Pneumatically agitated bioreactors in industrial and environmental bioprocessing: Hydrodynamics, hydraulics, and transport phenomena

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Major aspects of design and operation of pneumatically agitated bioreactors are reviewed. The focus is on considerations that are relevant to industrial practice. Airlift bioreactors are emphasized. The treatment covers hydraulics, hydrodynamics, gas-liquid and solid-liquid mass transfer, heat transfer, mixing, and suspension. Newtonian and non-Newtonian systems are discussed. Applications in microbial fermentations, animal and plant cell culture, biotransformations with immobilized enzymes, and treatment of wastewater are outlined. Comparisons with more conventional bioreactor technologies are made. Design features for sterile processing in airlift systems are detailed. The evidence for superior performance of airlift bioreactors is overwhelming. Excellent productivities have been demonstrated with yeasts, bacteria, and filamentous fungi. Processes that produce highly viscous broths, including several biopolymer producing fermentations, have been proven in airlift devices. Similarly, many hybridoma cultures and plant cell suspensions have given good results. As a general rule, volumetric productivity of airlift bioreactors equals or betters that of conventional stirred tanks. Typically, this level of performance is achieved at substantially lower power input than in stirred vessels. Furthermore, the probability of mechanical failure and likelihood of loss of sterility are lower with airlift bioreactors. In wastewater treatment, too, airlift devices have far outperformed conventional systems. Airlift bioreactors accept higher BOD loadings, produce less sludge, and the degradation rate is faster; performance improves with increasing scale of operation. This review article includes 328 references.

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part of *FLUID MECHANICS PROBLEMS IN BIOTECHNOLOGY*, edited by R Pohorecki

ASME Reprint No AMR233 $70

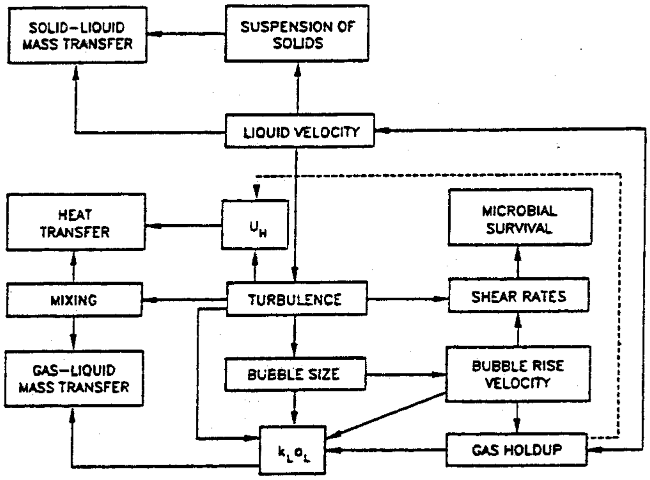

Fig 1. Interrelationships among the hydrodynamic parameters and other bioreactor performance indices.

1. INTRODUCTION

This review concentrates on hydrodynamics, related transport phenomena, and design of pneumatically agitated bioreactors. Especial emphasis is placed on airlift bioreactors as being the most important among the pneumatically agitated reactors. Design and operational features that are relevant to industrial practice are stressed. The treatment covers hydraulics, hydrodynamics, gas–liquid and solid–liquid mass transfer, heat transfer, mixing and suspension. Newtonian and non-newtonian systems are considered. Applications in microbial fermentations, animal and plant cell culture, biotransformations with immobilized enzymes and treatment of wastewater are discussed. Comparisons with more conventional bioreactor technologies are made. Design features for sterile processing are detailed.

Hydrodynamics, transport phenomena, microbial survival and production kinetics in bioreactors are interrelated. These interrelationships are quite complex as depicted in Figure 1. A variable may be directly or indirectly affected by several others in sometimes not so obvious ways. As an example, the induced liquid circulation in an airlift reactor depends on the gas holdup difference between the riser and the downcomer zones and on the geometry of the reactor. However, the gas holdup in individual zones, and the holdup difference, are in turn controlled by the rate of liquid circulation and the properties of the fluid. Liquid flow velocity and gas holdup combine to affect mixing, gas–liquid mass transfer and heat transfer. Suspension of solids and solid–liquid mass transfer are affected. Turbulence in the fluid generally determines the bubble size and hence the interfacial shear rate. Shear stresses associated with turbulence impact upon microbial morphology, floc size, and physical survival of fragile cells. The ability to heat and cool a bioreactor, supply oxygen and remove carbon dioxide depends on hydrodynamics. Ultimately the kinetics of the bioreaction are affected and so is the productivity of the reactor and the feasibility of the process (Chisti and Moo-Young, 1996). An understanding of the hydrodynamic factors and how they influence the productivity or performance is essential to bioreactor design, scale-up and operation. Insight into hydrodynamics and related phenomena is especially relevant to large scale bioprocesses for bulk products, wastewater treatment and environmental remediation. Performance and competitiveness in producing high–value–low–volume bioproducts may also be improved by attention to hydrodynamic phenomena.

1.1 Pneumatically agitated reactors

Airlift bioreactors and bubble columns are the two main types of pneumatically agitated reactors. Because of many operational advantages, airlift devices are far more significant than bubble columns in bioprocessing and bioremediation. Airlift reactors are pneumatically agitated gas–liquid and gas–liquid–solid contacting devices that are used in the chemical process industry, bioprocessing and waste treatment (Chisti, 1989; Onken and Weiland, 1983). Particularly attractive features of airlift reactors are simplicity of construction, good solid–liquid and gas–liquid mass transfer characteristics at low power consumption (Chisti, 1989; Mao *et al.*, 1992), a well-defined flow pattern, ease of long-term sterile operation, and a hydrodynamic environment suitable for fragile biocatalysts that are susceptible to physical damage by fluid turbulence or mechanical agitation. Airlift devices outperform bubble columns in suspending solids (Chisti and Moo-Young, 1994b). Evidence for superior heat transfer capability of airlift reactors relative to bubble columns has also been published (Ouyoung *et al.*, 1989). Some of these features have helped establish niche applications for airlift technology. For example, in high–intensity–low–volume activated–sludge type treatment of wastewater, high rates of oxygen transfer, mechanical simplicity and low operational costs are particularly valuable. The small volume and land area demands of airlift devices compared to conventional treatment processes are decided advantages in many locations (Moo-Young and Chisti, 1994). In certain other fermentations—production of monoclonal antibodies by

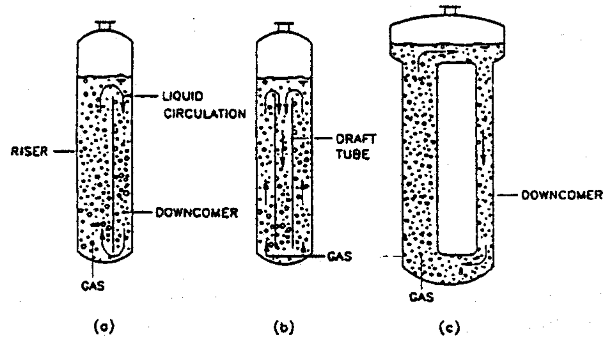

Fig 2. Airlift reactors: (a) split–cylinder internal–loop; (b) draft–tube internal–loop; and (c) external–loop.

suspension culture of hybridomas (Birch *et al.*, 1987) and plant cell culture (Doran, 1993), for instance—capabilities for sterile operation and a less damaging hydrodynamic environment are important features. New applications of airlift technology that are now being developed include anchorage–dependent culture of animal cells for biopharmaceuticals (Dai *et al.*, 1994; Ganzeveld *et al.*, 1995). Irrespective of the area of application, the predominant types of airlift reactors are the internal–loop devices shown in Figure 2. The external–loop configuration is less commonly used.

As shown in Figure 2, all airlift reactors consist of a gas–sparged riser and a downcomer which is usually not sparged. The riser and the downcomer are interconnected near the top and the bottom of the reactor. Difference in gas holdup between the gas–sparged riser and the downcomer causes a difference in bulk density of the fluid in these zones, resulting in circulation of fluid—up-flow in the riser and downflow in the downcomer (Chisti and Moo-Young, 1987a).

Literature on airlift bioreactors has been reviewed by Blenke (1979), Onken and Weiland (1983), Merchuk (1986), Chisti and Moo-Young (1987a), Chisti (1989) and Joshi *et al.* (1990). Much more information is available on bubble columns. Some major sources are Schügerl *et al.* (1977), Shah (1979), Shah *et al.* (1982), Deckwer and Schumpe (1987; 1993), and Deckwer (1992). In view of the extensive literature, this work concentrates mostly on developments since an earlier comprehensive treaties (Chisti, 1989).

2. PNEUMATIC AGITATION

2.1 Power input

The power input into a reactor controls most of its performance parameters and affects the economics of operation. In comparing the performance of bioreactors, specific power input is often used as the basis. Hence, correct calculation of power input is essential. Typically, all the energy input into pneumatically agitated reactors comes from the compressed gas. The gas imparts energy to the liquid as it expands isothermally from the hydrostatic pressure at the bottom of the reactor to the pressure in the head zone. A second source of power is the kinetic energy of the gas jet at the point of entry to the reactor. In most bioreactors, the sparger designs and the gas flow rates are such that the kinetic energy of the gas jet is negligible relative to the energy derived from isothermal expansion (Chisti, 1989). When the kinetic energy is ignored, the specific power input, or the power input per unit volume of the liquid (or slurry) is given by

$$\frac{P_G}{V_L} = \rho_L g U_G, \tag{1}$$

for bubble columns, and

$$\frac{P_G}{V_L} = \frac{\rho_L g U_{Gr}}{1+\frac{A_d}{A_r}}, \tag{2}$$

for airlift bioreactors. In equations (1) and (2) the superficial gas velocity (U_G or U_{Gr}) is actual height–averaged value, rather than a point value at any axial location (Chisti, 1989). The specific power input in pneumatically agitated bioreactors does not normally exceed 3 $kW{\cdot}m^{-3}$; a value of 2 $kW{\cdot}m^{-3}$ being a more typical maximum. In animal cell culture applications, power inputs of less than 100 $W{\cdot}m^{-3}$ are the norm.

2.2 Gas velocity

The actual height–averaged superficial gas velocity is calculated with the equation (Chisti, 1989)

$$U_G = \frac{Q_m RT}{h_L A \rho_L g}\ln\left(1+\frac{\rho_L g h_L}{P_h}\right), \tag{3}$$

which applies to bubble columns. In equation (3) Q_m is the molar flow rate of the gas, h_L is the static height of the gas–free liquid, R is the gas constant, T is the absolute temperature, A is the cross–sectional area of the column, and P_h is the pressure in the head zone. Equation (3) assumes ideal gas behavior which is an acceptable approximation in most cases.

Equation (3) may be used also for an airlift reactor to calculate a hypothetical superficial gas velocity (U_G) that can be converted to the superficial velocity (U_{Gr}) in the riser:

$$U_{Gr} = \frac{U_G A}{A_r}, \tag{4}$$

where A is the sum of the cross–sectional areas of the riser and the downcomer.

2.3 Pressurized operation

As seen in equation (3), correct calculation of the prevailing superficial gas velocity requires that the pressure (P_h) in the head zone of the reactor is taken into account. Two identical reactors operated at different headspace pressures but the same molar gas flow rate would have different specific power inputs. Sometimes the 'superficial gas velocity' reported in the literature does not correct for pressure as in equation (3). In such cases apparent effects of the headspace pressure on mixing time, gas holdup and other variables that are dependent on power input may be observed (Chisti and Moo-Young, 1987a). Correct calculation of the superficial gas velocity and power input eliminates these apparent effects. An example of the apparent pressure effect on oxygen transfer coefficients in draft–tube sparged airlift reactors was reported by Wu *et al.* (1992). Variations in headspace pressure over 1–2.36 atmospheres affected the mass transfer coefficients but the effect disappeared when the gas velocity was corrected for pressure as recommended by Chisti (1989, pp. 287–290).

3. GAS HOLDUP

3.1 Bubble columns

The volume fraction of gas in dispersion, or gas holdup (ε_G), is an important characteristic of bubble columns and other gas–liquid reactors. By definition,

$$\varepsilon_G = \frac{V_G}{V_G + V_L}, \tag{5}$$

where V_G and V_L are the volumes of the gas and the liquid (or slurry) phases respectively.. The reactor must be designed to accommodate the maximum expected gas holdup. The holdup affects many other performance characteristics in a reactor. Gas–liquid interfacial area for mass transfer is controlled by a combination of gas holdup and the mean bubble diameter. In bubble columns,

$$a_L = \frac{6\varepsilon_G}{d_B(1-\varepsilon_G)}, \tag{6}$$

where a_L is the gas–liquid interfacial area per unit volume of liquid, and d_B is the diameter of a sphere having the same surface–to–volume ratio as the gas bubble. In practice a distribution of bubble sizes is encountered and the mean diameter is calculated using

$$d_B = \frac{\sum_{i=1}^{N} n_i d_{Bi}^3}{\sum_{i=1}^{N} n_i d_{Bi}^2}, \tag{7}$$

where n_i is the frequency of occurrence of bubbles with a diameter d_{Bi}. Bubble size and frequency distributions may be measured by electrical resistivity probes and other similar sensors (Choi and Lee, 1990a). The efficiency of utilization of the gas depends on gas holdup because the contact time between the two phases is holdup dependent (Chisti, 1989, pp. 176–180),

$$t_G = \frac{h_L \varepsilon_G}{U_G(1-\varepsilon_G)}, \tag{8}$$

where t_G is the mean residence time of the gas in liquid. For columns larger than about 0.1 m diameter, gas holdup is independent of column diameter and the overall holdup is independent of the height of liquid (h_L). Hence, for a given superficial gas velocity, the gas phase residence time increases with column height. Aspect ratios of four or higher are not uncommon in bubble column bioreactors. In columns operated batchwise with respect to the liquid, the gas holdup can be analytically shown to depend on the mean rise velocity (U_b) of the bubbles,

$$\varepsilon_G = \frac{U_G}{U_b}. \tag{9}$$

Hence, at identical gas flow rates, sparger–fluid combinations that generate larger bubbles, or allow them to exist, may give rise to lower gas holdups than systems that produce smaller bubbles. In bubble columns gas holdup shows the following general dependence on gas velocity

$$\varepsilon_G = a U_G^b. \tag{10}$$

The parameters a and b depend on the flow regime and the properties of the fluid. As shown in Figure 3, for air–water a and b have been found to be 2.47 and 0.97, and 0.49 and 0.46, respectively, for bubble ($U_G < 0.05\ \mathrm{m{\cdot}s^{-1}}$) and churn-flow ($U_G > 0.05\ \mathrm{m{\cdot}s^{-1}}$) regimes in broad ranges of bubble column geometries (Chisti, 1989, pp. 161–164; Chisti and Moo-Young, 1988b). Addition of inorganic salts to water enhances gas holdup by a few percent up to an ionic strength corresponding to about 0.15 M sodium chloride (Chisti and Moo-Young, 1991; Kelkar *et al.*, 1983). This effect is due to coalescence inhibition that results from electrical repulsion between like ions at the surfaces of bubbles (Chisti and Moo-Young, 1991). For any given ionic strength, the type of ion does not influence gas holdup (Hikita *et al.*, 1980).

Numerous gas holdup correlations are available for bubble columns (Deckwer, 1992; Chisti, 1989). An example is the correlation (Akita and Yoshida, 1973)

$$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} = c\left(\frac{g d_o^2 \rho_L}{\sigma_L}\right)^{\frac{1}{8}}\left(\frac{g d_o^3 \rho_L^2}{\mu_L^2}\right)^{\frac{1}{12}}\left(\frac{U_G}{\sqrt{g d_o}}\right), \tag{11}$$

where c is either 0.20 (non-electrolytes) or 0.25 (electrolytes). Equation (11) covers column diameters (d_o) of 0.152 to 0.60 m and gas velocities from 0.004 to 0.33

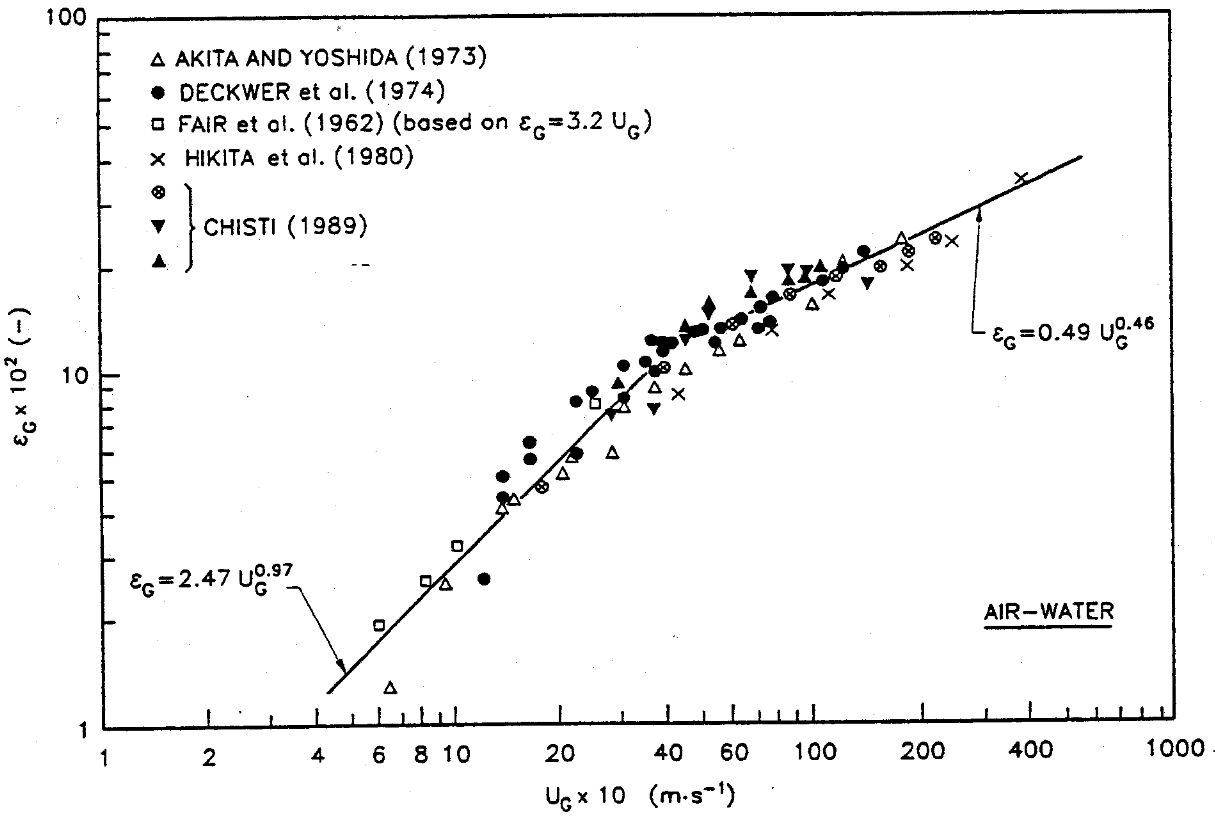

Fig 3. Gas holdup versus superficial gas velocity in bubble columns. The data shown cover column diameters of 0.10–1.067 m, and heights of 1.37–5.875 m (Chisti, 1989, p. 164).

m·s^{-1}. It applies to newtonian media such as water, glycerol and methanol.

In homogeneous non–newtonian systems the following equation may apply

$$\varepsilon_G = 0.207 U_G^{0.6} \mu_{ap}^{-0.19}. \tag{12}$$

Equation (12) was developed by Godbole *et al.* (1984) for aqueous carboxymethyl cellulose solutions (n = 0.440–0.697; K = 7.683–0.095 Pa·s^n) in churn–turbulent flow regime when the superficial gas velocity did not exceed 0.25 m·s^{-1}. A higher coefficient of 0.255 has been reported for equation (12) in presence of 0.8 M sodium sulfate.

The commonly employed shear rate expression γ = 5000·U_G, due to Nishikawa and coworkers (1977), was used by Godbole *et al.* (1984) to calculate the μ_{ap} in equation (12). Although this approach is widely used, it is fundamentally unsound and shear rates in bubble columns and other pneumatically agitated reactors are difficult to define as clearly demonstrated by Chisti and Moo-Young (1989).

As for newtonian media, the variation of gas holdup with gas velocity in non–newtonian fluids is generally consistent with equation (10). The exponent b seems to depend only on the flow regime and the n-value of the fluid, while the coefficient a is sensitive to the properties of the fluid (K, n, ρ_L and σ_L) as well as the flow regime (Chisti and Moo-Young, 1988c). Little experimental data exists on the relationship between a and the characteristics of the fluid. The exponent b has been observed to obey the empirical equation (Chisti and Moo-Young, 1988c)

$$b = 0.564 n^{-0.354}, \tag{13}$$

as illustrated in Figure 4. Equation (13) is only a rough guide which disregards any effects of the flow regime; nevertheless, the equation is based on data on a variety of

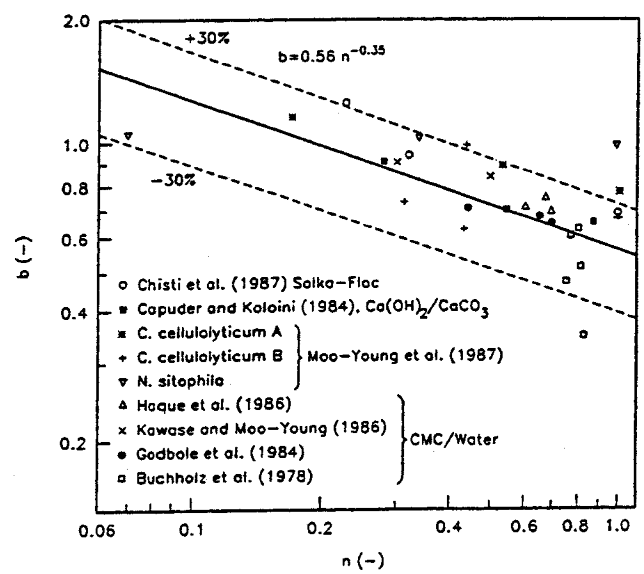

Fig 4. The exponent b (equation 10) versus the flow behavior index n for non-newtonian pseudoplastic media in pneumatically agitated reactors (Chisti and Moo-Young, 1988c).

fluids including fermentation broths of the fungi *Chaetomium cellulolyticum* and *Neurospora sitophila*. Additional data on gas holdup in slurries that simulate mycelial fermentation fluids are available elsewhere (Chisti and Moo-Young, 1988c; Chisti, 1989, pp. 149–201).

3.2 Airlift reactors

Using newtonian media (μ_L = 1–266 mPa·s) and a pseudoplastic fluid (K = 0.0256 Pa·s$^{0.902}$) in an external-loop airlift reactor (0.138 m^3; $A_r/A_d = 1$; aspect ratio ≈ 58; U_{Gr} = 0.002–0.12 m·s^{-1}), Akita *et al.* (1994) established that the gas holdup correlation proposed by Akita and Yoshida (1973) for bubble columns (see equation 11) applied to the riser of the airlift reactor when the superficial gas velocity (U_G) was replaced with $[U_{Gr} - (U_{Lr}\cdot\varepsilon_{Gr}/(1-\varepsilon_{Gr}))]$ as the correlating parameter; thus,

$$\frac{\varepsilon_{Gr}}{(1-\varepsilon_{Gr})^4} = \frac{c}{\sqrt{gd_r}}\left(\frac{gd_r^2\rho_L}{\sigma}\right)^{1/8}\left(\frac{gd_r^3\rho_L^2}{\mu_L^2}\right)^{1/12} \times\left(U_{Gr}-\frac{U_{Lr}\varepsilon_{Gr}}{1-\varepsilon_{Gr}}\right), \tag{14}$$

where c is 0.20 for non-electrolytes and 0.25 for electrolytes (Akita and Yoshida, 1973). The riser diameter d_r in equation (14) has no impact on the value of gas holdup. For water–like fluids, no gas was carried into the downcomer, because of an effective gas–liquid separator; however, in more viscous media, complete disengagement of the tiny gas bubbles could not be attained (Akita *et al.*, 1994).

For pseudoplastic aqueous solutions of carboxymethyl cellulose (K = 0.286–11.5 Pa·s^n; n = 0.441–0.617) in a draft–tube sparged internal–loop airlift reactor, Li *et al.* (1995) reported the following equations for gas holdup:

$$\varepsilon_{Gr} = 0.441\,U_{Gr}^{0.841}\,\mu_{ap}^{-0.135}, \tag{15}$$

and

$$\varepsilon_{Gd} = 0.297\,U_{Gr}^{0.935}\,\mu_{ap}^{-0.107}. \tag{16}$$

The apparent viscosity in equations (15) and (16) was calculated by assuming the shear rate to equal $5000\cdot U_{Gr}$ for $U_{Gr} \geq 0.04$ m·s^{-1}, or $1000\cdot(U_{Gr})^{0.5}$ for $U_{Gr} < 0.04$ m·s^{-1}. The apparent viscosity ranged over 0.020–0.85 Pa·s. At 0.055 m^3 the reactor was quite small; other geometric details were: $A_r/A_d = 0.618$ and aspect ratio ~ 26.

For newtonian and non-newtonian media in bubble columns and airlift bioreactors, Kawase *et al.* (1995) proposed the following equation for gas holdup in the riser:

$$\frac{\varepsilon_{Gr}}{1-\varepsilon_{Gr}} = \frac{(U_{Gr}/n)^{\frac{n+2}{2(n+1)}}}{2^{\frac{(3n+5)}{(n+1)}}\left(K/\rho_L\right)^{\frac{1}{2(n+1)}} g^{\frac{n}{2(n+1)}}\left(1+\frac{A_d}{A_r}\right)^{\frac{3(n+2)}{4(n+1)}}}. \tag{17}$$

Development of equation (17) assumed isotropic turbulence with the requisite energy dissipation being calculated as recommended by Chisti and Moo-Young (1989). As major limitations, equation (17) does not account for effects of liquid circulation on gas holdup, nor does it consider such factors as interfacial tension and solids concentration which affect holdup. Moreover, the 'theoretical' foundation of the equation is questionable because the assumption of isotropic turbulence in bubble columns and airlift reactors is not valid even at high energy inputs in water–like media (Lübbert and Larson, 1990; Lübbert *et al.*, 1990; Okada *et al.*, 1993). The supposed mechanistic foundation of equation (17) is especially doubtful when applied to fluids that are more viscous than water. Although counter claims have been made (Kawase and Moo-Young, 1990), they do not wholly support the existence of isotropic turbulence.

In an external–loop airlift reactor (A_r/A_d = 1.0; aspect ratio = 28) that achieved complete gas–liquid separation (that is, no gas in the downcomer) Pošarac and Petrovic (1991) conducted an exhaustive study of gas holdup. Tap water and aqueous solutions of alcohols (0.5 or 5% vol methanol; 0.5% vol n-butanol) were used. The superficial air velocity in the riser varied over 0.01–0.16 m·s^{-1} with corresponding liquid velocities (U_{Lr}) in the range 0.05–0.65 m·s^{-1}. In general, the presence of alcohols enhanced gas holdup relative to water. This effect was due to coalescence inhibitory properties of alcohols. Gas holdup in the riser followed the equation (Pošarac and Petrovic, 1991)

$$\varepsilon_{Gr} = 0.6\frac{\rho_G^{0.062}\,\rho_L^{0.069}\,\mu_G^{0.107}}{\mu_L^{0.053}\,\sigma_L^{0.185}}\frac{U_{Gr}^{0.936}}{(U_{Gr}+U_{Lr})^{0.474}}, \tag{18}$$

which agreed with 144 data points to within ±7%. Equation (18) does not included any geometry related effects because the geometry did not vary. Pošarac and Petrovic (1991) noted that the correlation proposed by Bello (1981),

$$\varepsilon_{Gr} = 0.16\left(\frac{U_{Gr}}{U_{Lr}}\right)^{0.56}\left(1+\frac{A_d}{A_r}\right), \tag{19}$$

gave consistently higher holdup values with an error that exceeded 70%. Similarly, the 'universal' equation of Kawase and Moo-Young (1987), which for newtonian fluids is

$$\varepsilon_{Gr} = 1.07\left(\frac{U_G^2}{gd_o}\right)^{1/3}, \tag{20}$$

showed yet poorer agreement with the data (Pošarac and Petrovic, 1991), overpredicting gas holdup by almost 100%.

Measurements in aqueous salt solutions (sodium chloride; potassium chloride; calcium chloride; sodium sulfate; and magnesium sulfate in the 0.01–0.2 kmol·m^{-3} concentration range) and sucrose solutions (0.5–8% wt/vol sucrose) in an external–loop airlift reactor showed that the type of salt or its concentration had very little effect on the measured holdup (Snape *et al.*, 1992), in keeping with similar results in bubble columns (Hikita *et al.*, 1980; Kelkar *et al.*, 1983). The holdup was not particularly sensitive to the concentration of sugar either; however, the liquid velocity did increase noticeably as the concentration increased from 0 to 1% wt/vol (Snape *et al.*, 1992). Further increase in sugar concentration reduced the rate of circulation apparently because of a viscosity effect. Several perforated spargers were employed for aeration (5–220 holes; 0.5–3.0 mm hole diameter; 0.2–0.5% free area). For the most part, the type of sparger did not influence the gas holdup or the liquid circulation rate (Snape *et al.*, 1992). When any sparger influences were observed, they were confined to the low aeration rate range (U_{Gr} = 0.03–0.09 $m \cdot s^{-1}$). Such influences disappeared as the aeration rate increased (Snape *et al.*, 1992) and as the turbulence in the fluid became the controlling influence on bubble size. The external–loop reactor used was 0.065 m^3; the other geometric details were: $A_r/A_d = 9$; and aspect ratio based on riser = 12.5 (Snape *et al.*, 1992).

In water, aqueous salt solution (10 $kg \cdot m^{-3}$ sodium chloride) and an algal culture medium, Fraser and Hill (1993) observed almost identical gas holdup values at identical conditions in the three media. An external–loop airlift reactor was used ($A_r/A_d = 2.3$; aspect ratio based on riser ≈ 18). The perforated plate sparger used could be rotated at various speeds. Compared to stationary operation, the rotating sparger (*ca* 430 rpm) produced somewhat higher gas holdup (Fraser and Hill, 1993), but not sufficiently high to justify the additional complexity of the rotating device. Liquid circulation rates in all media showed excellent agreement with equation (51) developed by Chisti *et al.* (1988).

For the external–loop airlift configuration shown in Figure 5, Ghirardini *et al.* (1992) obtained the following empirical equations

$$U_{Lr} = 6.87\, U_{Gr}^{0.34} F^{0.27} d_r^{0.41}, \tag{21}$$

$$\varepsilon_G = 0.55\, U_{Gr}^{0.78} F^{0.20} d_r^{0.42}, \tag{22}$$

and

$$a_L = 798.37\, U_{Gr}^{0.75} d_r^{0.19}. \tag{23}$$

Equations (21) and (22) are for air–water while equation (23) is for aqueous sulfite solutions. These equations were developed for $A_r/A_d = 1.0$. The riser diameter (d_r) varied over 0.025–0.1 m; the aspect ratio range used was 20–80, and the ratio of the separator–to–downcomer diameters varied over 3–6. The equations applied over a superficial air velocity range of 0.2–1.2 $m \cdot s^{-1}$. The 'filling factor,' F, was defined as the ratio of the non-aerated liquid volume in the separator to the unaerated volume in the loop. The F-values varied over 0.1–0.32.

The usefulness of equations (21)–(23) is limited outside the ranges used in establishing them and for configurations other than the one shown in Figure 5. Surprisingly, the specific interfacial area (a_L) was unaffected by the fill factor (equation 23) even though the fill volume influenced the overall gas holdup (equation 22).

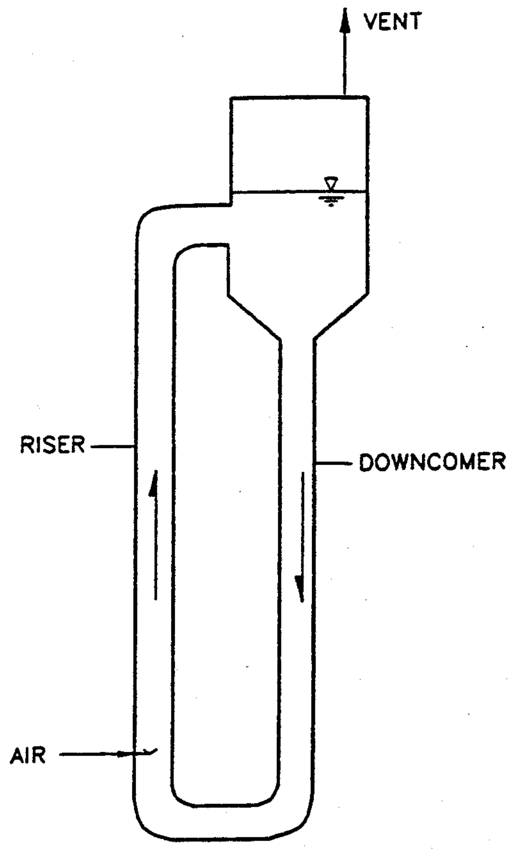

Fig 5. The external–loop airlift configuration used by Ghirardini *et al.* (1992).

The common use of the drift–flux model of Zuber and Findlay in correlating the riser gas holdup in air–water system in external–loop airlift reactors has been noted (Glennon *et al.*, 1993). The Zuber and Findlay model can be written as

$$\varepsilon_{Gr} = \frac{U_{Gr}}{\alpha(U_{Gr} + U_{Lr}) + \beta}, \tag{24}$$

where α, the distribution parameter, has typically ranged over 1.03–1.2; and β, the terminal rise velocity of a bubble, has varied over 0.25–0.36 $m \cdot s^{-1}$ (Glennon *et al.*, 1993). An α-value of close to unity indicates fairly flat radial velocity and gas holdup profiles in the riser (Chisti, 1989, p. 56); more parabolic profiles are obtained for higher values of α. The parameters α and β depend on the flow regime and the properties of the fluid.

Using measurements in two large external–loop airlift reactors (80 L, $A_r/A_d = 1$, aspect ratio = 19.5; 700 L, A_r/A_d = 1 or 1.8, aspect ratio (based on riser) = 35.9) and additional data from other sources, Kemblowski *et al.* (1993)

established the following empirical correlation for the riser gas holdup in external–loop reactors

$$\varepsilon_{Gr} = 0.203\frac{\mathrm{Fr}^{0.31}}{\mathrm{Mo}^{0.012}}\left(\frac{U_{Gr}\,A_r}{U_{Lr}\,A_d}\right)^{0.74}, \tag{25}$$

where the Morton number (Mo) and the Froude number (Fr) are, respectively,

$$\mathrm{Mo}=\frac{g(\rho_L-\rho_G)}{\sigma_L^3\rho_L^2}K^4\left(\frac{8U_{Lr}}{d_r}\right)^{4(n-1)}\left(\frac{3n+1}{4n}\right)^{4n}, \tag{26}$$

and

$$\mathrm{Fr}=\frac{(U_{Lr}+U_{Gr})^2}{gd_r}. \tag{27}$$

Equation (25) agreed with the data to within 20%; it covered the ranges: ε_{Gr} = 0.002–0.21; U_{Gr} = 0.001–0.50 m·s^{-1}; U_{Lr} = 0.07–1.3 m·s^{-1}; Fr = 0.005–14.1; Mo = 2.47 × 10^{-11}–0.39; and A_d/A_r = 0.11–1. Although not specifically developed or tested for internal–loop airlift reactors, equation (25) could well apply to them: the equation incorporates the liquid circulation velocity and, hence, indirectly, the effects of geometry.

Performance of a draft–tube sparged internal–loop airlift reactor (5 L; details unknown) that generated a fine-bubble dispersion was studied by Yuan *et al.* (1994). Experiments were done in water, aqueous ethanol (1%) and baker's yeast broth. Aeration was through a nucleopore membrane (3–5 μm pore size) sandwiched between a nylon wire screen and a perforated plate (Yuan *et al.*, 1994). Extremely small bubbles, 100–200 μm diameter, were produced in the alcohol solution; at 380 μm, the bubble size was somewhat larger in water. Because of large specific interfacial areas, the values of the gas–liquid volumetric mass transfer coefficient (k_La_L) were high. Thus, during yeast fermentations the mean value of k_La_L was about 0.46 s^{-1} (aeration rate of 1 vvm). In comparison, the average k_La_L during fermentation in a stirred tank (600 rpm agitation; 1 vvm aeration rate) was only 0.13 s^{-1}. When the membrane distributor in the airlift reactor was replaced with a 'jet' distributor, the k_La_L declined to 0.1 s^{-1} even though the reactor was aerated at 2.5 vvm compared to the 1 vvm with the membrane sparger. Despite these impressive results, membrane aerators with fine pore sizes are not suited to industrial fermentations because of potential difficulties with clogging.

The data presented showed that the overall gas holdup values were similar in water and the aqueous ethanol solution, but much lower in the yeast broth (Yuan *et al.*, 1994). The maximum gas holdup in aqueous ethanol was about 0.6; it was less than 0.2 in the yeast broth. Liquid circulation rates were about 4-fold higher in air–water than in the broth; the values in the ethanol solution were similar to those in the yeast broth. An interpretation of these results is that in the non-coalescing media (yeast broth; ethanol solution), many tiny bubbles must have recirculated into the downcomer, hence the driving force for liquid circulation was lower.

In draft–tube sparged concentric tube airlift reactors (A_r/A_d = 0.2–1.3; aspect ratios ≈ 10) operated with tap water, aqueous *n*-butanol (0.5% wt) and aqueous glycerol (46% wt), Petrovic *et al.* (1991) did not see much effect of the A_r/A_d ratio in water–like media. In the more viscous glycerol solution, there seemed to be an optimum A_r/A_d ratio of about 1/3 for maximum holdup (Petrovic *et al.*, 1991). The overall holdup agreed closely with the following correlation of Koide *et al.* (1983b)

$$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} = 0.124\left(\frac{U_G\mu_L}{\sigma_L}\right)^{0.996}\left(\frac{\rho_L\sigma_L^3}{g\mu_L^4}\right)^{0.294} \times\left(\frac{d_i}{d_o}\right)^{0.114}\left(1-0.276\left(1-\mathrm{e}^{-0.0386\mathrm{Ma}}\right)\right)^{-1}, \tag{28}$$

where Ma is the Marrucci's bubble coalescence parameter (equation 29) that characterizes the frothing ability of liquids (for non-frothing pure liquids Ma = 0). Koide *et al.* (1983b) developed equation (28) for the following ranges: 1.93 × 10^{-4} ≤ $U_G\mu_L/\sigma_L$ ≤ 2.85 × 10^{-2}; 1.31 × 10^6 ≤ $\rho_L\sigma_L^3/g\mu_L^4$ ≤ 6.04 × 10^{10}; 0.471 ≤ d_i/d_o ≤ 0.743; and 0 ≤ Ma ≤ 67.3. The equation was developed in draft–tube sparged concentric tube airlift reactors with aspect ratios of 5–15. The bubble coalescence parameter Ma is given as

$$\mathrm{Ma}=\frac{d_B}{\sigma_L}\frac{c_1}{\xi RT}\left(\frac{\mathrm{d}\sigma_L}{\mathrm{d}c_1}\right)^2\frac{(24\pi\sigma_L/\aleph d_B)^{2/3}}{\left(1+\dfrac{\mathrm{d}\ln f_1}{\mathrm{d}\ln c_1}\right)\left(1+\dfrac{x_1v_1}{x_2v_2}\right)}, \tag{29}$$

where $\aleph$ is the Hamaker constant; c_1 is the concentration of component 1 in liquid; ξ is the total number of moles of ions per mole of electrolyte (ξ is unity for non-electrolytes); f_1 is the activity coefficient of component 1 in liquid; x_1 and x_2 are mole fractions of components 1 and 2 in liquid; v_1 and v_2 are molar volumes of components 1 and 2 in liquid; R is the gas constant; and T is the absolute temperature of the liquid (Koide *et al.*, 1992).

For several newtonian media (ρ_L = 997–1179 kg·m^{-3}; μ_L = (0.894–16.81) × 10^{-3} Pa·s; σ_L = (53–72) × 10^{-3} N·m^{-1}) in draft–tube sparged concentric tube airlift devices (A_r/A_d = 0.3–1.2; aspect ratios = 5–16), Koide *et al.* (1988) recommended the following equations for gas holdup:

$$\varepsilon_{Gr} = \left(\frac{U_{Gr}}{\sqrt{gd_i}}\right) \times \left(0.415 + 4.27\left(\frac{U_{Gr}+U_{Lr}}{\sqrt{gd_i}}\right)\left(\frac{d_o^2 g\rho_L}{\sigma_L}\right)^{-0.188} + 1.13\left(\frac{U_{Gr}}{\sqrt{gd_i}}\right)^{1.22}\left(\frac{g\mu_L^4}{\rho_L\sigma_L^3}\right)^{0.0386}\right)^{-1}, \tag{30}$$

for the riser; and

$$\frac{\varepsilon_{Gd}}{(1-\varepsilon_{Gd})^5} = 279\,\varepsilon_{Gr}^{1.50}\left(\frac{A_r}{A_r+A_d}\right)^{2.98}\left(\frac{A_d}{A_r+A_d}\right)^{2.59} \times \left(\frac{L_d}{d_o}\right)^{0.352}\left(\frac{g\mu_L^4}{\rho_L\sigma_L^3}\right)^{0.0529}\left(1+0.433\left(\frac{U_{Gr}}{\sqrt{gd_i}}\right)\right)^{-1}. \tag{31}$$

for the downcomer. Equation (30) correlated 140 data within 12% of the measurements, while equation (31) applied to 109 data with an average error of less than 24%. The L_d in equation (31) was the length of the draft–tube. Equations (30) and (31) are apparently more refined forms of similar equations presented in an earlier paper by Koide *et al.* (1984a).

In an air–water system in a large (about 186 L; A_r/A_d = 0.77) draft–tube sparged internal–loop airlift device equipped with an elaborate gas–liquid separator, Cai and Nieuwstad (1991) noted that the gas holdup in the riser correlated well with the equation

$$\varepsilon_{Gr} = 2.47\,U_G^{0.97} \tag{32}$$

developed by Chisti (1989, p. 164) for bubble columns operating in the bubble flow regime. This agreement was probably because the liquid circulation velocities in the airlift reactor were relatively low (maximum U_{Lr} was less than 0.4 m·s^{-1}).

In a large draft–tube sparged concentric tube airlift reactor (A_r/A_d = 1.86; aspect ratio ≈ 11; 4 m^3) with an expanded head zone for gas liquid separation, Fröhlich *et al.* (1991a) noted that the gas holdup in the riser was independent of axial position, but the holdup declined down the downcomer. Tap water, aqueous ethanol solution (1% vol/vol ethanol) and tap water containing an antifoam agent were studied.

In studies with annulus sparged concentric–tube airlift vessels (aspect ratio = 6–15; A_d/A_r = 0.52–1.23), Koide *et al.* (1983a) obtained the following equation for overall gas holdup:

$$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} = 0.160\left(\frac{U_G\mu_L}{\sigma_L}\right)\left(\frac{\rho_L\sigma_L^3}{g\mu_L^4}\right)^{0.283}\cdot\left(\frac{d_i}{d_o}\right)^{-0.222} \times \left(\frac{\delta}{d_o}\right)^{-0.0237\exp(-0.00185\mathrm{Ma})} \times \left[1-1.61\left(1-\exp(-0.00565\mathrm{Ma})\right)\right]^{-1}, \tag{33}$$

where δ is the diameter of sparger nozzle. Equation (33) applied to the following ranges: $1.21 \times 10^{-4} \le U_G\mu_L/\sigma_L \le 1.57 \times 10^{-2}$; $1.18 \times 10^{6} \le \rho_L\sigma_L^3/g\mu_L^4 \le 5.93 \times 10^{10}$; $0.471 \le d_i/d_o \le 0.743$; $7.14 \times 10^{-3} \le \delta/d_o \le 2.86 \times 10^{-2}$; and $0 \le \mathrm{Ma} \le 93$. The average error in estimating the overall gas holdup with equation (33) was 9% for 175 measurements (Koide *et al.*, 1983a).

Clearly, many correlations have been developed for predicting gas holdup. The ones presented here supplement numerous others compiled in a previous work (Chisti, 1989). Despite this information, reliable prediction of gas holdup remains difficult. Most correlations are quite specific to particular fluid-reactor combinations and, even then, estimates should be used with caution. Two different fluids with identical densities, viscosities and interfacial tensions, may give very different results in the same reactor. For all but the simplest cases, reliable estimation of gas holdup for design and scale-up must rely on procedures such as that outlined in Section 3.2.3.

3.2.1 Relationship between riser and downcomer gas holdups

Internal–loop reactors. In airlift reactors without gas–liquid separators, the gas holdup in the downcomer depends mainly on the holdup in the riser and the velocity of the circulating liquid. Important aspects of performance of airlift reactors depend not only on the overall gas holdup, but also on the distribution of holdup between the riser and the downcomer; hence, a knowledge of this distribution is essential.

As noted earlier, the holdup distribution controls the rate of induced liquid circulation. In addition, the gas–liquid interfacial area for mass transfer and the residence time of gas in the liquid and the residence time of liquid in the riser and the downcomer are affected. Specifically, in internal–loop airlift devices, the ratio of the gas–liquid interfacial areas in the riser (A_{Lr}) and the downcomer (A_{Ld}) zones is (Chisti *et al.*, 1995)

$$\frac{A_{Lr}}{A_{Ld}} = \frac{A_r}{A_d}\cdot\frac{d_{Bd}}{d_{Br}}\cdot\frac{\varepsilon_{Gr}}{\varepsilon_{Gd}}, \tag{34}$$

where A_r and A_d are the cross–sectional areas of the riser and the downcomer, respectively. Generally, the mean bubble diameter in the downcomer (d_{Bd}) is slightly lower than that in the riser (d_{Br}) because smaller bubble are more easily and preferentially dragged into the downcomer

(Chisti, 1989, pp. 132–149). Similarly, the time spent by the recirculating liquid phase in the downcomer and the riser depend on gas holdup (Chisti *et al.*, 1995),

$$\frac{t_{Lr}}{t_{Ld}}=\frac{A_d}{A_r}\cdot\frac{(1-\varepsilon_{Gr})}{(1-\varepsilon_{Gd})}. \tag{35}$$

Because of its significance, the relationship between the gas holdups in the riser and the downcomer has received some attention in the published literature. Using air–water, Bello (1981) established the empirical relationship

$$\varepsilon_{Gd}=0.89\,\varepsilon_{Gr}, \tag{36}$$

which applied to annulus sparged concentric draft–tube airlift reactors (0.033 m^3 approximate dispersion volume; A_d/A_r = 0.13, 0.35 or 0.56; 0.152 m outer tube diameter in all cases; 1.8 m gas–liquid dispersion height; 1.55 m height of draft–tubes measured from the base of the reactors). In a much larger split–cylinder airlift device (0.23 m^3 working volume; 5.0 m liquid height; A_d/A_r = 0.411; 0.243 m column diameter; 0.229 m baffle width; 4.8 m baffle height), Chisti (1989, p. 218) found a similar correlation: Equation (37) was determined for air–water and air–salt

$$\varepsilon_{Gd}=0.997\,\varepsilon_{Gr}. \tag{37}$$

solution (0.15 M sodium chloride). Other investigations have revealed similar linear relationships between the gas holdup in the riser and that in the downcomer. Thus, there is significant empirical evidence spanning different types of airlift devices, ranges of reactor volumes and varieties of fluids, for linear dependence between the gas holdups in the riser and the downcomer. An exception was reported by Miyahara *et al.* (1986) who worked with draft–tube sparged concentric tube reactors of modest size (0.021 m^3 dispersion volume; 0.148 m outer column diameter; A_r/A_d = 0.128–0.808; 1.2 m height of dispersion; 1.0 m height of draft–tubes; ρ_L = 952–1168 kg·m^{-3}; μ_L = 1.0–14.9 mPa·s; σ_L = 34.1–72.0 mN·m^{-1}). Miyahara *et al.* (1986) observed the following relationships between gas holdups

$$\varepsilon_{Gd}=4.51\times10^{6}\,\mathrm{Mo}^{0.115}\left(\frac{A_r}{A_d}\right)^{4.2}\varepsilon_{Gr}^{4.2}, \tag{38}$$

when

$$\varepsilon_{Gr}<0.0133\left(A_r/A_d\right)^{-1.32}, \tag{39}$$

and

$$\varepsilon_{Gd}=0.05\,\mathrm{Mo}^{-0.22}\left(\left(A_r/A_d\right)^{0.6}\varepsilon_{Gr}\right)^{0.31\mathrm{Mo}^{-0.073}}, \tag{40}$$

when the gas holdup in the riser was greater than that obtained with the inequality (39). In these equations Mo is the Morton number defined as

$$\mathrm{Mo}=g\mu_L^4\big/\rho_L\sigma_L^3. \tag{41}$$

Equations (38) and (40) were established in reactors with cross–sectional areas of risers smaller than the areas of downcomers even though the opposite is the norm for practical designs of airlift reactors.

Equations (36)–(40) were empirically established. These equations are inconsistent with the observation that downcomer remains free of gas (that is ε_{Gd} = 0) until some gas holdup has been builtup in the riser. An alternative, more useful relationship between the riser and the downcomer gas holdups is the equation

$$U_{Lr}=\left[\frac{2gh_D(\varepsilon_{Gr}-\varepsilon_{Gd})}{K_B\left(\frac{A_r}{A_d}\right)^2\frac{1}{(1-\varepsilon_{Gd})^2}}\right]^{0.5}. \tag{42}$$

Equation (42) was developed by Chisti *et al.* (1988) for the prediction of liquid circulation velocity in internal–loop airlift reactors without gas–liquid separators. The equation was shown to apply to low–viscosity, water–like, fluids over a broad range of scales. Since initial publication, the equation has been repeatedly validated by other independent investigators (Cai and Nieuwstad, 1991; Choi and Lee, 1993; Fraser and Hill, 1993; García-Calvo and Letón, 1996; Petrovic *et al.*, 1991; Pošarac and Petrovic, 1991). In equation (42) K_B is the frictional loss coefficient for the bottom zone of the reactor. For internal–loop reactors, the parameter K_B is related to the configuration of the bottom zone in accordance with equation (54) (Section 4.1.1).

A simple rearrangement of equation (42) leads to

$$\frac{U_{Lr}^2K_B(A_r/A_d)^2}{2gh_D(\varepsilon_{Gr}-\varepsilon_{Gd})(1-\varepsilon_{Gd})^2}=1. \tag{43}$$

Equation (43) is an implicit relationship between the gas holdups in the riser and the downcomer. When the superficial liquid velocity (U_{Lr}) in the riser, the height of gas–liquid dispersion (h_D) and the gas holdup (ε_{Gr}) in the riser are known, the equation may be used for an iterative solution for the downcomer gas holdup (ε_{Gd}) in a reactor of given geometry (given A_r, A_d and A_b). The iterative procedure starts with an assumed value for the downcomer gas holdup (ε_{Gd}). The left–hand–side of equation (43) is computed and compared to the right–hand–side. An agreement between the two sides indicates a satisfactory value of ε_{Gd}, the desired variable.

Unlike such previous empirical correlations as equation (36) and equation (37), equation (43) shows the effects of the geometry of airlift reactors, that is the effects of K_B, h_D, A_r and A_d, on the relationship between the gas holdups in the riser and the downcomer. Geometry of the head zone has been clearly shown to affect the gas holdup distribution (Choi *et al.*, 1995b); mixing time and the liquid circulation velocity are also strongly affected (Choi *et al.*, 1995a, b). Because equation (43) accounts for reactor geometry–associated effects, it has the potential for general applicability to internal–loop reactors irrespective of scale. Equation (43) has been confirmed for prediction of the downcomer gas holdup in five internal–loop airlift reactors when the holdup in the riser and the velocity of liquid circulation were known (Chisti *et al.*, 1995). Remarkably good agreement between the measured and the calculated gas holdups was observed as illustrated in Figure 6 for two split–cylinder devices (Chisti *et al.*, 1995).

Using split–cylinder airlift reactors, Ganzeveld *et al.* (1995) showed that the relationship between the riser and downcomer gas holdups was unaffected by the loading of animal cell microcarriers (ρ_S = 1030–1050 kg·m^{-3}; 150–300 μm carrier diameter; 0–30 kg·m^{-3} solids loading; A_r/A_d = 1; aspect ratio = 7.6 and 14.5; U_{Gr} = (0–6.7) × 10^{-3}).

The gas holdup distribution in airlift devices operated such that the initial level of fluid is below the upper edge of the draft–tube or baffle is discussed in Section 11.1.

External–loop reactors. The head region of external–loop reactors usually achieves a high degree of separation of gas and liquid, hence, the downcomer gas holdup is much lower than in the riser. With well designed separators and water-like media, the downcomer gas holdup is negligible. In other cases, linear relationships between the riser and the downcomer holdups usually exist as noted for air–water systems (Bello, 1981; Chisti, 1989, pp. 184–186) as well as more complex fluids (Chisti *et al.*, 1987; Chisti and Moo-Young, 1987a). Such relationships are sensitive to the design of gas–liquid separator. Thus, for example, the expression

$$\varepsilon_{Gd} = 0.79\,\varepsilon_{Gr} - 0.057, \tag{44}$$

developed by Bello (1981) for air–water in external–loop airlift reactors did not apply to the measurements reported by Kawase *et al.* (1995). This discrepancy was certainly because of the very different configurations of the head zones for otherwise similar reactors that were used by Kawase *et al.* (1995) and Bello (1981).

3.2.2 Systems with suspended solids

With a limited amount of data acquired in gas–liquid–solid slurries in a draft–tube sparged internal–loop configuration, Lu *et al.* (1995) obtained the following empirical equation

$$\varepsilon_{Gd} = 0.8\,\varepsilon_{Gr}\left(1+20d_p\right)\left(1+\frac{W_S}{W_L}\right)^{-0.46}, \tag{45}$$

where W_S and W_L were respectively the total weight of solid and that of the liquid in the slurry. Equation (45) was obtained with suspensions of calcium alginate beads (ρ_p = 1030 kg·m^{-3}; d_p = 1–3.6 mm; loading = 0–30% vol).

Overall gas holdup in suspensions of calcium alginate beads in draft–tube sparged internal–loop airlift reactors was reported also by Koide *et al.* (1992). The properties of the systems were: 0–20% vol solids; 1.88–3.98 mm bead diameter; μ_L = (0.894–12.5) × 10^{-3} Pa·s; A_r/A_d = 0.3–1.2; and aspect ratios = 6–16. The beads were suspended in water, aqueous glycerol, and aqueous salt solutions (0.10 or 0.27 kmol·m^{-3} barium chloride; 0.4 kmol·m^{-3} sodium sulfate). The gas holdup declined with increasing concentration of solids, but was not affected by the size of the particles (Koide *et al.*, 1992). The holdup values in water and salt solutions were similar. The holdup was correlated with the following empirical equation (Koide *et al.*, 1992)

$$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} = \frac{0.13\left(\dfrac{d_i}{d_o}\right)^{0.057} \mathrm{Mo}^{-0.27}\left(\dfrac{U_G\mu_L}{\sigma_L}\right)^{0.89}}{\left(1-4.2\phi_S^{1.69}\right)\left(1-0.369\left(1-e^{-0.046\mathrm{Ma}}\right)\right)}, \tag{46}$$

where ϕ_S is the volume fraction of solids in the gas-free system, and the superficial gas velocity (U_G) is based on the diameter d_o of the outer column. In equation (46), the bubble coalescence parameter (Ma) is defined as in equation (29) while the definition given in equation (41) is used for the Morton number (Mo). Equation (46) estimated gas holdup with an average error of less than 10% for 260 data in the experimental ranges 2.48 × 10^{-4} ≤ $U_G\mu_L/\sigma_L$ ≤ 3.24 × 10^{-2}; 1.69 × 10^{-11} ≤ Mo ≤ 6.67 × 10^{-7}; 0.471 ≤ d_i/d_o ≤ 0.743; 0 ≤ Ma ≤ 53.9; and 0 ≤ ϕ_S ≤ 0.2 (Koide *et al.*, 1992).

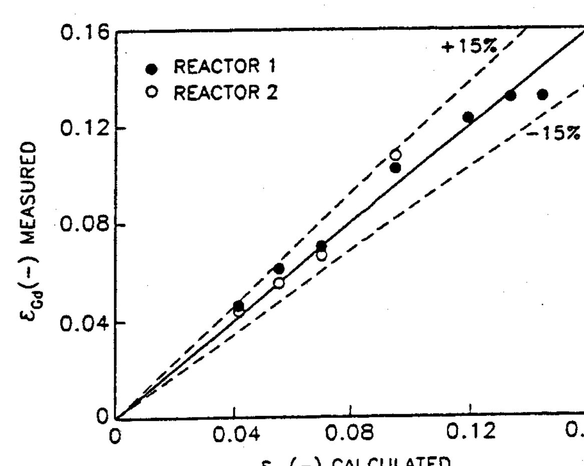

Fig 6. Measured downcomer gas holdup versus predicted values in two split–cylinder reactors. The calculated holdups were obtained with a best–fit K_B-value of 8.0. The solid line represents exact agreement. Air–water system; A_r/A_d = 2.44 (for both reactors); aspect ratios ~ 6.6 (reactor 1) and 13.2 (reactor 2); Chisti *et al.* (1995).

For suspensions of denser solids (glass or bronze spheres) in draft–tube sparged internal–loop vessels (A_r/A_d = 0.3–0.8; aspect ratio = 6–15), Koide *et al.* (1985) developed the following equation:

$$\frac{\varepsilon_G}{(1-\varepsilon_G)^4} = 0.127\left(\frac{U_G\mu_L}{\sigma_L}\right)^{0.934}\left(\frac{\rho_L\sigma_L^3}{g\mu_L^4}\right)^{0.292} \times\left(\frac{d_i}{d_o}\right)^{0.300}\left[1-0.196\left(1-\exp(-0.135\mathrm{Ma})\right)\right]^{-1} \times\left[1+0.17\left(\frac{C_S}{\rho_S}\right)^{0.091}\left(\frac{\rho_L\sigma_L^3}{g\mu_L^4}\right)^{0.043}\left(\frac{U_t}{U_G}\right)^{0.067}\right]^{-1}, \quad (47)$$

where C_S was the mean solids concentration in gas–free slurry and U_t was the terminal settling velocity of a single particle in stagnant liquid (Koide *et al.*, 1985). The average error in estimating the overall gas holdup with equation (47) was 12% for 383 data (Koide *et al.*, 1985). The equation applied to the following ranges: $3.59 \times 10^{-4} \le U_G\mu_L/\sigma_L \le 3.70 \times 10^{-2}$; $1.69 \times 10^{-11} \le g\mu_L^4/\rho_L\sigma_L^3 \le 2.55 \times 10^{-6}$; $0.471 \le d_i/d_o \le 0.743$; $0 \le \mathrm{Ma} \le 95.9$; $0 \le C_S/\rho_S \le 8.00 \times 10^{-2}$; and $1.17 \times 10^{-2} \le U_t/U_G \le 0.844$ (Koide *et al.*, 1985).

Equation (47) applied also to solids–free systems when the C_S–value was set to zero (Koide *et al.*, 1985). The physical properties ranges covered were: ρ_L = 997–1182 kg·m^{-3}; μ_L = (0.894–17.0) × 10^{-3} Pa·s; σ_L = (51.7–73.0) × 10^{-3} N·m^{-1}; and Ma = 0–95.9 (Koide *et al.*, 1985).

In suspensions of basalt particles in air–water in a draft–tube sparged internal–loop airlift reactor, Cai and Nieuwstad (1992) observed that the riser gas holdup showed excellent agreement with Chisti's (1989, p. 164) equation (32) which applied irrespective of the solids loading presumably because the loading did not exceed 4%. Similarly, the overall volumetric gas–liquid mass transfer coefficient was insensitive to the solids concentration (Cai and Nieuwstad, 1992). Details of the systems tested follow: A_r/A_d = 0.77; total volume = 0.186 m^3; ρ_S = 2760 kg·m^{-3}, d_p = 0.245 mm; the input solids loading = 27–134 kg·m^{-3} (Cai and Nieuwstad, 1992).

The overall gas holdup in a draft–tube sparged concentric draft–tube airlift reactor (A_r/A_d = 1.3; aspect ratio ≈ 5; U_{Gr} = (0.154–1.39) × 10^{-1} m·s^{-1}) for potential use in microbial desulfurization of coal was reported by Smith and Skidmore (1990). Measurements were made in a simulated basal salt medium (composition in distilled water, kg·m^{-3}: $(NH_4)_2SO_4$, 1.3; KH_2PO_4, 0.28; $MgSO_4{\cdot}2H_2O$, 0.25; $CaCl_2{\cdot}2H_2O$, 0.07; acidified to pH 2 by sulfuric acid) containing pulverized coal (74 × 10^{-6} m particle diameter; density = 1415 kg·m^{-3}) at 0–40% wt (equivalent to 0–0.315 volume fraction of solids in gas–free medium) at 30, 50 and 72°C. At any fixed gas flow rate, increases in fluid temperature led to small increases in gas holdup (Smith and Skidmore, 1990) possibly because of increased turbulence (lower viscosity). The overall holdup data were correlated with the empirical equation (Smith and Skidmore, 1990)

$$\varepsilon_G = \left(3.128 - 7.232\phi_S + 8.226\phi_S^2\right)U_{Gr}^{0.98}\,\mathrm{e}^{-284/T}, \quad (48)$$

where ϕ_S is the volume fraction of solids in the unaerated slurry and T is the absolute temperature.

Gas holdup data in a 0.35 m^3 airlift reactor of the type shown in Figure 7 has been published (Heijnen *et al.*, 1990). Over a superficial air velocity of 0.02–0.06 m·s^{-1} (based on outer tube), the gas holdup was somewhat higher than expected for bubble columns. The holdup was not significantly or systematically influenced by presence of sand particles in the 50–250 kg·m^{-3} concentration range.

Correlations for holdup of gas and solid phases in three–phase upflow by independent variation of slurry and gas flow rates were obtained by Douek *et al.* (1995). The procedure used had been detailed earlier (Chisti, 1989, pp. 269–271). For an air–water–glass beads (d_p = 500–600 μm; ρ_S = 2550 kg·m^{-3}) system the superficial gas (U_{Gr}) and liquid (U_{Lr}) velocities were independently varied over the respective ranges of 0.04–0.12 m·s^{-1} and 0.2–0.6 m·s^{-1} in a vertical pipe (diameter = 0.08 m). The gas and the solids holdups were correlated with the equations (Douek *et al.*, 1995)

$$\varepsilon_{Gr} = \frac{U_{Gr}}{1.11(U_{Gr}+U_{Lr}+U_{Sr})+0.26}, \quad (49)$$

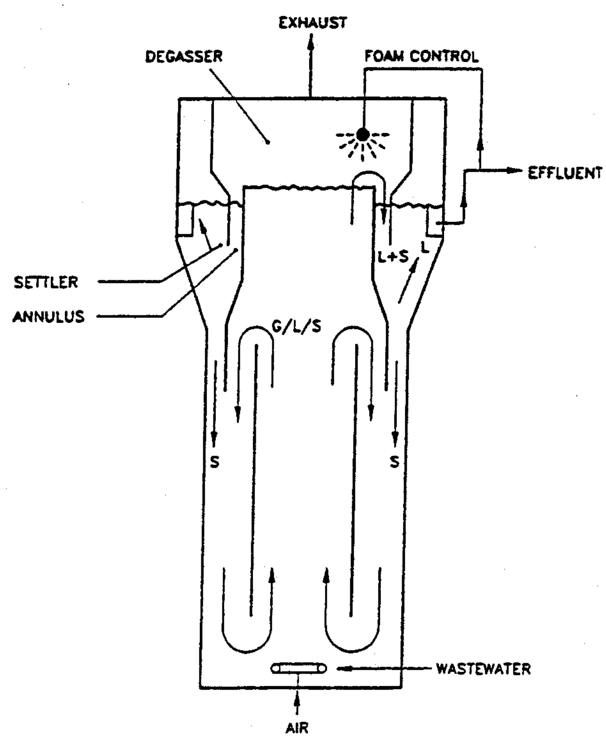

Fig 7. Microcarrier airlift bioreactor used in wastewater treatment with immobilized biofilm (Heijnen *et al.*, 1992).

and

$$\varepsilon_{Sr} = \frac{U_{Sr}}{0.93(U_{Gr} + U_{Lr} + U_{Sr}) - 0.12}, \qquad (50)$$

where U_{Sr} is the superficial velocity of solids, and ε_{Sr} is the solids holdup in the three-phase system. The velocity and holdup of solids were determined by collecting the overflow from the pipe into a bucket, quantifying the flow rate and the volumes of the individual phases. Based on equation (50), a fairly flat radial distribution of solids in the tube was inferred because the drift-flux parameter was close to unity. The constant, 0.12 $m \cdot s^{-1}$, in equation (50) amounted to the terminal settling velocity of solids in the three-phase system. As expected, this velocity was significantly higher than 0.08 $m \cdot s^{-1}$ that was estimated for settling in gas-free liquid. Correlations such as equations (49) and (50) are expected to apply to risers of airlift devices (Chisti, 1989, pp. 269–271).

3.2.3 A general approach for predicting gas holdup

As noted above and in prior reviews (Chisti and Moo-Young, 1987a; Chisti, 1989, pp. 58–65), many correlations are available for gas holdup in airlift reactors. However, as a rule, these correlations are system specific, being of little use in design or scaleup where the usual requirement is for estimation of expected performance in larger or geometrically different reactors or fluids. An alternative, scale-independent, method for prediction of gas holdup is now available (Chisti, 1989, pp. 266–272). This procedure is based on fundamental principles and it applies to the external-loop as well as the internal-loop geometric types of airlift reactors. The procedure has been demonstrated for newtonian and non-newtonian media (Chisti, 1989). Unlike most other correlations, Chisti's method employs simultaneous determination of gas holdup and the induced rate of liquid circulation. This is logical because in airlift reactors gas holdup and the velocity of liquid are interdependent: the holdup difference between the riser and the downcomer drives the circulatory flow which in turn affects the holdup (Section 4). The method works as follows: the parameters a and b in the Zuber and Findlay type expression (equation 24) are experimentally established with the fluid of interest. The parameters are determined over the range of U_{Gr} and independently varied U_{Lr} that are expected to occur in the riser of the airlift device. The experimental details and the model rig for the measurements have been described (Chisti, 1989). Once the riser gas holdup is known (equation 24) using an assumed value for U_{Lr} and a selected U_{Gr}, the downcomer gas holdup is calculated. The latter is usually a fraction of the riser gas holdup (see Section 3.2.1) if the reactor has no gas-liquid separator. If, on the other hand, an effective gas-liquid separator is used (see Section 5), the downcomer holdup is negligible. The holdup values and the assumed value of U_{Lr} are now used in equation (42) to compute a value for U_{Lr} (see Section 4). If the assumed and the calculated U_{Lr}-values agree, the corresponding gas holdups are the desired values. If no agreement is found, the iterative procedure is repeated with a new assumption for U_{Lr}. Figure 8 compares the riser gas holdup calculated using Chisti's procedure with measurements in a large annulus sparged concentric tube airlift reactor (~1 m^3; A_r/A_d = 3.61; aspect ratio = 3; air-water system).

3.3 Local events

In gas-liquid systems local properties such as bubbling frequency, bubble rise velocity, bubble diameter and gas holdup have sometimes been measured by electrical resistivity and optical probes. Several such studies are cited by Choi and Lee (1990a) who show that the design of the sensing probe has a strong influence on the measurements. Because different designs can give very different results (Choi and Lee, 1990a), probe-based measurements should be viewed with caution unless the data have been verified with an independent, well-established method. Choi and Lee (1990b) used an optical probe to measure radial distribution of gas holdup in air-water system in the annular downcomer of a concentric tube airlift reactor (~70 L; $A_r/A_d \approx 0.7$; aspect ratio ≈ 6.5; U_{Gr} = 0.05–0.20 $m \cdot s^{-1}$). The downcomer gas holdup increased with increasing gas flow rate in the riser, but the radial holdup profiles were fairly flat over the dimensionless radial distance $(r - r_i)/(r_o - r_i)$ = 0.2–0.8 for the entire range of gas flow rates (Choi and Lee, 1990b). Here, r is the radial position in the annulus, r_i is the outer radius of the draft-tube, and r_o is the radius of the outer tube. For all gas flow rates, the cross-sectional averaged local gas holdup declined down the downcomer (Choi and Lee, 1990b). The optical probe preferentially measured smaller bubbles (Choi and Lee, 1990b); thus, the reported axial decline in holdup was probably less pronounced than in the real situation. This was likely because smaller bubbles are known to penetrate deeper into the downcomer so that the proportion of smaller bubbles increases axially downward (Chisti, 1989; Chisti and Moo-Young, 1987a); consequently, higher than actual holdups are likely to have been measured in the lower zones. Data

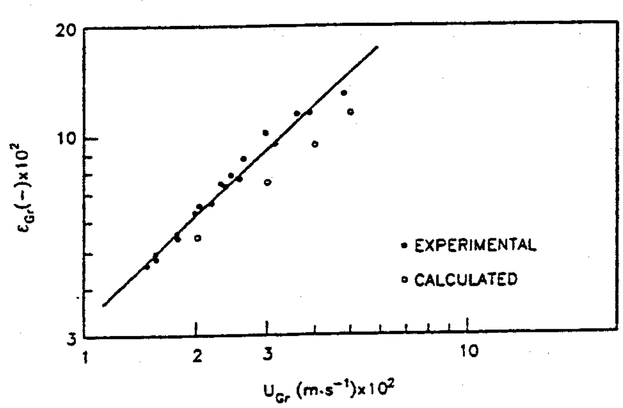

Fig 8. Measured and calculated riser gas holdups versus superficial gas velocity in air-water in a *ca* 1 m^3 airlift reactor. The line is based on the measurements (Chisti, 1989, p. 267).