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Gas Holdup Behaviour in Fermentation Broths and Other Non-Newtonian Fluids in Pneumatically Agitated Reactors

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ABSTRACT

Knowledge of gas holdup under various operating conditions is essential for the design and operation of gas–liquid–solid multiphase bioreactors. A hydrodynamic analysis of the gas holdup behaviour in pneumatically agitated bioreactors has been carried out for rheologically complex fluids. Gas holdup expressions of the form $\epsilon = \alpha U_G^{\beta}$ *are shown to have a fundamental basis. The dependence of the parameter* β *on the flow behaviour index* n *is elucidated for pseudoplastic media including several broths of commercially valuable mycelial fungal organisms. For all fluids,* β *increases with declining* n*. The implications of this phenomenon in bioreactor design are discussed.*

1. INTRODUCTION

Gas–liquid (or slurry) reactors are commonly used in chemical- and biotechnology-based industrial processes. Manufacture of antibiotics, biological treatment of wastes and hybridoma cell culture technology are some examples of established and more recently developed processes which are carried out in gas–slurry bioreactors [1] such as bubble columns and airlift devices. The volume fraction of gas in dispersion, or gas holdup, which exists under given operating conditions is an essential parameter [2, 3] for the design of all gas–liquid–solid reactors. The importance of gas holdup is multifold. The holdup determines the residence time of the gas in the liquid, and in combination with the bubble size, it controls the gas–liquid interfacial area available for mass transfer. Gas holdup predetermines reactor design also because the total design volume of the reactor for any range of operating conditions depends on the maximum holdup that must be accommodated. Furthermore, the magnitude of liquid circulation in airlift [4, 5] and bubble column [6] bioreactors is dependent on gas holdup. Table 1 provides a summary of the relationships between the various bioreactor performance indicators and gas holdup.

Because of their importance the gas holdup characteristics of pneumatically agitated reactors (bubble columns and airlifts) have been studied [1], but mainly in water-like fluids. Some theoretical insight into the gas holdup phenomena in low-viscosity newtonian media is also available [7]. However, for non-newtonian fluids such as biological suspensions of *Penicillia* and *Aspergilli* and solid-substrate fermentation broths (*e.g.* cellulose fibre slurries), the fundamentals of gas holdup behaviour are unknown. In this paper we analyse the gas–slurry hydrodynamics of non-newtonian systems and demonstrate the applicability of the concepts developed to several fluids including fermentation broths.

2. THEORY

Gas holdup may be expressed analytically [8 - 10] as the ratio of superficial gas velocity U_G to the mean terminal rise velocity U_T of the gas bubbles:

$$\epsilon = \frac{U_G}{U_T} \tag{1}$$

Equation (1) arises from the definition [11] of gas holdup:

$$\epsilon = \frac{A_G}{A} \tag{2}$$

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TABLE 1

Dependence of bioreactor performance on gas holdup

Parameter	*Reactor*	*Dependence on gas holdup*	*References*
Mass transfer k_La_L	Airlift, bubble column	$k_La_L = \psi \dfrac{\epsilon}{1-\epsilon}$ ψ = constant (s^{-1})	1
Gas–liquid specific interfacial area a_L	Airlift, bubble column	$a_L = \dfrac{6\epsilon}{d_B(1-\epsilon)}$	1
Dispersion (or reactor) volume V_D	Airlift, bubble column	$V_D = \dfrac{V_L}{1-\epsilon}$	20
Gas-phase residence time t_R	Bubble column	$t_R = \dfrac{h_L\epsilon}{U_G(1-\epsilon)}$	20
Liquid circulation velocity U_{Lr}	Airlift	$U_{Lr} = \left[\dfrac{2gh_D(\epsilon_r-\epsilon_d)}{K_T/(1-\epsilon_r)^2 + \{K_B/(1-\epsilon_d)^2\}(A_r/A_d)^2}\right]^{0.5}$	4

where A_G and A are the actual or the true cross-sectional area for gas flow and the total cross-section of the gas–liquid flow channel respectively, and from the continuity equation:

$$A_G U_T = A U_G \tag{3}$$

The actual or terminal rise velocity of the gas bubbles may be found by equating the buoyancy and the drag forces. The buoyancy force F_B on a bubble is [12]

$$F_B = (\rho_L - \rho_G)\frac{\pi}{6} d_B^3 g \tag{4}$$

where d_B is the Sauter mean bubble diameter. The drag force F_D is

$$F_D = \frac{C_D U_T^2 \rho_L A_p}{2} \tag{5}$$

assuming the bubble to be a rigid particle as a first approximation [12]. A_p and C_D are the projected area of the bubble and a dimensionless drag coefficient respectively. The substitution of

$$A_p = \frac{\pi d_B^2}{4} \tag{6}$$

into eqn. (5) gives

$$F_D = \frac{\pi C_D U_T^2 \rho_L d_B^2}{8} \tag{7}$$

Equating eqns. (4) and (7) followed by rearrangement leads to

$$U_T^2 = \frac{4(\rho_L - \rho_G) g d_B}{3\rho_L C_D} \tag{8}$$

The general form of the drag coefficient dependence on the particle Reynolds number may be expressed as [12]

$$C_D = \frac{i}{\mathrm{Re_p}^j} \tag{9}$$

where i and j depend on the flow regime. For non-newtonian flow a multiplier $\dot{Z}$ is incorporated [13, 14] in i in eqn. (9). The particle or the bubble Reynolds number is

$$\mathrm{Re_p} = \frac{(\rho_L - \rho_G) U_T d_B}{\mu_{Ls}} \tag{10}$$

where μ_{Ls} is the bubble–fluid interfacial viscosity. For power-law fluids, which are of interest here, a widely used (*e.g.* by El-Temtamy *et al.* [15]) expression for μ_{Ls} is

$$\mu_{Ls} = cK\left(\frac{U_T}{d_B}\right)^{n-1} \tag{11}$$

where c is a proportionality constant. Equations (10) and (11) lead to

$$\mathrm{Re_p} = \frac{(\rho_L - \rho_G) U_T^{2-n} d_B^n}{cK} \tag{12}$$

Recently, Walter and Blanch [16] showed that the bubble size in non-newtonian systems is given by

$$d_B \propto \frac{\sigma^{0.6}}{(P_G/V_L)^{0.4}\rho_L^{0.2}}\left(\frac{\mu_{ap}}{\mu_G}\right)^{0.1} \tag{13}$$

which is based on Kolmogoroff's turbulence theory. The shear rate in pneumatically agitated reactors has been correlated [17 - 19] by equations of the type

$$\dot{\gamma} = zU_G \tag{14}$$

and hence the apparent viscosity μ_{ap} for use in eqn. (13) becomes

$$\mu_{ap} = K(zU_G)^{n-1} \tag{15}$$

The shear rate expressions used in eqns. (11) and (15) are different because the interfacial shear, which influences the bubble rise velocity, and the shear fields in the bulk fluid, which determine the bubble size, are two very different phenomena [20]. Substitution of eqns. (9), (12), (13) and (15) along with the specific power input to the reactor

$$\frac{P_G}{V_L} = \rho_L g U_G \tag{16}$$

into eqn. (8) leads to the rather complex expression

$$U_T = \left\{\frac{4(\rho_L-\rho_G)^{1+j}g}{3\rho_L i\dot{Z}(cK)^j}\right\}\left\{f\frac{\sigma^{0.6}}{\rho_L^{0.6}g^{0.4}} \times \left(K\frac{z^{n-1}}{\mu_G}\right)^{0.1}\right\}^{(1+nj)/[2-(2-n)j]} U_G^{\delta} \tag{17}$$

with f being a dimensionless proportionality constant and δ standing for

$$\delta = \frac{\{0.1(n-1)-0.4\}(1+nj)}{2-(2-n)j} \tag{18}$$

Substitution of eqns. (17) and (18) in eqn. (1) leads to a correlation of the form

$$\epsilon = \alpha U_G^{\beta}$$

Gas holdup equations of this type have been derived from experiments by a number of investigators (Chisti and Moo-Young [1, 7] give many examples), although the hydroxynamical reasoning behind them was not established prior to this work.

For non-newtonian fluids, α and β in the above equation are respectively

$$\alpha = \left[\left\{\frac{4(\rho_L-\rho_G)^{1+j}g}{3\rho_L i\dot{Z}(cK)^j}\right\}^{1/[2-(2-n)j]}\left\{\frac{f\sigma^{0.6}}{\rho_L^{0.6}g^{0.4}} \times \left(K\frac{z^{n-1}}{\mu_G}\right)^{0.1}\right\}^{(1+nj)/[2-(2-n)j]}\right]^{-1} \tag{19}$$

and

$$\beta = 1-\delta = \frac{1-\{0.1(n-1)-0.4\}(1+nj)}{2-(2-n)j} \tag{20}$$

3. DISCUSSION

The expressions for α and β, although complex, nevertheless point to some important characteristics of these parameters: β is found to be dependent only on the flow regime (via j) and on the flow behaviour index of the fluid, whereas the parameter α is dependent not only on the flow regime, but also on n, K, the fluid densities and on the gravitational field.

In view of the prediction (eqn. (20)) that the exponent β on the gas velocity U_G in gas holdup correlations of the type $\epsilon = \alpha U_G^{\beta}$ should depend on n and j, we attempt to correlate the available empirical β and n data without regard to the prevailing flow regime (*i.e.* j). Thus, for several pseudoplastic fluids, including homogeneous media [21 - 24], cellulose fibre [2] and other slurries [25] and fermentation broths [26] of the fungi *Chaetomium cellulolyticum* and *Neurospora sitophila*, the values of β found in various bubble columns were calculated. These are shown in Table 2 along with the flow index n of the fluids. The data presented in Table 2 were correlated (Fig. 1) according to

$$\beta = 0.564n^{-0.354} \tag{21}$$

with a correlation coefficient of 0.73. This was satisfactory considering the large amount of data (24 results), the extremely large variety of fluids (biological and abiological suspensions and homogeneous solutions), the broad range of n and the fact that the possible flow regime variations were disregarded because they could not be calculated.

The increase in β (*i.e.* in the dependence of gas holdup on gas velocity) with increasing pseudoplasticity (*i.e.* decreasing n) of fermen-

TABLE 2

The hydrodynamic parameters of some non-newtonian fluids

Fluid	*K* (Pa s^n)	*n*	β	*Reference*
$Ca(OH)_2/CaCO_3$ suspensions (wt.%)				
5	0.0035	0.87	0.6482	[25]
10	0.037	0.54	0.6986	
15	0.33	0.28	0.9129	
Solka-Floc in 0.15 M NaCl (dry wt./vol.% SF)				
1	—	≈ 1	0.683	[2]
2	1.464	0.322	0.943	
3	6.127	0.237	1.268	
C. cellulolyticum A Hours since inoculation				
0	—	—	0.7651	[26]
36	0.131	0.53	0.8804	
90	2.95	0.17	1.1533	
C. cellulolyticum B Hours since inoculation				
0	—	—	0.6757	[26]
18	0.27	0.43	0.9816	
24	1.22	0.31	0.7251	
48	0.24	0.43	0.6309	
N. sitophila Hours since inoculation				
0	—	—	0.9808	[26]
23	0.62	0.34	1.0042	
91	0.97	0.07	1.0652	
CMC in water (wt.%)				
0.5	0.061	0.70	0.6862	[21]
1.0	0.102	0.67	0.74	
1.5	0.350	0.61	0.704	
CMC in water (wt.%)				
1	1.25	0.82	0.3398	[22]
1.2	1.6	0.81	0.5066	
1.4	2.5	0.8	0.6239	
1.56	3.0	0.778	0.6025	
1.7	4.45	0.754	0.4656	
CMC in water	7.683	0.440	0.706	[23] (β calculated using eqn. (23) of reference)
	0.184	0.654	0.666	
	0.095	0.697	0.658	
CMC in water + other fluids	—	0.3	0.896	[24] (based on eqn. (10) of reference)
	—	0.5	0.84	

tation fluids and other media is obvious in Fig. 1. Because the specific agitation power in a pneumatically mixed bioreactor is directly dependent on superficial gas velocity (*e.g.*, eqn. (16) for bubble columns), the sensitivity of gas holdup to bioreactor power consumption also depends on the flow behaviour index of the fluid being processed. Furthermore, in batch operations such as those in which a polymer is secreted or biomass is produced, the flow index is time dependent and consequently the relationship between power and gas holdup (and parameters dependent on holdup (Table 1)) changes continually. As a result, considerations of interrelationships between β and n and also the time history of n during a fermentation can be of significant importance for bioreactor design.

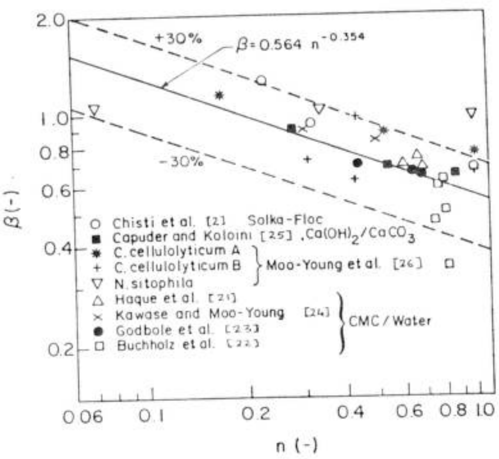

Fig. 1. The influence of flow behaviour index on gas holdup equations: the exponent β (eqn. (21)) *vs.* the flow behaviour index n of non-newtonian pseudoplastic media in pneumatically agitated reactors; β declines with n for an extremely large variety of fluids; the exponent β for the data of Buchholz *et al.* [22] was not included in the regression, to test the applicability of the equation to a data set not used in obtaining it.

4. CONCLUSIONS

According to hydrodynamic reasoning, the gas holdup in pneumatically mixed bioreactors shows a power-law-type dependence on gas velocity: $\epsilon = U_G^{\beta}$. The parameter β depends on the flow index n. A quantitative description of this behaviour, which is generally followed by a variety of media, is provided for use in bioreactor engineering.

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APPENDIX A: NOMENCLATURE

A	total cross-sectional area of reactor (m^2)
A_d	downcomer cross-sectional area (m^2)
A_G	actual cross-sectional area used for gas flow (m^2)
A_p	projected area of a bubble (m^2)
A_r	riser cross-sectional area (m^2)
a_L	interfacial area per unit liquid volume (m^{-1})
C_D	drag coefficient
CMC	carboxymethyl cellulose
c	constant of proportionality
d_B	Sauter mean bubble diameter (m)
F_B	buoyancy force on bubble (kg m s^{-2})
F_D	drag force on bubble (kg m s^{-2})
f	coefficient
g	gravitational acceleration (m s^{-2})
h_D	dispersion height (m)
h_L	liquid height (m)
i	a coefficient
j	a coefficient
K	consistency index (Pa s^n)
K_B	frictional loss coefficient (bottom)
K_T	frictional loss coefficient (top)
k_L	mass transfer coefficient (m s^{-1})
n	flow behaviour index
P_G	power input due to gassing (W)
Re_p	Reynolds number of bubble (eqn. 10))
SF	Solka-Floc cellulose fibre
t_R	gas-phase residence time in reactor (s)
U_G	superficial gas velocity (m s^{-1})
U_{Lr}	riser gas velocity (m s^{-1})
U_T	terminal rise velocity of bubbles (m s^{-1})
V_D	volume of dispersion (m^3)
V_L	volume of liquid (m^3)
$\dot{Z}$	multiplier for non-newtonian fluids
z	coefficient in eqn. (14) (m^{-1})

Greek symbols

α	a parameter ($m^{-\beta}$ s^{β})
β	exponent
$\dot{\gamma}$	shear rate (s^{-1})
δ	parameter defined by eqn. (18)
ϵ	overall gas holdup
ϵ_d	downcomer gas holdup
ϵ_r	riser gas holdup
μ_{ap}	apparent viscosity (Pa s)
μ_G	viscosity of gas (Pa s)
μ_{Ls}	bubble–fluid interfacial viscosity (Pa s)
ρ_G	density of gas (kg m^{-3})
ρ_L	density of liquid (kg m^{-3})
σ	interfacial tension (kg s^{-1})
ψ	k_L/d_B ratio (s^{-1})