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Some effects of rheology on the spatial distribution of gas hold-up in a mechanically agitated vessel☆

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Abstract

The effects of rheology on gas hold-up in non-Newtonian circulation flow has been rarely discussed in the literature, especially so far as local gas distribution in mechanically agitated vessels is concerned. This study analyses experimental gas dispersion behaviour for a wide range of complex-rheology conditions. Non-Newtonian aqueous solutions of xanthan gum with concentrations from 0.5 to 5 kg/m^3 and Newtonian glycerol solutions of 450 and 800 kg/m^3 containing also 5% electrolyte are studied. Hold-up is measured conductometrically at 36 spatial points within the stirred vessel mainly around and above the impeller for a conventional geometry (Rushton standard configuration with $T = 0.2$ m and $D/T = 0.33$). In addition to increasing viscosity, three aspects of rheology impact on gas hold-up and different dominances of component effects upon the overall gas dispersion mechanism are revealed. The results are presented as local and global gas hold-up vs. consistency, pseudoplasticity, and position. By comparing data obtained for Newtonian and non-Newtonian flow conditions and data obtained for the same polymer solution at constant plasticity and variable consistency, an attempt to reveal the net effects of consistency and pseudoplasticity is made. The boundaries of regions with large differences of mixing intensity reflected in terms of local gas availability are quantified. © 2002 Elsevier Science B.V. All rights reserved.

Keywords: Mixing; Gas hold-up; Non-Newtonian flow

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1. Introduction

Gas hold-up distribution is important for precise control of gas–liquid mass transfer where reaction processes take place in highly consistent media with complex rheology. It is known that such processes are most often met in polymerisation reactors and bioreactors [1]. Examples are the problems of fermentation suppression by aerobic micro-organisms when exposed to hypoxia [2] as well as the high sensitiveness of the metabolic pathways to oxygen availability in fungal bioreactors where macro- and micro-aeration are important for the control of the final bioproduct distribution (see, e.g. [3]). Such cases ideally require precise control of the local gas hold-up coupled with the knowledge of the gas–liquid flow structures.

In spite of its importance, relevant information on local point values is scarce. So far, this factor has been uncovered concisely only for water by Nienow et al. [4] and later by Nagase and Yasui [5]. Local gas hold-up in two-phase flow of highly viscous media has been mostly disregarded or treated only partially [6,7].

Possibly due to the general lack of information on properties of two-phase flow in complex-rheology fluids, studies concerning hold-up have been concentrated only on global averages [8–11]. Ranade and Ulbrecht [8] reported a decrease of overall gas hold-up in mechanically agitated CMC and PAA solutions in parallel with increasing polymer concentration which they assigned to the lower dispersion capability of drag-reducing polymers. Machon et al. [9] were the first to indicate a turning point hold-up behaviour for pseudoplastic solutions (e.g. CMC) which implied a parallel occurrence of at least two competing component effects, the one leading to ε-rise and the other leading to ε-fall. In their case, the ε-rise which prevailed at low polymer concentrations was attributed to the formation of non-coalescing small gas bubbles, while the ε-fall was explained further by generation of another class of very large bubbles which, rising fast, had very short residence times. Correlation of the results to the degree of pseudoplasticity was suggested, but was not confirmed separately because of the limited

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Nomenclature

Latin

d_b	size of bubbles (m)
D	impeller diameter (m)
K	consistency coefficient (Pa s^n)
n	flow behaviour index
N	impeller revolutions (s^{-1})
P	power (W)
Q	flow rate ($m^3\,s^{-1}$)
T	tank diameter (m)

Dimensionless parameters

P_o	power number ($= P/N^3D^5\rho$)
r/R	dimensionless radial coordinate
Re	impeller Reynolds number ($= ND^2\rho/\mu$)
z/H	dimensionless axial coordinate

Greek letters

γ	shear rate (s^{-1})
ε	gas hold-up
μ	viscosity (Pa s)
ρ	density (kg/m^3)
σ	surface tension (N m^{-1})
τ	shear stress (Pa)

Subscripts

a	apparent
av	average
G	gas
L	liquid

information obtained by the global approach involved. A local analysis is required to bring a wider comprehension. Other studies on mixing of complex-rheology fluids have considered the effects of rheology on gas hold-up only as a side effect in the analysis of power consumption or while tracing turbulent pseudo-caverns [12,13].

Regarding local gas hold-up as a practical parameter, it would be affected not only by the presence of polymer constituents, but also by electrolytes. Consequently, the physical model system adopted to be a better representation of reality for further studies should be a water-based system containing macromolecular components and electrolyte. In this study, we consider local gas hold-up in such an aqueous polymer-electrolyte system based on xanthan gum and potassium sulphate.

In what follows, the experiments were designed to reveal the effects of rheology on the gas hold-up with emphasis on the effect of pseudoplasticity. Following earlier work [4,6], we seek to study the separate effects of apparent viscosity, consistency and degree of pseudoplasticity. This has been accomplished by measuring local gas hold-up in several key regions of a turbine-agitated vessel, namely the impeller region, the wall region and the bulk region denoted by points C, W, and M, respectively (cf. Fig. 1(a)). Viscoelastic effects have been considered separately [14].

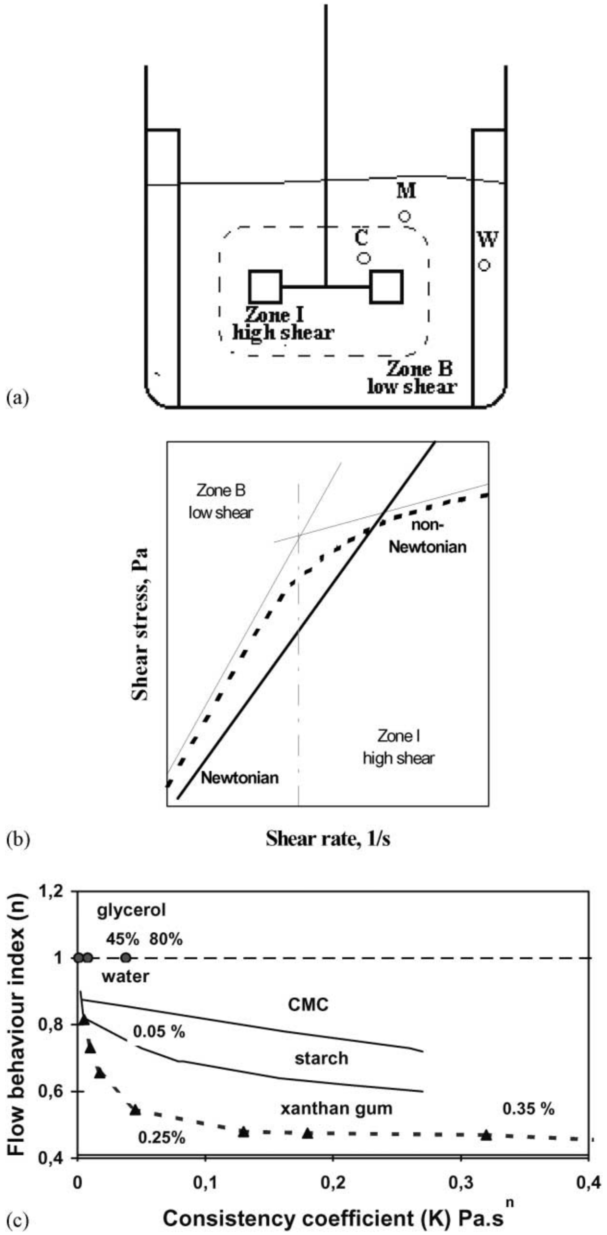

Fig. 1. Physical system: (a) vessel with zones; (b) distribution of shear and viscosity ($\mu = \tau/\gamma$) in zones; (c) power law parameters' relationships bearing shear effect on viscosity ($\mu = K\gamma^{n-1}$).

1.1. The analysis constraints

In terms of conventions from the literature, the analysis is limited to the following physical system (Fig. 1).

1. A laboratory vessel with P/V not exceeding 2 kW/m^3 and tip velocity of 1.5 m/s is used. In practice, such a case has been classified by Andrew [15] as a free bubble rise reactor.

2. In all cases mixing intensity corresponds to a regime of complete dispersion as defined by Nienow et al.[4]. However, since the bubble size distribution in polymer solutions is often bimodal, the complete dispersion regime here is defined in terms of small bubbles (up to 4 mm). During experiments, the large bubbles were seen not to disperse completely into the wall region.
3. The studies involve mainly xanthan gum colloidal dispersions. Because of the specific shape of their power law parameters' curves (Fig. 1(c)) depending on polymer concentration, these systems allow separate study of the influence of the plasticity and consistency property of the media involved. The degree of plasticity is estimated by the value of flow behaviour index n while consistency is indicated by consistency coefficient K. As seen from Fig. 1(c), depending on polymer concentration, xanthan gum solutions may exhibit nearly constant flow behaviour at $n \sim 0.52$–0.47 for a wide range of consistencies, i.e. K between 0.05 and 0.4 $\mathrm{Pa\,s}^n$, and in another range, variable flow behaviour (n between 0.9 and 0.65) at slightly variable consistencies ($K \sim 0.01\,\mathrm{Pa\,s}^n$). Thus, in contrast to carboxymethylcellulose solutions and starch solutions (whose power law parameters' curves are also illustrated in the figure), xanthan gum solutions allow a test of the effects of consistency and pseudoplasticity separately, and moreover present a more severe degree of non-Newtonian rheology.

   Aqueous solutions of glycerol 45 and 80% are used to represent the Newtonian system. For the sake of comparison, the polymer concentrations of the solutions employed (i.e. 0.05, 0.25, 0.35 and 0.5%) are chosen to exhibit apparent viscosities comparable with the viscosities of the Newtonian glycerol solutions employed (e.g. compare rows 4 and 5, and rows 1 and 6 in Table 1).
4. Elasticity is assumed negligible. For the fluids and concentrations treated, a Weissenberg number $Wi \sim 1$ is assumed to be based on the reported properties of similar xanthan gum solutions in other studies [16].
5. An analysis by DeKee et al. [17] has shown polymer solutions to hinder coalescence. Thus, non-coalescing media are tested in this paper. This analysis implies bubble generation and break-up or aggregation to prevail in the impeller region, while in the bulk bubble size distribution is assumed to change only with respect to the differences in bubble transportation (i.e. circulation and buoyancy). Such preservation of pattern could also be deduced by following bubble size distributions, as registered recently by Barigou and Greaves [18].
6. As the study considers gas hold-up distribution in a stirred vessel with non-homogeneous circulation flow, the relevance of the measurements to reported average values of N or μ is of informative value. In fact, distribution of shear and viscosity is meant which corresponds to the reported average values, $\gamma_{\mathrm{av}} = 11.6\,N$ [19], and $\mu_{\mathrm{av}} = \tau/\gamma_{\mathrm{av}}$.

Table 1
Systems studied and their rheological properties

System	x (%)	K ($\mathrm{Pa\,s}^n$)	n	μ_a (mPa s) at $N = 10\,\mathrm{s}^{-1}$
Xanthan	0.05	0.007	0.82	3
Xanthan	0.25	0.13	0.48	11
Xanthan	0.35	0.32	0.47	26
Xanthan	0.5	0.68	0.41	41
Glycerol	80	0.038	1	38
Glycerol	45	0.0055	1	5.5
Water		0.001	1	1

2. Experimental

The experimental conditions were:

1. A conventional Rushton turbine agitated reactor geometry with vessel size 0.2 m. The reactor volume was 6.3 l.
2. Measurement points, as illustrated in Fig. 1a. To account for the extent of: (a) gas recirculation; (b) bubble discharge and wall circulation; (c) gas distribution in the bulk. Measurements were concentrated at point C in the impeller region, with coordinates $z/H = 0.43$ and $r/R = 0.3$, point W in the wall region $z/H = 0.43$ and $r/R = 0.9$ and five points M in the middle of the bulk region at $z/H = 0.64$ or 0.74 and r/R variable between 0.3 and 0.9, respectively. (The two regions with significant difference of shear and circulation rates denoted as zone-I and zone-B are shown to correspond to the two linearized extremes of the pseudoplastic flow curve in Fig. 1(b).)
3. The measurements were accomplished by a conductivity probe tested previously [20,21] for sensitivity with respect to bubble size and fast changing flow direction at various degrees of turbulence. The probe performance is characterised in Fig. 2. The probe was calibrated against displacement in a quiescent uniformly bubbled column. The column diameter was 50 mm and liquid bed height was 200 mm. A porous glass distributor was used to obtain bubble populations with narrow bubble size distribution. In order to shift the dispersion to small bubbles, polyethylene glycol was added in concentration 100 g/l. The column was operated at flow rates below the critical values found to turn the gas flow pattern into a non-homogeneous one. Bubble size was found to be in the range 0.5–2.5 mm. The conductivity cell was a two-plate design with plate size 10 mm × 10 mm, plate separation 10 mm and cell constant $0.1\,\mathrm{mm}^{-1}$. It was positioned at 50, 100 and 150 mm (i.e. 0.25, 0.5 and 0.75) over the porous plate along the central axis of the vessel. Three levels of the void fraction, namely 11, 9 and 7% were set by bed expansion (the solid lines in the figure). The results are shown in Fig. 2(a). They show good coincidence (i.e. within 1% deviation) of the conductivity point values with the global displacement ones.

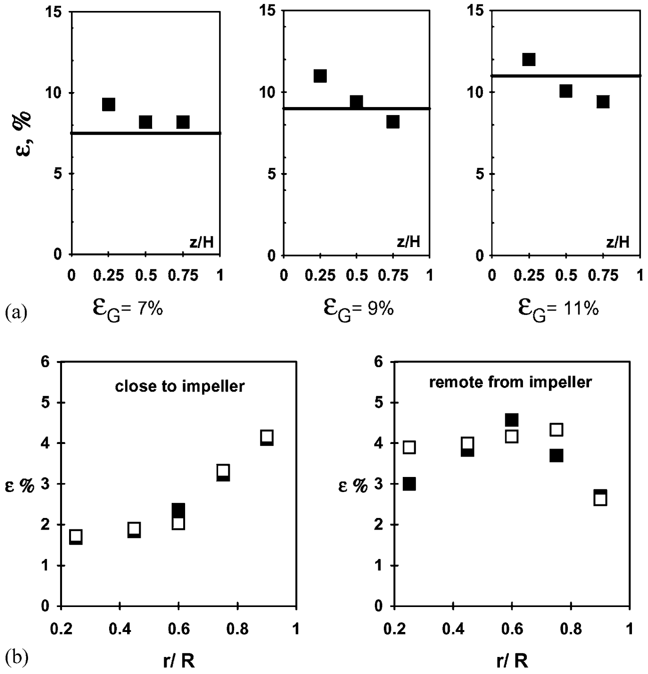

Fig. 2. Performance of probe: (a) calibration in quiescent uniformly bubbled column. Void fraction determined by displacement (solid lines) and by conductivity (■); (b) void fraction vs. radial position in CMC (1%) at 600 rpm and 800 l/h: ■ —cell 1; □—cell 2.

In order to check reproducibility of conductivity measurements exposed to bubbles of various size, the measurements of gas hold-up were repeated in 1% starch solution in the stirred vessel in two successive runs using two different cells, as follows:

- cell 1: plates 10 mm × 10 mm; distance between plates 10 mm; cell constant $0.1\,\mathrm{mm}^{-1}$,
- cell 2: plates 7 mm × 7 mm; distance between plates 7 mm; cell constant $0.14\,\mathrm{mm}^{-1}$.

Measurements were performed in zones close and remote from impeller. The profiles produced by both the cells in both the zones were essentially identical (Fig. 2(b)), in spite of the big bubbles present intermittently in both cases [21]. The mean square deviation of the individual gas hold-up measurement obtained in 55 measurements was found to be in the range 0.53–0.92%, the higher deviation being specific for the dispersions in water, while the lower value was more characteristic for pseudoplastic solutions. Based on these experiments, a probe with plate size 10 mm, plate separation 10 mm and probe constant $0.1\,\mathrm{mm}^{-1}$ was accepted for the further analysis.

4. In addition to local measurements, overall gas hold-up was calculated based on the total of 36 local hold-up measurements. These conductivity results were compared with reference data obtained from Chapman et al. [22]:

For water–electrolyte at 1 vvm and $N = 6.7$, 10, and 15 rps, it has been obtained:

ε (measured)	ε (Chapman et al.)
2.9	2.6
3.6	4
6.1	5.6

for water–glycerol at 0.5 vvm, and $N = 6.7$, 10, 15 rps, it has been obtained:

ε (measured)	ε (Chapman et al.)
2.4	2.0
2.2	2.9
2.7	3.7

The agreement is good.

5. In the studies, xanthan gum (Sigma) solutions with concentrations and parameters, as given in Table 1, were used. To separate the effect of pseudoplasticity from the effect of consistency, glycerol solutions were also tested. As seen in Fig. 1(c), a two fold increase of the degree of pseudoplasticity could be realised.
6. The following experimental conditions were tested: gas flow rates, Q_G, were varied between 200 and 1200 l/h corresponding to 0.5 up to 3 vvm in biological reactors. Impeller speed was varied between 5 and 15 rps. Vessel and impeller geometry were kept unchanged.
7. Fluid density was measured by floating-type density meter and the rheology flow curves were tested by a rotative viscometer Rheotest 2 (a Couette type).

3. Results and discussion

3.1. The effect of rheology at constant stirrer speed

Figs. 3(a) and (b), and Fig. 4 contain the relationships of gas hold-up vs. apparent viscosity at 400 and 900 rpm, respectively. The net effect of consistency K on localised and average gas hold-up is seen in Fig. 3(a) and (b) by the Newtonian glycerol curves for which $n = 1$ and K varies from 0.001 to 0.038 Pa s. In this case, gas hold-up is almost linear. In contrast, the non-Newtonian xanthan gum curves, for which both K and n vary, show extremal behaviour.

Comparing both the figures, one can see that the average gas hold-up in glycerol is the same, although the speed in Fig. 3(b) is more than twice as high as in Fig. 3(a). However, cross-sectional average gas hold-up may not always coincide with overall gas hold-up, the latter including all areas of the reactor. The cross-sectional average is determined based on 36 local values and thus some areas close to the shaft and under the impeller are excluded.

As stirrer speed N has been maintained constant, the average shear rate $\gamma_{\mathrm{av}}(= 11.6\,\mathrm{N})$ as well as its (apparent) spatial

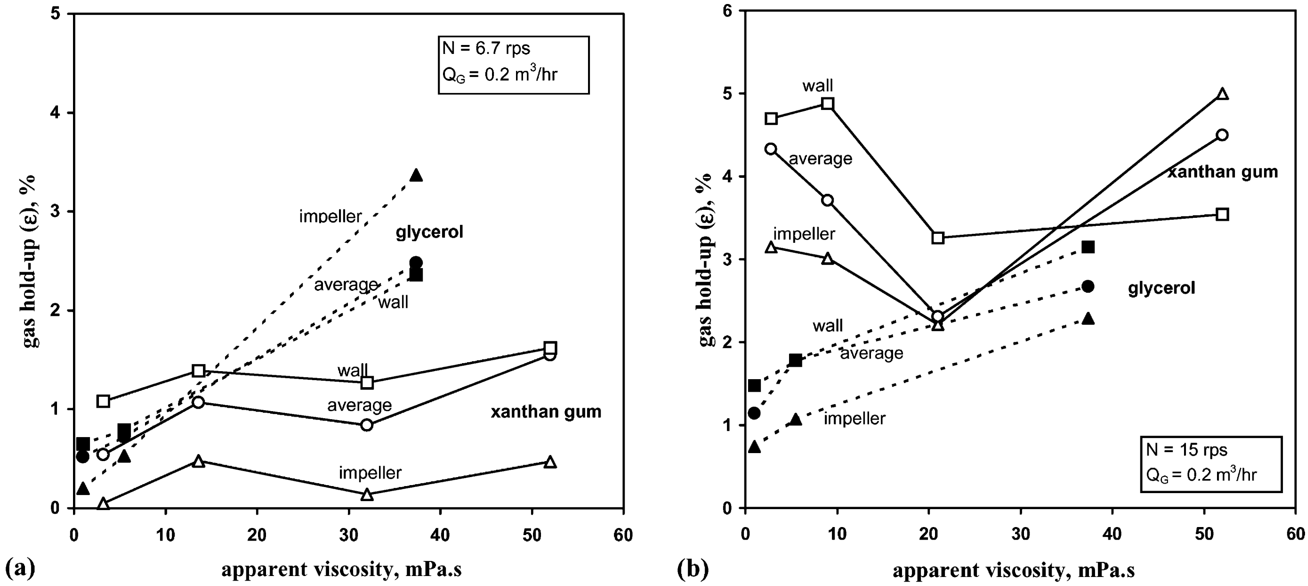

Fig. 3. Comparison of wall and impeller local gas hold-up with average gas hold-up vs. apparent viscosity in pseudoplastic xanthan gum solution for: (a) $N = 6.7$ rps and $Q_G = 0.2\,\mathrm{m^3/h}$; (b) $N = 15$ rps and $Q_G = 0.2\,\mathrm{m^3/h}$.

distribution would be the same [19] so that gas hold-up variation should be caused merely by the different fluid flow properties, i.e. K and n.

The cases can be discussed in terms of competing rates of bubble generation and transportation, both being enhanced or suppressed depending on rheology.

Discussing the initial low viscosity part of the curves in Fig. 3, bubble transportation is unhindered and ε is influenced mainly by bubble generation off the gas cavities behind the blades. At constant average shear rate γ, apparent viscosity (τ/γ) is rising in parallel to the rise of τ ($\tau = K\gamma^n$). Consequently, the rising parts of the gas hold-up curves in the figure correspond to increasing shear stress, which means dominance of bubble generation according to the condition [15,23]:

$$\tau > \frac{\sigma + \mu_G\sqrt{\tau/\rho_G}}{d_b} \qquad (1)$$

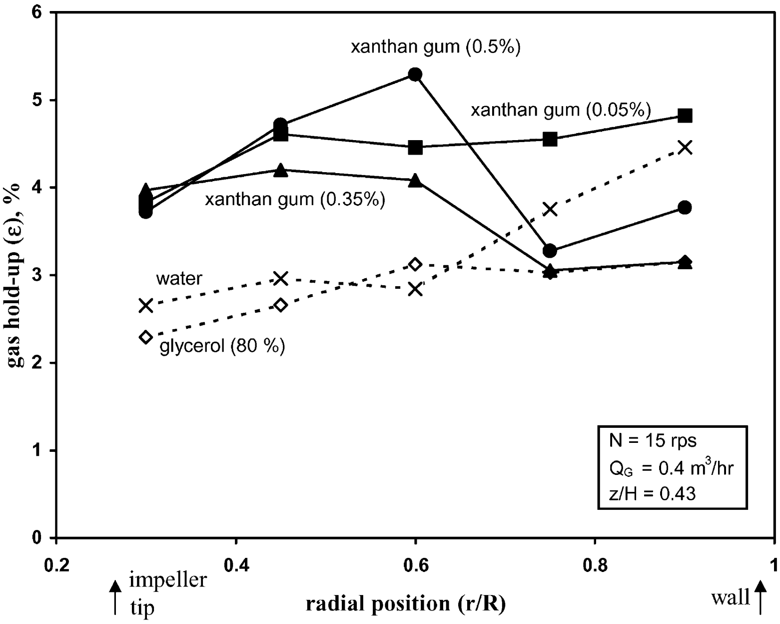

Fig. 4. Local gas hold-up vs. radial position at $z/H = 0.43$, $N = 15$ rps and $Q_G = 0.4\,\mathrm{m^3/h}$.

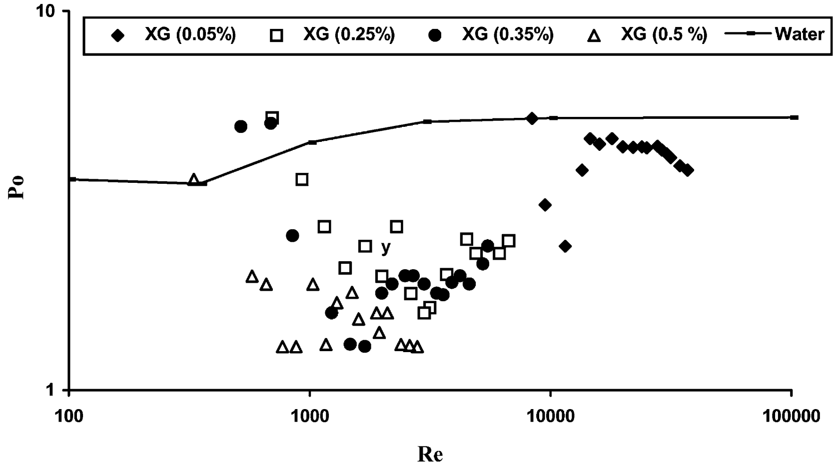

Fig. 5. Power number vs. Reynolds number for the xanthan gum and water–glycerol solutions used in the study.

In this inequality, τ is the shear stress, σ the surface tension, d_b the bubble size. μ_G and ρ_G stand for gas viscosity and density, respectively.

The descending parts of the curves in Fig. 3 may be due to drag reduction; as reported by Nienow et al. [12], the power dissipated in agitated polymer systems is being reduced depending on the concentration and plasticity of solutions. Consequently, eddies larger than d_b are formed which are not active for bubble splitting. Thus, bubble generation decreases. This is confirmed by the specific power characteristics registered for the particular polymer systems of this study. The reduction of power draw for the pseudoplastic xanthan gum solutions employed, is illustrated in Fig. 5. According to the figure, power is reduced twice when *Re* decreased between 7000 and 4000 proportional to a rise of viscosity. Thus, parallel suppression of gas dispersion by power reduction could be expected.

Discussing the final rising part of the non-Newtonian curves in Fig. 3, following the minimum gas hold-up at high viscosity, we assume that rheology changes the mechanism of liquid circulation and bubble transportation. At high *N*, such a change is evidenced in part by the increased gas hold-up in the impeller area while the gas hold-up at the wall remains the same (Fig. 3(b)). In fact, plasticity being increased twice in the viscosity range between 1 and 50 mPa s, non-Newtonian fluid circulation is not hindered equally for the different zones around the impeller. The liquid near the wall becomes more stagnant and the circulation loop around the impeller shrinks, the circulations becoming more intensive. The more concentrated the xanthan gum solutions, the higher is the degree of pseudoplasticity ($n < 0.5$) and the higher is the differences in mobility of the liquid phase in the region near the vessel wall compared to the impeller region. As a result, the gas hold-up increases in the vessel centre (point C), and decreases or remains unchanged at the vessel wall (point W).This is not the case for the Newtonian glycerol solutions where viscosity is constant throughout the vessel.

Postulated stagnancy and lower gas hold-up near the wall compared to higher circulation intensity and gas hold-up near the shaft, are seen to effect the quality of dispersion in Fig. 4. In this figure, the behaviour of local gas hold-up for a highly concentrated polymer solution (5 kg/m^3) is compared with the one obtained for low-concentrated polymer solutions (0.5–3.5 kg/m^3), glycerol and water. The formation of zones with high gas hold-up in the areas of higher liquid mobility (i.e. zone I, low *r*/*R*) and zones of low gas hold-up in the areas of low circulation intensity close to the wall (i.e. zone B, high *r*/*R*), is obvious in this figure. Gas hold-up behaviour depends largely on *K* and *n*. Indeed, for $n = 1$ and close to 1, i.e. when the liquid phase is Newtonian or weakly pseudoplastic (e.g. xanthan gum 0.5 kg/m^3), the radial profiles represent weak rising functions. In contrast, strongly pseudoplastic systems, such as xanthan gum of 3.5 and 5 kg/m^3, exhibit rising gas hold-up profiles in the area $r/R < 0.6$ and a sharp decrease of ε close to the wall, at $r/R > 0.6$. We believe that, parallel to plasticity rise, the liquid circulation loop shrinks and the liquid at the wall becomes more stagnant and receives less bubbles, while the liquid around the impeller circulates at a higher rate generating more bubbles and increasing the localised gas hold-up.

The deductions above are illustrated further by observing the changes of hold-up in three different planes along the vessel height, shown as radial profiles in Fig. 6 at *z*/*H* between 0.43 and 0.74. Fig. 6(a)– (c) contains the data for concentrated glycerol, diluted xanthan gum, and concentrated xanthan gum, respectively. The profiles for diluted xanthan gum solutions in Fig. 6(b) show the behaviour of a weak pseudoplastic substance ($K < 0.01$ Pa s^n, $n \sim 1$) The profiles in Fig. 6(c) correspond to a xanthan gum solution with strong pseudoplasticity ($K = 0.68$ Pa s^n, $n \sim 0.4$). Parallel to increasing pseudoplasticity (i.e. decreasing *n*), the locus of the maximum hold-up of these profiles is seen to shift from the wall region (Fig. 6(b)) to the shaft (Fig. 6(c)). As seen in the case of strong plasticity ($n < 0.5$), a zone with high gas content is formed near the impeller and a zone of low gas content is formed nearer to the wall. Besides, comparing different levels in Fig. 6(c), gas hold-up is seen to decrease from $z/H = 0.64$ to 0.74, i.e. from the impeller to the surface. Judging on the local gas hold-up distribution, the boundaries of a turbulent pseudo-cavern could be identified through the coordinates of locations marking the decrease of local gas hold-up values. No such distribution was observed for glycerol or diluted xanthan gum dispersions.

To generalise, the gas hold-up differences and formation of zones in constantly agitated pseudoplastic media can be explained by the fast (up to 200%) increase of plasticity beyond the impeller bound area towards the bulk zone and the wall. The effect is accompanied by decrease of power draw and shift from turbulent to transient mixing conditions Both effects result in poor liquid and gas circulation outside the impeller area.

Because the effects are due to the distribution of shear, it is interesting to see the effect of rheology in the context of variable mixing intensity.

3.2. The effect of rheology at variable stirrer speed

For the analysis of rheology effects to be complete, the effect of pseudoplasticity has been considered in the context of mixing intensity. Local gas hold-up variation at $z/H = 0.74$ for glycerol and xanthan gum solutions, $Q_G = 0.4\,m^3/h$, and for two different values of stirrer speed, namely 10 and

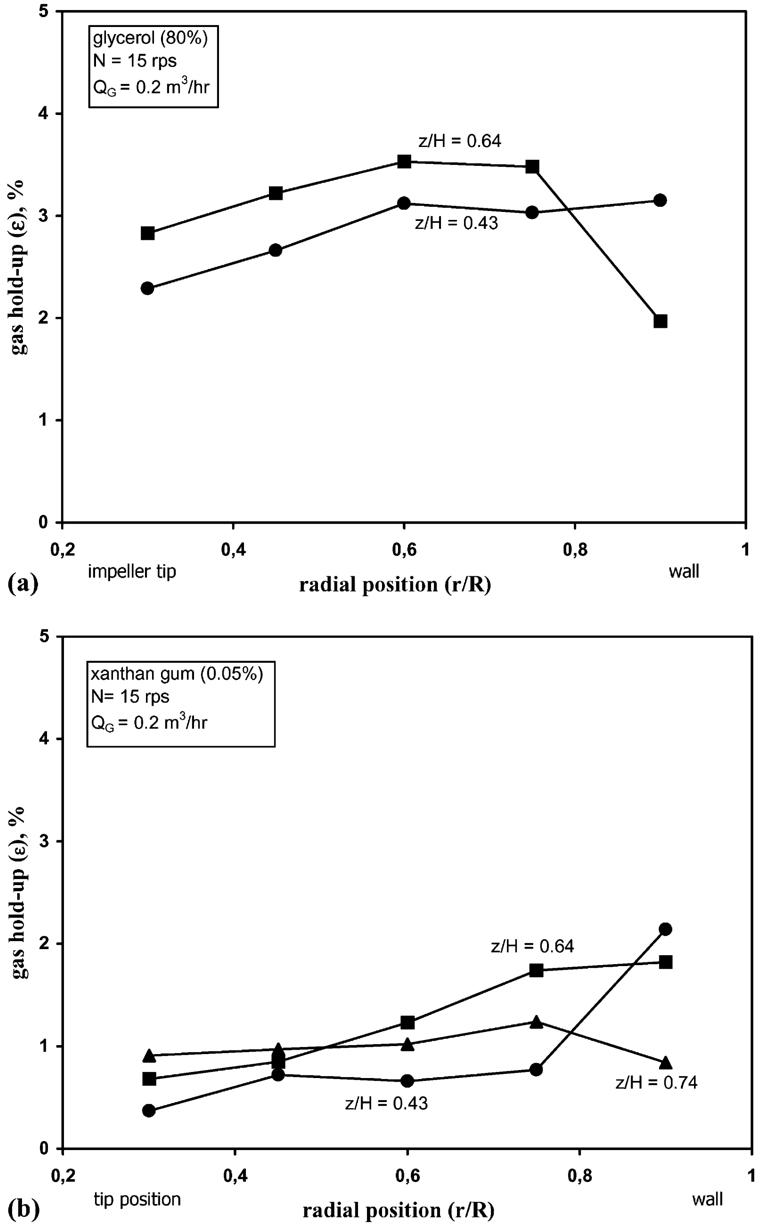

Fig. 6. Local gas hold-up vs. radial position for three axial heights in: (a) glycerol (80%); (b) xanthan gum (0.05%); (c) xanthan gum (0.5%).

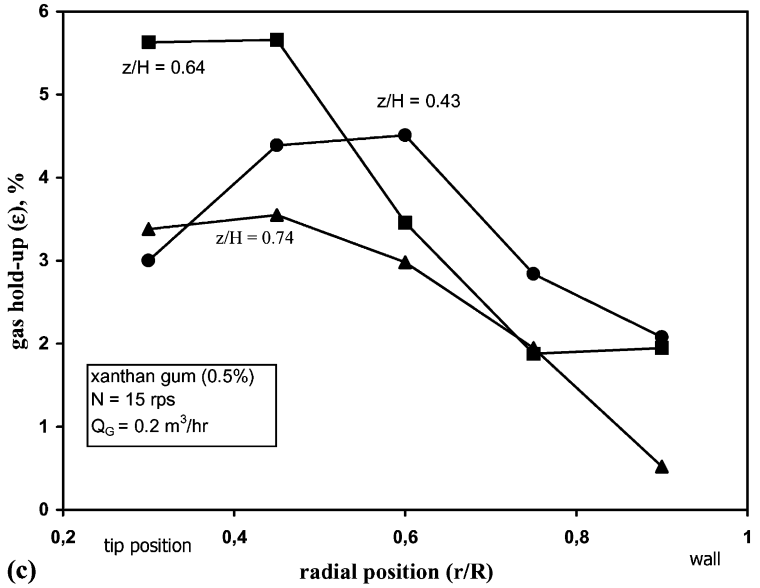

Fig. 6. (*Continued*).

15 rps, are plotted in Fig. 7. Fig. 7(a) contains the data for strong pseudoplasticity (xanthan gum 5 kg/m^3 (solid lines) and glycerol 800 kg/m^3 (the dotted lines). Fig. 7(b) contains the data for xanthan gum 0.5 kg/m^3 and glycerol 450 kg/m^3 for 10 and 15 rps.

The higher the stirrer speed *N*, the higher is the gas hold-up. It is seen that:

1. There is a remarkable difference between the radial profiles of the two types of fluids, i.e. Newtonian and non-Newtonian. Furthermore, there is a clear difference between the performance of fluids with various degrees of pseudoplasticity. For example, in the case of strong pseudoplasticity (xanthan gum 0.5% in Fig. 7(a)), one can identify clearly the formation of two zones, one below $r/R = 0.6$ with high gas content, and one nearer to the wall, with low gas hold-up. In contrast, the curves obtained for glycerol and weak non-Newtonian fluids (xanthan gum 0.05%) are relatively flat.
2. The higher the speed, the greater is the differences between the zones.

To facilitate an analysis, we have calculated apparent viscosity for three different values of fluid tangential velocities across the vessel radius. The adopted velocity range refers to the three-dimensional numerical analysis of flow behaviour in an agitated pseudoplastic liquid carried out by Kaminoyama et al. [24] and more recently to the LDA measurements of tangential velocity distributions carried out by Hirata et al. [13]. These authors show independently that one could expect a three fold velocity decrease in the area between the impeller tip and the wall. Table 2 contains these calculations. Clearly, a velocity change of 1:2 may bring spatial viscosity variation of 18% for $n = 0.8$ and 63% for $n = 0.4$ which is a significant difference. The higher the mixing velocity, the higher is the difference of apparent viscosity and the higher is the difference of gas hold-up due to bubble transportation. Also the lower the *n*-values, the is higher is the difference of gas hold-up in Fig. 7.

3.3. Comparison with other data for complex rheology systems

Using the above data, we should examine these results relative to other studies. It is seen clearly that zone I (described by C) outlined in this study for high polymer concentrations is identical to a turbulent pseudo-cavern described by Nienow and Elson [25]. One can see that the appearance of this zone can be well quantified by the value of the τ/γ ratio corresponding to the start of hindered bubble transportation. In this particular study, the value of 25–30 mPa s

Table 2

Point	Conditions		Xanthan gum 0.05% ($K = 0.01$ Pa s^n, $n = 0.8$)			Xanthan gum 0.55% ($K = 1$ Pa s^n, $n = 0.4$)		
	N (min^{-1})	γ (s^{-1})	τ (Pa s^n)	μ_a (mPa s)	*Re*	τ (Pa s^n)	μ_a (mPa s)	*Re*
C	900	172	0.61	3.5	19000	7.8	45	1500
Intermediate	600	115	0.45	3.9	12000	6.7	58	800
W	400	77	0.32	4.1	7000	5.7	74	400

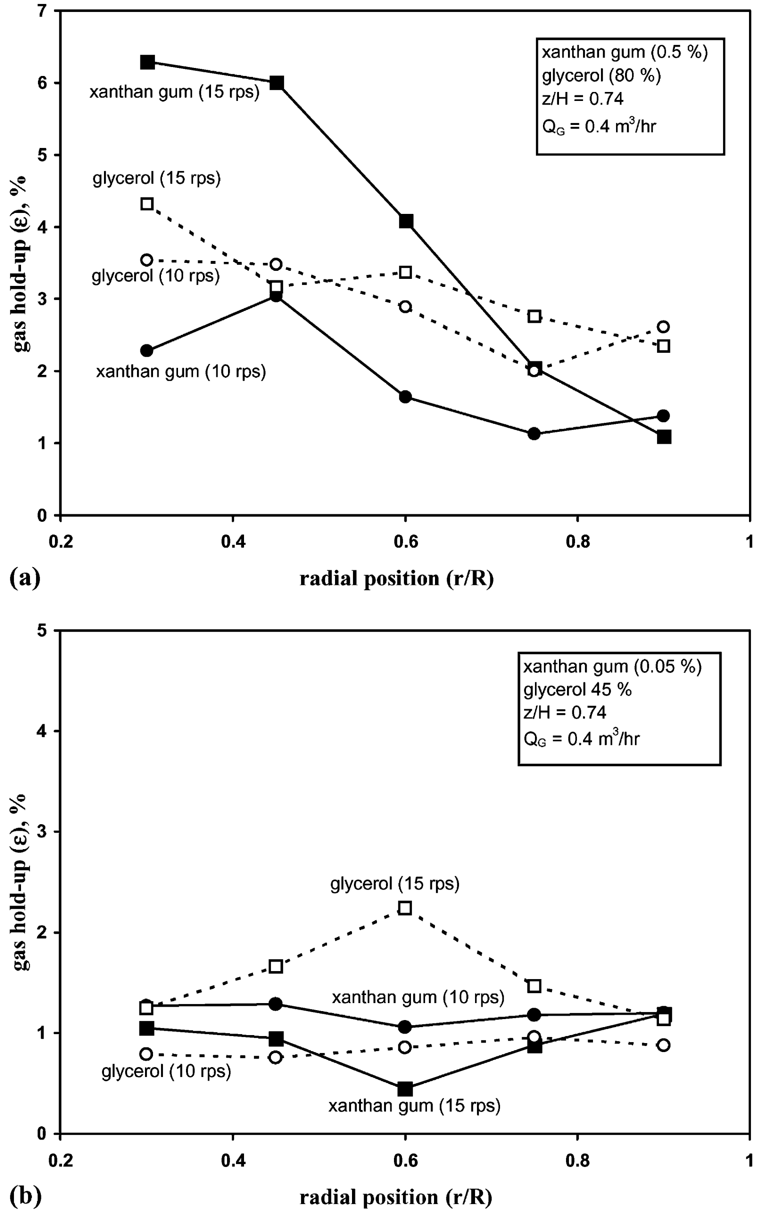

Fig. 7. The effect of stirrer speed on gas distribution in: (a) concentrated solutions of xanthan gum and glycerol; (b) dilute solutions of xanthan gum and glycerol.

corresponds to $K \sim 0.2$–$0.4\,\mathrm{Pa\,s}^n$ and $n \sim 0.55$–0.45. On the other hand, the gas hold-up distribution obtained could be used for sizing of turbulent pseudo-caverns. For example, following the case described in Fig. 6(c) (i.e. xanthan gum $5\,\mathrm{kg/m^3}$ at 15 rps), one can estimate the pseudo-cavern in this case to extend up to $z/H = 0.64$ axially and up to $r/R \sim 0.6$ radially, as these are the limits of high ε.

The lack of coalescence in our system (containing an electrolyte) did not allow comparison of our data with the data of Machon et al. [9]. Besides, plasticity in their case was very low. However, these results coincide with their results at low polymer concentrations, which could be related to the first part of the relationships illustrated in Fig. 3. As described in Fig. 2(c), the formation of bimodal size distribution was observed also in this study in transparent solutions at $\mu > 10\,\mathrm{mPa\,s}$ (i.e. following the ε maximum).

4. Conclusions

1. On the basis of point measurements of gas hold-up in a conventional agitated vessel at Reynolds number in the transition region and rheological conditions corresponding to strong pseudoplasticity ($n < 0.5$) at high consistency ($K > 0.1\,\mathrm{Pa\,s}^n$), where centrifugal forces and vortex systems differ substantially from those occurring in low-viscosity liquids, the structure of gas–liquid flow

in terms of the spatial distribution of gas hold-up has been determined so that the effects of rheology upon the gas–liquid mixing structure can be examined.

2. The results show significant non-uniformity of gas hold-up in both radial and axial directions. In terms of the typical division of the vessel into impeller and bulk zones, the distribution of gas hold-up is characterised by the relationships as follows:
    - At low polymer concentrations (i.e. low K and high n) $K \sim 0.005\,\mathrm{Pa\,s}^n$ and $n \sim 0.7$–0.8, the gas distribution in pseudoplastic media in the impeller region as well as in the bulk is similar to that obtained for Newtonian flow.
    - At high polymer concentrations (i.e. $K > 0.2$–$0.3\,\mathrm{Pa\,s}^n$ and $n < 0.5$), the primary circulation is localised closer to the shaft. The locus of maximum gas hold-up shifts to the shaft, while the region near the wall becomes a trap for small bubbles with increasing concentration at high resistance to bubble rise. Near the impeller, this dispersion regime is characterised by an increase of primary gas circulation registered by a significant gas hold-up rise.
3. The gas dispersion patterns observed are explained by the changes in shear and flow distribution in the pseudoplastic solutions due to drag reduction and hindered gas dispersion, as well as due to occurrence of a range of viscosity beyond which buoyancy rise becomes insignificant and gas hold-up is merely the result of flow and circulation generated by the impeller rotation.
4. The parameter boundaries of previously postulated restructuring from uniformity to well-delineated non-uniformity in highly concentrated and highly viscous non-Newtonian polymer dispersions are quantified for pseudoplastic flow as follows: at gas flow rates between 0.5 and 1 vvm and impeller tip velocities greater than 0.5 m/s, pseudoplasticity induced formation of zones with different mixing intensity starts at $K \sim 0.2$–$0.3\,\mathrm{Pa\,s}^n$ and $n \sim 0.5$.

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