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The Design of Airlift Fermenters for use in Biotechnology

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Introduction

Rapid progress is being made in the discovery and commercial development of genetic engineered products. New biotechnology has provided pharmaceutical products such as synthesized human insulin, human growth hormone (HGH), tissue plasminogen activator (TPA), animal growth hormones, the α-, β- and γ-interferons, synthesized vaccines for hepatitis, herpes, malaria, polio, cholera, rabies and new agricultural bioengineered products, such as frost-resistant additives, environmentally safe pesticides, fungicides and herbicides.

Traditional fermentation products also span the wide range from vitamins, enzymes and amino acids to alcoholic beverages, artificial sweeteners and single-cell protein.

As we become more aware of the adverse impact of technology on our environment, the need to protect it becomes ever more urgent. Biological treatment of polluted water and industrial wastewater is but one further example of the application of biotechnology.

Common to all the above is the use of bioreactors or fermenters for biomass production, which require efficient multi-phase heat transfer and metabolite transport between the biomass and its environment for it to sustain life, grow and reproduce. Among the many fermenters are the airlift types, which use air as the principal means to achieve the various transport requisites.

Since 1969 (Hatch, Cuevas and Wang, 1969; Wang, Hatch and Cuevas, 1971), airlift fermenters without mechanical agitation have attracted increased interest. Airlift fermenters particularly feature simplicity and low capital investment, which are most important when fermenter efficiency and processing costs are of major concern. They have higher efficiency in mass transfer compared to mechanically stirred fermenters at the same power

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inputs (Onken and Weiland, 1983). However, except for a few instances, such as wastewater treatment processes and the ill-fated single-cell protein processes, for most other bioengineered products the fermenter cost has not been a major concern. This situation is expected to change when new products are mass produced.

Airlift fermenters, their hydrodynamics, mass transfer and heat transfer characteristics as well as their uses were reviewed previously by Onken and Weiland (1983).

Airlift fermenter designs

Airlift fermenters generally consist of two sections, with only one of the sections sparged with air. The difference in the hydrostatic head between the sections provides the driving force for liquid circulation. Compared to bubble columns, airlift fermenters provide an additional mode of liquid flow which increases mass and heat transport efficiency. Compared to mechanically agitated stirred-tank fermenters, airlift fermenters are easier to scale up, require less energy to operate and are more adaptable to cultures that are shear sensitive. As stirred-tank systems grow in size, their mixing quality suffers. The mixing time in airlift systems, on the other hand, increases only slightly with increased size because the liquid circulation rate can be kept almost constant (Sittig and Faust, 1982).

In most cases, the rate of mass transport of oxygen from the gas bubbles to the fermenter broth determines the maximum rate of biomass production. In practice, the efficiency of a given fermenter is measured by its volumetric production rate and its specific power consumption, the energy spent to create interfacial surface areas of contact between the various phases involved in the biochemical reactions.

Many variables affect the rate of mass transfer in an airlift fermenter. Key physical parameters include the vessel geometry, height to width or diameter ratio, internal baffle arrangement, sparger type and location. The rate of mass transfer is dependent on the interfacial area, which is determined by the gas hold up, the average size and the size distribution of the air bubbles dispersed in the liquid and the rate of bubble break-up and coalescence. These are functions of the physical parameters of the fermenter, the superficial air velocity and the physical, rheological and chemical properties of the liquid, such as temperature, density, viscosity, interfacial tension, substrate composition and concentration, metabolite and biomass concentration, etc.

Commercial airlift fermenter designs

CONCENTRIC DRAFT TUBE AIRLIFT FERMENTER

Most industrial airlift fermenters are of the concentric draft tube type. In most designs, the coaxial draft tube functions as the aerated section. Air-sparged liquid rises up the draft tube, is partially degassed and flows down the annulus. In some designs, the flow pattern is reversed by sparging the air at the base of the annulus between the draft tube and the outer wall. The choice appears to be a matter of individual preference, although differences in fermenter foaming and heat transfer rate have been noted by some investigators (Cooper, Silver and Boyle, 1975).

The ratio of the diameter of the draft tube to that of the fermenter, or the ratio of the cross-sectional area of the riser to that of the downcomer, are important design characteristics in maximizing liquid circulation rate and in minimizing mixing time. Diameter ratios in the range of 0·6–0·8 have often been quoted to give optimal performance.

Another feature of industrial fermenters of the concentric draft tube type is their slim design, i.e. very large height to diameter ratios, as large as 10. Increasing the hydrostatic head by increasing liquid height in the fermenter increases the partial pressure of oxygen at the base of the aerated section and provides a larger driving force for oxygen transfer. A large difference in hydrostatic head between the aerated section and the downcomer section also promotes liquid circulation. To redisperse gas bubbles in the upflowing fluid, baffles are sometimes placed in the draft tube.

TOWER LOOP AIRLIFT FERMENTER

Tower loop airlift fermenters, also known as external loop airlift fermenters or tubular loop fermenters, differ from the concentric tube fermenters by having the air-sparged riser column physically separated from the downcomer. A typical design consists of two vertical columns, usually of different diameters, connected at the top with a degassing zone and at the bottom with a liquid return line. In contrast to well-mixed systems, the reaction conditions in a tower airlift fermenter are space-dependent. Luttmann and his coworkers (Buchholz *et al.*, 1980; Luttmann *et al.*, 1982, 1983a, b) have developed a quasi-steady-state model for the mass production of bacteria and yeasts. From the results of comparing the model with experimental data, they concluded that the volumetric oxygen transfer coefficient must vary along the column due to bubble coalescence.

Another characteristic of the tower loop airlift fermenters is in the cross-sectional area ratios of riser to downcomer. They vary much more than in concentric draft tube fermenters, ranging from 1 to 30 or more. The relative ease in obtaining data from a tower loop airlift reactor made this type of reactor the choice of many laboratory studies.

ICI DEEP SHAFT AIRLIFT FERMENTER

ICI developed, as a spin-off from their single-cell protein fermenter, a biological effluent treatment process using a below-ground deep-shaft airlift fermenter. Constructed with oil drilling and mining techniques, these wells have diameters up to 3 m (10 ft) and are cased in concrete, plastics or coated carbon steel. They are normally 45–150 m (150–500 ft) deep, with liquid circulating at 0·9–1·5 m s^{-1} (3–5 ft s^{-1}) with air sparging both upward in the riser section and downward in the downcomer section (Bolton, Hines and Bouchard, 1977).

The first commercial unit, measuring 1·1 m (3·6 ft) in diameter and 100 m (330 ft) deep, was cased in carbon steel. It was used to treat continuously the wastewater streams from a potato starch plant, at about 2 h average residence time.

The airlift fermenter has an oxygen transfer rate of 60–90 mmol l^{-1} h^{-1} (see the equation p. 387). Because the fermenter is deep and the oxygen demand of the wastewater streams is generally quite low, the power requirement for such a unit can be minimized by locating the air spargers not at the bottom of the riser but at as high as 20 m (70 ft) from the top, sufficient to overcome the resistance to liquid circulation. Power economy is claimed to be in the range of 2·7–5·4 kg (6–12 lb) of oxygen transferred per kWh and the oxygen in the injected air is nearly completely consumed in the fermenter.

LOW PRESSURE AIRLIFT FERMENTER

To produce single-cell protein from hydrocarbons, the high demand of oxygen with hydrocarbon substrates and the rising cost of energy and capital investment made it apparent that economical fermenter designs could have a significant impact on the cost of production. In addition to high oxygen demand, the process is exothermic. The amount of heat to be removed is related to the cell yield on carbon. The lower the yield, the greater is the demand on oxygen and heat removal (Cooney and Makiguchi, 1977). The minimum amount of heat to be removed during yeast propagation can be estimated from the theoretical mass balance described above, using the heats of combustion of the substrates and dry cells. For each gram of dry cell produced, a minimum of about 27·6 kJ of heat must be removed from the fermenter. This value is in general agreement with the values reported by Holve (1976).

The cost of heat removal is a major component of total fermentation cost. However, in most airlift fermenter designs the exothermic heat generated during fermentation is removed by using cooling coils or via an external circulation loop through heat exchangers (Wang, Hatch and Cuevas, 1971; Cooney and Makiguchi, 1977; Onken and Weiland, 1983). Because most fermentation systems operate not far above ambient temperatures, they are therefore cooled with chilled water. As a result, fermentation plants usually include large refrigeration systems, adding to the operating cost.

Low-pressure airlift fermenters are designed to couple heat and mass transfer processes using a single vehicle, viz. low-pressure air. The same air is used to supply oxygen to the micro-organisms, provide broth agitation, disperse hydrocarbon and remove heat of fermentation. Fermenter cooling is accomplished by evaporating water in the fermenter with air, which gains in humidity. Heat transfer by direct contact of air with the fermenter liquid is more efficient than indirect heat exchange which requires a finite temperature differential across the heat exchange surfaces.

The low-pressure airlift concept was first described by Cooper in two US Patents in 1975 (Cooper, 1975a, b; Cooper and Silver, 1975). The concept considers not just the costs for oxygen transfer but the overall costs of the fermenter, including the cost of heat removal and capital investment. In the conceptual design, the fermenter consisted of an array of draft tubes, 2 m (6 ft) high and 2·7 m (9 ft) in diameter, placed in a large shallow concrete basin, 2·4 m (8 ft) tall. It was designed to hold about 1·2 m (4 ft) depth of unaerated liquid. Air was sparged in the annulus and the liquid circulated up the annulus and down the draft tubes.

In another design (Chen, Kondis and Srinivasan, 1987; Chen *et al.*, 1987), consideration was given to the constraints on the fermenter when air was the only vehicle for heat and mass transport. To support a desired rate of cell production, a minimum superficial air velocity directly related to the rate of oxygen transfer must be maintained. When oxygen transfer is the only concern, enough air to meet this velocity requirement is sufficient. Thus, for a given diameter vessel, the required air rate per unit volume of liquid may be varied by changing the liquid height. In this situation, a deeper tank is sometimes preferred over a shallower tank (Urza and Jackson, 1975) because the power requirement does not increase linearly with liquid depth. However, if the same air is also used to cool the fermenter, the air flow rate per unit volume of liquid must also be sufficient to fulfil the evaporative cooling needs.

The amount of air required for fermenter cooling depends on the water carrying capacity of the cooling air, the fermenter temperature and the dry cell productivity. For example, if the fermenter is operating at 38–40°C, saturated air at that temperature has a water carrying capacity of 0·034–0·046 g per g of air. Ambient air at 25°C holds about 0·01 g of water per g of air. Thus to remove 12 g of water for each g of dry cell would require 280–420 litres of air, or 40–60 times the volume essential for biomass production. If the superficial air velocity needed to support the metabolic oxygen requirement for a given dry cell productivity is known, and if all the air is supplied at the bottom of the fermenter, then the minimum height of the fermenter liquid can be determined by the air rate per unit volume of liquid required for cooling. For example, if a dry cell productivity of 1 g l^{-1} h^{-1} can be met with a superficial air velocity of 350 m h^{-1}, 40–60 times the stoichiometric air requirement would require a minimum liquid height of 1·25 m and 0·83 m, respectively. Thus, the concept of using air to perform all the mass transport functions and fermenter cooling could best be practised in shallow pool or low-pressure airlift fermenters.

The conceptual design was tested in 190 l and 640 l rectangular fermenters. The 190 l tank measured 30·5 cm by 91·4 cm in cross-section and was 1·22 m high. It was divided into three sections by 51 cm high baffles. The center section housed the air sparger; the two side sections were used for liquid return. Liquid height above the sparger could be varied from 0 to 60 cm by either changing the sparger position, or by varying liquid volume in the tank.

The 640 l tank, measuring 91·4 cm by 91·4 cm in cross-section, had the same design as the 190 l tank except that the liquid height above the sparger could be varied up to 1·1 m. The 640 l tank also represents a single module of a commercial-size fermenter.

Different spargers were tested. However, the only major variable affecting the volumetric oxygen transfer coefficient (K_La) was found to be the superficial air velocity. Changing the sparger from a low-pressure drop bubble cap to a high-pressure drop fritted disc air diffuser did not increase the rate of oxygen transfer significantly. This indicated that the rate of oxygen transfer at the high air rate was determined primarily by the shear action of turbulent liquid on the rising gas bubbles, rather than by the size or shape of the primary bubbles formed by the sparger. Power consumption to support a cell productivity of 1 g l^{-1} h^{-1} was estimated to be about 0·65 kWh kg^{-1} of dry cell, a saving of more than 70% when compared to a conventional stirred-tank fermenter (Chen, Kondis and Srinivasan, 1987).

Liquid circulation and mixing characteristics

The gas holdup in the liquid in the riser section and the volumetric mass transfer coefficients are influenced by the circulation velocity of the liquid in the airlift fermenter.

For a given liquid medium, the liquid circulation velocity increases with increasing superficial gas velocity in the riser, and is also increased when the height of the liquid in the air-sparged riser column and the ratio of the cross-sectional area of the downcomer to the riser are increased (Bello, Robinson and Moo-Young, 1984).

The liquid circulation rate can be estimated experimentally with tracer techniques, for example, by injecting pulses of a concentrated KCl solution into the fermenter and measuring the changes in conductivity with distance using a pair of conductivity probes. Experimental data generally support the theoretically derived expression that the liquid circulation velocity, or more precisely, the riser linear liquid velocity, increases proportionally to the cubic root of the riser superficial gas velocity (Lee, 1987). However, experimental data also indicate that among these three variables, the ratio of the cross-sectional area of the downcomer to riser has the largest influence on the liquid circulation velocity.

Mixing in airlift fermenters is characterized by a dimensionless Bodenstein number, $Bo = U_l L_c / D_{eff}$. Where U_l is the average liquid velocity, m s^{-1}; L_c is the length of circulation path, m; and D_{eff} is the effective liquid phase axial dispersion coefficient, m^2 s^{-1}.

If we define specific mixing time as time per unit volume for a step change in concentration needed to achieve a new steady-state concentration in a batch fermenter, for a given power input the specific mixing time of an airlift fermenter, being a better mixer, is significantly shorter than that of a bubble column.

Gas holdup in airlift fermenters

Because of the importance of gas holdup in mass transfer and operation of airlift fermenters, the effects of gas distribution on gas holdup in water and various solutions are the subjects of many investigations.

Gas holdup depends mainly on gas velocity and liquid properties, including suspended solids concentration and the presence of baffles in the column (O'Dowd *et al.*, 1987), and at low gas velocities also depends on sparger geometry. At low superficial gas velocities (less than about 70 m h^{-1}), in the absence of bubble coalescence, there is a homogeneous distribution of rising bubbles throughout the riser. In this regime, gas holdup is equal to the superficial gas velocity divided by the single bubble rise velocity, which is determined by its size. The average bubble size depends on the sparger and the liquid medium, ranging from less than 1 mm with porous plates in a non-coalescing liquid to 2–4 mm in a coalescing liquid. The influence of sparger type is quite complex, depends on the coalescence behavior of the liquid and the initial bubble size. Among the variables which determine the coalescence property of a liquid medium are its surface tension and viscosity.

Additives are known to affect gas holdup because they can alter bubble size, which depends both on bubble breakup and coalescence. For pure liquids, which promote coalescence, the bubble size is mainly influenced by coalescence and not by the primary bubble size at the gas distributor, whereas when coalescence is hindered by additives, the primary bubble size distribution from the aerator is largely preserved. The addition of salts is known to hinder coalescence (Marrucci and Nicodemo, 1967; Lessard and Zieminski, 1971). Addition of organics, such as 1% ethanol, also hinders coalescence (Schuegerl, Luecke and Oels, 1977); on the other hand, the addition of an antifoaming compound to a non-coalescing medium increases its coalescing properties (Schuegerl *et al.*, 1978).

Not much is known about the effect of surface tension of the liquid on its coalescence properties. Calderbank found that droplet size is directly proportional to the 0·6 power of the interfacial tension. Thus a decrease in the surface tension of the liquid would be expected to decrease droplet size and increase gas holdup (Calderbank, 1958). However, this was not found to be the case by Hikita *et al.* (1980), probably due to the fact that droplet coalescence was also affected by other properties of the liquids. The addition of surfactants to a liquid has often been used to lower the interfacial tension, particularly between two liquid phases, such as in hydrocarbon fermentation. The effects of surfactants on the coalescence properties of the liquid and the gas holdup have not received much attention. Liquid properties which affect bubble breakup and coalescence were reviewed by Heijnen and van't Reit (1984).

Most unicellular fermentation broths behave as low-viscosity Newtonian fluids. Mycelial broths, on the other hand, are viscous non-Newtonian fluids, i.e. their apparent viscosity (the ratio of shear stress to the rate of shear) is not constant but depends on the rate of shear. High liquid viscosity promotes bubble coalescence and helps prevent the accumulation of tiny bubbles (Bukur and Patel, 1989).

A recent study (Moo-Young *et al.*, 1987) on the rheological properties of mycelial suspensions showed that not only are they non-Newtonian fluids, but some of them also exhibit properties of Bingham plastics, which exhibit a finite yield stress before flow starts.

The gas holdup of non-Newtonian carboxy methylcellulose solutions (Godbole *et al.*, 1984; Kawase and Moo-Young, 1986) is generally lower than that of water. This was also found to be the case with mycelial suspensions. In actual batch fermentation experiments the gas holdup was found to decrease with time due to cell growth (Moo-Young *et al.*, 1987). Use of a power law model to represent the rheological properties of actual fermenter broths was successful only when they had zero yield stress. Substituting the power law model with a Bingham model improved the correlation of fluids with finite yield stress (Kawase and Moo-Young, 1989).

In a more recent study, Bovonsombut, Wilhelm and Riba (1987) reinvestigated the effect of gas distributor design on gas holdup and oxygen transfer characteristics, in the region of low superficial gas velocity (less than 20 m h^{-1}). There were no strikingly new revelations. As reported by earlier investigators (Marrucci and Nicodemo, 1967; Zieminski and Whittemore, 1971), porous plate spargers which generate small bubbles need a coalescence suppressor to maintain their small size, and therefore with small-bubble-generating spargers, gas holdup and mass transfer coefficient will depend on the nature of the medium much more than with large-bubble-generating spargers. Large bubbles are much less prone to coalescence. This is true only when the superficial gas velocity is low and the bubbles generated by the sparger maintain their size throughout the riser.

Fermenter foaming is not desirable because it decreases fermenter output and thus increases capital cost. Recent studies show that higher liquid viscosity promotes bubble coalescence, although foaming mixtures have low coalescence rates (Bukur and Patel, 1989). As a result, in the bubbly flow regime, the sparger design has a definite effect on the gas holdup of foaming mixtures. However, in fully developed churn turbulent flow regimes, gas holdup is essentially independent of distributor type.

Saxena, Vadivel and Saxena (1989) observed a hysteresis in the gas holdup values when the superficial gas velocity was increased and then decreased. They attributed this to foaming. The magnitude of this hysteresis was found to decrease with increasing column diameter.

Thus the actual gas holdup in an airlift fermenter involves many complex variables and remains poorly understood. Furthermore, it can vary during the different stages of the fermentation reaction when changes in liquid properties occur.

From the mass transport point of view, airlift fermenters should be designed and operated in such a way that carry-over of air from the riser into the downcomer is kept as low as possible. Thus, the gas in the downcomer liquid contributes little to oxygen transfer and, by reducing the effective density difference between the contents of the riser and downcomer, reduces the liquid circulation rate. Thus the mixing performance of the fermenter is impaired. However, as pointed out by McManamey *et al.* (McManamey *et al.*, 1984; McManamey, Wase and Raymahasay, 1985), in some fermentations even momentary oxygen lack can very seriously affect productivity. Thus sufficient air to maintain aerobic conditions is essential in the down draft section to maintain the fermentation in an optimal condition.

Mass transfer in airlift fermenters

OXYGEN TRANSFER

Estimation of metabolic oxygen requirement

The stoichiometric oxygen requirement may be estimated from the overall stoichiometry of the conversion of a substrate to dry cells. Take yeast for example, where both theoretical analysis and experimental data show that the yield of dry cells containing about 60% crude protein from *n*-paraffin substrates such as *n*-hexadecane is about 1 g dry cells per g of *n*-hexadecane. Typically, a yeast containing about 60% crude protein has a formula composition of $C_4H_7O_2N_{0\cdot 7}$ per 100 g of dry cells (Holve, 1976). Therefore, the stoichiometry of converting 100 g of *n*-hexadecane (n-$C_{16}H_{34}$) to 100 g of dry cells may be written as:

$$0{\cdot}44n\text{-}C_{16}H_{34} + 0{\cdot}7NH_3 + 6{\cdot}5O_2 = C_4H_7O_2N_{0\cdot 7} + 3CO_2 + 5H_2O$$

That is, to produce 1 g of dry cells, 65 mmol of oxygen would be consumed. This corresponds to a minimum stoichiometric requirement of about 7 l of air per g of dry cells.

For carbohydrates, similar calculations lead to a minimum air requirement of about 2·25 l per g of dry cells, or about one-third of that required by hydrocarbons (Darlington, 1964).

Estimation of volumetric oxygen transfer coefficient, K_La

Oxygen transfer in different airlift designs has been studied by many investigators. They include the use of concentric tubes (Hatch, Cuevas and Wang, 1969; Wang, Hatch and Cuevas, 1971; Orazem and Erickson, 1979; Bello, Robinson and Moo-Young, 1985a) and external loops (Hatch, Cuevas and Wang, 1969; Onken and Weiland, 1983; Bello, Robinson and Moo-Young, 1985a; McManamey, Wase and Raymahasay, 1985).

Experimentally, K_La is often determined by measuring the response of the dissolved oxygen concentration in the liquid to a step change in the oxygen concentration of the inlet gas:

$$K_La = -d\ln(C_c - C_L)/dt$$

where C_e is the equilibrium oxygen concentration in the liquid and C_L the measured concentration. Rapid response oxygen electrodes with time constants of less than 2 s are now available (Chapman, Gibilaro and Nicnow, 1982).

If the oxygen concentration in the gas phase changes with column height, the local C_e can be estimated if the outlet gas concentration is simultaneously measured. However, the method assumes that C_L is uniform throughout the vessel at any particular time. The difficulty is that this condition is often not met in tower airlift fermenters.

In the case of an active fermentation broth the dissolved oxygen concentration in the broth depends on the oxygen transfer rate and the oxygen uptake rate by the microbes, which must be determined separately. Kamp *et al.* (1987) reported recently that the method proposed in 1967 by Bandyopadhyay and coworkers (Bandyopadhyay, Humphrey and Taguchi, 1967) was satisfactory. The method requires stopping the aeration for a very brief period to measure the oxygen uptake rate by the microbes.

Oxygen transfer coefficient

The hydrodynamics of the flowing of air in liquid columns can be characterized by several flow regimes. Two of these regimes are important to commercial-size airlift fermenters, namely the bubbly flow regime and the churn turbulent regime (Shah *et al.*, 1982).

The bubbly flow regime is characterized by almost uniformly sized bubbles with equal radial distribution. Superficial gas velocity is less than 0·05 m s^{-1} (180 m h^{-1}). The churn turbulent regime is when homogeneous dispersion cannot be maintained and channelling occurs at higher superficial gas velocities.

Volumetric mass transfer coefficients experimentally determined by different investigators varied a great deal. However, it is interesting to note that for a given fermenter and liquid medium, K_La varies linearly on a log–log scale over a reasonable range of superficial gas velocities. Since the gas holdup in the riser also increases with increasing superficial gas velocity, Kamp *et al.* (1987) concluded that gas holdup in the riser section, or the *active gas holdup*, which varies with each particular system, rather than the total gas holdup in the fermenter, is the major variable in determining the effective mass transfer coefficient, a relationship first proposed by Andrew (1982) for airlift fermenters and also found to be true experimentally in bubble columns by Akita and Yoshida (1973, 1974). Such a dependence would be expected, because both the volumetric mass transfer coefficient and gas holdup have similar forms of relationship with the superficial gas velocity in bubble columns.

A general relationship between the volumetric mass transfer coefficient and gas holdup (McManamey and Wase, 1986) was found to exist for all ratios of downcomer to riser.

Bello, Robinson and Moo-Young (1985a, b) pointed out that the main parameters that influence the riser gas holdup in airlift contactors must be taken into account in order to achieve a meaningful correlation of the data in different equipment. These parameters include the riser superficial gas velocity and the riser superficial liquid velocity, as well as the cross-sectional area ratio between downcomer and riser (A_d/A_r). Other possible parameters that they did not study include the liquid viscosity, the surface tension and the ionic concentrations.

These investigators found that the riser gas holdup and the volumetric mass transfer coefficient in their laboratory airlift fermenters depended not only on the superficial gas velocity but also on the circulating liquid velocity in the riser. For specified liquid-phase physicochemical properties, column configuration, and probably sparger type, liquid circulation velocity, U_L, is established by the downcomer-to-riser cross-sectional area ratio at a given gas rate. The values of gas holdup and volumetric mass transfer coefficient both increase with increasing gassing power input, but an increase in the circulating liquid velocity, i.e. increased A_d/A_r, causes both to decrease. Airlift mass transfer performance was poorer than that of bubble column because of the effect of the high circulating liquid velocity in the riser as well as the negligible mass transfer capabilities of the downcomer section. Extending these findings to commercial-scale airlift fermenters remains to be undertaken.

Viscosity effect and static mixer

The volumetric oxygen transfer coefficient was found to decrease with increasing viscosity and increase with the presence of a motionless static mixer in the riser, particularly with viscous non-Newtonian fluids (Stejskal and Potucek, 1985).

Gas holdup also decreased with increasing viscosity, and was further decreased with the presence of the static mixer. The static mixer created a very fine dispersion of gas bubbles in the downcomer which reduced the liquid circulation rate. This is consistent with the finding that high liquid circulation rate leads to lower volumetric oxygen transfer coefficients (Bello, Robinson and Moo-Young, 1985a, b).

Non-Newtonian fluids

Recognizing that, compared to mechanically stirred fermenters, airlift fermenters are most suitable for the cultivation of shear-sensitive microbes, a number of recent studies have been conducted on the mass transport properties of non-Newtonian fluids in airlift fermenters.

Because the flow properties of non-Newtonian fluids are not well understood, especially under turbulent flow conditions, it is not surprising that there is little quantitative information on the volumetric oxygen transfer coefficient in non-Newtonian fluids. However, measurements have been made by using simulated non-Newtonian fluids, such as carboxy methylcellulose solutions (Gbewonyo and Wang, 1983; Godbole *et al.*, 1984; Wase *et al.*, 1985; Kawase and Moo-Young, 1986). As is to be expected, the volumetric oxygen transfer coefficients in mycelial fermentation broths were lower than those in water. However, meaningful correlations with the physical properties of different types of non-Newtonian fluids remain to be developed.

HYDROCARBON TRANSPORT

In addition to oxygen transfer, the fermentation system involves the transport of substrates and essential minerals from the aqueous phase to the micro-organisms. The kinetics of the growth of micro-organisms, even with adequate supply of oxygen could be rate limited by the supply of substrates and many other factors, such as temperature, pH and toxic metabolite concentrations.

The use of hydrocarbons as the carbon source creates an additional liquid phase. The hydrocarbon phase is usually dispersed in aqueous media as droplets, which continuously coalesce and redisperse, giving rise to a size distribution (Shah *et al.*, 1972).

Except when using water-in-oil emulsion (Coty *et al.*, 1971), use of hydrocarbon substrates presents a special mass-transport problem because hydrocarbons generally have low water solubilities. For example, the solubility of dodecane in water is less than 4 p.p.b. at room temperature and, because of this, some researchers postulated that in addition to the utilization of the hydrocarbon substrate through the aqueous phase, other more direct pathways, such as direct attachment of submicron oil droplets to the microbial cells (Aiba *et al.*, 1969; Moo-Young and Shimizu, 1971; Moo-Young, Shimizu and Whitworth, 1971; Hug, Blanch and Fiechter, 1974), or direct attachment of cells to large droplets of oil, also happen (Nakahara and Erickson, 1975).

Microscopic examinatioin provided some evidence for direct contact between hydrocarbon droplets and microbial cells (Wang and Ochoa, 1972; Gutierrez, 1977). The specific growth rate was found to increase with increasing interfacial area, which was greatly affected by the presence of surface-active agents generated by the microbes during fermentation. Thus, emulsification is essential to obtain a good rate of growth. However, some artificial surfactants, while promoting hydrocarbon emulsification, could inhibit oxygen transport and thus cell growth (Tanaka and Fukui, 1971; Moo-Young and Viswanatha, 1973).

Heat transfer

Aerobic fermentation reactions are exothermic. Thus fermenter cooling is an important variable in the design of fermenters (Cooney, Rha and Tannenbaum, 1980). As mentioned earlier when discussing the design of low-pressure airlift fermenters, the amount of heat to be removed is related to the cell yield on carbon. In fact, the lower the biomass yield, the greater is the demand on oxygen and therefore the more heat to be removed, when the carbon in the substrate is oxidized to carbon dioxide.

Unlike the low-pressure airlift fermenter, most types of airlift fermenter require external cooling. The cooling duty can be provided by an internal plate heat exchanger, or by a circulation loop through an external heat exchanger.

The thermal conductivity of a gas–liquid mixture, an important property of the fermenter fluid, in bubble columns was studied by Chen and McMillan (1982) and Chen (1989). Assuming that the two-phase mixture can be treated as a homogeneous fluid possessing a unique thermal conductivity (the 'effective thermal conductivity'), they found that for bubble columns operating in the bubbly flow regime, this assumption appears to hold true for both batch and continuous flow systems. The effective thermal conductivity was found to be independent of liquid flow rate but depended strongly on the gas flow rate and the physical properties of the liquid phase, including density, viscosity and surface tension.

The presence of microbial particles in the fermenter fluid introduces unexpected effects on the rate of heat transfer from the fluid to the cooling surfaces. For example, Blakebrough and his coworkers (Blakebrough, McManamey and Tart, 1978; Blakebrough, McManamey and Walker, 1983a, b) found that in airlift fermenters, mycelial suspensions increase the value of the heat transfer coefficient in the laminar flow regime. The observed higher heat transfer coefficients were attributed to the disturbance of the liquid film near the tube wall caused by the motion of the mycelial aggregates. However, the effect is expected to be less pronounced in the turbulent flow regime.

Heat transfer data corresponding to the churn turbulent regime differ appreciably with different size columns, under constant operating conditions (Saxena, Vadivel and Saxena, 1989). Heat transfer coefficients are higher for the larger column. The exact source of this difference has not been identified and cannot be predicted by known correlations and heat transfer models. Scale-up from laboratory and pilot facilities to commercial units could be quite difficult.

Mathematical modelling and correlations

Shah *et al.* (1982) wrote an extensive critical review of a large number of correlations developed by different investigators over the past 20 years, useful for the estimation of operating parameters and the scale-up of different types of bubble columns, including airlift systems. They concluded that because of the complexity of flow characteristics of these systems, bubble columns are difficult to scale-up and most correlations are based on little data.

For less viscous fluids, flow regime characteristics have been correlated (Richardson and Zaki, 1954). Unfortunately, no systematic effort has been made to study the effect of fluid properties or the nature of the sparger on the flow regime characteristic.

Gas holdup is often very sensitive to the kind of liquid media, and the number of correlations proposed indicates that no single unified equation is available. Properties such as density, viscosity and interfacial tension are not sufficient to describe the system. A large scatter in the reported data is due mainly to the extreme sensitivity of holdup to the material system and to the trace impurities, a matter that is not well understood.

In modelling the single-cell protein process, Luttman *et al.* (1982, 1983) developed a distributed parameter model which took into consideration the mass balances of oxygen, carbon substrate and microbial cell mass in the liquid phase as well as oxygen and CO_2 in the gas phase. This distributed parameter model considers the change of process variables and the variation of the volumetric mass transfer coefficient along the column due to bubble coalescence, and assumes a twofold substrate Monod kinetics as follows:

$$\frac{dX(t)}{dt} = \mu_m X(t) \frac{S(t)}{K_s + S(t)} \times \frac{O_F(x,t)}{K_o + O_F(x,t)}$$

where X is the cell concentration; t, time; μ_m, maximum specific growth rate; S, substrate concentration; K_s, substrate saturation constant of Monod kinetics; O_F, oxygen concentration in the liquid; and K_o is the oxygen saturation constant of Monod kinetics.

The unknown parameters of the model, namely the volumetric oxygen transfer coefficient at the air inlet, K_La^E, the coalescence factor K_{ST} and the two kinetic parameters for the Monod growth kinetics, R_{omax} (maximum oxygen consumption rate, $ML^{-3}T^{-1}$) and K_o (saturation constant of Monod kinetics with regard to oxygen, ML^{-3}) (where M = mass, L = length, T = time) were calculated using experimentally determined values of longitudinal dissolved oxygen concentration profiles, oxygen consumption rates, CO_2 production rates and oxygen transfer rates, with yeast cultures in a 15 cm diameter bench-scale airlift tower reactor.

Assuming both unlimited and oxygen transfer limited conditions and Monod kinetics, the process has been simulated with this model, using an iterative procedure, on an analog–digital hybrid computer under quasi-steady-state conditions. Satisfactory values of the four unknown model parameters were obtained for a coalescence-suppressing substrate, ethanol, and a coalescence-neutral substrate, glucose.

Isaacs, Thoma and Munack (1987) proposed the use of an orthogonal collocation approximation for parameter identification and control optimization of this model. Their method allows relatively fast, yet reasonably accurate solutions and has the advantage of easy programmability. They gave an example of the use of the collocation approximation in a parameter identification task.

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