Evaluation of the dual digestion system: Part 3: Considerations in the process design of the aerobic reactor[1]

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Abstract

In this paper, design considerations for pure oxygen and air oxygenated aerobic reactors in dual digestion are outlined based on the steady state heat balance. Accepting that the objectives of the aerobic reactor are sludge pretreatment through oxygen limitation and pasteurisation, it is demonstrated that 3 parameters are of crucial importance to design which centres around minimising the reactor retention time: (1) the oxygen consumption rate of the sludge (OCR_{bio}) which fixes the maximum biological heat generation rate; (2) the oxygen transfer rate (OTR) of the oxygenation system which should be less than OCR_{bio} to ensure oxygen limitation; and (3) the oxygen transfer efficiency of the oxygenation system (OTE) which determines the vent gas volumetric flow rate at a given OTR. 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 is lost via hot effluent sludge thereby allowing pasteurisation temperatures to be attained at short retention times; if OTE is low (which usually is 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 to attain pasteurisation temperatures. However, with heat exchange between reactor effluent and influent sludge flows and with increased mechanical heat input, heat sources are increased without increasing heat losses in the vent gas and pro rata reductions in retention time are allowed.

Nomenclature and abbreviations

(For those not listed hereunder, see Messenger et al., 1993a on p. 185-191 in this issue)

A = wall and pipework heat loss surface area (m^2).
a = vent gas mass to moles multiplying factor for air, oxygen enriched air and pure oxygen systems.
f_{mo_2} = mass fraction of oxygen in influent gas.
H'_{se}, H'_{ge} = net sensible heat loss rate between influent and effluent sludge and vent gas flow rates (i.e. H_{se}-H_{si} and H_{ge}-H_{gi}) respectively (MJ/h).
H'_{ve} = total vent gas heat loss rate (MJ/h).
= sum of the net sensible H'_{ge} and water vapour (H_{ve}) heat loss rates.
h_{fg} = latent heat of vaporisation of water (MJ/kg).
M_{we} = vent gas water vapour mass flow rate (kg/h).
M_{dg} = dry vent gas molar flow rate (kmoles/h).
P_T = total atmospheric (barometric) pressure (mmHg).
p = partial pressure (mmHg). Subscripts w and dg denote saturation water vapour and dry gas respectively.
Q_s = influent and effluent sludge flow rate (m^3/d).
T_{diff} = difference between reactor sludge and exit vent gas temperatures (°C) (= T_{se}-T_{ge}).

Design and simulation of the aerobic reactor

Aerobic reactor design and simulation procedures were derived from the results of the Milnerton aerobic reactor performance (Messenger et al., 1992;1993a). These are founded on fundamental heat and mass balance principles and therefore they are general and apply to reactors oxygenated with air, oxygen enriched air or pure oxygen.

The design procedure is based on the solution of the steady state heat balance over the reactor. Such a heat balance yields a constant temperature for the reactor sludge and is applicable only to reactors that are continuously fed. In practice, however, to avoid recontamination of pasteurised sludge, the reactor is batch fed which causes the sludge temperature to vary with time in a saw-tooth pattern between 2 and 4°C per cycle. To predict the cyclically varying temperature profile requires a moment by moment solution of an unsteady state heat balance. However, this cyclic temperature variation is so small relative to the difference between the feed and reactor sludge temperatures (i.e. ± 40°C) that the mean reactor sludge temperature is completely adequate for design. Accordingly the steady state approach is adopted for design because it greatly simplifies the design procedure. The small error that this introduces is readily corrected with the aid of a simulation computer program which solves the moment by moment unsteady heat balance, using the steady state estimate as a starting point. The general simulation algorithm and computer program ATASIM for solution of the unsteady state heat balance is presented in the fourth and final paper in this series (Messenger and Ekama, 1993). A list of symbols is given by Messenger et al. (1993a).

Aerobic reactor design objectives

Two objectives need to be met by the aerobic reactor in dual digestion, viz.

- **Pasteurisation**: By exposure of the sludge to a minimum temperature for a minimum length of time, generally above 60°C for 2 h or above 70°C for 30 min.
- **Pretreatment**: The thermophilic aerobic environment in the reactor apparently pretreats the sludge leading to an enhanced performance of the anaerobic digester. Mason (1986) suggests that this pretreatment is improved by operating the reactor

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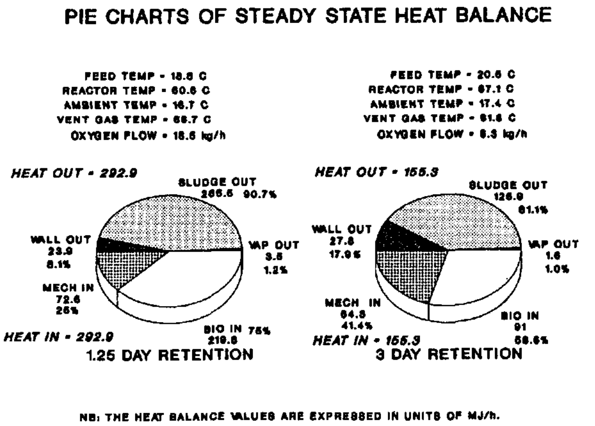

Figure 1

Typical steady state heat balance results of the Milnerton pure oxygen oxygenated reactor at 1,25 and 3 d retention time illustrating the relative proportions of the heat sources (biological generation and mechanical input) and heat losses (hot effluent sludge, water vapour and wall heat loss – vent gas sensible hfeat loss negligible)

under oxygen limiting conditions - this apparently increases the production of extracellular enzymes and improves solubilisation of particulate material which, according to Eastman and Ferguson (1981), is the rate limiting step in anaerobic digestion of sewage sludge.

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 the specified pasteurisation temperature and times are maintained in the reactor, and that the sludge is oxygen limited, i.e. the oxygen transfer rate [OTR, kgO/(m³·h)] to the sludge by the oxygenation system is controlled to a lower value than the sludge's oxygen consumption rate [OCR_{bio}, kgO/(m³·h)]. Not only is this important for sludge pretreatment but it also enables reactor sludge temperature control (Messenger et al., 1993a). In the aerobic reactor, sludge stabilisation, i.e. VS reduction, is **not** an objective; this is accomplished in the anaerobic digester. Therefore calculation of the oxygen supply rate (OSR) is based on the requirement to produce and control the rate of biological heat generation to achieve pasteurisation conditions.

Steady state heat balance over the aerobic reactor

In the general case for pure oxygen and air oxygenation, the steady state heat balance over the aerobic reactor is given by:

$$H_{sources} = H_{sinks} \quad \text{(MJ/h)}$$
$$H_{bi} + H_{mi} + H_{gi} + H_{si} = H_{se} + H_{ge} + H_{we} + H_{ve} \qquad (1)$$

Equation (1) can be simplified considerably by making the reasonable assumptions that density (ρ_s), specific heat (C_{ps}) and volumetric low rate Q_s, of the influent and effluent sludge liquor streams, are the same. For sewage sludge, C_{ps} and ρ_s can be accepted to be equal to that of water i.e. C_{ps} = 4,184 MJ/(t·°C) and ρ_s = 1,00 t/m³; this tends to overestimate the heat requirements but insignificantly so. With pure oxygen oxygenation the influent and effluent volumetric sludge flow can be assumed equal, because the flow of water vapour lost in the vent gas is negligible; at Milnerton less than 0,01% of the influent flow (Messenger et al., 1993a). In air oxygenated systems this loss is much higher (~ 5 to 10% depending on retention time, Pitt and Ekama, 1993) so that the volumetric flow rate of the effluent may be significantly lower than that of the influent. However, insofar as the heat balance is concerned, whether water leaves the aerobic reactor as water vapour or as effluent liquid, it still requires to be heated to the reactor temperature. Therefore, if the sensible heat of the water vapour is excluded in the heat balance, which is the case in Eq. (1), the influent and effluent sludge volumetric flow rates can be assumed equal with no loss of accuracy to the heat balance. Accepting the 3 assumptions, the difference between H_{si} and H_{se} in Eq. (1) may be combined into a single term H'_{se} representing the net rate of heat loss via hot sludge leaving the reactor i.e.:

$$H'_{se} = (H_{se} - H_{si}) = [V_p/R_H \cdot 24)]\ \rho_s\, C_{ps}\, (T_{se} - T_{si})\ \{\text{MJ/h}\} \qquad (2)$$

The two terms representing the sensible heats of the influent and vent gas streams H_{gi} and H_{ge} also can be combined into a single term, if it can be assumed that the specific heats (C_{pg}) of the various gas components O_2, N_2 and CO_2 are constant and equal over the vent gas temperature range 50 to 65°C and that the molar flow rates (m_{dg}) of the dry influent and dry vent gas streams are equal. The reasonableness of the first assumption can be seen from the negligible difference between the specific heats of O_2, N_2 and CO_2 at 20°C and 60°C i.e. O_2: 0,0299 and 0,0302 MJ/(kmol·°C); N_2: 0,0291 and 0,0293 MJ/(kmol·°C); CO_2: 0,0379 and 0,0392 MJ/(kmol·°C). Taking due consideration of the vent gas composition of air and oxygen oxygenated reactors, average C_{pg} values of 0,0302 and 0,0368 MJ/(kmol·°C) were calculated for air and oxygen oxygenated reactors respectively. With regard to the second assumption, this clearly is not true because the respiration quotient (Y_{CO_2} moles CO_2 produced per mole O_2 consumed), is not 1,0 (Messenger et al., 1993a; Pitt and Ekama, 1993). However, for pure oxygen reactors, where the Y_{CO_2} value plays a significant role in determining the dry vent gas molar flow rate (m_{dg}), the magnitude of the influent and vent gas sensible heat terms H_{gi} and H_{ge} is negligibly small; H_{ge} is less than 13% of the vent gas water vapour heat loss (H_{ve}) which in turn is less than 1% of the total heat losses in Eq. (4) below (see Fig. 1). In contrast, for air oxygenated reactors, the terms H_{gi} and H_{ge} are not negligible in the heat balance, but now the volume of inert N_2 gas passing through the reactor is so large in comparison to the O_2 consumed 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 the vent gas molar flow rate. Therefore it can be accepted that, insofar as the heat balance is concerned, the molar flow rates of the influent and vent gas streams can be assumed equal, without any significant loss of accuracy. This assumption does not hold for accurate oxygen mass balances! (Messenger et al., 1992; 1993a).

Accepting the two assumptions, the net rate of sensible heat loss in the vent gas H'_{ge} is given by:

$$H'_{ge} = (H_{ge} - H_{gi}) = m_{dg}\, C_{pg}\, (T_{ge} - T_{gi})\ \text{(MJ/h)} \qquad (3)$$

Substituting Eq. (2) for H_{se} and H_{si} and Eq. (3) for H_{ge} and H_{gi} into Eq. (1) and rearranging, yields the simplified steady state heat balance useful for design:

$$H_{sources} = H_{losses}$$
$$H_{bi} + H_{mi} = (V_p/R_H) \cdot \rho_s\, C_{ps}\, (T_{si} - T_{se}) + H'_{ge} + H_{we} + H_{ve} \qquad (4)$$

From Eq. (4) it can be seen that aside from the wall heat loss H_{we} which can be kept very small and constant by adequate lagging of the reactor, the heat generated in the reactor (H_{bi} + H_{mi}) is lost via only 2 variable routes: the hot effluent sludge H_{se}, and the vent gas via the sum of the sensible H'_{ge} and water vapour H_{ve} heat losses. From Eq. (4), to achieve short retention times requires maximising the heat sources H_{bi} and H_{mi} and minimizing the heat losses H_{ve} and H'_{ge}. However, H_{bi}, H_{mi} and H_{ve} and H_{ge} are interrelated through the oxygenation system; H_{bi} is proportional to the amount of oxygen consumed by the sludge while H_{ve} is proportional to the volume of vent gas generated during oxygen transfer. Some of the electrical power consumed by the oxygenation system in order to effect a given OTR contributes to heating the sludge. Details of the interrelationship between the various heat balance terms are discussed below.

Throughout this paper, where reference is made to the Milnerton (pure oxygen) and Athlone (air) dual digestion systems, the information was obtained from De Villiers et al. (1992) and Messenger et al. (1992;1993a,b) for the former and from Pitt (1990) and Pitt and Ekama (1993) for the latter.

Interrelationship between heat balance terms

The biological heat generation rate, H_{bi}

The biological heat generation rate is directly proportional to the biological oxygen consumption rate (Messenger et al., 1992; 1993a). In maximising H_{bi} there are 3 constraints, the first a biological one and the remaining two design and operational ones, viz:

Biological oxygen consumption rate (OCR_{bio})

The maximum biological heat generation rate $H_{bi\ max}$ is attained when the sludge oxygen consumption rate is not oxygen limited [OCR_{bio}, kgO/(m³·h)] i.e.

$$H_{bi\ max} = Y_H\ OCR_{bio}\ V_p\ \text{(MJ/h)} \qquad (5)$$

Under oxygen sufficient conditions, increasing the OSR to the reactor will not increase the OCR_{bio} of the sludge and therefore the biological heat generation rate cannot increase.

From a design point of view, the OCR_{bio} 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 digetion, OCR_{bio} estimation is based on VS removal rates, but this approach is not appropriate for the aerobic reactor in dual digestion (Messenger et al., 1990; 1992): At Milnerton at 1,25 d retention time, 75% of the heat required to maintain temperatures above 60°C (81 kW) was biologically generated with a negligible VS removal (1,5%). Until a reliable method of OCR_{bio} estimation can be developed for the aerobic reactor, it is recommended that it is measured in pilot-scale trials. At Milnerton, where the sludge treated was a mixture of primary and humus tank sludge, a steady state OCR_{bio} of about 0,38 kgO/(m³·h) was measured at a sludge concentration of 30 kgVS/m³. The temptation of converting the OCR_{bio} to a VS concentration specific rate should be resisted at this stage because the effect of VS concentration on OCR_{bio} is uncertain; it may be thought that doubling the VS concentration would double OCR_{bio}, but this may not necessarily be so. Biological oxygen consumption is related to the kinetics of substrate utilisation 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. A very accurate estimate of OCR_{bio} is not required for design because, as discussed below, the reactor needs to be operated under oxygen limiting conditions, with the result that the system oxygen consumption rate (OCR_{sys}) will be considerably lower than the OCR_{bio}.

Oxygen limitation

This is desirable for 2 reasons; pretreatment for enhancing subsequent anaerobic digestion and temperature control. In order to achieve oxygen limitation the sludge should consume oxygen at a rate (OCR_{sys}) lower than its OCR_{bio}. Therefore the oxygen transfer rate (OTR) must be less than the OCR_{bio} and under these circumstances the H_{bi} is given by:

$$H_{bi} = Y_H \cdot OTR \cdot V_p = Y_h\ OCR_{sys}\ V_p\ \text{(MJ/h)} \qquad (6)$$

As mentioned earlier, the degree of oxygen limitation (i.e. defined as $1\text{-}OTR/OCR_{bio}$ or $1\text{-}OCR_{sys}/OCR_{bio}$) for proper pretreatment could not be established from the literature, nor could it be quantified at Milnerton. At Milnerton, at 1,25 and 3 d R_H, the oxygen limitation was about 0,20 and 0,60 respectively. Under oxygen limiting conditions, it was found that the Milnerton reactor sludge temperature could be completely and virtually instantaneously controlled with the OSR.

Maximum oxygen transfer rate of oxygenation system (OTR_{max})

This parameter is very important in the design in the aerobic reactor - if OTR_{max} is too low, pasteurisation temperatures will not be achieved at the design retention time (unless supplementary heat sources are applied - see below). Associated with the OTR of the oxygenation system, is its oxygen transfer efficiency (OTE) where OTE = OTR/OSR. Generally, OTE of a system decreases as OTR increases up to OTR_{max}. The OSR fixes the vent gas molar flow rate which in turn fixes the vent gas heat loss rates H_{ve} and H'_{ge}. Once the system reaches its OTR_{max} further increases in OSR will not increase OTR but serve only to cause a greater H_{ve} and H'_{ge} without increasing H_{bi}. Preferably the design OTR_{max} should be somewhat greater than OCR_{bio} of the sludge. In that way, the biological kinetics (i.e. OCR_{bio}) will govern the minimum retention time rather than the OTR_{max} of the oxygenation system. This was the case at Milnerton where a temporary OCR_{bio} (as opposed to a sustained OCR_{bio}), and therefore also an OTR of at least as high, of 0,44 kgO/(m³·h) was observed. At Athlone, a 184 m³ reactor was oxygenated with a 19 kW liquid ring air compressor with a maximum airflow rate of 780 m³ (STP)/h through coarse bubble diffusers set 6 m below the sludge surface and mixed with a 7,5 kW centrifugal pump rated at 396 m³/h. A standard (tap water) unsteady state aeration test at ambient temperature yielded an OTR_{max} (at 20°C) of 0,145 kgO/(m³·h) and OTE of 0,125. Generally the OTR_{max} and the OTR-OTE interrelationship is a function of the oxygenation system as well as the mixing and geometry of the reactor.

Mechanical heating rate H_{mi}

Electrical power is required to mix and oxygenate the reactor sludge. Depending on the type of oxygenation and mixing system much of this power can contribute directly to heating the sludge. At Milnerton a continuously operating 20 kW Vitox pumped recirculation system provided both oxygenation and mixing, with more than 90% of the power drawn (19 kW≡70 MJ/h) contributing directly to heating the sludge; this contribution formed a significant portion of the heat sources, typically 24% and 42% at 1,25 and 3 d retention times respectively (see Fig. 1). Even though H_{mi} was constant at about 70 MJ/h, H_{mi} becomes a larger proportion of the heat sources as R_H increases because, as H'_{se} decreases with increasing R_H so H_{bi} must decrease to avoid overheating (Eq. 4).

The volume specific power consumption (W/m^3) to achieve a desired OTR_{max} and good mixing and the proportion of this power which contributes to heating the sludge will be different for different reactors and oxygenation systems. At Milnerton an OTR of at least 0,44 kgO/(m^3·h) was achieved at 450 W/m^3, 90% of which contributed to heating the sludge. At Athlone, an OTR of around 0,14 kgO/(m^3·h) was achieved at a total of 144 W/m^3 with only about 25% (90% of the 7,5 kW mixing pump) compressor and mixing power contributing to heating the sludge. At Palmersford (UK), Booth and Tramontini (1984) reported achieving an OTR of 0,134 kgO/(m^3·h) at 130 W/m^3 with a pure oxygenation system. Morgan and Gunson (1989) presented an air oxygenated design which they claim was capable of achieving an OTR of 0,43 kgO/(m^3·h) at 120 W/m^3 (substantiation of this claim could not be found in the written literature). With the last 2 mentioned, the proportion of the power contributing to heating the sludge is not known.

Owing to the large differences in oxygenation system and mixing designs, an equation to estimate the mechanical heat input rate H_{mi} is not given. In this paper H_{mi} will be dealt with by assigning to it a fixed and constant value, i.e. 70 MJ/h as measured on the Milnerton reactor.

Vent gas heat loss rate (H'_{ve}) – sensible (H'_{ge}) water vapour (H'_{ve})

For the sake of convenience, the vent gas heat loss rate H'_{ve} will denote the sum of the sensible (H'_{ge}) and water vapour (H_{ve}) heat loss rates. To calculate H_{ge} with Eq. (3), the dry vent gas molar flow rate (m_{dg}) needs to be known. For air and oxygen oxygenated reactors an equation for m_{dg} can be derived from first principles in terms of the OTR and the OTE of the oxygenation system (Messenger et al., 1992) i.e.:

$$m_{dg} = [(a\text{-}OTE+OTE \cdot Y_{CO_2})/(32 \cdot OTE)]OTR \cdot V_p \text{ (kmol/h)} \qquad (7)$$

where:

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

$f_{mo_2} =$ 0,2317 for air and 1,00 for pure oxygen

and so:

$a =$ 4,79 for air and 1,00 for pure oxygen

32 and 28 = molar masses of O_2 and N_2 respectively.

Accepting the 2 assumptions regarding H'_{ge} stated earlier, i.e. constant specific heat C_{pg} and Y_{CO_2} = 1,00 and substituting Eq. (7) into Eq (3), yields for H'_{ge}:

$$H'_{ge} = (a\ OTR \cdot V_p)/(32 \cdot OTE) \cdot C_{pg}\ (T_{ge}\text{-}T_{gi}) \text{ (MJ/h)} \qquad (8)$$

In the Milnerton reactor it was found that the vent gas temperature at the point of exit from the reactor (T_{ge}) was 2 to 5°C below that of the sludge (T_{ge}), and was due to heat losses through the walls of the reactor head space. This will probably arise in most aerobic reactors but because this temperature difference (T_{diff}) represents a very small heat loss (a minor fraction of the sludge wetted wall heat loss H_{we}, see below), for the purposes of design, it can be accepted that the vent gas temperature is a constant value less than that of the sludge. For the purposes of illustration in this paper, a T_{diff} of 3°C is accepted, i.e. $T_{ge} = T_{se}$ - 3°C.

The vent gas water vapour heat loss rate H_{ve} is given by the product of the latent heat (h_{fg}, MJ/kg) and mass flow rate (M_{we}, kg/h) of the water vapour i.e.:

$$H_{ve} = h_{fg} \cdot M_{we} \text{ (MJ/h)} \qquad (9)$$

Although h_{fg} is a function of temperature, its variation in the temperature range encountered in an aerobic reactor (50 to 65°C) is so small that a constant value of 2,358 MJ/kg can be accepted with negligible loss of accuracy.

The M_{we} is a function of the vent gas temperature (T_{ge}) molar flow rate of the dry vent gas (m_{dg}), and the degree of water vapour saturation of the vent gas. At both Milnerton and Athlone it was found that the vent gas was saturated at the vent gas temperature T_{ge}. Accepting saturation, an equation to calculate M_{we} can be derived from first principles and steam tables, viz:

$$M_{we} = 18\ m_{dg}\ p_w/p_{dg} \text{ (kg/h)} \qquad (10)$$

where:

$$p_w = 10^{[8,896 - 2238/(T_{ge}+273)]} \text{ mm Hg} \qquad (11)$$

$$p_{dg} = P_T - p_W \qquad (12)$$

Substituting the simplified form of Eq. (7) for m_{dg} (i.e. Y_{CO_2} = 1,0) into Eq. (10) for M_{we}, and the resulting equation for M_{we} into Eq. (9) yields:

$$H_{ve} = h_{fg} \cdot 18(a \cdot OTR \cdot V_p)/(32 \cdot OTE) \cdot (p_w/p_{dg}) \text{ (MJ/h)} \qquad (13)$$

Adding Eqs. (8) and (13) for H'_{ge} and H_{ve} respectively gives the total vent gas sensible and water vapour heat loss H'_{ve} i.e.:

$$\begin{aligned} H'_{ve} = {} & (a \cdot OTR \cdot V_p)/(32 \cdot OTE)[C_{pg}(T_{ge}\text{-}T_{gi}) \\ & + h_{fg}\ 18(p_w/p_{dg})] \text{ (MJ/h)} \end{aligned} \qquad (14)$$

From Eq. (14) it can be seen that H'_{ve} is proportional to OTR and inversely proportional to OTE. If the vent gas is saturated with water vapour, H_{ve} is by far the greater part of H'_{ve} being about 7,7 and 9,3 times larger than H'_{ge} for oxygen and air oxygenated reactors respectively under typical conditions.

The wall heat loss rate, H_{we}

This heat loss is readily minimised by providing thick high quality lagging on all the reactor surfaces and hot sludge transport pipework and/or locating the reactor in an enclosed building to eliminate exposure to wind, rain and cold ambient temperatures.

Very simplistically, the wall heat loss H_{we} can be described as follows:

$$H_{we} = U_{oa}\ (T_{se}\text{-}T_{amb})\ A \text{ (MJ/h)} \qquad (15)$$

where:

U_{oa} = overall heat transfer coefficient {MJ/(m^2·°C·h)}

A = reactor and hot pipework surface area (m^2)

T_{amb} = ambient temperature (°C)

At Milnerton, where the reactor was not enclosed in a building, the use of 50 mm polyurethane lagging and the mild climate kept H_{we} to a low value of 10 to 20 MJ/h, which is about 15 to 30% of the mechanical heat input rate H_{mi}, or 4 to 9% of the heat losses at 1,25 and 3 d retention time respectively (Fig. 1). Unless lagging is insufficient and the variations in ambient temperature and weather extreme, H_{we} can be accepted for design to remain approximately constant. In this paper H_{we} will be accepted to remain constant at 20 MJ/h, the value measured in the Milnerton reactor. At Milnerton a U_{oa} value of 0,00733 MJ/(m^2·°C·h) was measured. For a particular design H_{we} will be quite difficult to estimate, but a very accurate value is not required because generally H_{we} can be designed to be less than 5% of the heat losses through the use of thick quality lagging and insulation. At Athlone, the aerobic reactor was built inside the anaerobic digester to minimise H_{we}; here H_{we} was estimated around 5% of the heat losses.

Discussion of design interrelationships

All the terms required for the steady state heat balance (Eq. 4) have been defined above in Eqs. (5) to (15) with the exception of the retention time R_H and the oxygenation system OTR and OTE. Consequently, by specifying certain reactor design and operating conditions, R_H can be plotted versus OTE for selected OTR. This is done in Fig. 2 taking the Milnerton aerobic reactor as a basis, viz:

- Reactor volume $V_p = 45$ m^3
- Influent and effluent sludge (T_{si} and T_{se}) and gas (T_{gi} and T_{ge}) temperatures 20°C, 60°C, 20°C and 57°C respectively
- Total atmospheric (barometric) pressure at sea level (P_T) 760 mm Hg
- Mechanical heat input (H_{mi}) and wall heat loss (H_{we}) constant at the measured values of 70 MJ/h and 20 MJ/h respectively
- Sludge $OCR_{bio} = 0,40$ kgO/(m^3·h).

The lines in Fig. 2 have the shape of hyperbolae, 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 H_{bi} and H_{mi} are fixed and this heat is lost principally via 2 varying routes, i.e. via hot effluent sludge H'_{se} which is related to the retention time (Eq. 2) or via the vent gas (H'_{ve}) which is related to the OTE (Eq. 14). Accordingly, in Fig. 2, along a particular hyperbola (i.e. constant OTR), longer 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 is not significant 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. 2) represents low heat loss via the vent gas and high heat loss via the hot effluent sludge and a low OTE and long R_H (top left of Fig. 2) represents low heat loss via the hot effluent sludge and high heat loss via the vent gas. With this general background to Fig. 2, the design of air and oxygen oxygenated reactors is now examined in a more practical way below.

Relative significance of OTR and OTE

A high OTR will yield a high rate of biological heat generation, H_{bi}, provided OTR < OCR_{bio}. High OTRs also result in large vent gas flow rates so that it is only by maximising OTE that the heat loss H'_{ve} can be minimised. In general, the OTR and OTE achieved by an oxygenation system, whether it be pure oxygen or air, are entirely dependent on the characteristics and type of system, as well as the mixing and geometry of the reactor. In order to transfer sufficient oxygen to enable the generation of the required biological heat, the relationship between the OTR and OTE of the oxygenation system needs to be known and well defined. If the OTR-OTE relationship is not known, then H_{bi} and H'_{ve} cannot be defined and hence also not the minimum retention time.

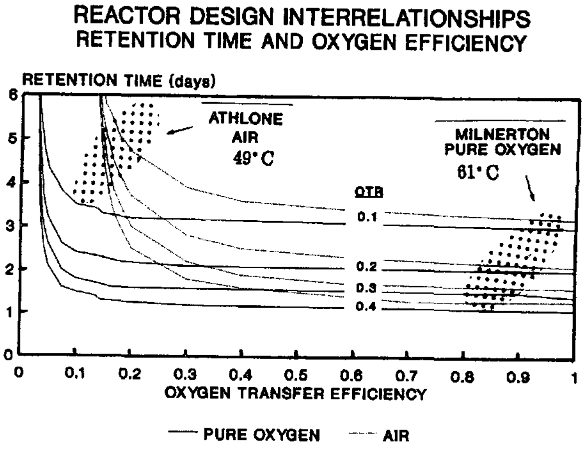

Figure 2

Hydraulic retention time (d) versus oxygen transfer efficiency for OTRs of 0,10; 0,20; 0,30 and 0,40 kgO/(m^3·h) for pure oxygen (solid lines) and air (dashed lines) oxygenated aerobic reactors to maintain a sludge temperature of 60°C. Approximate results for the Milnerton and Athlone reactors are plotted to illustrate operating performance of pure oxygen (Milnerton, bottom right) and air (Athlone, top left) oxygenated reactors. The actual Milnerton and Athlone average reactor sludge temperatures are 61 and 49°C

Pure oxygen versus air

High OTR and OTE values can be achieved in a reactor oxygenated with pure oxygen and the resulting H'_{ve} will generally be small enough to be neglected. This clearly maximises OTE and minimises H'_{ve}. 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 implemented (see below). The large volumes of nitrogen gas introduced into the sludge with air vaporise large masses of water, a situation which is exacerbated by even a slight decrease in OTE. Although OTEs of 1 have been reported by Wolinski (1985) for an air system, these efficiencies were achieved at low OTRs [0,076 to 0,132 kg/(m^3·h)], which necessarily imply long retention times (Fig. 2). In contrast, at Athlone where a 184 m^3 air oxygenated aerobic reactor has been operated over the past two years, OTEs ranging from 0,09 to 0,23 and averaging around 0,12 were achieved at OTRs ranging from 0,12 to 0,19 kgO/(m^3·h) and averaging around 0,14 kgO/(m^3·h). The higher OTRs and OTEs generally occurred together at times when a large foam layer (3 m) built up on the reactor. When the large foam layer was absent, attempts to increase OTR by increasing the air supply rate led to a reduction in OTE, and generally temperatures above 50° C could not be maintained at retention times less than 4 d.

To demonstrate the design implications of a pure oxygen and

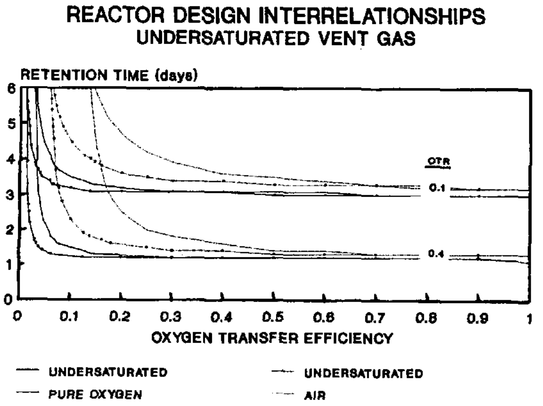

Figure 3
Hydraulic retention time (d) versus oxygen transfer efficiency at oxygen transfer rates of 0,10 and 0,40 kgO/(m^3·h) for pure oxygen (solid lines) and air (dashed lines) oxygenated reactors with vent gas water vapour saturation (lines without dots) and undersaturation (lines with dots) to maintain a sludge temperature of 60°C

air oxygenated aerobic reactor, the results observed at the Milnerton pure oxygen and Athlone air oxygenated reactors are plotted in Fig. 2. At Milnerton OTEs ranging between 0,80 to 1,00 at retention times 1,25 to 3 d were obtained with OTE increasing as R_H increased; at Athlone, OTEs between 0,09 and 0,23 averaging at 0,12 and retention times between 3 and 6 d were achieved with OTE increasing as R_H increased (although the foam layer had a greater influence on this than R_H). These 2 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, much greater heat losses via the vent gas take place thereby requiring longer retention times.

From Fig. 2, clearly the greater the OTR and OTE, the shorter the retention time R_H. This demonstrates the importance of achieving high OTRs, with both pure oxygen and air, if the reactor is to be operated at short R_H; at high OTRs, high rates of biological heat generation are achieved. Now as mentioned earlier, the rate of biological heat generation is limited by the sludge biological oxygen consumption rate (OCR_{bio}) and increasing the OTR above the OCR_{bio}, does not increase the rate of biological heat generation; indeed, it will only serve to decrease the OTE and hence increase H'_{ve}. Furthermore when OTR reaches OCR_{bio}, 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 and reactor temperature control via the oxygen/air supply rate are compromised.

Vent gas water vapour undersaturation

In Fig. 2, vent gas water vapour saturation is assumed. This was found to be the case not only at Milnerton but also at Athlone, the latter with air oxygenation, where it would have seemed less likely to occur. Nevertheless, vent gas saturation may not necessarily apply in every case where low OTRs and OTEs are encountered. Figure 3 shows plots of R_H vs OTE for pure oxygen and air oxygenated systems at OTRs of 0,1 and 0,4 kgO/(m^3.h) for saturated and undersaturated vent gas conditions, where the undersaturated case was modelled on the basis that as OTE decreased from 1,0 to 0,0 so the degree of saturation decreased linearly from 100% to 33%. From Fig. 3 it can be seen that if the degree of vent gas vapour saturation, and thus the rate of vent gas heat loss, decreases with OTE, then the point where the vent gas heat loss begins to become significant with respect to the effluent hot sludge heat loss shifts to lower OTE values. That this is so is clear because reducing the degree of water vapour saturation reduces the vent gas heat loss and therefore allows stable operation at lower OTEs than if the vent gas were saturated. However, undersaturation does not reduce the minimum retention time operation at high OTEs because at these OTEs, the heat loss via the hot effluent sludge is much greater than the heat loss via the vent gas; here as before the OTR plays the major role in fixing the minimum retention time.

The influence of mechanical heat input rate (H_{mi})

The H_{mi} contributes directly to the heat sources. Increasing H_{mi} does not cause increased vent gas heat loss like increasing H_{bi} does. Although process economics require H_{mi} to be minimised, it is likely that high OTRs will be achieved only at the expense of a high power consumption to dissolve the large mass of oxygen per m^3 reactor volume and mix the reactor sludge. Thus the advantage of a high H_{bi} arising from a high OTR will be supplemented by a high H_{mi}. Conversely, the combination of a low OTR and a low H_{mi}, will severely compromise any effort at achieving a short retention time operation because this reduces the two available heat sources. This is illustrated in Figs. 4a and 4b where reducing H_{mi} (from 70 to 35 MJ/h) can be seen to have a much greater effect on retention time at a low OTR [0,2 kg/(m^3·h)] (Fig. 4a) than at a high OTR [0,4 kg/(m^3·h) (Fig. 4b)].

Supplementary heat sources, e.g. heat exchange

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 oxygen in less oxygen.
(4) Reduced vent gas heat loss.
(5) Lower sensitivity of reactor to OTR limitations of the oxygenation system.
(6) Less need to know OCR_{bio} accurately.

or none of the above but instead:

(7) Shorter retention time, i.e. higher sludge volume throughput.

Two readily exploitable supplementary heat sources available for dual digestion systems are heat exchange between reactor effluent and influent sludge and consumption of anaerobic digester gas. In practice these heat sources might be employed to preheat the influent sludge so their contribution to the heat balance can most easily be accounted for by increasing the reactor sludge feed temperature T_{si}. If R_H remains unchanged, then the increased T_{si} reduces H_{se} (Eq. 2). This allows H_{bi} (Eq. 1) and hence also OTR (Eq. 6) and OSR to be reduced and advantages (1) and (6) above result. On the other hand, if OTR (and OTE) are kept constant, H_{bi} and H'_{ve} do not change with the result that H'_{se}, (Eq. 3) also is unchanged. With H'_{se} unchanged, the increased T_{si} allows R_H to be reduced (advantage 7).

To demonstrate the effect of heat exchange on the aerobic reactor, a rather extreme case was accepted where the reactor influent sludge was heated 16°C from 20°C to 36°C by cooling the effluent sludge by 16°C from 60°C to 44°C, i.e. the influent and effluent sludge temperature difference (T_{se}-T_{si}) was reduced by 40% (Fig. 5). At constant OTR and OTE, H_{bi} and H'_{ve} remain unchanged and, with constant H_{mi} and H_{we}, the hot effluent sludge heat loss H'_{se} remains unchanged through the steady state heat balance. The 40% reduction in (T_{se}-T_{si}) therefore allows a 40% reduction in retention time: For the pure oxygen case (high OTR and OTE; 0,40 kgO/(m^3·h) and >0,80 respectively) the retention time is reduced from 1,25 to 0,75 d; for the air case (low OTR and OTE; 0,10 kgO/(m^3·h) and 0,20 respectively), the retention time decreases from 4,7 to 2,8d (Fig. 5). It should be noted also from Fig. 5 that for the air case with heat exchange, the 0,10 and 0,40 kgO/(m^3·h) OTR lines cross each other at an OTE of 0,13 making the lower OTR the better design. This arises because at these low OTEs, less heat is lost via the vent gas at lower OTRs than at higher OTRs. Heat exchange has the same effect as increasing the mechanical heat input H_{mi}; it adds heat to the reactor without contributing to the heat losses.

In practice a heat exchanger will be more useful as a means of controlling the anaerobic digester temperature than as a provider of supplementary heat to the aerobic reactor. The only heat source to maintain and control the digester temperature is the hot effluent sludge from the reactor with the heat input rate being proportional to both the flow rate and temperature of this sludge. Without heat exchange the digester heating rate cannot be varied and the heat input rate may be either excessive (in summer) or insufficient (in winter) to maintain the digester temperature within the optimum range of 35 to 38°C. At Milnerton, in the winter, all the heat of the hot (60°C) aerobic reactor effluent sludge at 1,25 d retention time was required to maintain 37°C in the digester. Whereas, in the summer, this was too much with the result that the digester overheated unless the aerobic reactor retention time was increased to reduce the heat flow to the digester. At Athlone, where heat exchange also was not practiced and the aerobic reactor was built inside the digester to conserve heat, the average reactor and digester temperatures were 49,1°C and 31,6°C respectively. In this case the optimum mesophilic temperature in the digester was not attained due to the low temperature of the reactor effluent sludge.

Where a heat exchanger serves to control digester temperature, the need for a varying heat flow to the digester will result in it being used intermittently. This impacts on temperature control in the aerobic reactor because it will require the reactor heat sources to vary; sometimes supplementary heat will be available and sometimes not. In order to keep the reactor sludge at a constant temperature, the heat sources must remain approximately constant and hence another heat source will need to be increased when the supplementary heat source decreases and vice versa. Of the mechanical and biological heat sources, only the biological source can be readily varied through the OSR but only while the reactor is oxygen limited. Alternatively, another supplementary heat source, such as the combustion of anaerobically produced methane, may be exploited to compensate for changes in the availability of heat resulting from heat exchange. Combustion of gas alone might be sufficient to produce all of the heat required, giving rise at the extreme to a situation where biological heat generation could be completely dispensed with. In such a situation, the two-stage process would no longer be a dual

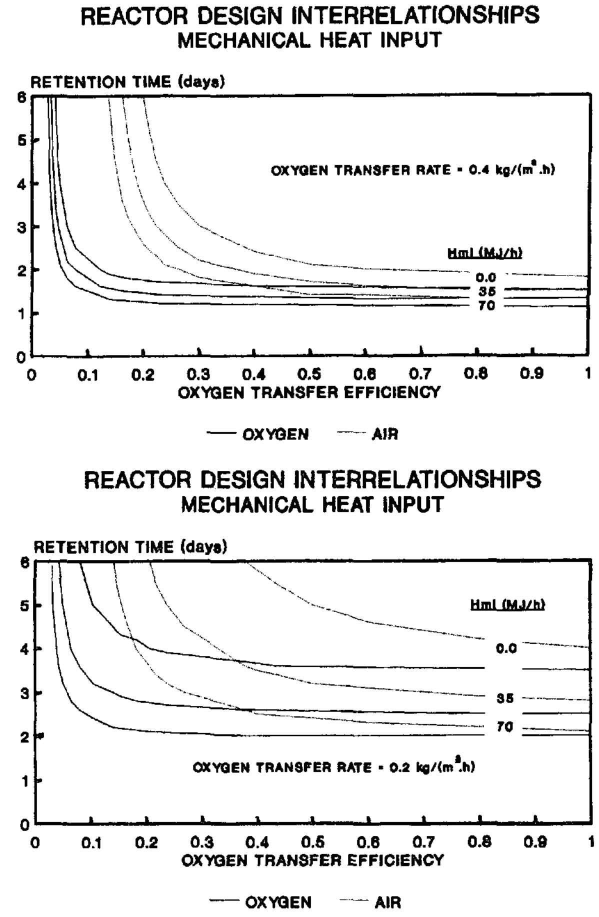

Figure 4

Hydraulic retention time versus oxygen transfer efficiency at oxygen transfer rates (OTR) of 0,40 (Fig. 4a, top) and 0,20 (Fig. 4b, bottom) kgO/(m^3·h) for pure oxygen (solid lines) and air (dashed lines) oxygenated reactors for mechanical heat input rates (H_{mi}) of 0,35 of 0,35 and 70 MJ/h to maintain a reactor temperature of 60°C

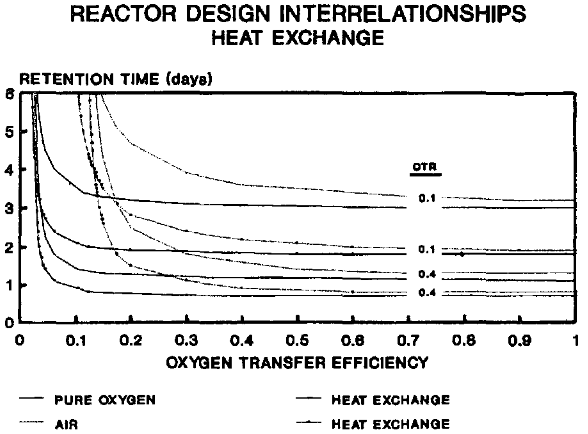

Figure 5

Hydraulic retention time (d) versus oxygen transfer efficiency at oxygen transfer rates (OTR) of 0,10 and 0,40 kgO/(m^3.h) for pure oxygen (solid lines) and air (dashed lines) oxygenated reactors for no heat exchange (without dots on lines) and 40% heat exchange (with dots on lines) between effluent and influent sludge streams to maintain a reactor temperature of 60°C digestion system but would become the combination of a sludge pre-pasteuriser and an anaerobic digester as implemented at Cape Flats Waste-water Treatment Plant (Morrison, 1990). However, dispensing with biological heat generation, and therefore sludge oxygenation, might have an impact on the claimed sludge pretreatment effect in the reactor. While sludge pasteurisation could still be accomplished, the claimed potential of the reactor to significantly enhance the performance of the subsequent anaerobic digester might be lost, a claim which could not be conclusively verified in the Milnerton research.

Conclusions

In this paper process 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 pure oxygen and the Athlone air reactors, the practical aspects and experience gained are discussed by Messenger et al. (1992) and Pitt and Ekama (1993).

Accepting that the objectives of the aerobic reactor are sludge pretreatment through oxygen limitation and pasteurisation 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, which centres on minimising the retention time through maximising the heat sources (i.e. biological heat generation and mechanical heat input) and minimising the heat losses (i.e. wall heat loss and vent gas water vapour and sensible heat losses). The 3 parameters are:

- The sludge OCR_{bio}. This fixes the maximum biological heat generation rate.
- The 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 OCR_{bio} to ensure sludge pretreatment through oxygen limitation and reactor temperature control through control of the oxygen supply rate (OSR).
- The 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 2 d); if OTE is low (0,10 to 0,20) and oxygenation is with air, 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 to 6 d) to maintain thermophilic temperatures (> 60° C). For example at Milnerton, where high OTRs and OTEs of 0,37 $kgO/(m^3 \cdot h)$ and > 0,80 respectively were achieved with pure oxygen, the vent gas heat losses were very low enabling operation at 1,25 d retention time and reactor temperatures above 60°C (Messenger et al., 1992, 1993a). In contrast, at Athlone, where low OTRs and OTEs of around 0,14 $kgO/(m^3 \cdot h)$ and 0,12 respectively were achieved with air, a larger part of the heat generated was lost via the vent gas with the result that a smaller part of the heat could be lost via hot effluent sludge thereby forcing 4 to 6 d retention time operation to maintain temperatures above 50°C (Pitt, 1990; Pitt and Ekama, 1993).

Biological heat generation has the drawback that increasing its rate requires increases in OTR. Increases in OTR require increases in oxygen or air supply rates which, together with the reduction in OTE usually accompanied by increases in OTR, cause larger vent gas heat losses. In contrast to biological heat generation, increased mechanical heat input and supplementary heat sources such as heat exchange between reactor effluent and feed sludge or anaerobic digester 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 considerations presented in this paper are based on the steady state heat balance which accepts that the reactor is continuously fed and sludge temperature is constant. Pasteurisation requires the reactor to be batch fed, resulting in a continuously changing sludge temperature over about 3°C in a saw-tooth pattern, 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 the last paper of this four-part series (Messenger and Ekama, 1993), which presents an algorithm and a computer program (ATASIM) for solving the unsteady heat balance to simulate the temperature profile of the batch fed aerobic reactor.

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