concentrations measured in the aerobic reactor sludge.

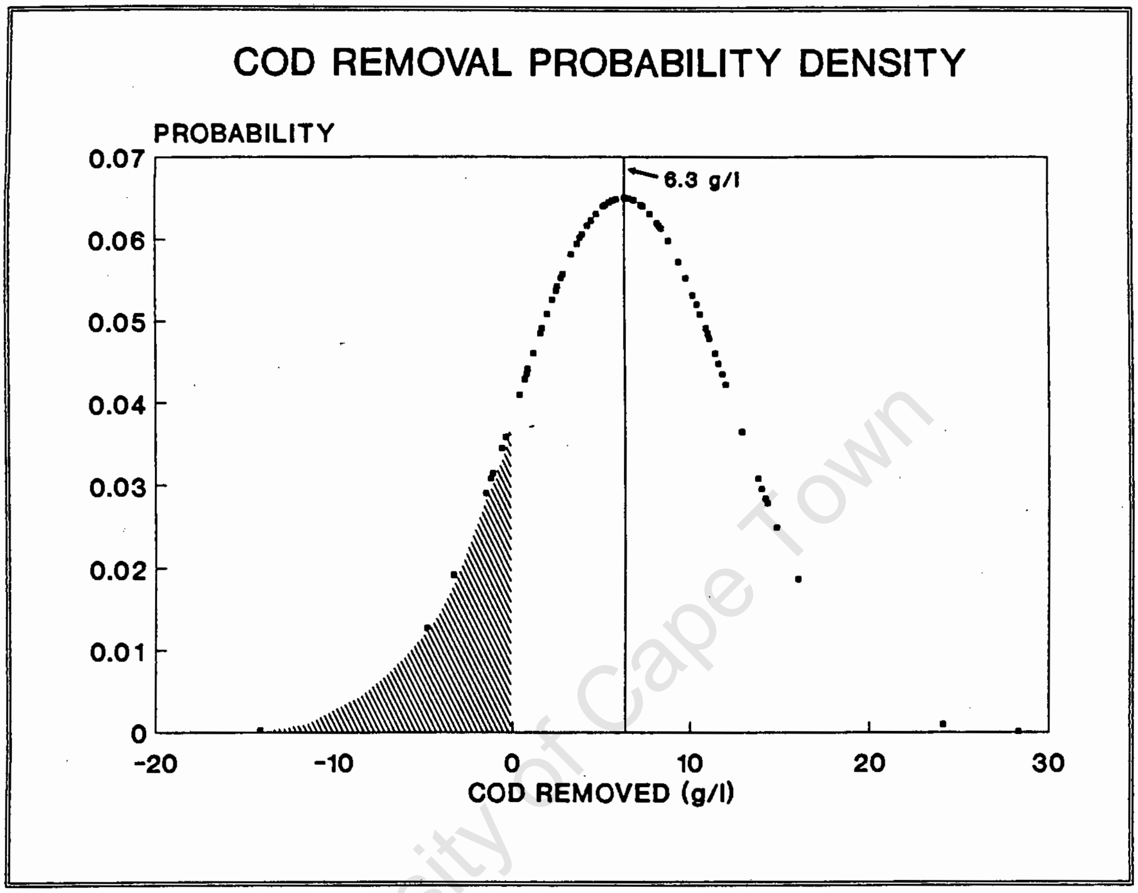


Fig.4.21: A probability density plot of COD removal (g/ℓ) in the aerobic reactor sludge. The shaded area represents the data where COD removal was negative ie the COD concentration in the effluent sludge was higher than the COD concentration in the feed sludge.

4.10.2.2 *Analytical error*

The COD test is more complex than the VS test and is subject to interference by compounds resistant to oxidation. In a study to determine the comparative reliability of the COD test, Smith reported a coefficient of variation of 31.1% in the analysis by 18 laboratories of the same sample of filtered sewage effluent. In a later study involving 37 laboratories, the same author reported smaller coefficients of variation of 5.6, 8.7, 11.8 and 14.7% for the COD determinations of 4 samples of filtered (4μm) raw sewage.

Because sewage sludge contains particulate matter, the COD of a sample of primary sludge is more difficult to measure accurately than a sample of filtered sewage. This results in large variations between COD analysis on the same sample and for this reason 3 COD tests were done on each sample. In the 75 influent and effluent COD concentrations, it was found that the 3 COD's of one triplicate set showed variations of up to 10% of the mean of the 3 values. In contrast, the variation between a pair of VS tests on a sludge sample rarely exceeded 1%.

As mentioned above for the VS results, 12 of the influent and effluent samples were tested also for COD for comparative purposes at both the Milnerton and DWT laboratories. The results of the 12 comparative COD tests are listed in Table 4.7 and a statistical analysis is given in Appendix 4D, Table 4D.4.

From Table 4D.4 it can be seen that the difference between the means of DWT and Milnerton influent and effluent COD's is less than 1,7% and that at the 95% confidence interval, there is no significant difference between these means. Consequently it can be accepted the COD test was accurately and reliably conducted.

4.10.2.3 *Conclusion*

Although the COD test gives an accurate measure of the influent and effluent sludge COD, when a large set of COD data is available, it, like the VS, is very imprecise on a single daily sample. Even though theoretically it is closely related to biological energy changes, from a point of view of practice, the COD does not provide a better substitute energy measure than the VS.

4.10.3 Influent and effluent COD/VS ratio

From the discussion of the VS and COD results in Sections 4.10.1 and 4.10.2 above, the aerobic reactor produced on average a larger reduction in sludge COD than in sludge VS i.e. 1.3% VS and 11.9% COD. The removal of more COD than VS caused the sludge COD/VS ratio to decrease across the reactor. For 75 paired influent and effluent VS and COD values, the mean COD/VS ratios for the influent and effluent sludge are 1.721 and 1.548 kgCOD/kgVS respectively i.e. a decrease of 10%. As expected from the large removal of COD compared to the VS, a statistical t-test confirms a significant difference between the means at a 95% confidence level.

In models of autothermal thermophilic aerobic digestion which describe the rate of biological heat generation in terms of VS removal kinetics and the stoichiometry in terms of oxygen consumption, the COD/VS ratio forms the link between the kinetics and the oxygen consumption - see Eq (2.9b). From the above and COD results it is clear that this approach is inappropriate for the aerobic reactor in dual digestion, because the COD/VS ratio of the sludge changes through the reactor. Therefore, apart from the observation that the VS removal was very small, to estimate the oxygen consumption on the basis of VS removal kinetics through a constant COD/VS ratio is clearly very unreliable for the aerobic reactor design.

4.10.4 The mass of oxygen consumed per kg COD removed

For 65 of the 75 influent and effluent COD data pairs, it was possible to compare the daily mass of COD removed with the daily mass of oxygen consumed by the reactor sludge and hence to calculate the ratio (kgO consumed)/(kgCOD removed). The ratio ranged from -24.9 to 29.6 kgO/(kgCOD removed), with an average of 2.34 ± 5.88 kgO/(kgCOD removed). It is worth commenting on 2 aspects of this data, (1) the deviation of the measured ratio from that predicted from stoichiometry and (2) the imprecision of the data.

4.10.4.1 The stoichiometry of oxygen consumption and COD removal

From the definition of COD, the ratio kgO/(kgCOD removed) should be unity, but the data shows that on average, 2.34 times more oxygen was consumed than COD removed. Differences between the expected stoichiometric ratio and that measured may possibly be due to (1) electron donors in the sludge which are not oxidised in the COD test, (2) interference with the COD test by chlorides, and (3) nitrification.

However none of these two reasons are probable. In the first instance, the concentration of electron donors not oxidised in the COD test would be too small to account for the large discrepancy between oxygen consumed and COD removed. The most common of these electron donors found in sewage are short chain fatty acids (SCFA's) but the COD test procedure employed at Milnerton was adapted to oxidize SCFA's. In the second instance, chlorides were complexed by the addition of mercuric sulphate and in the third instance, both the high temperature (60°C) and oxygen limiting conditions (1.25d) suppress nitrification - that nitrification did not occur is supported by the fact that the ammonia concentration increased, rather than decreased across the reactor.

The fact that a COD balance was not obtained over the aerobic reactor was a perplexing problem in the investigation; for this reason a number of measures were adopted to check the apparently spurious COD data eg fastidious compositing of samples, triplicate testing and interlaboratory cross checks. Despite these measures, no explanation for the apparently poor COD balance across the reactor can be advanced.

4.10.4.2 Imprecision of the measured ratio kgO/(kgCOD removed)

Because the measurement of oxygen consumption in the Milnerton reactor was shown to be accurate and precise, the wide scatter in the kgO/kgCOD removed ratio results from the large variation in the daily measurement of the COD change across the aerobic reactor. The large variation arises because the COD change or removal is the difference between two large numbers; small variations in each of the daily influent and effluent COD concentrations, cause large variations in the daily COD removal. This variability is evident when examining the 12 duplicate Milnerton and DWT COD analyses mentioned in Section 4.10.1.1 above and listed in Table 4.7. The COD removal calculated from a pair of influent and effluent DWT COD's correlates very poorly with the COD removal calculated from the Milnerton COD's. Indeed the percentage difference between the 12 corresponding DWT and Milnerton COD removals ranges from 3.6% to 470% with an average of 161%; this in spite of (1) the statistical F and t tests showing no significant differences between the Milnerton and DWT influent and effluent COD's, and (2) the good comparison between corresponding individual COD's from the two laboratories (the average percentage difference between the 12 Milnerton and

DWT influent and effluent COD's is 7,5% and 6,9% respectively).

From the above results, it is clear that for the short retention time aerobic reactor in dual digestion any stoichiometric parameter which is based on COD or VS removal, such as kgO or MJ, will show very wide variation on a day to day basis. Over a long period of time, accurate values of the parameters may be determined, but because of their wide variation on a day to day basis, the parameters are basically useless for process operation and temperature control. Essentially the VS and COD (and TS) have value only in providing information on the quantity and quality of sludge processed daily.

4.10.5 Conclusion

VS. The apparent absence of VS removal in the aerobic reactor makes it impossible to derive a relationship between the biological heat generation and the VS removal in the reactor. The fact that 44% of the VS analyses showed a higher VS concentration in the effluent than in the influent sludge suggests that such a relationship would be highly unreliable. In the literature figures of 21 MJ biological heat per kg of VS removed are cited. If such a calculation is made for the Milnerton reactor, where typically 5275 MJ/d of biological heat was generated at an oxygen consumption of around 408 kgO/d and a VS removal of 14 kgVS/d, values of 377 MJ/kg removed and 29 kgO/kgVS removed are obtained, values completely unrealistic in terms of VS degradation stoichiometry.

COD. While the stoichiometry between biological heat generation and oxygen consumption has been proven for the aerobic reactor of the Milnerton dual digestion system, the stoichiometry between oxygen consumption and COD removal in this reactor has been shown to be highly variable and therefore unreliable. This variability is due to the difficulty in obtaining a precise measure of the sludge COD removal across the reactor, which arises because the removal is given by a small difference between two large numbers which themselves can vary more than the difference between them. Hence, in spite of the theoretical soundness of linking biological heat generation to COD removal as discussed in Chapter 2, the practicality of precisely determining the COD removal in the aerobic reactor of dual digestion makes the use of the COD as a substitute energy parameter impractical and unreliable.

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CHAPTER 5

CONSIDERATIONS IN THE PROCESS DESIGN OF THE AEROBIC REACTOR OF A DUAL DIGESTION SYSTEM

ABSTRACT

In this chapter, design considerations for pure oxygen and air oxygenated aerobic reactors based on the steady state heat balance, are outlined. Accepting that the objectives of the aerobic reactor are sludge pretreatment through oxygen limitation and pasteurization, it is demonstrated that three parameters are of crucial importance to the design procedures which centre around minimizing the reactor retention time: (1) the oxygen utilization rate (OUR) of the sludge which fixes the maximum biological heat generation rate of the sludge, (2) the oxygen transfer rate (OTR) of the oxygenation system (which should be somewhat less than the OUR to ensure oxygen limitation) and (3) the oxygen transfer efficiency, which determines the volumetric flow rate of the vent gas. For a given OTR, if OTE is high (attainable with pure oxygen), vent gas heat losses (through water vapour and sensible heats) are small and most of the heat generated can be lost via hot effluent sludge thereby allowing short retention times; if OTE is low (which is usually the case with air), vent gas heat losses are high and much less heat can be lost via hot effluent sludge thereby forcing longer retention times. Heat exchange between reactor effluent and influent sludge flows and increased mechanical heat input, increase heat sources without increasing heat losses and therefore allow *pro rata* reductions in retention time.

5.1 INTRODUCTION

The aims of this chapter are to discuss the various factors which need to be considered in the process design of the aerobic reactor of a dual digestion system and to illustrate some of these factors by means of graphs with a design example. From the results of the investigation into the performance of the aerobic reactor of the Milnerton dual digestion system, the single most important aspect relevant to design is the direct proportionality between the rates of biological heat generation (H_{bi}) and oxygen consumption (O_c) through the specific heat yield Y_H. This proportionality is fundamental both to the design procedures, and to an understanding of how design objectives may be achieved.

The process design procedure presented in this chapter was derived from the results of the Milnerton aerobic reactor presented in Chapter 4. The derived procedure is founded on the basic principles observed in the Milnerton results and accordingly is suitable for general application in the design of the aerobic reactor of a dual digestion system. Therefore, whereas only pure oxygen was used to oxygenate the Milnerton aerobic reactor, the design procedure applies to reactors which are oxygenated either by pure oxygen or by air. As an example, the Milnerton reactor is adopted to outline the principles and demonstrate the design procedure.

The design procedure is based on the solution of the steady state heat balance across the reactor. Such a heat balance (for details see Chapter 3, Section 3.4.2) is applicable to a reactor which is continuously fed. In practice however, the requirements of sludge pasteurization impose a system of batch feeding on the reactor (see below). Nevertheless, the steady state approach simplifies the design procedure and serves as a useful starting point. The small error which it introduces into the design is easily corrected with the aid of a simulation programme. A general simulation programme based on the unsteady heat balance, which allows minute by minute prediction of the reactor sludge temperature for a wide range of conditions, is presented in Chapter 6.

5.2 THE OBJECTIVES FOR THE DESIGN OF THE REACTOR

The objectives which have to be met in the design of the reactor are largely defined by the role of the reactor which is twofold, viz.

1) *Pasteurization:* Pasteurization is a method of disinfecting sludge by exposing the sludge to a specified minimum temperature for a minimum length of time. For example, at Milnerton these minima were 55° C and 2h, although operation at 60° C and 2.5h was generally adhered to.

   The design of the thermophilic aerobic reactor for pasteurization must ensure that sufficient heat is provided in order to maintain the reactor sludge at the specified thermophilic temperature for the required length of time. Because a possibility of feed to effluent short circuiting exists in a continuously fed reactor, batch draw and fill feeding is essential to guarantee that the entire reactor sludge contents are indeed exposed to the pasteurization temperature for the specified length of time.

2) *Pretreatment:* The thermophilic oxidative environment in the thermophilic aerobic reactor apparently pretreats the sludge, leading to an enhanced performance of the mesophilic anaerobic digester (the second stage of the dual digestion process) compared to conventional mesophilic digestion. The enhanced performance may be due both to the biological oxidation reactions occurring in the sludge and to the thermal effect on the sludge caused by operation at thermophilic temperature. Research by Mason (1986) suggests that the biological pretreatment effect in the reactor is improved by operating the reactor under conditions of oxygen limitation – this increases production of extracellular enzymes and hence greater solubilization of particulate material occurs. According to Eastman and Ferguson (1981), the hydrolysis of particulate material to soluble substrate is the rate limiting step in anaerobic digestion.

The literature does not define the temperature and the degree of oxygen limitation at which sludge pretreatment in the reactor is best accomplished, nor could this be established at Milnerton. Thus in the absence of explicit pretreatment specifications, the approach adopted to the reactor design is simply to ensure that (1) the specified pasteurization temperature and times are maintained in the reactor, and (2) the sludge is oxygen limited. From the literature it appears that long hydraulic retention times of the sludge in the reactor (> 2d) are not required because (1) sludge stabilization is not an objective of the reactor – this is accomplished in the anaerobic digester, and (2) sufficient heat apparently can be generated to maintain the pasteurization specifications at short retention times (< 2d) provided the heat losses could be minimized, e.g. by using pure oxygen. Short hydraulic retention times are generally desirable as the constraints of available capital and land space often dictate that the reactor be kept small. Typically, in cases where heat losses are minimized, a retention time of 1 day is proposed [Drnevich and Matsch (1978)] and so a retention time of 1 day was accepted as the design value for Milnerton. Data from full scale plants in Switzerland, Germany and Austria indicate that operation at 1 to 1.8 days is generally accomplished [Baier and Zwiefelhofer (1989)].

In aerobic reactors where heat losses are not minimized, e.g. those with air oxygenation, retention times necessarily will be longer. To take account of these types of aerobic reactors also, the design procedure below will consider both short and long retention times i.e. from 1 to 6 days. It is important to note that insofar as heat generation is concerned, there is no difference between short and long retention time reactors because, as was shown in Chapters 3 and 4, heat generation is proportional to oxygen utilization and not VS removal; the maximum oxygen utilization rate of the sludge governs the minimum retention time to achieve pasteurization conditions. Thus the most important consideration in design is to understand that the maximum heat generation rate is a characteristic of the particular sludge, and the greater the measures that are taken account of in the design to minimize heat losses and to maximize heat sources, the shorter the retention time of the reactor can be. This is the governing principle in the design procedure and considerations outlined below.

5.3 HEAT GENERATION IN THE REACTOR

The temperature of the sludge feed to the aerobic reactor of a dual digestion system generally falls into the lower mesophilic range of 10 to 25° C. The feed sludge temperature (T_{si}) at Milnerton ranged from 17° C in winter to 26° C in summer. The temperature of the sludge in the reactor, T_{se}, must be high enough to pasteurize the sludge - in a biological process such as dual digestion, a temperature in the thermophilic range, that is 50 to 65° C, is considered to be sufficient for these purposes provided the minimum time criteria are complied with. In the design example presented in this chapter, 20 and 60° C are accepted for T_{si} and T_{se} respectively.

To increase the sludge temperature from T_{si} to T_{se}, heat is required to increase the heat content of the sludge. The rate of increase in sludge heat content is proportional to the flow rate of sludge through the reactor and may be calculated with the following equation:

$$H'_{se} = F_s \times \rho_s \times Cp \times (T_{se} - T_{si})$$

$$= \frac{V_p}{R_H} \times \rho_s \times Cp \times (T_{se} - T_{si}) \qquad (MJ/h) \qquad (5.1)$$

where,

H'_{se} = the rate of increase in sludge heat content {MJ/h}
F_s = the flow rate of sludge through the reactor {m³/h)
Vp = the operating volume of the reactor (m³}
ρ_s = the sludge density {ton/m³}
R_H = the hydraulic retention time in the reactor {days}
Cp = the specific heat of the sludge {MJ/(ton.°C)}
T_{se} = the temperature of the reactor sludge {°C}
T_{si} = the temperature of the feed sludge {°C}

The heat that is required to effect the rate of increase in sludge heat content, H'_{se}, may be derived from a number of heat sources. If the sludge in the reactor is to be maintained at a constant temperature, T_{se}, then heat must be supplied to the reactor at a rate which is greater than H'_{se} in order to compensate for the rate at which heat is lost to a number of heat sinks. Accordingly, the rate of increase in heat content of the sludge, H'_{se}, is equal to the difference between the rate of heat supply from the heat sources and the rate of heat loss to the heat sinks:

$$H'_{se} = H_{sources} - H_{losses} \quad (MJ/h) \qquad (5.2)$$

where,

$H_{sources}$ = the rate of heat generation in the reactor from the heat sources in order to maintain the desired temperature in the sludge {MJ/h}
H_{losses} = the sum of the rates of all the heat losses from the reactor in various forms excluding only the heat lost via the hot effluent sludge, H'_{se} {MJ/h}

Equations (5.1) and (5.2) clearly show that to operate the reactor at short retention times necessitates maximizing the rate $H_{sources}$, and minimizing the rate, H_{losses}. This task may best be accomplished by considering the steady state heat balance across the reactor.

5.4 THE STEADY STATE HEAT BALANCE ACROSS THE REACTOR

The steady state heat balance across the aerobic reactor presented in Chapter 3 [see Eq (3.4)] was simplified for the Milnerton reactor by the omission of a term describing the rate of heat loss from the reactor in the sensible heat of the vent gas because with pure oxygen oxygenation this rate is negligible. However in the general case where air oxygenation may be employed it needs to be included; including it in Eq (3.4), the steady state heat balance becomes:

$$H_{sources} = H_{sinks}$$
$$H_{bi} + H_{mi} + H_{gi} + H_{si} = H_{se} + H_{ge} + H_{we} + H_{ve} \qquad (5.3)$$

where,

H_{sinks} = the sum of the rates of heat losses from the reactor in all forms including the net rate of heat loss via the hot effluent sludge H'_{se} {MJ/h}

H_{bi} = the rate of biological heat generation {MJ/h}

H_{mi} = the rate of heat generation due to mechanical agitation {MJ/h}

H_{gi} = the rate of heat gain with respect to a defined reference temperature due to the sensible heat of influent gas {MJ/h}

H_{si} = the rate of heat gain with respect to a defined reference temperature due to the sensible heat of influent feed sludge {MJ/h}

H_{se} = the rate of heat loss with respect to a defined reference temperature due to the sensible heat of effluent reactor sludge {MJ/h}

H_{ge} = the rate of heat loss with respect to a defined reference temperature due to the sensible heat of vent gas {MJ/h}

H_{we} = the rate of heat loss from the reactor surfaces {MJ/h}

H_{ve} = the rate of heat loss due to the vaporization of water from the reactor sludge by the vent gas {MJ/h}

Equation (5.3) above is an alternative representation to Eq (5.2) of the steady state heat balance across the aerobic reactor and can be considerably simplified by making the reasonable assumptions that the (1) mass flow rate, (F_s), (2) density (ρ_s) and (3) specific heat (C_p) of the influent and effluent sludges are the same. The reasonableness of accepting the density and specific heat of the sludge liquor to be constant during aerobic thermophilic treatment and equal to that of water was discussed in Chapter 3, Section 3.4.2; these assumptions tend to overestimate the heat requirements but insignificantly so. With regard to the volumetric sludge flow rate remaining constant, this was found to be the case at Milnerton because the volume of water vapour lost in the vent gas amounted to at highest 3,35 kg/h out of 45 m^3 i.e. 0,007% per hour. In air oxygenated systems this would be considerably higher and not necessarily negligibly small so that the volumetric flow rate in the effluent may be measurably lower than the influent. However, insofar as the heat balance is concerned, whether water leaves the aerobic reactor as water vapour or as effluent liquid, it nevertheless requires to be heated to the reactor temperature. So if the sensible heat of the water vapour is excluded in the heat balance, which in this development is the case, the influent and effluent sludge volumetric flow rates can be assumed equal with no loss of accuracy.

Accepting these 3 assumptions, the difference between H_{si} and H_{se} in Eq (5.3) may be combined into a single term H'_{se} representing the net rate of heat loss via hot sludge leaving the reactor where H'_{se} is given by Eq (5.1) i.e. :

$$(H_{se} - H_{si}) = \frac{V_p}{R_H} \text{ x } \rho_s \text{ x } Cp \text{ x } (T_{se} - T_{si}$$

$$= H'_{se} \qquad (MJ/h) \qquad (5.4)$$

In the same way as the two terms representing the heat gain and loss via the influent and effluent sludge flows, H_{si} and H_{se}, are combined into a single term by making reasonable simplifying assumptions, so also the two terms representing the sensible heats of the influent and vent gas streams H_{gi} and H_{ge} can be combined into a single term if it is assumed that (1) the specific heats of the various gas components O_2, N_2 and CO_2 are assumed constant over the temperature range 20 – 60°C and (2) the molar flow rates of the influent and vent gas streams are equal. In Appendix 5A the reasonableness of accepting the first assumption without any significant loss of accuracy is outlined and values of 0.0302 and 0.0368 MJ/(kmol.°C) were calculated for air and pure oxygen oxygenated reactors. With regard to the second assumption, this clearly is not true because the respiration quotient Y_{CO_2}, is not necessarily unity. However, for pure oxygen reactors, where the Y_{CO_2} value plays a significant role in determining the vent gas molar flow rate, the magnitude of the influent and vent gas sensible heat terms H_{gi} and H_{ge} are negligibly small (which is why these two terms were ignored in the heat balance as applied to the Milnerton reactor). In contrast, for air oxygenated reactors, the terms H_{gi} and H_{ge} are not negligible, but now the volume of inert nitrogen gas passing through the reactor is so large in comparison to the O_2 utilized and CO_2 generated, that a value of Y_{CO_2} of 0,66 or 1,0 makes a negligible difference to the magnitude of molar flow rates. Therefore it can be accepted that insofar as the heat balance is concerned, the molar flow rates of the influent and vent gas streams are the same, (ie $Y_{CO_2} = 1$) without any significant loss of accuracy.[1]

Hence

$$H'_{ge} = H_{ge} - H_{gi}$$
$$= m_{dg}\, C_{pg}\, (T_{ge} - T_{gi}) \quad (MJ/h) \qquad (5.5)$$

where,

H'_{ge} = the net rate of sensible heat loss in the vent gas {MJ/h}

m_{dg} = molar flow rate of the dry vent gas {kmol/h}

C_{pg} = specific heat of the vent gas {MJ/(kmol.°C)}

Accepting Eq (5.5) for H'_{ge} reduces Eq (5.3) to:

$$H_{sources} = H_{sinks}$$
$$H_{bi} + H_{mi} = H'_{se} + H'_{ge} + H_{we} + H_{ve} \qquad (5.6)$$

Substituting Eq (5.1) for H'_{se} into Eq (5.6) and rearranging, yields Eq (5.7) i.e.

$$\frac{V_p}{R_h} \rho_s C_p (T_{si} - T_{se}) = H_{bi} + H_{mi} - H'_{ge} - H_{we} - H_{ve}$$
$$= H_{sources} - H_{losses} \qquad (5.7)$$

From Eq (5.7) it can be seen that in order to minimize the retention time, R_H, $H_{sources}$ needs to be maximized and H_{losses} minimized. To accomplish the task

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[1] If this is so then the question comes to mind as to why so much trouble was taken at Milnerton to measure the vent gas volumetric flow rate and composition. The principal reason was to measure the oxygen consumption rate to accurately determine the specific heat yield for which the oxygen mass flow rate in the vent gas was required to be known; it was not required for the accuracy of the heat balance. The high degree of sensitivity of the specific heat yield to error in the measurement of the parameters required to obtain the oxygen consumption rate such as the oxygen supply rate and the oxygen mass flow rate in the vent gas, is discussed in Chapter 4, Section 4.4.4. The equipment set up at Milnerton to measure the oxygen mass flow rate in the vent gas is described in detail in Chapter 3, Section 3.4.1.2.

Methods for OUR prediction that are detailed in the literature are of doubtful reliability. An example of such a method is presented in Chapter 2 [Eq (2.3)], and is based on VS removal rates in the sludge. That the sludge in the Milnerton investigation was found to consume significant amounts of oxygen without showing a loss of VS, implies that this approach to predicting OUR is unreliable.; indeed on the basis of even very high thermophilic VS removal rates it was shown in Chapter 2, Section 2.5, that sufficient heat cannot be generated to operate at retention times of less than 3 days, yet the reactor at Milnerton showed that thermophilic temperatures could readily be achieved through biological heating at 1,25 d retention time. For this reason it is recommended that the OUR of the sludge is best quantified by measurement on pilot scale trials.

Regarding the second of the above 3 constraints, i.e. oxygen limitation, it is desirable for pretreatment purposes that the reactor sludge consumes oxygen at a rate lower than its OUR in order to keep the sludge oxygen limited. Thus, if oxygen limitation is required, the rate of oxygen transfer (OTR) to the reactor sludge must be less than the sludge OUR and accordingly, biological heat generation cannot be exploited at its maximum rate. Therefore under oxygen limitation, the biological heat generation rate H_{bi} is constrained by the OTR i.e.

$$H_{bi} = Y_H \times OTR \times V_p \quad (MJ/h) \tag{5.9}$$

where,

OTR = the rate of oxygen transfer to the sludge mixed liquor {kg/(m³.h)}

Oxygen limitation is desirable for reasons other than sludge pretreatment. It also allows the reactor sludge temperature to be controlled by means of oxygen supply rate. While the reactor sludge is oxygen limited, increasing oxygen supply rate causes more oxygen to be consumed and thus more biological heat to be generated. The increased biological heat generation rate raises the reactor sludge temperature via Eq (5.7). Conversely decreasing the supply of oxygen to the reactor sludge, lowers the reactor sludge temperature. The response of biological heat generation rate to both an increase and a decrease in oxygen supply rate is immediate (see Chapter 4, Section 4.8).

Having defined a suitable oxygen transfer rate (OTR) as some value less than the

OUR, it is then necessary to consider the third of the 3 constraints defined above – the capacity of the oxygenation system to transfer oxygen to the reactor sludge. This capacity is generally not favoured by the viscosity, particulate nature and high temperature of the sludge in the reactor, all of which hinder the rapid transfer of oxygen. Moreover, while oxygen transfer is generally proportional to the amount of electrical power consumed by the oxygenation system, economic considerations in sludge treatment require excessive use of electrical power to be avoided. Although higher oxygen transfer rates can be achieved with pure oxygen rather than with air, the use of pure oxygen must also be assessed in the light of its higher operational costs. During the investigation at Milnerton, the oxygenation system used pure oxygen and consumed about 20 kW of electrical power via the recirculation pump and achieved a maximum measured OTR of 0.44 kg/(m^3.h) in the 45 m^3 reactor. However, it appeared that this maximum was constrained by the OUR of the sludge and not by the oxygen transfer capacity of the oxygenation system. Qualitatively good reactor mixing seemed to enhance oxygen transfer at Milnerton although quantitative evidence was not gathered to support this observation (see Appendix 5B, Section 5B.1.4).

5.4.2 The rate of mechanical heat input, H_{mi}

Power is consumed to mix and oxygenate the reactor sludge. Much of this power may be transferred to the sludge as heat, especially when mixing and oxygenation is via a pumped sidestream. If the mixing and oxygenation is continuous, then heat will be derived from this source at a constant rate. At Milnerton 90% of the electrical power consumed by the Vitox system (i.e. 19 kW $\equiv$ 70 MJ/h) was converted to heat in the reactor sludge and H_{mi} formed a significant portion of the net rate of sludge heat content increase, H'_{se} typically 27% and 51% at 1.25 and 3 day retention times respectively. Note that although H_{mi} was constant at around 70 MJ/h at Milnerton, because H'_{se} decreases with increasing retention time (see Eq 5.4), H_{mi} becomes a larger and larger proportion of H'_{se} as retention time increases.

The specific power consumption (W/m^3) to achieve given oxygen transfer rates can differ from reactor to reactor. With pure oxygen, an OTR of 0.44 kg/(m^3.h) was achieved at Milnerton with 450 W/m^3 (further increase in OTR was limited by the sludge OUR). At Palmersford, U.K., Booth and Tramontini (1984) reported achieving an OTR of 0.134 kg/(m^3.h) with 130 W/m^3. Gunson and

Morgan (1989) presented a design based on air oxygenation which they claimed capable of achieving an OTR of 0.43 kg/(m³.h) at 120 W/m³.

A fundamental equation to predict H_{mi} is not offered in the design considerations presented in this chapter. In general, however, it will tend to be proportional to the oxygen transfer rate that is required by the design, i.e. the greater the mass of oxygen transferred to the liquid per m³ per hour, the greater the power required per m³.

5.4.3 The net rate of sensible heat loss in the vent gas, H'_{ge}

The oxygen or air supply entering the reactor and the vent gas leaving the reactor, differ in temperature, composition and molar flow rate. Nevertheless, as outlined in Section 5.4 above, for the purposes of calculating the net rate of vent gas sensible heat loss from the reactor, the differences in composition and molar flow rate may be ignored. As mentioned earlier the error in the heat balance arising from this simplification is small and is justified by the increased ease with which the net vent gas sensible heat loss term can be calculated. An equation describing H'_{ge} in terms of the molar flow rate, (m_{dg}) and specific heat (C_{pg}) of the dry vent gas and the temperature difference between influent and vent gas ($T_{ge} - T_{gi}$) was given above in Eq (5.5). In Appendix 5A a general equation linking m_{dg} to the oxygen transfer rate (OTR) and the oxygen transfer efficiency (OTE) for air and pure oxygen oxygenated reactors is derived, i.e.

$$m_{dg} = \frac{(a - OTE + OTE.Y_{co_2})}{32\ OTE} \cdot OTR.V_p \qquad \text{(kmol/h)} \qquad (5A.10)$$

where,

a $= 1 + (1-f_{mo_2})32/(28f_{mo_2})$

f_{mo_2} = mass fraction of oxygen in the influent gas

= 0.2317 for air so that a = 4.79

= 1,000 for pure oxygen so that a = 1.00

32 & 28 = molar masses of oxygen and nitrogen respectively.

Substituting this equation for m_{dg} into Eq (5.5) and accepting as explained in Section 5.4 above that $Y_{CO_2} = 1.00$, yields for H'_{ge}

$$H'_{ge} = \frac{a}{32} \times \frac{OTR \times V_p}{OTE} \times C_{pg} \times (T_{ge} - T_{gi}) \quad (MJ/h) \tag{5.10}$$

where,

OTE = oxygen transfer efficiency {fraction}

T_{gi} = the temperature of the influent gas {°C}

T_{ge} = the temperature of the vent gas {°C}

With regard to the specific heat yield of the vent gas C_{pg}, it was mentioned in Section 5.4 above that not only could a constant value be accepted without a significant loss of accuracy for vent gas temperatures in the range of 20–60°C typical for pure oxygen oxygenated aerobic reactors, where the vent gas is predominantly CO_2, but also for air oxygenated reactors where the vent gas is predominantly N_2. Hence the following C_{pg} values are recommended for design and are used in the design example given in Section 5.6 below.

C_{pg} = specific heat for vent gas
= 0.0302 MJ/(kmol.°C) for air
= 0.0368 MJ/(kmol.°C) for pure oxygen.

In the design example presented later in this chapter, it will be seen that the contribution of H'_{ge} to H_{losses} is small, but because its magnitude is proportional to oxygen transfer rate, OTR, it follows from Eq (5.9) that H'_{ge} will increase with H_{bi}. Equation (5.10) shows that H'_{ge} can be reduced by maintaining a high oxygen transfer efficiency (OTE) at any specified OTR. Although minimizing this small heat loss by maintaining as high an OTE as possible will be of little assistance in achieving short retention times in the reactor, it will be demonstrated below (in Section 5.4.5) that at the same time as minimizing H'_{ge} a much greater heat loss i.e. the vent gas water vapour heat loss, H_{ve}, also is minimized.

At Milnerton, the temperature of the vent gas at the point of exit from the reactor, T_{ge}, was found to be 2 to 5°C lower than that of the reactor sludge, T_{se}, depending on whether the flow rate of vent gas was high or low respectively. This temperature difference was the result of heat loss from the vent gas in the reactor head space through the reactor head space walls before leaving the reactor. Because this temperature difference is small relative to the difference between the vent gas and the influent gas temperatures, for the purpose of this design example, it will be assumed that the vent gas temperature (T_{ge}) is the same as the reactor sludge temperature (T_{se}), i.e.

$$T_{ge} = T_{se} \quad (^{\circ}C) \tag{5.11}$$

5.4.4 The rate of wall heat loss from the reactor surfaces, H_{we}

The rate at which heat is lost to the ambient environment by conduction through and convection away from, the walls and surfaces of the reactor is typically described by an equation of the following form:

$$H_{we} = U \times A \times (T_{se} - T_{amb}) \quad (MJ/h) \tag{5.12}$$

where,

U	= the overall heat transfer coefficient	$\{MJ/(m^2 \cdot {}^{\circ}C.h)\}$
A	= the reactor surface area	$\{m^2\}$
T_{amb}	= the temperature of the ambient surroundings	$\{^{\circ}C\}$

This heat loss may easily be minimized by improving the quality and thickness of reactor lagging and even by locating the reactor within an enclosed building – exposure to rain and wind serve to increase H_{we} more than low ambient temperatures alone. At Milnerton, the use of 50mm polyurethane lagging and the mild climate kept H_{we} to a low value of about 20 MJ/h which is around a third of the mechanical pump heat input H_{mi}. Unless the lagging is insufficient and the variations in ambient climate are extreme, the heat loss rate, H_{we}, remains effectively constant over time.

5.4.5 The rate of heat loss due to the vaporization of water from the reactor sludge by the vent gas, H_{ve}

The rate of vent gas vapour heat loss, H_{ve} is described by:

$$H_{ve} = h_{fg} \times M_{wp} \quad (MJ/h) \tag{5.13}$$

where,

h_{fg} = the latent heat of vaporization of water = 2.358 {MJ/kg}

M_{wp} = the predicted rate of water vaporization {kg/h}

Although the latent heat of vaporization of water, h_{fg}, is a function of temperature, its variation over the vent gas temperatures encountered in dual digestion (50–65°C) is so slight that a constant value may be assumed. The average value recommended for design and which is used in the design example in Section 5.6 below is 2.358 MJ/kg.

The mass flow rate of water vapour leaving the reactor in the vent gas, M_{we}, is a function of the temperature, molar flow rate and the degree of water vapour saturation of the vent gas. Equations to calculate M_{wp} have been derived in Appendices 3D and 5A on the basis that the vent gas is saturated with water vapour and these equations satisfactorily predicted the measured M_{we} rates observed in the Milnerton aerobic reactor – see Section 4.7 and Fig 4.13. The relevant equations have been abstracted from the Appendices and are given below,

$$M_{wp} = 18 \text{ x } m_{dg} \text{ x } \frac{p_w}{p_{dg}} \quad \text{(kg/h)} \tag{5.14}$$

$$p_w = 10^{\left(-\frac{2238}{T_{ge} + 273} + 8.896\right)} \tag{5.15}$$

$$p_{dg} = P_T - p_w \tag{5.16}$$

where,

m_{dg} = the molar flow rate of dry vent gas {kmol/h}

p_w = the partial pressure of water in the vent gas {mmHg}

p_{dg} = the partial pressure of dry vent gas {mmHg}

P_T = the total absolute pressure of the vent gas {mmHg}

18 = the molar mass of water {kg/kmol}

As mentioned in Section 5.4.3 above, in Appendix 5A an equation is derived which allows calculation of the molar flow rate of the dry vent gas, m_{dg}, as a function of the oxygen transfer rate, OTR, the oxygen transfer efficiency, OTE, and the respiration quotient, Y_{CO_2}. The equation takes into account oxygenation with air, pure oxygen or oxygen enriched air and was given above as Eq (5A.10):

At Milnerton, H_{ve} was very small - at a 1.25 day retention time where the rate of increase in sludge heat content, H'_{se}, amounted to typically 250 MJ/h, H_{ve} was measured as 3.5 MJ/h (see Fig.4.16). The above Eqs (5.13) to (5.16) indicate that, as with H'_{ge}, H_{ve} is proportional to the rate of biological heat generation, H_{bi} through the OTR; larger rates of biological heat generation require larger oxygen supply rates which, in turn, produce more vent gas and hence more vaporization of water. This proportionality was observed at Milnerton and is illustrated in Fig 4.17.

At Milnerton, H_{ve} was a very small part of H_{losses} because pure oxygen was used and high transfer efficiencies (OTE > 0.8) were obtained which resulted in low molar flow rates of vent gas. Under these circumstances, minimizing H_{ve} further would not only have been difficult, but also of little benefit insofar as reducing the retention time was concerned. However, such small vapour heat losses are not inevitable in the design of the reactor. In a design for an air oxygenated aerobic reactor presented by Morgan and Gunson (1989), the rate of increase in sludge heat content was calculated to be 118 kW and the vent gas vapour heat loss as 580 kW. This is a rather extreme example where it was hoped that the vaporization of large amounts of water would serve to dewater and thicken the sludge. It is clear from Eq (5.7) that the loss of so much heat relative to the amount required to effect the desired increase in sludge heat content is completely contrary to achieving short retention times in the reactor.

5.5 MAXIMIZING HEAT SOURCES AND MINIMIZING HEAT SINKS

Having defined and quantified the heat balance terms in $H_{sources}$ and H_{losses} in Eqs (5.6) and (5.7), it is possible to approach the task of maximizing the $H_{sources}$ and minimizing the H_{losses} in order to achieve short retention time operation in the reactor. In this exercise, the rates of vent gas sensible heat loss, H'_{ge}, and reactor surface heat loss, H_{we}, are not considered. As will be demonstrated in the design example, H'_{ge} is insignificant and those factors that reduce its value also reduce the far larger heat loss, H_{ve}. Design specifications, such as good lagging, can cause the value of H_{we} to be small.

Like H_{we}, the rate of mechanical heat input, H_{mi}, is also determined by design specifications, in this instance by the size of a mixer or recirculation pump. Process economics require it to be kept small in order to reduce reactor operating costs, while larger values serve as a valuable heat source for operation at short retention times and also tend to favour high oxygen transfer rates. The relationship between H_{mi} and reactor retention time will be examined in the design example given in Section 5.6 below.

Much of the effort in the task of maximizing the heat sources and minimizing the heat losses therefore must be focused on maximizing the biological heat source, H_{bi}, and on minimizing the vapour heat loss, H_{ve}. However, because of the proportionality between these two rates, maximizing H_{bi} and minimizing H_{ve} appears to be paradoxical; should the OUR of a sludge be high, then to exploit this as best as possible, OTR should correspondingly also be high - with a high OUR and OTR in Eqs (5.8) and (5.9) respectively, increases H_{bi} but also increases the molar flow rate of vent gas [Eq (5A.10)] and hence H_{ve} [Eqs 5.13 to 5.16)].

In considering maximizing H_{bi} and minimizing H_{ve}, 4 factors merit attention:

(1) The relative significance of the oxygen transfer rate (OTR) and the oxygen transfer efficiency (OTE).
(2) The use of pure oxygen as opposed to air.
(3) The degree of water vapour saturation in the vent gas.
(4) The exploitation of a supplementary heat source such as heat exchange between reactor sludge effluent and feed.

5.5.1 The relative significance of oxygen transfer rate (OTR) and oxygen transfer efficiency (OTE)

The two parameters, oxygen transfer rate (OTR) and oxygen transfer efficiency (OTE), are fundamental to the paradox described earlier, viz. that of maximizing H_{bi} and minimizing H_{ve}. A high OTR will increase the rate of biological heat generation, H_{bi}, as long as the reactor remains oxygen limited [Eq (5.9)]. Accepting that high OTR's also result in large vent gas flow rates (and hence large H_{ve}), the cardinal role of oxygen transfer efficiency (OTE) in minimizing

H_{ve} [via m_{dg} in Eq (5A.10)] becomes apparent. It is only by maximizing OTE that m_{dg} (and hence H_{ve}) can be minimized. Failure to maximize OTE may lead to the situation where attempts to increase H_{bi} by increasing OTR cause H_{ve} to increase to a level where it has a significant effect on the retention time in the reactor. The relative effects of high OTR and low OTE in respectively increasing and decreasing the retention time are examined in the design example below.

5.5.2 Pure oxygen versus air

High OTR and OTE values can be achieved in a reactor oxygenated with pure oxygen and the resultant vent gas vapour heat losses will generally be small enough to be neglected. This clearly maximizes biological heat generation and minimizes vapour heat losses. In this respect air oxygenated reactors will be unable to match the performance i.e. shortness of retention time, of pure oxygen reactors unless supplementary heat sources are explored (see Section 5.5.4 below). The large volumes of nitrogen introduced into the sludge with air vaporize large masses of water, a situation which is exacerbated by even a slight decrease in OTE. Although OTE's of 100% have been reported by Wolinski for an air system, these efficiencies were achieved at low OTR's [0.076 to 0.132 kg/(m^3.h)], which necessarily imply long retention times. In contrast, at Athlone (Cape) where a 160 m^3 air oxygenated aerobic reactor has been operated over the past 2 years (Pitt, 1990), OTE's ranging from 0.10 to 0.26 and averaging around 0.14 were achieved at OTR's ranging from 0.14 to 0.21 kgO/(m^3.h) and averaging around 0.16 kgO/(m^3.h). The higher OTR's and higher OTE's apparently occurred together at times where a large foam layer (3m) built up on the reactor. At times when a large foam layer was absent, attempts to increase OTR by increasing the air supply rate led to a reduction in OTE. In general, the OTR and OTE achieved by an oxygenation system whether it be pure oxygen or air, is entirely dependent on the characteristics, mixing and geometry of the reactor. In order to transfer sufficient oxygen to enable the generation of the required biological heat, the OTR and OTE of the oxygenation system needs to be known and well defined. If the OTR and OTE are not known, then in the operation of the reactor, it will be found that the OTR and OTE will govern the biological heat generation and vent gas water vapour heat loss rates and hence also the minimum retention time. At Athlone, it was found that temperatures above 50° C could not be maintained at retention times less than 4 days.

5.5.3 The degree of water vapour saturation in the vent gas

The equations in Section 5.4.5 for calculating vent gas water vapour heat loss rate, H_{ve}, are based on the assumption that the vent gas is saturated with water vapour. Should this not be the case, then the actual value of H_{ve} will be smaller than the calculated values, in proportion to the degree of saturation. The conditions which increase the probability of vent gas water vapour saturation, also favour high oxygen transfer efficiencies, namely intimate contact between the sludge and the oxygenation gas i.e. fine bubble dispersion, good sludge mixing and long retention of the oxygenation gas in the reactor. Conversely, it is probable that at low oxygen transfer efficiencies the vent gas will be undersaturated with respect to water vapour and, accordingly, the vent gas vapour heat losses will not be as high as those predicted.

At Milnerton, where pure oxygen was used, the finding that the vent gas was saturated with water vapour at vent gas temperatures about 3°C lower than the reactor sludge is not unexpected due to the low vent gas flow rates per m³ reactor volume. At the air oxygenated Athlone plant, it was also found that the vent gas was saturated with water vapour at vent gas temperatures about 5°C lower than the sludge (Pitt, 1990). This observation seems to indicate that even with air at an OTR of 0.16 kgO/(m³.h) and OTE of 0.14, water vapour saturation of the vent gas can take place with the consequent large influence on H_{losses} by high values of H_{ve}.

5.5.4 Supplementary heat sources

The use of air for oxygenation may limit the minimum retention time at which the reactor can be operated. A large proportion of the heat sources may be consumed to compensate for the heat lost in the vent gas vapour rather than being used to increase the heat content of the sludge. Attempts to generate more biological heat to overcome this problem by increasing the influent air flow rate may, instead, exacerbate it if the increased air flow rate does not increase the biological heat generation rate but increases the vent gas water vapour heat loss rate instead; in this situation, the temperature of the reactor sludge will decrease unless the retention time is increased.

A means of avoiding high vent gas water vapour heat loss rates is to reduce the dependence on biological heat generation by exploiting a supplementary heat source. Such a heat source may be derived from the exchange of heat between the hot reactor effluent and cold influent feed sludges. A number of advantages are gained by reducing dependence on biological heat generation:

(1) Shorter retention times can be maintained.
(2) Oxygen limitation can be achieved more easily at lower oxygen transfer rates.
(3) Reactor operating costs can be decreased by reducing the oxygenation power requirements. In a pure oxygen system, lower operating costs will also result from reduced consumption rates of pure oxygen.

There are 2 constraints which limit the amount of heat available for heat exchange. The first concerns the anaerobic stage of the dual digestion system, which, up to now, has not been considered in the discussion of aerobic reactor design and the second concerns the method of operation of the heat exchanger.

In a dual digestion system, the anaerobic digestor is operated at mesophilic temperatures. As is the case with the reactor, the mesophilic digestor loses heat to its surroundings at a rate which increases in cold weather and decreases in hot weather. In order to maintain mesophilic temperatures, heat must be provided to the digestor at a rate which is equivalent to this rate of heat loss. This heat is supplied by the hot reactor effluent sludge. If heat exchange cools the reactor sludge too much, then not enough heat will be available for the anaerobic digestor and it will cool to below mesophilic temperatures. During cold winter months, when the heat loss rate from the digestor is large, the required digestor heat input rate may be so high that heat exchange is no longer possible.

Following the above reasoning the rate at which heat is available from heat exchange can be calculated with the following equation:

$$H_{ex} = \frac{V_p}{R_H} \text{ x } \rho_s \text{ x } Cp \text{ x } (T_{se} - T_{min}) \quad \text{(MJ/h)} \qquad (5.17)$$

where,

H_{ex} = the rate of heat available for heat exchange {MJ/h}
T_{min} = the minimum temperature at which the heat input rate to the digestor is sufficient to maintain mesophilic temperatures {°C}

With regard to the second constraint the heat exchanger can be operated either in a continuous flow or batch fed way. Batch heat exchange is considered to be more suitable for 2 reasons:

(1) Because batch feeding is employed in the reactor as a prerequisite for sludge pasteurization (see Section 5.2 above), batch heat exchange integrates more readily into the dual digestion system.

(2) Batch heat exchange allows easier control of the heat exchanged between reactor influent and effluent – it is easier to avoid cooling the reactor sludge to below T_{min}.

The maximum amount of heat that can be exchanged in a batch heat exchanger occurs when the temperatures of the hot reactor and cold feed sludges have equilibrated which for an ideal heat exchanger (no heat loss from it) is midway between the reactor influent (T_{si}) and effluent (T_{se}) sludge temperatures, i.e. for T_{si} and T_{se} at 20°C and 60°C respectively, about 40°C. This temperature is likely to be too low to maintain mesophilic temperatures in the anaerobic digestor. At Milnerton, in the winter all of the heat in the reactor effluent sludge at 1,25 d retention time and 60°C was required to maintain the digestor at 38°C at 21 days retention time; however in the summer, the digestor overheated under these conditions requiring the use of a heat exchanger (see Appendix 5B).

5.6 GRAPHICAL EXAMINATION OF THE DESIGN INTER-RELATIONSHIPS

In this section, a design example is presented to illustrate the effect of various reactor operating conditions on the retention time in the reactor. This is accomplished by plotting a series of curves displaying the retention time against oxygen transfer efficiency for 4 different oxygen utilization rates and for oxygenation with pure oxygen and with air. These curves are altered by considering:

(1) The effect of reducing the rate of mechanical heat input, H_{mi}.

(2) The degree of water vapour saturation in the vent gas.

(3) The utilization of heat exchange.

The design example is based on the 45 m³ reactor of the Milnerton dual digestion system. Mechanical heat is supplied to the reactor and heat is lost from the reactor surfaces at constant rates of H_{mi} = 70 MJ/h and H_{we} = 22.5 MJ/h. Feed sludge, reactor sludge, influent gas and vent gas temperatures are T_{si} = 20°C, T_{se} = 60°C, T_{gi} = 20°C and T_{ge} = 60°C. The total pressure of the vent gas, P_T, is 760 mmHg and the specific heat yield, Y_H, is 12.9 MJ/kgO.

5.6.1 Retention time as a function of oxygen transfer rate for different values of oxygen transfer efficiency

From Eq (5.7) it is possible to express retention time R_H in terms of oxygen transfer rate, OTR, and oxygen transfer efficiency, OTE, i.e.

$$R_H = \frac{1}{24} \times \frac{V_p \times \rho_s \times Cp \times (T_{se} - T_{si})}{H_{bi} + H_{mi} - H'_{ge} - H_{we} - H_{ve}} \tag{5.18}$$

Substituting the accepted values for this design example into the equations for H_{bi}, H_{mi}, H'_{ge}, H_{we} and H_{ve}, which describe the terms of the heat balance in terms of OTR and OTE, yields for;

The rate of biological heat generation, H_{bi}

From Equation (5.9), H_{bi} = 12.9 * 45 * OTR
= 580.5 * OTR

The rate of mechanical heat generation, H_{mi}

Accepted constant at H_{mi} = 70 MJ/h

The net rate of vent gas sensible heat loss, H'_{ge}

From Eq (5.10):

For pure oxygen: H'_{ge} = (0.0368.1/32) * OTR * 45/OTE * (60 – 20)
= 2.1 * OTR/OTE

For air: H'_{ge} = 0.0302.4.79/32) * OTR * 45/OTE * (60 – 20)
= 8.1 * OTR/OTE

The rate of heat loss from the reactor surfaces, H_{we}

Accepted constant at $H_{we} = 22.5$ MJ/h

The rate of vent gas vapour heat loss, H_{ve}

From Eq (5.15): $p_w = 10^{(-2238/(57+273) + 8.896)}$
$= 150$ mmHg

Hence, from Eq (5.16): $P_{dg} = 760 - 150$
$= 610$ mmHg

The dry vent gas molar flow rate, m_{dg}, from Eq (5A.10):

For pure oxygen:

$$m_{dg} = (1/32) * OTR * 45/OTE * (1 - OTE + OTE * 1)$$
$$= 1.41 * OTR/OTE$$

For air:

$$m_{dg} = 1/32 * OTR * 45/OTE * (4.79 - OTE + OTE * 1)$$
$$= 6.74 * OTR/OTE$$

Hence the value of H_{ve} can be calculated from Eqs (5.13) and (5.14):

For pure oxygen:

$$H_{ve} = 2.358 * 18 * 1.41 * OTR/OTE * 150/610$$
$$= 14.7 * OTR/OTE \quad \text{MJ/h}$$

For air:

$$H_{ve} = 2.358 * 18 * 6.74 * OTR/OTE * 150/610$$
$$= 70.3 * OTR/OTE \quad \text{MJ/h}$$

Substituting the above expressions and values for H_{bi}, H_{mi}, H'_{ge}, H_{we} and H_{ve} into Eq (5.18) yields 2 equations, 1 for pure oxygen and 1 for air, which express R_H in terms of OTR and OTE:

For pure oxygen:

$$R_H = \frac{1}{24} \times \frac{7531.1}{580.5 \times OTR + 70 - 2.1 \times \dfrac{OTR}{OTE} - 22.5 - 14.7 \times \dfrac{OTR}{OTE}} \text{ (days)} \qquad (5.19)$$

For air:

$$R_H = \frac{1}{24} \times \frac{7531.1}{580.5 \times OTR + 70 - 8.1 \times \dfrac{OTR}{OTE} - 22.5 - 70.3 \times \dfrac{OTR}{OTE}} \text{ (days)} \qquad (5.20)$$

From Eqs (5.19) and (5.20) retention time is plotted against oxygen transfer efficiency in Fig 5.1 for 4 different oxygen transfer rates: 0.1, 0.2, 0.3 and 0.4 kg/(m^3.h). The results shown in Fig 5.1 are discussed below at the end of this chapter together with 3 further graphs which are developed first.

5.6.2 The effect of reducing the rate of mechanical heat input, H_{mi}, on retention time

The mechanical heat input rate of 70 MJ/h at Milnerton was used to achieve oxygen transfer rates of ±0.4 kg/(m^3.h) and accounted for 27% of the heat sources at a 1.25 day retention time. In a design based on lower OTR's it is likely that a lower value of H_{mi} will be required to transfer the oxygen. Consequently, both heat sources, H_{bi} and H_{mi}, will contribute less to the rate of change of sludge heat content, H'_{se}. The relative effects on retention time, of H_{mi}'s of 0, 35 and 70 MJ/h for design OTR's of 0.4 and 0.2 kg/(m^3.h) are examined in Figs 5.2a and 5.2b.

5.6.3 The degree of saturation of the water vapour in the vent gas

In Chapter 4, Section 4.7, the degree of saturation of water vapour in the vent gas from the Milnerton reactor was examined, and it was concluded that the gas was fully saturated. In Section 5.5.3 of this chapter, it was suggested that those circumstances which cause low oxygen transfer efficiencies, such as poor mixing and insufficient sludge/gas contact, may lead to undersaturation of water vapour in the vent gas. The effect of such undersaturation would be to reduce the rate of vapour heat loss, H_{ve}, at low OTE values and to allow shorter retention time operation in the reactor than predicted by Eqs (5.19) and (5.20).

To observe the effect of this undersaturation theoretically, the terms (14.7 * OTR/OTE) and (70.3 * OTR/OTE) which calculate the rates of vapour heat loss in Eqs (5.19) and (5.20) respectively, each are multiplied by the term (1+2*OTE)/3. This term progressively decreases the water vapour heat loss down to a minimum of 1/3 of the saturation value, as the oxygen transfer efficiency decreases to zero. The effects of full saturation and progressively decreasing saturation are examined in Fig. 5.3 for OTR's of 0.1 and 0.4 kg/(m³.h).

5.6.4 Heat exchange

The rate at which heat may be transferred during heat exchange between effluent and feed sludges, H_{ex}, may be estimated with Eq (5.17). If the heat exchanger is assumed to be perfectly lagged, with no heat losses to the surroundings, then the respective decrease and increase in the effluent and influent sludge temperatures will be equal. At Milnerton, the potential for heat exchange was restricted by the relatively high rate of heat loss from the anaerobic digestor – see Section 5.5.4 – and heat exchange could only be used during the hot summer months. Even then, cooling the effluent reactor sludge to less than 44°C might have jeopardized the maintenance of mesophilic conditions in the digestor. Thus, in this example, the reactor effluent sludge is cooled by 16°C in the heat exchanger. The corresponding 16°C increase in feed sludge temperature, causes the heat content of this feed sludge to increase at a rate which is 40% of the required rate of heat content increase, H'_{se} [see Eq (5.1)].

There are 2 methods of incorporating H_{ex} into Eq (5.7) [and hence Eqs (5.19) and (5.20)]. In the first, H_{ex} may be treated as a supplementary heat source and included as a term in $H_{sources}$. In the second method the feed sludge temperature, T_{si}, is increased to account for the feed sludge heat content increase due to heat exchange. In this example, T_{si} increases by 16°C from 20 to 36°C in

Eq (5.7) causing the numerator of 7531.1 in Eqs (5.19) and (5.20) to decrease by 40% to 4518.7. The second method is preferred because it is more straightforward than the first and results in more accurate prediction of the reactor sludge temperature/time profile in computer simulations (which is described in Chapter 6). The plot of retention time against OTE with and without heat exchange is presented in Fig 5.4 for oxygen transfer rates of 0.1 and 0.4 kg/(m^3.h).

5.7 DISCUSSION OF THE DESIGN INTERRELATIONSHIPS

In Fig 5.1, the retention time R_H is plotted versus the oxygen transfer efficiency (OTE) for oxygen transfer rates (OTR), of 0.10, 0.20, 0.30 and 0.40 kgO/(m^{3}h) for pure oxygen (solid lines) and air (dotted lines).

The lines have the shape of a hyperbola, one for each particular OTR, and the greater the OTR, the closer the hyperbola to the horizontal and vertical axes. The lines are hyperbolic because for a fixed OTR the heat sources are constant because H_{bi} and H_{mi} are constant and this heat is lost principally via two varying routes, i.e. via hot effluent sludge which is related to the retention time or via the vent gas as water vapour and gas sensible heats, which is related to the OTE. So in Fig 5.1, increasing retention time means reduced heat loss via hot effluent sludge, and increasing OTE means reduced heat loss via the vent gas. The wall heat loss does not really come into this because like the mechanical heat input it is constant. This means that at a particular OTR, i.e. constant heat input, a high OTE and short R_H (bottom right of Fig 5.1) implies that the heat loss via the vent gas is small and that most of the heat is lost via the hot effluent sludge and a low OTE and long R_H (top left of Fig 5.1) implies that the heat loss via the hot effluent sludge is small and that most of the heat is lost via the vent gas. With this theoretical background to Fig 5.1 in mind, the design of air and oxygen oxygenated reactors is now examined in a more practical way below.

From Fig 5.1, clearly the greater the OTR the shorter the retention time R_H. This demonstrates the importance of achieving high OTR's, with both pure oxygen and air, if the reactor is to be operated at short R_H; at high OTR's, high rates of biological heat generation are achieved. Now as outlined in Section 5.4.1 above, the rate of biological heat generation is limited by the sludge biological oxygen utilization rate (OUR) and increasing the OTR above this oxygen utilization rate, does not increase the rate of biological heat generation. Indeed, as will be discussed below, it will only serve to increase the rate of vent gas heat loss and decrease the oxygen transfer efficiency. Furthermore when OTR reaches OUR, a condition of oxygen sufficiency in the sludge is created instead of the desired condition of oxygen limitation and therefore the apparent sludge pretreatment ability of the reactor for anaerobic digestion enhancement is lost.

For a given OTR (and H_{mi}), if the oxygen transfer efficiency (OTE) is > 0.15 for pure oxygen and > 0.40 for air, practically all the heat generated will be lost via the hot effluent sludge and very little heat is lost via the vent gas (as water vapour and sensible heat). Hence retention time is independent of OTE (the controlling factor in the vent gas heat loss) and is fixed by the amount of heat generated which practically all can be lost via hot sludge in the effluent. Consequently at high OTE's the shortest retention times are achieved.

For OTE's less than 0.15 and 0.40 for pure oxygen and air oxygenation respectively, retention time becomes increasingly sensitive to decreasing OTE. This happens because below these OTE's, the vent gas water vapour and sensible heat losses, where the former is 7.0 and 8.7 times higher than the latter for oxygen and air oxygenated systems respectively, become increasingly larger with respect to the heat lost via the hot effluent sludge and this increased vent gas heat loss needs to be compensated for by a reduction in heat loss via hot effluent sludge which means an increase in retention time. As OTE decreases further, requiring more and more air or oxygen to pass through the reactor to achieve the specified OTR, the heat loss via the vent gas becomes the dominant heat loss, and the heat loss via the hot effluent sludge to maintain sludge temperature of necessity decreases to zero. In Fig 5.1 for pure oxygen, if OTE is 0.05 then all the heat generated is lost via the vent gas and R_H tends to infinity; for air, this occurs at an OTE of about 0.15. The significance of this point is that it represents the state of operation where the rate of vent gas heat loss begins to exceed the rates of heat generation. In practice, it might occur when attempts to increase the OTR by increasing the air or oxygen supply rate are unsuccessful, either due to oxygen transfer limitations of the aeration equipment or due to the desired OTR exceeding the sludge biological oxygen utilization rate (OUR). In these events, an increase in oxygen/air supply would cause the vent gas heat losses (water vapour and sensible) to increase while H_{bi} remains constant, and as an increase in retention time will no longer compensate sufficiently for the additional heat lost via the higher air/oxygen supply rate, the reactor sludge temperature will decrease below 60° C.

To demonstrate the design implications of a pure oxygen and air oxygenated aerobic reactor, the results observed at Milnerton reported in this thesis for pure oxygen and those reported by Pitt (1990) for the Athlone air oxygenated reactors are plotted in Fig 5.1: At Milnerton OTE's ranging between 0.80 to 1.00 at retention times 1.25 to 3 days were obtained with OTE increasing as R_H increased (see right side of Fig 5.1); at Athlone, OTE's between 0.10 and 0.26 averaging at 0.14 and retention times between 3 and 6 days were achieved with OTE increasing as R_H increased (although the foam layer had a greater influence on this than R_H) (see left side of Fig 5.1). These two situations of oxygen and air oxygenated systems show that for an oxygen oxygenated reactor, the heat losses are principally via hot effluent sludge allowing short retention times whereas with an air oxygenated reactor, the heat losses are principally via the vent gas thereby requiring longer retention times.

The plots in Fig 5.1 are based on the assumption that the vent gas is saturated with water vapour, an assumption which, even though found to be the case for the Athlone air oxygenated rate at OTE's of 0.10 to 0.25, may not necessarily apply in every case where low OTE's are encountered. Figure 5.3 shows plots of R_H vs OTE for pure oxygen and air oxygenated systems at OTR's of 0.1 and 0.4 kg/(m^3.h) for saturated and undersaturated vent gas conditions, where the undersaturated case was modelled on the assumption that as OTE decreased from 1.0 to 0.0 so the degree of saturation decreased from 100% to 33%. From Fig 5.3 it can be seen that if the degree of vent gas vapour saturation, and thus the rate of vent gas heat losses, decrease with oxygen transfer efficiency, then the point where the vent gas heat losses began to become significant with respect to the hot effluent sludge heat losses, shifts to lower OTE values. That this is so is clearly because reducing the degree of water vapour saturation reduces the vent gas heat losses and therefore allows stable operation at lower OTE's than if the vent gas were saturated. However, undersaturation does not reduce the minimum retention time operation at high OTE's because at these OTE's, the heat lost via the hot effluent sludge is much greater than the heat lost via the vent gas; here as before the OTR plays the major role in fixing the minimum retention time.

The rate of mechanical heat input, H_{mi}, contributes directly to the heat sources. Increasing this rate does not trigger increased heat losses in the way that increasing H_{bi} triggers increased vent gas heat losses. Although process economics require H_{mi} to be minimized, it is likely that high oxygen transfer rates will only be achieved at the expense of a high power consumption to dissolve the large mass of oxygen per m^3 reactor volume and to mix the reactor sludge. Thus the advantage of a high biological heat generation rate arising from a high OTR will be supplemented by a high mechanical heat input rate. Conversely, the combination of a low OTR and a low H_{mi}, will severely compromise any effort at achieving a low retention time operation because this reduces the two available heat sources. This is illustrated in Figs 5.2a and 5.2b where reducing H_{mi} can be seen to have a much greater effect on retention time at a low OTR [0.2 kg/(m^3.h)] (Fig 5.2a) than at a high OTR [0.4 kg/(m^3.h)] (Fig 5.2b).

In cases where a low OTR and a low H_{mi} are combined in design, short retention operation will not be possible unless a supplementary heat source, such as heat exchange, is exploited. The heat exchange situation is illustrated in Fig 5.4. and the effect of heat exchange is to reduce the heat sources by about 40%, while still maintaining a pasteurization temperature of 60° C, because the heat content of the sludge entering the reactor is increased. This increased heat content of the feed sludge is sufficient to enable the retention time to be reduced by 40%. In this example, where the full rate of mechanical heat input of 70 MJ/h was applied, heat exchange reduced the retention time for air oxygenation at an OTR of 0.1 kg/(m^3.h) and an OTE of 0.3, from 3.7 to 2.2 days. Had an H_{mi} of 35 MJ/h been applied under the same conditions (as in Fig 5.2a), heat exchange would have reduced the retention time from 6.1 to 3.5 days.

As with H_{mi}, heat exchange contributes directly to increasing the sludge heat content without leading to increased rates of heat loss. While this contribution was limited to 40% of the required heat sources, additional heat could be derived from other sources such as the combustion of all, or part of, the methane produced in the anaerobic stage of the process. This source alone might be sufficient to produce all of the heat required, giving rise to a situation where biological heat generation could be completely dispensed with. In such a situation, the 2 stage process would no longer be a dual digestion system but would become the combination of a sludge pre–pasteurizer and an anaerobic digestor.

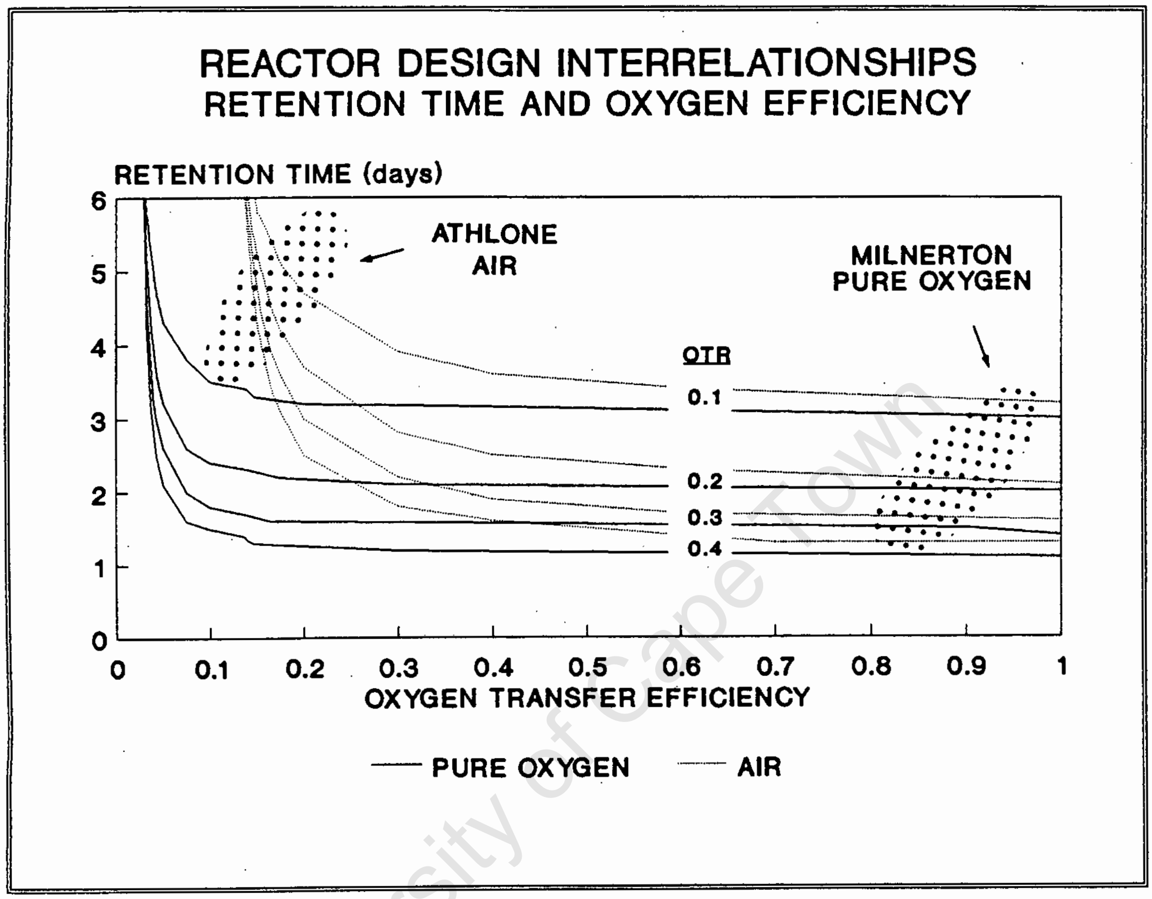


Fig 5.1: Hydraulic retention time in the aerobic reactor as a function of oxygen transfer efficiency, OTE, for 4 different oxygen transfer rates using pure oxygen and air. The temperature of the reactor sludge is 60°C. The operating ranges of R_H and O_{eff} for the pure oxygen reactor at Milnerton and the air oxygenated reactor at Athlone are denoted by the shaded areas – note that the reactor sludge temperature in the Athlone reactor is 50°C.

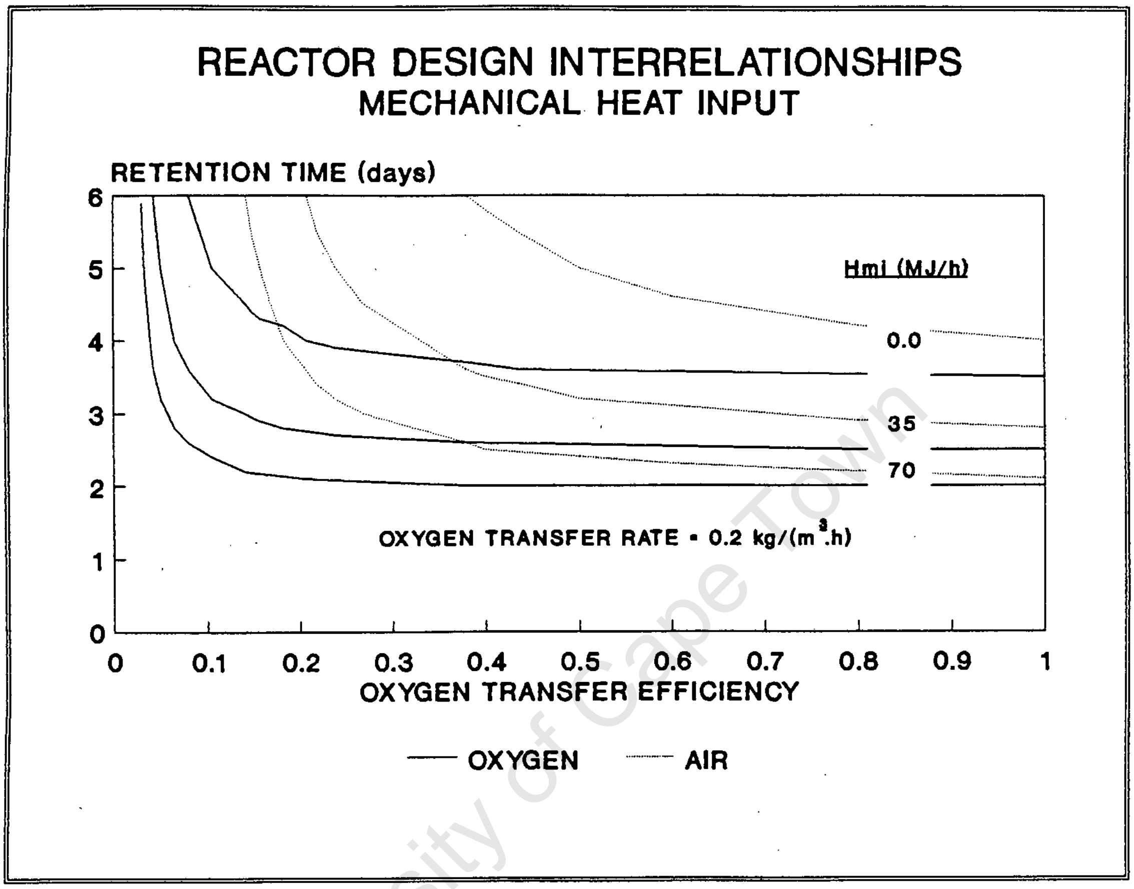
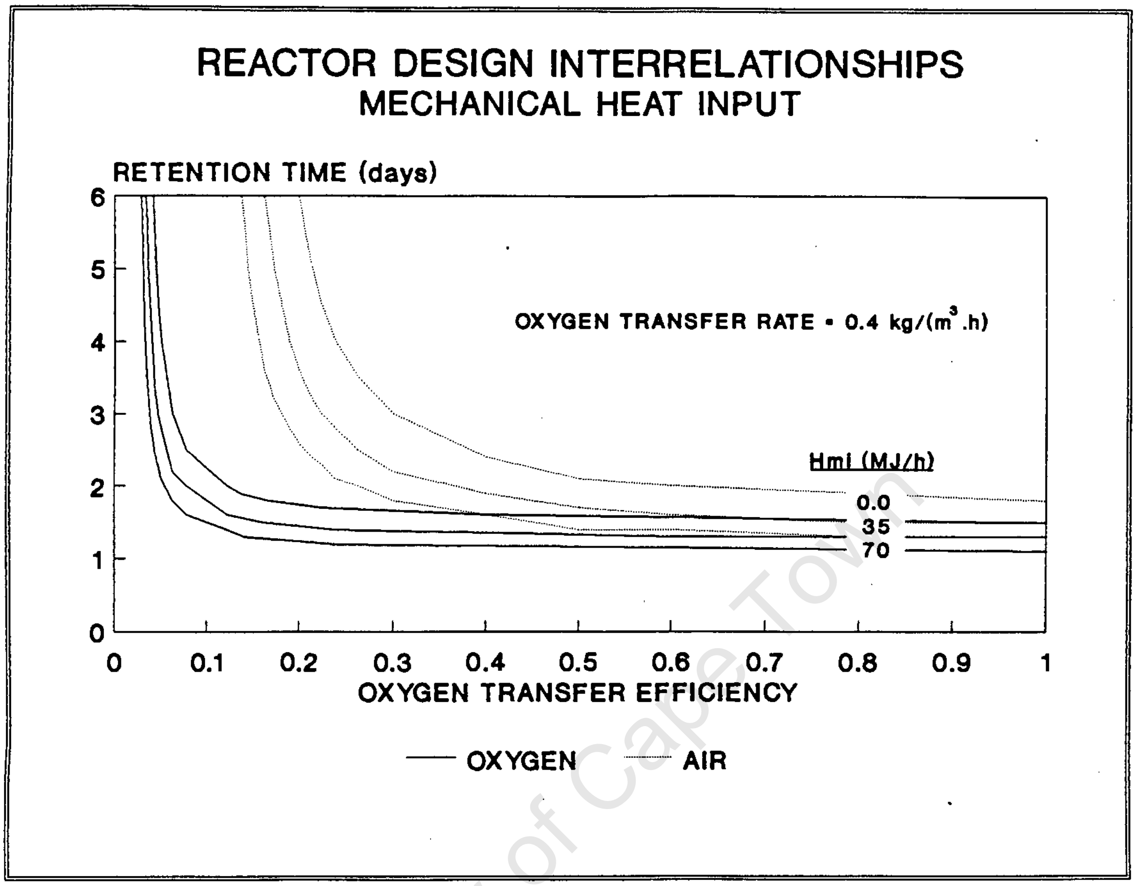


Fig 5.2(a): The effect of mechanical heat input, H_{mi}, on hydraulic retention time in the aerobic reactor with an oxygen transfer rate of 0.2 kg/(m³.h) using pure oxygen and air.



<u>Fig 5.2(b)</u>: The effect of mechanical heat input, H_{mi}, on hydraulic retention time in the aerobic reactor with an oxygen transfer rate of 0.4 kg/(m^3.h) using pure oxygen and air.

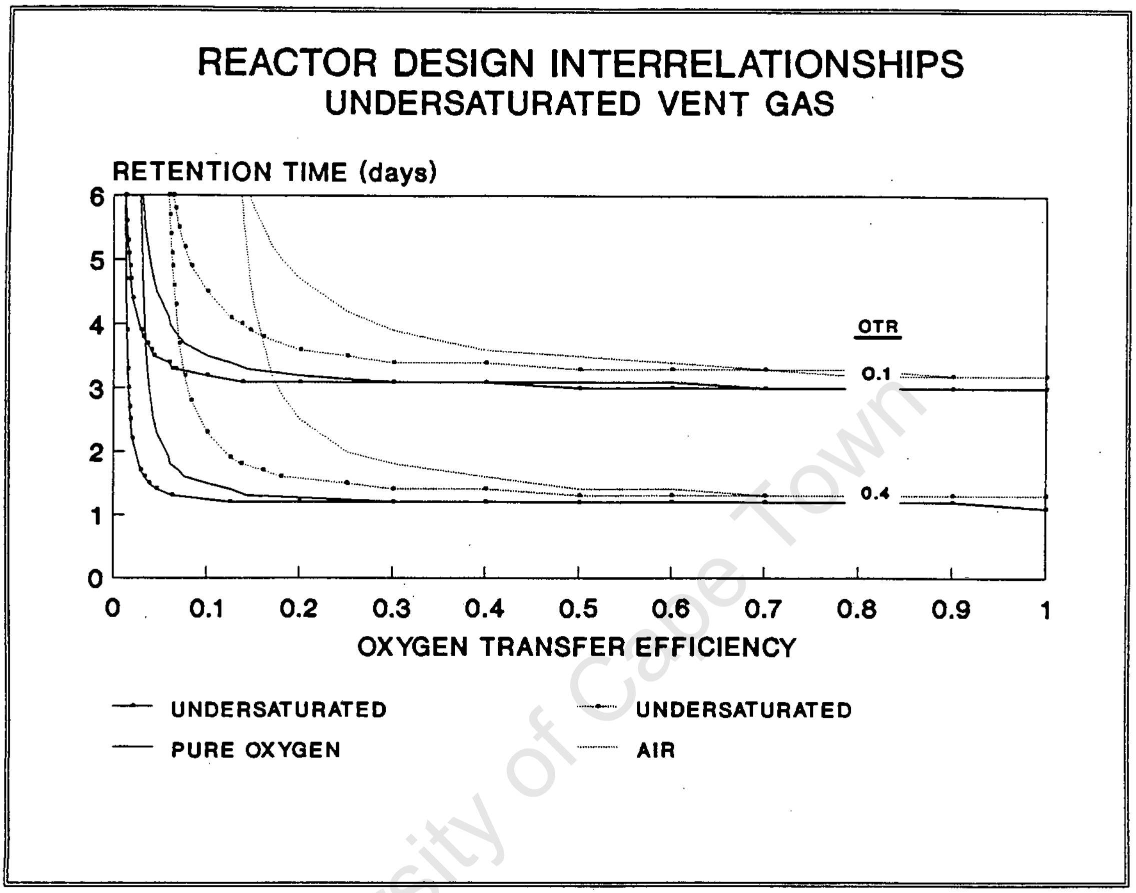


Fig 5.3: The comparative effects of fully and partially saturated reactor vent gas on hydraulic retention time in the aerobic reactor for oxygen transfer rates of 0.1 and 0.4 kg/(m^3.h) using pure oxygen and air.

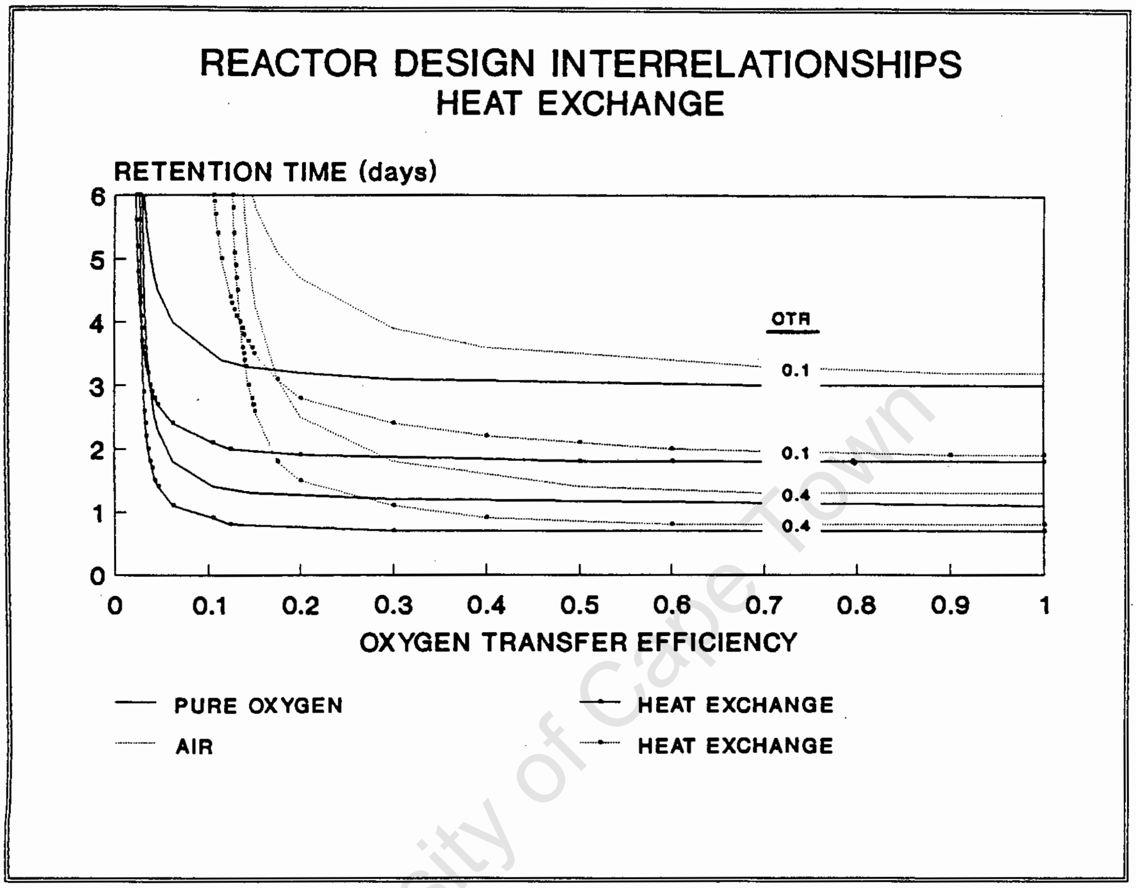


Fig 5.4: The effect of employing heat exchange between the reactor feed and effluent sludges, on hydraulic retention time in the aerobic reactor for oxygen transfer rates of 0.1 and 0.4 kg/(m³.h) using pure oxygen and air.

Dispensing with biological heat generation, and therefore sludge oxygenation, might have an impact on the sludge pretreatment in the reactor. While sludge pasteurization could still be accomplished, the reported potential of the reactor to significantly enhance the performance of the subsequent anaerobic stage might be lost. It is on this potential that much of the justification for employing the biologically heated aerobic reactor in the dual digestion process rests.

5.8 CONCLUSIONS

In this chapter design considerations for pure oxygen and air oxygenated aerobic reactors in dual digestion have been outlined.

Practical aspects of design such as sludge level and temperature control, dealing with foam, etc. were not discussed; for the Milnerton reactor, the practical aspects and experience gained are discussed in detail in Appendix 5B.

Accepting that the objectives of the aerobic reactor are sludge pretreatment through oxygen limitation and pasteurization by heating to thermophilic temperatures (~ 60° C), it is demonstrated with the aid of the steady state heat balance that 3 parameters are of crucial importance to design. The design centres on minimizing the retention time through maximizing the heat sources (biological heat generation and mechanical heat input) and minimizing the heat losses (vent gas water vapour and sensible heats). The 3 parameters are:

(1) The oxygen utilization rate (OUR) of the sludge which fixes the maximum biological heat generation rate of the sludge.

(2) The oxygen transfer rate (OTR) of the oxygenation system which is directly related to the actual biological heat generation rate through the specific heat yield (Y_H). The OTR should be less than the OUR to ensure sludge pretreatment through oxygen limitation.

(3) The oxygen transfer efficiency (OTE) which for a particular pure oxygen or air oxygenation system determines the volumetric flow rate of the vent gas ; the lower the OTE the greater the vent gas flow rate for a given OTR and hence the greater the vent gas heat losses via water vapour and gas sensible heats.

For a certain OTR and mechanical heat input which fix the heat sources, if OTE is high vent gas heat losses are small with the result that most of the heat generated can be lost via hot effluent sludge thereby allowing short retention times; if OTE is low, vent gas heat losses are high with the result that much less heat can be lost via the hot effluent sludge thereby forcing operation at longer retention times. At Milnerton where the sludge OUR was observed to be around 0.4 kg/(m^{3}h) and high OTR's of 0.35 kg/(m^3.h) were maintained, the use of pure oxygen and the attainment of high OTE's (0.8 to 1.0) with the pure oxygen, resulted in very low vent gas heat losses with the result that most of the heat generated could be lost via hot effluent sludge thereby allowing short 1.25d retention time operation and still readily meeting the pretreatment oxygen limitation and pasteurization temperature (>60°C) requirements. In contrast, at the air oxygenated Athlone reactor, where the oxygenation system achieved OTR's of around 0.16 kg/(m^{3}h) with OTE's around 0.14, the major part of the heat generated was lost via the vent gas with the result that only a minor part of the heat could be lost via hot effluent sludge thereby forcing long 4-6 days retention time operation to maintain temperatures above 50°C.

Biological heat generation has the drawback that increasing its rate requires increases in OTR, and increases in OTR require increases in oxygen or air supply rates which, even if OTE is not reduced by the increase, causes larger vent gas heat losses. This is particularly the case with air, where a large proportion of the gas passing through the system is inert nitrogen gas, coupled with the problem that increases in air supply usually cause decreases in OTE. In contrast to biological heat generation, alternative heat 'sources' such as heat exchange between reactor effluent and feed sludge or increased mechanical heat input, increase the heat sources without increasing the heat losses, and therefore allow a *pro rata* reduction in retention time.

The design considerations are based on the steady state heat balance which accepts that the reactor sludge temperature is constant and therefore applies only to a continuously fed reactor. Pasteurization requires the reactor to be batch fed, resulting in a continuously changing sludge temperature over about 3°C in a saw-tooth pattern, decreasing at times of feeding and increasing between the feeding times, with the result that insofar as the heat balance is concerned the reactor is not at steady state. However the error arising from assuming steady state conditions is small and well justified by the simplifications in the heat balance it affords. The saw-tooth temperature profile can only be determined by solving the unsteady state heat balance by a forward integration method. This is done in Chapter 6, where a computer programme is described which is able to simulate the temperature profile of the batch fed aerobic reactor. The simulation programme also allows the specified conditions for sludge pasteurization to be defined - in the steady state design it was possible to specify only the required pasteurization temperature, not the undisturbed sludge detention period. The unsteady state simulation permits this detention time to be defined in terms of the interval between consecutive batch feeds.

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CHAPTER 6

A MATHEMATICAL MODEL TO SIMULATE THE CHANGE OF TEMPERATURE WITH TIME IN A BATCH FED AEROBIC STAGE OF A DUAL DIGESTION SYSTEM

ABSTRACT

In this chapter, procedures are developed which enable the profile of the reactor sludge temperature versus time in a batch fed aerobic stage of a dual digestion system to be simulated. By the solution of an unsteady state heat balance across the reactor over a small discrete time interval, Δt, the increment or decrement in reactor sludge temperature may be calculated over this time interval. Repetitive computation of the unsteady state heat balance over a large number of time intervals allows the profile of the reactor sludge temperature to be plotted against time. From those procedures a computer model is developed which calculates and plots the sludge temperature versus time in a reactor which is oxygenated either by air/oxygen enriched air or by pure oxygen. The model is shown to accurately predict the temperature/time profiles in the aerobic reactor of the Milnerton dual digestion system for a range of different operating conditions.

6.1 INTRODUCTION

In the previous Chapter, design considerations and a design procedure for the design of the aerobic reactor in the dual digestion system were presented. The considerations and the procedure were developed from principles observed in the results of the research on the Milnerton aerobic reactor. Although this reactor was oxygenated with pure oxygen, by building on the heat balance principles observed to operate in the Milnerton aerobic reactor, the design procedure was extended and generalized to include oxygenation with air. In the interests of simplicity for design, the design procedure was based on the steady state heat balance which in effect accepts that the sludge flow through the digestor is continuous. This approach allowed a constant temperature for the aerobic reactor to be calculated by balancing the heat sources and heat sinks. However, due to the requirements of pasteurization, the aerobic reactor will not be operated under continuous flow conditions but under batch draw and fill conditions. This method of operation causes a moment by moment imbalance between the heat sources and heat sinks and results in a continually changing reactor sludge temperature, which takes the form of a distinctive saw-tooth temperature versus time profile (see Fig 6.1); during the addition of a batch of cold feed sludge (phase 1) the reactor sludge temperature decreases rapidly (phase 2) and during the heating phase (phase 3), it increases gradually through biological and mechanical heating until the next batch of feed sludge is added.

Although of great value for design, the steady state heat balance is incapable of predicting the temperature versus time profile under batch draw and fill conditions; that is, even though it is capable of accurately predicting the average reactor temperature, it cannot predict the changes of temperature resulting from the batch draw and fill feeding. To do this, the unsteady state heat balance needs to be solved moment by moment, with the heat surplus or deficit increasing or decreasing the reactor sludge temperature respectively. In this investigation, a mathematical model in the form of a computer programme was developed which is capable of predicting the moment by moment reactor sludge temperature versus time profile of the batch fed aerobic reactor. The aim of this Chapter is to describe the algorithm on which this computer model is based.

Although the computer model has been developed on the basis of information observed on the aerobic reactor of the Milnerton dual digestion system, it is completely general and may be applied to the aerobic reactor of any dual digestion system provided that the following conditions are met:

(1) The reactor is batch fed using a draw and fill system.[1]
(2) The reactor is continuously mixed.
(3) The reactor is continuously oxygenated with air, pure oxygen, or with pure oxygen enriched air.

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[1] Although developed for the aerobic reactor in dual digestion, which for pasteurization reasons is batch fed, the computer model also can be used to simulate continuous flow aerobic reactors (short retention time) and digestors (long retention time) at any temperature between 10° C and 80° C by making very short batch cycles (say 15 minutes) and letting the feeding phase take up the major portion of the cycle (say 10 minutes) and the transfer phase the remainder of the cycle so that there is no linear heating phase. The oxygen and heat balances calculated by the model under these conditions would be valid. The only drawback is that for the digestion case, the VS removal is not given because the biological heat generation rate is based on oxygen consumption rate through the specific heat yield and not on VS removal.

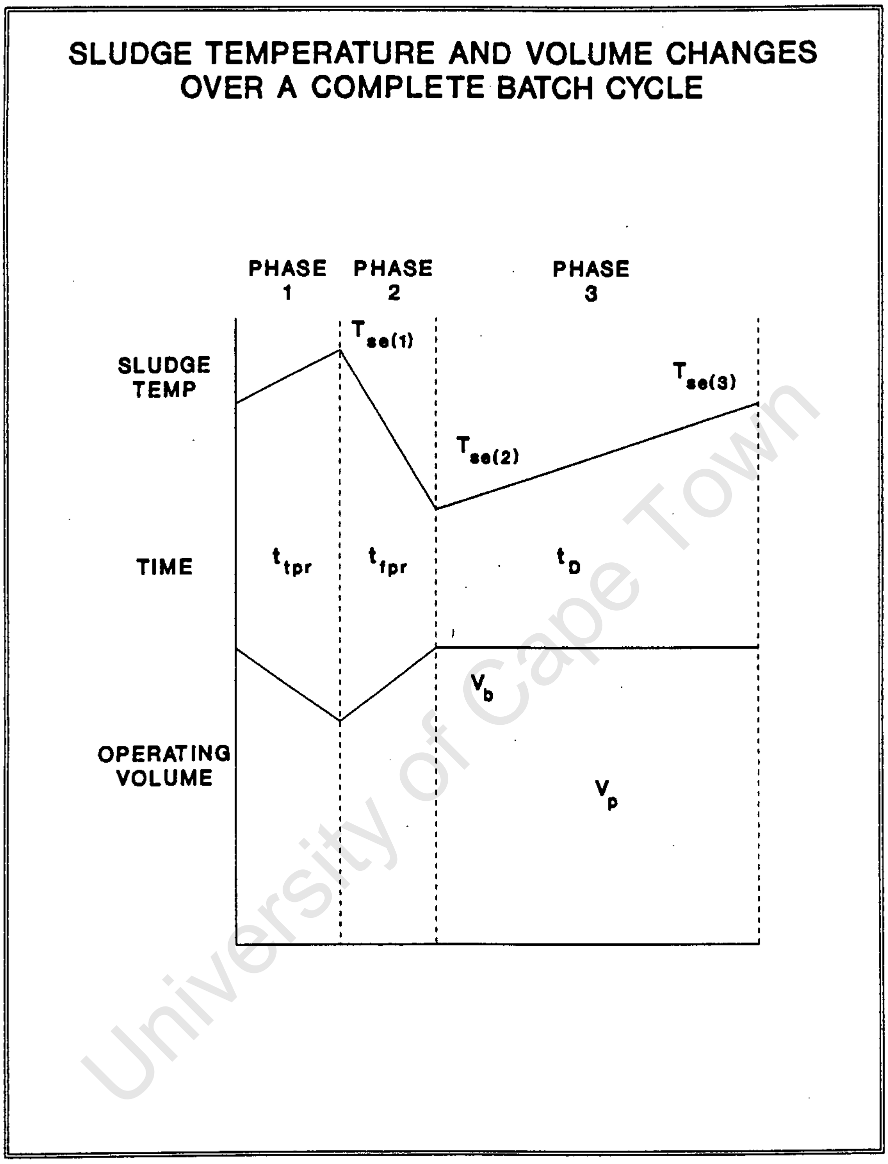


Fig.6.1: The typical sludge temperature and volume changes which occur in a batch fed aerobic reactor over the 3 phases of a complete batch cycle.

6.2 THE BATCH CYCLE

6.2.1 The phases of a batch cycle

A single, complete batch cycle is illustrated in Fig.6.1. The cycle consists of 3 separate phases – a transfer phase (phase 1), a feeding phase (phase 2) and a heating phase (phase 3). Each phase differs from the other 2 in respect of the changes in both the reactor sludge temperature and operating volume which occur during that phase. Also, the duration of the phases may be different, with the heating phase typically lasting for a longer period of time than the transfer and feeding phases.

Phase 1 – the transfer phase. A measured batch volume, V_b, of hot, treated sludge is pumped out of the reactor and transferred to the anaerobic stage of the dual digestion system. During this time the reactor operating volume, V_p, decreases by the amount V_b m³. The reactor sludge temperature, T_{se}, continues to increase gradually. The maximum temperature in a steady state batch cycle occurs at the end of this phase and is termed $T_{se(1)}$ in Fig 6.1.

Phase 2 – the feeding phase. Immediately upon cessation of phase 1, a batch volume of V_b m³ of cold feed sludge is pumped into the reactor restoring the sludge volume to the full operating volume. The sludge temperature, T_{se}, decreases rapidly as the cold feed sludge mixes with the hot reactor sludge. The minimum temperature in a steady state batch cycle occurs at the end of this phase and is termed $T_{se(2)}$ in Fig 6.1.

Phase 3 – the heating phase. For the remainder of the batch cycle sludge is neither transferred from nor fed to the reactor. Over phase 3 the sludge operating volume, V_p, remains constant. The sludge temperature, T_{se}, increases gradually as the sludge enthalpy increases with the input of energy from mixing and biological reactions in the sludge.

By examining the appearance of the saw-tooth profiles in single and successive batch cycles in a well mixed aerobic reactor, it is possible to derive useful qualitative and quantitative information about the operation of the reactor.

6.2.2 Qualitative information derived from the saw tooth profile

6.2.2.1 *Steady state operation*

It was stated in Chapter 3 that the aerobic reactor of the Milnerton dual digestion system was never at a steady state at a defined moment in time due to the system of batch feeding. Steady state in such a reactor can only be defined in terms corresponding times in successive batch cycles. Hence if the specific temperature at corresponding points in successive batch cycles, (such as $T_{se(1)}$ in Fig 6.1) are equal, then the reactor may be considered to be at steady state. This is illustrated in Fig 6.2(a).

6.2.2.2 *Biological heat generation and biological death in the reactor*

During phase 3, the enthalpy of the sludge, H_s, and hence the temperature of the sludge, T_{se}, increases with time due to the flow of heat into the sludge. This heat is derived from mechanical (mixing) and biological sources. The greater the rate of addition of this heat, the steeper the slope of the sludge temperature versus time profile in phase 3. If the mechanical heat source is continuous, then the above mentioned slope will vary as the rate of biological heat addition is changed, which, if the reactor is oxygen limited, can be controlled by controlling the oxygen supply rate.

For example, the slope may be increased by increasing the rate of oxygen supply to the sludge, as this will cause biological heat to be generated at a greater rate. The saw-tooth profiles of successive batches will curve upwards as shown in Fig 6.2(b). The steepest slope is achieved when biological heat is being generated at its maximum rate – see Section 6.3.2.1.

Conversely, if biological 'death' of the sludge occurs due to the action of a biocide or from the cessation of oxygenation, then biological heat generation ceases and only the mechanical heat input increases the reactor sludge temperature. Under these conditions the saw-tooth profiles of successive batches will decrease in a step wise–manner as shown in Fig 6.2(c).

6.2.2.3 *Reactor sludge mixing*

The use of multiple temperature probes at different locations in the reactor sludge to gauge the quality of sludge mixing has been discussed in detail in Chapter 3, Section 3.3.4.3 and Appendix 5B, Section 5B.1.4.1. The reading of temperature with time from each one of the probes will generate a saw tooth profile. If each of these profiles is identical, then mixing energy will be evenly dispersed throughout the reactor. The rapidity with which feed sludge is blended with the reactor

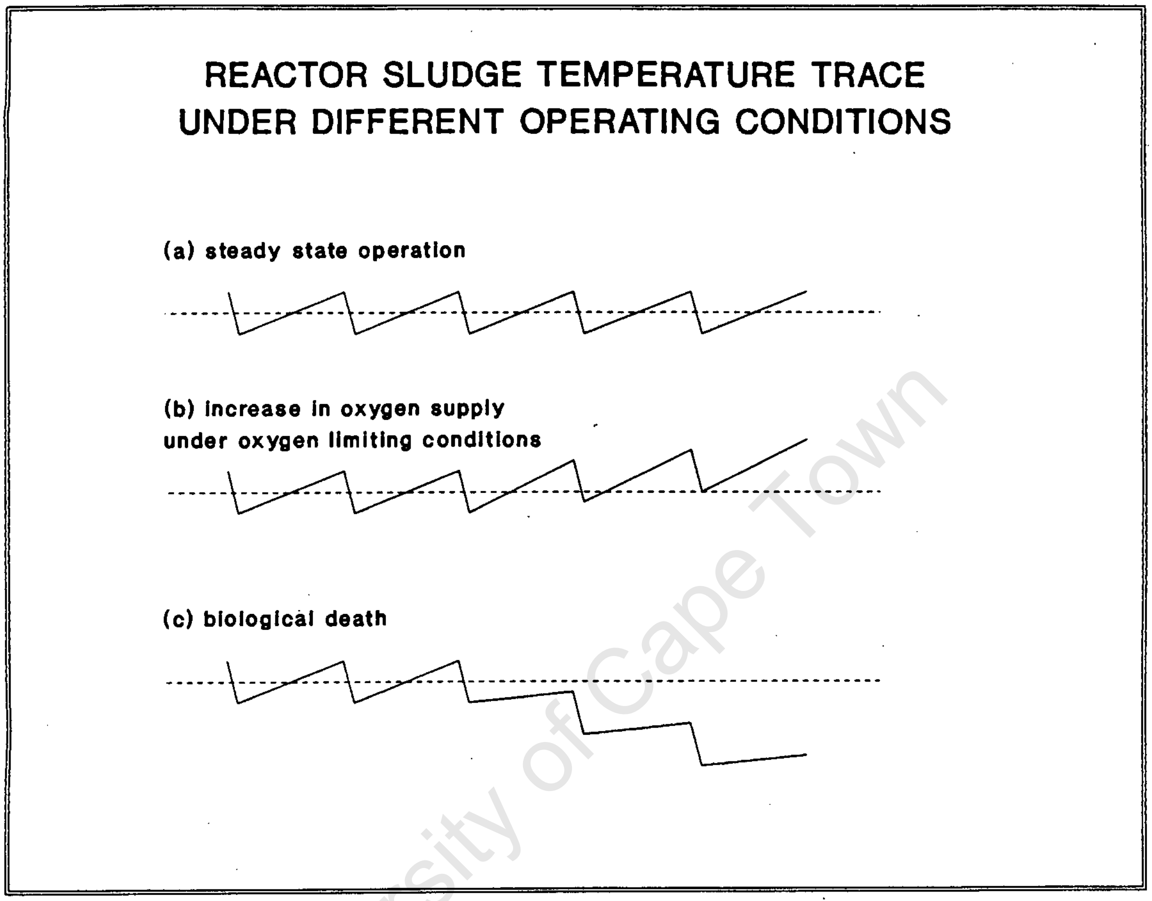


Fig.6.2: The manner in which 3 different operating conditions are reflected in the saw tooth profile of sludge temperature versus time in the batch fed aerobic reactor of a dual digestion system.

contents is indicated by the temperature profile from an individual probe during phase 2 and the early part of phase 3. Rapid mixing will cause a decrease in sludge temperature to commence and cease immediately at the start and end respectively of phase 2.

6.2.3 Quantitative information derived from the saw tooth profile

6.2.3.1 *The batch fraction, f_b, and sludge temperature decrease during phase 2*

The batch fraction is the ratio of the volume of a single batch of feed sludge and the full operating volume of the reactor. The amount by which the reactor sludge temperature, T_{se}, decreases during the feeding phase is a function of the batch fraction, f_b, and the difference between the temperature at the end of the transfer phase $T_{se(1)}$ and the feed sludge temperature, T_{si}. For given values of $T_{se(1)}$ and T_{si}, the decrease in T_{se} is directly proportional to f_b. The larger the value of f_b, the larger the value of the temperature decrease.

6.2.3.2 *The cycle time, t_c, and sludge pasteurization*

The cycle time, t_c, is equal to the duration of a complete batch cycle as shown in Fig 6.1. The value of the cycle time, t_c, determines the duration of the heating phase of the batch cycle, t_D, and this is an important criteria in defining effectiveness of sludge pasteurization in the reactor. Pasteurization is generally only deemed to be effective if the sludge is maintained at a specified temperature for a specified length of time, t_D. At Milnerton, for example, this meant keeping the sludge in the reactor for at least 2h at a temperature of not less than 55° C. This period of sludge retention is accomplished during the heating phase of a batch cycle, as sludge neither enters nor leaves the reactor during this phase. Hence the cycle time duration must be sufficiently long so as to ensure that the heating phase is at least as long as the minimum required pasteurization time, t_D. Thus:

$$t_c = t_{tpr} + t_{fpr} + t_D \quad \text{(h)} \tag{6.1}$$

where,

t_c = the cycle time i.e. the duration of a full batch cycle {h}
t_{tpr} = the duration of phase 1 {h}
t_{fpr} = the duration of phase 2 {h}
t_D = the minimum sludge detention period required for pasteurization {h}

6.2.3.3 *The cycle time, t_c, and hydraulic retention time, R_H*

The duration of t_c determines the number of times that the reactor is fed in a day. The number of batch feeds in a day and the volume of each of these batches define the daily sludge feed volume and hence the reactor retention time. As the batch volume is the product of the batch fraction, f_b, and the full operating volume, V_p, the retention time can be calculated as follows:

$$R_H = \frac{V_p}{(24/t_c) \times (f_b \times V_p)}$$

$$= \frac{t_c}{24 \times f_b} \quad \text{(d)} \qquad (6.2)$$

where;

V_p = the full reactor operating volume {m³}

f_b = the batch fraction

R_H = the hydraulic retention time {d}

6.3 THE PROCEDURE FOR SIMULATING THE TEMPERATURE–TIME PROFILE OF THE REACTOR SLUDGE

The change in sludge temperature with time in the reactor is caused by changes in the enthalpy of the sludge. By constructing an unsteady state heat balance across the reactor, these changes can be calculated over a small, arbitrary time interval Δt, allowing the sludge temperature change over this time interval to be determined. An algorithm comprising 4 calculation steps was developed to simulate the saw tooth sludge temperature–time profile viz:

Step 1: The enthalpy, $H_{s(t)}$, of a defined operating volume of sludge, $V_{p(t)}$, is calculated at time t by assuming a reactor sludge temperature, $T_{se(t)}$.

Step 2: An unsteady state heat balance is constructed across the reactor over a small time interval Δt, accepting that the reactor volume and sludge temperature at time t, $V_{p(t)}$ and $T_{se(t)}$, remain constant. From the heat

$H_{s(t)}$ = the enthalpy of the reactor sludge at time t {MJ}
$V_{p(t)}$ = the sludge operating volume at time t {m³}
ρ_s = the sludge density taken equal to that of water (1.0 ton/m³)
Cp = the sludge specific heat taken equal to that of water {4.182 MJ/(ton.°C)}
T_{ref} = a specified reference temperature {°C}

6.3.2 The change in the reactor sludge enthalpy over the time interval Δt, ΔH_s

The reactor sludge enthalpy changes as the reactor loses heat to a number of sinks or gains heat from a number of sources. In the aerobic reactor and hence so also in the computer programme model, these sinks and sources are the following:

Heat sinks:
(1) heat lost to the ambient environment through the reactor walls
(2) heat lost due to the vaporization of water from the sludge by vent gases
(3) heat lost in the vent gas sensible heat
(4) heat lost in the sensible heat of sludge transferred out of the reactor

Heat sources:
(1) heat released by from biological reactions in the sludge
(2) heat gained by mixing the sludge
(3) heat gained through the sensible heat of the oxygen or air used to oxygenate the reactor
(4) heat gained through the sensible heat of the sludge fed to the reactor

The change in reactor sludge enthalpy, ΔH_s, over a time interval, Δt, is calculated as the difference between the heat sources and heat sinks across the reactor during this interval. Hence,

$$\Delta H_s = \sum_{t=i}^{t+\Delta t} \text{heat sources} - \sum_{t=i}^{t+\Delta t} \text{heat sinks} \quad \text{(MJ)} \tag{6.4}$$

Equation (6.4) is solved by constructing an unsteady state heat balance across the reactor over the time interval Δt. The unsteady state heat balance equation is as follows:

$$\Delta H_s = \Delta t \times (H_{bi} + H_{mi} + H_{gi} + H_{si} - H_{se} - H_{ge} - H_{ve} - H_{we}) \tag{6.5}$$

where,

Y_H = the specific heat yield {MJ/kg O}

O_C = the oxygen consumption rate {kg/h}

OTR = oxygen transfer rate {kgO/(m^3.h)}

In order to calculate the biological heat generation rate H_{bi} with Eq (6.6) the biological oxygen consumption rate O_c needs to be known.

Because the reactor is accepted to be oxygen limited, all the oxygen transferred to the liquid by the oxygenation system is assumed to be consumed biologically, and hence O_c is obtained by equating it to the product of the oxygen transfer rate (OTR) and the volume of the reactor (V_p): The OTR, in turn, is calculated from the input conditions i.e. from the oxygen supply rate and the oxygen transfer efficiency. Only when these two input parameters change, does the mass of oxygen transferred to the liquid i.e. OTR.V_p change. In conformity with this, when the reactor volume changes during the sludge transfer and feeding cycles, it is assumed that the OTR changes correspondingly to give the same constant mass oxygen transferred rate OTR.V_p. The reasons for assuming this are two fold:

(1) the oxygenation system is unlikely to be operating at its maximum OTR so that a small decrease in volume is unlikely to decrease the mass of oxygen transferred to the reactor

(2) the sludge is oxygen limited so that a small reduction in volume is not likely to reduce the biological oxygen consumption rate O_c (kgO/h).

It is important to note however that the rate at which biological heat can be generated is not unlimited. The maximum biological heating rate occurs when the reactor sludge is no longer oxygen limited and under conditions of oxygen sufficiency. In this event the sludge is consuming oxygen at the biological oxygen utilization rate, OUR. Hence, the maximum value of O_c i.e. O_{cmax} that may be substituted into Eq (6.6a) is

$$O_{cmax} = OUR \times V_p \quad \text{(kg/h)} \qquad (6.6b)$$

where,

OUR = the biological oxygen utilization rate of the sludge {kg/(m³.h)}

If the sludge is operating at its OUR , then it is no longer oxygen limited, and a change in volume is likely to change O_c with the result that during the transfer and feeding phases, the biological heat generation rate is likely to decrease and increase respectively. However, because these two phases are generally much shorter than the heating phase, and because the volume changes are small, the decrease in H_{bi} over this very short period will not affect significantly the temperature of the reactor sludge, so H_{bi} can be assumed constant over all three phases without loss of accuracy.

In Chapter 4, it was shown that Y_H decreased with increasing oxygen consumption in the Milnerton reactor, and it was postulated that Y_H was proportional to the degree of oxygen limitation in the sludge. On the basis of the Milnerton data, this relationship may be represented by the following equation:

$$Y_H = 14.245 - 1.648 \left(\frac{O_c}{O_{c\,max}} \right) \tag{6.7}$$

6.3.2.2 *The mechanical heat input rate, H_{mi}*

The mechanical heat input rate at Milnerton was calculated with Eq (3.8) which is specific for the pumped recirculation mixing and oxygenation system installed in the aerobic reactor. In general, the nature and means by which mechanical heat enters the reactor sludge will vary depending upon the reactor mixing/oxygenation design. This makes it difficult to develop an equation which is general in application for predicting H_{mi}. Accordingly, in the computer model a simple approach has been adopted which requires that a value for H_{mi} which is appropriate to the design which is to be simulated, is selected. It is accepted that the mechanical heat input rate is constant over all 3 phases because this is reasonable for most mixing/oxygenation applications and also simplifies calculations.

6.3.2.3 *The wall heat loss rate, H_{we}*

For the Milnerton aerobic reactor Eq (3.9) was derived to predict the rate of wall heat loss, H_{we}:

$$H_{we} = U \times A \times (T_{se} - T_{amb}) \quad \text{(MJ/h)} \tag{6.8}$$

where,

U = the overall heat transfer coefficient {MJ/(m².h.°C)}
A = the surface area of the reactor {m²}
T_{amb} = the ambient temperature {°C}

Theoretically H_{we} changes during the 3 phases due to the change in the reactor sludge temperature T_{se} with time. Practically however, not only is H_{we} itself very small relative to the other heat sinks, but also the changes in it, due to the 2–5°C changes in T_{se}, are very small. Consequently for simulation purposes H_{we} can be assumed constant with no loss of accuracy. This together with the fact that the wall heat loss depends on the reactor geometry, weather conditions and the degree of reactor logging led to the decision that in the computer model the wall heat loss rate H_{we} is constant and for a given situation is estimated as a rate in MJ/h or kW given as input to the model.

6.3.2.4 *The rate of vent gas water vapour heat loss, H_{ve}*

The rate of vent gas water vapour heat loss, H_{ve} is described by:

$$H_{ve} = h_{fg} \times M_{wp} \quad (\mathrm{MJ/h}) \tag{6.9}$$

where,

h_{fg} = latent heat of water vaporization (MJ/kg)
M_{wp} = predicted mass flow rate of water vapour in the vent gas (kg/h).

The latent heat of water vaporization was obtained from steam tables and in general can be closely approximated by

$$h_{fg} = 2.510 - 2.386.10^{-3}\, T_{ge} \quad \{\mathrm{MJ/kg}\} \tag{6.10}$$

where,

T_{ge} = vent gas temperature as it leaves the reactor {°C}

The mass flow rate of water vapour leaving the reactor in the vent gas, M_{wp}, is a function of the vent gas temperature, flow rate and the degree of vapour saturation of the vent gas. Assuming the vent gas is saturated with water vapour, equations describing M_{wp} have been derived in Appendices 3B and 5A (Eqs 3B.6

to 3B.9 and 5A.22 to 5A.26). These equations satisfactorily predicted water vapour condensate collected in the vent gas primary and secondary water traps indicating that at Milnerton the vent gas was saturated with water vapour at vent gas temperatures 2 to 5°C below the reactor sludge temperature. At Athlone (Pitt, 1990) an air oxygenated system, it was also found that the vent gas was saturated, with the vent gas temperature also about 5°C below the sludge temperature. Therefore for the computer model it is reasonable to accept that the vent gas is saturated with water vapour at the vent gas temperature (T_{ge}) and the temperature difference between the reactor sludge and vent gas (T_{diff}) is given as input. The equations to predict the mass flow rate of water vapour incorporated in the model are given below (see Appendix 5A for details):

$$M_w = 18 \times m_{dg} \times \frac{p_w}{p_{dg}} \quad \text{(kg/h)} \tag{6.11}$$

$$P_w = 10^{\left(-\frac{2238}{T_{ge} + 273} + 8.896\right)} \tag{6.12}$$

$$T_{ge} = T_{se} - T_{diff} \quad (^\circ\text{C}) \tag{6.13}$$

$$P_{dg} = P_T - P_w \tag{6.14}$$

where,

m_{dg} = the molar flow rate of dry vent gas {kmol/h}
p_w = the partial pressure of water in the vent gas {mmHg}
p_{dg} = the partial pressure of dry vent gas {mmHg}
T_{diff} = the temperature difference between the vent gas and the reactor sludge {°C}
P_T = the total absolute pressure of the vent gas {mmHg}
18 = the molar mass of water {kg/kmol}

In Appendix 5A, an equation has been derived which allows calculation of the molar flow rate of the dry vent gas, m_{dg}, as a function of the oxygen transfer rate, OTR, the oxygen transfer efficiency, OTE, and the respiration quotient (mole CO_2 produced/mole O_2 consumed), Y_{CO_2}. The equation accounts for sludge oxygenation using both air and pure oxygen and is as follows:

$$m_{dg} = \frac{1}{32} \text{ x } \frac{OTR \, . \, V_p}{OTE} \text{ x } (a - OTE + OTE \text{ x } Y_{CO_2}) \qquad (6.15)$$

where,

a $= 1 + (1-f_{mo_2})32/(28\ f_{mo_2})$ for pure oxygen
f_{mo_2} = mass fraction of oxygen in feed supply
= 1.000 for pure oxygen
= 0.2317 for air
so that
a = 1 for pure oxygen
and
a = 4.79 for air
Y_{CO_2} = the respiration quotient.

For oxygen enriched air, f_{mo_2} is increased appropriately depending on the mass flow rate of oxygen injected into the air supply.

6.3.2.5 *The rate of vent gas sensible heat loss/gain, $H_{ge} - H_{gi}$*

The feed oxygen or air entering the reactor and the vent gas leaving the reactor differ in composition and mass flow rate. Nevertheless, for the purposes of calculating the net rates of sensible heat loss and gain from the reactor, these differences may be ignored. The effect on the heat balance of the calculation error arising from this step is small and is justified by the increased ease with which the net sensible heat loss term can be calculated (see Section 5.4). An equation describing $H_{ge} - H_{gi}$ is listed below. The derivation of this equation is described in Appendix 5A.

$$H'_{ge} = (H_{ge} - H_{gi}) = C_{pg}\, m_{dg}\, (T_{ge} - T_{gi}) \quad \text{(MJ/h)} \qquad (6.16)$$

where,

m_{dg} = molar flow rate of dry vent gas (kmol/h) (see Eq 6.15)

C_{pg} = specific heat of the vent gas
= 0.0368 MJ/(kmol.°C) for pure oxygen
= 0.0302 MJ/(kmol.°C) for air
(see Appendix 5A for derivation of these values)
T_{ge}, T_{gi} = vent and influent gas temperatures {°C}

For continuous oxygenation Eq (6.16) applies over all three phases of the batch cycle.

6.3.2.6 *The sensible heat loss/gain due to sludge leaving/entering the reactor, H_{se}, H_{si}*

The volume of sludge leaving the reactor during the transfer phase and entering the reactor during the feeding phase of the batch cycle is equal to the batch volume, V_b. The loss and the gain in sensible heat due to this sludge flow may be calculated with the following equations:

$$H_{se} = \frac{V_b}{t_{tpr}} \rho_s \times C_p \times (T_{se} - T_{ref}) \quad \text{(MJ/h)} \tag{6.17}$$

$$H_{si} = \frac{V_b}{t_{tpr}} \rho_s \times C_p \times (T_{si} - T_{ref}) \quad \text{(MJ/h)} \tag{6.18}$$

The density and specific heat of the sludge ρ_s and C_p are accepted to be equal to that of water, i.e. 1.000 ton/m³ and 4.182 MJ/(ton.°C); the justification for this was discussed in Chapter 3, Section 3.4.2. Also in Eqs (6.17) and (6.18) V_b is divided by t_{tpr} and t_{fpr} respectively in order to express them in time units of per h. and so by multiplying them by Δt (see Eq 6.5) will give the amount of heat gained/lost in the time interval Δt over which the heat balance is constructed.

6.3.3 The reactor sludge enthalpy at time t+Δt, H_s (t+Δt)

The enthalpy in the reactor at time t, $H_s(t)$, is calculated with the aid of Eq (6.3). After an interval of time, Δt, has elapsed the enthalpy of the sludge changes to a new value:

$$H_{s(t+\Delta t)} = H_{s(t)} + \Delta H_s \quad \text{(MJ)} \tag{6.19}$$

where,

ΔH_s = the change in reactor enthalpy over the time interval Δt {MJ}

The change in reactor sludge enthalpy, ΔHs, is calculated with Eq (6.5) after solving the heat balance described above.

6.3.4 The reactor sludge temperature at time t+Δt, $T_{se}(t+\Delta t)$

By rearranging Eq (6.3), the reactor sludge temperature at time t+Δt can be calculated as follows:

$$T_{se(t+\Delta t)} = \frac{H_{s(t+\Delta t)}}{V_{p(t+\Delta t)} \times \rho_s \times Cp} + T_{ref} \quad (^{\circ}C) \qquad (6.20)$$

In order to accomplish this calculation, the sludge operating volume at time t+Δt must be determined. The operating volume, Vp, only changes during transfer and feeding. During the heating phase, V_p remains constant. In each of the transfer and feeding phases, the amount by which V_p changes over the entire duration of the phase is equal to the volume of a single batch of sludge, V_b – see Fig.6.1. The value of the batch volume, V_b, is constant for a given design and is equal to the product of the batch fraction, f_b, and the full operating volume, V_p. For example, at Milnerton a batch fraction of 1/12 was set by the level control system (see Section 3.3.2, Fig 3.3) resulting in a batch volume of 3.75m³. Hence the rate of change in V_p in the transfer, feeding and heating phases may be generalized with the following equations:

$$\text{Phase 1 : } F = -f_b \times \frac{V_p}{t_{t\,pr}} \quad (m^3/h)$$

$$\text{Phase 2 : } F = f_b \times \frac{V_p}{t_{f\,pr}} \quad (m^3/h) \qquad (6.20)$$

$$\text{Phase 3 : } F = 0 \quad (m^3/h)$$

where,

F = the rate of change of the operating volume during one of the 3 phases of a batch cycle {m³/h}

f_b = the batch fraction i.e. the ratio of the batch volume, V_b, and the operating volume, V_p

Thus the operating volume at time t+Δt is given by:

$$V_{p(t+\Delta t)} = V_{p(t)} + F \times (\Delta t) \quad (m^3) \tag{6.21}$$

where,

$V_{p(t)}$ = the operating volume of the reactor at time t {m³}

$V_{p(t+\Delta t)}$ = the operating volume of the reactor at time (t+Δt) {m³}

6.3.5 Performing the four calculation steps

In order to proceed with the calculations in steps 1 to 4 above, values for $T_{se(t)}$ and $V_{p(t)}$ in step 1, and the time interval Δt, need to be defined. Furthermore, to solve the heat balance equations in step 2, information regarding the design and operation of the reactor is required and accordingly, the values of 15 input variables need to be specified.

6.3.5.1 *The duration of the time interval, Δt*

The values of the time interval Δt need not be constant over a complete batch cycle. In the computer program discussed in the following section, 3 different values are used – 1 for each phase of the batch cycle. In the computer program, the value of the Δt is 1/10 of the duration of the phase. Therefore 30 iterations, 10 per phase, of steps 2 to 4 above, will generate 30 values of T_{se} with which the temperature–time profile of a single batch cycle can be constructed. The accuracy of the simulation is increased by iterations with small values of Δt.

6.3.5.2 *The initial values of T_{se} and V_p*

At the start of the simulation, the precise value of T_{se} at any given time is unknown and cannot be specified. Thus a value which will serve as a reasonable starting point for these iterations needs to be assumed. While any value in the thermophilic range is adequate, a value which closely approaches the temperatures which constitute the steady state temperature– time profile will reduce the number of iterations required to achieve steady state. The computer programme incorporates features which allow estimation of the T_{se} starting temperature (see Section 6.5 below – Model Features).

Unlike the sludge temperature, T_{se}, the operating volume, V_p, can be specified at a given time, and therefore an exact value can be specified for $V_{p(t)}$ in step 1. The value required as input to the programme is the full operating volume, i.e. the volume during the heating phase.

6.3.5.3 *The initial time, t*

Because the value entered into the programme for $V_{p(t)}$ is the full operating volume, i.e. the volume during the heating phase, the simulation must be started at some point in the batch cycle when the sludge in the reactor is in the heating phase. The end of the heating phase, or the start of the transfer phase is taken as the starting point of the simulation at a time of 0 hours in the computer program (see Appendix 6A for listing).

6.3.5.4 *The reference temperature, T_{ref}*

A convenient value for T_{ref} is 0°C as this simplifies Eqs (6.17) and (6.18). With T_{ref} at 0°C, the enthalpy of the sludge and gases is 0 MJ when T_{se} is equal to 0°C.

6.3.5.5 *The design and operating specifications*

To solve Eq (6.5), 15 different input variables have to be specified. These variables constitute the design specifications and operating conditions of the reactor. These 15 variables are:

Y_H = the specific heat yield i.e. the amount of biological heat produced by the sludge per kg of oxygen consumed {MJ/kg O}

Y_{CO_2} = the respiration quotient i.e. the moles of carbon dioxide produced per mole of oxygen consumed

T_{si} = the temperature of the feed sludge {°C}

T_{gi} = the temperature of the supply air or oxygen {°C}

T_{diff} = the temperature difference between the vent gas and the sludge in the reactor {°C}

P_T = the total pressure in the vent gas {mmHg}

O_c = the biological oxygen consumption rate {kg O/h}

O_{eff} = the fraction of feed supply oxygen which is utilized

V_p = the full operating volume of the reactor {m^3}

f_b = the batch fraction i.e. the ratio of the batch and full operating volumes

t_c = the duration of a complete batch cycle {h}

t_{tpr} = the duration of phase 1 {h}
t_{fpr} = the duration of phase 2 {h}
H_{mi} = the rate of mechanical heat input {MJ/h} or {kW}
H_{we} = the rate of wall heat loss {MJ/h} or {kW}

6.4 THE USE OF A COMPUTER PROGRAM TO SIMULATE THE CHANGE OF TEMPERATURE WITH TIME IN THE REACTOR

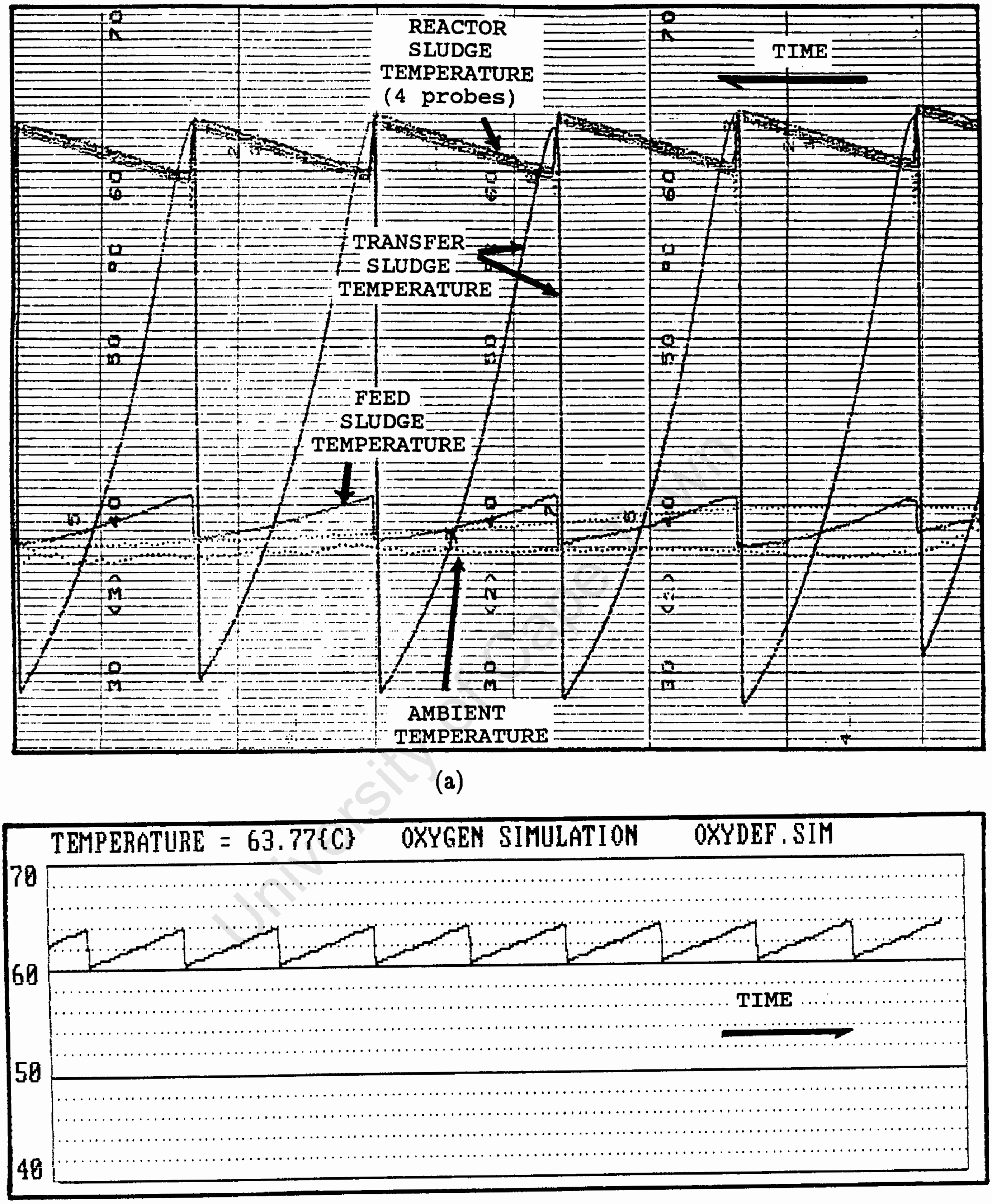
A computer program was written to perform the simulation of reactor sludge temperature with time, displaying the temperature - time profile in a graphical plot. The program has been written in Turbo Pascal Ver.4.0 and is listed in full in Appendix 6A.

While the program is based on the 4 steps listed above, it does not calculate the heat loss through the reactor walls, H_{we}, with the aid of Eq (6.8). Hence values of the overall heat transfer coefficient, U, the ambient temperature, T_{amb}, and the reactor surface area A, are not required as data input into the program. An average value for H_{we} must be calculated outside of the program and entered into the program as one of the operating data as in Section 6.3.3 above. Although this might be a less accurate means of expressing H_{we} than Eq (6.8), H_{we} makes a very small contribution to the heat balance (see Chapter 4, Section 4.2.3) if the reactor is lagged, and the error introduced into the simulation by this simplification is unlikely to be significant.

The accuracy with which the computer model is able to predict reactor sludge temperatures is demonstrated in Table 6.1 where the actual temperatures recorded at Milnerton are compared with those predicted under equivalent operating conditions by the model. The temperatures quoted in the table are those measured at the end of phase 1 of the batch cycle i.e. the highest temperature within a batch cycle [temperature $T_{se(1)}$ in Fig.6.1] . In Figs 6.3 and 6.4 the actual and simulated temperature - time profiles of steady state reactor operation are presented - note that the direction of time on the x-axis of the simulated profile is opposite to that on the actual profile.

6.5 FEATURES OF THE COMPUTER SIMULATION MODEL

The computer simulation model has been termed "ATASIM" (autothermal thermophilic aerobic simulation) and is coded in Turbo Pascal Ver.4.0 for IBM PC or compatibles. It comprises 11 separately compiled units and is initiated from DOS by the command "ATASIM". Some features of the program and



(a)

(b)

Fig.6.3: Actual (a) and simulated (b) saw tooth temperature profiles of sludge temperature versus time for the aerobic reactor of the Milnerton dual digestion system.

TABLE 6.1: Predicted vs actual reactor temperatures.

Retention R_H (days)	O_2 supply (kg/h)	%O_2 consumed	Feed temp T_{si} (°C)	Actual T_{se} (°C)	Predicted T_{se} (°C)
1.25	24.3	69.2	17.9	61.0	60.1
1.25	22.0	77.9	19.0	60.5	60.8
1.25	18.5	83.5	18.3	57.1	56.8
1.5	18.5	84.4	19.7	65.8	66.6
1.5	16.5	85.4	20.8	62.4	62.0
1.25	14.5	92.6	18.8	53.0	54.0
3.0	6.3	100.0	20.8	67.8	67.0

elementary operating instructions are given below.

The program consists of 3 main sections, namely a steady state design window, a data input window and an unsteady state simulation window. The design window allows the user to perform a rough steady state process design of the reactor with the reactor sludge temperature fixed at 60°C. The default data in the window represents a reactor at steady state and may be altered to suit the requirements of the program user according to the instructions listed in the key menu on the RHS of the screen. After the data has been altered, it no longer represents reactor steady state conditions at 60°C but, by "marking" one of the unaltered parameters this parameter will change to restore the steady state.

Upon leaving the design window, the program allows the design data to be transferred to the data window for either a pure oxygen or an air/enriched air simulation. If data transfer is not required then a default data file provides data to the window. The data window allows the design data to be more precisely specified for simulation purposes. A key menu on the RHS of the screen lists various functions which may be performed, such as data modification and file routines. Certain of the parameters specified in Section 6.3.5.5. above, may not be input from this window but are specified in the program source code. To change them requires that the source code be modified and, consequently, that the program be recompiled. The parameters and their locations in the source code re:

Y_H – the specific heat yield, Y_H, is represented by an equation [see Eq (6.7)] in Procedure 'effluent gas moles and mass' of unit CALCULAT.

Y_{CO_2} – the respiration quotient, Y_{CO_2}, is specified in Procedure 'load data' of unit SIMULDAT.

From the data window it is possible to perform a quick simulation. This quick simulation calculates the steady state temperature for the data set displayed in the data window and records this temperature for use as a starting point in the unsteady state simulation – see Section 6.3.5.2.

After completing the process of data input, the full unsteady state simulation window is entered and the temperature versus time profile for successive 24h periods is plotted. During simulation a key menu on the lower left corner of the screen allows a number of functions to be performed such as data modification and heat and mass balance displays.

The following features of the program are worthwhile noting:

Help: Help routines are available in the design and data windows and may be activated by the key sequence Ctrl–H. These routines supply information about the parameters displayed in the windows.

File operations: Files may be saved and retrieved from the data and the simulation windows. From the data window, the program automatically retrieves two default files OXYDEF.SIM for pure oxygen and AIRDEF.SIM for air/enriched air simulation. These default files can be altered from ATASIM or via a standard text editor. Note that during file operations, unless otherwise specified, the default file path and drive are those from which the ATASIM program is run.

Graphics compatibility: ATASIM employs graphics routines from Borland's Turbo Graphix Toolbox Ver.4.0. The program will only operate on a PC fitted with a Hercules graphics card.

Printer compatibility: In the simulation window, the key menu allows both steady state data and a hard copy of the temperature versus time profile to be printed. The printing routines will only operate on Epson or compatible printers.

Leaving ATASIM: From the design and data windows the program may be exited by the key sequence Ctrl–Break, while from the simulation window, simply pressing the key 'Q' is sufficient.

CHAPTER 7

CONCLUSIONS

7.1 INTRODUCTION

With the increasing awareness of the health risks associated with the use of sewage sludge for agricultural purposes as a soil conditioner, sludge treatment systems which not only effect sludge stabilization but also pasteurization are receiving more attention. Dual digestion is such a system. It treats sludge in two sequential stages, first in an autothermal thermophilic aerobic reactor of short retention time (1-3 days) for pasteurization, then in a mesophilic anaerobic stage for stabilization. The combination of these two stand-alone sludge treatment systems is claimed to offer a number of advantages over conventional mesophilic anaerobic digestion viz. in the aerobic reactor,

(1) the thermophilic temperatures pasteurize the sludge making it safer for disposal,

(2) the sludge is 'pretreated' through partial solubilization of particulate organic matter allowing short sludge ages (10 days) in the anaerobic stage,

(3) solubilization produces alkalinity through ammonification of proteins lending greater pH stability to the anaerobic stage,

(4) very little sludge stabilization takes place – only to the degree that the heat generated biologically maintains thermophilic temperatures; final and full stabilization takes place in the anaerobic stage – methane generation is reduced only marginally by the aerobic pretreatment stage,

(5) the major portion of the heat required to achieve thermophilic temperatures is biologically generated,

and in the anaerobic stage

(6) the heat required to maintain mesophilic temperatures is derived solely

from the hot aerobic sludge feed with the result that the methane generated can be used for purposes other than digestor heating.

From a practical point claim (2) is of particular importance especially in South Africa where mesophilic anaerobic digestion of sewage sludge is very commonly practiced; if substantiated, then dual digestion would be a viable system for upgrading existing fully loaded anaerobic digestors. This single feature would justify implementation of dual digestion with sludge pasteurization being an additional benefit.

7.2 RESEARCH INTO DUAL DIGESTION

Recognizing the potential benefits of dual digestion, the Water Research Commission and Milnerton Municipality, in 1987 commenced supporting a full scale evaluation of dual digestion (45 m^3 pure oxygen oxygenation aerobic reactor and 600 m^3 anaerobic digestor) at Milnerton's Potsdam Wastewater Treatment Works, with scientific and academic input by the Division of Water Technology of the CSIR and the Department of Civil Engineering of UCT. In this evaluation a number of important considerations needed to be addressed such as,

(1) oxygen requirements and oxygen utilization efficiency with pure oxygen and air oxygenation, to achieve thermophilic temperatures by means of biological heat generation.

(2) minimum retention time which can be achieved in the aerobic reactor and in the anaerobic digestor without compromising the attainment of thermophilic temperature in the reactor and VS removal, gas production and sludge stability in the digestor.

(3) control of temperature in the aerobic reactor and anaerobic digestor,

(4) efficacy of pathogen inactivation and mode of operation to prevent recontamination,

(5) dewaterability of the anaerobically digested sludge,

(6) economic evaluation of the system.

As part of the research programme, a detailed investigation into the aerobic reactor performance and behaviour was undertaken. Central to this investigation was the study of biological heat generation by the reactor sludge. The determination of the stoichiometry and kinetics of biological heat generation in the reactor would enable design criteria such as the reactor oxygen requirements and retention time to be quantified. This involved measuring unsteady state heat and oxygen mass balances over the aerobic reactor from which the biological heat generation rates (H_{bi}) and biological oxygen consumption rates (O_C) could be determined. Also various chemical parameters such as total solids (TS), volatile solids (VS) and COD of the influent and effluent sludge were measured to check (1) the removal rates of these materials in the aerobic reactor, and (2) whether the biological heat generation rate and oxygen consumption rate could be defined in terms of the VS or COD removal rates. The conclusions from the investigation into the aerobic reactor are given below.

7.3 CONCLUSIONS FROM THE AEROBIC REACTOR INVESTIGATION

7.3.1 Biological heat generation and oxygen consumption

From 116 oxygen mass and heat balance tests conducted over 8 months, it was established that (1) biological heat generation is directly proportional to oxygen consumption; the constant of proportionality being the specific heat yield (Y_H, MJ/kgO) and (2) Y_H varied from 11.4 to 14.6 MJ/kgO and gave an average value of 12.77 ± 0.58 MJ/kgO. The Y_H value was found to be independent of reactor sludge temperature and retention time, which ranged from 54 – 67°C and 1.2 to 3 days respectively. However, Y_H showed a slight linear dependence on oxygen limitation with Y_H increasing as oxygen limitation increased, where oxygen limitation is defined as (1 - OTR/OUR) where OTR [kgO/(m^3.h)] is the oxygen transfer rate to the sludge and OUR [kgO/(m^3.h)] the biological oxygen utilization rate. This dependence of Y_H on oxygen limitation is consistent with bioenergetics from which it is reasonable to accept that energy utilization for microbial growth decreases under the increasing oxygen limited conditions, resulting in a greater energy loss through the release of heat.

The average measured Y_H value of 12.77 MJ/kgO agrees favourably with values predicted from bioenergetics and from thermodynamic values derived from the work of McCarty (1972). Also experimental measurements relating the heat of combustion to the COD of various organic substrates yield Y_H values closely similar to the average value measured in this investigation. Reliable direct comparative measurements of Y_H are not available from research into autothermal thermophilic aerobic digestion. However, Y_H measurements have been made in the field of microbiology by Cooney *et al.* (1968) and by Rothbaum and Stone (1961); the average values of the measurements by Cooney *et al.* agree favourably with that measured in this investigation but the values reported by Rothbaum and Stone are higher, ranging from 16.3 to 26.7 MJ/kgO. Confidence in the accuracy of the measured Y_H value of 12,77 MJ/kgO comes not so much from the favourable comparison with other measured values as from (1) the favourable comparison with McCarty's calculated values and Kambu's (1971) and Wright's (1975) heat of combustion values and (2) the thoroughness and uncompromising vigilance and attention to every detail in the experimental procedure employed to measure Y_H in the Milnerton reactor.

7.3.2 Maximum biological heat generation rate

By virtue of the proportionality between the biological heat generation rate and the sludge oxygen consumption rate, the maximum biological heat generation rate $H_{bi(max)}$ is obtained when the sludge is consuming oxygen at its maximum rate ie the biological oxygen utilization rate [OUR, kgO/(m³.h)]. The $H_{bi(max)}$ will be achieved when the sludge is not oxygen limited. Under these conditions increasing the oxygen supply rate (OSR) to the reactor will not increase the oxygen consumption rate of the sludge and therefore not increase the biological heat generation rate.

At Milnerton, where the sludge treated was a mixture of primary and humus tank sludge, a steady state OUR of about 0.38 kgO/(m³.h) was measured at a sludge concentration of 30 kgVS/m³. At this stage, the temptation of converting the OUR to a VS concentration specific rate should be resisted because the effect of VS concentration on OUR is uncertain; it may be thought that doubling the VS concentration would double the OUR, but this may not necessarily be so. Biological oxygen consumption is related to the kinetics of substrate utilization by aerobic thermophilic organisms and therefore its rate depends more on the concentration of these organisms in the reactor than on the concentration of the substrate (sludge) feed[1]. Until a reliable method of OUR estimation can be developed for the aerobic reactor, it is recommended that it is measured in laboratory or pilot scale trials. From a design point of view, a very accurate estimate of OUR is not required because, as discussed below, the reactor needs to be operated under oxygen limiting conditions, with the result that the actual biological oxygen consumption rate necessarily needs to be lower than the maximum. Nevertheless, the OUR of a sludge is an important parameter because it gives an indication of the approximate maximum oxygen transfer rate (OTR_{max}) required by the oxygenation system. For autothermal thermophilic aerobic *digestion*, OUR estimation is based on Volatile Solids (VS) removal rates, but this approach is not appropriate for the aerobic reactor in dual digestion principally because VS removal (1) is not an objective in the aerobic reactor and (2) was negligible in the Milnerton aerobic reactor (1.5%) although over 0.35 kgO/(m^3.h) oxygen was consumed and 4.59 MJ/(m^3.h) heat was generated at 1.25d retention time.

7.3.3 Oxygen limitation

Oxygen limitation is desirable for two reasons (i) pretreatment for enhancing subsequent anaerobic digestion and (ii) temperature control in the aerobic reactor. In order to achieve oxygen limitation the sludge should consume oxygen at a rate lower than its OUR. Therefore the oxygen transfer rate (OTR) must be less than the OUR.

The degree of oxygen limitation, defined as 1–OTR/OUR, for proper pretreatment could not be established from the literature, nor could it be quantified at Milnerton. At Milnerton, at 1.25 and 3 days retention time, the oxygen limitation in terms of the above definition was about 0.10 and 0.57 respectively.

7.3.4 Reactor temperature control

Under oxygen limiting conditions, it was found that the Milnerton reactor sludge

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[1] While the above comments may question the prudence of linking OUR and BVS removal in a kinetic equation, they do not refute the possibility that higher VS concentrations in the reactor will sustain higher OUR's. Furthermore the economic benefits of treating thick as opposed to thin sludges are self evident and thus it remains a sound operating practice to ensure that the highest possible VS concentration is fed to the reactor.

temperature could be completely and instantaneously controlled by the oxygen supply rate (OSR). Provided the increased OTR was less than the OUR, step increases even as high as 330% in OSR [from 0,13 to 0,43 kgO/(m^3.h)], which increased the OTR, and hence the biological oxygen consumption rate, by 300% [from 0.13 to 0.39 kgO/(m^3.h)] caused an immediate (<2h) increase in the biological heating rate (H_{bi}) which in turn caused an increasing reactor sludge temperature. In the same way, decreasing the OSR caused an immediate decrease in H_{bi} and a decreasing reactor sludge temperature. It should be noted that an increase (decrease) in OSR did not cause an equivalent increase (decrease) in OTR because with oxygenation systems, as was the case with the Milnerton pure oxygen system, an increase (decrease) in OSR usually causes a decrease (increase) in the oxygen transfer efficiency (OTE) where OTE is the ratio of the oxygen transfer and supply rates i.e. OTE = OTR/OSR.

7.3.5 Maximum oxygen transfer rate of oxygenation system (OTR_{max})

At Milnerton, a temporary maximum OUR of 0.44 kgO/(m^3.h), and therefore also an OTR_{max} of at least as high, was observed. This OTR was achieved with a 20 kW Vitox pumped recirculation pure oxygen oxygenation system; the OTE at this OTR_{max} was 0.83. The 20 kW power consumption was not wasted because more than 90% of it contributed directly to heating the reactor sludge. The maximum oxygen transfer rate of the oxygenation system is very important in the design and operation in the aerobic reactor - if the oxygenation system cannot transfer sufficient oxygen, pasteurization temperatures will not be achieved at the design retention time because OTR_{max} directly controls the maximum biological heating rate (provided that OTR_{max} < OUR).

Associated with the OTR of the oxygenation system, is its oxygen transfer efficiency, (OTE). Generally the OTE of a system decreases as OTR increases up to OTR_{max}. This means that for a certain increase in OTR, a greater increase in the oxygen supply rate, (OSR) is required. The OSR fixes the vent gas molar flow rate which in turn fixes the vent gas heat loss rate (H_{ve}) via sensible and water vapour heat losses. Once the system reaches its OTR_{max}, further increases in OSR will not increase OTR but serve only to cause greater vent gas heat losses without increasing the rate of biological heat generation (H_{bi}). Preferably the design OTR_{max} should be somewhat greater than the OUR of the sludge. In that way, the biological kinetics (i.e. OUR) will govern the minimum retention time stoichiometric formulæ for VS e.g. $C_5H_7O_2N$, it appears that the respiration quotient – mole CO_2 generated per mole O_2 utilized, denoted Y_{CO_2} – should be unity. However from the 116 vent gas measurements required for the heat and oxygen mass balances, Y_{CO_2} ranged from 0.53 to 0.85 with a mean of 0.66 ± 0.04. From this it would appear that the heat generating oxidation reactions are not simply those of VS degradation, as already indicated by the low VS removal.

Measurement of Y_{CO_2} was not a specific objective in the Milnerton research on the aerobic reactor; it emerged as a result of analyzing the vent gas volumetric flow rate and oxygen composition. This vent gas analysis was not required for the accuracy of the heat balance but for the accuracy of the oxygen mass balance from which the oxygen consumption rate (O_C), or equivalently the oxygen transfer rate (OTR), was measured in order to determine the specific heat yield Y_H. Had a Y_{CO_2} of unity been assumed in order to avoid having to measure the vent gas volumetric flow rate, then incorrect values of oxygen consumption rates and specific heat yields would have been measured.

As a parameter in the design of the aerobic reactor, Y_{CO_2} is of little significance. Its value influences the rate of heat loss in the vent gas from the aerobic reactor by fixing the vent gas volumetric flow rate for a given oxygen supply rate. However in pure oxygen aerated reactors like Milnerton, the rate of the heat losses in the vent gas is negligible and in air oxygenated reactors, while the vent gas heat loss rate is not negligible, the volume of inert nitrogen gas passing through the reactor is so large in comparison to the O_2 utilized and CO_2 generated that a value of Y_{CO_2} of 0.66 or 1.00 makes a negligible difference to the vent gas volumetric flow rate.

7.3.8 Degree of vent gas water vapour saturation

In order to accurately measure the vent gas volumetric flow rate it was necessary to partially dry the vent gas by condensing out most of the water vapour. By measuring the water vapour condensate accumulation rate in the water traps and calculating the remaining water vapour in the cooled vent gas, it was possible to determine the vent gas water vapour flow rate. Assuming vent gas saturation, the vent gas water vapour flow rate was calculated from the vent gas temperature and compared with that measured. The calculated and measured vent gas water vapour flow rates correlated very well and it was concluded that the vent gas was saturated with water vapour at the vent gas temperature measured at the point of exit from the reactor. This temperature was a constant 2 to 5°C below that of the sludge depending on the vent gas flow rate.

7.4 DESIGN AND SIMULATION OF THE AEROBIC REACTOR

The close correlation, and rapidity of response, between the biological heat generation rate and the oxygen transfer rate make the OTR a pivotal parameter in the design and simulation of the autothermal thermophilic aerobic reactor in dual digestion.

Aerobic reactor design and simulation procedures were derived from the results of the Milnerton aerobic reactor performance. These are founded on the basic heat and mass balance principles and accordingly are suitable for general application to aerobic reactor design. Therefore, whereas pure oxygen oxygenation was used on the Milnerton reactor, the derived design and simulation procedures are general and apply to reactors oxygenated also with air or oxygen enriched air.

The design procedure is based on the solution of the steady state heat balance across the reactor. Such a heat balance yields a constant temperature for the reactor sludge and is applicable only to reactors that are continuously fed. Even though, in practice, the reactor is batch fed (to avoid recontamination of pasteurized sludge) causing the reactor temperature to vary between 2 and 4°C per batch cycle, the steady state approach is adopted for design because it greatly simplifies the design procedure.

Two objectives need to be met by the aerobic reactor in dual digestion, viz. (1) pasteurization; by exposure of the sludge to a temperature above 60°C for 2h or above 70°C for 30 min and (2) pretreatment through oxygen limitation; for enhanced performance of the anaerobic digestor. The literature does not define the temperature and the degree of oxygen limitation at which sludge pretreatment is best accomplished, nor could this be established at Milnerton. Thus the approach to design is to ensure that (1) the specified pasteurization temperature and times are maintained in the reactor, and (2) the sludge is oxygen limited, i.e. the OTR [kgO/(m^3.h)] to the sludge by the oxygenation system is controlled at a lower value than the sludge's maximum OUR [kgO/(m^3.h)]. Not only is oxygen limitation important for sludge pretreatment but, as mentioned above, also for reactor sludge temperature control.

Accepting that the design of the aerobic reactor centres on achieving the shortest possible retention time through maximizing the heat sources (i.e. biological heat generation and mechanical heat input) and minimizing the heat losses (i.e. wall heat loss and vent gas water vapour and sensible heat losses) it is demonstrated with the aid of the steady state heat balance that 3 parameters are of crucial importance, viz:

(1) The sludge oxygen utilization rate (OUR), which fixes the maximum biological heat generation rate through the specific heat yield (Y_H).

(2) The oxygen transfer rate (OTR) of the oxygenation system. The actual biological heat generation rate is directly proportional to the OTR through the specific heat yield (Y_H). The OTR should be less than the OUR to ensure (1) sludge pretreatment through oxygen limitation and (2) reactor temperature control through control of the oxygen supply rate (OSR).

(3) The oxygen transfer efficiency (OTE) of the oxygenation system. For a particular pure oxygen or air oxygenation system at a given OTR, the OTE controls the vent gas volumetric flow rate; the lower the OTE the greater the vent gas flow rate and hence the greater the vent gas heat losses (via water vapour and sensible heats).

At a certain OTR and mechanical heat input, the heat sources are fixed. If OTE is high (>0.80) and oxygenation is with pure oxygen, the vent gas heat losses are small with the result that most of the heat generated can be lost via hot effluent sludge thereby allowing short retention times (~ 1.25 to 2d); if OTE is low (0.10–0.20) and oxygenation is with air then vent gas heat losses are high with the result that much less heat can be lost via the hot effluent sludge thereby forcing long retention times (4–6d). At Milnerton, where the sludge OUR was observed to be around 0.38 kgO/(m^3.h) and high OTR's and OTE's of 0.37 kgO/(m^3.h) and > 0.80 respectively were achieved with the use of pure oxygen, the vent gas heat losses were very low allowing 1.25d retention time operation and reactor temperatures above 60°C. In contrast, with air oxygenation, OTR's and OTE's tend to be considerably lower [e.g. at Athlone, 0.14 kgO/(m^3.h) and 0.12 respectively, Pitt, 1990] and the major part of the heat generated is lost via the vent gas with the result that only a minor part of the heat can be lost via hot effluent sludge thereby forcing long retention time operation to maintain temperatures above 60° C.

Biological heat generation has the drawback in that increasing its rate requires increases in OTR. Increases in OTR require increases in oxygen or air supply rates (OSR) which, together with the reduction in OTE usually caused by increases in OTR, causes larger vent gas heat losses. While these heat losses may be insignificant in pure oxygen oxygenated reactors, they can be substantial in air oxygenated reactors. In contrast to biological heat generation, supplementary heat sources such as (1) heat exchange between reactor effluent and feed sludge (2) increased mechanical heat input, or (3) anaerobic digestor gas combustion, have the equivalent effect of increasing the heat sources without increasing the heat losses, and therefore allow a *pro rata* reduction in retention time.

The design procedure described above is based on the steady state heat balance which accepts that the reactor sludge temperature is constant and therefore applies only to a continuously fed reactor. Pasteurization requires the reactor to be batch fed, resulting in a continuously changing sludge temperature over about 3° C in a saw-tooth pattern, decreasing at times of feeding and increasing between the feeding times, with the result that the reactor is not at steady state. The error arising from assuming steady state conditions is small and well justified by the simplifications in the heat balance it affords. The saw-tooth temperature profile can only be determined by solving the unsteady state heat balance by a forward integration method. This is dealt with in this thesis and an algorithm and computer programme are presented which solve the unsteady heat balance in order to simulate the temperature profile of the batch fed aerobic reactor. For specified input conditions identical to those observed in the Milnerton reactor, the simulation programme is shown to simulate the reactor temperature profile to within 0.5° C. Whereas the steady state design allows only the required pasteurization temperature to be specified, the simulation programme allows both the sludge pasteurization temperature and the undisturbed sludge detention period to be defined, the latter in terms of the interval between consecutive batch feeds.

7.5 EVALUATION OF AEROBIC REACTOR INVESTIGATION

Evaluating the results achieved in the investigation of the aerobic reactor described above, it can be seen that the objectives set for the aerobic reactor investigation, listed in Section 7.2 above, were met in that

(1) oxygen requirements and oxygen utilization efficiency with pure oxygen and air oxygenation, to achieve thermophilic temperatures by means of biological heat generation,
(2) minimum retention time, and
(3) a method for temperature control

in the aerobic reactor were determined. In addition, a design procedure was developed based on a steady state heat balance allowing *inter alia* the retention time in the reactor for specified influent and effluent sludge temperatures to be determined in terms of the oxygen transfer rate and the oxygen transfer efficiency of the oxygenation system. Also for the purposes of simulating the fluctuating reactor sludge temperature profile under the batch feed draw and fill conditions (to avoid recontamination of the treated sludge) an algorithm and simulation computer programme was developed which, moment by moment, solves the unsteady heat balance under specified operating conditions.

With these design and simulation procedures the attention of the design engineer needs to be focused on correctly designing and estimating the oxygen transfer rate (OTR) and efficiency (OTE) of the pure oxygen or air oxygenation system. In achieving this, knowledge of the sludge OUR is important in order to establish the biological upper limit to the rate at which oxygen can be transferred to the sludge. No numerical method for determining the sludge OUR was developed in this research, but, noting the likely complexity of such a method, it is suggested instead that laboratory or pilot scale tests be employed to determine its value for a particular sludge. As discussed below in Section 7.7, it is possible to avoid requiring knowledge of the sludge OUR if heat sources other than biological heat are exploited viz supplementary heat sources.

The Milnerton aerobic reactor, oxygenated with pure oxygen could be operated at 1 day retention time during hot summer months when feed sludge (26°C) and ambient temperatures (30°C) were high. In winter, at cold feed sludge (17°C) and ambient (10°C) temperatures, a retention time of 1.25 days was required to maintain reactor temperatures above 60°C. At 1.25 day retention time, the biological heat generation rate was 75% of the total heat required.

In operation, the aerobic reactor was stable, reliable and easy to operate. For 8 hours per day (24h00–08h00), it was left unattended by operating staff, who, when on duty during the 16h per day, had little more to do than take sludge samples and check that feed sludge was available in the feed sludge thickener. Technical failure seldom occurred; the major maintenance problem arose from erosion/abrasion of the Vitox oxygenation system pump which continuously recirculated sludge through the oxygen injection system and provided mixing of the reactor sludge. The casing of this pump needed to be rebuilt at about 5 month intervals.

7.6 EVALUATION OF DUAL DIGESTION INVESTIGATION

Although an evaluation of the dual digestion investigation results is not part of this thesis, a brief summary of the results of the Milnerton dual digestion system as a whole is given in the interests of completeness. The results are discussed in the context of the claimed advantages for the system given in Section 7.1 above.

(1) Pasteurization. This was satisfactory with zero viable *Ascaris* ova in the final (anaerobic effluent) sludge. Faecal coliforms were reduced by about 9 orders of magnitude. Also *Salmonella* was not detected in the final sludge. Interestingly, *Salmonella* also was not detected in the feed (to aerobic reactor) sludge and apparently this is due to rapid *Salmonella* die off at low pH conditions in the feed sludge thickener.

(2) Sludge pretreatment. The lowest stable anaerobic retention time was 15 days. At 12 days the gas production declined and volatile fatty acid concentration increased sharply but reverted to normal upon increasing the retention time again to 15 days. These results require further confirmation because the absence of inter stage heat exchange (between reactor effluent and feed sludge) caused the anaerobic digestor to overheat (44°C) during summer months if the aerobic reactor retention time was not increased (to reduce the digestor heat inflow) or if sludge transfer from the reactor was not diverted. This affected the digestor retention time, preventing a

comprehensive evaluation of the digestor performance at short retention times.

(3) Digestor stability. The aerobic reactor did generate a considerable increase in $H_2CO_3^*$ alkalinity, which when the digestor retention time, temperature and volume were adequately controlled, provided stability to the digestor. The pH of the aerobically treated sludge was around 6.7 whereas that of the aerobic reactor feed sludge was about 5.2.

(4) Sludge stabilization. The claim that very little sludge stabilization takes place in the aerobic reactor and most of it in the anaerobic digestor was verified – during the investigation the average VS and COD removals in the aerobic reactor were 1.5% and 12% respectively, whereas the VS and COD removals in the anaerobic digestor were both about 54% at 20 days retention time. The gas production was not influenced by the aerobic reactor with about $0.5 m^3$ (STP) methane per kgVS removed or $0.36\ m^3$ (STP) methane per kgCOD removed being produced. The aerobic reactor effluent sludge was very difficult to dewater; even with centrifugation, a large mass of fine material remained suspended which necessitated drying the complete sludge sample, without prior centrifugation, in order to get accurate results in the solids analysis (VS, TS). The final (anaerobic effluent) sludge was stable and odourless. It settled well but, due to the presence of some fine material, it tended to blind the sludge drying beds. Its dewatering characteristics in terms of Capillary Suction Time (CST) and Specific Resistance to Filtration (SRF) are approximately 350 sec. and $600 \cdot 10^{12}$ m/kg which reflect a poorly dewaterable sludge, but this is no worse than a conventional mesophilic anaerobically digested primary and activated sludge mixture.

(5) Digestor heating requirements. In the winter months, all the heat of the hot effluent sludge from the aerobic reactor at 1,25d retention time was required to maintain 37°C in the digestor. In the summer, this was too much with the result that the digestor overheated unless the aerobic reactor retention time was increased or the hot aerobic reactor sludge was diverted in order to reduce the heat flow to the digestor. This substantiates the claim that the heat required to maintain mesophilic temperatures in the digestor

is derived solely from the hot aerobic reactor sludge with the result that the methane generated can be used for purposes other than digestor heating.

7.7 EXPLOITATION OF SUPPLEMENTARY HEAT SOURCES

When supplementary heat sources are exploited then, irrespective of pure oxygen or air oxygenation, the following advantages are obtained:

(1) reduction in dependence on biological heat generation,
(2) oxygen limitation can be achieved more easily,
(3) reduction in operating costs – for air in reduced power and for pure oxygen in less oxygen,
(4) reduced vent gas heat loss,
(5) lower sensitivity of reactor to transfer rate limitations of oxygenation system,
(6) less need to know the maximum biological oxygen utilization rate (OUR) accurately.

or none of the above but instead

(7) higher sludge load, i.e. reduced retention time.

Two readily exploitable supplementary heat sources available for dual digestion systems are (1) heat exchange between reactor effluent and influent sludge and (2) combustion of anaerobic digestor gas. Influence of these supplementary heat sources on the steady state heat balance is most easily appreciated by considering that its effect is to increase the reactor sludge feed temperature (T_{si}). If the retention time remains unchanged, then the increased T_{si} reduces the hot effluent sludge heat loss rate (H_{se}). This allows the biological heat generation rate (H_{bi}) and hence oxygen transfer rate (OTR) to be reduced by reducing oxygen supply rate (OSR) and advantages (1) and (6) above result. On the other hand, if OTR (and oxygen transfer efficiency, OTE) are kept constant, H_{bi} and vent gas heat loss rate through water vapour (H_{ve}) and sensible heat (H_{ge}) do not change with the result that H_{se} also is unchanged. With H_{se} unchanged, the increased T_{si} allows the retention time (R_H) to be reduced (advantage 7).

With regard to the heat exchange, the amount of heat that can be extracted from effluent sludge depends on the anaerobic digestor wall heat losses. Because the only heat source for the digestor is the hot effluent sludge from the reactor, this hot sludge must not be cooled excessively lest it become difficult to maintain mesophilic temperatures (37° C) in the digestor.

As mentioned above, at Milnerton, in the winter, all the heat of the hot effluent sludge at 1.25 days retention time was required to maintain 37° C in the digestor. However, in the summer, this was too much heat with the result that the digestor overheated unless the retention time in the aerobic reactor was increased or the hot reactor sludge was diverted to reduce the heat flow to the digestor. Therefore at Milnerton, a heat exchanger would only have been used during the hot summer months.

At constant aerobic reactor retention time, heat exchange has the same effect as increased mechanical heat input H_{mi}; it adds heat to the reactor (through hotter feed sludge) without contributing to the heat losses, which leads to advantages (1) to (6) cited above. However, because of the need to maintain 37° C in the digestor solely by means of the hot effluent sludge, this supplementary heat source may be very limited depending on the heat losses from the digestor (as was the case at Milnerton during winter). Additional heat could be derived from an alternative heat source such as the combustion of all, or part of, the methane produced in the anaerobic stage of the process. This source alone might be sufficient to produce all of the heat required, giving rise to a situation where biological heat generation could be completely dispensed with. In such a situation, the 2 stage process would no longer be a dual digestion system but would become the combination of a sludge pre-pasteurizer and an anaerobic digestor. However dispensing altogether with biological heat generation, and therefore sludge oxygenation, might have an impact on the claimed sludge pretreatment effect in the reactor. While sludge pasteurization could still be accomplished, the reported potential of the reactor to significantly enhance the performance of the subsequent anaerobic digestor might be lost. It is on this potential that much of the justification for employing the biologically heated aerobic reactor in the dual digestion process rests.

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APPENDIX 3A

THE CALIBRATION OF THE OXYGEN SUPPLY ROTAMETERS AND THE VENT GAS METER

The oxygen feed rotameters and the Parkinson Cowan vent gas meter were key items of equipment in the oxygen mass balance for the specific heat yield test. In order to ensure accurate results for this test, it was necessary to accurately calibrate these 2 items of equipment. This appendix describes the methods and results of the calibration of the oxygen feed rotameters and the vent gas meter.

OXYGEN ROTAMETERS

Two rotameters, each fitted with its own needle valve, controlled and measured the flow of pure oxygen to the Vitox oxygenation system, one for high flows of oxygen (10 to 55 kgO/h) and one for low flows of oxygen (6 to 20 kgO/h). The former was an ASV 50–500ℓ model and the latter an ASV 15-150ℓ model. A calibration certificate was provided for each rotameter but it was decided to check these by testing. The results obtained in these tests justified this decision as the tests showed the supplier's calibrations to differ from measured values by up to 80%.

Method of calibration

A cylinder of pure oxygen was placed on a beam balance scale and connected to the Vitox oxygen supply panel. The oxygen pressure was regulated both at the cylinder and just prior to the rotameter needle valves in the oxygen control panel. The oxygen temperature was assumed to be at ambient (18°C) after having passed through approximately 20 metres of copper piping. At a fixed system pressure of 400 kPa, a rotameter scale setting was selected and, with a stopwatch, the time was measured for a known mass of gas to pass through the rotameter. This process was repeated for a number of rotameter scale settings.

The results for the 2 rotameters are summarized in Tables (3A.1) and (3A.2) respectively, from which the calibration curves shown in Figs (3A.1) and (3A.2) were obtained.

TABLE 3A.1: ASV 50-500 ROTAMETER CALIBRATION.

% SCALE	CYLINDER MASS (kg)		TEST MASS (kg)	TIME min'sec"	FLOWRATE (kg/h)
	BEFORE	AFTER			
70	78.9	73.9	5	7'32.68"	39.76
60	73.9	70.9	3	5'14.31"	34.36
60	77.4	73.4	4	7'12.42"	33.30
50	77.6	73.4	4.2	8'48.00"	28.64
50	74.0	70.6	3.4	7'01.33"	29.05
40	73.4	70.4	3	7'25.00"	24.27
30	70.4	67.4	3	9'18.00"	19.35
20	73.0	70.0	3	12'25.37"	14.5

TABLE 3A.2: ASV 15-150 ROTAMETER CALIBRATION.

% SCALE	CYLINDER MASS (kg)		TEST MASS (kg)	TIME min'sec"	FLOWRATE (kg/h)
	BEFORE	AFTER			
100	80.4	78.4	2	6'00.00"	20.00
70	78.4	77.0	1.4	5'15.80"	15.95
50	78.85	78.1	0.75	3'29.00"	12.92
30	77.0	76.0	1	5'52.16"	10.22
20	76.0	75.0	1	7'24.72"	8.09
10	75.0	74.0	1	9'34.19"	6.27

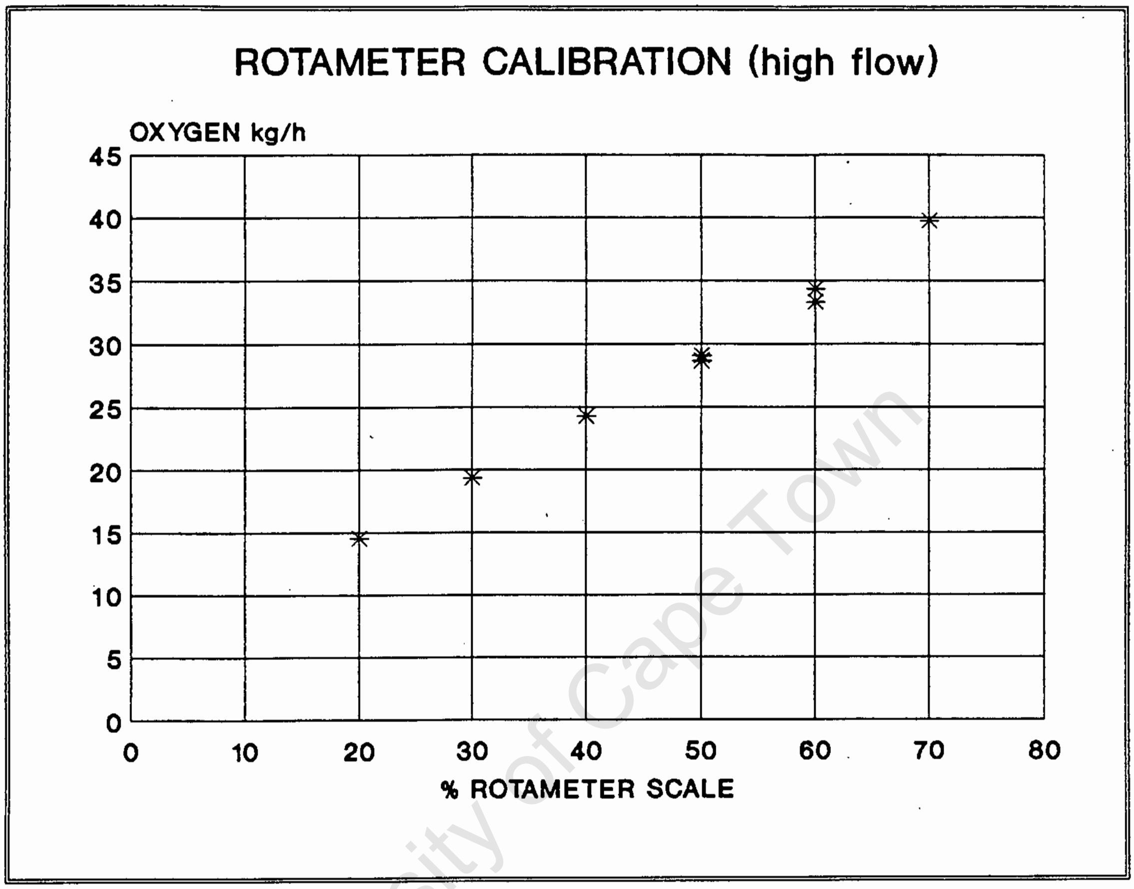

Fig 3A.1: High flow (ASV 50–500) oxygen supply rotameter calibration curve.

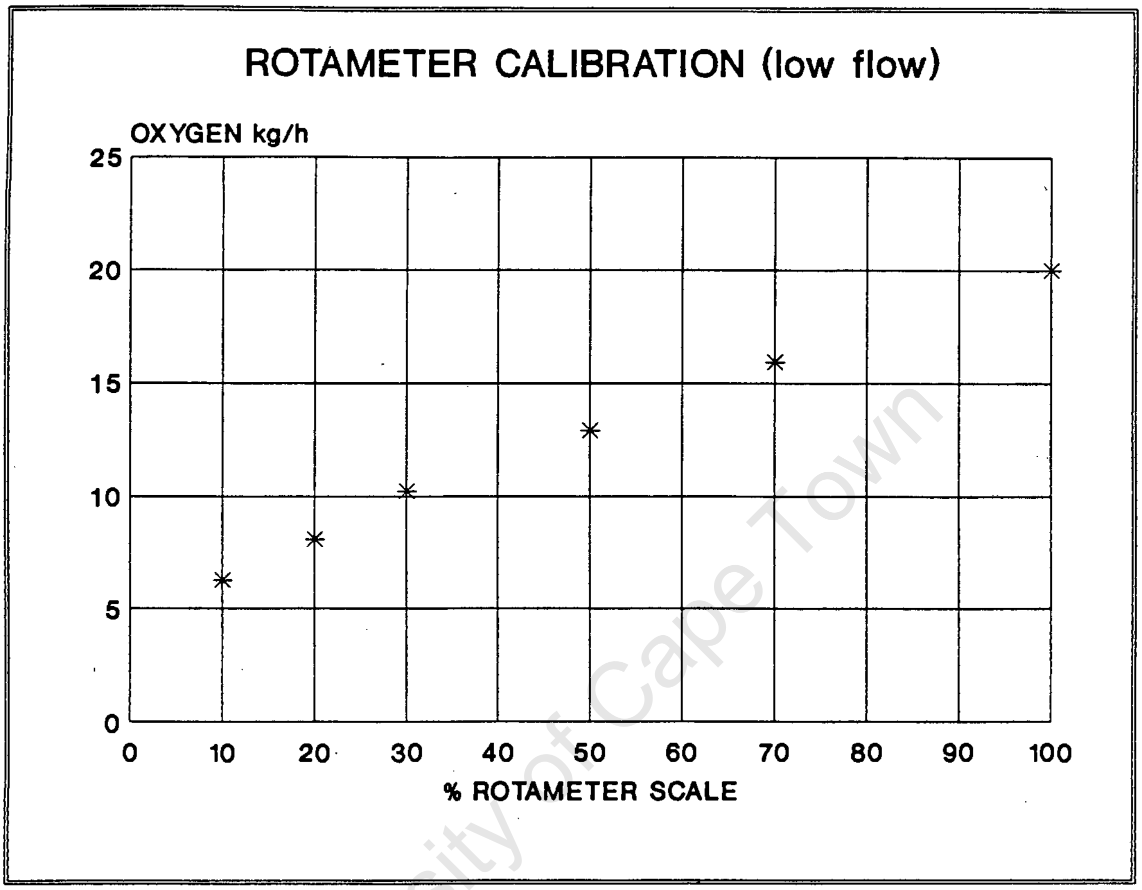

__Fig 3A.2__: Low flow (ASV 15–150) oxygen supply rotameter calibration curve.

PARKINSON COWAN METER CALIBRATION

The Parkinson Cowan towns gas meter displays a totalizing readout in m³. The meter was calibrated in a similar manner to that of the rotameters discussed above, i.e. by passing known masses of a gas through the meter and recording the meter readout. The gas selected was nitrogen and the mass of nitrogen delivered was converted to a volume of nitrogen at ambient temperature and pressure. This volume was compared with the meter readout. In order to simulate the pressure of the vent gas that needed to be monitored, the pressure of the metered gas was maintained at a similar level, ie atmospheric. The test was performed twice in order to verify the accuracy of the gas meter at both low and high gas flow rates. The results obtained in the calibration are summarized in Tables (3A.3) and (3A.4).

Vent gas meter calibration results

The test volume is obtained by converting the mass of gas measured during each time interval (1 kg) to a volume at the test ambient temperature (22° C). The conversion is:

$$\text{Test vol} = 1\ (\text{kg})/28\ (\text{kg/kmol}) * 22.4\ (\text{m}^3/\text{kmol}) * 295\ (\text{K})/273\ (\text{K})$$
$$= 0.86\ \text{m}^3$$

The test and the metered volumes for both the low [Table (3A.3)] and the high [Table (3A.4)] test flowrates are very close: In the low flowrate case ($9\text{m}^3/\text{h}$) a total of 5 kg of N_2 gas was passed through the meter giving a calculated gas volume of 4.30m^3; the metered volume was 4.42m^3 giving an overestimate of 2.8%. In the high gas flowrate case ($36\ \text{m}^3/\text{h}$), a total of 8 kg of nitrogen gas was passed through the meter giving a calculated volume of $6.88\ \text{m}^3/\text{h}$; the metered volume was $6.83\ \text{m}^3$ giving an underestimate of 0.7%. From these results it was accepted that the gas meter gave reliable gas volume readings.

TABLE 3A.3: VENT GAS METER LOW FLOW CALIBRATION

TIME min'sec"	CYLINDER MASS kg	TEST FLOW kg/h	TEST VOL m^3	GAS METER READING m^3	METER VOL m^3
0'00"	78			1134.51	
4 18	77	14.0	0.86	1135.47	0.96
9 45	76	11.0	0.86	1136.36	0.89
15 19	75	10.8	0.86	1137.20	0.84
20 47	74	11.0	0.86	1138.06	0.86
27 16	73	9.3	0.86	1138.93	0.87
TOTAL	5		4.30		4.42

TABLE 3A.4: VENT GAS METER HIGH FLOW CALIBRATION

TIME	CYLINDER MASS kg	TEST FLOW kg/h	TEST VOL m^3	GAS METER READING m^3	METER VOL m^3
0'00"	77			1127.02	
2 22	76	25.4	0.86	1127.85	0.83
4 51	75	24.2	0.86	1128.74	0.89
0 00	75			1128.74	
2 10	74	27.7	0.86	1129.56	0.82
4 32	73	25.4	0.86	1130.43	0.87
6 52	72	25.7	0.86	1131.27	0.84
9 12	71	25.7	0.86	1132.13	0.86
11 36	70	25.0	0.86	1133.00	0.87
13 56	69	25.7	0.86	1133.85	0.85
TOTAL	8		6.88		6.83

APPENDIX 3B

CALCULATIONS TO DETERMINE THE REACTOR VENT GAS COMPOSITION

VENT GAS HANDLING AND MEASUREMENTS

During the specific heat yield test, the vent gas from the reactor was collected and cooled to ±20° C in a double jacket, water cooled condenser (see Chapter 3, Section 3.4.1.2). The water vapour that condensed from the vent gas by this cooling was collected and from it the condensate accumulation rate M_{wc} (kg/h) over the test period was calculated. The gas leaving the condenser was only partially dried in that it was still saturated at the condenser temperature (20° C). In order to determine the total water vapour mass flow rate [M_{we} in Eq (3.11)], the mass flow rate of the remaining uncondensed water vapour in the vent gas M_{wr} needed to be determined. The development which follows, sets out how this was calculated. In this development it was accepted that vent gas left the condenser at 20° C and was saturated with water vapour at this temperature.

After cooling the vent gas to 20° C, it was reheated slightly (by 7-15° C) in order to prevent water vapour condensation in the Parkinson Cowan gas meter which followed. The temperature of the vent gas and its O_2 content were measured after reheating but before metering. The pressure of the gas meter was atmospheric and assumed to be always at 760 mm Hg (i.e. barometric pressure variations were ignored). The dry vent gas composition was analyzed with an Orsat apparatus which was graduated in tenths of a percent and which enabled the fractional volume of the CO_2 and O_2 in the vent gas to be measured. When both the CO_2 and the O_2 were absorbed by the Orsat, the remaining volume of gas was too small to be measurable. Consequently all components of the dry vent gas other than CO_2 and O_2 were assumed to be negligible. From this it was reasonable to assume that the vent gas from the aerobic reactor comprised only CO_2, O_2 and water vapour.

The following information was available from the vent gas measurements:

(1) the temperature of the vent gas at the point of exit from the reactor, T_{ge}

(2) the volume of gas passing through the gas meter V_{gm}, over a selected period of time (the specific heat yield test)

(3) the temperature of the gas passing through the gas meter T_{gm}, and
(4) the fractional volume of oxygen in the dry vent gas, f_{O2}.

With the aid of the above 4 measurements and also knowing the oxygen supply rate, O_s, it is possible to calculate the following information required in the unsteady state heat and oxygen mass balances from which, in turn, the specific heat yield was calculated:

(1) the volume of dry gas leaving the reactor at STP (0°C, 760 mm Hg)
(2) the mass of water remaining in the partially dried vent gas
(3) the mass of oxygen consumed by the aerobic reactor.

The 4 measurements also allowed calculation of the following additional information not directly related to the specific heat yield, i.e.

(4) the respiration quotient, Y_{CO_2}, i.e. the moles of CO_2 produced per mole of O_2 consumed, and
(5) the degree of water vapour saturation of the vent gas at the point of exit from the reactor.

The assumptions on which the calculations of the above 5 parameters are based have been mentioned earlier but are summarized below in the interests of clarity:

(1) The vent gas passing through the gas meter contains only oxygen, carbon dioxide and water vapour.

(2) The pressure of the vent gas is 760 mm Hg and the gas is saturated with water vapour at 20°C after the condenser.

(3) Ideal gas laws apply, i.e. 1 kmole of gas occupies 22.4 m^3 at 0°C and 760 mm Hg.

(4) The molar masses of oxygen, carbon dioxide and water are 32, 44 and 18 kg/kmole respectively.

1. The volume of dry gas leaving the reactor at STP (0°C, 760mm Hg)

The saturation vapour pressure of water vapour at 20°C is 18.1 mm Hg. Hence 1 m^3 of gas saturated with water at 20°C and 760 mm Hg contains:

1 m^3 of water vapour at 18.1 mm Hg and 20°C
1 m^3 of dry gas at 741.9 mm Hg and 20°C

At 760 mm Hg and 20°C, the volume of dry gas is:

$$1 * 741.9/760 = 0.9762 \text{ m}^3$$

Hence every m^3 of vent gas passing through the gas meter contains 0.9762 m^3 of dry gas and 0.0238 m^3 of water vapour.

The standard volume (0°C and 760 mm Hg) of dry vent gas passing through the meter is:

$$0.9762 * V_{gm} * 273/(273 + T_{gm})$$

$$= 266.5 * V_{gm}/(273 + T_{gm}) \quad (\text{m}^3/\text{h}) \qquad (3B.1)$$

where

V_{gm} =volumetric flow rate of vent gas (m^3/h) passing through the gas meter, i.e. the volume of vent gas passing through the gas meter during a specific heat yield test divided by the test time period.

T_{gm} =the temperature of the vent gas passing through the gas meter (°C).

2. The mass of water remaining in the partially dried vent gas

From the calculations in (1) above, each m^3 of gas passing through the gas meter contains 0.0238 m^3 of water vapour. The standard volume of this water vapour is:

$$0.0238 * 273/(273 + T_{gm}) \quad (\text{m}^3/\text{m}^3)$$

Hence the mass of this water vapour is:

$$0.0238 * 273/(273 + T_{gm}) * 18/22.4$$
$$= 5.22/(273 + T_{gm}) \quad (\text{kg/m}^3)$$

Therefore the mass flow rate of water vapour in the partially dried vent gas is:

$$M_{wr} = 5.22 * V_{gm}/(273 + T_{gm}) \quad \text{(kg/h)} \qquad (3B.2)$$

This mass flow rate was added to the measured condensate accumulation rate M_{wc} in the primary and secondary water traps to give the total water vapour mass flow rate, M_{we}.

3. The mass of oxygen consumed by the reactor, O_c

From Eq (3B.1), the standard volume of dry gas passing through the meter is:

$$266.5 * V_{gm}/(273 + T_{gm}) \quad (m^3/h)$$

The standard volume of oxygen passing through the meter is:

$$f_{O_2} * 266.5 * V_{gm}/(273 + T_{gm}) \quad (m^3/h)$$

The mass flow rate of oxygen in the vent gas is:

$$f_{O_2} * 266.5 * V_{gm}/(273 + T_{gm}) * 32/22.4$$

$$380.7 * f_{O_2} * V_{gm}/(273 + T_{gm}) \quad \text{(kgO/h)}$$

Hence the mass of oxygen consumed in the reactor, O_c, is:

$$O_c = O_s - 380.7 * f_{O_2} * V_{gm}/(273 + T_{gm}) \quad \text{(kgO/h)} \qquad (3B.3)$$

where

O_s = the oxygen supply rate (kgO/h) [Eq. (3.1) of Chapter 3]

4. The respiration quotient, Y_{CO_2}

The respiration quotient, Y_{CO_2}, is the number of moles of CO_2 produced per mole of O_2 consumed.

The molar flow rate of CO_2 in the vent gas is calculated as follows:

From Eq (3B.1), the standard volume of dry gas passing through the meter is:

$$266.5 * V_{gm}/(273 + T_{gm}) \quad (m^3/h)$$

The standard volume of carbon dioxide passing through the meter is:

$$(1 - f_{O2}) * 266.5 * V_{gm}/(273 + T_{gm}) \quad (m^3/h)$$

The molar flow rate of carbon dioxide in the vent gas is:

$$(1 - f_{O2}) * 266.5 * V_{gm}/(273 + T_{gm}) * 1/22.4$$

$$= (1 - f_{O2}) * 11.9 * V_{gm}/(273 + T_{gm}) \quad (kmol/h)$$

The molar oxygen consumption rate by the reactor is $O_c/32$ (kmol/h)

Thus, the respiration quotient,

$$Y_{CO_2} = [(1 - f_{O2}) * 11.9 * V_{gm}/(273 + T_{gm})]/[O_c/32]$$

$$= 380.7 * (1 - f_{O2}) * V_{gm}/[(273 + T_{gm}) \times O_c] \qquad (3B.4)$$

[Equation (3.3) in Chapter 3]

5. The degree of water vapour saturation of the vent gas leaving the reactor

The degree of water vapour saturation in the vent gas was assessed by comparing the measured total water vapour mass flow rate in the reactor vent gas, M_{we}, with a predicted total water vapour mass flow rate, assuming the vent gas is saturated. For the predicted water vapour mass flow rate the molar flow rates of the CO_2 and O_2 in the vent gas and the temperature and the pressure of the vent gas were accepted to be equal to those of the measured vent gas for each of the 116 specific heat yield tests.

In the Milnerton reactor, the temperature of the vent gas at the point of exit from the reactor, T_{ge}, increased linearly with time concomitantly with the reactor sludge temperature increase over the specific heat yield test (see Table 3.1). This change was small (≈3°C) and for the purpose of calculating the predicted saturated water vapour mass flow rate, the average T_{ge} over the specific heat yield test period was accepted for the temperature of the vent gas. The total pressure of the vent gas was atmospheric and in these calculations a constant pressure of 760 mmHg was accepted for all 116 tests.

An equation to determine the volumetric flow rate of the dry vent gas at STP from the specific heat yield test data was derived in Section (1) above. This volumetric flow rate is converted to a molar flow rate (kmol/h) by dividing by the molar volume of an ideal gas, i.e. 22.4 m³ (STP)kmol. Hence the molar flow rate of the dry vent gas is:

$$m_{dg} = 266.5/22.4 * V_{gm}/(273 + T_{gm})$$

$$= 11.9 * V_{gm}/(273 + T_{gm}) \qquad (3B.5)$$

where

m_{dg} = the molar flow rate of dry vent gas (kmol/h)

Predicting the water vapour mass flow rate for the vent gas if saturated

When a 2 phase system consisting of gas and water are in a state of equilibrium, the gas will be saturated with water vapour. According to ideal gas laws, in this state of equilibrium, the following ratios will be equal:

$$m_w/m_{dg} = p_w/p_{dg} \qquad (3B.6)$$

where

m_w = the molar flow rate of water vapour in the saturated gas (kmol/h)
m_{dg} = the molar flow rate of dry gas (kmol/h)
p_w = the partial pressure of the water vapour (mmHg)
p_{dg} = the partial pressure of the dry gas (mmHg)

Note that in Eq (3B.6):

$$m_w = M_{wp}/18$$

where

M_{wp} = the predicted mass flow rate of water vapour in the vent gas if saturated (kg/h)

18 = the molar mass of water (kg/kmol)

Thus the predicted mass flow rate of water in the vent gas if saturated is:

$$M_{wp} = 18 * m_{dg} * p_w/p_{dg} \quad \text{(kg/h)} \qquad (3B.7)$$

The partial pressure of water vapour in the saturated vent gas may be calculated with the aid of the Antoine equation:

$$\text{Log}(p_w) = B - A/(T_{ge} + 273)$$

where

B = 8.896

A = –2238

T_{ge} = the temperature of the gas/vapour at the point of exit from the reactor (°C).

Hence,

$$p_w = 10^{(8.896 - 2238/(Tge + 273))} \quad \text{(mmHg)} \qquad (3B.8)$$

Note that, from Dalton's law, the total pressure of the saturated vent gas is equal to the sum of the partial pressures in the gas, so that,

$$\begin{aligned} p_{dg} &= P_T - p_w \\ &= 760 - p_w \quad \text{(mmHg)} \qquad (3B.9) \end{aligned}$$

where

P_T = the total pressure of the saturated vent gas (760 mmHg)

Thus the predicted mass flow rate of water vapour in the vent gas if saturated was calculated with the aid of the following equations:

$$M_{wp} = 18 * m_{dg} * p_w/p_{dg} \quad \text{(kg/h)} \quad \text{[Eq (3B.7)]}$$

where

$m_{dg} = 11{,}9 * V_{gm}/(273 + T_{gm})$ (kmol/h) [Eq (3B.5)]

$p_w = 10^{(8.896 - 2238/(Tge + 273)}$ (mmHg) [Eq (3B.5)]

$p_{dg} = 760 - p_w$ (mmHg) [Eq (3B.9)]

APPENDIX 3C

DETERMINATION OF THE HEAT INPUT FROM THE VITOX RECIRCULATION PUMP (H_{mi})

The Vitox recirculation pump operated continuously. The kinetic energy imparted to the pumped sludge was converted to heat in the aerobic reactor and the Vitox pipework.

Of the total electrical power consumed by the recirculation pump, a portion was lost due to inefficiency in the electric motor and a portion due to friction in the transmission system. The balance of the power was all converted to heat imparted to the pumped sludge. This heat was calculated with the aid of the following equation:

$$H_{mi} = f_{mech} \times V \times I \times \sqrt{3} \times \cos\phi \times \frac{3.6}{1000} \quad (\mathrm{MJ/h}) \qquad (3C.1)$$

where

H_{mi} = the rate of heat generation in the sludge due to the mechanical action of the recirculation pump (MJ/h)

f_{mech} = the fraction of electrical power, consumed by the recirculation pump's electric motor, which was converted to heat in the sludge

V = the phase voltage (volts)

I = the phase current (amps)

$\cos\phi$ = conversion from watts to MJ/h

At Milnerton the phase voltage was 380 V and the current drawn by the recirculation pump (I) was read at the pump's ammeter. Hence in Eq (3C.1), f_{mech} and $\cos\phi$ are the only unknowns. For the power factor, $\cos\phi$, a value of 0.82 was estimated from engineering data for the recirculation pump's electric motor. To obtain accurate estimation of H_{mi}, required the power to heat conversion factor, f_{mech}, to be measured on the reactor/Vitox system. The method in which this was done and the results obtained are discussed in this Appendix.

The method of measuring the power to heat conversion factor, f_{mech}

The aerobic reactor was filled with 45m³ of sludge at ambient temperature. The Vitox recirculation pump was operated but no oxygen was injected into the recirculated sludge. In the absence of oxygen addition, no biological heat was generated in the reactor and no vent gas was produced - without vent gas production there was no heat loss via water vaporization from the sludge. Also, heat losses through the reactor walls were negligible because the temperature difference between the reactor sludge and ambient surroundings remained small throughout the test period. Under these operating conditions the unsteady state heat balance about the reactor reduces to:

$$H_{net} = H_{mi}$$

$$= V_p \times \rho_s \times Cp \times \frac{dT_{se}}{dt} \quad (MJ/h) \qquad (3C.2)$$

Hence from Eqs (3C.1) and (3C.2):

$$V_p \times \rho_s \times Cp \times \frac{dT_{se}}{dt} = f_{mech} \times V \times I \times \sqrt{3} \times \cos\phi \times \frac{3.6}{1000} \quad (MJ/h) \quad (3C.3)$$

The values of dT_{se}/dt and I were measured during the test. The other values required in Eq (3C.3) were as follows:

V_p = 45m³ : ρ_s = 1 ton/m³ : Cp = 4.184 MJ/ton : V = 380 volts : $\cos\phi$ = 0.82.

Substituting these values into Eq (3C.3) allows f_{mech} to be solved from measured values of I and dT_{se}/dt with the following equation:

$$f_{mech} = 96.9 \times \frac{dT_{se}/dt}{I} \qquad (3C.4)$$

Results of f_{mech} determination:

Three f_{mech} tests were performed on 23/10/87, 11/11/87 and 17/11/87. In the 1st and 2nd tests, dT_{se}/dt was determined by measuring the initial and final values of T_{se} over 6.25 and 4 hour periods respectively. The linearity of temperature change with time was checked by placing a rule on the chart recorder temperature trace. Readings from all 4 reactor PT100's were taken and each of the 4 sets were found to be linear. In the 3rd test, the T_{se} value displayed by 1 of the 4 PT100 probes was recorded at intervals and a linear regression was used to calculate dT_{se}/dt and to confirm its linearity. The results obtained for the 1st, 2nd and 3rd tests are listed below in Tables (3C.1) to (3C.3) respectively.

The results of dT_{se}/dt and I for each of the 3 tests were substituted into Eq (3C.4) and the following values of f_{mech} were obtained for tests 1 to 3 respectively: 0.88, 0.9, 0.78. The average of these 3 values, i.e. 0.85, was used in all the heat balance calculations presented in this thesis. With f_{mech} = 0.85, the mechanical heat input via the recirculation pump reduces to an equation in terms of the electric current drawn, I, as follows:

$$H_{mi} = 1.65 \times I \quad (MJ/h) \qquad (3C.5)$$

[Equation (3.8) in Chapter 3]

TABLE 3C.1. RESULTS OF TEST 1.

TIME (h)	PT100 No.	T_{se} (°C)	T_{amb} (°C)	$d(T_{se})/dt$ (°C/h)	PUMP AMPS
10h45	1	23.2	19.0		42
	2	23.0			
	8	23.4			
	10	23.0			
17h00	1	25.6	20.0	0.384	42
	2	25.4		0.384	
	8	25.8		0.384	
	10	25.4		0.384	

TABLE 3C.2. RESULTS OF TEST 2.

TIME (h)	PT100 No	T_{se} (°C)	T_{amb} (°C)	$d(T_{se})/dt$ (°C/h)	PUMP AMPS
11h00	1	23.7	27.3		43
	2	23.3			
	8	23.7			
	10	23.4			
15h00	1	25.3	30.9	0.4	43
	2	24.8		0.375	
	8	25.4		0.425	
	10	25.0		0.4	

TABLE 3C.3. RESULTS OF TEST 3.

TIME	PERIOD	T_{se} (°C)	PUMP AMPS	DATA
13h00	0	25.6	41	
17h00	4	27.1	41	
01h00	12	29.9	41	
02h00	13	30.2	40	
03h00	14	30.5	41	
04h00	15	30.8	41	
05h00	16	31.1	40	
06h00	17	31.4	41	
07h00	18	31.7	41	<u>$d(T_{se})/dt$</u>
08h00	19	32.2	41	0.33 (°C/h)
09h00	20	32.4	41	
11h00	22	33.0	41	<u>CORRELATION</u>
12h00	23	33.3	41	0.9991

APPENDIX 3D

DETERMINATION OF THE REACTOR WALL HEAT LOSSES, H_{we}

Heat was lost from the aerobic reactor and the Vitox system by conduction through and convection away from the system walls. Both the reactor and Vitox system were lagged to reduce these losses.

The rate of heat loss can be described by the following equation:

$$H_{we} = U \times A \times (T_{se} - T_{amb}) \qquad (3D.1)$$

where

Hwe = the rate of heat loss from the surfaces of the reactor and Vitox oxygenation system (MJ/h)

U = the overall heat transfer coefficient (MJ/(m^2.°C.h)]

A = surface area of the reactor and the Vitox oxygenation system (m^2)

T_{se} = the temperature of the reactor sludge (°C)

T_{amb} = the ambient temperature (°C)

The overall heat transfer coefficient, U can be estimated from engineering data but, due to the inaccuracies of this approach, it was decided that a better result would be obtained by directly measuring H_{we} and then estimating U from the measured H_{we} values. The method with which this was done and the results obtained are described in this Appendix.

The method of determining the overall heat transfer coefficient, U

With the reactor filled to its operating volume of 45m^3 with sludge at its normal pasteurization temperature (±60°C) and the Vitox recirculation pump in operation, the injection of oxygen into the sludge was stopped. In the absence of oxygen injection, no biological heat was generated in the reactor and no vent gas was produced - without vent gas production there was no heat loss via water vaporization from the sludge. Under these conditions of operation the unsteady state heat balance about the reactor [Eqs (3.5) and (3.6)] reduces to:

$$H_{net} = H_{mi} - H_{we} \ \text{(MJ/h)} \qquad (3D.2)$$

where

H_{net} = the rate of increase of reactor sludge enthalpy (MJ/h)

H_{mi} = the rate of heat generation in the sludge due to the mechanical action of the recirculation pump (MJ/h)

This procedure requires the recirculation pump heat input, H_{mi}, to be known. This was obtained from the calibrated Eq (3.8) i.e. H_{mi} = 1.65*I (see Appendix 3C). Substituting Eqs (3.6), (3.8) and (3D.1) for H_{net}, H_{mi} and H_{we} respectively into Eq (3D.2) yields:

$$V_p \times \rho_s \times Cp \times \frac{d(T_{se})}{dt} = 1.65 \times I - U \times A \times (T_{se} - T_{amb}) \text{ (MJ/h)} \qquad (3D.3)$$

By recording, over the time period of a test, (1) the change in reactor sludge temperature with time i.e. $d(T_{se}/dt)$, (2) the ambient temperature, T_{amb}, and (3) the electric current drawn by the pump motor, I, the value of U in Eq (3D.3) could be calculated. The other values required in Eq (3D.3) were as follows:

$V_p = 45m^3$: ρ_s = 1 ton/m : Cp = 4.184 MJ/ton : A = $77m^2$:

Substituting these values into Eq (3D.3) allows U to be solved from the following equation:

$$U = \frac{1.65 \times I - 188.28 \times \frac{dT_{se}}{dt}}{77 \times (T_{se} - T_{amb})} \quad [(MJ/(m^2.{}^\circ C.h)] \qquad (3D.4)$$

<u>**Results of U determination**</u>

A single test to determine U was performed over a 10 hour period on 13/6/1988. At 11h00 and 21h00, the reactor sludge temperature, T_{se}, was determined using 3 of the 4 reactor PT100's. The ambient temperature, T_{amb}, was recorded over the same interval and the current drawn by the recirculation pump was constant at 46 amps. The linearity of dT_{se}/dt was checked on the chart recorder trace of T_{se} and found to be satisfactory. A value of 0.25°C/h was obtained for dT_{se}/dt. Table (3D.1) below summarizes the results obtained in the U determination test.

Substituting the measured values of dT_{se}/dt (0.25°C/h), I (46 amps), T_{se} (63.7° C[1]) and T_{amb} (12.5° C) into Eq (3D.4) and solving yields a value for U of 0.0073 $MJ/\{m^2.^\circ C.h\}$. Accepting this value for U, Eq (3D.1) describing the wall heat loss simplifies to:

$$H_{we} = 0.564 \times (T_{se} - T_{amb}) \quad (MJ/h) \tag{3B.5}$$

This simplified equation above [Eq (3D.5)] was used to calculate the wall heat losses from the reactor from the observed T_{se} and T_{amb} in all the heat balance results reported in this thesis.

TABLE 3D.1. TEST DATA USED TO CALCULATE COEFFICIENT U.

TIME	PT100 No	T_{se} (°C)	T_{amb} (°C)	$d(T_{se})/dt$ (°C/h)	PUMP AMPS
11h00	1	62.0	13.2		46
	8	62.1			
	10	62.4			
21h00	1	64.5	11.7	0.25	46
	8	64.6		0.25	
	10	64.9		0.25	

Comments

Before accepting Eq (3D.5) as a valid means by which H_{we} may be predicted during specific heat yield tests, 2 factors which influence the accuracy of this equation are discussed below.

(1) During the above test, overcast weather and strong winds prevailed. During the specific yield tests, weather conditions were often windless and sunny with half of the reactor exposed to the sun and the other half to shade. Ambient

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[1] The values of T_{se} and T_{amb} changed during the test as shown in Table 1. The changes were small enough to allow average values to be used in the calculation of U.

temperatures during these tests always were measured in the shade. Under these conditions it is doubtful that heat losses would have been as high as predicted by Eq (3D.5).

(2) During normal operation of the reactor, water vapour was conveyed from the sludge by the vent gases arising from sludge oxygenation. A small amount of this vapour condensed on the walls of the $5m^3$ reactor headspace, releasing heat during condensation, some of which was conducted out of the reactor through the headspace walls. As a result of this heat loss the vent gas was cooled by ±2 to 5°C depending on the vent gas flow rate through the headspace. This heat loss through the headspace walls did not occur during the test to determine the overall heat transfer coefficient, U, as the reactor sludge was not oxygenated during the test. However, this heat loss is negligibly small; it is equal to the heat of vaporization of water in cooling the vent gas flow from the reactor sludge temperature (T_{se}) to the vent gas temperature at the point of exit from the reactor (T_{ge}); in cooling from T_{ge} to the condenser temperature (20°C) the heat loss is less than 10% of the total heat losses, so the heat loss in cooling from T_{se} to T_{ge} is negligible and can be ignored. Consequently even though the value of U would have been slightly underestimated by the particular test method compared to normal operation, the difference is small enough to be of no practical significance.

The above 2 inaccuracies in the calculation of H_{we} by Eq (3D.5) are not likely to have significantly affected the outcome of the determination of the specific heat yield. In the particular wall heat loss test (T_{se} = 63.7°C, T_{amb} = 12.5°C), the heat loss rate H_{we} amounts to 28.8 MJ/h which is less than 10% of the heat sources during normal operation. Consequently, the accuracy of Eq (3D.5) is not important, as even large errors in the prediction of H_{we} by Eq (3D.5) have only minor influence on the overall heat balance.

APPENDIX 4A

DATA AND RESULTS FOR THE 116 SPECIFIC HEAT YIELD TESTS

DATE	T_{amb} °C	T_{si} °C	T_{se} °C max	T_{se} °C avg	R_H days	V_{gm} m^3/h	T_{gm} °C	f_{O2} %	O_s kgO/h	M_{wc} kg/h	T_{ge} °C low	T_{ge} °C high	$\Delta T_{se}/\Delta t$ °C/h	I amps	M_{we} kg/h	O_c kgO/h	O_{eff}	Y_{CO2}	H_{net} MJ/h	H_{mi} MJ/h	H_{ve} MJ/h	H_{we} MJ/h	H_{bi} MJ/h	Y_H MJ/kgO
7/6/88	13.1	19.4	61.7	60.0	1.25	11.90	26.0	18	22.0	1.512	57.0	59.3	1.477	43	1.72	19.27	87.6	0.64	278.1	71.0	4.1	26.45	237.7	12.3
9/6/88	11.0	19.1	63.4	61.7	1.25	12.10	30.0	17	22.0	1.450	56.8	61.0	1.615	44	1.66	19.42	88.3	0.65	304.1	72.6	3.9	28.59	264.0	13.6
15/6/88	13.0	17.8	67.5	66.2	1.25	14.96	27.0	32	22.0	2.720	63.7	64.9	1.250	44	2.98	15.93	72.4	0.81	235.4	72.6	7.0	30.00	199.8	12.5
20/6/88	15.9	19.2	63.0	61.4	1.25	12.42	27.0	23	22.0	1.815	58.5	60.6	1.500	44	2.03	18.37	83.5	0.66	282.4	72.6	4.8	25.66	240.3	13.1
21/6/88	14.1	19.1	63.5	61.8	1.25	11.95	26.0	21	22.0	1.750	58.6	61.3	1.523	44	1.96	18.80	85.5	0.64	286.8	72.6	4.6	26.90	245.7	13.1
23/6/88	15.8	18.9	63.5	62.5	1.25	12.50	28.5	17	22.0	1.830	58.6	61.4	1.523	44	2.05	19.32	87.8	0.68	286.8	72.6	4.8	26.34	245.3	12.7
24/6/88	10.3	19.2	63.2	61.6	1.25	11.90	26.5	21	22.0	1.780	58.3	61.1	1.523	44	1.99	18.82	85.5	0.63	286.8	72.6	4.7	28.93	247.8	13.2
24/6/88	19.4	19.1	63.2	61.1	1.25	7.05	35.0	10	14.5	1.025	58.1	60.1	1.154	44	1.14	13.63	94.0	0.58	217.2	72.6	2.7	23.52	170.8	12.5
25/6/88	20.3	19.3	57.9	56.6	1.25	7.90	33.5	9	14.5	0.835	53.0	55.7	1.200	43	0.97	13.62	93.9	0.66	225.9	71.0	2.3	20.47	177.7	13.0
27/6/88	13.1	18.4	54.5	53.2	1.25	7.55	26.5	10	14.5	0.725	49.4	51.8	1.200	44	0.86	13.54	93.4	0.64	225.9	72.6	2.0	22.62	177.9	13.1
28/6/88	14.9	18.8	54.5	53.2	1.25	8.38	31.0	10	14.5	0.705	49.9	52.3	1.200	46	0.85	13.45	92.8	0.70	225.9	75.9	2.0	21.60	173.6	12.9
28/6/88	6.9	18.9	54.5	53.1	1.25	8.00	24.0	10	14.5	0.725	49.4	51.8	1.200	45	0.87	13.47	92.9	0.69	225.9	74.3	2.1	26.06	179.8	13.3
29/6/88	9.0	19.3	54.5	53.2	1.25	8.07	21.0	10	14.5	0.720	49.5	51.5	1.200	45	0.86	13.46	92.8	0.70	225.9	74.3	2.1	24.93	178.6	13.3
29/6/88	12.8	19.4	55.2	53.5	1.25	13.81	26.0	24	24.5	0.880	50.6	53.0	1.600	43	1.12	20.28	82.8	0.66	301.2	71.0	2.7	22.95	255.9	12.6
30/6/88	14.4	19.2	55.2	54.0	1.25	8.25	27.0	10	14.5	0.940	51.0	52.0	1.150	44	1.08	13.45	92.8	0.70	216.5	72.6	2.6	22.33	168.8	12.6
30/6/88	19.9	19.5	56.0	54.2	1.25	14.40	32.0	23	24.3	1.085	50.5	53.3	1.546	43	1.33	20.17	83.0	0.69	291.1	71.0	3.2	19.35	242.7	12.0
1/7/88	17.2	19.4	61.6	59.9	1.25	14.13	29.0	23	24.3	1.755	56.6	59.4	1.500	42	2.00	20.20	83.1	0.68	282.4	69.3	4.7	24.08	241.9	12.0

DATE	T_{amb} °C	T_{si} °C	T_{se} °C max	T_{se} °C avg	R_H days	V_{gm} m^3/h	T_{gm} °C	f_{O2} %	O_s kgO/h	M_{wc} kg/h	T_{ge} °C low	T_{ge} °C high	$\Delta T_{se}/\Delta t$ °C/h	I amps	M_{we} kg/h	O_c kgO/h	O_{eff}	Y_{CO2}	H_{net} MJ/h	H_{mi} MJ/h	H_{ve} MJ/h	H_{we} MJ/h	H_{bi} MJ/h	Y_H MJ/kgO
4/7/88	15.6	18.8	61.7	60.3	1.25	14.20	24.0	26	24.3	1.900	57.5	60.1	1.400	43	2.15	19.57	80.5	0.69	263.6	71.0	5.1	25.21	222.9	11.4
7/7/88	17.9	18.8	61.8	60.1	1.25	14.40	28.5	26	24.3	1.800	57.0	59.4	1.500	42	2.05	19.57	80.5	0.69	282.4	69.3	4.8	23.80	241.7	12.4
11/7/88	21.0	17.9	60.0	58.2	1.25	13.95	25.5	29	24.3	1.845	55.1	57.1	1.431	42	2.09	19.14	78.8	0.66	269.4	69.3	4.9	20.98	226.0	11.8
13/7/88	15.9	18.0	62.4	60.8	1.25	14.30	28.0	36	24.3	2.200	57.7	60.3	1.431	43	2.45	17.79	73.2	0.65	269.4	71.0	5.8	25.32	229.5	12.9
15/7/88	16.9	18.8	62.5	60.8	1.25	14.10	27.5	28	24.3	1.930	57.5	60.1	1.523	43	2.17	19.30	79.4	0.67	286.8	71.0	5.1	24.76	245.7	12.7
18/7/88	15.8	17.9	61.1	59.5	1.25	15.05	25.0	38	24.3	2.035	56.7	59.0	1.408	46	2.30	16.99	69.9	0.70	265.0	75.9	5.4	24.65	219.2	12.9
19/7/88	15.0	18.6	59.7	58.2	1.25	15.15	23.5	40	24.3	1.790	55.2	57.3	1.292	44	2.06	16.52	68.0	0.71	243.3	72.6	4.9	24.36	200.0	12.1
20/7/88	12.5	18.2	59.0	57.5	1.25	14.90	22.0	40	24.3	1.700	53.6	56.2	1.362	44	1.96	16.61	68.4	0.69	256.4	72.6	4.7	25.38	213.9	12.9
21/7/88	15.5	17.8	59.9	58.3	1.25	12.70	25.0	28	22.0	1.360	54.9	56.7	1.385	44	1.58	17.46	79.4	0.67	260.7	72.6	3.7	24.14	215.9	12.4
22/7/88	16.3	18.3	60.3	58.6	1.25	12.50	27.0	24	22.0	1.565	55.3	57.6	1.454	44	1.78	18.19	82.7	0.66	273.7	72.6	4.2	23.86	229.2	12.6
1/8/88	16.7	18.3	60.6	59.0	1.25	11.25	27.5	23	22.0	1.270	55.6	58.0	1.408	44	1.47	18.72	85.1	0.59	265.0	72.6	3.5	23.86	219.8	11.7
2/8/88	16.0	18.6	60.8	59.1	1.25	12.40	27.0	27	22.0	1.550	55.9	58.1	1.431	44	1.77	17.75	80.7	0.65	269.4	72.6	4.2	24.31	225.3	12.7
3/8/88	10.0	19.0	60.6	59.0	1.25	12.35	24.0	30	22.0	1.575	55.9	58.7	1.408	43	1.79	17.25	78.4	0.64	265.0	71.0	4.2	27.64	225.8	13.1
5/8/88	16.7	18.9	60.4	58.9	1.25	12.45	30.0	29	22.0	1.395	55.8	58.2	1.338	43	1.61	17.46	79.4	0.64	252.0	71.0	3.8	23.80	208.6	11.9
23/8/88	16.4	19.3	59.4	57.9	1.25	13.15	27.5	34	22.0	1.425	54.5	57.0	1.315	45	1.65	16.34	74.3	0.67	247.7	74.3	3.9	23.41	200.7	12.3
24/8/88	20.6	19.3	59.3	57.7	1.25	12.70	29.5	31	22.0	1.325	54.7	56.6	1.408	44	1.54	17.05	77.5	0.65	265.0	72.6	3.7	20.92	217.0	12.7
25/8/88	15.5	19.6	59.8	58.1	1.25	12.65	24.0	30	22.0	1.350	54.6	57.0	1.431	45	1.57	17.14	77.9	0.66	269.4	74.3	3.7	24.03	222.8	13.0

DATE	T_{amb} °C	T_{si} °C	T_{se} °C max	T_{se} °C avg	R_H days	V_{gm} m^3/h	T_{gm} °C	f_{O2} %	O_s kgO/h	M_{wc} kg/h	T_{ge} °C low	T_{ge} °C high	$\Delta T_{se}/\Delta t$ °C/h	I amps	M_{we} kg/h	O_c kgO/h	O_{eff}	Y_{CO2}	H_{net} MJ/h	H_{mi} MJ/h	H_{ve} MJ/h	H_{we} MJ/h	H_{bi} MJ/h	Y_H MJ/kgO
25/8/88	19.1	19.6	59.7	58.1	1.50	12.90	28.5	30	22.0	1.380	55.0	57.4	1.408	45	1.60	17.11	77.8	0.67	265.0	74.3	3.8	22.00	216.5	12.7
26/8/88	16.1	19.6	59.6	58.0	1.25	12.65	25.0	30	22.0	1.340	54.5	56.6	1.408	43	1.56	17.15	78.0	0.66	265.0	71.0	3.7	23.63	221.3	12.9
29/8/88	15.6	18.4	57.8	56.2	1.25	12.85	26.0	33	22.0	1.185	52.7	54.9	1.385	44	1.41	16.60	75.5	0.66	260.7	72.6	3.3	22.90	214.3	12.9
30/8/88	13.8	19.0	56.6	55.1	1.25	13.55	24.0	43	22.0	1.220	52.2	53.8	1.292	45	1.46	14.53	66.0	0.68	243.3	74.3	3.5	23.29	195.8	13.5
30/8/88	14.2	18.8	56.5	55.1	1.25	13.15	24.0	40	22.0	1.300	52.9	54.0	1.246	46	1.53	15.26	69.4	0.66	234.6	75.9	3.6	23.07	185.4	12.1
1/9/88	13.3	18.1	56.9	55.4	1.25	12.60	26.0	36	22.0	1	52.2	54	1.292	44	1.45	16.22	73.7	0.63	243.3	72.6	3.5	23.74	197.9	12.2
1/9/88	16.8	18.2	56.8	55.4	1.25	9.60	29.0	30	18.5	1	52.0	54	1.269	45	0.80	14.87	80.4	0.57	239.0	74.3	1.9	21.77	188.4	12.7
2/9/88	15.1	18.6	56.0	54.5	1.25	11.05	27.0	29	18.5	1	51.0	53	1.292	46	1.14	14.43	78.0	0.69	243.3	75.9	2.7	22.22	192.3	13.3
5/9/88	17.8	19.1	57.8	56.2	1.25	9.55	32.0	17	18.5	0.865	52.3	55.1	1.385	45	1.03	16.47	89.0	0.60	260.7	74.3	2.4	21.66	210.5	12.8
6/9/88	19.2	19.4	58.1	56.6	1.25	10.70	32.0	25	18.5	1.135	53.9	56.0	1.292	46	1.32	15.16	81.9	0.66	243.3	75.9	3.1	21.09	191.6	12.6
7/9/88	14.6	19.5	57.9	56.4	1.25	10.10	27.0	23	18.5	0.900	53.0	55.3	1.338	45	1.08	15.55	84.1	0.63	252.0	74.3	2.6	23.58	203.9	13.1
7/9/88	19.3	19.4	57.9	56.4	1.25	10.55	29.0	23	18.5	1.050	53.1	55.3	1.362	45	1.23	15.44	83.5	0.66	256.4	74.3	2.9	20.92	205.9	13.3
12/9/88	13.5	18.3	56.8	55.3	1.25	10.20	27.0	23	18.5	0.870	51.6	54.2	1.338	45	1.05	15.52	83.9	0.64	252.0	74.3	2.5	23.58	203.8	13.1
13/9/88	16.6	18.4	56.7	55.3	1.25	10.60	31.0	26	18.5	0.905	51.8	54.2	1.269	45	1.09	15.05	81.4	0.65	239.0	74.3	2.6	21.83	189.1	12.6
14/9/88	15.8	19.2	61.1	59.3	1.50	10.92	27.0	24	18.5	1.376	55.7	58.3	1.294	44	1.57	15.17	82.0	0.69	243.6	72.6	3.7	24.53	199.2	13.1
15/9/88	15.3	18.8	63.2	61.4	1.50	10.52	27.0	20	18.5	1.480	57.7	60.6	1.313	44	1.66	15.83	85.6	0.67	247.1	72.6	3.9	26.00	204.4	12.9
16/9/88	15.3	18.8	64.0	62.1	1.50	10.68	27.0	23	18.5	1.672	58.7	61.4	1.275	45	1.86	15.38	83.1	0.68	240.1	74.3	4.4	26.40	196.6	12.8

DATE	T_{amb} °C	T_{si} °C	T_{se} °C max	T_{se} °C avg	R_H days	V_{gm} m^3/h	T_{gm} °C	f_{O2} %	O_s kgO/h	M_{wc} kg/h	T_{ge} °C low	T_{ge} °C high	$\Delta T_{se}/\Delta t$ °C/h	I amps	M_{we} kg/h	O_c kgO/h	O_{eff}	Y_{CO2}	H_{net} MJ/h	H_{mi} MJ/h	H_{ve} MJ/h	H_{we} MJ/h	H_{bi} MJ/h	Y_H MJ/kgO
19/9/88	20.3	19.3	64.9	63.1	1.50	10.84	31.0	24	18.5	1.688	59.4	62.1	1.313	46	1.87	15.24	82.4	0.68	247.1	75.9	4.4	24.14	199.7	13.1
20/9/88	17.6	19.8	65.8	64.0	1.50	10.40	33.0	22	18.5	1.732	60.5	63.3	1.313	46	1.91	15.65	84.6	0.64	247.1	75.9	4.5	26.17	201.9	12.9
21/9/88	17.3	19.7	65.9	64.1	1.50	10.68	31.0	21	18.5	1.790	60.2	63.0	1.275	44	1.97	15.69	84.8	0.67	240.1	72.6	4.6	26.40	198.5	12.7
22/9/88	18.2	19.5	65.9	64.1	1.50	10.52	34.0	19	18.5	1.740	59.9	63.1	1.313	44	1.92	16.02	86.6	0.66	247.1	72.6	4.5	25.89	204.9	12.8
4/10/88	17.3	20.1	64.9	62.8	3.00	4.76	32.0	0	8.1	0.732	56.1	61.5	0.770	39	0.81	8.10	100.0	0.73	145.0	64.4	1.9	25.66	108.2	13.4
5/10/88	15.9	20.0	68.5	66.3	3.00	4.86	34.0	0	8.1	0.852	59.3	64.5	0.760	39	0.93	8.10	100.0	0.74	143.1	64.4	2.2	28.43	109.3	13.5
6/10/88	16.8	20.2	68.8	66.9	3.00	3.86	35.0	0	6.7	0.558	57.9	63.1	0.640	38	0.62	6.70	100.0	0.71	120.5	62.7	1.5	28.26	87.6	13.1
7/10/88	17.4	20.5	68.7	66.7	3.00	3.74	37.0	0	6.3	0.638	59.5	63.9	0.670	39	0.70	6.30	100.0	0.73	126.1	64.4	1.6	27.81	91.1	14.5
8/10/88	16.6	21.2	69.1	67.2	3.00	3.72	36.0	0	6.0	0.558	58.6	63.6	0.640	38	0.62	6.00	100.0	0.76	120.5	62.7	1.5	28.54	87.8	14.6
10/0/88	16.0	20.2	69.1	67.1	3.00	3.60	35.0	0	6.0	0.524	59.0	62.8	0.620	38	0.59	6.00	100.0	0.74	116.7	62.7	1.4	28.82	84.2	14.0
11/10/88	20.0	20.8	68.1	66.2	3.00	3.72	39.0	0	6.5	0.564	59.7	63.5	0.660	39	0.63	6.50	100.0	0.70	124.3	64.4	1.5	26.06	87.5	13.5
12/10/88	22.0	20.5	67.9	65.9	3.00	3.68	37.0	0	6.4	0.600	59.8	63.7	0.640	38	0.66	6.40	100.0	0.71	120.5	62.7	1.6	24.76	84.2	13.2
13/10/88	23.0	20.7	68.1	66.2	3.00	3.72	39.0	0	6.4	0.630	59.3	64.2	0.660	38	0.69	6.40	100.0	0.71	124.3	62.7	1.6	24.36	87.6	13.7
14/10/88	21.7	22.1	69.0	67.1	3.00	3.92	40.0	0	6.4	0.698	60.1	65.4	0.680	38	0.76	6.40	100.0	0.74	128.0	62.7	1.8	25.61	92.7	14.5
15/10/88	20.3	22.0	69.4	67.7	3.00	3.53	37.0	0	6.4	0.710	62.7	65.0	0.667	40	0.77	6.40	100.0	0.68	125.5	66.0	1.8	26.73	88.0	13.8
16/10/88	17.0	21.7	67.0	65.1	3.00	3.46	34.0	0	5.9	0.522	59.0	63.2	0.620	40	0.58	5.90	100.0	0.73	116.7	66.0	1.4	27.13	79.2	13.4
17/10/88	15.9	21.5	67.7	65.8	3.00	3.50	34.0	0	6.0	0.542	60.7	63.2	0.640	39	0.60	6.00	100.0	0.72	120.5	64.4	1.4	28.14	85.6	14.3

DATE	T_{amb} °C	T_{si} °C	T_{se} °C max	T_{se} °C avg	R_H days	V_{gm} m^3/h	T_{gm} °C	f_{O2} %	O_s kgO/h	M_{wc} kg/h	T_{ge} °C low	T_{ge} °C high	$\Delta T_{se}/\Delta t$ °C/h	I amps	M_{we} kg/h	O_c kgO/h	O_{eff}	Y_{CO2}	H_{net} MJ/h	H_{mi} MJ/h	H_{ve} MJ/h	H_{we} MJ/h	H_{bi} MJ/h	Y_H MJ/kgO
18/10/88	19.8	21.0	64.9	63.4	1.20	9.05	34.0	17	19.4	1.600	59.0	62.3	1.350	37	1.75	17.49	90.2	0.53	254.2	61.1	4.1	24.59	221.8	12.7
18/10/88	20.5	20.6	64.4	63.2	1.20	10.40	34.0	15	19.4	1.450	59.6	61.6	1.320	38	1.63	17.47	90.1	0.63	248.5	62.7	3.8	24.08	213.7	12.2
18/10/88	15.3	21.2	63.0	61.7	1.20	10.75	26.0	19	19.4	1.410	57.6	59.8	1.250	37	1.60	16.80	86.6	0.66	235.4	61.1	3.8	26.17	204.3	12.2
19/10/88	20.0	20.2	65.5	64.7	1.20	13.80	28.0	35	19.4	1.810	61.5	62.4	1.013	37	2.05	13.29	68.5	0.85	190.6	61.1	4.8	25.21	159.5	12.0
19/10/88	20.3	20.3	60.8	59.5	1.20	11.05	33.0	22	19.4	1.340	56.0	58.4	1.250	38	1.53	16.38	84.4	0.65	235.4	62.7	3.6	22.11	198.4	12.1
20/10/88	20.3	20.5	59.9	58.4	1.20	10.91	35.0	22	19.4	1.386	55.0	57.5	1.340	38	1.57	16.43	84.7	0.64	252.3	62.7	3.7	21.49	214.8	13.1
20/10/88	30.5	20.5	59.7	58.2	1.20	11.14	36.0	22	19.4	1.305	55.3	57.4	1.292	38	1.49	16.38	84.4	0.65	243.3	62.7	3.5	15.62	199.7	12.2
21/10/88	18.9	21.2	59.9	58.4	1.20	10.55	32.0	16	19.4	1.280	54.1	56.7	1.390	38	1.46	17.29	89.1	0.64	261.7	62.7	3.5	22.28	224.8	13.0
21/10/88	22.0	21.5	59.9	58.5	1.20	10.90	37.0	16	19.4	1.225	54.5	57.2	1.370	38	1.41	17.26	89.0	0.65	257.9	62.7	3.3	20.59	219.1	12.7
22/10/88	13.9	21.4	60.3	58.9	1.20	10.10	27.0	19	19.4	1.310	54.6	57.3	1.330	38	1.49	16.96	87.4	0.61	250.4	62.7	3.5	25.38	216.6	12.8
24/10/88	17.1	20.4	58.3	57.0	1.20	11.00	31.0	27	19.4	1.175	54.3	56.3	1.200	38	1.36	15.68	80.8	0.64	225.9	62.7	3.2	22.50	188.9	12.0
26/10/88	19.4	20.4	56.6	55.5	1.20	11.80	33.0	35	19.4	1.065	52.6	54.2	1.200	40	1.27	14.26	73.5	0.67	225.9	66.0	3.0	20.36	183.3	12.9
27/10/88	18.3	20.5	59.2	57.7	1.50	11.68	31.0	36	19.4	1.100	54.8	57.3	1.120	39	1.30	14.13	72.8	0.66	210.9	64.4	3.1	22.22	171.8	12.2
1/11/88	20.1	20.8	62.5	61.0	1.50	8.60	34.0	22	16.5	1.255	58.0	59.9	1.160	38	1.40	14.15	85.8	0.59	218.4	62.7	3.3	23.07	182.1	12.9
2/11/88	16.5	21.2	62.5	61.0	1.50	9.40	30.0	24	16.5	1.295	57.6	59.7	1.120	38	1.46	13.67	82.8	0.66	210.9	62.7	3.4	25.10	176.7	12.9
2/11/88	18.9	21.2	62.4	61.0	1.50	9.70	31.0	25	16.5	1.220	57.6	59.4	1.120	38	1.39	13.46	81.6	0.68	210.9	62.7	3.3	23.74	175.2	13.0
3/11/88	15.9	20.8	62.1	60.6	1.50	9.35	30.0	27	16.5	1.250	57.2	59.4	1.120	40	1.41	13.33	80.8	0.64	210.9	66.0	3.3	25.21	173.4	13.0

DATE	T_{amb} °C	T_{si} °C	T_{se} °C max	T_{se} °C avg	R_H days	V_{gm} m^3/h	T_{gm} °C	f_{O2} %	O_s kgO/h	M_{wc} kg/h	T_{ge} °C low	T_{ge} °C high	$\Delta T_{se}/\Delta t$ °C/h	I amps	M_{we} kg/h	O_c kgO/h	O_{eff}	Y_{CO2}	H_{net} MJ/h	H_{mi} MJ/h	H_{ve} MJ/h	H_{we} MJ/h	H_{bi} MJ/h	Y_H MJ/kgO
3/11/88	17.0	20.7	61.9	60.4	1.50	9.20	30.0	25	16.5	1.185	57.1	58.9	1.100	39	1.34	13.61	82.5	0.64	207.1	64.4	3.2	24.48	170.4	12.5
4/11/88	16.6	20.5	61.8	60.2	1.50	9.50	29.0	28	16.5	1.260	57.0	59.3	1.100	39	1.42	13.15	79.7	0.66	207.1	64.4	3.4	24.59	170.7	13.0
4/11/88	17.4	20.8	61.7	60.3	1.50	9.55	31.0	26	16.5	1.220	57.5	59.1	1.120	39	1.38	13.39	81.2	0.66	210.9	64.4	3.3	24.20	174.0	13.0
5/11/88	21.1	21.3	61.7	60.3	1.50	9.75	36.0	27	16.5	1.110	57.0	58.6	1.120	40	1.27	13.26	80.4	0.66	210.9	66.0	3.0	22.11	170.0	12.8
7/11/88	16.9	21.7	61.7	60.2	1.50	9.50	31.0	25	16.5	1.270	57.2	59.3	1.140	40	1.43	13.53	82.0	0.66	214.6	66.0	3.4	24.42	176.4	13.0
7/11/88	19.9	21.8	61.9	60.6	1.50	9.75	34.0	25	16.5	1.175	57.4	59.4	1.140	39	1.34	13.48	81.7	0.67	214.6	64.4	3.2	22.95	176.4	13.1
8/11/88	15.6	21.9	62.0	60.4	1.50	9.65	27.0	30	16.5	1.280	57.7	59.3	1.060	39	1.45	12.83	77.8	0.67	199.6	64.4	3.4	25.27	163.9	12.8
8/11/88	19.0	22.0	62.0	60.5	1.50	10.00	31.5	30	16.5	1.285	57.7	59.4	1.090	39	1.46	12.75	77.3	0.69	205.2	64.4	3.4	23.41	167.6	13.1
9/11/88	16.9	22.0	62.0	60.5	1.50	9.35	31.0	27	16.5	1.200	57.2	58.9	1.100	40	1.36	13.34	80.8	0.64	207.1	66.0	3.2	24.59	168.9	12.7
9/11/88	20.0	21.6	62.1	60.6	1.50	9.65	35.0	27	16.5	1.175	57.4	59.0	1.100	40	1.34	13.28	80.5	0.66	207.1	66.0	3.2	22.90	167.2	12.6
10/11/88	19.0	21.7	62.4	60.9	1.50	9.25	34.0	26	16.5	1.155	57.1	59.0	1.100	39	1.31	13.52	81.9	0.63	207.1	64.4	3.1	23.63	169.4	12.5
10/11/88	23.3	21.9	62.9	61.2	1.50	12.95	37.5	33	22.0	1.675	57.5	59.4	1.280	39	1.89	16.76	76.2	0.63	241.0	64.4	4.5	21.38	202.5	12.1
10/11/88	26.1	21.8	63.3	61.7	1.50	12.77	37.0	33	22.0	1.650	57.7	59.6	1.260	39	1.87	16.82	76.5	0.62	237.2	64.4	4.4	20.08	197.3	11.7
11/11/88	21.9	21.6	64.9	63.2	1.50	12.60	35.0	35	22.0	1.880	61.7	62.1	1.280	39	2.09	16.55	75.2	0.61	241.0	64.4	4.9	23.29	204.8	12.4
11/11/88	28.9	21.8	65.1	63.5	1.50	13.10	41.0	35	22.0	1.975	60.0	62.2	1.240	39	2.19	16.44	74.7	0.63	233.5	64.4	5.2	19.51	193.8	11.8
12/11/88	32.5	21.7	66.7	65.0	1.50	12.70	38.0	31	22.0	2.350	61.6	63.8	1.320	40	2.56	17.18	78.1	0.62	248.5	66.0	6.0	18.33	206.8	12.0
14/11/88	18.5	21.6	67.3	65.4	1.50	12.05	31.0	26	22.0	2.085	62.2	64.3	1.340	38	2.29	18.08	82.2	0.62	252.3	62.7	5.4	26.45	221.5	12.3

DATE	T_{amb} °C	T_{si} °C	T_{se} °C max	T_{se} °C avg	R_H days	V_{gm} m^3/h	T_{gm} °C	f_{O2} %	O_s kgO/h	M_{wc} kg/h	T_{ge} °C low	T_{ge} °C high	$\Delta T_{se}/\Delta t$ °C/h	I amps	M_{we} kg/h	O_c kgO/h	O_{eff}	Y_{CO2}	H_{net} MJ/h	H_{mi} MJ/h	H_{ve} MJ/h	H_{we} MJ/h	H_{bi} MJ/h	Y_H MJ/kgO
17/11/88	30.0	22.7	65.6	64.3	1.50	9.00	40.0	16	16.1	1.475	60.7	62.8	1.180	39	1.63	14.35	89.1	0.64	222.2	64.4	3.8	19.35	181.0	12.6
17/11/88	34.7	22.8	64.0	62.6	1.20	9.25	49.0	15	16.1	1.250	58.8	61.7	1.200	39	1.40	14.46	89.8	0.64	225.9	64.4	3.3	15.74	180.5	12.5
18/11/88	18.2	23.2	61.0	60.0	1.20	8.45	29.0	16	16.1	0.985	56.1	58.8	1.150	39	1.13	14.40	89.4	0.62	216.5	64.4	2.7	23.58	178.4	12.4
18/11/88	21.2	23.2	60.9	59.5	1.20	8.45	35.0	16	16.1	1.000	56.0	58.3	1.150	39	1.14	14.43	89.6	0.61	216.5	64.4	2.7	21.60	176.4	12.2
19/11/88	21.7	23.1	59.4	58.0	1.20	8.25	35.0	15	16.1	0.859	54.7	57.5	1.180	40	1.00	14.57	90.5	0.59	222.2	66.0	2.4	20.47	179.1	12.3
20/11/88	25.1	23.2	58.9	57.5	1.20	8.45	40.0	12	16.2	0.740	54.2	56.7	1.225	40	0.88	14.97	92.4	0.60	230.6	66.0	2.1	18.27	185.0	12.4
21/11/88	24.0	23.3	58.5	57.3	1.20	8.65	37.0	16	16.2	0.880	53.9	56.4	1.200	40	1.03	14.50	89.5	0.62	225.9	66.0	2.4	18.78	181.1	12.5
21/11/88	27.1	23.3	58.5	57.3	1.20	8.60	41.0	16	16.3	0.790	53.9	56.3	1.200	40	0.93	14.63	89.8	0.60	225.9	66.0	2.2	17.03	179.1	12.2
22/11/88	18.5	23.5	58.2	57.0	1.20	8.55	31.0	20	16.0	0.830	54.1	56.2	1.125	39	0.98	13.86	86.6	0.62	211.8	64.4	2.3	21.71	171.4	12.4
23/11/88	19.3	23.3	57.7	56.6	1.20	8.50	31.0	18	16.3	0.810	53.2	55.6	1.175	39	0.96	14.38	88.2	0.61	221.2	64.4	2.3	21.04	180.1	12.5
24/11/88	18.2	23.1	58.2	56.8	1.20	8.30	31.0	13	16.3	0.870	53.5	56.2	1.250	39	1.01	14.95	91.7	0.60	235.4	64.4	2.4	21.77	195.2	13.1
25/11/88	21.9	23.4	59.2	57.8	1.20	9.05	33.0	10	16.3	0.950	54.5	57.1	1.275	39	1.10	15.17	93.1	0.67	240.1	64.4	2.6	20.25	198.6	13.1
27/11/88	26.7	23.5	59.4	58.0	1.20	8.75	38.0	12	16.3	0.870	54.5	56.9	1.225	40	1.02	15.01	92.1	0.63	230.6	66.0	2.4	17.65	184.7	12.3
28/11/88	19.3	23.2	59.2	58.0	1.20	8.35	33.0	12	16.3	0.850	54.0	56.4	1.225	40	0.99	15.05	92.3	0.61	230.6	66.0	2.4	21.83	188.8	12.5

APPENDIX 4B

AEROBIC REACTOR TS, VS AND COD CONCENTRATIONS

DATE	TS_{in} g/ℓ	TS_{out} g/ℓ	VS_{in} g/ℓ	VS_{out} g/ℓ	COD_{in} g/ℓ	COD_{out} g/ℓ	COD_{in}/ VS_{in}	COD_{out}/ VS_{out}
30/11/87	32.9	35.8	27.5	29.4				
1/12/87	34.2	32.8	28.7	27.2				
7/12/87	37.8	33.9	31.9	27.7				
8/12/87	32.8	27.2	27.0	22.5				
9/12/87	37.8	34.4	31.5	28.2				
10/12/87	41.9	35.7	34.3	29.2				
11/12/87	35.4	35.5	28.7	28.9				
14/12/87	40.3	34.0	32.7	27.3				
15/12/87	35.0	35.1	28.8	28.5				
16/12/87	33.9	30.7	28.4	25.4				
17/12/87	37.2	31.5	31.1	26.1				
4/1/88	34.6	28.9	29.1	24.4				
5/1/88	33.2	34.9	27.7	28.8				
6/1/88	30.5	39.1	25.2	31.9				
7/1/88	36.4	33.0	30.0	27.0				
8/1/88	27.8	30.7	23.3	25.3				
9/1/88	32.3	38.9	26.7	31.8				
8/2/88	28.9	28.1	23.4	23.2				
23/2/88	31.9	32.5	26.2	26.6				
24/2/88	36.4	34.5	29.4	27.8				
25/2/88	41.6	36.3	33.8	29.4				
29/2/88	27.9	22.3	22.3	28.1				
1/3/88	39.8	34.0	32.1	27.3				
2/3/88	39.0	36.3	31.4	29.0				

DATE	TS_{in} g/ℓ	TS_{out} g/ℓ	VS_{in} g/ℓ	VS_{out} g/ℓ	COD_{in} g/ℓ	COD_{out} g/ℓ	COD_{in}/VS_{in}	COD_{out}/VS_{out}
3/3/88	37.5	39.4	30.1	31.4				
4/3/88	36.3	37.8	29.2	30.2				
5/3/88	35.4	37.8	28.5	29.9				
6/3/88	35.4	35.8	28.9	28.6				
7/3/88	32.7	33.3	26.6	26.6				
8/3/88	34.8	37.3	28.2	29.6				
9/3/88	36.1	36.8	29.2	29.1				
10/3/88	37.3	37.9	30.5	30.4				
11/3/88	35.5	38.0	29.3	30.9				
12/3/88	37.5	37.8	31.0	30.8				
13/3/88	35.2	36.0	29.2	29.4				
14/3/88	33.1	35.5	27.4	28.7				
15/3/88	34.1	34.7	28.2	27.9				
16/3/88	38.2	34.7	31.8	28.2				
20/3/88	32.8	36.3	27.4	29.6				
21/3/88	37.8	36.8	31.3	29.9				
23/3/88	37.2	38.7	30.9	31.5				
24/3/88	37.2	39.3	30.6	32.1				
27/3/88	36.6	41.2	29.9	33.5				
28/3/88	40.2	41.4	32.7	33.1				
29/3/88	38.1	41.2	30.9	33.2				
30/3/88	39.4	42.1	32.5	33.7				
31/3/88	40.9	43.0	33.4	35.0				
1/4/88	39.5	44.2	32.5	36.0				

DATE	TS_{in} g/ℓ	TS_{out} g/ℓ	VS_{in} g/ℓ	VS_{out} g/ℓ	COD_{in} g/ℓ	COD_{out} g/ℓ	COD_{in}/VS_{in}	COD_{out}/VS_{out}
2/4/88	37.4	42.0	30.9	34.6				
3/4/88	36.3	43.0	29.7	35.2				
6/4/88	39.4	39.5	32.2	31.9				
7/4/88	37.3	32.9	37.3	32.9				
9/4/88	35.6	35.3	29.5	29.2				
10/4/88	32.8	38.2	27.1	31.2				
15/4/88	34.3	41.8	28.4	34.3				
16/4/88	35.0	39.5	29.1	32.3				
17/4/88	32.2	36.1	26.7	29.8				
18/4/88	33.5	35.2	27.9	28.9				
19/4/88	30.7	34.0	25.8	28.2	41.7	40.1	1.616	1.422
21/4/88	31.7	32.4	26.5	26.6	43.3	37.6	1.634	1.414
23/4/88	29.9	30.3	24.8	25.0	40.1	35.8	1.617	1.432
25/4/88	35.2	32.1	29.3	26.2				
26/4/88					49.9	40.4		
27/4/88	37.6	35.6	30.6	29.0	51.3	42.5	1.676	1.466
28/4/88	31.2	31.9	25.6	25.9	44.6	40.1	1.742	1.548
30/4/88	34.4	35.3	28.1	28.0				
1/5/88	34.4	33.8	28.0	27.4				
4/5/88	36.5	36.9	29.7	29.9				
6/5/88	36.6	37.8	30.6	30.9				
7/5/88	33.5	39.8	27.8	32.7				
8/5/88	35.3	39.5	29.4	32.3				
9/5/88	34.3	38.6	28.3	31.9				

DATE	TS_{in} g/ℓ	TS_{out} g/ℓ	VS_{in} g/ℓ	VS_{out} g/ℓ	COD_{in} g/ℓ	COD_{out} g/ℓ	COD_{in}/VS_{in}	COD_{out}/VS_{out}
10/5/88	38.7	41.0	31.5	33.4				
11/5/88	37.5	41.3	30.7	33.6				
12/5/88	35.5	41.5	29.1	34.0				
22/5/88	36.6	38.5	30.8	32.0				
29/5/88	35.6	37.6	29.3	30.5				
30/5/88	37.4	38.3	30.7	31.2				
31/5/88	38.2	35.2	31.4	28.7				
1/6/88	38.0	37.1	30.7	30.2				
2/6/88	35.3	34.6	28.6	28.0				
3/6/88	36.1	34.6	29.9	28.0				
4/6/88	38.2	37.0	31.2	29.9				
5/6/88	36.9	36.5	30.0	29.5				
6/6/88	38.1	34.1	30.9	27.4	42.5	34.3	1.375	1.252
7/6/88	36.9	34.3	30.3	27.6				
8/6/88	35.6	35.2	29.1	28.3	51.1	40.9	1.756	1.445
9/6/88	35.4	38.0	28.4	30.6	49.5	42.9	1.743	1.402
10/6/88	39.9	36.2	30.4	27.8				
11/6/88	42.8	37.9	34.1	29.2				
12/6/88	40.2	36.5	32.0	28.4	54.6	40.8	1.706	1.437
16/6/88	38.3	37.3	31.1	29.5	50.6	46.1	1.627	1.563
17/6/88	35.2	38.9	28.8	31.5				
18/6/88	36.8	34.4	27.9	30.4				
19/6/88	34.8	36.2	28.7	29.2	46.4	41.6	1.617	1.425
21/6/88	37.7	36.1	30.7	29.0				

DATE	TS_{in} g/ℓ	TS_{out} g/ℓ	VS_{in} g/ℓ	VS_{out} g/ℓ	COD_{in} g/ℓ	COD_{out} g/ℓ	COD_{in}/VS_{in}	COD_{out}/VS_{out}
22/6/88	37.2	34.8	30.3	27.8				
23/6/88	35.0	34.6	28.5	27.9	54.4	43.4	1.909	1.556
24/6/88	36.6	33.0	30.0	26.5				
25/6/88	34.2	32.4	28.2	26.3				
26/6/88	34.3	31.5	28.5	25.8				
27/6/88	36.3	34.2	30.1	27.9	47.9	44.2	1.591	1.584
28/6/88	37.1	31.7	30.7	26.0	56.0	41.8	1.824	1.608
29/6/88	36.8	35.8	30.3	29.3	54.4	48.7	1.795	1.662
30/6/88	34.3	33.5	28.2	27.4				
3/7/88	35.9	33.1	29.7	26.9	50.4	40.0	1.697	1.487
4/7/88	35.7	30.6	29.5	24.9	50.9	36.6	1.725	1.470
5/7/88	38.6	34.2	32.0	27.8	52.3	40.5	1.634	1.457
6/7/88	39.5	34.7	32.8	28.4	55.7	42.8	1.698	1.507
7/7/88	38.7	35.7	31.9	29.3				
8/7/88	36.9	34.9	30.2	28.3				
10/7/88	57.9	44.5	35.6	30.0	57.7	45.7	1.621	1.523
11/7/88	42.1	43.4	30.5	29.1	50.6	43.3	1.659	1.488
12/7/88	41.1	39.7	31.3	27.7				
15/7/88	41.3	44.2	32.5	33.2				
17/7/88	36.9	38.0	30.0	30.0	51.6	43.3	1.720	1.443
18/7/88	43.4	42.9	34.6	34.1	57.7	49.9	1.668	1.463
19/7/88	42.4	42.7	34.0	33.9	57.2	54.3	1.682	1.602
20/7/88	42.9	41.1	34.1	32.7	59.2	49.8	1.736	1.523
21/7/88	52.5	42.4	37.9	31.8	63.7	52.3	1.681	1.645

DATE	TS_{in} g/ℓ	TS_{out} g/ℓ	VS_{in} g/ℓ	VS_{out} g/ℓ	COD_{in} g/ℓ	COD_{out} g/ℓ	COD_{in}/VS_{in}	COD_{out}/VS_{out}
22/7/88	37.9	42.6	27.9	31.0				
31/7/88	35.4	38.9	28.7	31.0	48.5	48.1	1.690	1.552
1/8/88	35.6	38.3	28.8	30.7	48.5	47.3	1.684	1.541
2/8/88	37.7	35.6	30.5	28.6	53.8	45.5	1.764	1.591
3/8/88	37.1	38.7	29.8	31.0				
4/8/88	38.9	41.7	31.0	33.2	55.5	51.5	1.790	1.551
6/8/88	39.7	40.7	31.5	32.1				
7/8/88	38.6	40.3	30.9	31.9				
22/8/88	40.3	36.4	33.4	29.7	55.0	43.4	1.647	1.461
23/8/88	37.1	36.8	30.9	30.3	50.6	44.2	1.638	1.459
24/8/88	38.5	37.0	31.9	30.5	55.6	45.0	1.743	1.475
25/8/88	38.7	39.5	31.5	32.2	55.9	49.0	1.775	1.522
26/8/88	37.4	36.8	27.8	30.3				
28/8/88	35.5	41.4	27.4	31.1	43.0	44.5	1.569	1.431
29/8/88	42.9	45.1	32.7	34.2	51.4	49.1	1.572	1.436
31/8/88	43.0	44.0	32.2	33.2	52.7	49.3	1.637	1.485
1/9/88	41.8	43.9	31.8	33.0	53.6	54.2	1.686	1.642
2/9/88	36.3	45.8	27.7	34.6				
5/9/88	34.8	36.3	26.9	28.2	44.1	42.4	1.639	1.504
6/9/88	47.0	37.9	36.0	29.1	58.4	42.4	1.622	1.457
11/9/88	32.2	36.8	25.8	28.4	41.9	42.3	1.624	1.489
12/9/88	34.1	32.3	26.9	25.7	45.1	39.2	1.677	1.525
13/9/88	37.7	35.2	30.3	28.0	52.2	45.3	1.723	1.618
14/9/88	34.2	37.1	27.4	29.4	46.8	47.9	1.708	1.629

DATE	TS_{in} g/ℓ	TS_{out} g/ℓ	VS_{in} g/ℓ	VS_{out} g/ℓ	COD_{in} g/ℓ	COD_{out} g/ℓ	COD_{in}/VS_{in}	COD_{out}/VS_{out}
15/9/88	36.2	36.7	28.9	28.9	52.1	46.6	1.803	1.612
16/9/88	38.5	36.0	30.2	28.4				
18/9/88	36.3	35.6	28.6	27.8	49.5	42.9	1.731	1.543
20/9/88	41.7	36.7	33.3	28.8	56.3	45.4	1.691	1.576
21/9/88	37.5	36.8	29.6	28.8	51.7	43.5	1.747	1.510
5/10/88	50.4	36.4	39.8	29.0	69.5	41.2	1.746	1.421
6/10/88	55.8	37.2	39.1	28.9	68.4	44.3	1.749	1.533
10/10/88	40.3	43.1	31.2	32.0	54.7	52.7	1.753	1.647
11/10/88	41.8	42.5	32.1	31.8	56.2	55.3	1.751	1.739
12/10/88	42.8	41.9	33.7	31.7	57.4	49.0	1.703	1.546
13/10/88	41.2	39.9	32.1	30.3	57.5	46.4	1.791	1.531
14/10/88	48.4	41.6	37.4	31.8				
15/10/88	42.1	41.9	32.4	32.3	56.9	51.6	1.756	1.598
16/10/88	38.6	41.8	30.4	32.1	53.4	56.7	1.757	1.766
17/10/88	40.1	40.1	31.7	30.7	55.2	50.0	1.741	1.629
19/10/88	36.5	38.4	28.8	29.9	50.6	48.1	1.757	1.609
20/10/88	44.8	39.9	34.9	30.7				
23/10/88	39.8	48.3	31.1	37.2	55.5	60.3	1.785	1.621
26/10/88					62.4	55.3		
27/10/88	39.1	43.1	30.7	33.7	58.2	54.8	1.896	1.626
31/10/88	42.9	42.8	33.9	33.6	56.5	52.0	1.667	1.548
1/11/88	42.1	37.7	33.7	29.6	59.3	49.5	1.760	1.672
2/11/88	42.2	42.8	34.0	33.8	59.8	55.9	1.759	1.654
3/11/88	39.8	37.3	31.4	30.0	53.9	51.6	1.717	1.720

DATE	TS_{in} g/ℓ	TS_{out} g/ℓ	VS_{in} g/ℓ	VS_{out} g/ℓ	COD_{in} g/ℓ	COD_{out} g/ℓ	COD_{in}/VS_{in}	COD_{out}/VS_{out}
4/11/88	43.6	40.3	35.2	32.0				
6/11/88	36.0	42.8	28.6	33.8	52.7	52.0	1.843	1.538
7/11/88	36.4	38.4	29.5	30.4	51.0	48.4	1.729	1.592
8/11/88	41.8	39.4	33.9	31.6	55.1	47.7	1.625	1.509
9/11/88	46.3	40.6	36.6	32.2	67.5	53.5	1.844	1.661
10/11/88	42.1	34.5	33.2	24.9	61.4	58.6	1.849	2.353
12/11/88	44.0	35.7	44.1	35.0				
14/11/88	31.0	36.1	24.9	28.7				
16/11/88	34.8	37.2	27.9	29.6	52.0	46.8	1.864	1.581
18/11/88	37.3	36.3	30.0	28.9	52.3	45.7	1.743	1.581
19/11/88	33.6	35.2	26.8	28.2	47.7	46.9	1.780	1.663
21/11/88	37.7	35.9	29.8	28.3	59.5	44.7	1.997	1.580
22/11/88	35.2	39.3	27.8	30.8	50.2	51.4	1.806	1.669
24/11/88	28.8	40.8	21.9	31.4	44.3	58.4	2.023	1.860
22/11/88	35.2	39.3	27.8	30.8	50	51	1.806	1.669
24/11/88	28.8	40.8	21.9	31.4	44	58	2.023	1.860

APPENDIX 4C

TS, VS AND COD CONCENTRATIONS MEASURED ON THE AEROBIC REACTOR BY THE BELVILLE LABORATORY OF THE DIVISION OF WATER TECHNOLOGY OF THE CSIR.

DATE	TS_{in} g/ℓ	TS_{out} g/ℓ	VS_{in} g/ℓ	VS_{out} g/ℓ	COD_{in} g/ℓ	COD_{out} g/ℓ	COD_{in}/VS_{in}	COD_{out}/VS_{out}
29/2/88	35.7	30.5	28.7	24.8	45.5	39.5	1.585	1.593
7/3/88	28.7	31.4	23.4	25.2	39.8	38.1	1.701	1.512
15/3/88	42.1	31.8	33.3	25.9	57.1	40.8	1.715	1.575
22/3/88	36.9	31.9	29.8	25.7	49.8	38.5	1.671	1.498
28/3/88	43.5	33.7	34.5	26.8	51.6	39.6	1.496	1.478
11/4/88	33.9	31.4	28.3	26.0	49.0	38.0	1.731	1.462
19/4/88	30.4	31.0	25.7	26.3	41.6	40.0	1.619	1.521
25/4/88	30.8	31.5	25.3	25.9	34.1	37.2	1.348	1.436
2/5/88	30.4	32.7	24.9	26.6	40.0	36.3	1.606	1.365
9/5/88	33.2	32.3	28.1	26.8	54.9	47.0	1.954	1.754
17/5/88	34.3	34.6	28.4	28.9	49.1	46.6	1.729	1.612
7/6/88	38.2	33.5	31.0	26.4	49.5	39.7	1.597	1.504
15/6/88	39.4	36.9	31.6	28.9	55.4	49.5	1.753	1.713
20/6/88	38.5	32.1	32.1	26.1	51.4	36.0	1.601	1.379
28/6/88	36.0	34.0	30.4	28.1	53.0	42.0	1.743	1.495
5/7/88	36.1	31.6	29.8	25.8	48.0	38.0	1.611	1.473

DATE	TS_{in} g/ℓ	TS_{out} g/ℓ	VS_{in} g/ℓ	VS_{out} g/ℓ	COD_{in} g/ℓ	COD_{out} g/ℓ	COD_{in}/VS_{in}	COD_{out}/VS_{out}
12/7/88	43.5	43.6	31.7	29.7	48.4	42.3	1.527	1.424
19/7/88	43.8	42.0	34.8	33.0	56.5	48.5	1.624	1.470
3/8/88	35.6	38.4	29.2	31.2	47.7	45.6	1.634	1.462
17/8/88	38.1	31.7	33.0	26.2	65.5	45.7	1.985	1.744
25/8/88	39.2	35.9	32.3	29.6	52.2	50.2	1.616	1.696
6/9/88	34.6	35.2	26.5	27.6	49.2	39.5	1.857	1.431
13/9/88	35.7	33.8	27.9	26.2	45.9	42.0	1.645	1.603
12/10/88	41.4	43.9	31.8	32.7	56.0	49.2	1.761	1.505
18/10/88	41.2	40.3	33.1	31.2	72.7	49.8	2.196	1.596
1/11/88	42.9	42.8	33.6	33.5	57.6	68.8	1.714	2.054
10/11/88	35.4	36.9	28.1	28.5	50.4	44.5	1.794	1.561
24/11/88	38.1	36.9	30.5	29.5	49.8	49.4	1.633	1.675
31/1/89	29.9	29.5	24.2	24.1	47.6	46.5	1.967	1.929
16/2/89	38.2	27.7	31.4	22.6	42.7	31.9	1.360	1.412
2/3/89	43.4	31.3	32.3	25.5	48.5	37.0	1.502	1.451

APPENDIX 4D.

STATISTICAL ANALYSES OF THE SPECIFIC HEAT YIELD AND OF CHANGES IN VS & COD CONCENTRATIONS AND IN THE COD/VS RATIO.

In this Appendix, pairs of means belonging to various sets of data viz specific heat yields (Y_H), volatile solids concentrations (VS), chemical oxygen demand concentrations (COD) and COD/VS ratios, are compared with the aid of statistical F and t tests at the 95% confidence level. The t test determines whether there is a significant difference between 2 means and hence whether these means are derived from different data populations.

For the sets of data examined in this Appendix, the objectives of the test differ slightly. In the case of the specific heat yield data, the t test is used to determine whether a significant inverse proportionality exists between the specific heat yield and the rate of oxygen consumption by the reactor sludge. For the VS and COD analyses and the COD/VS ratios, the test ascertains whether their mean values differ significantly for the feed and effluent sludge analyses, and hence whether their concentrations in the sludge have been reduced by treatment in the aerobic reactor. Also, COD and VS analyses performed by the DWT and by the writer at the Milnerton plant on the same samples are compared using the t test in order to cross check the analytical procedures.

The following procedure is used in each case:

(1) The data is checked, using a probability plot, to see if it obeys a normal distribution. If the plotted points lie on a straight line then normality is indicated, and a F and a t–test may be constructed on the data.

(2) A 2–sided F–test at the 95% confidence level is performed on each pair of means to determine if their variances are equal. For this to be so, the ratio of their sample variances (F_{obs}) must be less than a calculated statistic, $F_{n1-1;n2-2}$, which is obtained from statistical tables. Only if equality between the variances is shown, may the t–test on each pair of means proceed.

(3) In the t–test, a pooled variance is used to calculate an observed statistic, t_{obs}. The means are considered to differ significantly if this statistic lies outside of the interval defined by a t probability function, such that $-t_{n1+n2-2} < t_{obs} > t_{n1+n2-2}$.

The equations used in the above tests are listed below:

The sample mean:

$$\overline{X} = \frac{1}{n} \sum_{i-1}^{n} X_i$$

The sample variance:

$$s^2 = \frac{1}{n-1} \sum_{i-1}^{n} (X_i - \overline{X})$$

Observed F statistic:

$$F_{obs} = \frac{s_1^2}{s_2^2} \quad \text{where } s_1^2 > s_2^2$$

Pooled variance:

$$s = \sqrt{\frac{(n_1 - 1)s_1^2 + (n_2 - 1)s_2^2}{n_1 + n_2 - 2}}$$

Observed t statistic:

$$t_{obs} = \frac{\overline{X}_1 - \overline{X}_2}{s\sqrt{\frac{1}{n_1} + \frac{1}{n_2}}}$$

4D.1 The specific heat yield, Y_H

In Chapter 4, Section 4.4.3.3, the set of 116 measured specific heat yield values were ordered in terms of increasing oxygen consumption rate (O_c) and divided into 3 groups falling into the following ranges of O_c: 5.9 – 8.1 kgO/h, 12.7 – 15.9 kgO/h, 16.0 – 20.3 kgO/h. The mean specific heat yield measured for each of these groups is given in Table 4D.1 and can be seen to decrease with increasing O_c. A probability plot of each of the 3 ranges of data indicated that data in each range was normally distributed and hence F and t tests may be performed.

The test results for groups 5.9 – 8.1 kgO/h and 12.7 – 15.9 kgO/h and groups 12.7 – 15.9 kgO/h and 16.0 – 20.3 kgO/h are given in Table 4D.1. For both pairs of groups, the calculated values of t_{obs} lie outside the range of values for $t_{n1+n2-2}$ obtained from statistical tables. Hence it may be concluded that, at a 95% level of confidence, the means of the 3 groups of data are derived from different data populations and that an inverse proportionality between the specific heat yield and the oxygen consumption rate exists.

TABLE 4D.1: Statistical analysis of specific heat yield data.

	O_c 5.9 - 8.1 kgO/h	O_c 12.7 - 15.9 kgO/h	O_c 12.7 - 15.9 kgO/h	O_c 16.0 - 20.3 kgO/h
MEAN Y_H	13.81	12.75	12.75	12.5
s^2	0.2724	0.1316	0.1316	0.230
# of data	13	56	56	47
F_{obs}	2.07		1.75	
$F_{n1-1;n2-1}$	2.19		1.78	
s	0.396		0.42	
t_{obs}	-8.69		-3.01	
$t_{n1+n2-2}$	-1.991 to 1.991		-1.984 to 1.984	

4D.2 Milnerton VS and COD analyses of feed and reactor effluent sludge

This test was to determine if the VS and COD concentrations in the feed sludge and the COD/VS ratio were changed by treatment in the aerobic reactor. Table 4D.2 lists the statistics relevant to the 3 pairs of data. The t–tests were performed at the 5% confidence interval, single sided tests being used for the VS and COD data, and double sided tests for the COD/VS ratios.

The observed F values for each pair lie inside the confidence interval defined by $F_{n1-1;n2-1}$ and thus equality of variance may be assumed. The observed t values for the VS concentrations lie inside the defined confidence interval, while those for the COD concentrations and the COD/VS ratios fall outside. Hence, of the 3 parameters, only the VS concentration in the feed sludge does not appear to have been significantly changed in the aerobic reactor.

TABLE 4D.2: Statistical data for Milnerton analyses.

	VS_{in}	VS_{out}	COD_{in}	COD_{out}	COD/VS OUT	COD/VS IN
MEAN	30.28	29.89	52.99	46.69	1.721	1.548
s^2	9.24	6.71	35.39	30.86	0.009	0.010
# of data	180	180	75	75	75	74
F_{obs}	1.337		1.62		1.62	
$F_{n1-1;n2-1}$	1.44		1.15		1.11	
s	2.792		5.755		0.0975	
t_{obs}	1.325		6.698		10.829	
$t_{n1+n2-2}$	1.645		1.645		1.96	

4D.3 DWT analyses of sludge VS and COD concentrations

In this test the VS and COD concentrations of 31 sludge samples analysed by the CSIR (see Appendix 4C) are examined to ascertain whether they, and their ratio, were changed by the action of the aerobic reactor. Table 4D.3 lists the statistics relevant to the 3 pairs of data. The t–tests were performed at the 95% confidence interval, single sided tests being used for the VS and COD data, and double sided tests for the COD/VS ratios.

The observed F values for each pair lie inside the confidence interval defined by $F_{n1-1;n2-1}$ and thus equality of variance may be assumed. The observed t values for both the VS and COD concentrations and their ratio lie outside the defined confidence interval. Hence, all 3 parameters were significantly reduced in the aerobic reactor.

TABLE 4D.3: Statistical data for CSIR analyses.

	VS_{in}	VS_{out}	COD_{in}	COD_{out}	COD/VS OUT	COD/VS IN
MEAN	29.86	27.59	50.34	43.15	1.686	1.561
s^2	9.59	7.15	55.02	47.22	0.031	0.024
Nº of data	31	31	31	31	31	31
F_{obs}	1.341		1.165		1.284	
$F_{n1-1;n2-1}$	2.07		2.07		2.07	
s	2.893		7.150		0.167	
t_{obs}	3.04		3.89		2.90	
$t_{n1+n2-2}$	1.671		1.671		2.000	

4D.4 Comparison between duplicate sludge TS, VS and COD analyses performed by the DWT (CSIR) and the writer at Milnerton

Twelve samples of sludge were tested for TS, VS and COD by both the writer at Milnerton and the DWT at the CSIR in order to cross-check the analytical techniques (see Table 4.7). Table 4D.4 lists the statistics relevant to the 3 pairs of data. Single sided t-tests at the 95% confidence interval were used in all cases.

The F-test indicates variance equality, allowing the t-test to proceed. All of the observed t values fall inside the confidence interval defined by $t_{n1+n2-2}$. Hence the influent and effluent analyses for TS, VS and COD performed at Milnerton and the DWT show good agreement at the 95% confidence level.

TABLE 4D.4: Statistical data for duplicate Milnerton and CSIR analyses.

	TS_{in} CSIR MILN	TS_{out} CSIR MILN	VS_{in} CSIR MILN	VS_{out} CSIR MILN	COD_{out} CSIR MILN	COD_{in} CSIR MILN
MEAN	39.0 38.5	38.2 38.1	31.0 30.5	29.8 29.6	53.1 52.2	47.1 45.6
s^2	11.7 10.5	18.0 19.2	5.78 5.64	6.64 8.04	52.8 25.6	65.2 24.4
№ of data	12 12	12 12	12 12	12 12	12 12	12 12
F_{obs}	1.12	1.07	1.02	1.21	2.06	2.57
$F_{n1-1;n2-1}$	3.48	3.48	3.48	3.48	3.48	3.48
s	3.33	4.31	2.49	2.71	6.26	6.69
t_{obs}	0.368	0.057	0.490	0.181	0.352	0.549
$t_{n1+n2-2}$	2.074	2.074	2.074	2.074	2.074	2.074

APPENDIX 4E

THE EFFECT OF CARBON DIOXIDE DISSOLUTION INTO THE SLUDGE LIQUOR ON THE VENT GAS RESPIRATION QUOTIENT (Y_{CO_2})

During each of the 116 specific heat yield tests, vent gas from the reactor was collected and cooled to ± 20°C and the volume and composition of this partially dried gas was measured. From these measurements it was possible to calculate, amongst other parameters, the ratio of the molar carbon dioxide flow rate in the vent gas to the molar rate of oxygen utilized by the reactor sludge. Details of this calculation are listed in Appendix 3B. This calculated ratio was assumed to be equal to the respiration quotient (Y_{CO2}) of the reactor sludge i.e. the moles of carbon dioxide produced by the reactor sludge for each mole of oxygen that it utilized, and in this Appendix this calculated ratio will be termed the apparent respiration quotient Y'_{CO2}.

Theoretically, Y_{CO2} values of 0.95 and 1.0 may be calculated by considering the oxidation of typical formulations of sludge elemental composition reported in the literature (see Page 4.32) and it was expected that a value of 1.0 would be measured at Milnerton. However, the 116 specific heat yield tests gave an average Y'_{CO2} of 0.66 ie the reactor sludge generated only 0.66 moles of CO_2 for each mole of O_2 that it consumed. The lower than expected measured yield of CO_2 could be due to (1) the composition of the oxidized substrate differing from the typical formulations of sludge elemental composition or (2) some of the CO_2 remaining dissolved in the sludge liquor as a result of the increase in $H_2CO_3{}^*$ alkalinity through the ammonification of proteins (NH_3) to ammonium (NH_4^+). The aim of this Appendix is to examine the second of these explanations ie would the average measured value of Y'_{CO2} increase from 0.66 to about 1.0 if the measured CO_2 production from the reactor is increased by the amount of CO_2 calculated to remain dissolved in the reactor sludge liquor.

In order to determine how much CO_2 remains dissolved in the reactor sludge liquor, the increase in the carbonic species molar concentration (C_T) across the reactor is calculated. The molar flow rate of CO_2 calculated to be lost from the reactor via the sludge liquid is added to the average measured molar flow rate of CO_2 in the vent gas to give the total molar flow rate of CO_2 produced biologically. This total is divided by the average measured molar oxygen utilization in the reactor to yield a revised respiration quotient equal to the biological respiration quotient Y_{CO2}.

CALCULATION PROCEDURE

The following calculation procedures are employed in examining the effect of dissolved CO_2 on Y'_{CO2}:

(1) The *in situ* pH of the reactor sludge liquor is calculated from the measured CO_2 partial pressure in the sludge which was assumed to be equal to that measured in the head space vent gas and from the estimated reactor sludge liquor $H_2CO_3^*$ alkalinity. ($H_2CO_3^*$ alkalinity is the alkalinity of the carbonate species with respect to the carbonic acid reference species or equivalent solution). The $H_2CO_3^*$ alkalinity was estimated from the routine alkalinity titrations on the reactor sludge (listed in Appendix 5B, Table 5B.2(b)) to pH 5.75 and pH 4.3, taking due account of the effect of SCFA's; the effect of NH_4^+ on the $H_2CO_3^*$ alkalinity could be ignored because all the sludge liquor samples had pH values less than 7.5. The estimated $H_2CO_3^*$ alkalinity lay between 0.008 and 0.016 mol/ℓ, and these two values represented the extremes of the range of $H_2CO_3^*$ alk values estimated for the sludge liquor. It is necessary to calculate the *in situ* pH of the reactor sludge liquor because this value was not measured; reactor sludge liquor pH values quoted in this thesis [see Table 5B.2(b)] were measured on a composite sample which, over a 12 to 24 h period, had cooled to ambient temperature and which had equilibrated with atmospheric air probably leading to considerable CO_2 loss. The following equation is employed to calculate the *in situ* pH from estimated $H_2CO_3^*$ alk and p_{CO2}

$$H_2CO_3^*\ \text{alk} = p_{CO2}\cdot K_{CO2}\cdot 10^{pH-pK_1}.(2\cdot 10^{pH-pK_2}+1) + 10^{pH-pK_w} - 10^{-pH}/f_m \qquad (4E.1)$$

where,

$H_2CO_3^*$ Alk = carbonic acid alkalinity of reactor sludge liquor (mol/ℓ)
p_{CO2} = partial pressure of carbon dioxide in the reactor headspace (0.8 atm)

pH $=$ *in situ* pH of the reactor sludge liquor
K_{CO2} $=$ Henry's Law Constant ($K_{CO2} = 0.0149$ at 60° C)
K_w $=$ thermodynamic ionic product for water ($pK_w = 14$)
K_1,K_2 $=$ thermodynamic equilibrium constants ($pK_1 = 6.3$ and $pK_2 = 10.14$, both at 60° C)
f_m $=$ activity coefficient for monovalent species ($f_m = 0.8$)

(2) The $CO_3^=$ acidity of the reactor sludge liquor is determined from the calculated *in situ* pH and the estimated $H_2CO_3{}^*$ alkalinity of the reactor sludge liquor. The following equation is employed in this calculation:

$$H_2CO_3{}^*\ Alk = \left[\frac{1+2.10^{(pH-pK_2)}}{1+2.10^{(pK_1-pH)}}\right] \cdot [CO_3^=\ Acid + 10^{(pH-pK_w)} - 10^{-pH}] \tag{4E.2}$$

$H_2CO_3{}^*$ alkalinity and calculated $CO_3^=$ acidity of the reactor sludge liquor: i.e.

$$2C_T = H_2CO_3{}^*\ Alk + CO_3^=\ Acidity \quad (mol/\ell) \tag{4E.3}$$

where C_T $=$ the molar concentration of carbonic species in the reactor sludge liquor (mol/ℓ).

(4) The molar concentration of dissolved CO_2 remaining in the reactor sludge liquor is equated to C_T because the influent C_T for the low $H_2CO_3{}^*$ alk wastewaters characteristic of the Western Cape is negligible compared to the C_T increase in the reactor (see Table 5B.2b).. The daily amount of CO_2 lost from the reactor dissolved in the sludge liquor is calculated and added to the daily average measured amount of carbon dioxide in the vent gas to yield the total daily CO_2 production. A respiration quotient is obtained by dividing the daily average measured oxygen consumption into the total daily CO_2 production. The following equation is employed:

$$Y_{CO2} = \frac{36000 \cdot C_T + 7920}{12000} \tag{4E.4}$$

where,
Y_{CO2} $=$ the respiration quotient

36000 = the daily sludge liquor flow rate (ℓ/d)
7920 = the daily average measured CO_2 flow rate in the vent gas (mol/d)
12000 = the daily average measured O_2 consumption rate (mol/d)

THE RESPIRATION QUOTIENT (Y_{CO2})

The values of Y_{CO2} calculated using the above procedure are listed in Table 4E.1.

TABLE 4E.1: CALCULATIONS FOR THE RESPIRATION QUOTIENT

$H_2CO_3^*$ Alkalinity (mole/ℓ)	In situ pH	Acidity (mole/ℓ)	C_T (mole/ℓ)	Y_{CO2}
0.008	6.13	0.0316	0.0198	0.72
0.016	6.43	0.0397	0.0279	0.74

The values of 0.72 and 0.74 represent 9 and 12% increases respectively on the average apparent Y'_{CO2} value of 0.66 but the corrected values remain well below the expected Y_{CO2} value of 1.0. Although the value calculated from Eq 4E.4 is higher than the average apparent value, the increase is too slight to necessitate the modification of the Y_{CO2} values reported in Chapter 4 in order to account for carbon dioxide which remained dissolved in the sludge liquor and hence which was not measured as a constituent of the vent gas. The justification for this is as follows:

> During the specific heat yield test the increase in C_T applies only to 3.75m^3 of the 45m^3 reactor sludge liquor, this being the volume of sludge fed to the reactor at the start of the test. The C_T increases of 0.0198 and 0.0279 mol/ℓ (see Table 4E.1) represent 74 and 104 moles respectively of CO_2 dissolved during a specific heat yield test. Now, noting that the average rate of CO_2 generation during the specific heat yield test was 330 mol/h, it can be calculated that the above C_T increases would have occurred within 13 and 19 minutes respectively of the commencement of a sludge feed to the reactor. Generally the commencement of vent gas volume measurement for the specific heat yield test only occurred about 15 minutes after the commencement of sludge feeding, by which time the increase in C_T in the reactor would have been completed. Hence, the measurement of CO_2 production by means of vent gas volume and composition analysis during the specific yield test and the resulting calculated values of Y_{CO2}, would not have been affected by the dissolution of CO_2 in the reactor sludge liquor.

APPENDIX 5A

THE MOLAR FLOW AND HEAT BALANCE FOR GASES ENTERING AND LEAVING THE AEROBIC REACTOR

5A.1 INTRODUCTION

The air or pure oxygen that is used to oxygenate the aerobic reactor of a dual digestion system both adds heat to, and removes heat from, the reactor. This heat takes the form of the sensible heat of the gas, and the latent heat of water vaporized by the gas. The steady state heat balance across the aerobic reactor of a dual digestion system is discussed in Chapter 3. The mathematical formulation of this balance is represented by the following equation:

Heat sources = Heat sinks

$$H_{bi} + H_{mi} + H_{gi} + H_{si} - H_{se} + H_{ge} + H_{ve} + H_{we} \quad \{MJ/h\} \qquad (5A.1)$$

where,

H_{bi} = the rate of biological heat generation {MJ/h}

H_{mi} = the rate of heat generation due to the mechanical action of mixers and pumps {MJ/h}

H_{si},H_{se} = the rates of sensible heat gain and loss due to sludge entering and leaving the reactor respectively {MJ/h}

H_{gi},H_{ge} = the rates of sensible heat gain and loss due to gases entering and leaving the reactor respectively {MJ/h}

H_{ve} = the rate of heat loss due to the vaporization of water from the reactor sludge {MJ/h}

H_{we} = the rate of heat loss from the surfaces and walls of the reactor {MJ/h}

The net rate of sensible heat loss in the reactor vent gas (H_{ge}- H_{gi}) is proportional to the molar flow rates and the temperatures of the influent and the vent gases. The rate of heat loss due to water vapour in the vent gas is proportional to the temperature, flow rate and degree of water vapour saturation of the vent gas. The aim of this appendix is to develop equations which

(1) describe the molar flow rates of the reactor influent and vent gases

(2) calculate the net loss of sensible heat in the vent gas for aerobic reactors that are oxygenated with pure oxygen or air. The equations can be readily adapted to suit reactors aerated with pure oxygen enriched air.

(3) calculate the rate of heat loss from water vaporization in a reactor vent gas that is saturated with water vapour.

5A.2 GAS MOLAR FLOW RATES

The molar flow rate calculations presented below are based on 4 assumptions:

(1) The gases obey the ideal gas laws.

(2) Air is assumed to contain only oxygen and nitrogen, i.e. the percentage oxygen in air by mass is 23,17% and the remainder, 76,83% is assumed to be nitrogen. In conformity to this, the argon, carbon dioxide, and water vapour in the influent are ignored. Hence it is assumed that the influent air is dry and that all the carbon dioxide present in the vent gas has been generated in the reactor.

(3) The ratio of the moles of carbon dioxide produced to the moles of oxygen consumed in the aerobic reactor is a fixed value which is termed the respiration quotient and denoted by the symbol Y_{CO2}.

(4) The molar masses (kg/kmol) of the gases described in the calculations are:

Oxygen (O_2) = 32:
Carbon dioxide (CO_2) = 44:
Nitrogen (N_2) = 28:
Water (H_2O) = 18:

In the derivations below, separate equations are presented for the cases of reactor oxygenation by means of pure oxygen and by means of air. The equations are derived in terms of 3 parameters only, i.e. (1) oxygen consumption rate O_c, (2) oxygen transfer/utilization efficiency O_{eff}, and (3) respiration quotient Y_{CO2}.

5A.2.1 <u>Influent gas molar flow rate</u>

The molar flow rate of oxygen supplied to the reactor is:

$$O_i = O_c/(32 * O_{eff}) \quad \{kmol/h\} \qquad (5A.2)$$

The molar flow rate of nitrogen supplied to the reactor is:

$$N_i = 0.7683 * O_c/(28 * O_{eff} * 0.2317) \quad \{kmol/h\} \tag{5A.3}$$

5A.2.2 Vent gas molar flow rate

The molar flow rate of oxygen in the reactor vent gas is:

$$\begin{aligned} O_e &= O_i - O_c/32 \quad \{kmol/h\} \\ &= O_c/32\ [\ (1-O_{eff})/O_{eff}] \quad \{kmol/h\} \end{aligned} \tag{5A.4}$$

The molar flow rate of carbon dioxide in the vent gas is:

$$C_e = (O_c/32) * Y_{CO2} \quad \{kmol/h\} \tag{5A.5}$$

The molar flow rate of nitrogen in the vent gas is:

$$\begin{aligned} N_e &= 0.7683\ (\ O_c/(28 * O_{eff} * 0.2317) \quad \{kmol/h\} \\ &= N_i \end{aligned} \tag{5A.6}$$

5A.2.3 Summary

The dry influent (m_i) and dry vent (m_{dg}) total molar gas flow rates for pure oxygen and air oxygenated reactors are as follows:

For pure oxygen:

$$\begin{aligned} m_{io} &= O_i \qquad \{kmol/h\} \\ &= 1/32 \times \frac{0_c}{0_{eff}} \end{aligned} \tag{5A.7}$$

$$\begin{aligned} m_{dgo} &= O_e + C_e \qquad \{kmol/h\} \\ &= 1/32 \times \frac{0_c}{0_{eff}} \times (1 - O_{eff} + O_{eff} * Y_{CO2}) \end{aligned} \tag{5A.8}$$

For air:

$$m_{ia} = O_i + N_i \qquad \{kmol/h\}$$
$$= 0.15 \times \frac{O_c}{O_{eff}} \qquad (5A.9)$$
$$= 4.79\ m_{io}$$

$$m_{dga} = O_e + C_e + Ne \qquad \{kmol/h\}$$
$$= 1/32 \times \frac{0_c}{0_{eff}} \times (4.79 - O_{eff} + O_{eff} * Y_{CO2}) \qquad (5A.10)$$

where,

m_i = the molar flow rate of dry influent gas {kmol/h}
m_{dg} = the molar flow rate of dry vent gas {kmol/h}

Additional subscripts o and a denote pure oxygen and air oxygenated systems.

The above equations can be adapted for oxygen enriched air oxygenation by reducing constant, 4.79 appropriately to reflect the increased oxygen mass in the influent gas. The constant 4.79 denoted generally by a arises out of the derivation of Eq (5A.10), i.e.

$$a = 1 + (1-f_{mo_2})\ 32/\{28 \cdot f_{mo_2}\}$$

where,

f_{mo_2} = fraction by mass of oxygen in influent gas
= 0.2317 for air
= 1.00 for pure oxygen.

5A.3 THE NET DRY GAS SENSIBLE HEAT LOSS, $(H_{ge} - H_{gi})$

Lengthy calculations are required to determine the net sensible heat loss in the dry vent gas, $(H_{ge} - H_{gi})$. In this section, simplified calculations are presented in order to facilitate the determination of what, in reality, amounts to a very small portion of the heat losses from the reactor. However, a synopsis of the complete calculation procedure is first presented so that the modifications in the simplified procedure may be more clearly understood.

5A.3.1 Complete calculations

The sensible heat of a gas may be calculated with the aid of the following equation:

$$H \quad = m * Cp * (T_g - T_{ref}) \quad \{MJ\} \tag{5A.11}$$

where,

H = the gas sensible heat {MJ}
m = the moles of the gas {MJ}
Cp = the specific heat yield {MJ/(mol.°C)}
T_g = the temperature of the gas {°C}
T_{ref} = an arbitrary reference temperature {°C}

The specific heat yield, Cp, of a gas is a function of temperature, T_g, and may be calculated with the following general equation:

$$Cp = a + b.(T_g) + c.(T_g)^2 + d.(T_g)^3 \quad \{MJ/(mol.^\circ C)\} \tag{5A.12}$$

where,
a, b, c, and d are constants specific to the particular gas.

Hence Eq (5A.11) becomes:

$$H = m * \int_{T_{ref}}^{T_g} (a + b.T + c.T^2 + d.T^3) \quad \{MJ\} \tag{5A.13}$$

To compute the net sensible heat loss in the vent gas, Eq (5A.13) above must be solved for each component of the influent and vent gases.

5A.3.2 Simplified calculations

The following assumptions may be made in order to simplify the above procedure:

(1) An average specific heat yield, Cp, may be accepted in Eq (5A.11). This average value will be a composite value for both the influent and vent gases and does not change with temperature. A value of 0.0368 MJ/(kmol.°C)

has been assumed for a pure oxygen system with a dry vent gas composition of 20% oxygen and 80% carbon dioxide. A value of 0.0302 MJ/(kmol.°C) has been selected for an air system with a dry vent gas composition of 80% nitrogen, 10% oxygen and 10% carbon dioxide.

(2) The molar flow rate of the vent gas is the same as that of the influent gas.

These simplifying assumptions are justified because:

(1) Over the range of gas temperatures encountered in the aerobic stage of dual digestion, the specific heats change only slightly. For example, the specific heats of oxygen, nitrogen and carbon dioxide at 20° C and 60° C are:

oxygen:	0.0299 and 0.0302 MJ/(kmol.°C) respectively
nitrogen:	0.0291 and 0.0293 MJ/(kmol.°C) respectively
carbon dioxide	0.0379 and 0.0392 MJ/(kmol.°C) respectively

(2) The molar flow rate of the vent gas will be smaller than that of the influent gas if the respiration quotient, Y_{CO_2}, is less than unity. For example, at Milnerton where pure oxygen was used to oxygenate the sludge, 0.44 kmol of vent gas typically corresponded to 0.61 kmol of influent gas with a respiration quotient, Y_{CO_2}, of 0.66. However, had air been used to oxygenate the sludge, the difference between the flow rates would have been much smaller due to the large molar flow rate of nitrogen which passed through the reactor unchanged. Thus the assumption of equimolar influent and vent gas flow rates when Y_{CO_2} is less than unity, is subject to greater error when the sludge is oxygenated with pure oxygen than with air. However, the sensible heat losses in the vent gas from a reactor oxygenated with pure oxygen constitute a negligible portion of the total heat losses and the slight error in their calculation will have a negligible effect on the heat balance given in Eq (5A.1) above.

Thus the net rate of sensible heat loss in the dry vent gas is given by

$$(H_{ge} - H_{gi}) = m_i * Cp * (T_{ge} - T_{gi}) \quad \{MJ/h\} \qquad (5A.14)$$

Substituting the composite values for Cp listed in assumption (1) above and Eqs (5A.7) and (5A.9) describing m_i for pure oxygen and air respectively into Eq (5A.14), yields the following:

Pure oxygen:

$$(H_{ge} - H_{gi}) = 0.00115 \times \frac{0_c}{0_{eff}} \times (T_{ge} - T_{gi}) \quad \text{(MJ/h)} \qquad (5A.15)$$

Air:

$$(H_{ge} - H_{gi}) = 0.00453 \times \frac{0_c}{0_{eff}} \times (T_{ge} - T_{gi}) \quad \text{(MJ/h)} \qquad (5A.16)$$

where,

T_{ge} = the temperature of the vent gas {°C}
T_{gi} = the temperature of the influent gas {°C}

5A.4 THE RATE OF HEAT LOSS DUE TO THE VAPORIZATION OF WATER BY THE REACTOR VENT GAS

The rate of heat loss in the reactor vent gas, H_{ve}, may be calculated with the following equation:

$$H_{ve} = h_{fg} * M_{wp} \quad \{MJ/h\} \qquad (5A.17)$$

where,

h_{fg} = the latent heat of vaporization of water {MJ/kg H_2O}
M_{wp} = the predicted mass rate of water vaporization {kg H_2O/h}

5A.4.1 <u>The latent heat of water vaporization, h_{fg}</u>

The latent heat of water vaporization is a function of temperature. It may be calculated with the following equation which has been derived from data presented in steam tables:

$$h_{fg} = 2.510 - 2.386*10^{-3} * T_{ge} \quad \{MJ/kg\ H_2O\} \qquad (A5.18)$$

where,

T_{ge} = the vent gas temperature {°C}

In the aerobic reactor of a dual digestion system, the temperatures typically range between 55 and 65°C and as the variation in h_{fg} is very slight over this temperature range, the errors arising from the use of an average value will be insignificant.

5A.4.2 <u>The predicted rate of water vaporization</u>, M_{wp}

When a 2 phase system consisting of gas and water are(is) in a state of equilibrium, the gas will be saturated with water vapour. According to ideal gas laws, at this stage of equilibrium, the following ratios will be equal:

$$m_w/m_{dg} = p_w/p_{dg} \tag{5A.19}$$

where,

m_w = the moles of water vapour in the gas phase
m_{dg} = the moles of dry gas
p_w = the partial pressure of the water vapour (mm Hg)
p_{dg} = the partial pressure of the dry gas (mm Hg)

Note that, from Dalton's law, the total pressure of the gas phase P_T is equal to the sum of the partial pressures in the gas phase. Hence:

$$P_T = p_w + p_{dg} \quad \text{(mm Hg)} \tag{5A.20}$$

The partial pressure of water vapour in air may be calculated with the aid of the Antoine equation:

$$\text{Log}(p_w) = B - A/(T_{ge} + 273) \tag{5A.21}$$

where,

B = 8.896
A = 2238
T_{ge} = vent gas temperature at the point of exit from the reactor {°C}.

5A.4.3 <u>The rate of vapour heat loss in a saturated vent gas</u>

Thus to calculate the rate of heat loss due to water vaporization in a *saturated* vent gas from the aerobic reactor of a dual digestion system, H_{ve} by means of Eq

(5A.17), the following equations may be used to calculate the required latent heat of water vaporization h_{fg} and water vapour mass flow rate M_{wp}:

From Eq (5A.18), the latent heat of water vaporization is found i.e.

$$h_{fg} = 2.501 - 2.386 \times 10^{-3} \times T_{ge} \quad (MJ/kg) \tag{5A.22}$$

and from Eq (5A.19), the water vapour mass flow rate is found, i.e.

$$M_{wp} = (p_w/p_{dg}).18.m_{dg} \quad (kg\ H_2O/h)$$

where, from Eq (5A.21):

$$p_w = 10^{\left[-\frac{2238}{T_{ge}+273}+8,896\right]} \quad (mm\ Hg) \tag{5A.24}$$

From Eq (5A.20):

$$p_{dg} = P_T - p_w \quad (mm\ Hg) \tag{5A.25}$$

and from Eqs (5A.8) and (5A.10):

$$m_{dg} = 0.03125 \times \frac{0_c}{0_{eff}} \times (a - O_{eff} + O_{eff} \times Y_{CO2}) \quad (kmol/h) \tag{5A.26}$$

where,

a = 1 when the reactor is oxygenated with pure oxygen

a = 4.79 when the reactor is oxygenated with air.

Should the vent gas not be saturated, H_{ve} can still be calculated by the above method, but is reduced in proportion to the degree of water vapour saturation.

APPENDIX 5B.

OPERATIONAL ASPECTS OF THE MILNERTON DUAL DIGESTION SYSTEM.

INTRODUCTION

The principal objective of this thesis was to investigate and evaluate biological heat generation in the aerobic reactor of the Milnerton dual digestion system and the body of this thesis focuses on this aspect of the project. However, in undertaking this investigation, a wealth of experience with the dual digestion system as a whole was obtained which, while not directly relevant to this thesis, is of significant value to prospective designers and operators of a dual digestion system. In this Appendix, some of this operational experience is discussed, covering the following topics:

Aerobic reactor
- Start–up
- Foaming
- Biological "death" of sludge
- Mixing
- Level control
- Abrasive corrosion of the Vitox recirculation pump
- Aerobic reactor collapse
- Energy considerations

Anaerobic Digestor
- Gas pressure fluctuations in the headspace
- Overheating

Dual digestion system
- Disinfection of sludge
- Overall performance

The discussion focuses on operational details, including the initial failures and problems which formed as instructive (if not as agreeable) a part of the research as did the eventual successes. It is hoped that the information in this chapter will assist future operators of a dual digestion system and enable them to avoid repeating some of the mistakes made at Milnerton.

In addition to the operational details, two research topics, which are not the subject of this thesis, are briefly presented in this Appendix. They deal with the sludge disinfection capability of the process and the performance of the anaerobic stage. The data pertaining to these topics are summarized, with only cursory comment as to their significance. The reader is directed to the Division of Water Technology, CSIR[1], reports by H.A. De Villiers *et al.* for more detailed information on both of these topics.

5B.1 THE AEROBIC REACTOR

5B.1.1 Start up of the aerobic reactor

Over a period of 2 years (1987 and 1988), the aerobic reactor was started up on 4 different occasions with 45 m^3 of raw, ambient temperature sludge. The start up phase usually lasted for about 4 days and was considered to be completed once an active aerobic culture had been established at a reactor sludge temperature of ±60°C. Fig 5B.1 illustrates the change in reactor sludge temperature with time over a typical start up.

During the start up phase the reactor was periodically fed with raw sludge to ensure that sufficient substrate was available to establish an active aerobic culture. In Fig.5B.1 the times at which feeds took place can be seen as the times of sudden temperature decrease.

As was the case for normal operation, sludge recirculation and oxygenation were continuous during start up. In the first year of operation (1987), the rate of oxygen supply over the start up phase ranged from 15 to 20 kgO/h but in subsequent start ups, this rate was reduced to 6 kgO/h in an attempt to control foaming. Foam was generated in the reactor throughout the start up phase, but ceased once an active aerobic culture had been established in the sludge (see Section 5B.1.2 below).

It was easy to identify at which point an active aerobic culture had been

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[1] Council for Scientific and Industrial Research.

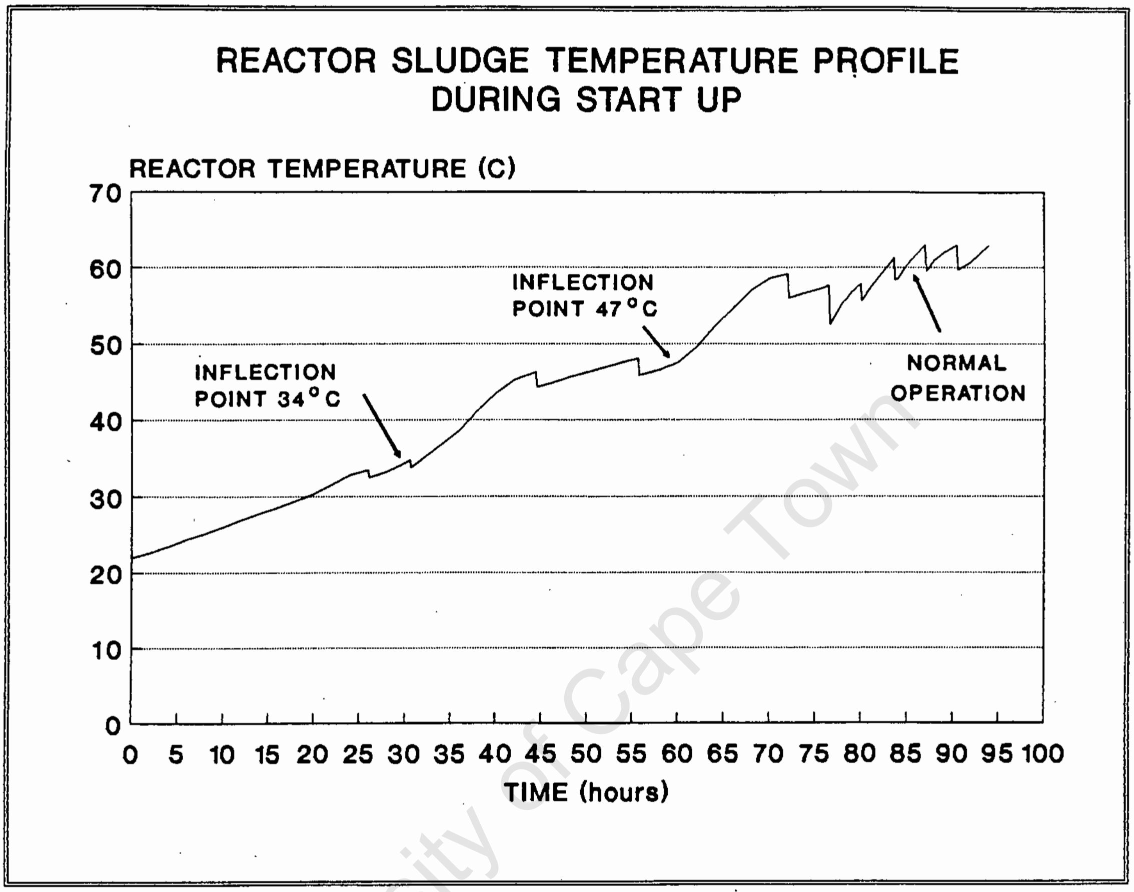


Fig.5B.1: The change in reactor sludge temperature with time during a typical start up of the aerobic reactor from ambient temperature. While the temperature change over the first 24 hours is approximately linear and is due to heat input from the Vitox recirculation pump, inflection points on the temperature trace indicate heat generation associated with incipient biological activity. The introduction of sludge feed to the reactor is shown by sharp decreases in reactor sludge temperature. In the above illustration, about 90 hours (3.75 days) elapsed between starting up with raw sludge and the development of a fully active biological culture.

established. Besides the cessation of foaming, another indicator of incipient biological activity was the rate of change of reactor sludge temperature. Immediately after start up this rate of reactor sludge temperature change was consistent with the rate of mechanical heat input from the recirculation pump. However, when biological activity commenced, oxygen consumption began, and the concomitant generation of biological heat caused the rate of sludge temperature change to show a sudden increase. These times are noticeable as inflection points on the reactor sludge temperature curve and in Fig 5B.1 there are 2 such points on the curve. These points correspond to temperatures of 34 and 47°C and might indicate points of transition for mesophilic and thermophilic cultures. Within 35 hours of commencing sludge oxygenation, the reactor sludge was biologically active and by 90 hours it had reached the desired operating temperature of about 60°C.

On one occasion 2 weeks were required for the start up instead of the usual 4 days. In this instance biological activity only commenced after the reactor sludge pH had been raised from a pH of ±5.2 to 7 by means of lime addition. On previous occasions pH correction of the feed sludge (pH = ±5.1) had not been necessary during start up; the pH had increased, without lime addition, to pH = 7 within about 4 days, due to alkalinity generated by the biological activity. Nevertheless, subsequent to this occasion, the precaution was usually taken of correcting the feed sludge pH to pH = 7 before starting up the reactor.

5B.1.2 Foaming in the aerobic reactor

In all but one of the occasions when the aerobic reactor produced foam, the sludge in the reactor was biologically dead. *Foam production was never observed during normal operation*, when oxygen was being consumed and biological heat was being generated – both indicators of a fully developed and active biological culture.

Systematic foaming occurred when the aerobic reactor was started up with fresh sludge, and continued until the sludge became biologically active (and consumed most of the oxygen supplied). It could not be avoided, short of starting up without oxygen injection – a step which is at odds with the aim of a start up i.e. to develop an active aerobic culture in the sludge. Operation with low flows of oxygen (6 kgO/h) reduced the volume of foam produced but did not prevent it occurring. This foam was probably caused by large amounts of undissolved oxygen passing through the sludge; as the sludge began to consume more oxygen during start up, more was dissolved, and the foaming tapered off. This explanation possibly accounts for the continuous foaming which is observed in the normal operation of air oxygenated reactors – in this instance large volumes of undissolved nitrogen probably cause the foaming. If this explanation is correct, then it may have been possible to avoid foaming on reactor start up at Milnerton by means of dissolved oxygen control in the reactor sludge – oxygen supply on start up would then have been intermittent, depending on the sludge oxygen concentration. However this control would only have been used during start up, as continuous oxygenation is desirable for normal operation.

Foaming was also noticed when the sludge "died" after having been deprived of oxygen for an more than an hour. This foaming ceased after the sludge became active (see Section 5B.3 on biological death below). On only one occasion did the reactor produce foam when the sludge culture was active. This was after a step increase in oxygen supply from 6 to 19.4 kgO/h and a step decrease in retention time from 3 to 1.25 days (see Chapter 4, Section 4.8). Foaming was very slight, lasting for about 30 minutes. It appeared to be the result of an acclimatization phase in which, after its growth had been constrained by the 6 kgO/h oxygen supply for about 14 days at a 3 day retention time, the sludge culture adapted to the higher supply rate of oxygen and shorter retention time.

Foaming did not cause undue problems at Milnerton – it seldom occurred and was highly predictable. When the reactor did produce foam, the operating staff were able to take steps to prevent it from becoming a nuisance. These steps involved simply diverting the foam to drain. A 100mm sludge overflow pipe fitted with a 300mm water seal and feeding to a drain, was fitted to the reactor at a point above the 45m^3 operating mark (see Fig 5B.2). At the start of foaming, the 75mm vent gas pipe situated at the top of the reactor, was isolated and the foam flowed to drain through the overflow pipe. It is interesting to note, however, that the foam did not flow to drain of its own accord, but required some persuasion. Even without the 300mm water seal on the sludge overflow pipe, if the vent gas pipe was not isolated, then the foam filled up the entire headspace, crept along the vent gas pipe and fouled the vent gas monitoring equipment. Although this was prevented by isolating the vent gas pipe, the foam still did not flow freely down the overflow pipe. Instead, it was blown down this pipe in regular bursts as

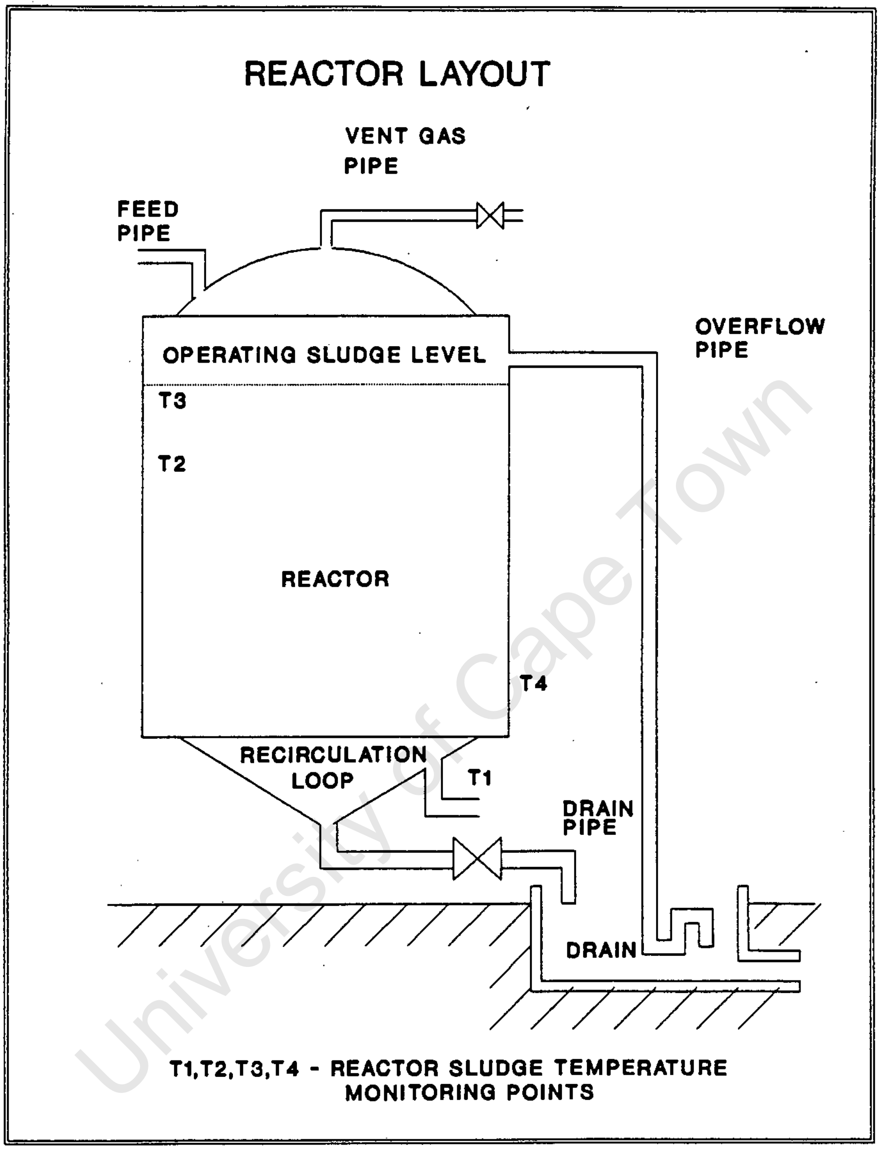


Fig.5B.2: A schematic diagram of the aerobic reactor showing pipework and the approximate location of temperature monitoring points.

pressure from the oxygenation gas built up in the headspace. Had an open topped reactor been used, the foam would most certainly have flowed over the top of the reactor rather than down the overflow pipe.

When the reactor produced foam, precautions had to be taken against creating a vacuum in the reactor headspace. After isolating the vent gas pipe as described above, only the overflow pipe connected the reactor to the ambient atmosphere. In the event of a sudden vacuum in the reactor headspace, air was drawn through the overflow pipe to relieve the vacuum. However during foaming, this vacuum relief was often obstructed by an accumulation of sludge and foam in the overflow pipe, and consequently the vacuum could have caused the glass fibre reactor to collapse (implode) (see Section 5B.1.7 below). It was possible to create such a vacuum by transferring sludge out of the reactor, and therefore sludge transfers were not allowed during periods of foaming.

5B.1.3 Biological 'death' of the reactor sludge

The reactor sludge was considered to be 'dead' when

(1) the sludge generated foam and
(2) no biological heat generation could be discerned and therefore the reactor sludge temperature increased at a rate which was consistent with the addition of mechanical heat (from the recirculation pump) only.

On only 1 occasion in 3 years of operation, did the reactor sludge 'die' for no apparent reason. This death occurred during normal operation and lasted for about 50 minutes, after which the plant recovered completely. All other instances of sludge death were caused by stopping sludge oxygenation while mechanical repairs to the plant were undertaken. It appeared that without oxygenation the sludge would 'die' after 45 to 60 minutes. The time taken to recover from this death depended on for how long the sludge was deprived of oxygen. It was observed that if the sludge was without oxygen for an hour, it would recover in about 30 minutes, whereas after 3 hours without oxygen, the sludge could take up to 24 hours to recover.

The observation of sludge 'death' precludes the use of intermittent, rather than continuous, reactor sludge oxygenation. For example, the description of operating procedures for an autothermal thermophilic aerobic digestor at Castle Donnington in Great Britain, stated that digestor sludge recirculation and oxygenation were stopped when the digestor sludge temperature exceeded the operating temperature of 55° C. On the basis of the Milnerton experience, this practice probably caused sludge death whenever the reactor reached operating temperature.

5B.1.4 Reactor sludge mixing

The mixing of the reactor sludge was accomplished with the Vitox oxygenation system (see Fig 3.2) which formed a recirculation loop about the reactor. As previously described (see Section 3.3.4.3), the quality of mixing was judged by the uniformity of response to temperature change of the 4 PT100 resistance thermometers located at different points in the reactor (see Fig 5B.2 for temperature probe positions in the reactor).

5B.1.4.1 *Sludge temperature and reactor mixing*

The reactor sludge temperature was monitored at 3 points on the reactor and at 1 point in the Vitox recirculation loop (see Fig 5B.2). When the sludge was well mixed, all 4 sludge temperature readings corresponded closely (allowing for calibration discrepancies between individual probes) and the 4 temperature traces on the chart recorder formed a tight bunch. When the mixing was poor, then 1 or more of the temperature probe readings did not correspond with the others and the chart recorder traces diverged, particularly following the addition of a batch of cold feed sludge. The difference between good and bad mixing for a single temperature probe has been illustrated in Fig 3.5 (see Chapter 3). In Fig 5B.5 good and bad mixing are illustrated with a copy of the actual Milnerton chart recorder trace of the 4 reactor sludge temperature readings.

The mixing quality was most apparent from both the individual and collective responses of the four temperature probes at times of sudden decrease in the reactor sludge temperature, such as when a batch of cold feed sludge was introduced into the hot reactor sludge. A rapid response by each individual probe and a simultaneous and uniform response by all 4, indicated good mixing. An example of this is illustrated in Fig 5B.3 by a plot of the 4 temperature readings recorded at 34.5 second intervals following a sludge feed. The most rapid response to temperature change was recorded with thermometer N^0 3 because it was located just below the sludge surface and directly beneath the point at which feed sludge entered the reactor.

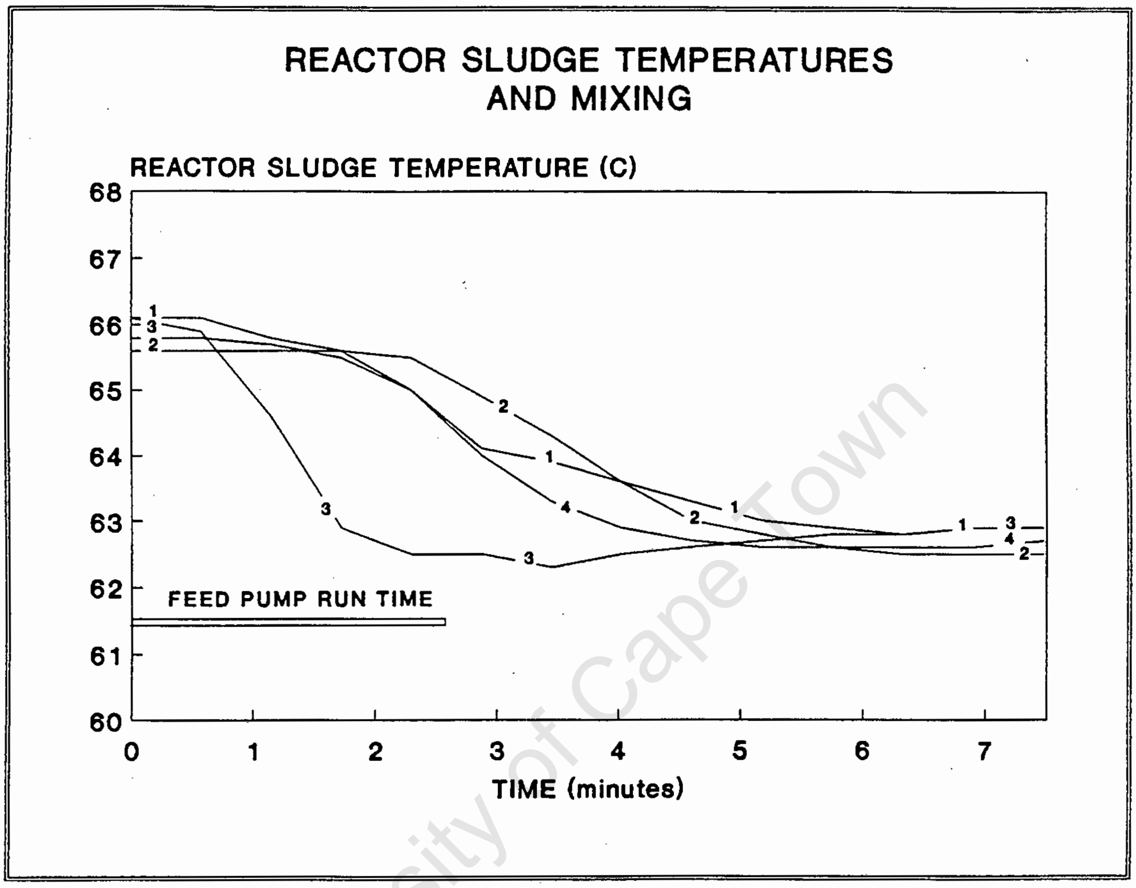


Fig.5B.3: The change in reactor sludge temperature with time following the addition of a batch of feed sludge to the reactor. The rapid and simultaneous response to temperature change by the 4 temperature probes indicate good mixing in the reactor. Probe 3 registered the change most rapidly because it was located directly below the feed entry port.

In the first couple of months following the commissioning of the Milnerton dual digestion system, the reactor was badly mixed and temperature traces similar to those on the right hand side of Fig 5B.5 were obtained. Coincidentally, during this period it was difficult to consistently maintain the reactor sludge temperature at the desired operating temperature of about 60° C. Although the research and operating staff were intrigued by the unusual temperature trace, at this early stage they were unaware that it was caused by poor sludge mixing. However, this became apparent when it was discovered that the Vitox sludge distribution manifold inside the reactor was almost completely blocked and that, consequently, the sludge recirculation rate was too low to achieve effective mixing. An improved manifold design resulted in the desired reactor sludge temperatures and also produced a chart recorder temperature trace similar to that on the left hand side of Fig.5B.4, indicating good mixing. The development of the recirculation sludge distribution manifold that eventually produced the desired result is discussed below.

5B.1.4.2 *The sludge distribution manifold and reactor mixing*

The Vitox oxygenation system imparted 400 W/m^3 of mixing power to the reactor sludge and recirculated the sludge at a rate of about 150 m^3/h, turning the reactor sludge contents over approximately every 20 minutes. Although 400 W/m^3 is a more than adequate mixing power density by most waste water design guidelines, it nevertheless was no guarantee of good mixing unless attention was paid to the correct distribution of recirculated sludge in the reactor. This was accomplished by means of a sludge distribution manifold fitted to the discharge end of the Vitox recirculation loop and located about 1.5m below the reactor sludge surface. In the early stages of the project, 2 different manifold designs proved unsuitable before a third design yielded satisfactory performance. The 3 sludge distribution manifolds are illustrated in Fig 5B.4.

The initial manifold design [Fig 5B.4(a)] consisted of a 3 nozzle sparger directed vertically downwards in the reactor. Nozzle blockages were considered unlikely due to the high sludge discharge velocities (12 – 15 m/s), but experience proved the contrary because rags solidly blocked 2 of the nozzles and partially blocked the third. The blocked manifold was removed and, while a new manifold was constructed, the sludge was discharged into the reactor through the open 100mm steel recirculation pipe. The unobstructed flow increased the specific mixing

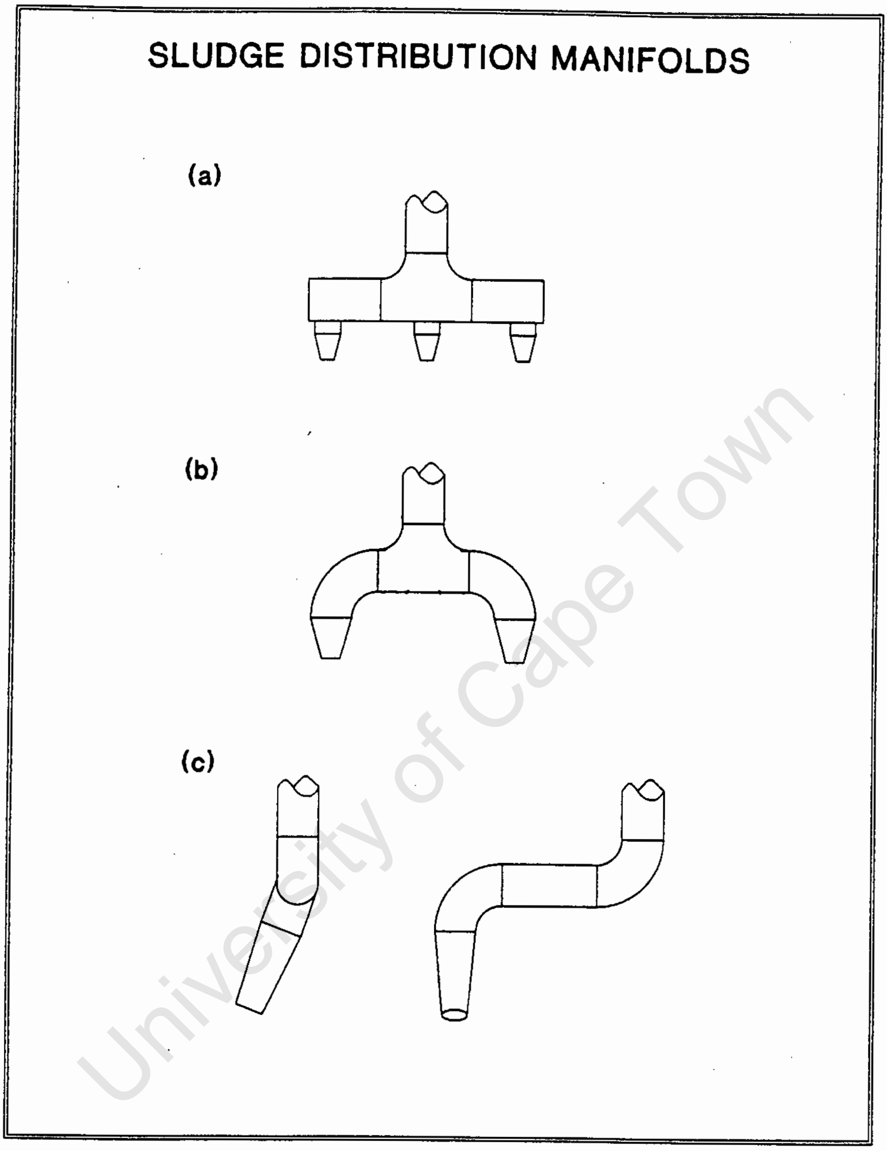


Fig.5B.4: The 3 successive designs of the sludge distribution manifold which was fitted to the discharge of the Vitox recirculation loop. Neither of the designs (a) and (b) provided satisfactory mixing and both were susceptible to blockage by material in the sludge. Manifold (c) performed satisfactorily and was designed after good sludge mixing was noticed following the blockage of one of the two nozzles in (b) - see Fig.5B.5.

power density from 400 to 450 W/m^3 but in spite of this, the sludge mixing remained poor.

The replacement manifold [Fig 5B.4(b)] discharged sludge through 2 nozzles and was more streamlined than its predecessor in the hope that sludge blockages would not recur. After its installation, the mixing power density reverted to 400 W/m^3 and mixing showed no improvement. Furthermore, the new manifold also proved to be no less resistant to blockage than the previous one, but this time the blockage served to point the way to a third, and this time successful, manifold design. After installation of the second sludge distribution manifold, one of the 2 nozzles on the manifold blocked within 48 hours and, immediately, mixing showed a substantial improvement. This event is clearly illustrated by the sludge temperature traces in the chart recording in Fig.5B.5. It is surmised that the sudden convergence of the temperature traces (marked in the illustration) coincided with the nozzle blockage. Certainly when the manifold was removed a few hours later, one of the nozzles was indeed blocked with rags. The resulting improvement in sludge mixing due to sludge discharge through a single nozzle is especially apparent in the responses of the 4 probes to sludge feeding subsequent to the blockage.

Thus, the third manifold design [Fig 5B.4(c)] consisted of a single nozzle, located to one side of the reactor axis and offset from the vertical by about 15°. This manifold provided good mixing and was impervious to blockage, operating successfully throughout the following 2 years. Subsequent to its installation, the reactor performance improved in that it became easier to consistently maintain the desired reactor sludge operating temperature.

5B.1.5 Sludge level control in the reactor

The functions of the sludge level control system were:

(1) To maintain a constant reactor operating volume of 45m^3.

(2) To control the draw and fill batch feeding system, ensuring that a correct volume of sludge was transferred from the reactor to the anaerobic digestor, and that an equivalent volume of sludge was fed into the reactor from the sludge thickener.

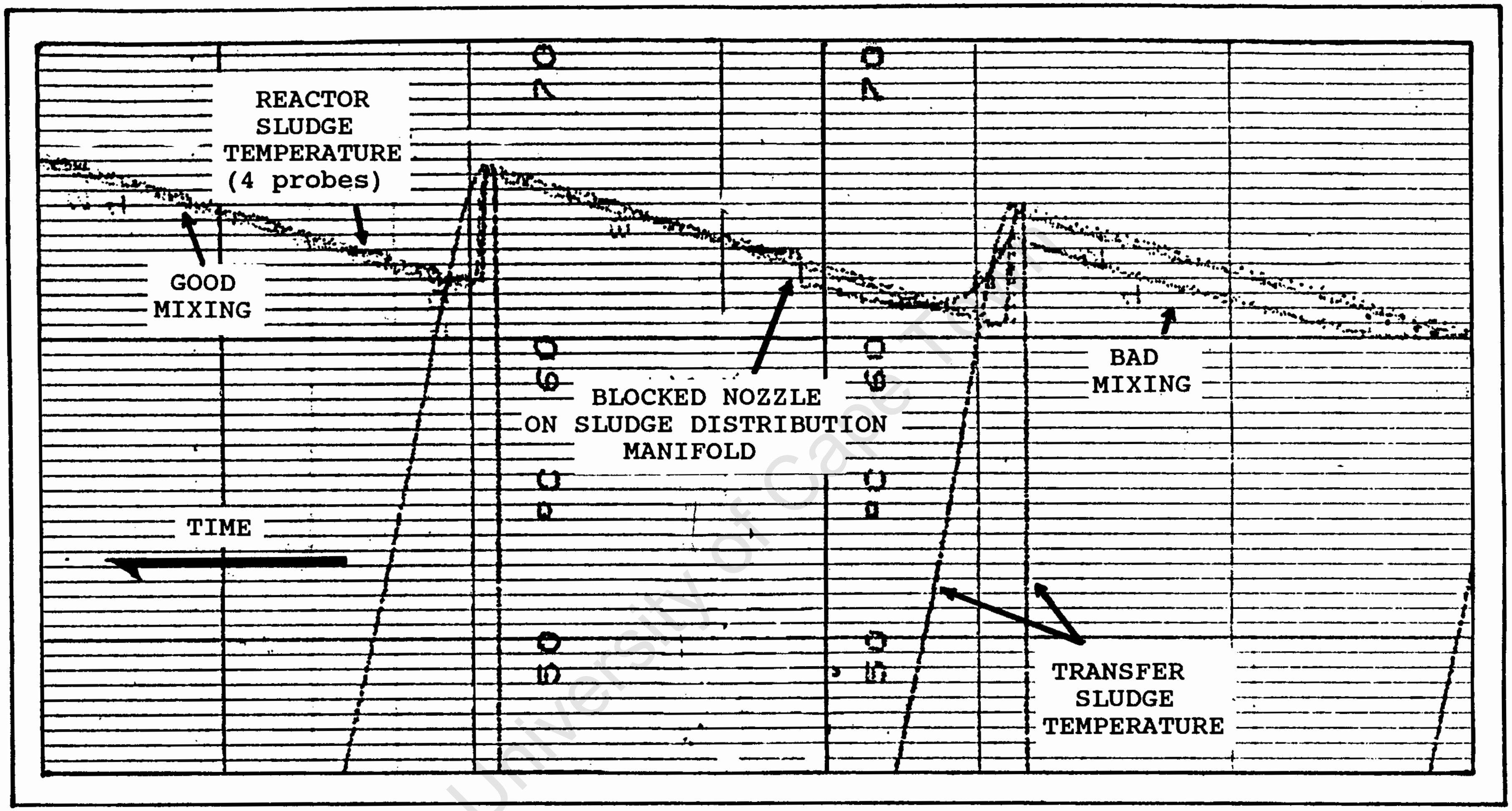


Fig.5B.5: A chart recorder trace of the temperatures in the dual digestion system. The saw tooth temperature-time profile on the RHS of the illustration is typical of the bad mixing achieved with the sludge distribution manifold illustrated in Fig.5B.4(b). One of the two nozzles in this manifold blocked at a point shown in the illustration and the saw tooth temperature-time profile to the left of this point indicates that good mixing was achieved with the remaining nozzle.

The level control system was mounted on top of the reactor so that its level detection sensors protruded through the reactor headspace and onto the sludge surface. Two types of level sensor, viz (1) conductivity probes and (2) capacitance probes, were tried and found to be unsuitable. Both systems were unable to provide reliable control of the draw and fill batch feeding system, resulting in an unacceptably large variability in the reactor operating volume. It appeared that sludge turbulence, trace amounts of foam and the high humidity of the vent gas in the reactor headspace interfered with their operation.

A third level control system (described in Chapter 3, Section 3.3.2) which proved to be successful, was entirely "home made", designed and constructed by the research and operating staff for a fraction of the cost of the more sophisticated, but unsuitable, systems. The description and illustration of the system presented in Chapter 3 is repeated below.

It comprised a heavy stainless steel float, which rested on the sludge liquid surface, and to which was connected a thin rigid rod with a plastic bottle containing a magnet fastened to its top end. The magnet would activate reed switches fixed to a rule at measured points. The reed switch at the high level corresponded to the 45m^3 operating volume and at the low level to 41.25m^3, the difference between the two levels being the 3.75m^3 batch volume. During normal operation the float held the magnet at the high level switch.

At intervals of not less than 2h, a quartz pulse timer activated the transfer pump and sludge was pumped from the reactor to the digestor causing the stainless steel float to drop until the low level reed switch was activated by the magnet. This terminated the transfer operation with exactly 3.75m^3 having been transferred. Immediately after termination of sludge transfer to the digestor, sludge feeding from the thickener would commence. This feeding was terminated by the high level reed switch when the operating volume was once again 45m3. The pulse timer merely initiated this transfer/feeding cycle and after initiation, the level control system controlled the sludge transfer and feed operations. Additional reed switches were attached to the rule above and below the high and low levels. These were alarm switches which would be activated in the event of a failure of the high or low level switches.

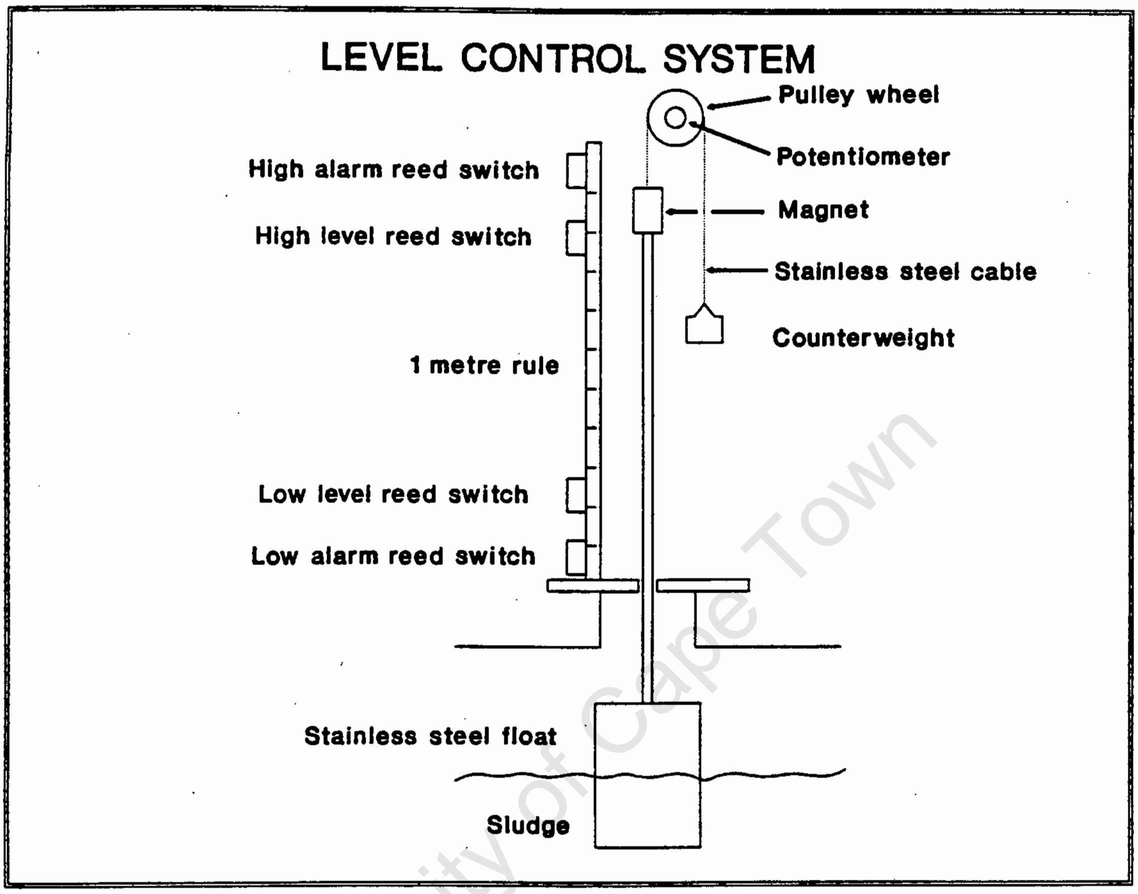


__Fig 3.3__: The aerobic reactor level control system which maintained a constant sludge level of 45m^3, and which controlled the batch draw and fill system to ensure that 3.75m^3 batches of sludge were transferred from and fed to the reactor.

A 1.5mm stainless steel cable was connected to the plastic bottle containing the magnet at the top of the vertical stainless steel rod. The cable was looped over a pulley wheel fitted onto the shaft of a potentiometer, and was connected to a counterweight on its other end. Vertical movement of the stainless steel rod rotated the pulley wheel and the potentiometer shaft. The rotation of the potentiometer changed the flow of electrical current in a circuit of which the potentiometer was a component. The change in electrical current was displayed on an ammeter, which was located in the control hut. This arrangement provided a visual indication, in the control hut, of the sludge level in the reactor. This level control system proved to be reliable, robust, and impervious both to foam and to fouling caused by material in the sludge.

5B.1.6 Abrasive erosion of the Vitox recirculation pump

In the 6 months following the commissioning of the Milnerton aerobic reactor, a large amount of attention and effort was dedicated to overcoming numerous mechanical 'teething troubles'. The single item which required the most attention was the Vitox recirculation pump, an Allis Chalmers NSR 4x4x12 model. The problems experienced with this pump fell into 4 categories, 2 of which were without remedy. These 2 were

(1) Poor standard of manufacture and assembly
(2) Cavitation - the pump appeared to cavitate (in the absence of oxygen injection) in spite of (i) being operated at its optimum design duty (ii) a flooded suction with a 7m static head and (iii) a low suction velocity over a short (2m) length of 200mm suction pipe. However, oxygen injection eliminated the external symptoms (eg noise) of cavitation entirely, apparently due to the entrainment of fine bubbles into the pump suction.

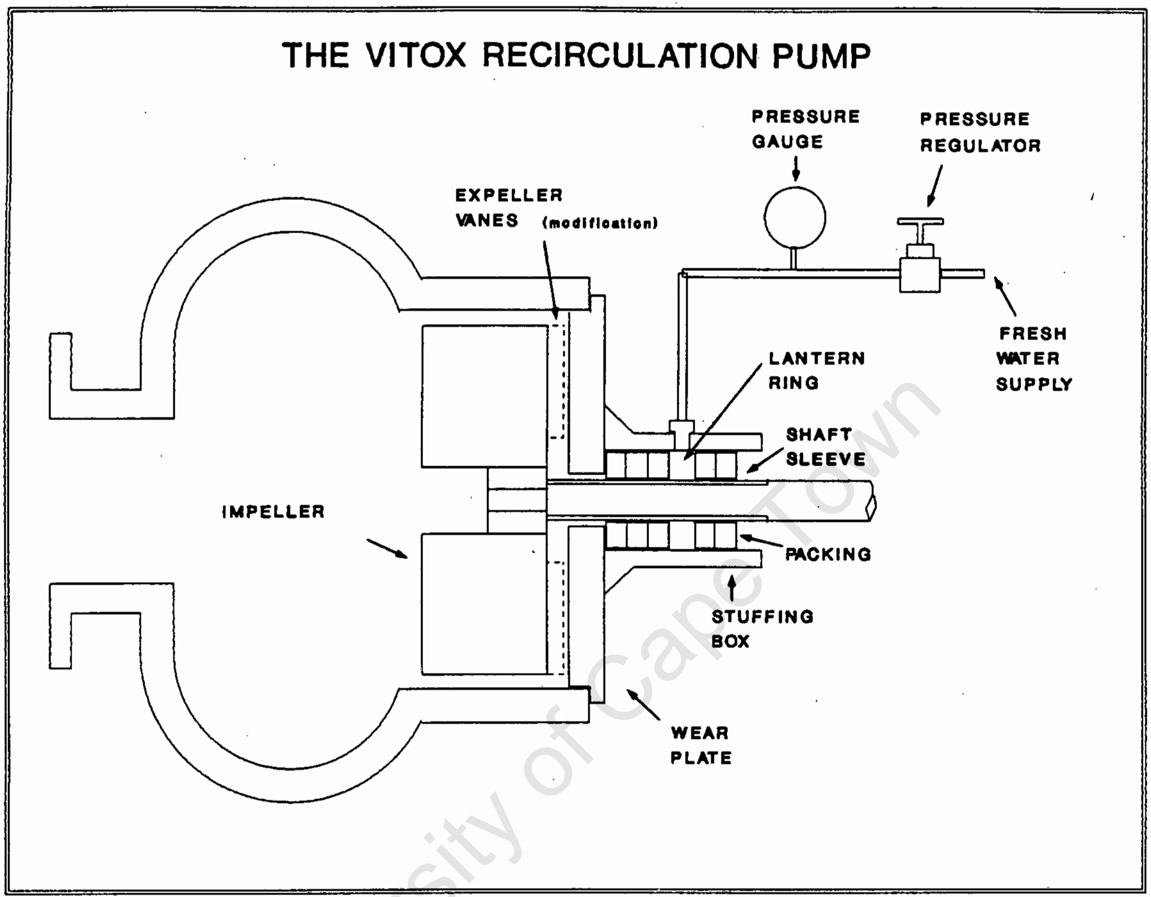
The two problems for which solutions were found were
(3) shaft sleeve wear and gland leakage
(4) erosion of the impeller and pump casing.

A section through the Vitox recirculation pump is illustrated in Fig 5B.6. The diagram is not drawn to scale and has been considerably simplified for the purposes of this discussion. The pump and impeller were both manufactured from cast iron, although after 6 months of operation the cast iron impeller was replaced with a stainless steel one.

During this first 6 months, the leakage of sludge past the pump shaft proved to be a nuisance. The shaft was sealed with a packed gland comprising 5 packing rings and a lantern ring. A 6mm pipe had connected the pump casing to the stuffing box in order to provide water from the pump discharge as a coolant and lubricant to the packing, but this was removed as it was clearly inappropriate considering that the pumped medium was a sludge with a high grit and solids content. Pressure behind the impeller forced sludge into the stuffing box and past the packing, causing a steady drip of sludge on the outboard side of the stuffing box. Grit in the sludge became embedded in the packing and rapidly abraded the stainless steel shaft sleeve. Within days the drip of sludge became a trickle. Tightening the gland became necessary to reduce the leakage but this only exacerbated the problem by damaging the packing and increasing the rate of shaft sleeve wear. After 6 months the pump had to be opened for repairs because the cast iron wear plate fractured. Inspection revealed that the stainless steel shaft sleeve had been worn through and a 3 mm deep groove had been cut into the shaft – a consequence of excessive gland tightening – necessitating shaft replacement.

In addition to the shaft damage, the impeller and the inside of the pump casing had been severely eroded. The worn impeller was replaced with one made of stainless steel and this new impeller proved to be more robust than its predecessor – after the following 12 months of continuous operation, wear was so slight that eroded surfaces could be filled in with welding. The problem of casing wear, however, was not solved as easily. Rebuilding and coating the inside surfaces with a variety of hard surfacing materials failed to reduce the rate of wear and consequently it became the practice to rebuild the casing at 6 to 8 month intervals. A standby pump was purchased to enable continuous plant operation while these repairs were being made. In Chapter 4, Section 4.2.2, the effects of pump wear on reactor performance are discussed.

When pressurized grease injection into the stuffing box and a variety of different packings all failed to reduce shaft sleeve wear, a more radical approach yielded the solution to this problem. Expeller vanes were welded to the back of the impeller shroud to reduce the pressure behind the impeller and in the stuffing box.



<u>Fig.5B.6</u>: A section through the Vitox recirculation pump which illustrates how shaft sleeve wear was overcome by (1) welding expeller vanes to the back of the impeller shroud and (2) by fitting a fresh water gland cooling and lubrication system.

Hydraulic testing with the reactor filled with treated sewage effluent showed this modification to be singularly successful, the stuffing box pressure dropping from 270 to 25 kPa. Clean tap water was injected into the stuffing box at a pressure of about 35 kPa and this served both as a packing coolant and lubricant, and as a barrier against the intrusion of abrasive grit from the sludge. The stainless shaft sleeve was replaced with a mild steel sleeve plated with a wear resistant hard chrome surfacing.

As a result of this approach, shaft sleeve and packing replacement was extended to 4 – 6 month intervals. Throughout these intervals, gland leakage was controlled to a steady drip of fresh water. In fact, by measuring the rate of gland leakage and the flow rate of fresh water to the stuffing box, the flow rate of fresh water to the inboard side of the stuffing box could be determined – a large inboard flow rate indicated that the shaft sleeve had worn excessively and needed to be replaced.

Altogether, the first 8 months of the project were punctuated with Vitox recirculation pump problems, but thereafter the modifications described above ensured relatively trouble free operation for the rest of the research project. Rebuilding the pump casing every 6 to 8 months was accepted by operating staff as normal maintenance and not inconsistent with over 5000 hours of pumping an abrasive sludge.

5B.1.7 Implosion of the aerobic reactor under vacuum

Of all the experiences during the research project, the one that caused the most consternation was the collapse under vacuum of the dome of the aerobic reactor. This incident occurred while emptying the reactor of sewage effluent water after completing the recirculation hydraulic test mentioned above in Section 5B.6.

Both the reactor overflow pipe and the reactor drain pipe (see Fig 5B.2) discharged to the same drain. Due to an oversight by project staff, the vent gas pipe at the top of the reactor was closed while the sewage effluent water emptied via the reactor drain pipe into the drain. Thus the only relief of vacuum in the reactor headspace was via the overflow pipe which terminated in the same drain. In itself, this would not have been a problem but, unfortunately, a new sludge dewatering system which had also completed hydraulic testing, was simultaneously being emptied to the same drainage system. The capacity of the drainage system was exceeded causing water to surcharge in all the drains. The level of effluent in the reactor drain rose until it sealed the overflow pipe, blocking the only vacuum relief. The headspace vacuum increased and the reactor dome collapsed (imploded) with a loud thud which startled the project staff. Fortunately, the situation was saved by the presence of mind of one of the project staff who quickly removed the 50mm plug from the vent gas pipe - normally this plug was removed with the aid of a wrench, but under the circumstances bare hands and a surfeit of adrenalin accomplished the task.

The fiberglass with which the reactor was constructed, proved to be a forgiving material, and once the vacuum was broken, the collapsed dome popped back into its original shape. The inside of the reactor was inspected but no structural damage could be discerned and the project was able to continue.

5B.1.8 Energy considerations in the aerobic reactor

To maintain thermophilic temperatures in the aerobic reactor of a dual digestion system requires an adequate supply of energy. With the generation of energy in modern society having an increasing impact on the global environment, due to both its demand on non-recoverable resources, and its production of thermal and chemical pollution, energy optimization is of profound importance. The generation of 'clean' energy with renewable biological sources enjoys a high level of attention among the environmentally conscious. It is an interesting exercise to examine the gearing of energy in the aerobic reactor at Milnerton ie the ratio of biologically generated energy to the electrical energy purchased from the supply network.

The maximum biological power output from aerobic stage at Milnerton was 73 kW (as low grade heat). This energy was liberated by oxidation reactions in the sludge consuming 20 kg/h of pure oxygen. To dissolve this oxygen in the sludge at an 83 % efficiency required 19 kW of electrical power and a supply of 24 kg/h of oxygen. Of this 19 kW about 17 kW was directly recovered as heat (see Appendix 3A). Thus it appears that the energy gearing was 4.7 ie 90 kW net heating power at the expense of a 19 kW input.

However the production of pure oxygen in liquid form typically consumes 1100 kWh of electrical energy per ton of product, which means that each kg of pure oxygen introduced (not dissolved) in the reactor sludge, was equivalent to 1.1 kWh electrical energy. The 24 kg/h oxygen supply therefore represents a further power input of 26 kW which is non-recoverable. More realistically then, 45 kW of electrical power is converted to a net 90 kW of heat ie a gearing of 2.0.

It is possible to develop this reasoning further, albeit at a more abstract level. The generation of electricity by an average coal-fired power station in South Africa is only about 35% efficient ie coal which represents say, 100 kWh of energy, results in only 35 kWh being delivered to the end user. Applying this to the above discussion suggests that the 45 kW of electrical power consumed actually represents a true external power base of 128 kW. This reduces the gearing to 0.7 and thus the Milnerton aerobic reactor was not a net producer of energy but was, in fact, only 70 % energy efficient in that 90 kW of power was generated at the expense of 128 kW of externally supplied power. This efficiency is peculiar to the Milnerton plant and some simple process modifications can alter it quite considerably.

For example, the supply of pure oxygen to the reactor may be derived from the on-site generation of gaseous oxygen as opposed to the bulk liquid supply used at Milnerton. A liquid oxygen supply represents a large waste of energy as the energy required to liquefy the gas is completely lost when the oxygen is vaporized. The generation of gaseous oxygen at a 95% purity can be achieved at the expense of approximately 300 kWh of energy (as opposed to 1100 kWh for liquid oxygen) using a pressure swing absorption plant (PSA) delivering 10 tons per day. Supplying the oxygen in this way to the Milnerton plant would lead to the consumption of 79 kW of power derived from non-renewable sources, for a return of 90 kW of heat ie a gearing of 1.14. As a dual digestion plant gets larger this efficiency will increase due to scale effects resulting in a smaller power requirements to achieve oxygen dissolution.

Any process modification which reduces the dependence of the aerobic reactor on biologically generated heat (and therefore on oxygen and dissolving energy) without substituting it with another primary non-renewable energy source, will improve the gearing further. Examples of such modifications are energy recovery using heat exchange and energy generation from the methane produced in the anaerobic stage. These aspects are discussed in a design example presented in Chapter 6.

5B.2 OPERATION OF THE ANAEROBIC DIGESTOR

As mentioned earlier the start up and operation of the anaerobic digestor is not covered in this thesis – these can be found in the project progress and final reports (De Villiers *et al*, 1987, 1988, 1989, 1990) – but two aspects peculiar to the Milnerton anaerobic digestor are presented. These are (1) gas pressure fluctuations in the headspace and (2) overheating.

5B.2.1 Gas pressure fluctuations in the anaerobic digestor headspace

The operating volume of the 1000 m^3 anaerobic digestor in the Milnerton dual digestion system varied between 400 and 600 m^3, resulting in a headspace volume of around 500 m^3. A 150 mm water seal was installed on the digestor roof in order to relieve a sudden increase in headspace pressure, or to break a sudden vacuum. In early stages of the project, the sludge transferred from the aerobic reactor was discharged into the digestor from a port in the roof, from where it cascaded through the headspace gas. Following the commissioning of the anaerobic digestor, a problem became apparent: whenever sludge was discharged into the digestor, the pressure in the headspace increased, causing the water seal to break. The pressure increase was caused by the hot reactor sludge (±60°C), falling through and heating up the cooler headspace gas. Although the increase in the headspace gas temperature was slight, it was sufficient to raise the headspace pressure by as much as 300 mmWg. The pressure increase posed no threat to the digestor structure but air was drawn into the digestor as the headspace gas cooled and contracted, and concern arose over the introduction of oxygen into the methane rich digestor gas.

The problem was corrected by extending a pipe from the sludge discharge port on the digestor roof to below the sludge surface, thereby limiting the transfer of heat from the hot reactor sludge to the headspace gas. A break had to be included in this pipe in order to prevent establishing a syphon between the aerobic reactor and the anaerobic digestor.

5B.2.2 Anaerobic digestor overheating

The heat required to maintain the anaerobic digestor of the dual digestion system at mesophilic temperatures was derived from the hot (±60° C) sludge transferred from the aerobic reactor. During the latter half of summer, when ambient temperatures frequently exceeded 30° C, the rate of heat loss from the digestor to the ambient surroundings was too small to cool the digestor contents sufficiently and consequently the anaerobic digestor overheated to as much as 44° C. In the absence of a sludge cooler or a reactor feed to effluent sludge heat exchanger, the only way of preventing this occurrence was to reduce the sludge flow to the digestor. This interfered with the research on the digestor as stable short retention times could not be maintained during summer. In an attempt to maintain as short as possible anaerobic retention times, the liquid volume in the digestor was reduced as the sludge flow to the digestor was reduced. However the volume could not be reduced below 380 m^3 without adversely affecting the mixing action of the submersible mixer. The topic of short anaerobic retention time is discussed further in Section 5B.3.3. below.

5B.3 OVERALL DUAL DIGESTION SYSTEM PERFORMANCE

Although the focus in this thesis was the evaluation of biological heat generation in the aerobic reactor, a summary of microbiological and chemical analyses of the aerobic reactor feed and transfer sludge and the anaerobic digestor effluent sludge is given in Tables 5B.1 and 5B.2 respectively. The analyses were performed by the Division of Water Technology (DWT) of the CSIR and are reported for the period May 1988 to November 1988. A summary of weekly anaerobic gas production figures for the same period are given in Table 5B.2. More detailed analyses can be obtained from the project progress and final reports (de Villiers *et al.*

In the introduction to this thesis (Chapter 1), a number of process advantages asserted by proponents of the dual digestion system were mentioned. These included claims that

(1) sludge is disinfected in the aerobic stage,
(2) the aerobic stage pretreated or 'conditioned' the sludge, enabling more stable operation at short retention times of the anaerobic stage than could be achieved in conventional mesophilic anaerobic digestion
(3) methane production in the anaerobic stage would not be compromised by sludge pretreatment in the aerobic stage.

The overall performance of the Milnerton dual digestion system in the context of the above claims is very briefly discussed below. This allows the reader to place the work on the aerobic reactor presented in this thesis in the context of the overall project. The brief discussion does not do justice to all the work done on the anaerobic digestor regarding disinfection, stabilization and dewaterability, and the reader is directed to the progress and final reports on the project for further information.

5B.3.1 Sludge disinfection in the Milnerton dual digestion system

The pasteurization specifications applied to the Milnerton dual digestion system required the sludge to be detained at a temperature greater than 55° C for at least 2 hours. These minimum specifications were exceeded at Milnerton with aerobic reactor temperatures ranging from 60.0 to 66.1° C and a batch cycle time of 2.5 hours.

TABLE 5B.1: Microbiological analysis of Dual Digestion sludge.

Date 1988	Faecal Coli (per g dry)			*Ascaris* ova per 10g dry					
	Feed	Reactor	Digestor	Feed		Reactor		Digestor	
				Viable	Non Viable	Viable	Non Viable	Viable	Non Viable
7/3	$8.01*10^{7}$	0	0	0	321	0	0	0	33
15/3	$1.19*10^{8}$	$7.86*10^{1}$	0	0	428	0	63	82	1224
22/3	$5.69*10^{7}$	0	$1.79*10^{2}$	0	108	0	31	0	143
28/3	$5.15*10^{7}$	0	$1.37*10^{4}$	0	368	0	158	0	24
11/4	$1.87*10^{8}$	$1.25*10^{4}$	$7.65*10^{4}$	246	197	0	100	0	167
19/4	$5.66*10^{9}$	$6.55*10^{3}$	$1.53*10^{4}$	122	244	0	98	0	175
25/4	$4.97*10^{10}$	0	$5.31*10^{3}$	403	151	0	427	0	170
2/5	$3.59*10^{9}$	0	$8.11*10^{4}$	301	0	0	311	0	349
9/5	$2.02*10^{11}$	$1.09*10^{4}$	$2.17*10^{4}$	31	217	0	146	0	359
17/5	$2.16*10^{12}$	$2.56*10^{5}$	$3.16*10^{3}$	102	256	0	81	0	101
7/6	$1.08*10^{12}$	0	$2.51*10^{3}$	196	65	0	675	0	806
15/6	$1.27*10^{11}$	$1.08*10^{3}$	$2.77*10^{3}$	0	53	0	163	0	350
20/6	$1.28*10^{13}$	$1.96*10^{3}$	$1.59*10^{4}$	40	80	0	116	0	267
28/6	$4.11*10^{12}$	$5.09*10^{3}$	$4.98*10^{2}$	164	817	0	735	0	697
5/7	$2.38*10^{12}$	$2.53*10^{2}$	$1.53*10^{5}$	203	254	35	348	57	848
19/7	$2.68*10^{12}$	$4.52*10^{4}$	$1.75*10^{4}$	228	913	0	730	0	947
3/8	$1.89*10^{12}$	$6.82*10^{3}$	$3.30*10^{4}$	102	153	0	210	0	251
17/8	$4.45*10^{12}$	$3.01*10^{4}$	$1.05*10^{4}$	210	105	0	588	0	600

The results of microbiological analyses for Faecal coliform and *Ascaris Ova* are reported in Table 5B.1 for the period 7 March 1988 to 17 August 1988. Salmonella were detected in neither the feed sludge to, nor the effluent sludge from the dual digestion system. Apparently Salmonella are inactivated by low pH values and as pH values of 5.3 were routinely measured for the Milnerton feed sludge, it is reasonable to expect no viable Salmonella in the feed sludge.

The results show a decline of about 10^9 in Faecal coliforms and, with the exception of 3 samples, complete inactivation of *Ascaris Ova.* Regrowth of Faecal coliforms can be seen to occur in the anaerobic digestor, from which it may be concluded that little information regarding the hygienic quality of a sludge can be derived from this type of analysis – substantial regrowth of Faecal coliforms is likely to occur on sludge drying beds and in sludge stockpiles. No explanation can be offered for the 3 counts of viable *Ascaris Ova.* Given that zero viable counts were consistently measured in the other 16 analyses, it is possible that the positive readings were caused by sample contamination.

5B.3.2 Sludge pretreatment in the aerobic stage

The data in Tables 5B.2 a), b) and c) clearly shows that the aerobic reactor did not stabilize the sludge, and that this was accomplished by the digestor. On average the dual digestion system removed 54% and 55% of the influent sludge VS and COD respectively of which 6% of the VS and 14% of the COD were removed by the reactor. Note that the 18 DWT data for reactor indicate higher VS removal than the 180 VS data measured by the writer (see Chapter 4, Section 4.10.1).

The pretreatment or 'conditioning' ability of the reactor is reflected by the data which shows, on average, increases in the following parameters:

pH – from 5.2 to 6.7
ammonium ion concentration – from 220 to 297 mgN/ℓ
alkalinity (titration to pH 5.75) – from 12 to 614 mg/ℓ
alkalinity (titration to pH 4.3) – from 733 to 892 mg/ℓ.

The aerobic reactor appeared to solubilize particulate material in the sludge as substantial amounts of colloidal material were observed in the reactor effluent (transfer) sludge. These colloids caused this sludge to dewater poorly, blinding

TABLE 5B.2(a): Chemical analyses of the feed sludge.

Date	TS (g/ℓ)	VS (g/ℓ)	NH_4-N (mg/ℓ)	COD (g/□)	pH	Alk 5.75 (mg/ℓ)	Alk 4.3 (mg/ℓ)
17/5	34.3	28.4	168	49.1	5.7	0	575
7/6	38.2	31.0	255	49.5	5.8	0	663
15/6	39.4	31.6	166	55.4	5.8	25	700
20/6	38.5	32.1	248	51.4	6.0	63	600
28/6	36.0	30.4	268	53.0	5.8	19	669
5/7	36.1	29.8	253	48.0	5.5	0	800
12/7	43.5	31.7	276	48.4	5.8	25	850
19/7	43.8	34.8	253	56.5	5.8	25	813
3/8	35.6	29.2	234	47.7	5.7	0	731
17/8	38.1	33.0	194	65.5	5.8	38	594
25/8	39.2	32.3	211	52.2	5.4	0	656
6/9	34.6	26.5	225	49.2	5.7	0	744
13/9	35.7	27.9	222	45.9	5.8	25	663
12/10	41.4	31.8	234	56.0	5.2	0	950
18/10	41.2	33.1	220	72.7	5.3	0	825
1/11	42.9	33.6	144	57.6	5.7	0	831
10/11	35.4	28.1	200	50.4	5.2	0	783
24/11	38.1	30.5	197	49.8	5.4	0	744

TABLE 5B.2(b): Chemical analyses of the aerobic digestor sludge.

Date	TS (g/ℓ)	VS (g/ℓ)	NH_4-N (mg/ℓ)	COD (g/□)	pH	Alk 5.75 (mg/ℓ)	Alk 4.3 (mg/ℓ)
17/5	34.6	28.9	196	46.6	6.7	444	713
7/6	33.5	26.4	238	38.7	7.1	475	675
15/6	36.9	28.9	237	49.5	6.9	1731	813
20/6	32.1	26.1	375	36.0	7.5	669	988
28/6	34.0	28.1	349	42.0	7.1	563	875
5/7	31.6	25.8	304	38.0	7.1	550	875
12/7	43.6	29.7	293	42.3	6.9	425	763
19/7	42.0	33.0	352	48.5	6.9	538	1075
3/8	38.4	31.2	277	45.6	7.0	300	463
17/8	31.7	26.2	220	45.7	6.9	450	825
25/8	35.9	29.6	206	50.2	6.7	325	738
6/9	35.2	27.6	259	39.5	7.3	644	1031
13/9	33.8	26.2	282	42.0	6.8	529	950
12/10	43.9	32.7	431	49.2	7.5	663	1019
18/10	40.3	31.2	451	49.8	7.1	750	1238
1/11	42.8	33.5	259	68.8	7.0	650	850
10/11	36.9	28.5	352	44.5	7.4	925	1263
24/11	36.9	29.5	270	49.4	7.0	425	900

TABLE 5B.2(c): Chemical analyses of the anaerobic digestor sludge.

Date	TS (g/ℓ)	VS (g/ℓ)	NH_4-N (mg/ℓ)	COD (g/□)	pH	Alk 5.75 (mg/ℓ)	Alk 4.3 (mg/ℓ)	Vol Acid (mg/ℓ)
17/5	23.4	16.6	512	28.5	7.8	2650	3350	317
7/6	22.3	15.8	599	25.5	7.5	2113	2563	285
15/6	22.0	16.2	513	25.8	7.4	1669	2219	192
20/6	20.8	15.2	622	24.1	7.9	1863	2375	192
28/6	20.1	14.9	690	24.2	7.7	2038	2569	294
5/7	18.8	13.4	639	22.0	7.5	1875	2400	390
12/7								
19/7	19.8	13.4	591	21.3	7.3	1813	2513	240
3/8	19.9	13.5	721	22.0	7.5	1513	1988	282
17/8	20.0	14.3	656	25.2	7.3	1963	2575	72
25/8	18.5	13.2	636	20.6	7.3	1900	2488	120
6/9	17.8	12.3	721	22.1	7.4	1919	2831	250
13/9	18.3	12.1	698	21.8	7.6	1913	2725	90
12/10	19.1	13.3	722	21.8	7.5	2113	2663	180
18/10	20.3	13.8	732	23.7	7.5	2088	2725	120
1/11	19.5	13.0	653	23.6	7.6	2250	2619	180
10/11	21.3	13.9	662	25.8	7.4	1850	2475	135
24/11	20.0	14.0	783	28.5	7.6	1488	2700	195

filter media and resisting centrifugation. For this reason it was not possible to perform suspended solids tests on the reactor effluent sludge and total solids were measured instead.

With regard to biological pretreatment making the sludge more readily anaerobically digestible, little can be said regarding observations on the sludge itself. This would best be tested by comparing the performances of two anaerobic digestors in operated in parallel, one with sludge pretreatment and the other without.

5B.3.3 Hydraulic retention time in the digestor

Although it was hoped to operate the digestor at a retention time of 10 – 12 days, this was prevented by digestor overheating problems (see Section 5B.2.2 above). Nevertheless, on one occasion in the winter of 1989, when an attempt was made to operate the digestor at a 12 day retention time, it caused a substantial decrease in the rate of digestor gas production and an increase in the volatile acids concentration. To obviate digestor failure, the retention time was again increased to 15 days. Although, on the basis of this observation, it appears that the desired retention time of 10 –12 days will not be achieved in the digestor, short retention operation will need to be investigated more thoroughly in further research.

5B.3.4 VS and COD reduction and gas production

Methane production by the digestor is summarized in Table 5B.3 in the form of weekly gas volumes. If the methane content of the digestor gas is assumed to 65% (this was measured on only one occasion and found to be 67%) then, at an average retention time of 21.4 days, the specific methane production in the anaerobic digestor amounted to 0.36 m^3 (STP)/(kg COD removed). If COD removal across the system is considered (not only across the digestor) then this figure decreases to 0.27 m^3 (STP)/(kg COD removed) [0.47 m^3 (STP)/(kg VS removed)].

Considering the gross heat of combustion of digestor gas (CO_2 and CH_4), combustion of the digestor gas production with air (at 100% efficiency) would have generated heat at a rate of 130 kW – a rate of heat supply of about 90 kW was required to maintain a sludge temperature of 60°C in the reactor at a 1 day retention.

With regard to the anaerobic digestor sludge quality, the final effluent sludge was very fine and thin, with the appearance of black ink. During dewatering on sludge beds the sludge neither attracted flies nor generated offensive odours – the sludge odour was characteristic of a well stabilized anaerobically digested sludge. Although it settled well, it did not exhibit good dewatering characteristics as the very fine sludge particles tended to blind sludge drying beds.

In spite of the high VS reduction in the sludge (about 55% across the dual digestion system), and its apparent stability, the sludge continued to generate gas after discharge to the sludge drying beds. The gas bubbles floated some of the sludge forming a 'crust' on the sludge surface, while the remainder of the sludge settled and blinded the drying bed media. Sandwiched between these two layers of sludge was a clear supernatant which could be decanted within a day of sludge discharge to the beds. In hot, dry weather, the practice of decanting this

TABLE 5B.3. ANAEROBIC DIGESTOR - WEEKLY OPERATING SUMMARY
30/4/1988 TO 2/12/1988

WEEK	MEAN DIGESTOR TEMP °C	FEED VOL (m^3/week)	MEAN DIGESTOR VOL (m^3)	MEAN RETENTION (days)	GAS PRODUCTION (m^3/week)	MEAN GAS TO SLUDGE RATIO
1	29.4	180	569	22	1593	8.85
2	31.1	139	662	33	2514	18.12
3	33.2	244	865	25	3435	14.09
4	33.5	251	897	25	3295	13.11
5	34	251	630	18	3187	12.68
6	34.7	244	488	14	2439	10.00
7	33.8	191	582	21	2234	11.68
8	35	251	533	15	2346	9.34
9	31.7	41	531	26	-	-
10	32.7	251	659	18	-	-
11	34.6	146	637	17	-	-
12	34.9	252	706	20	-	-
13	34.1	96	731	23	-	-
14	32.6	184	621	24	-	-
15	34.7	255	740	20	-	-
16	36.1	251	711	20	-	-
17	37.2	251	708	20	-	-
18	35.3	49	685	28	-	-
19	30.9	131	666	20	1740	12.57
20	30.5	109	694	25	1448	13.31
21	33.1	228	564	17	2000	8.77
22	-	0	420	-	263	-
23	28.8	56	423	30	575	10.23
24	31	105	442	29	1603	15.27
25	34.1	173	428	17	2674	15.5
26	37.8	248	436	12	3016	12.19
27	36.2	150	410	19	2566	17.10
28	37.2	180	413	16	2877	15.98
29	39.6	188	588	22	2730	14.56
30	37.8	128	432	24	1756	13.77

supernatant enabled the sludge to be manually removed from the beds 6 to 8 days after it had been discharged from the anaerobic digestor.

In conclusion, the dual digestion system holds great promise as a relatively simple and easy to operate sludge pasteurization and stabilization system. Certain aspects, principally the final sludge dewatering characteristics and the operation of the anaerobic digestor at short retention times need to be evaluated more thoroughly before the technical and economic viability of the system can be fully appraised.

APPENDIX 6A

LISTING OF THE SIMULATION PROGRAM ·ATASIM·

This appendix contains a listing of the source code for the program "ATASIM" which simulates the temperature versus time profile in a batch fed aerobic stage of a dual digestion system. The program is coded using Turbo Pascal Ver 4.0 and comprises 8 separately compiled units which appear below in the following order:

VERYABLE
TITLES
UTILITI
FILER
DESIGN
SIMULDAT
CHANGE
CALCULAT

Two additional units, Gkernel and Gprinter, are employed by the program to accomplish graphical simulation of a reactor temperature trace on the screen. These units form part of the graphics application package Turbo Graphix Toolbox and are not listed below. The main program ATASIM follows after the above 8 units.

UNIT VERYABLE.

```
UNIT veryable;

INTERFACE

USES
    DOS,
    CRT,
    Gdriver,
    Printer,
    Gkernel;


VAR
```

```
 label1,dlabel1                 :ARRAY [1..18] OF STRING[25];
label2,dlabel2                 :ARRAY [1..18] OF STRING[15];
values,dvalues                 :ARRAY [1..18] OF REAL;
helplines                      :ARRAY [1..70] OF STRING[54];
a                              :ARRAY [1..40] OF CHAR;
ops, sistim                    :STRING;
file_name, dummy_file_name     :STRING[30];
error_message                  :STRING[55];
newvalue                       :PACKED ARRAY [1..10] OF CHAR;
f                              :TEXT;
correct,air,balance,quit,
o2xs, steady_state, no_error,
file_saved, speed, mass_2_on,
escape                         :BOOLEAN;
move,run,finish,page,transfer,o,
fov, fnv, cfsv, fsv, flv       :CHAR;
posy,x,y,II,I,PP, marker,
phase_1, phase_2,
check, check2, check3 ,check4,
check5, system_choice,
increment, stored_increment       :INTEGER;

mole_o2_in, mole_o2_out, mole_n2_in, mole_n2_out,
mole_co2_out, mole_air_in, mole_h2o_in, mole_H2O_out,
mole_o2_used, mole_o2_used_max, o2_resp_max,
mass_N2_in, Os, Oc, Oc_max,
mass_h2o_in,
o2_pc_in, o2_pc_out, Oeff,
YCO2, rel_humid, hfg, log_pw, pw,
pp_dry_air, pp_dry_vent_gas,
mole_dry_air_in, mole_wet_air_in, mole_dry_vent_gas, mole_wet_vent_gas,
vol_dry_air_in, vol_dry_vent_gas, vol_O2_in,
air_temp_in, gas_temp_in, Tse, Tse_t, temp_diftl,
Tge, Tgi, z, o2_temp_in, Tsi, Tref,
reactor_temp, vent_temp,
gas_press_in, PT,
vol_air_in,
Hmech, Hwall,
Fb, Vb, Vp, Vp_i, Vp_t, batch_fract, t_c, t_tpr, t_fpr, retention_time,
sludge_flow, ps, Cp, aa, bb,
Hbi, Hmi, Hsi, Hse, Hwe, Hge, Hve, Mwe,
Hbi_out, Hmi_out, Hsi_out, Hse_out, Hwe_out, Hge_out,
Hve_out, Mwe_out,
Hbi_print, Hmi_print, Hse_print, Hwe_print,
Hge_print, Hve_print, Mwe_print,
Hs_i, Hs_t, delta_Hs,
xold,yold,xnew,ynew,delta_t,number,
sldg_ht, wall_ht, bio_ht, mech_ht, Y_H           :REAL;

IMPLEMENTATION

END.
```

UNIT TITLES.

```
UNIT titles;

INTERFACE
         USES
              DOS,
              CRT,
              Gdriver,
              Printer,
              Gkernel,
              utiliti,
              veryable;
         PROCEDURE title;


IMPLEMENTATION

{***************}
PROCEDURE TITLE;
{***************}

BEGIN
CLRSCR;

lowvideo;
FOR I := 1 TO 8 DO
 BEGIN
  GOTOXY(8,5+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(14,5+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(23,5+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(32,5+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(38,5+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(56,5+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(62,5+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(66,5+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(70,5+I);
  WRITELN(CHR(219),CHR(219));
 END;

FOR I := 1 TO 2 DO

BEGIN
GOTOXY(8,5+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(8,9+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(20,5+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(32,5+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(32,9+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(44,5+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(44,8+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(44,11+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(62,5+I);
```

```
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
END;

GOTOXY(44,8);WRITELN(CHR(219),CHR(219));
GOTOXY(50,11);WRITELN(CHR(219),CHR(219));

normvideo;
highvideo;
FOR I := 1 TO 8 DO
 BEGIN
  GOTOXY(10,4+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(16,4+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(25,4+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(34,4+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(40,4+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(58,4+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(64,4+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(68,4+I);
  WRITELN(CHR(219),CHR(219));
  GOTOXY(72,4+I);
  WRITELN(CHR(219),CHR(219));
 END;

FOR I := 1 TO 2 DO

BEGIN
GOTOXY(10,4+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(10,8+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(22,4+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(34,4+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(34,8+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(46,4+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(46,7+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(46,10+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
GOTOXY(64,4+I);
WRITELN(CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219),CHR(219));
END;

GOTOXY(46,7);WRITELN(CHR(219),CHR(219));
GOTOXY(52,10);WRITELN(CHR(219),CHR(219));

normvideo;

mark;
box(8,17,71,20,1);
unmark;

GOTOXY(9,18);
WRITELN('PROCESS DESIGN AND SIMULATION OF THE AUTOTHERMAL AEROBIC STAGE');
GOTOXY(9,19);
WRITELN('                 OF A DUAL DIGESTION SYSTEM');

GOTOXY(74,23);WRITELN('THESIS');
box(73,22,80,24,1);
```

```
GOTOXY(5,23); WRITELN('PRESS ANY KEY TO CONTINUE.....');
REPEAT UNTIL KEYPRESSED;
o:=READKEY;

CLRSCR;

GOTOXY(5,2);
WRITELN(' ','ATASIM  is  the  joint  property  of  AFROX  LIMITED  and  the');
WRITELN('     ','Department of Civil Engineering, University of Cape Town. This');
WRITELN('     ','software was designed, written and coded by  John Messenger as');
WRITELN('     ','part of a post graduate research program at the  University of');
WRITELN('     ','Cape Town.');
WRITELN(' ','');
WRITELN('     ','ATASIM may not be used or copied without the permission of:');
WRITELN(' ','');
WRITELN(' ','');
GOTOXY(5,10);
WRITELN(' ','Dept Civil Engineering             AFROX Market Development');
WRITELN('     ','University of Cape Town             P.O. Box 4100');
WRITELN('     ','Private Bag                         Germiston South');
WRITELN('     ','Rondebosch                    and   1411');
WRITELN('     ','7700                                South Africa');
WRITELN('     ','South Africa                        ATTN. Dr. D.Ossin');
WRITELN('     ','ATTN. Prof G.A.Ekama');
WRITELN('     ','');
WRITELN('     ','AFROX LIMITED, the Universtity of Cape Town and John Messenger');
WRITELN('     ','make no warranties,  either expressed or implied, with respect');
WRITELN('     ','to the quality and performance of this  software,   and do not');
WRITELN('     ','accept liability for direct, indirect,incidental or consequen-');
WRITELN('     ','tial damages resulting from the use of this software for the ');
WRITELN('     ','design and simulation of process plant.');
box(5,9,29,17,1);
box(40,9,65,16,1);
GOTOXY(20,24); WRITELN('PRESS ANY KEY TO CONTINUE...');
REPEAT UNTIL KEYPRESSED;
o:=READKEY;
CLRSCR;
END;

END.  {of UNIT titles}
```

UNIT UTILITI

```
UNIT UTILITI;

INTERFACE
   USES
        DOS,
        CRT,
        Gdriver,
        Printer,
        Gkernel,
        veryable;

   PROCEDURE MARK;
   PROCEDURE UNMARK;
   PROCEDURE BOX(x1,y1,x2,y2,a:INTEGER);
   PROCEDURE NOISE;
   PROCEDURE NUMCHECK;


IMPLEMENTATION


{**************}
PROCEDURE MARK;
{**************}

BEGIN
 TEXTBACKGROUND(WHITE);
 TEXTCOLOR(BLACK);
END;



{****************}
PROCEDURE UNMARK;
{****************}

BEGIN
 TEXTBACKGROUND(BLACK);
 TEXTCOLOR(WHITE);
END;



{**************************************}
PROCEDURE BOX(x1,y1,x2,y2,a:INTEGER);          { draws a frame around}
{**************************************}        { specified co-ordinates}

VAR
  II:INTEGER;

BEGIN

  a:=x2-x1-1;
  FOR II:=1 TO a DO          {draw the horizontal lines}
  BEGIN
    GOTOXY(x1+1,y1);
    WRITELN(CHR(196));
    GOTOXY(x1+1,y2);
    WRITELN(CHR(196));
    x1:=x1+1;
  END;

  x1:=x1-a;                 {restore x1}
  a:=y2-y1-1;
```

```
  FOR II:=1 TO a DO          {draw the vertical lines}
  BEGIN
    GOTOXY(x1,y1+1);
    WRITELN(CHR(179));
    GOTOXY(x2,y1+1);
    WRITELN(CHR(179));
    y1:=y1+1;
  END;

  y1:=y1-a;                 {restore y1}

GOTOXY(x1,y1);            {draw the the corners}
WRITELN(CHR(218));
GOTOXY(x1,y2);
WRITELN(CHR(192));
GOTOXY(x2,y1);
WRITELN(CHR(191));
GOTOXY(x2,y2);
WRITELN(CHR(217));

END;

{***************}
PROCEDURE noise;
{***************}

BEGIN
SOUND(800);DELAY(500);SOUND(650);DELAY(500);
SOUND(800);DELAY(500);SOUND(480);DELAY(400);
SOUND(800);DELAY(500);SOUND(650);DELAY(500);
SOUND(800);DELAY(500);SOUND(480);DELAY(400);
NOSOUND;
END;


{******************}
PROCEDURE numcheck;                  { check that the input is real }
{******************}

VAR
decpnt,blank,i                :INTEGER;
dec,divide                    :REAL;

BEGIN

 decpnt:=0;
 divide:=1;
 blank:=0;
 number:=0;
 dec:=0;
 correct:=true;


 FOR i:=1 TO 10 DO
  BEGIN
   IF newvalue[i]='.' THEN decpnt:=decpnt+1
   ELSE IF (ord(newvalue[i]) IN [48..57]) AND (decpnt=0) THEN      {determine non-decimal part}
                   number:=number*10+(ord(newvalue[i])-48)
   ELSE IF (ord(newvalue[i]) IN [48..57]) AND (decpnt>0) THEN      {determine decimal part}
                 BEGIN
                   dec:=dec*10+(ord(newvalue[i])-48);
                   divide :=divide*10;
                 END
   ELSE IF newvalue[i]=' ' THEN blank:=blank+1

   ELSE IF (ord(newvalue[i]) IN [13]) THEN i:=10

   ELSE correct:=false;
```

```
  END;

 IF blank=10 THEN correct:=false;
 IF correct=true THEN number:=number+dec/divide;          {combine the 2 halves of the number}

END; {of numcheck}


END.  {of UNIT}
```

```
{************************}
PROCEDURE check_file_save;  {checks whether the current file must}
{************************}                    {be saved}

BEGIN

IF NOT (file_saved) THEN
  BEGIN
       GOTOXY(15,21);
       WRITELN('THE CURRENT FILE HAS NOT BEEN SAVED - SAVE (Y/N) ?');
        REPEAT cfsv := READKEY
        UNTIL UPCASE(cfsv) IN ['Y','N',#27];
       IF UPCASE(cfsv) = #27 THEN escape := true;

     IF NOT escape THEN
       BEGIN

           GOTOXY(15,21);
           WRITELN('                                                   ');
           IF UPCASE(cfsv) ='Y' THEN file_saver;
       END;  {of escape IF}
   END;

END;  { of check_file_save }



{**********************}
PROCEDURE file_operator;  {checks whether a file must}
{**********************}                    {be saved or loaded}

BEGIN

menu;
escape := false;
dummy_file_name := file_name;
check5 := 0;
GOTOXY(25,21);
WRITELN('LOAD FILE  L   SAVE FILE  S ');
GOTOXY(36,21); mark; WRITELN('L');
GOTOXY(51,21); WRITELN('S'); unmark;
REPEAT fov := READKEY
UNTIL UPCASE(fov) IN ['L','S',#27];
IF UPCASE(fov) = #27 THEN escape := true;
IF NOT escape THEN
  BEGIN
       GOTOXY(25,21);
       WRITELN('                        ');
       IF UPCASE(fov) = 'L' THEN
         BEGIN
              check_file_save;
              IF NOT escape THEN file_loader;
         END
        ELSE
         BEGIN
            file_saver;
         END;

  END;  {of escape IF}

wiper;

END;  {of file_operator}
```

```
{*****************}
PROCEDURE warning;   {warns the user of a file error}
{*****************}

BEGIN

       no_error := false;
       SOUND(640);
       GOTOXY(15,21);
       WRITELN (error_message);
       DELAY(500);
       NOSOUND;
       DELAY(1000);
       GOTOXY(15,21);
       WRITELN('                                               ');
       file_namer;
       IF escape THEN no_error := true;

END;  {of warning}


{********************}
PROCEDURE file_namer;   {allows a file to be named by the user}
{********************}


BEGIN

dummy_file_name:='';
GOTOXY(27,21);
WRITELN('NAME THE FILE TO BE ',OPS);
GOTOXY(15,22);
WRITELN('(INCLUDE DRIVE AND PATH IF DEFAULT IS NOT ASSUMED)');
I:=1;
  REPEAT
     a[I]:= READKEY;
    IF NOT (a[I] IN [#13,'.']) THEN dummy_file_name := dummy_file_name
                                         + UPCASE (a[I]);
     GOTOXY(37+I,23);
     WRITELN (UPCASE(a[I]));
     I := I+1;
  UNTIL a[I-1] IN ['.',#13,#27];

IF a[I-1] = #27 THEN escape := true;
IF NOT escape THEN
  BEGIN
     dummy_file_name := dummy_file_name + '.SIM';
     GOTOXY(27,21);
     WRITELN('                             ');
     GOTOXY(15,22);
     WRITELN('                                               ');
     GOTOXY(15,23);
     WRITELN('                                               ');
   END;  {of IF}

END;  {of file_namer}


{********************}
PROCEDURE file_saver;   {saves the current file to disk}
{********************}

BEGIN
ops := 'SAVED:';
GOTOXY(25,21);
WRITELN('SAVE FILE AS ',dummy_file_name,' (Y/N) ?');
  REPEAT fsv := READKEY
  UNTIL UPCASE(fsv) IN ['Y','N',#27];
```

```
IF UPCASE(fsv) = #27 THEN escape := true;

IF NOT escape THEN
BEGIN

     GOTOXY(25,21);
     WRITELN('                                 ');
     IF UPCASE(fsv) = 'N' THEN file_namer;

        {$I-}
        REPEAT
        no_error := true;
        error_message :=
        'FILE OR DRIVE SPECIFICATION IS INCORRECT - TRY AGAIN';
        ASSIGN(f,dummy_file_name);
        REWRITE(f);
        IF IORESULT<>0 THEN warning;
        UNTIL no_error;
        {$I+}

IF NOT escape THEN
BEGIN

     WRITELN(f,system_choice);
     FOR I := 1 TO 18 DO
       BEGIN
            WRITELN(f,values[I]);
       END;

     CLOSE(f);
     file_saved := true;

END;  {of escape IF 1}
END;  {of escape IF 2}

END;  {of filesaver}


{*********************}
PROCEDURE file_loader;  {loads and reads a given file}
{*********************}

BEGIN

REPEAT
  ops := 'LOADED:';
  file_namer;

IF NOT escape THEN
  BEGIN

    REPEAT
      no_error := true;
      error_message :=
      'FILE OR DRIVE SPECIFICATION IS INCORRECT - TRY AGAIN';
      ASSIGN(f,dummy_file_name);
      {$I-}
      RESET(f);
      {$I+}
      IF IORESULT=0 THEN no_error := true ELSE warning;
    UNTIL no_error;

  END;  {of escape IF 1}

      IF NOT escape THEN READLN(f,check5) ELSE check5 := 101;
      error_message := 'FILE DATA IS INAPPROPRIATE FOR THIS PROGRAM';
      IF NOT (check5) IN [101,105] THEN warning;
```

```
UNTIL (check5 IN [101,105]);

IF NOT escape THEN
BEGIN
       system_choice := check5;
       IF system_choice = 101 THEN
         BEGIN
              air := true;
              sistim := 'AIR SIMULATION    ';
              FOR I := 1 TO 18 DO
                  BEGIN
                       READLN(f,values[I]);
                  END;
              file_name := dummy_file_name;
          END;

       IF system_choice = 105 THEN
         BEGIN
              air := false;
              sistim := 'OXYGEN SIMULATION';
              FOR I := 1 TO 14 DO
                  BEGIN
                       READLN(f,values[I]);
                  END;
              file_name := dummy_file_name;
         END;


CLOSE(f);

END;  {of escape IF 2}

END;   {of file_loader}

END.  {of unit}
```

UNIT DESIGN.

```
UNIT design;

INTERFACE
    USES
        DOS,
        CRT,
        Gdriver,
        Printer,
        Gkernel,
        utiliti,
        veryable;

    PROCEDURE deeinitvalues;
    PROCEDURE deehelp;
    PROCEDURE deecalculate;
    PROCEDURE sidemenu;
    PROCEDURE deerestore;
    PROCEDURE deedisplay;


IMPLEMENTATION

{***********************}
PROCEDURE deeinitvalues;          { initialise labels and variables }
{***********************}

BEGIN
 dlabel1[1]:='PROCESS TEMPERATURE......';
  dlabel2[1]:=' (C)        ';
   dvalues[1]:=60;
 dlabel1[2]:='SLUDGE FLOWRATE..........';
  dlabel2[2]:=' (M3/d)     ';
   dvalues[2]:=1000;
 dlabel1[3]:='RETENTION TIME...........';
  dlabel2[3]:='  (d)       ';
   dvalues[3]:=2.0;
 dlabel1[4]:='PROCESS VOLUME...........';
  dlabel2[4]:=' (m3)       ';
   dvalues[4]:=2000;
 dlabel1[5]:='RESPIRATION RATE.........';
  dlabel2[5]:=' (g/M3/h)   ';
   dvalues[5]:=229.3;
 dlabel1[6]:='FEED SLUDGE TEMPERATURE..';
  dlabel2[6]:=' (C)        ';
   dvalues[6]:=21;
 dlabel1[7]:='MECHANICAL HEAT INPUT....';
  dlabel2[7]:=' (kW)       ';
   dvalues[7]:=300;
 dlabel1[8]:='WALL HEAT LOSSES.........';
  dlabel2[8]:=' (kW)       ';
   dvalues[8]:=80;
 dlabel1[9]:='OXYGEN REQUIREMENTS......';
  dlabel2[9]:=' (KG/h)     ';
   dvalues[9]:=573.2;
 dlabel1[10]:='OXYGEN EFFICIENCY........';
  dlabel2[10]:=' (%)        ';
   dvalues[10]:=80;
 dlabel1[11]:='AIR REQUIREMENTS.........';
  dlabel2[11]:=' (M3(STP)/h)';
   dvalues[11]:=5095;
 dlabel1[12]:='AIR EFFICIENCY...........';
  dlabel2[12]:=' (%)        ';
   dvalues[12]:=30;

marker:=0;
```

```
check:=0;
check2:=0;
check4:=0;
transfer := ' ';

END;{of deeinitvalues }


{*****************}
PROCEDURE deehelp;
{*****************}


BEGIN
CLRSCR;

GOTOXY(1,5);
WRITELN('       ','The  ATASIM  design window contains operating parameters for an');
WRITELN('       ','ATAD plant which is at steady state at 60 C. With the exception');
WRITELN('       ','of  the  PROCESS TEMPERATURE,  all  of the  parameters  may  be');
WRITELN('       ','CHANGED in order to achieve a steady state design for your  own');
WRITELN('       ','data. To CHANGE a parameter,  first select it, then  ERASE  it,');
WRITELN('       ','and then type in your desired value and  ENTER  it.  Certain of');
WRITELN('       ','the parameters are linked and a CHANGE in one  of  these  will ');
WRITELN('       ','trigger a simultaneous change in the others. For example OXYGEN');
WRITELN('       ','REQUIREMENTS is linked to AIR REQUIREMENT and RESPIRATION RATE.');
WRITELN('     ','');
WRITELN('       ','By having CHANGED values of parameters in the window,  you will');
WRITELN('       ','have  "unbalanced"  the window  and it will no longer reflect a');
WRITELN('       ','steady state condition. To return it to steady state (1) SELECT');
WRITELN('       ',' a parameter which must change to restore steady state (2) MARK');
WRITELN('       ','the parameter.   It will automatically change to a steady state');
WRITELN('       ','value.  Note that certain parameters cannot be MARKED and these');
WRITELN('       ','are indicated by a (',chr(17),') symbol on the RHS of the window.');
box(5,3,73,23,1);
box(32,2,44,4,1);
GOTOXY(33,3);WRITELN('ATASIM HELP');
GOTOXY(30,23);WRITELN('Press any key for more help....');
REPEAT UNTIL KEYPRESSED;
o:=READKEY;
CLRSCR;
GOTOXY(1,5);
WRITELN('       ','If a  parameter  at  steady state is  MARKED,  then it will be');
WRITELN('       ','indicated on the LHS of the  window by  an  arrow.   CHANGING a');
WRITELN('       ','parameter  will  cause both the  arrow to disappear immediately');
WRITELN('       ','and the marked  parameter to revert to a  steady  state  value.');
WRITELN('       ','Such  a  marker  may  be  cancelled  by  SELECTING  the  MARKED');
WRITELN('       ','parameter and pressing the CANCEL key.');
WRITELN('');
WRITELN('       ','The design window offers an approximate design.This approximate');
WRITELN('       ','design can be "fine-tuned" in the  quick  and  full simulations');
WRITELN('       ','which follow the design window.  Upon leaving the design window');
WRITELN('       ','by opting to exit,  you will be offered the chance to  transfer');
WRITELN('       ','your design parameters through to the simulation window. Should');
WRITELN('       ','you not choose to transfer these parameters then the simulation');
WRITELN('       ','default values  will be  chosen.  Note that the simulation data');
WRITELN('       ','is much more  comprehensively  defined than the  design  window');
WRITELN('       ','data.');
box(5,3,73,22,1);
box(32,2,44,4,1);
GOTOXY(33,3);WRITELN('ATASIM HELP');
GOTOXY(30,22);WRITELN('Press any key to return.....');

REPEAT UNTIL KEYPRESSED;
o:=READKEY;
END;
```

```
{**********************}
PROCEDURE deecalculate;
{**********************}

BEGIN

CASE marker OF
    3: BEGIN
         bio_ht := dvalues[9] * dvalues[10]/100 * 13.1 * 1E6;
         mech_ht := dvalues[7] * 3.6 * 1E6;
         wall_ht := dvalues[8] * 3.6 * 1E6;
         sldg_ht := bio_ht + mech_ht - wall_ht;
         dvalues[3] := dvalues[4]/(sldg_ht*24) * 4.184 * 1E6 * (60-dvalues[6]);
       END;
    4: BEGIN
         bio_ht := dvalues[9] * dvalues[10]/100 * 13.1 * 1E6;
         mech_ht := dvalues[7] * 3.6 * 1E6;
         wall_ht := dvalues[8] * 3.6 * 1E6;
         sldg_ht := bio_ht + mech_ht - wall_ht;
         dvalues[4] := dvalues[3]*sldg_ht *24/ (4.184 * 1E6 * (60-dvalues[6]));
       END;
    6: BEGIN
         bio_ht := dvalues[9] * dvalues[10]/100 * 13.1 * 1E6;
         mech_ht := dvalues[7] * 3.6 * 1E6;
         wall_ht := dvalues[8] * 3.6 * 1E6;
         sldg_ht := bio_ht + mech_ht - wall_ht;
         dvalues[6] := 60 - sldg_ht*24 * dvalues[3]/(dvalues[4] * 4.184 * 1E6);
       END;
    7: BEGIN
         bio_ht := dvalues[9] * dvalues[10]/100 * 13.1 * 1E6;
         sldg_ht := dvalues[4]/dvalues[3] * 4.184 * 1E6 * (60-dvalues[6])/24;
         wall_ht := dvalues[8] * 3.6 * 1E6;
         mech_ht := sldg_ht + wall_ht - bio_ht;
         dvalues[7] := mech_ht/(3.6*1E6);
       END;
    8: BEGIN
         bio_ht := dvalues[9] * dvalues[10]/100 * 13.1 * 1E6;
         sldg_ht := dvalues[4]/dvalues[3] * 4.184 * 1E6 * (60-dvalues[6])/24;
         mech_ht := dvalues[7] * 3.6 * 1E6;
         wall_ht := bio_ht + mech_ht - sldg_ht;
         dvalues[8] := wall_ht/(3.6*1E6);
       END;
    9: BEGIN
         mech_ht := dvalues[7] * 3.6 * 1E6;
         wall_ht := dvalues[8] * 3.6 * 1E6;
         sldg_ht := dvalues[4]/dvalues[3] * 4.184 * 1E6 * (60-dvalues[6])/24;
         bio_ht := sldg_ht + wall_ht - mech_ht;
         dvalues[9] := bio_ht / (dvalues[10]/100 * 13.1 * 1E6);
         dvalues[11] := bio_ht / (0.21 * dvalues[12]/(100*22.4) * 32 *13.1
                       *1E6);
         dvalues[5] := dvalues[9]*1000*dvalues[10]/100/dvalues[4];
       END;
END;

marker:=0;
check:=0;
check2 :=0;

FOR I:=1 TO 12 DO               {write the table}
 BEGIN
  IF I=posy THEN mark;
    GOTOXY(29,9+I);
    WRITELN(dvalues[I]:6:1);
  IF I=posy THEN unmark;
    GOTOXY(1,9+I);
    WRITELN('  ');
 END;
```

```
END;


{******************}
PROCEDURE sidemenu;
{******************}

BEGIN

mark;
box(33,2,47,4,1);
unmark;
GOTOXY(34,3);
WRITELN('ATASIM DESIGN');

GOTOXY(54,12);WRITELN('  KEY       FUNCTION');
GOTOXY(54,14);WRITELN('  ',CHR(24),CHR(25),'    SELECT PARAMETER');
GOTOXY(54,15);WRITELN('<ENTER>    ERASE VALUE');
GOTOXY(54,16);WRITELN('<ENTER>  RETURN NEW DATA');
GOTOXY(54,17);WRITELN('   M     MARK PARAMETER');
GOTOXY(54,18);WRITELN('   C     CANCEL MARK');
GOTOXY(54,19);WRITELN('ctrl H       HELP');
GOTOXY(54,20);WRITELN('   D         EXIT');
box(53,11,79,13,1);
box(53,11,61,21,1);
box(53,11,79,21,1);
GOTOXY(48,10);WRITELN(chr(17));
GOTOXY(48,11);WRITELN(chr(17));
GOTOXY(48,14);WRITELN(chr(17));
GOTOXY(48,19);WRITELN(chr(17));
GOTOXY(48,21);WRITELN(chr(17));

END;


{********************}
PROCEDURE deerestore;              { display the data and values }
{********************}

BEGIN

CLRSCR;
mark;
box(3,9,47,22,1);             {draw the table borders}
unmark;

FOR I:=1 TO 12 DO             {write the table}
 BEGIN
  IF I=posy THEN mark;
  GOTOXY(4,9+I);
  WRITELN(dlabel1[I]);
    GOTOXY(29,9+I);
    WRITELN(dvalues[I]:6:1);
      GOTOXY(35,9+I);
      WRITELN(dlabel2[I]);
  IF I=posy THEN unmark;
 END;

sidemenu;

IF marker<>0 THEN
 BEGIN
   GOTOXY(1,9+check);
   WRITELN(chr(26));
 END;

END;
```

```
{*******************}
PROCEDURE deedisplay;                { display the data and values }
{*******************}

BEGIN

CLRSCR;
mark;
box(3,9,47,22,1);              {draw the table borders}
unmark;

FOR I:=1 TO 12 DO               {write the table}
 BEGIN
  IF I=2 THEN mark;
  GOTOXY(4,9+I);
  WRITELN(dlabel1[I]);
    GOTOXY(29,9+I);
    WRITELN(dvalues[I]:6:1);
      GOTOXY(35,9+I);
      WRITELN(dlabel2[I]);
  IF I=2 THEN unmark;
 END; {of FOR}

sidemenu;

posy:=2;
REPEAT
REPEAT move := READKEY UNTIL (UPCASE (move) IN ['M','C','D']) OR
                                               (ORD(move) IN [72,80,13,8]);
                      {allow data to be changed}

IF ORD(move) = 80 THEN              { select the data value to be changed }
 BEGIN
   GOTOXY(4,9+posy);
   WRITELN(dlabel1[posy]);
     GOTOXY(29,9+posy);
     WRITELN(dvalues[posy]:6:1);
       GOTOXY(35,9+posy);
       WRITELN(dlabel2[posy]);
 posy:=posy+1;
  IF posy=13 THEN posy:=2;
 GOTOXY(4,9+posy);
 mark;
 WRITELN(dlabel1[posy]);
   GOTOXY(29,9+posy);
   WRITELN(dvalues[posy]:6:1);
     GOTOXY(35,9+posy);
     WRITELN(dlabel2[posy]);
 unmark;
END;  {of move = 80}

IF ORD(move) = 72 THEN              { select the data value to be changed }
 BEGIN
   GOTOXY(4,9+posy);
   WRITELN(dlabel1[posy]);
     GOTOXY(29,9+posy);
     WRITELN(dvalues[posy]:6:1);
       GOTOXY(35,9+posy);
       WRITELN(dlabel2[posy]);
 posy:=posy-1;
  IF posy=1 THEN posy:=12;
 GOTOXY(4,9+posy);
 mark;
 WRITELN(dlabel1[posy]);
   GOTOXY(29,9+posy);
   WRITELN(dvalues[posy]:6:1);
     GOTOXY(35,9+posy);
     WRITELN(dlabel2[posy]);
```

```
 unmark;
END;  {of move = 72}


IF ORD(move) = 13 THEN                         { change selected data }
 BEGIN

  REPEAT
    GOTOXY(29,9+posy);
    mark;
    WRITELN('         ');
    GOTOXY(29,9+posy);
    {$I-}
    READ(number);
    {$I+}
  UNTIL IORESULT = 0;
  check2:=1;
  dvalues[posy]:=number;
  GOTOXY(29,9+posy);
  WRITELN(dvalues[posy]:6:1);
  unmark;
  CASE posy OF
     2: BEGIN
         dvalues[4]  := dvalues[3] * dvalues[2];
         dvalues[5]  := dvalues[9]*dvalues[10]/100*1000/dvalues[4];
         GOTOXY(29,9+4);WRITELN(dvalues[4]:6:1);
         GOTOXY(29,9+5);WRITELN(dvalues[5]:6:1);
        END;
     3: BEGIN
         dvalues[4]  := dvalues[3] * dvalues[2];
         dvalues[5]  := dvalues[9]*dvalues[10]/100*1000/dvalues[4];
         GOTOXY(29,9+4);WRITELN(dvalues[4]:6:1);
         GOTOXY(29,9+5);WRITELN(dvalues[5]:6:1);
        END;
     4: BEGIN
         dvalues[3]  := dvalues[4] / dvalues[2];
         GOTOXY(29,9+3);WRITELN(dvalues[3]:6:1);
        END;
     9: BEGIN
         dvalues[11]  := dvalues[9]*dvalues[10]/(dvalues[12]*32)*22.4/0.21;
         dvalues[5]  := dvalues[9]*dvalues[10]/100*1000/dvalues[4];
         GOTOXY(29,11+9);WRITELN(dvalues[11]:6:1);
         GOTOXY(29,5+9);WRITELN(dvalues[5]:6:1);
        END;
    11: BEGIN
         dvalues[9]  := dvalues[11]*0.21*dvalues[12]/dvalues[10]*32/22.4;
         dvalues[5]  := dvalues[9]*dvalues[10]/100*1000/dvalues[4];
         GOTOXY(29,9+9);WRITELN(dvalues[9]:6:1);
         GOTOXY(29,5+9);WRITELN(dvalues[5]:6:1);
        END;
    10: BEGIN
         dvalues[12]  := dvalues[9]*dvalues[10]*22.4/32/(dvalues[11]*0.21);
         dvalues[5]  := dvalues[9]*dvalues[10]/100*1000/dvalues[4];
         GOTOXY(29,12+9);WRITELN(dvalues[12]:6:1);
         GOTOXY(29,5+9);WRITELN(dvalues[5]:6:1);
        END;
    12: BEGIN
         dvalues[10]  := dvalues[11]*0.21*dvalues[12]/22.4*32/dvalues[9];
         dvalues[5]  := dvalues[9]*dvalues[10]/100*1000/dvalues[4];
         GOTOXY(29,10+9);WRITELN(dvalues[10]:6:1);
         GOTOXY(29,5+9);WRITELN(dvalues[5]:6:1);
       END;
  END;
END;  { of move = 13 }


IF ORD (move) = 8 THEN
 BEGIN
   deehelp;
```

```
   deerestore;
 END;  {of move = 8}
IF (UPCASE (move) = 'M') AND (marker=0) THEN          { mark changing data }
 BEGIN
  IF posy IN [3,4,6,7,8,9,11] THEN
   BEGIN
    unmark;
    GOTOXY(1,9+posy);
    WRITELN(chr(26));
    IF posy=11 THEN marker :=9 ELSE marker:=posy;
    check:=posy;
   END;  {of IF}
 END;  {of move = M}

IF (UPCASE (move) ='C') AND (posy=check) THEN       { cancel data marker }
 BEGIN
   GOTOXY(1,9+posy);
   WRITELN('  ');
   marker:=0;
   check:=0;
 END;  {of move = C}

IF (marker<>0) AND (check2<>0) THEN deecalculate;

IF (UPCASE (move) ='D') AND (check2>0) THEN
BEGIN
 GOTOXY(20,24);
 WRITELN('** RESTORE STEADY STATE (ie MARK) BEFORE EXITING **');
 noise;
 DELAY(600);
 GOTOXY(20,24);
 WRITELN('                                                  ');
END;  {of move = D}

IF (UPCASE (move) = 'D') AND (check2=0) THEN

BEGIN
GOTOXY(6,11);
WRITELN('                                                                  ');
GOTOXY(6,12);
WRITELN(' Do you wish to transfer this data to simulation window..(Y/N) ? ');
GOTOXY(6,13);
WRITELN('                                                                  ');
mark;
box(5,10,71,14,1);
unmark;
REPEAT transfer := READKEY UNTIL UPCASE(transfer) IN ['Y','N'];

IF UPCASE(transfer) = 'Y' THEN
BEGIN
  FOR I := 1 TO 12 DO
    BEGIN
     IF dvalues[I] < 0 THEN check4 := 1;
    END;
  IF check4 = 1 THEN
    BEGIN
     GOTOXY(6,12);
WRITELN(' Are you sure you wish to transfer negative numbers......(Y/N) ? ');
     REPEAT transfer := READKEY UNTIL UPCASE(transfer) IN ['Y','N'];
     IF UPCASE(transfer) = 'N' then
       BEGIN
        deerestore;
        transfer:= ' ';
        check4 := 0;
       END;
    END;
END;
END;
```

```
UNTIL UPCASE(transfer) IN ['Y','N'];
CLRSCR;

END; {of PROCEDURE}

END.
```

UNIT SIMULDAT.

```
UNIT simuldat;

INTERFACE

USES
      DOS,
      CRT,
      Gdriver,
      Printer,
      Gkernel,
      design,
      veryable,
      filer,
      utiliti;

PROCEDURE initlabels;
PROCEDURE air_initlabels;
PROCEDURE data_screen_menu;
PROCEDURE helplabels;
PROCEDURE help_select;
PROCEDURE help1;
PROCEDURE help2;
PROCEDURE help3;
PROCEDURE restore1;
PROCEDURE air_restore1;
PROCEDURE air_allocate_data;
PROCEDURE allocate_data;
PROCEDURE display;
PROCEDURE air_display;
PROCEDURE load_data;
PROCEDURE choice;
PROCEDURE reset_variables;


IMPLEMENTATION


{********************}
PROCEDURE initlabels;           { initialise labels and variables }
{********************}

BEGIN
 label1[1]:='PROCESS VOLUME........';
  label2[1]:=' (M3)   ';
 label1[2]:='BATCH FRACTION........';
  label2[2]:='        ';
 label1[3]:='CYCLE INTERVAL........';
  label2[3]:=' (HRS)  ';
 label1[4]:='FEED SLUDGE TEMP......';
  label2[4]:=' (C)    ';
 label1[5]:='TRANS PUMP RUN TIME...';
  label2[5]:=' (HRS)  ';
 label1[6]:='FEED PUMP  RUN TIME...';
  label2[6]:=' (HRS)  ';
 label1[7]:='PROC/VENT TEMP DIFF...';
  label2[7]:=' (C)    ';
 label1[8]:='OXYGEN FEED RATE......';
  label2[8]:=' (KG/HR)';
 label1[9]:='OXYGEN TEMPERATURE....';
  label2[9]:=' (C)    ';
 label1[10]:='% OXYGEN EFFICIENCY...';
  label2[10]:='        ';
 label1[11]:='VENT GAS PRESSURE.....';
  label2[11]:=' (kPag) ';
   values[11]:=3;
```

```
 label1[12]:='MECHANICAL HEAT IN....';
  label2[12]:=' {kW}   ';
 label1[13]:='SYSTEM HEAT LOSSES....';
  label2[13]:=' {kW}   ';
 label1[14]:='RESPIRATION RATE......';
  label2[14]:=' {g/M3} ';


END;{of initlabels }


{************************}
PROCEDURE air_initlabels;         { initialise labels and variables }
{************************}

BEGIN
 label1[1]:='PROCESS VOLUME........';
  label2[1]:=' {M3}   ';
 label1[2]:='BATCH FRACTION........';
  label2[2]:='        ';
 label1[3]:='CYCLE INTERVAL........';
  label2[3]:=' {HRS}  ';
 label1[4]:='FEED SLUDGE TEMP......';
  label2[4]:=' {C}    ';
 label1[5]:='TRANS PUMP RUN TIME...';
  label2[5]:=' {HRS}  ';
 label1[6]:='FEED PUMP  RUN TIME...';
  label2[6]:=' {HRS}  ';
 label1[7]:='PROC/VENT TEMP DIFF...';
  label2[7]:=' {C}    ';
 label1[8]:='AIR FEED RATE.........';
  label2[8]:=' {M3/HR}';
 label1[9]:='% OXYGEN IN FEED AIR..';
  label2[9]:=' {C}    ';
 label1[10]:='AIR DELIVERY PRESS....';
  label2[10]:=' {kPag} ';
 label1[11]:='DELIVERY AIR TEMP.... ';
  label2[11]:=' {C}    ';
 label1[12]:='FEED AIR TEMPERATURE..';
  label2[12]:=' {C}    ';
 label1[13]:='FEED AIR REL HUMIDITY.';
  label2[13]:=' {%}    ';
 label1[14]:='% OXYGEN EFFICIENCY...';
  label2[14]:='        ';
 label1[15]:='VENT GAS PRESSURE.....';
  label2[15]:=' {kPag} ';
 label1[16]:='MECHANICAL HEAT IN....';
  label2[16]:=' {kW}   ';
 label1[17]:='SYSTEM HEAT LOSSES....';
  label2[17]:=' {kW}   ';
 label1[18]:='RESPIRATION RATE......';
  label2[18]:=' {g/M3} ';


END;{of air_initlabels }



{********************}
PROCEDURE helplabels;
{********************}

BEGIN

CLRSCR;
GOTOXY(15,4);
helplines[1]:='The aerobic reactor is batch fed.  In other words';
helplines[2]:='at the start of a "cycle ",a measured volume of';
helplines[3]:='sludge is pumped out of the aerobic reactor by';
```

```
helplines[4]:='the TRANSFER PUMP,  following which exactly the';
helplines[5]:='same amount of feed sludge is pumped into the';
helplines[6]:='reactor by the FEED PUMP.  This measured amount';
helplines[7]:='of sludge divided by the process volume is the';
helplines[8]:='BATCH FRACTION.  The time that elapses between';
helplines[9]:='these "cycles" is the CYCLE INTERVAL.';
helplines[10]:='Generally a cycle interval should not be less than';
helplines[11]:='2 hours because of disinfection guidelines.';
helplines[12]:=' ';
helplines[13]:='Note that CYCLE INTERVAL/BATCH FRACTION = retention.';
helplines[14]:='Note also that the cycle interval includes the  ';
helplines[15]:='FEED PUMP RUN TIME and the TRANSFER PUMP RUN TIME';
helplines[16]:='and therefore the cycle interval should always be';
helplines[17]:='a minimum of 2 hrs + FEED PUMP RUN TIME + TRANS  ';
helplines[18]:='PUMP RUN TIME.';
helplines[19]:=' ';
helplines[20]:='PROC/VENT TEMP DIFF. This is the temperature';
helplines[21]:='difference between the vent gas leaving the reactor';
helplines[22]:='and the sludge in the reactor.  If this is comp-';
helplines[23]:='letely unknown then values of 3 to 7 should be';
helplines[24]:='acceptable.';
helplines[25]:='VENT GAS PRESSURE. The gauge pressure of the vent ';
helplines[26]:='gas as it leaves the reactor.';
helplines[27]:='% OXYGEN EFFICIENCY.  The % of the oxygen supplied';
helplines[28]:='which is consumed.';
helplines[29]:='MECHANICAL HEAT IN. The heat added to the process';
helplines[30]:='by non-biological means eg pumps, steam injection';
helplines[31]:='etc.  For instance, a recirculation pump using';
helplines[32]:='22 kW of power can add 19 kW of heat to the';
helplines[33]:='reactor.';
helplines[34]:='SYSTEM HEAT LOSSES. The heat lost through the walls';
helplines[35]:='of the reactor. It does not include sensible and';
helplines[36]:='latent heat lost in the vent gas.';
helplines[37]:='RESPIRATION RATE. The maximum uptake of oxygen by';
helplines[38]:='the sludge in the reactor.';
helplines[39]:='AIR FEED RATE. The volumetric flowrate of air feed  ';
helplines[40]:='into the reactor at the DELIVERY AIR PRESS (gauge)  ';
helplines[41]:='and the DELIVERY AIR TEMP.';
helplines[42]:=' ';
helplines[43]:='FEED AIR REL HUMIDITY. The relative humidity of the ';
helplines[44]:='air feed into the reactor as determined at the ';
helplines[45]:='FEED AIR TEMPERATURE ie prior to compression etc.';
helplines[46]:='  ';
helplines[47]:='% OXYGEN IN FEED AIR. This allows for using air ';
helplines[48]:='enriched with pure oxygen ';


END;{ of helplabels }



{************************}
PROCEDURE allocate_data;
{************************}

BEGIN

Vp := values[1];
batch_fract := values[2];
t_c := values[3];
Tsi := values[4];
t_tpr := values[5];
t_fpr := values[6];
temp_diftl := values[7];
Os := values[8];
Tgi := values[9];
Oeff := values[10];
PT := values[11];
Hmech := values[12];
```

```
Hwall := values[13];
o2_resp_max := values[14];

aa := 0.001012;  { variables a & b are used to calculate}
bb := 1.0;       { Hge and Hve in the heat balance}
Vb := batch_fract * Vp;
END;  {of allocate_data}



{****************************}
PROCEDURE air_allocate_data;  { transfers data from the arrays }
{****************************}          { to the program variables }

BEGIN

Vp := values[1];
batch_fract := values[2];
t_c := values[3];
Tsi := values[4];
t_tpr := values[5];
t_fpr := values[6];
temp_diftl := values[7];
vol_air_in := values[8];
o2_pc_in := values[9];
gas_press_in := values[10];
Tgi := values[11];
air_temp_in := values[12];
rel_humid := values[13];
Oeff := values[14];
PT := values[15];
Hmech := values[16];
Hwall := values[17];
o2_resp_max := values[18];

aa := 0.004372;  { variables a & b are used to calculate}
bb := 4.79;      { Hge and Hve in the heat balance}
Vb := batch_fract * Vp;
END;  {of air_allocate_data}


{***************}
PROCEDURE help1;
{***************}

BEGIN

CLRSCR;
box(12,2,66,23,1);

GOTOXY(14,4);

FOR I:= 1 TO 19 DO
 BEGIN
  WRITELN(helplines[I]);
  GOTOXY(14,4+I);
 END;
mark;
GOTOXY(14,11);
WRITELN('BATCH FRACTION');
GOTOXY(36,12);
WRITELN('CYCLE INTERVAL');
GOTOXY(14,18);
WRITELN('FEED PUMP RUN TIME');
GOTOXY(41,18);
WRITELN('TRANSFER PUMP RUN TIME');
GOTOXY(32,2);
WRITELN('ATASIM  HELP');
GOTOXY(32,23);
```

```
WRITELN('ATASIM  HELP');

unmark;


REPEAT UNTIL KEYPRESSED;
o:=READKEY;

END; { of help1 }

{***************}
PROCEDURE help2;
{***************}

BEGIN

CLRSCR;
box(12,2,66,23,1);

GOTOXY(14,4);

FOR I:= 20 TO 38 DO
 BEGIN
  WRITELN(helplines[I]);
  GOTOXY(14,I-15);
 END;

mark;
GOTOXY(14,4);
WRITELN('PROC/VENT TEMP DIFF');
GOTOXY(14,9);
WRITELN('VENT GAS PRESSURE');
GOTOXY(14,11);
WRITELN('% OXYGEN EFFICIENCY');
GOTOXY(14,13);
WRITELN('MECHANICAL HEAT IN');
GOTOXY(14,18);
WRITELN('SYSTEM HEAT LOSSES');
GOTOXY(14,21);
WRITELN('RESPIRATION RATE');

GOTOXY(32,2);
WRITELN('ATASIM  HELP');

GOTOXY(32,23);
WRITELN('ATASIM  HELP');
unmark;

REPEAT UNTIL KEYPRESSED;
o:=READKEY;

END;  { of help2 }


{***************}
PROCEDURE help3;
{***************}

BEGIN

CLRSCR;
box(12,5,66,18,1);

GOTOXY(14,7);

    FOR I:= 39 TO 48 DO
 BEGIN
  WRITELN(helplines[I]);
  GOTOXY(14,I-31);
```

```
 END;

mark;
GOTOXY(14,7);
WRITELN('AIR FEED RATE');
GOTOXY(38,8);
WRITELN('DELIVERY AIR PRESS');
GOTOXY(22,9);
WRITELN('DELIVERY AIR TEMP');
GOTOXY(14,11);
WRITELN('FEED AIR REL HUMIDITY');
GOTOXY(14,13);
WRITELN('FEED AIR TEMPERATURE');
GOTOXY(14,15);
WRITELN('% OXYGEN IN FEED AIR');

GOTOXY(32,5);
WRITELN('ATASIM  HELP');
GOTOXY(32,18);
WRITELN('ATASIM  HELP');
unmark;

REPEAT UNTIL KEYPRESSED;
o:=READKEY;

END; { of help3}


{*********************}
PROCEDURE help_select;
{*********************}

BEGIN

IF air THEN
  BEGIN
       CASE posy OF
         1,2,3,4,5,6: BEGIN
                        help1
                      END;

         7,14,15,16,17,18: BEGIN
                             help2
                           END;
         8,9,10,11,12,13: BEGIN
                            help3
                          END
       END;
END;   { of IF }

IF NOT air THEN
  BEGIN
       CASE posy OF
         1,2,3,4,5,6: BEGIN
                        help1
                      END;
         7,10,11,12,13,14: BEGIN
                             help2
                           END
       END;    { of CASE }

  END;    { of IF }

END;    { of help_select }


{***************************}
PROCEDURE data_screen_menu;
```

```
{***************************}

BEGIN

mark;
box(56,2,79,5,1);
box(56,5,79,9,1);
unmark;
GOTOXY(59,3);
WRITELN('ATASIM SIMULATION');
GOTOXY(59,4);
WRITELN('   DATA WINDOW   ');
GOTOXY(57,6);
WRITELN('FILE:');
GOTOXY(57,7);
WRITELN(file_name);
GOTOXY(57,8);
WRITELN(sistim);


GOTOXY(54,12);WRITELN('  KEY        FUNCTION');
GOTOXY(54,14);WRITELN('  ',CHR(24),CHR(25),'    SELECT PARAMETER');
GOTOXY(54,15);WRITELN('<ENTER>    ERASE VALUE');
GOTOXY(54,16);WRITELN('<ENTER>  RETURN NEW DATA');
GOTOXY(54,17);WRITELN('   K     QUICK SIMULATE');
GOTOXY(54,18);WRITELN('   D     FULL SIMULATE');
GOTOXY(54,19);WRITELN('ctrl H       HELP');
GOTOXY(54,20);WRITELN('   F     FILE OPERATIONS');
box(53,11,79,13,1);
box(53,11,61,21,1);
box(53,11,79,21,1);

END; {of PROCEDURE}


{******************}
PROCEDURE restore1;
{******************}

BEGIN

CLRSCR;
mark;
box(7,8,46,23,1);             {draw the table borders}
unmark;

FOR I:=1 TO 14 DO              {write the table}
 BEGIN
  GOTOXY(8,8+I);
  WRITELN(label1[I]);
    GOTOXY(30,8+I);
    WRITELN(values[I]:8:2);
      GOTOXY(38,8+I);
      WRITELN(label2[I]);
 END;

 GOTOXY(8,8+posy);
 mark;
 WRITELN(label1[posy]);
   GOTOXY(30,8+posy);
   WRITELN(values[posy]:8:2);
     GOTOXY(38,8+posy);
     WRITELN(label2[posy]);
 unmark;

 data_screen_menu;

 END; { of restore1 }
```

```
{**********************}
PROCEDURE air_restore1;
{**********************}

BEGIN

CLRSCR;
mark;
box(7,4,46,23,1);              {draw the table borders}
unmark;

FOR I:=1 TO 18 DO               {write the table}
 BEGIN
  GOTOXY(8,4+I);
  WRITELN(label1[I]);
    GOTOXY(30,4+I);
    WRITELN(values[I]:8:2);
      GOTOXY(38,4+I);
      WRITELN(label2[I]);
 END;

 GOTOXY(8,4+posy);
 mark;
 WRITELN(label1[posy]);
   GOTOXY(30,4+posy);
   WRITELN(values[posy]:8:2);
     GOTOXY(38,4+posy);
     WRITELN(label2[posy]);
 unmark;

 data_screen_menu;

 END; { of air_restore1 }


{*************************}
PROCEDURE reset_variables;      { set some program variables to zero }
{*************************}

BEGIN

mass_h2o_in := 0;
mole_n2_in := 0; mole_n2_out :=0;
vol_air_in := 0; air_temp_in := 0;

END;


{*******************}
PROCEDURE load_data;   {loads default file}
{*******************}

BEGIN

CLRSCR;
GOTOXY(25,10);
SOUND(700);
DELAY(300);
NOSOUND;
WRITELN('Reading default file........');
ASSIGN(f,file_name);
RESET(f);
READLN(f,system_choice);
FOR I := 1 TO 18 DO
   BEGIN
        READLN(f,values[I]);
   END;
CLOSE(f);
```

```
DELAY(2000);

IF (AIR) AND (UPCASE(transfer) = 'Y') THEN
 BEGIN
   values[1] := dvalues[4];
   values[3] := dvalues[3]*2;
   values[4] := dvalues[6];
   values[8] := dvalues[11]*(values[11]+273)/273*101.3/(101.3+values[10]);
   values[14] := (1-dvalues[12]/100)*21;
   values[16] := dvalues[7];
   values[17] := dvalues[8];
   file_name := 'noname.sim';
 END;

IF NOT (AIR) AND (UPCASE(transfer) = 'Y') THEN
 BEGIN
   values[1] := dvalues[4];
   values[3] := dvalues[3]*2;
   values[4] := dvalues[6];
   values[8] := dvalues[9];
   values[10] := (100-dvalues[10])/0.93;
   values[12] := dvalues[7];
   values[13] := dvalues[8];
   file_name := 'noname.sim';
 END;

balance := true;
mass_2_on := false;
speed := false;
quit    := false;
o2xs := false;
phase_1 := 0;
phase_2 := 0;
ps := 1.0;
Cp := 4.184;
tref := 0;
increment := 10;
stored_increment := 10;
YCO2 := 0.7;
Tse := 58;
xold := 0; yold := (Tse-30) * 17.5;
check3 := 0;
file_saved := true;
Hbi_out := 0; Hmi_out := 0; Hsi_out := 0; Hse_out := 0; Hge_out := 0;
Hwe_out := 0; Hve_out := 0; Mwe_out := 0;


END;  {of load_data}



{****************}
PROCEDURE choice;        { choose between air & oxygen systems }
{****************}

BEGIN

REPEAT
  air := false;
  CLRSCR;
  mark;
  box(23,8,51,10,1);
  box(23,8,51,12,1);
  unmark;
  GOTOXY(24,9);
  WRITELN('1.  AIR/ENRICHED AIR SYSTEM');
  GOTOXY(24,11);
  WRITELN('2.   PURE OXYGEN SYSTEM');
  GOTOXY(26,17);
```

```
  WRITELN('SELECT SYSTEM BY ENTERING');
  GOTOXY(28,18);
  WRITELN('THE RELEVANT NUMBER....');
  GOTOXY(51,18);
  REPEAT move := READKEY UNTIL move IN ['1','2'];
  IF move='1' THEN
    BEGIN
         air := true;
         sistim := 'AIR SIMULATION    ';
         air_initlabels;
         helplabels;
         system_choice := 101;
         file_name := 'AIRDEF.SIM';
    END
   ELSE
    BEGIN
         air := false;
         sistim := 'OXYGEN SIMULATION';
         initlabels;
         helplabels;
         system_choice := 105;
         file_name :='OXYDEF.SIM';
    END;
  CLRSCR;
  IF move='1' THEN
    BEGIN
         GOTOXY(19,8);
         WRITELN('THE AIR/ENRICHED AIR SYSTEM HAS BEEN SELECTED.');
    END
   ELSE
    BEGIN
         GOTOXY(21,8);
         WRITELN('THE PURE OXYGEN SYSTEM HAS BEEN SELECTED.');
    END;

  GOTOXY(17,10);
  WRITELN('DO YOU WISH TO CHANGE YOUR SELECTION ? (Y/N)....');
  GOTOXY(65,10);
  REPEAT move := READKEY UNTIL UPCASE(move) IN ['Y','N'];

UNTIL UPCASE(move) = 'N';

load_data;

IF NOT air THEN reset_variables;

END;  { of PROCEDURE }




{*****************}
PROCEDURE display;                  { display the data and values }
{*****************}

BEGIN

CLRSCR;
mark;
box(7,8,46,23,1);                 {draw the table borders}
unmark;

FOR I:=1 TO 14 DO                 {write the table}
 BEGIN
  IF I=1 THEN mark;
  GOTOXY(8,8+I);
  WRITELN(label1[I]);
```

```
    GOTOXY(30,8+I);
    WRITELN(values[I]:8:2);
      GOTOXY(38,8+I);
      WRITELN(label2[I]);
  IF I=1 THEN unmark;
 END;

data_screen_menu;

posy:=1;
REPEAT
REPEAT move := READKEY UNTIL (UPCASE(move) IN ['K','D','F']) OR
                                                (ORD(move) IN [8,72,80,13]);
                       {allow data to be changed}

IF ORD(move) = 80 THEN        { select the data value to be changed }
 BEGIN
   GOTOXY(8,8+posy);
   WRITELN(label1[posy]);
     GOTOXY(30,8+posy);
     WRITELN(values[posy]:8:2);
       GOTOXY(38,8+posy);
       WRITELN(label2[posy]);
 posy:=posy+1;
  IF posy=15 THEN posy:=1;
 GOTOXY(8,8+posy);
 mark;
 WRITELN(label1[posy]);
   GOTOXY(30,8+posy);
   WRITELN(values[posy]:8:2);
     GOTOXY(38,8+posy);
     WRITELN(label2[posy]);
 unmark;
END;

IF ORD(move) = 72 THEN        { select the data value to be changed }
 BEGIN
   GOTOXY(8,8+posy);
   WRITELN(label1[posy]);
     GOTOXY(30,8+posy);
     WRITELN(values[posy]:8:2);
       GOTOXY(38,8+posy);
       WRITELN(label2[posy]);
   posy:=posy-1;
   IF posy=0 THEN posy:=14;
   GOTOXY(8,8+posy);
   mark;
   WRITELN(label1[posy]);
     GOTOXY(30,8+posy);
     WRITELN(values[posy]:8:2);
       GOTOXY(38,8+posy);
       WRITELN(label2[posy]);
   unmark;
  END;

IF ORD(move) = 13 THEN                           { change selected data }
 BEGIN
   file_saved := false;

  REPEAT
    GOTOXY(30,8+posy);
    mark;
    WRITELN('        ');
    GOTOXY(30,8+posy);
    {$I-}
    READ(number)
    {$I+}
  UNTIL IORESULT = 0;
```

```
  values[posy]:=number;
  GOTOXY(30,8+posy);
  WRITELN(values[posy]:8:2);
  unmark;
 END;

IF ORD(move) = 8 THEN                  { change selected data }
 BEGIN
   help_select;
   restore1;
 END;

UNTIL UPCASE(move) IN ['K','D','F'];


IF UPCASE(move) = 'F' THEN
  BEGIN
       file_operator;
  END;

IF air THEN
  BEGIN
       air_allocate_data;
       air_initlabels;
  END
 ELSE
  BEGIN
       allocate_data;
       initlabels;
       reset_variables;
  END;

END;   {of display}


{********************}
PROCEDURE air_display;                 { display the data and values }
{********************}

BEGIN

CLRSCR;
mark;
box(7,4,46,23,1);              {draw the table borders}
unmark;

FOR I:=1 TO 18 DO               {write the table}
 BEGIN
  IF I=1 THEN mark;
  GOTOXY(8,4+I);
  WRITELN(label1[I]);
    GOTOXY(30,4+I);
    WRITELN(values[I]:8:2);
      GOTOXY(38,4+I);
      WRITELN(label2[I]);
  IF I=1 THEN unmark;
 END;

data_screen_menu;

posy:=1;
REPEAT
REPEAT move := READKEY UNTIL (UPCASE(move) IN ['K','D','F']) OR
                                                 (ORD(move) IN [8,72,80,13]);
                       {allow data to be changed}

IF ORD(move) = 80 THEN         { select the data value to be changed }
 BEGIN
   GOTOXY(8,4+posy);
```

```
   WRITELN(label1[posy]);
     GOTOXY(30,4+posy);
     WRITELN(values[posy]:8:2);
       GOTOXY(38,4+posy);
       WRITELN(label2[posy]);
 posy:=posy+1;
  IF posy=19 THEN posy:=1;
 GOTOXY(8,4+posy);
 mark;
 WRITELN(label1[posy]);
   GOTOXY(30,4+posy);
   WRITELN(values[posy]:8:2);
     GOTOXY(38,4+posy);
     WRITELN(label2[posy]);
 unmark;
END;

IF ORD(move) = 72 THEN         { select the data value to be changed }
 BEGIN
   GOTOXY(8,4+posy);
   WRITELN(label1[posy]);
     GOTOXY(30,4+posy);
     WRITELN(values[posy]:8:2);
       GOTOXY(38,4+posy);
       WRITELN(label2[posy]);
 posy:=posy-1;
  IF posy=0 THEN posy:=18;
 GOTOXY(8,4+posy);
 mark;
 WRITELN(label1[posy]);
   GOTOXY(30,4+posy);
   WRITELN(values[posy]:8:2);
     GOTOXY(38,4+posy);
     WRITELN(label2[posy]);
 unmark;
END;

IF ORD(move) = 13 THEN                            { change selected data }
 BEGIN
   file_saved := false;

  REPEAT
    GOTOXY(30,4+posy);
    mark;
    WRITELN('        ');
    GOTOXY(30,4+posy);
    {$I-}
    READ(number)
    {$I+}
  UNTIL IORESULT = 0;

  values[posy]:=number;
  GOTOXY(30,4+posy);
  WRITELN(values[posy]:8:2);
  unmark;
 END;

IF ORD(move) = 8 THEN                           { change selected data }
 BEGIN
      help_select;
      air_restore1;
 END;

UNTIL UPCASE(move) IN ['K','D','F'];


IF UPCASE(move) = 'F' THEN
  BEGIN
       file_operator;
```

```
  END;

IF air THEN
  BEGIN
       air_allocate_data;
       air_initlabels;
  END
 ELSE
  BEGIN
       allocate_data;
       initlabels;
  END;

END; { of air_display }

END.  {of unit}
```

UNIT CHANGE.

```
UNIT change;

INTERFACE

USES
     DOS,
     CRT,
     Gdriver,
     Printer,
     Gkernel,
     utiliti,
     simuldat,
     veryable;

PROCEDURE changedata;
PROCEDURE air_changedata;


IMPLEMENTATION


{********************}
PROCEDURE changedata;          { allows the program data to be changed }
{********************}

BEGIN

GOTOXY(1,24);
WRITELN('                                                                      ');
check3 :=0;

mark;
box(4,22,43,24,1);
unmark;

GOTOXY(5,18);WRITELN('  ',CHR(24),CHR(25),'    - SELECT   <ENTER> - RETURN');
GOTOXY(5,20);WRITELN('<ENTER> - CHANGE         D - ESCAPE');

mark;

posy:=1;
 GOTOXY(5,23);                 { write the initial data values and labels }
 WRITELN(label1[posy]);
   GOTOXY(27,23);
   WRITELN(values[posy]:8:2);
     GOTOXY(35,23);
     WRITELN(label2[posy]);
REPEAT
REPEAT move := READKEY UNTIL (UPCASE(move) ='D') OR (ORD(move) IN [72,80,13]);

IF ORD(move) = 80 THEN             { select the data value to be changed }
 BEGIN
   GOTOXY(5,23);
   WRITELN('                                     ');
 posy:=posy+1;
  IF posy=15 THEN posy:=1;
 GOTOXY(5,23);
 WRITELN(label1[posy]);
   GOTOXY(27,23);
   WRITELN(values[posy]:8:2);
     GOTOXY(35,23);
     WRITELN(label2[posy]);
END;

IF ORD(move) = 72 THEN             { select the data value to be changed }
 BEGIN
```

```
   GOTOXY(5,23);
   WRITELN('                                       ');
 posy:=posy-1;
  IF posy=0 THEN posy:=14;
 GOTOXY(5,23);
 WRITELN(label1[posy]);
   GOTOXY(27,23);
   WRITELN(values[posy]:8:2);
     GOTOXY(35,23);
     WRITELN(label2[posy]);
END;


IF ORD(move) = 13 THEN              { change selected data }
 BEGIN
   file_saved := false;

  REPEAT
    GOTOXY(27,23);
    WRITELN('        ');
    GOTOXY(27,23);
    FOR II:=1 to 10 DO
     BEGIN
       newvalue[II] := ' ';
     END;
    II := 0;
    REPEAT
    II:=II+1;
    REPEAT UNTIL KEYPRESSED;
    newvalue[II] := READKEY;
    GOTOXY(26+II,23);
    WRITELN(newvalue[II]);
    UNTIL ORD(newvalue[II]) = 13;
    numcheck;
  UNTIL correct=true;

  values[posy]:=number;
  GOTOXY(27,23);
  WRITELN(values[posy]:8:2);
 END;

UNTIL UPCASE(move) = 'D';

unmark;

GOTOXY(4,22);                                        { clear the data window }
WRITELN('                                        ');
GOTOXY(4,23);
WRITELN('                                        ');
GOTOXY(4,24);
WRITELN('                                        ');
GOTOXY(5,18);
WRITELN('                             ');
GOTOXY(5,20);
WRITELN('                             ');

allocate_data;

END; { of PROCEDURE }


{************************}
PROCEDURE air_changedata;          { allows the program data to be changed }
{************************}

BEGIN

GOTOXY(1,24);
WRITELN('                                                                        ');
```

```
check3 :=0;

mark;
box(4,22,43,24,1);
unmark;

GOTOXY(5,18);WRITELN('  ',CHR(24),CHR(25),'    - SELECT   <ENTER> - RETURN');
GOTOXY(5,20);WRITELN('<ENTER> - CHANGE        D - ESCAPE');

mark;

posy:=1;
 GOTOXY(5,23);                 { write the initial data values and labels }
 WRITELN(label1[posy]);
   GOTOXY(27,23);
   WRITELN(values[posy]:8:2);
     GOTOXY(35,23);
     WRITELN(label2[posy]);
REPEAT

REPEAT move := READKEY UNTIL (UPCASE(move) ='D') OR (ORD(move) IN [72,80,13]);

IF ORD(move) = 80 THEN           { select the data value to be changed }
 BEGIN
   GOTOXY(5,23);
   WRITELN('                   ');
 posy:=posy+1;
  IF posy=19 THEN posy:=1;
 GOTOXY(5,23);
 WRITELN(label1[posy]);
   GOTOXY(27,23);
   WRITELN(values[posy]:8:2);
     GOTOXY(35,23);
     WRITELN(label2[posy]);
END;

IF ORD(move) = 72 THEN           { select the data value to be changed }
 BEGIN
   GOTOXY(5,23);
   WRITELN('                   ');
 posy:=posy-1;
  IF posy=0 THEN posy:=18;
 GOTOXY(5,23);
 WRITELN(label1[posy]);
   GOTOXY(27,23);
   WRITELN(values[posy]:8:2);
     GOTOXY(35,23);
     WRITELN(label2[posy]);
END;

IF ORD(move) = 13 THEN           { change selected data }
 BEGIN
   file_saved := false;

  REPEAT
    GOTOXY(27,23);
    WRITELN('        ');
    GOTOXY(27,23);
    FOR II:=1 to 10 DO
     BEGIN
       newvalue[II] := ' ';
     END;
    II := 0;
    REPEAT
    II:=II+1;
    REPEAT UNTIL KEYPRESSED;
    newvalue[II] := READKEY;
    GOTOXY(26+II,23);
    WRITELN(newvalue[II]);
```

```
    UNTIL ORD(newvalue[II]) = 13;
    numcheck;
  UNTIL correct=true;

  values[posy]:=number;
  GOTOXY(27,23);
  WRITELN(values[posy]:8:2);
 END;

UNTIL move IN ['d','D'];

unmark;

GOTOXY(4,22);                                        { clear the data window }
WRITELN('                                         ');
GOTOXY(4,23);
WRITELN('                                         ');
GOTOXY(4,24);
WRITELN('                                         ');
GOTOXY(5,18);
WRITELN('                                    ');
GOTOXY(5,20);
WRITELN('                                    ');

air_allocate_data;

END;

END.
```

UNIT CALCULAT.

```
UNIT calculat;

INTERFACE


USES
     Dos,
     CRT,
     Gdriver,
     Printer,
     Gkernel,
     veryable,
     utiliti,
     filer,
     simuldat,
     change;


PROCEDURE steam(z:REAL);
PROCEDURE influent_gas_moles_and_mass;
PROCEDURE effluent_gas_moles_and_mass;
PROCEDURE check_oxygen_excess;
PROCEDURE outprint;
PROCEDURE heat_display_labels;
PROCEDURE mass_display;
PROCEDURE mass_display_2;
PROCEDURE clear_heat_display;
PROCEDURE key_menu(a,b:INTEGER);
PROCEDURE clear_key_menu(a,b:INTEGER);
PROCEDURE options;
PROCEDURE quick_sim_data;
PROCEDURE quick_sim;
PROCEDURE graph_paper;
PROCEDURE heat_display(a,b,c,d,e,f:REAL);
PROCEDURE output_accumulate;
PROCEDURE reset_output_accumulate;
PROCEDURE output_format;
PROCEDURE update_screen_windows;
PROCEDURE change_phase;
PROCEDURE heat_balance;
PROCEDURE new_temperature;

IMPLEMENTATION



PROCEDURE steam(z:real);
{************************************************************************
 This procedure calculates the vapour pressure and latent heat of
 water at temperature = z C, and delivers a result in kPa
 ***********************************************************************}

BEGIN
         { vapour pressure }
  log_pw := (-2238/(z+273) + 8.896) * ln(10);
  pw     := exp(log_pw);  {mm Hg}
      {now convert from mm Hg to kPa}
  pw     := pw * 101.3/760;  {kPa}

         { latent heat }
  hfg        := (-2.386*z + 2501)/1000;     {MJ/kg}

END;
```

```
PROCEDURE influent_gas_moles_and_mass; { not process temperature dependent }
{************************************************************
 This procedure calculates the molar and mass flowrates of
 oxygen, nitrogen and water in the feed air to the reactor
 ************************************************************}
BEGIN

IF air THEN
  BEGIN

    mole_wet_air_in :=
          vol_air_in/22.4*(gas_press_in+101.3)/101.3*273/(Tgi+273);

    steam(values[12]);  {get the vapour pressure of the water vapour
                             in the influent air}
    pp_dry_air  := 101.3 - pw;  { calculate the partial pressure
                  of the dry air assuming absolute air pressure = 101.3 kPa}
    mole_H2O_in := mole_wet_air_in/(pp_dry_air/pw + 1);
    mass_H2O_in := mole_H2O_in * 18;    {kg/h}

    mole_dry_air_in := mole_wet_air_in - mole_H2O_in;

    mole_O2_in  := O2_pc_in/100 * mole_dry_air_in;
    Os  := mole_O2_in * 32;      {kg/h}

    mole_N2_in  := mole_dry_air_in - mole_O2_in;
    mass_N2_in  := mole_N2_in * 28;      {kg/h}

    vol_dry_air_in := (mole_N2_in + mole_O2_in) * 22.4  {m3/h STP}


  END

    ELSE

  BEGIN
    mole_O2_in  := Os/32;
    vol_O2_in   := mole_O2_in * 22.4;
  END;

END; { of PROCEDURE }



PROCEDURE effluent_gas_moles_and_mass;  { not process temperature dependent }
{*******************************************************************
 This procedure calculates the molar and mass flowrates of
 oxygen, nitrogen and carbon dioxide in the vent gas.  It also
 checks whether the oxygen consumptions calculated from the
 OUR data and from the oxygen utilization efficiency data, compare.
 *******************************************************************}

BEGIN
   { calculate the maximum mass of oxygen that the reactor is able to
     consume - this is governed by the repiration rate of the sludge }

    { kg.mole }       { g/m3 }     { m3 }    { kg/kg.mole * g/kg }
mole_O2_used_max := O2_resp_max * Vp / 32000;
Oc_max := mole_O2_used_max * 32;

{ PERFORM THE CALCULATIONS FOR AN AIR OPERATED REACTOR }

  IF air THEN
    BEGIN
      mole_N2_out  := mole_N2_in;
      mole_O2_used := mole_O2_in * Oeff/100;

      IF mole_O2_used > mole_O2_used_max THEN
        BEGIN
```

```
          O2xs         := true;   { according to the data entered, the
                          reactor will consume more oxygen then the
                          sludge respiration rate will allow - this
                          cannot occur }
          mole_O2_used := mole_O2_used_max;  { set the oxygen
                          consumption to the maximum allowed by the
                          respiration rate }
                          { then recalculate the oxygen utilization
                           efficiency}
          Oeff := mole_O2_used/mole_O2_in * 100;
          Oc := Oc_max;
          mole_CO2_out := YCO2 * mole_O2_used;
          mole_O2_out  := mole_O2_in - mole_O2_used;
        END

           ELSE

        BEGIN
          O2xs         := false;
          mole_O2_out  := mole_O2_in - mole_O2_used;
          Oc := mole_O2_used*32;
          mole_CO2_out := YCO2 *mole_O2_used;
        END;

    mole_dry_vent_gas := mole_N2_out + mole_CO2_out + mole_O2_out;
    O2_pc_out         := 100 * mole_O2_out / mole_dry_vent_gas;

    END

    { PERFORM THE CALCULATIONS FOR AN OXYGEN OPERATED REACTOR }

      ELSE

    BEGIN
      mole_O2_used := mole_O2_in * Oeff/100;  {kg/h}
      IF mole_O2_used > mole_O2_used_max THEN
        BEGIN
          O2xs              := true;
          mole_O2_used      := mole_O2_used_max;
          Oeff := mole_O2_used/mole_O2_in * 100;
          Oc        := Oc_max;
          mole_CO2_out      := YCO2 * mole_O2_used;
          mole_O2_out       := mole_O2_in - mole_O2_used;
        END

          ELSE

        BEGIN
          O2xs              := false;
          mole_O2_out       := mole_O2_in - mole_O2_used;
          Oc        := mole_O2_used*32;
          Oeff := mole_O2_used/mole_O2_in * 100;
          mole_CO2_out      := YCO2 * mole_O2_used;
        END;

    mole_dry_vent_gas := mole_CO2_out + mole_O2_out;
    O2_pc_out         := 100 * mole_O2_out / mole_dry_vent_gas;

    END;

vol_dry_vent_gas := mole_dry_vent_gas * 22.4;     {m3/h STP}
Y_H := 14.245 - 1.648 * mole_O2_used/mole_O2_used_max;

END; { of PROCEDURE }


PROCEDURE check_oxygen_excess;
{************************************************************
 This procedure checks the variable O2xs - see PROCEDURE
```

```
 effluent_gas_moles_and_mass - and either displays a message
 at, or clears that message from, the base of the screen.
 **********************************************************}

BEGIN
  IF (O2xs = true) AND (check3 = 0) THEN
    BEGIN
      noise;
      GOTOXY(1,24);
      WRITELN
 ('** MAX O2 UTILISATION - INCREASE IN O2 SUPPLY WILL NOT INCREASE TEMP **');
      check3 :=1;
    END;
  IF (O2xs = false) AND (check3 = 1) THEN
    BEGIN
      GOTOXY(1,24);
      CLREOL;
      check3 :=0;
    END;
END;  {of PROCEDURE}


PROCEDURE OUTPRINT;
{**********************************************************************
 This procedure prints out the steady state simulation data when the
 "Z" option is selected from the options menu on the simulation screen.
 **********************************************************************}

BEGIN

{$I-}

vent_temp := reactor_temp - temp_diftl;
retention_time := t_c/batch_fract/24;
sludge_flow := Vp/retention_time;

REPEAT
no_error := true;
WRITELN
(LST,'     THE STEADY STATE DATA FOR SIMULATION OF A COMPLETELY MIXED, BATCH FED');
WRITELN(LST,'                   AEROBIC STAGE OF A DUAL DIGESTION SYSTEM');
WRITELN(LST,'');
WRITELN(LST,'                   THE DATA FILE USED IS ',file_name);
WRITELN
(LST,'     ********************************************************************');
WRITELN(LST,'');
WRITELN(LST,'PROCESS VOLUME = ',Vp:6:2,' M3');
WRITELN(LST,'   SLUDGE FLOW = ',sludge_flow:6:2,' M3/h');
WRITELN(LST,'RETENTION TIME = ',retention_time:6:2,' days');
WRITELN(LST,'    ........................................');
WRITELN(LST,'FEED SLUDGE TEMPERATURE = ',Tsi:4:2,' C');
WRITELN(LST,'    REACTOR TEMPERATURE = ',reactor_temp:4:2,' C');
WRITELN(LST,'   VENT GAS TEMPERATURE = ',vent_temp:4:2,' C');
WRITELN(LST,'    ........................................');
IF air THEN WRITELN(LST,'  SUPPLY AIR VOL AT ',Tgi:4:2,' C  AND ',
            gas_press_in:4:1,' kPag = ',vol_air_in:5:1,' M3/h' );

IF air THEN WRITELN(LST,'SUPPLY AIR TEMPERATURE BEFORE COMPRESSION = ',
            air_temp_in:5:1,' C');
IF air THEN WRITELN(LST,' SUPPLY AIR TEMPERATURE AFTER COMPRESSION = ',
            Tgi:5:1,' C');
IF air THEN WRITELN(LST,'                SUPPLY DRY AIR VOL AT STP =',
            vol_dry_air_in:5:1);
IF air THEN WRITELN(LST,'    ........................................');
IF NOT air THEN WRITELN(LST,'SUPPLY OXYGEN VOL AT STP =',vol_O2_in:5:1);
WRITELN(LST,'      MASS OF OXYGEN IN = ',Os:6:1,' kg/h');
WRITELN(LST,'    MASS OF OXYGEN USED = ',Oc:6:1,' kg/h');
WRITELN(LST,'      OXYGEN EFFICIENCY = ',Oeff:6:1,' %');
WRITELN(LST,'   % OXYGEN IN VENT GAS = ',O2_pc_out:6:1,' %');
```

```
WRITELN(LST,'    .......................................');
WRITELN(LST,'     VOL OF DRY VENT GAS AT STP =',vol_dry_vent_gas:5:1,' M3/h' );
WRITELN(LST,'           RESPIRATION QUOTIENT = ',YCO2:4:3);
WRITELN(LST,'  MASS OF INFLUENT WATER VAPOUR = ',mass_H2O_in:5:1,' kg/h');
WRITELN(LST,'  MASS OF EFFLUENT WATER VAPOUR = ',Mwe_print:5:1,' kg/h');
IF air THEN WRITELN(LST,'% RELATIVE HUMIDITY OF FEED AIR = ',
            rel_humid:5:1,' %');
WRITELN(LST,'    .......................................');
WRITELN(LST,'   GAS SENSIBLE HEAT LOSS = ',Hge_print:6:1,' MJ/h');
WRITELN(LST,'     GAS VAPOUR HEAT LOSS = ',Hve_print:6:1,' MJ/h');
WRITELN(LST,'SLUDGE SENSIBLE HEAT LOSS = ',Hse_print:6:1,' MJ/h');
WRITELN(LST,'           WALL HEAT LOSS = ',Hwe_print:6:1,' MJ/h');
WRITELN(LST,'   BIOLOGICAL HEAT SOURCE = ',Hbi_print:6:1,' MJ/h');
WRITELN(LST,'   MECHANICAL HEAT SOURCE = ',Hmi_print:6:1,' MJ/h');

{$I+}

IF IORESULT<>0 THEN
  BEGIN
       no_error := false;
       GOTOXY(5,24);
     WRITELN('** PRINTER ERROR - <Esc> TO CANCEL - ANY KEY TO CONTINUE **');
       noise;
       REPEAT UNTIL KEYPRESSED;
       o:=READKEY;
       IF o = #27 THEN no_error := true;
       GOTOXY(5,24);
     WRITELN('                                                           ');
  END;

UNTIL no_error = true;

END; {of outprint}


PROCEDURE heat_display_labels;
{*****************************************************
 This procedure writes the labels for the heat balance
 displayed on the lower RHS of the simulation screen.
 *****************************************************}

BEGIN
box(46,17,78,23,1);
GOTOXY(49,18);WRITELN('GAS  ');GOTOXY(49,19);WRITELN('VAP  ');
GOTOXY(49,20);WRITELN('WALL ');GOTOXY(49,21);WRITELN('SLDG ');
GOTOXY(63,18);WRITELN('MECH ');GOTOXY(63,20);WRITELN('BIO  ');
GOTOXY(49,22);WRITELN('TOTAL');GOTOXY(63,22);WRITELN('TOTAL');
mark;
GOTOXY(47,19);WRITELN('O');
GOTOXY(47,20);WRITELN('U');
GOTOXY(47,21);WRITELN('T');
GOTOXY(61,18);WRITELN('M');
GOTOXY(61,19);WRITELN('J');
GOTOXY(61,20);WRITELN('/');
GOTOXY(61,21);WRITELN('H');
GOTOXY(61,22);WRITELN('R');
GOTOXY(75,19);WRITELN('I');
GOTOXY(75,20);WRITELN('N');
unmark;


END;


PROCEDURE mass_display;
{***************************************************************
 This procedure displays the mass balance on the lower RHS of
 the screen if the "E" option is selected from the options menu.
 ***************************************************************}
```

```
BEGIN
box(46,17,78,22,1);
GOTOXY(49,18);WRITELN('MASS O2 IN    ',Os:6:1,' {kg/h}');
GOTOXY(49,19);WRITELN('MASS O2 USED  ',Oc:6:1,' {kg/h}');
GOTOXY(49,20);WRITELN('O2 EFFICIENCY ',Oeff:6:1,' {%}');
GOTOXY(49,21);WRITELN('MAX O2 AVAIL. ',Oc_max:6:1,' {kg/h}');

END;  {of mass_display}


PROCEDURE mass_display_2;
{*******************************************************************
 This procedure displays the alternate mass balance on the lower RHS
 of the screen if the "G" option is selected from the options menu.
 *******************************************************************}


BEGIN
box(46,17,78,22,1);
retention_time := t_c/batch_fract/24;
sludge_flow := Vp/retention_time;
GOTOXY(49,18);WRITELN('SLUDGE FLOW   ',sludge_flow:6:1,' {m3/d}');
GOTOXY(49,19);WRITELN('RETENTION     ',retention_time:6:1,' {d}');
GOTOXY(49,20);WRITELN('%O2 IN VENT   ',O2_pc_out:6:1,' {%}');
GOTOXY(49,21);WRITELN('SPEC HT YLD   ',Y_H:4:1,' {MJ/kg}');

END;  {of mass_display}


{*****************************}
PROCEDURE clear_heat_display;
{*****************************}


BEGIN

GOTOXY(46,17);WRITELN('                                  ');
GOTOXY(46,18);WRITELN('                                  ');
GOTOXY(46,19);WRITELN('                                  ');
GOTOXY(46,20);WRITELN('                                  ');
GOTOXY(46,21);WRITELN('                                  ');
GOTOXY(46,22);WRITELN('                                  ');
GOTOXY(46,23);WRITELN('                                  ');


END;




PROCEDURE key_menu(a,b:INTEGER);
{******************************************
 This procedure displays the options menu
 on the lower LHS of the simulation screen.
 ******************************************}

BEGIN

mark;
GOTOXY(a,b);  WRITELN('H - HARDCOPY   W - WAIT   E - MASS ON');
GOTOXY(a,b+1);WRITELN('Z - DATACOPY   Q - QUIT   G - MASS 2 ');
GOTOXY(a,b+2);WRITELN('C - CONTINUE   D - DATA   B - HEAT ON ');
GOTOXY(a,b+3);WRITELN('S - SPEED      F - FILE              ');
unmark;
box(a-1,b-1,a+38,b+4,1);
box(a+13,b-1,a+24,b+4,1);

GOTOXY(29,2); WRITELN(sistim);
```

```
GOTOXY(50,2); WRITELN(file_name);


END;



PROCEDURE clear_key_menu(a,b:INTEGER);
{*****************************************************
 This procedure clears the options menu on the screen.
 *****************************************************}

BEGIN

GOTOXY(a-1,b-1);WRITELN('                                        ');
GOTOXY(a-1,b);WRITELN('                                        ');
GOTOXY(a-1,b+1);WRITELN('                                        ');
GOTOXY(a-1,b+2);WRITELN('                                        ');
GOTOXY(a-1,b+3);WRITELN('                                        ');
GOTOXY(a-1,b+4);WRITELN('                                        ');

END;



PROCEDURE options;
{**********************************************
 This procedure executes the options selected
 from the options menu in the simulation screen.
 **********************************************}

BEGIN

CASE UPCASE (run) OF

{ change the simulation data }

   'D':  BEGIN
            clear_key_menu(4,19);
            IF air THEN air_changedata
                   ELSE changedata;
            influent_gas_moles_and_mass;
            effluent_gas_moles_and_mass;
            key_menu(4,19);
         END;

{ display a heat balance on the lower RHS of the simulation screen }

   'B':  BEGIN
              clear_heat_display;
              balance := true;
              mass_2_on := false;
              heat_display_labels;
         END;

{ display a mass balance on the lower RHS of the screen }

   'E':  BEGIN
              clear_heat_display;
              balance := false;
              mass_2_on := false;
         END;
{ print out the steady state simulation data }

   'Z':  BEGIN
              IF steady_state THEN outprint
              ELSE
                  BEGIN
                      GOTOXY(1,24);
```

```
                      WRITELN('** SYSTEM NOT AT STEADY STATE **');
                      noise;
                      GOTOXY(1,24);
                      WRITELN('                                ');
                  END;
         END;

{ print a copy of the entire screen  }

   'H':  BEGIN
              hardcopy(false,6);
         END;

{ stop execution of the program until a key is pressed }

   'W':  BEGIN
              REPEAT UNTIL KEYPRESSED;
              o:=READKEY;
         END;

{ call up the file system to save or load a simulation file }

   'F':  BEGIN
              file_operator;
              IF air THEN
                         BEGIN
                              air_initlabels;
                              air_allocate_data;
                         END
                       ELSE
                         BEGIN
                              initlabels;
                              allocate_data;
                              reset_variables;
                         END;
              IF balance THEN heat_display_labels ELSE mass_display;
              influent_gas_moles_and_mass;
              effluent_gas_moles_and_mass;
              key_menu(4,19);
         END;

{ continue the heat balance without updating the screen plot
  until steady state is achieved }

   'S':  BEGIN
              speed := true;
              increment := 3;
         END;

{ display an alternative mass balance on the lower RHS of the screen }

   'G':  BEGIN
              clear_heat_display;
              balance := false;
              mass_2_on := true;
              mass_display_2;
         END;

   'Q':  BEGIN
              quit := true;
         END

END;

END;


PROCEDURE quick_sim_data;
{*****************************************************
```

```
 This procedure displays useful data on the screen
 during the continuous flow simulation.
 *************************************************}

BEGIN

retention_time := t_c/batch_fract/24;
sludge_flow := Vp/retention_time;
GOTOXY(22,2);WRITELN('CONTINUOUS FLOW STEADY STATE SIMULATION OF');
GOTOXY(22,3);WRITELN('      DUAL DIGESTION AEROBIC STAGE');
GOTOXY(22,4);WRITELN('******************************************');

GOTOXY(5,6);WRITELN('Sludge flowrate = ',sludge_flow:5:1,' m3/d');
GOTOXY(5,7);WRITELN('Hydraulic retention = ',retention_time:6:2,' d');
GOTOXY(5,8);WRITELN('Reactor volume = ',Vp:5:1,' m3');
GOTOXY(45,6);WRITELN('Oxygen supply = ',Os:5:1,' kg/h');
GOTOXY(45,7);WRITELN('Oxygen consumption = ',Oc:5:1,' kg/h');
GOTOXY(45,8);WRITELN('Oxygen efficiency = ',Oeff:4:1,' %');
GOTOXY(20,9);WRITELN('% Oxygen in the vent gas = ',O2_pc_out:4:1,' %');
GOTOXY(5,11);WRITELN('Bio  ht = ',Hbi:5:1,' MJ/h');
GOTOXY(5,12);WRITELN('Mech ht = ',Hmi:5:1,' MJ/h');
GOTOXY(45,11);WRITELN('Sldg ht = ',Hse:5:1,' MJ/h');
GOTOXY(45,12);WRITELN('Wall ht = ',Hwe:5:1,' MJ/h');
GOTOXY(45,13);WRITELN('Gas  ht = ',Hge:5:1,' MJ/h');
GOTOXY(45,14);WRITELN('Vap  ht = ',Hve:5:1,' MJ/h');

END;  { of PROCEDURE }


PROCEDURE quick_sim;
{************************************************************************
 This procedure allows a continuous flow steady state simulation of the
 heat balance in order to determine the steady state reactor temperature.
 ***********************************************************************}
var    great,small   :BOOLEAN;
       I             :INTEGER;
       Tcorr         :REAL;

BEGIN

CLRSCR;

YCO2 := 0.7;
Tse := 58;
Tcorr := 10;
great := false;
small := false;
influent_gas_moles_and_mass;
effluent_gas_moles_and_mass;
check_oxygen_excess;
box(25,16,55,18,1);
I:=0;

REPEAT
  I:=I+1;
  Tge := Tse - temp_diftl;
  Hbi := Y_H * Oc;
  Hmi := Hmech * 3.6;  {convert kW to MJ/h}
  Hse := 1/t_c * batch_fract * Vp * ps * Cp * (Tse - Tsi);
  Hwe := Hwall * 3.6;   {convert kW to MJ/h}
  Hge := aa * Oc/(Oeff/100) * (Tge - Tgi);
  steam(Tge);  {calculate the partial pressure of water ie pw}
  Mwe := 18 * 0.03125 * Oc/(Oeff/100)
         * (bb - Oeff/100 + Oeff/100 * Yco2) * pw/((PT+101.3)-pw);
  Hve := hfg *  (Mwe - mass_H2O_in);

  {now calculate the change in reactor sludge enthalpy}

  delta_Hs :=  (Hbi + Hmi - Hse - Hwe - Hge - Hve);
```

```
  IF (delta_Hs >0) AND (great = false) THEN
  BEGIN
    Tcorr := ABS(Tcorr/10);
    great := true;
    small := false;
  END;
  IF (delta_Hs <0) AND (small = false) THEN
  BEGIN
    Tcorr := -Tcorr/10;
    small := true;
    great := false;
  END;

  Tse := Tse + Tcorr;
  GOTOXY(26,17);WRITELN('PROCESS TEMPERATURE = ',Tse:6:4);
UNTIL (ABS(Tcorr)<0.001) OR (I=1000) OR (delta_Hs = 0);

IF I = 1000 THEN
  BEGIN
       Tse := 58;
       SOUND(660);DELAY(400);NOSOUND;
       GOTOXY(15,20);
    WRITELN('THE QUICK SIMULATION WILL NOT CONVERGE - USE FULL SIMULATION');
  END
  ELSE
  yold := (Tse-30)*17.5;

quick_sim_data;
GOTOXY(15,22);WRITELN('Press any key to return to data window....');
REPEAT UNTIL KEYPRESSED;
o:=READKEY;

END;  { of quick_sim }



PROCEDURE graph_paper;
{************************************************************
 This procedure generates a new set of screen axes whenever
 the temperature trace has reached the RHS of the screen.
 ************************************************************}

BEGIN

  xnew := 0; xold := 0;

  FOR y := 1 TO 16 DO
     BEGIN
          GOTOXY(1,y);
          CLREOL;
     END;


  DRAWBORDER;


  FOR y:=1 TO 19 DO
   BEGIN
    FOR x:=1 TO 123 DO
     BEGIN
      DRAWPOINT(8*x,2*y*17.5);
     END;
   END;

  FOR y:=1 TO 3 DO
   BEGIN
     DRAWLINE(0,175*y,984,175*y);
   END;
```

```
  y:=3;x:=70;
  WHILE y<>18 DO
   BEGIN
     GOTOXY(1,y);
      WRITELN(x);
     y:=y+3;x:=x-10;
   END;

 GOTOXY(4,2);
 WRITELN('TEMPERATURE =      {C}');
 GOTOXY(29,2); WRITELN(sistim);
 GOTOXY(50,2); WRITELN(file_name);

 END;



 PROCEDURE draw;
 {**************************************************************
  This procedure plots the reactor temperature to the screen
  graphics display unless the speed option is exercised.
  **************************************************************}

 BEGIN

 IF NOT speed THEN
   BEGIN

     IF KEYPRESSED THEN
                    BEGIN
                         run := READKEY;
                         options;
                    END;

     ynew:= (Tse - 30) * 17.5;
     xnew:= xnew + delta_t * 41;
     DRAWLINE(xold,yold,xnew,ynew);
     xold:=xnew;yold:=ynew;

   END;

 END;    {of draw}



 PROCEDURE heat_display(a,b,c,d,e,f:REAL);
 {******************************************************************
  This procedure displays the heat balance on the lower RHS of
  the screen if the "B" option is selected from the options menu.
  ******************************************************************}

 VAR
 g,h                         :REAL;

 BEGIN
 box(46,17,78,23,1);
 g:=a+b+c+d;h:=e+f;
 GOTOXY(54,18);WRITELN(a:6:1);GOTOXY(54,19);WRITELN(b:6:1);
 GOTOXY(54,20);WRITELN(c:6:1);GOTOXY(54,21);WRITELN(d:6:1);
 GOTOXY(68,18);WRITELN(e:6:1);GOTOXY(68,20);WRITELN(f:6:1);
 GOTOXY(54,22);WRITELN(g:6:1);GOTOXY(68,22);WRITELN(h:6:1);

 END;
```

```
PROCEDURE output_accumulate;
{*****************************************************
 This procedure calculates cumulative totals of
 the heat balance data for a complete batch cycle.
 These totals are used in PROCEDURE output_format.
*****************************************************}


BEGIN

Hbi_out := Hbi_out + Hbi * delta_t;
Hmi_out := Hmi_out + Hmi * delta_t;
Hsi_out := Hsi_out + Hsi * delta_t;
Hse_out := Hse_out + Hse * delta_t;
Hwe_out := Hwe_out + Hwe * delta_t;
Hge_out := Hge_out + Hge * delta_t;
Hve_out := Hve_out + Hve * delta_t;
Mwe_out := Mwe_out + Mwe * delta_t;

END ; {of output_accumulate}

PROCEDURE reset_output_accumulate;
{*******************************************************
 This procedure resets the cumulative heat balance
 totals at the end of a complete batch cycle.
*******************************************************}

BEGIN

Hbi_out := 0;
Hmi_out := 0;
Hsi_out := 0;
Hse_out := 0;
Hwe_out := 0;
Hge_out := 0;
Hve_out := 0;
Mwe_out := 0;

END;  {of PROCEDURE reset_output_accumulate}


PROCEDURE output_format;
{***************************************************
 This procedure calculates the heat balance
 data on an hourly basis over a complete batch
 cycle so that it may be displayed on screen
 or sent to the printer.
***************************************************}


BEGIN

Hbi_print := Hbi_out/t_c;
Hmi_print := Hmi_out/t_c;
Hse_print := (Hse_out - Hsi_out)/t_c;
Hwe_print := Hwe_out/t_c;
Hge_print := Hge_out/t_c;
Hve_print := Hve_out/t_c;
Mwe_print := Mwe_out/t_c;

{now check to see if the reactor is at steady state}

IF ABS(Hbi_print + Hmi_print - Hse_print - Hwe_print -
       Hge_print - Hve_print) < 0.1 THEN
   BEGIN
     steady_state := true;
     IF speed THEN
        BEGIN
          speed := false;
```

```
          increment := stored_increment;
        END;
   END

   ELSE
   steady_state := false;

END;  {of PROCEDURE output_format}



PROCEDURE update_screen_windows;
{**************************************************
 This procedure calls the procedures that write
 heat and mass balance data to the screen.
**************************************************}


BEGIN

IF balance THEN
   BEGIN
     heat_display_labels;
     heat_display(Hge_print, Hve_print, Hwe_print, Hse_print, Hmi_print,
                  Hbi_print);

   END

   ELSE

IF mass_2_on THEN mass_display_2

   ELSE
   mass_display;

END;   {of PROCEDURE update_screen_windows}



PROCEDURE change_phase;
{**********************************************
 This procedure terminates a phase of the
 batch cycle and commences a new phase. The
 integers phase_1 and phase_2 are set to
 ensure that slugde sensible heat enters and
 leaves the reactor at the appropriate times
**********************************************}

BEGIN


IF (phase_1 = 0) AND (phase_2 = 0) THEN
   BEGIN            {commence phase 1}
     output_format;  {prepare heat balance data for the screen}
     update_screen_windows;  {write heat or mass balance data to the screen}
     reset_output_accumulate;
     phase_1 := 1;     {terminate phase 3 and commence phase 1}
     delta_t := t_tpr/increment;
     Fb := -Vb/Vp * Vp/t_tpr;    {this is the sludge volume change factor}
     Hs_i := Vp * ps * Cp * (Tse - Tref);
     Vp_i := Vp;
     check_oxygen_excess;
   END

   ELSE

IF phase_1 = 1 THEN
```

```
   BEGIN
     phase_1 := 0;       {terminate phase 1}
     phase_2 := 1;       {commence phase 2}
     delta_t := t_fpr/increment;
     Fb := Vb/Vp * Vp/t_fpr;   {this is the sludge volume change factor}
     GOTOXY(18,2); WRITELN(Tse:5:2);    {print the sludge temperature}
                                        {at the top of the screen}
     reactor_temp := Tse;   {this sets the value of reactor sludge
                             temperature for a data print out}
   END

   ELSE

IF phase_2 = 1 THEN
   BEGIN
     phase_2 := 0;       {terminate phase 2 and commence phase 3}
     delta_t := (t_c - t_tpr - t_fpr)/increment;
     Fb := 0;   {this is the sludge volume change factor}
     Vp_i := Vp;     {restore process volume to compensate for any slight
                      incremental error}
   END;


END; {of PROCEDURE change_phase}



PROCEDURE heat_balance;
{*********************************************
This procedure performs the heat balance in
order to determine the change in reactor
sludge enthalpy over the time interval delta_t.
*********************************************}



BEGIN

Tge := Tse - temp_diftl;
Hbi := Y_H * Oc;
Hmi := Hmech * 3.6;  {convert kW to MJ/h}
Hsi := (Vb/t_fpr) * ps * Cp * (Tsi - Tref) * phase_2;
Hse := (Vb/t_tpr) * ps * Cp * (Tse - Tref) * phase_1;
Hwe := Hwall * 3.6;   {convert kW to MJ/h}
Hge := aa * Oc/(Oeff/100) * (Tge - Tgi);
steam(Tge);  {calculate the partial pressure of water ie pw}
Mwe := 18 * 0.03125 * Oc/(Oeff/100)
      * (bb - Oeff/100 + Oeff/100 * Yco2) * pw/((PT+101.3)-pw);
Hve := hfg *  (Mwe - mass_H2O_in);

{now calculate the change in reactor sludge enthalpy}

delta_Hs := delta_t * (Hbi + Hmi + Hsi - Hse - Hwe - Hge - Hve);

END;


PROCEDURE new_temperature;
{*********************************************************************
This procedure controls both the simulation itself and the display
of the simulation.  It performs heat balances for the transfer, feed and
heating cycle.  The heat balances are performed over a discrete interval
of time determined by the variable "increment".  After the heat balance
is performed for 1 increment, the new process temperature is calculated
and plotted to screen by PROC "draw".  This procedure also allows the
simulation data to be changed, the screen display of results/data to be
changed and file operations through the PROC "options".
 *********************************************************************}
```

```
BEGIN

change_phase;     {select the phase of the batch cycle}

FOR PP := 1 TO increment DO   {now perform the heat balance and calculate}
                             {and plot the temperatures for the phase}
   BEGIN
     IF xnew >= 984 THEN graph_paper;  {the temperature trace has reached
                                        the end of the screen - get new screen}
     heat_balance;              {get the change in reactor sludge enthalpy}
     output_accumulate;    {allow the heat balance data for output to screen and
                          printer to be accumulated}


     Hs_t := Hs_i + delta_Hs;  {calculate the new reactor sludge enthalpy}
     Vp_t := Vp_i + Fb * delta_t;         {calculate the new reactor volume}
     Tse_t := Hs_t/(Vp_t * ps * Cp) + Tref;      {calc the new sludge temp}
     IF Tse_t > 69.8 THEN Tse_t := 69.8;  {ie don't allow the temp trace to
                                           leave the screen}
     Hs_i := Hs_t; Vp_i := Vp_t; Tse := Tse_t;
     draw;    {update the screen temperature trace}

   END;


END; {of PROCEDURE new_temperature}

END.  {of UNIT}
```

MAIN PROGRAM.

```
PROGRAM ATASIM;

USES
    DOS,
    CRT,
    Gdriver,
    Printer,
    Gkernel,
    utiliti,
    titles,
    design,
    simuldat,
    veryable,
    calculat;

BEGIN
title;
initgraphic;
leavegraphic;
REPEAT
CLRSCR;

deeinitvalues;
deedisplay;
choice;

IF air THEN air_display ELSE display;
   REPEAT
         WHILE UPCASE(move) IN ['F','K'] DO
             BEGIN
                  IF UPCASE(move) IN ['K'] THEN quick_sim;
                  IF air THEN air_display ELSE display;
             END;
   UNTIL UPCASE(move) IN ['D'];

influent_gas_moles_and_mass;
effluent_gas_moles_and_mass;

CLRSCR;

entergraphic;
DEFINEWORLD(1,0,700,984,0);
DEFINEWINDOW(1,3,30,76,212);
SELECTWORLD(1);
SELECTWINDOW(1);

graph_paper;

key_menu(4,19);

REPEAT
      new_temperature;
UNTIL quit = true;

leavegraphic;

CLRSCR;
GOTOXY(20,12);
WRITELN('Do you wish to restart ATASIM....(Y/N) ?');
REPEAT finish := READKEY UNTIL UPCASE(finish) IN ['Y','N'];

UNTIL UPCASE(finish) ='N';

END.
```