Chemical and Process Engineering: New Frontiers
DOI: 10.24425/cpe.2024.149473

Gas hold-up for gas-liquid and biophase-gas-liquid systems agitated in a vessel equipped with vertical tubular baffles

Marta Major-Godlewska*, Magdalena Cudak

West Pomeranian University of Technology in Szczecin, Faculty of Chemical Technology and Engineering, Piastów Ave. 42, 71-065 Szczecin, Poland;
Marta Major-Godlewska ORCID: https://orcid.org/0000-0001-8407-029X
Magdalena Cudak ORCID: https://orcid.org/0000-0002-3727-1651

*Corresponding author: Marta Major-Godlewska, e-mail mmajor@zut.edu.pl

Abstract. The influence of the agitator type, agitator speed, superficial gas velocity, type of sugar (glucose or sucrose) and the presence of yeast in the system on the gas hold-up in an agitated vessel with 24 vertical tubular baffles (located on the circuit in the vessel) has been presented in this paper. The measurement of gas hold-up was conducted in an agitated vessel with inner diameter of D = 0.288 m and liquid height of H = 0.288 m. Three different agitators were used in the experimental study. Five gas-liquid and two biophase-gas-liquid systems were agitated in an agitated vessel. Air was used as gas. The influence of gas flow number, Weber number, the mass fraction of aqueous sugar solution c_i, and mass fraction of yeast suspension y_s for gas-liquid and biophase-gas-liquid systems on the gas hold-up φ was described mathematically. These equations do not have equivalents in the literature.

Keywords: mixing, gas-liquid or biophase-gas-liquid, gas hold-up, vertical tubular baffles, Newtonian or non-Newtonian liquid

This article has been accepted for publication and undergone full peer review but has not been through the copyediting, typesetting, pagination and proofreading process which may lead to differences between this version and the version of record. Please cite this article as DOI: 10.24425/cpe.2024.149473.

Received: 08 August 2024 | Revised: 18 October 2024 | Accepted: 14 November 2024

1. INTRODUCTION

Single-, two- and three-phase processes are carried out in an agitated vessel of various scales. (Stręk, 1981; Kamieński, 2004; Major-Godlewska et al., 2012; Petricek et al., 2018; Cudak, 2020; Cudak and Rakoczy, 2022). Adding another phase makes it much more difficult to ensure proper hydrodynamics of such systems. An additional difficulty occurs when one of the phases is the biological phase. This difficulty results from: the very often complicated non-linear kinetics of such a process, the possibility of occurrence of transient states, the variability of the structure of biological material, the required high accuracy of process parameter regulation (temperature, pH, pO_2, concentration of substrates or products) and sterility, as well as: the non-Newtonian nature of the media and its variability over time process (Bednarski and Fiedurka, 2007). In order to properly conduct tests in such systems, it is necessary to meet several conditions:

- selecting appropriate microorganisms,
- selecting appropriate culture medium (medium)
- selection of appropriate geometric, operational and physical parameters
- selecting appropriate process conditions (temperature, pH).

Cultures of microorganisms can be used for: multiplying cells to produce food (proteins from single cells), vaccines; production of primary metabolites, e.g. acids, alcohols, enzymes, polysaccharides; production of secondary metabolites, e.g. antibiotics; bioremediation, e.g. wastewater treatment, metal bioleaching; biotransformation of organic compounds; various processes in molecular biology, e.g. production of recombinant proteins, gene cloning (Nair, 2008*).* The growth rate of microorganisms depends on: the type and strain of microorganisms; composition of the medium (type and amount of nutrients, amount of harmful metabolites); physical and chemical growth conditions (temperature, pH, water activity, redox potential) (Bednarski and Fiedurka, 2007).

Media used for industrial breeding should, whenever possible, meet the following conditions: maximize the yield of the product or biomass relative to the substrate, maximize the concentration of the product or biomass, maximize the speed of product production, minimize the production of undesirable products, be cheap and available all year round, minimize technological difficulties - aeration , mixing, product purification, sewage and waste generation (Szewczyk, 2003). Therefore, the media used for the cultivation of microorganisms must contain many ingredients necessary for the proper course of metabolic processes. The main components of the media are: water, substances constituting a source of carbon, nitrogen, oxygen, minerals, especially phosphorus compounds, various substances that are growth

promoters or precursors of desired metabolic products (Szewczyk, 2003). The carbon source may be: glucose, lactose, starch, sucrose, sugar beet or sugar cane molasses, whey, glycerol, ethanol, corn and maltose syrup, citric acid and others. The sources of nitrogen used are primarily yeast extract, fish and plant meal (rapeseed, soybean, corn germ, etc.), bovine blood and others.

Some of the devices used in bioprocesses are bioreactors with (one or more) agitators. The use and selection of an appropriate agitator or agitator configuration is necessary to evenly distribute nutrients and microbial populations throughout the entire volume of the vessel (Gogate et al., 2000; Newell and Grano, 2007; Zhu et al., 2009; Bustamante et al., 2013; Gelves et al., 2014; Xie et al., 2014; Cudak, 2016; de Jesus et al., 2017). In the case of aerobic processes, it is necessary to facilitate gas exchange supplying the appropriate amount of oxygen to the system, i.e. appropriate aeration (Devi et al.,2014). One of the parameters on the basis of which we can determine the state of the system is the share of gas retained in the liquid. Obtaining optimal values of the share of gas retained in the liquid requires appropriate selection of geometric parameters (both the agitated vessel and the agitator) and operational conditions prevailing in the system (Brusciglio et al., 2013; Wan et al., 2016; Amiraftabi et al., 2020).

In the case of geometric parameters, the greatest influence on the gas hold-up is the type of agitator in the case of vessels with slender (H/D > 1) configuration of the agitators (Moucha et al., 2003; Bao et al., 2015; Busciglio et al., 2017). Comparing different systems of single agitators or agitator configurations, it can be concluded that the highest values of the gas hold-up are obtained for systems with a single Rushton turbine agitator or a configuration of several such agitators (Newell and Grano, 2007; Mueller and Dudukovic, 2010). In the case of gas-liquid or gas-solid-liquid systems, this is beneficial. However, when there is a biophase in the system, it may lead to its destruction due to the generation of high shear stresses by the agitator (Campesi et al., 2009; Collignon et al., 2010; Yang et al., 2014). In this case, it is necessary to replace the Rushton turbine agitator (or agitators) with agitators with lower shear stresses, e.g. Smith turbine agitator (CD6) or A315. However, obtaining comparable values of the gas hold-up requires supplying more energy to the system. In addition to the agitators, the gas hold-up is influenced by the baffles: their number, type and location in the bioreactor. The most important operating parameters include the agitator speed and the volumetric gas flow rate through the system (superficial gas velocity).

It can be clearly stated that the gas hold-up increases both with the increase in the agitator speed and with the increase in the volumetric gas flow rate through the agitated vessel. However, how big this increase is depends on the other parameters of a given system (Chinnasamy et al., 2015;

Major-Godlewska and Radecki, 2018; Jamshed et al., 2018; Barros et al., 2022). Additionally, the gas hold-up is influenced by the physical properties of the individual phases included in the system: density, viscosity, surface tension, concentration (Saravanan et al., 2009; Major-Godlewska et al., 2003, 2011; Cudak, 2014; Khalili et al., 2018; Jamshidzadeh et al., 2020; Liu et al., 2020; Major-Godlewska and Cudak, 2022). The selection of the most advantageous system of vessel-agitator-baffle-type of individual phases requires many studies on the influence of various parameters on the gas hold-up (Major-Godlewska and Cudak, 2022).

The research results presented in this paper aim to determine the influence of the agitator type, agitator speed, superficial gas velocity, type of sugar (glucose or sucrose) and the presence of yeast in the system on the gas hold-up in an agitated vessel with 24 vertical tubular baffles.

2. MATERIALS AND METHODS

The gas hold-up φ was measured in an agitated vessel with vertical tubular baffles. The vertical tubular baffles can be an alternative to flat baffles. They constitute baffles in the agitated vessel, and after supplying a hot or cold medium to them, they can work as vertical tubular coils (Karcz et al., 2001). Their purpose in such a case may be to maintain a constant temperature in a process; for example, in multiplication of microorganisms.

The diameter of the agitated vessel was D = 0.288 m. The agitated vessel was filled by liquid or bioliquid up to the $H = D$. The vertical tubular baffles consisted of J = 24 vertical tubes. Tubes were arranged symmetrically in the circuit of diameter $D_B = 0.7D$ inside the agitated vessel. The outer diameter of a single tube was $B = 0.02D$. Three different agitators: Rushton turbine (RT) agitator, Smith turbine (CD6) agitator and A315 agitator were used in the experimental study. The diameter of the agitators used for the tests was the same $d = 0.33D$ but the number of blades was Z = 6 for the Rushton turbine (RT) and Smith turbine (CD6) and Z = 4 for A315 agitator. When choosing the Rushton turbine (RT) agitator, the reason was that it is a standard agitator used in many processes, characterized by good mixing intensity. Unfortunately, this agitator is also characterized by the generation of high shear stresses, which is not favorable in the case of biofluids. Therefore, it seems appropriate to replace the Rushton turbine (RT) agitator with a Smith turbine (CD6) agitator or an A315 agitator, which are characterized by a modified blade shape. These agitators are characterized by much lower shear stress for a given blade shape (curves or large surface area). The gas sparger was formed in the shape of the ring with diameter $d_g = 0.7d$. The gas sparger of-bottom clearance was $e = 0.5h$, where h was the distance of the agitator from the bottom $h = 0.17H$.

The experimental study was carried out for gas-liquid and biophase-gas-liquid systems. The liquid phase for gas-liquid system was: distilled water, aqueous solution of glucose with a mass fraction of c_g = 0.03 kg_A/kg and c_g = 0.06 kg_A/kg, aqueous solution of sucrose with a mass fraction of c_s = 0.03 kg_A/kg and c_s = 0.06 kg_A/kg. Similarly, for the biophase-liquid system: aqueous solution of glucose of c_g = 0.06 kg_A/kg or sucrose of c_g = 0.06 kg_A/kg as liquid and the fresh pressed baker`s yeast *Saccharomyces cerevisiae* produced by Lesaffre Polska S.A. with mass fraction y_s = 0.02 kg_A/kg as biophase. The gas phase was air. The volumetric gas flow rate was V_G = <$2.78 \cdot 10^{-4}$; $5.56 \cdot 10^{-4}$> m^3/s and its corresponding superficial gas velocity w_{og} = $4V_G/\pi D^2$ = <$4.27 \cdot 10^{-3}$; $8.53 \cdot 10^{-3}$> m/s. The gas hold-up measurements were conducted for the range of good dispersion of gas bubbles in liquid. The smallest agitator speeds for all measurement series are shown in Fig. 1 in the form of a dependence of n = f(type of agitator). On the other hand, the highest agitator speeds were these values at which surface aeration of the liquid in the vessel did not occur yet.

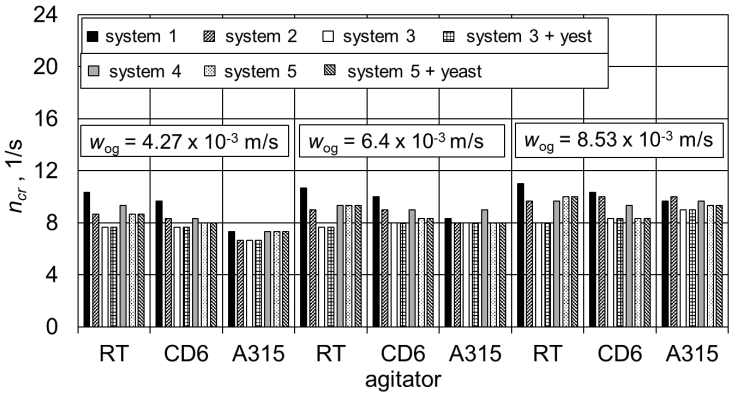


Fig. 1. Dependence of n_{cr} = f(type of agitator)

The dynamic viscosity coefficient η for Newtonian liquids was determined using a Höppler viscometer. Values of the dynamic viscosity coefficient η, density ρ and surface tension σ for fluids are shown in Table 1.

Table 1. The values of dynamic viscosity coefficient η, density ρ and surface tension σ for Newtonian liquids

systems	liquid	η, Pa·s	ρ, kg/m^3	σ, N/m
1	water	0.001	998	0.072
2	aqueous solution of glucose c_g = 0.03 kg_A/kg	0.00102	1010	0.073
3	aqueous solution of glucose c_g = 0.06 kg_A/kg	0.00105	1019.5	0.0792
4	aqueous solution of sucrose c_s = 0.03 kg_A/kg	0.00102	1011	0.0697
5	aqueous solution of sucrose c_s = 0.06 kg_A/kg	0.00107	1022.5	0.0744

For a given system, each time its rheological parameters (m and K) were measured, which enabled the calculation of the dynamic viscosity coefficient from the equation:

$$\eta_{b-l} = K \cdot \gamma^{m-1} \quad (1)$$

The rheological parameters of the biofluid varied in the range of m and K, density ρ and surface tension σ for biofluids shown in Table 2.

Table 2. Ranges of rheological parameters, density ρ and surface tension σ of the biofluid (non-Newtonian liquids)

systems	Biophase-liquid system	Ranges m	Ranges K, Pa·s^m	ρ, kg/m^3	σ, N/m
3 + yeast	Aqueous solution of glucose c_g = 0.06 kg$_A$/kg and traditional yeast y_s = 0.02 kg$_A$/kg	0.755-0.837	0.00725-0.00379	1021	0.0802
5 + yeast	Aqueous solution of sucrose c_s = 0.06 kg$_A$/kg and traditional yeast y_s = 0.02 kg$_A$/kg	0.821-0.847	0.00469-0.00418	1024	0.0812

Rheological parameters were determined using a rheoviscometer of RT 10 manufactured by Haake. The measurements were carried out using a system of two coaxial cylinders (DG 41).

The gas hold-up φ was calculated from Equation (2) using the values h_g – determined as height of a gas liquid (biophase-gas-liquid) mixture in the agitated vessel (which was calculated as average with 10 values) and H – liquid height in the agitated vessel.

$$\varphi = \frac{h_g}{h_g + H} \quad (2)$$

3. RESULTS AND DISCUSSION

On the basis of the experimental study, the set of gas hold-up φ values was obtained for seven different gas-liquid or biophase-gas-liquid systems. An analysis was carried out on the basis of 10560 measurement points. The agitator speed ranged from n_{cr} (Fig. 1), for a given type of agitator, to $n \leq 14$ 1/s. The results corresponded to turbulent regime for gas-liquid systems Re $\in$ <58450; 126200> and for biophase-gas-liquid systems Re $\in$ <28830; 73800>.

Based on the conducted research, it was found that the gas hold-up φ increased with the increase of the agitator speed n and the superficial gas velocity w_{og}. Analyzing the data presented in Fig. 2, for an agitated vessel with Rushton turbine (RT) agitator, taking into account the type of two-phase gas-liquid system, it was observed that at constant values of the agitator speed n = 11 1/s,

the lowest values of φ were obtained for the air-system 1 amounting to: φ = 4.57 % for w_{og} = $4.27 \cdot 10^{-3}$ m/s, φ = 6.16 % for w_{og} = $6.4 \cdot 10^{-3}$ m/s and φ = 6.49 % for w_{og} = $8.53 \cdot 10^{-3}$ m/s. Comparable values of φ, by about 7%, were observed when the gas was dispersed in the systems 3, 5 and 3 + yeast. By comparing the values φ obtained for the air-system 5 with the values φ obtained for the air-system 5 + yeast at a constant value of n = 11 1/s and w_{og} = $4.27 \cdot 10^{-3}$ m/s, it was found that the addition of yeast into the system resulted in approximately 11% lower values of gas hold-up.

Increasing the conventional value of the superficial gas velocity w_{og} = $4.27 \cdot 10^{-3}$ m/s for n = const increases the value of φ by approximately 10% on average for w_{og} = $6.4 \cdot 10^{-3}$ m/s and by approximately 20% for w_{og} = $8.53 \cdot 10^{-3}$ m/s for systems 3, 3 + yeast, 5 and 5 + yeast. A greater influence of w_{og} was found for the air-system 1 and it was approximately 35% and 40%. The addition of biophase to the gas-liquid system had little significance when yeast was added to an aqueous solution of glucose (system 3). However, the influence of the system was visible when the yeast was added to an aqueous solution of sucrose (system 5). In this case, the difference between the gas-liquid system, where the liquid is the system 5, and the biophase-gas-liquid system, where the biophase-liquid is the system 5 + yeast is approximately 14% (n = 11 1/s, w_{og} = $6.4 \cdot 10^{-3}$ m/s).

Adding glucose or sucrose to the model system (system 1) causes an increase in the φ value in all analyzed cases (Fig. 2). This influence decreases with an increase in the agitator speed n and with an increase in w_{og}. Assuming n = const, adding sugar to the system increased the value of φ for an aqueous glucose solution (system 3) by approximately 53 %, 29 %, 28 % respectively for w_{og} = $4.27 \cdot 10^{-3}$ m/s, w_{og} = $6.4 \cdot 10^{-3}$ m/s w_{og} = $8.53 \cdot 10^{-3}$ m/s (n = 11 1/s), by about 27 %, 22 %, 17 % respectively for w_{og} = $4.27 \cdot 10^{-3}$ m/s, w_{og} = $6.4 \cdot 10^{-3}$ m/s w_{og} = $8.53 \cdot 10^{-3}$ m/s (n = 13 1/s), and for an aqueous sucrose solution (system 5) by about 28 %, 25 %, 22% respectively for w_{og} = $4.27 \cdot 10^{-3}$ m/s, w_{og} = $6.4 \cdot 10^{-3}$ m/s w_{og} = $8.53 \cdot 10^{-3}$ m/s (n = 12 1/s), by about 26 %, 23 %, 17 % respectively for w_{og} = $4.27 \cdot 10^{-3}$ m/s, w_{og} = $6.4 \cdot 10^{-3}$ m/s w_{og} = $8.53 \cdot 10^{-3}$ m/s (n = 13 1/s). However, assuming w_{og} = $8.53 \cdot 10^{-3}$ m/s, adding glucose to the system (air-system 1) increased the value of φ by approximately 28%, 22%, 17%, and in the case of adding sucrose by approximately 26%, 22%, 17% for n = 11 1/s, 12 1/s, 13 1/s, respectively.

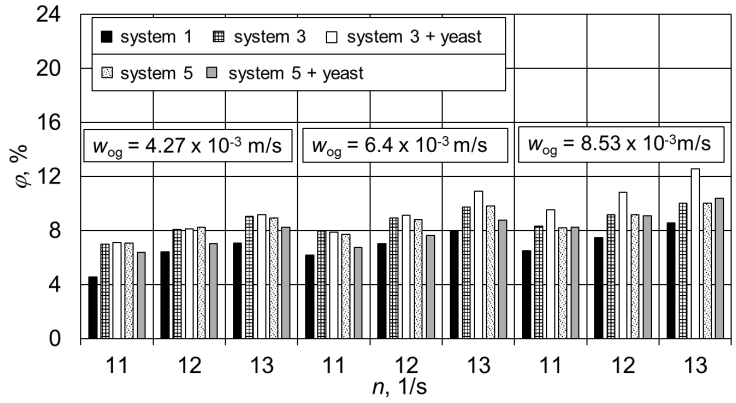


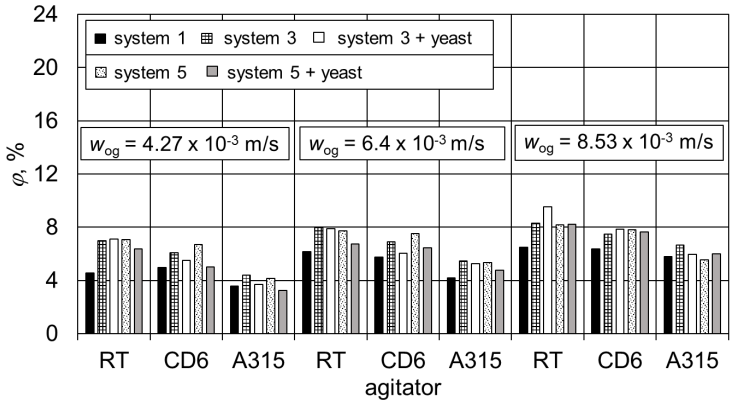
Fig. 2. Dependence $\varphi = f(n)$ for the Rushton turbine (RT) agitator

Analyzing the data presented in Fig. 2, it can be concluded that adding yeast (system 3 + yeast) to an aqueous glucose solution (system 3) caused an increase in the value of φ by approximately 15% (n = 11 1/s, $w_{og} = 8.53 \cdot 10^{-3}$ m/s), 18% (n = 12 1/s, $w_{og} = 8.53 \cdot 10^{-3}$ m/s) and 25% (n = 13 1/s, $w_{og} = 8.53 \cdot 10^{-3}$ m/s), and also for n = 13 1/s, $w_{og} = 6.4 \cdot 10^{-3}$ m/s by about 12%. The biophase-gas-liquid systems (air-system 5 + yeast) and the gas-liquid system (air-system 5) behaved differently. In this case, higher values of φ were obtained only at lower gas flow rates ($w_{og} = 4.27 \cdot 10^{-3}$ m/s and $w_{og} = 6.4 \cdot 10^{-3}$ m/s). Adding yeast to such a system reduced the hold-up φ by approximately 10%, 15%, 8% for n = 11 1/s, 12 1/s, 13 1/s, respectively ($w_{og} = 4.27 \cdot 10^{-3}$ m/s) and by about 13%, 13%, 11% for n = 11 1/s, 12 1/s, 13 1/s, respectively ($w_{og} = 6.4 \cdot 10^{-3}$ m/s).

Fig. 3 shows the influence of the type of agitator used for testing the gas hold-up φ. The influence of the type of agitator on the gas hold-up φ depends on the agitator speed n of the agitator, the superficial gas velocity w_{og}, the type of liquid and the presence or absence of yeast suspension in the system. The greatest influence of the type of agitator on the gas hold-up φ was found, in most cases, for the lowest agitator speed n = 11 1/s and the lowest values of $w_{og} = 4.27 \cdot 10^{-3}$ m/s. In this case, replacing the Rushton turbine agitator with a Smith turbine agitator resulted in a decrease in the value of the hold-up φ by approximately 15-30%, depending on the type of liquid and the presence or absence of yeast suspension. Even greater differences in the obtained values of the gas hold-up φ can be seen when replacing the Rushton or Smith turbine agitators with the A315 agitator. In this case, for lower agitator speed n = 11 1/s and lower value superficial gas velocity $w_{og} = 4.27 \cdot 10^{-3}$ m/s, the decrease in the gas hold-up φ is approximately 30-90% - replacing the Rushton turbine agitator (RT) with the agitator A315 and approximately 40-60% - replacing the Smith turbine agitator (CD6) with the A315 agitator. The influence of the type of agitator on the gas hold-up decreased both with increasing n and w_{og}. In this case, this effect also depends on the type of liquid and the presence of yeast in the system.

Comparing the values of φ at a constant value of $w_{og} = 8.53 \cdot 10^{-3}$ m/s and n = 13 1/s, it was observed that for the tested systems there were no large differences in the values of φ when agitators (RT or CD6) were used for mixing. Only when the agitated system was the air-system 3 + yeast, the type of agitator used (RT and CD6) was important. Higher φ values by approximately 24% were obtained when the system: air- system 3 + yeast was agitated with a turbine agitator with straight blades (RT). However, comparing the values φ obtained for the Smith turbine agitator (CD6) with the values φ obtained for the A315 agitator, characterized by a larger blade surface, it was found that at the same $w_{og} = 8.53 \cdot 10^{-3}$ m/s and n = 13 1/ s higher values by approximately 15% - 40% depending on the type of agitated system were obtained for the Smith turbine agitator. In such a case, when selecting the type of agitator, the type of system to be agitated should be taken into account.

a)



b)

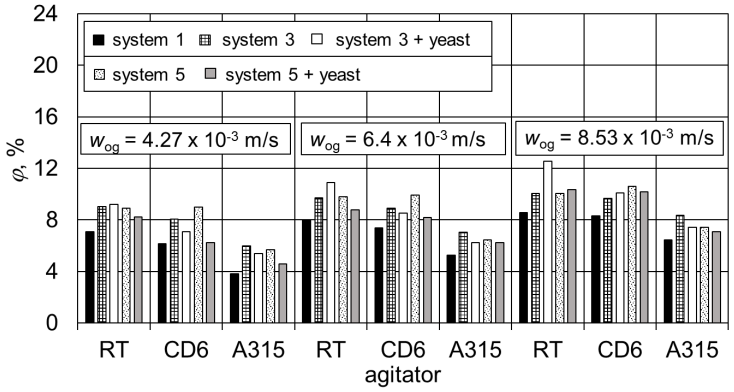


Fig. 3. Dependence φ = f(types of agitator); n = 11 1/s (a) and n = 13 1/s (b)

The influence of the type of agitator on φ depends on the type of liquid phase in the system and the presence or absence of yeast. In the case of changing the Rushton (RT) or Smith (CD6)

turbine agitator, a greater influence of the type of agitator on the value of φ was found for the system with an aqueous glucose solution (system 3). However, when comparing the results obtained for an agitator with a Rushton (RT) or Smith (CD6) turbine agitator with the results obtained for an agitator with an A315 agitator it was found that a greater influence of the type of agitator on the gas hold-up φ was obtained for the system with an aqueous sucrose solution (system 5). In most analyzed cases, a greater influence of the type of agitator, regardless of the liquid phase used, on the gas hold-up was found when a biological phase was added to the system.

The influence of fluid properties on gas retention in the liquid is shown in Fig. 4. It was observed that at the same agitator speed n = const. (Fig. 4a) the gas hold-up φ increased with the change of the system and the value of superficial gas velocity w_{og}. For example, for the biophase-gas-liquid system 3 + yeast for n = const = 12 1/s, an approximately 33% higher value φ was obtained for superficial gas velocity $w_{og} = 8.53 \cdot 10^{-3}$ m/s compared to the value φ obtained for $w_{og} = 4.27 \cdot 10^{-3}$ m/s. Analyzing the values of φ for different systems with the same constant value n = 12 1/s and for the value of superficial gas velocity $w_{og} = 8.53 \cdot 10^{-3}$ m/s, it was observed that the value of φ was influenced by the type of liquid used in the systems. The highest value of φ = 10.8% was obtained for the biophase-gas-liquid system 3 + yeast, and the lowest φ = 7.5% for the system in which the liquid was distilled water (system 1) (Fig. 4a).

a) b)

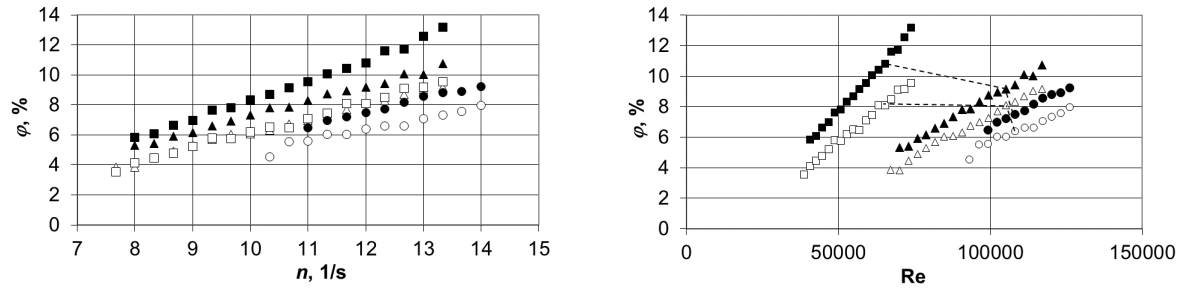


Fig. 4. Dependence of $\varphi = f(n)$ (a) or $\varphi = f(\mathrm{Re})$ (b) for the Rushton turbine agitator; the gas-liquid system: ○,● - system 1, Δ, ▲ – system 3; the biophase-gas-liquid system: □, ■ – system 3 + yeast; two different superficial gas velocity w_{og}: ○, Δ, □ - $4.27 \cdot 10^{-3}$ m/s; ●,▲, ■ - $8.53 \cdot 10^{-3}$ m/s.

Analyzing the data presented in Fig. 4b, it was found that the range of the Reynolds number Re varied depending on the properties of the fluid used in the tests. Similar values of the Reynolds number Re are observed for the gas-liquid system in which system 1 or system 3 was used as

the liquid. Much lower values of the Reynolds number (Re = 65291), at the same value of the agitator speed n = 12 1/s (lines for the eye in Fig. 4b), are observed when yeast (system 3 + yeast) was added to the system 3 compared to the values Re = 105219 obtained for the system with aqueous glucose solution (system 3). The lack of a biological phase in the system in these studies means that similar hold-up φ values can be obtained for higher Reynolds numbers. For w_{og} = 4.27·10^{-3} m/s, the value of φ equal to approximately 8% for the gas – system 1 was obtained for a Reynolds value approximately twice as high Re = 126203 compared to the biophase-liquid systems (system 3 + yeast), where Re = 65291.

The influence of gas flow number Kg, Weber number We, the mass fraction of aqueous sugar solution c_i, and mass fraction of yeast suspension y_s on the gas hold-up φ, for a two- and three-phase systems, was presented in the form of an equation:

$$\varphi = a_1 \cdot \mathrm{Kg}^{a_2} \cdot \mathrm{We}^{a_3} \cdot (1 + c_i)^{a_4} \cdot (1 + a_5 \cdot y_s) \quad (3)$$

The values of the coefficients (a_1, a_5) and exponents (a_2, a_3, a_4), the average relative error of Equation (3) are given in Table 3.

Table 3. Values of coefficients a_1, a_5 and exponents a_2, a_3 and a_4 in Eq. (3), the average relative error and ranges of gas flow number Kg, Weber number We

No	Agitator	a_1	a_2	a_3	a_4	a_5	$\pm\Delta$	c_i, kg$_A$/kg	Ranges of Kg; We
1.	RT	2.38 x 10^{-4}	0.28	0.88	4.22	3.19	6	water and sucrose $c_s \in$ <0; 0.06>	Kg ∈ <0.023; 0.072>; We ∈ <480; 2445>.
2.	CD6	9.50 x 10^{-5}	0.37	1.03	5.11	-0.75	8		
3.	A315	3.01 x 10^{-4}	0.51	0.90	2.74	-3.26	7		
4.	RT	2.32 x 10^{-4}	0.31	0.90	5.15	5.43	4	water and glucose $c_g \in$ <0; 0.06>	
5.	CD6	1.15 x 10^{-4}	0.37	1.01	4.53	-0.10	5		
6.	A315	1.88 x 10^{-4}	0.53	0.97	5.55	-3.58	7		

The differences between the values of the a_5 coefficient in Equation (3) may result from the influence of the liquid circulation generated by the individual agitators in the agitated vessel with vertical tubular baffles on the growth of microorganisms. In the case of the agitated vessel with the Rushton turbine agitator (RT), this circulation was the closest to radial circulation. On the other hand, in the agitated vessel with the Smith turbine (CD6) or A315 agitator, radial-axial circulation occurred. However, the influence of these components changed. A greater influence of the axial component over the radial component was observed for the agitated vessel with the A315 agitator than for the agitated vessel with Smith turbine agitator (CD6). For this

reason, the circulation generated in the agitated vessel also influenced the value of the share of gas retained in the liquid in these vessels. The highest values of the share of gas retained in the liquid were found for the agitated vessel with the Rushton turbine agitator (RT), lower - for the vessel with the Smith turbine agitator (CD6) and the lowest - for the A315 agitator. It can be assumed that the higher the φ values, the more favorable the conditions in the agitated vessel for the growth of microorganisms.

4. CONCLUSIONS

Based on the data obtained, it was found that the type of (gas-liquid, biophase- gas-liquid) system used in the tests influenced the gas hold-up φ. Differences in the obtained φ values are also visible depending on the type of agitator used (RT, CD6, A315). The value of the conventional linear gas velocity w_g and the agitator speed n are also important.

It can be clearly stated that the share of gas hold-up increased with an increase in the agitator speed and the superficial gas velocity. However, this increase depended on other variables (type of agitator, type of liquid, presence of biophase).

It is not possible to propose one best solution: system - agitator - parameters w_{og}, n. When selecting such a solution, one should take into account what type of system is mixed (gas-liquid, biophase-gas-liquid), what parameters φ are required (or must be the highest) and the consumption of power depending on the type of agitators.

The obtained results are presented in this article and described mathematically using Equation (3), where the values of constants: a_1, a_5 and exponents: a_2, a_3, a_4, presented in Table 3, can be used to design and model multiphase systems characterized by identical physicochemical properties.

SYMBOLS

B - width of the baffle, m

c_i - sugar (sucrose or glucose) mass fraction, kg_A/kg

D - inner diameter of the agitated vessel, m

d - diameter of the agitator, m

d_g - sparger diameter, m

e - off-bottom clearance of gas sparger, m

H - liquid height in the agitated vessel, m

h - distance between the agitator and the bottom, m

h_g - the height of a gas-liquid (gas-biophase-liquid) mixture in the agitated vessel, m

J - number of baffles

K - consistency index, $Pa \cdot s^m$

m - flow index

n - agitator speed, 1/s

n_{cr} - critical agitator speed, 1/s

w_{og} - superficial gas velocity, m/s

V_G - volumetric gas flow rate, m^3/s

y_s - yeast mass fraction, kg_A/kg

Z - number of agitator blades

Greek symbols

η - dynamic viscosity, $Pa \cdot s$

φ - gas hold-up

ρ - density, kg/m^3

σ - surface tension, N/m

Dimensionless numbers

$Kg = \frac{V_G}{nd^3}$ gas flow number

$We = \frac{n^2 d^3 \rho}{\sigma}$ Weber number

$Re = \frac{nd^2 \rho}{\eta}$ Reynolds number

REFERENCES

1. Amiraftabi M., Khiadani M., Mohammed H.A., 2020. Performance of a dual helical ribbon impeller in a two-phase (gas-liquid) stirred tank reactor. C*hem. Eng. Process.: Process* Intensif., 148, 107811. DOI: 10.1016/j.cep.2020.107811.

2. Bao Y., Yang J., Wang B., Gao Z., 2015. Influence of impeller diameter on local gas dispersion properties in a sparged multi-impeller stirred tank. *Chin. J. Chem. Eng.*, 23, 615-622. DOI: 10.1016/j.cjche.2014.12.006.
3. Barros P.A., Ein-Mozaffari F., Lohi A., 2022. Gas Dispersion in Non-Newtonian Fluid with Mechanically Agitated Systems: A review. *Process.*, 10, 275-304. DOI: 10.3390/pr10020275.
4. Bednarski W. and Fiedurka J., 2007. *Basics of industrial biotechnology* (in Polish). WNT, Warszawa
5. Busciglio A., Grisafi F., Scargiali F., Brucata A., 2013. On the measurement of local gas hold-up, interfacial area and bubble size distribution in gas-liquid contactors *via* light sheet and image analysis: Imaging technique and experimental results. *Chem. Eng. Sci*., 102, 551-566. DOI: 10.1016/j.ces.2013.08.029.
6. Bustamante M.C.C., Cerri M.O., Badino A.O., 2013. Comparison between average shear rates in conventional bioreactor with Rushton and Elephant ear impellers. *Chem. Eng. Sci*., 90, 92-100. DOI: 10.1016/j.ces.2012.12.028.
7. Campesi A., Cerri M.O., Hokka C.O., Badino A.C., 2009. Determination of the average shear rate in a stirred and aerated tank bioreactor. *Bioprocess Biosyst Eng*., 32, 241-248. DOI: 10.1007/s00449-008-0242-4.
8. Chinnasamy G., Kaliannan S., Eldho A., Nadarajan D., 2016. Development and performance analysis of a novel agitated vessel. *Korea. J. Chem. Eng.*, 33, 1181-1185. DOI: 10.1007/s11814-015-0264-y.
9. Collignon M.L., Delafosse A., Crine M., Toye D., 2010. Axial impeller selection for anchorage dependent animal cell culture in stirred bioreactors: methodology based on the impeller comparison at just-suspended speed of rotation. *Chem. Eng. Sci*., 65, 5929-5941. DOI: 10.1016/j.ces.2010.08.027.
10. Cudak M., 2014. Hydrodynamic characteristics of mechanically agitated air-aqueous sucrose solutions. *Chem. Process Eng*., 35, 97-107. DOI: 10.2478/cpe-2014-0007.
11. Cudak M., 2016. *Experimental and numerical analysis of transfer processes in a biophase-gas-liquid system in a bioreactor with an impeller* (in Polish). BEL Studio Sp. z o.o., Warszawa, Poland.
12. Cudak M., 2020. The effect of vessel sale on gas hold-up in gas-liquid systems. *Chem. Process Eng*., 41, 241-256. DOI: 10.24425/cpe.2020.136010.

13. Cudak M., Rakoczy R., 2022. Hydrodynamics of gas-liquid and biophase-gas-liquid systems in stirred tanks of different scales. *Korean J. Chem. Eng*., 39, 2959-2971. DOI: 10.1007/s11814-022-1281-2.
14. de Jesus S.S., Moreira Neto J., Filho R.M., 2017. Hydrodynamics and mass transfer in bubble column, conventional airlift, stirred airlift and stirred tank bioreactors, using viscous fluid: A comparative study. *Biochem. Eng. J*., 118, 70-81. DOI: 10.1016/j.bej.2016.11.019.
15. Devi T.T., Kumar B., 2014. Effects of superficial gas velocity on process dynamics in bioreactors. *Thermophys. Aeromech.*, 21, 365-382. DOI: 10.1134/S0869869431403010X.
16. Gelves R., Dietrich A., Takors R., 2014. Modeling of gas-liquid mass transfer in a stirred tank bioreactor agitated by Rushton turbine or a new pitched blade impeller. *Bioprocess Biosyst Eng.*, 37, 365-375. DOI: 10.1007/s00449-013-1001-8.
17. Gogate P.R., Beenackers A.A.C.M., Pandit A.B., 2000. Multiple-impeller systems with a special emphasis on bioreactors: a critical review. *Biochem. Eng. J*., 6, 109-144. DOI: 1016/S1369-703X(00)00081-4.
18. Jamshed A., Cooke M., Ren Z., Rodgers T.L., 2018. Gas–liquid mixing in dual agitated vessels in the heterogeneous regime. *Chem. Eng. Res. Des*., 133, 55-69. DOI: 10.1016/j.cherd.2018.02.034.
19. Jamshidzadeh M., Ein-Mozaffari F., Lohi A., 2020. Local and overall gas holdup in an aerated coaxial mixing system containing a non-Newtonian fluid. *AIChE J.*, 66, e17016. DOI: 10.1002/aic.17016.
20. Kamieński J., 2004. *Agitation of multiphase systems* (in Polish). WNT, Warszawa, Poland.
21. Karcz J., Major M., 2001. Experimental studies of heat transfer in an agitated vessel equipped with vertical tubular coil. *Inż. Chem. i Proc*., 22, 445-459.
22. Khalili F., Nasr M.R.J., Kazemzadeh A., Ein-Mozaffari F., 2018. Analysis of gas holdup and bubble behavior in a biopolymer solution inside a bioreactor using tomography and dynamic gas disengagement techniques. *J. Chem. Technol. Biotechnol.,* 93, 340-349. DOI: 10.1002/jctb.5356.
23. Liu B., Xiao Q., Gao P., Sunder B., Fan F., 2020. Investigation of gas–liquid dispersion and mass transfer performance of wide-viscosity-range impellers in water solutions of xanthan gum. *Chem. Eng. Res. Des*., 154, 60-69. DOI: 10.1016/j.cherd.2019.12.005.
24. Major-Godlewska M., Cudak M., 2022. Gas hold-up in vessel with dual impellers and different baffles. *Energies*, 15, 8685. DOI: 10.3390/en15228685.

25. Major-Godlewska M., Karcz J., 2003. Gas hold-up and power consumption for gas-liquid system agitated in a stirred tank equipped with vertical coil. *Chem. Pap*., 57, 432-437.
26. Major-Godlewska M., Karcz J., 2011. Process characteristics for a gas-liquid system agitated in a vessel equipped with a turbine impeller and tubular baffles. *Chem. Pap*., 65, 132-138. DOI: 10.2478/s11696-010-0080-0.
27. Major-Godlewska M., Karcz J., 2012. Agitation of a gas-solid-liquid system in a vessel with high-speed impeller and vertical tubular coil. *Chem. Pap*., 66, 566-573. DOI: 10.2478/s11696-012-0148-0.
28. Major-Godlewska M., Radecki D., 2018. Experimental analysis of gas hold-up for gas-liquid system agitated in a vessel equipped with two impellers and vertical tubular baffles. *Pol. J. Chem. Technol*., 20, 7-12. DOI: 10.2478/pjct-2018-0002.
29. Moucha T., Linek V., Prokopova E., 2003. Gas hold-up, mixing time and gas-liquid volumetric mass transfer coefficient of various multiple-impeller configurations: Rushton turbine, pitched blade and techmix impeller and their combinations. *Chem. Eng. Sci.*, 58, 1839-1846. DOI: 10.1016/S0009-2509(02)00682-6.
30. Mueller S.G., Dudukovic M.P., 2010. Gas hold-up in gas-liquid stirred tanks. *Ind. Eng. Chem. Res.*, 49, 10744-10750. DOI: 10.1021/ie100542a.
31. Nair J., 2008. Introduction to Biotechnology and Genetic Enineering. Infinity Science Press LLC, New Delhi, India.
32. Newell R., Grano S., 2007. Hydrodynamics and scale up in Rushton turbine flotation cells: Part 1 — Cell hydrodynamics. *Int. J. of Miner. Process*., 81, 224-236. DOI: 10.1016/j.minpro.2006.06.007.
33. Petricek R., Moucha T., Rejl F.J., Valenz L., Haidl J., Cmelikova T. 2018. Volumetric mass transfer coefficient, power input and gas hold-up in viscous liquid in mechanically agitated fermenters. Measurements and scale-up. *Int. J. Heat Mass Transf*., 124, 1117-1135. DOI: 10.1016/j.ijheatmasstransfer.2018.04.045.
34. Saravanan K., Ramamurthy V., Chandramohan K., 2009. Gas hold-up in multiple impeller agitated vessels. *Mod.* Appl. Sci., 3, 49-59. DOI: *10.5539/mas.v3n2p49*.
35. Stręk F., 1981. *Agitation and agitated vessels* (in Polish). WNT, Warszawa, Poland.
36. Szewczyk K.W., 2003. *Biochemical Technology* (in Polish). Oficyna Wydawnicza Politechniki Warszawskiej, Warsaw, Poland.
37. Wan X., Takahata Y., Takahashi K., 2016. Power consumption, gas holdup, and mass-transfer coefficient of triple-impeller configurations in a stirred vessels with vertical tubular coils. *Can. J. of Chem. Eng*., 94, 349-354. DOI: 10.1002/cjce.22385.

38. Xie M., Xia J., Zhou Z., Chu J., Zhuang Y., Zhang S., 2014. Flow pattern, mixing, gas hold-up and mass transfer coefficient of triple-impeller configurations in stirred tank bioreactors. *Ind. Eng. Chem. Res*., 53, 5941-5953. DOI: 10.1021/ie400831s.
39. Yang S., Li X., Deng G., Yang Ch., Mao Z., 2014. Application of KHX impeller in a low-shear stirred bioreactor. *Chin. J. Chem. Eng*., 22, 1072-1077. DOI: 10.1016/j.cjche.2014.09.001.
40. Zhu H., Nienow A.W., Bujalski W., Simmons M.J.H., 2009. Mixing studies in a model aerated bioreaktor equipped with an up- or a down-pumping 'Elephant Ear' agitator: Power, hold-up and aerated flow field measurements. *Chem. Eng. Res. Des*., 87, 307-317. DOI: 10.1016/j.cherd.2008.08.013.