PAPER • OPEN ACCESS

Experimental study of liquid velocity profiles in large-scale bubble columns with particle tracking velocimetry

To cite this article: G Besagni et al 2019 J. Phys.: Conf. Ser. 1224 012036

View the article online for updates and enhancements.



IOP ebooks[™]

Bringing you innovative digital publishing with leading voices to create your essential collection of books in STEM research.

Start exploring the collection - download the first chapter of every title for free.

Experimental study of liquid velocity profiles in large-scale bubble columns with particle tracking velocimetry

G Besagni¹, F Inzoli¹, T Zieghenein² and D Lucas²

¹ Politecnico di Milano, Department of Energy, Via Lambruschini 4a, 20156, Milano

² Helmholtz-Zentrum Dresden-Rossendorf e. V. Institute of Fluid Dynamics, 01314 Dresden, Germany

E-mail: giorgio.besagni@polimi.it

Abstract. A complete knowledge of the bubble column fluid dynamics relies on understanding the global and the local fluid dynamic properties. Unfortunately, most of the previous literature focused on the "global-scale" fluid dynamics, whereas a limited attention was devoted to the "local-scale". We contribute to present-day discussion by proposing an experimental study concerning the local liquid velocity profiles within the pseudo-homogeneous flow regime. The experimental study, based on a particle-identification and particle-tracking algorithm, was conducted in a large-diameter and large-scale bubble column (height equal to 5.3 m; inner diameter equal to 0.24 m) operated in the counter-current mode. We considered gas superficial velocities in the range of 0.37-1.88 cm/s and liquid superficial velocities up to -9 cm/s. Timeaveraged and transient liquid velocity fields were obtained for five superficial gas velocities and four superficial liquid velocities at two measuring heights. Subsequently, the liquid velocity observations were coupled with previously measured bubble size distributions and local void fractions, to provide a complete description of the "local-scale" fluid dynamics. These data would help in the validation procedure of numerical codes, to support the prediction of industrial-scale relevant conditions.

1. Introduction

In the broader framework of multiphase reactors, bubble columns are widely used as contacting devices in industrial applications, owing to their many advantage in both design and operation (i.e., low initial cost, low maintenance, high contact area between the phases, as discussed in ref. [1]). The simplest bubble column layout consists of a cylinder where the gas phase is dispersed—through a gas sparger-into a liquid phase, which is supplied in the batch mode or it may be led in either cocurrently or counter-currently to the upward gas phase [2]. On the practical point of view, batch-mode bubble columns are applied in hydrogenation and fermentation processes; conversely, waste-water treatment, water ozonation, and three-phase inverse fluidized bed involve a continuous flow of the liquid phase [3]. In these cases, counter-current bubble columns are employed thanks to the higher interfacial mass transfer compared with batch-mode bubble columns [4].

The gas phase, introduced through the gas sparger, evolves in the axial direction of the bubble column either in the form of "dispersed bubbles" or in the form of "coalescence-induced structures", depending on the prevailing flow regime (please refer to the discussion proposed by Besagni et al. [2]). The complete understanding of the flow regime characteristics, which is essential to correctly design and operate bubble columns, depends on the precise knowledge of the multi-scale fluid

Content from this work may be used under the terms of the Creative Commons Attribution 3.0 licence. Any further distribution of this work must maintain attribution to the author(s) and the title of the work, journal citation and DOI. Published under licence by IOP Publishing Ltd 1

dynamics, encompassing the "bubble-scale" (i.e., break-up/coalescence, swarm effects, ...) and the "reactor-scale" (i.e., large-scale circulation, global scale properties, ...). The fluid dynamics in bubble columns is subject to this multi-scale connection since the bubble motion drive the liquid phase, which in reverse influence the rising characteristic and break-up/coalescence of the bubbles. The most important "reactor-scale" parameter is the gas holdup, ε_G (defined as the volume of the gas phase divided by the total volume of the system). Conversely three parameters characterize the fluid dynamics at the "local-scale": (a) liquid velocity, (b) local gas fraction, and (c) bubble size distributions (BSDs). This concept is displayed in Figure 1, which represents a typical liquid velocity profile in batch-mode bubble columns, as outlined in the pioneering study of Ueyama and Miyauchi [5]: the liquid phase flows upward in the center of the column, it becomes zero at an intermediate radial position and, finally, it flows downward near the wall. As a consequence of such liquid velocity profile, bubbles move either towards the center of the column or towards the wall, owing to the wellknown effect of the lift force [7]. Subsequently, the distribution of the gas phase determines the local void fraction profiles: for example, the local void fraction profile is center-pecked if large bubbles move towards the center of the bubble columns, as observed by Besagni and Inzoli [7]; conversely, the local void fraction profile is flat is the BSD is prevailing composed by small-bubbles.



Figure 1. Radial liquid velocity profile in bubble columns (batch-mode).

In the last decades, a large amount of studies concerning bubble column fluid dynamics have been proposed; unfortunately, most of the literature investigated the batch/co-current modes and a limited attention was devoted to counter-current flows, as summarized by Trivedi et al. [6] (in particular, the reader may refer to Table 1 in ref. [6]). These studies mostly investigate the gas holdup curves [7-9], the flow regime transitions [7] and, a limited number provided BSDs and local flow properties measured by needle probes [7, 10]. Generally speaking, the gas holdup increases while increasing the counter-current liquid velocity, whereas the degree of this effect depends on the system design, phase properties and phase velocities [6, 7]. As stated above, three local fluid dynamic parameters (viz. liquid velocity, local gas fraction, and BSDs) together characterize the "local-scale"; however, none of the above-mentioned studies [7-10] measured the liquid velocity. Such experimental data are rare in the literature, especially if considering counter-current flows and large-scale experimental facilities; nevertheless this information is essential to completely understand the prevailing fluid dynamics in bubble columns. This concept has been stated by Trividi et al. [5], who contributed to the lack of experimental data by a lumped-parameter modeling approach. In summary, nowadays there is no experimental evidence regarding the influence of the effect of the counter-current mode on the profile proposed in Figure 1. This lack of knowledge is caused by the challenges in measuring the liquid velocity profiles, especially in dense bubbly flows in "large-scale" systems. In the broader group of

the *PIV* methods with [11] and without [12] laser illumination falls the Particle Tracking Velocimetry (*PTV*) method, which has been used in the present study. When no lasers are used for *PIV* measurements, a background light may be applied so that the shadows of particles are recorded, viz. Particle Shadowgraphy Velocimetry (*PSV*) [13]. *PSV* methods are not commonly used, but they have some advantages in bubbly flows, i.e., using a permanent light source so that no extra trigger is needed, which simplifies the experimental setup.

This study contributes to the existing discussion concerning the "local-scale" fluid dynamics in bubble columns, by applying a *PTV* method in a shadowgraphy setup (*PSTV*). The measurements have been conducted at the bubble column built at the Department of Energy of Politecnico di Milano. This bubble column has been developed concerning the scaling-up criteria behind the "large-diameter" and the "large-scale" concept (viz. the bubble columns has an inner diameter equal to 0.24 m, a height equal to 5.3 m, and a spider gas sparger with large openings [2, 14]). For this reason, this experimental setup can be considered as a pilot-scale experimental facility to scale-up results from "laboratoryscale", towards the "industrial-scale". We considered gas superficial velocities in the range of U_G = 0.37–1.88 cm/s and liquid superficial velocities up to $U_L = -9$ cm/s, which are within the boundaries of the pseudo-homogeneous flow regime. Time-averaged and transient liquid velocity fields were obtained for five superficial gas velocities and four superficial liquid velocities at two measuring heights. The investigated cases correspond to dense bubbly flow conditions, as the gas holdup is in the range of $\varepsilon_G = 1.02 - 7.55$ %. Such velocity measurements in pilot-plant scale are rarely reported in the literature, in particular for a counter-current bubble column, especially considering the "large-scale" concept. The liquid velocity measurements will complete a previous dataset concerning both "globalscale" (i.e., gas holdup) and "local-scale" (i.e., bubble size and shape, local volume fraction, ...) measurements [7], under the same operating conditions, aiming to provide a precise understanding of bubble column fluid dynamics and to offer a comprehensive validation database for Computational Fluid-Dvnamic (CFD) model [15].

This paper proceeds as follows. First, the experimental setup and methods are described; subsequently, the experimental results are presented and coupled with the previously measured bubble size distributions and local void fractions. Finally, the main outcomes and outlooks are discussed.

2. Experimental setup and methods

Herein, the experimental setup, the measurements methods and the experimental matrix are presented and discussed.

2.1. Experimental setup

The experimental facility, displayed in Figure 2, is a non-pressurized vertical pipe made of Plexiglas[®] (ADPlast s.r.l., Milano, Italy) with a height of 5.3 m and an inner diameter of 0.24 m. A pressure reducer controls the pressure upstream of the rotameters ((1) and (2) in Figure 2), which are used to measure the gas flowrate (accuracy $\pm 2\%$ f.s.v., E5-2600/h, manufactured by ASA, Italy). A pump, controlled by a bypass valve, provides water recirculation and a rotameter (3) in Figure 2) measures the liquid flowrate (accuracy $\pm 1.5\%$ f.s.v., G6-3100/39, manufactured by ASA, Italy). The gas sparger is a "spider-gas sparger", which is similar to the distributors used in industrial application; the gas sparger has hole-diameters in the range of $d_0 = 2-4$ mm. Owing to the gas sparger openings, the produced bubble size distribution (BSD) is polydispersed (see the measurements obtained by Besagni and Inzoli [7]). Filtered air and filtered, deionized water were used; during the experiments, the air and water temperatures were maintained constant at room temperature (T = 22 °C). The values of gas density (used to compute the superficial gas velocity), are based upon the operating conditions existing at the column midpoint, owing to the large scale of the experimental setup. The initial liquid level is set at 3.0 m above the gas sparger, thus leading to an aspect ratio (viz. the ratio of the initial liquid level to the bubble column diameter) above the critical value of 5, reported by Wilkinson et al. [14] and discussed by Besagni et al [2].



Figure 2. The experimental setup.

2.2. Experimental method

All the liquid velocity measurements were performed 0.7 m and 1.9 m above the gas sparger, to obtain void fraction profiles comparable with the local void fraction profiles reported by Besagni and Inzoli [7]. The experimental method is a shadowgraphy technique, based on a Q-VIT high-speed camera (AOS Technologies AG) with a backlight (air-cooled 400W halogen lamp); an example of the outcome of this method is displayed in Figure 3 and in Figure 4. The PTV method proceeds in two steps (particle identification and particle tracking); the interested reader may refer to the description proposed by Hessenkemper and Ziegenhein [13]. The sampling bias, which is found in multiphase flows owing to the coupling between the dispersed and the continuous phases, is treated with a hold processor, accordingly with the proposal of Ziegenhein and Lucas [16]. Prior to measurements the applied method and the sampling bias treatment were verified by considering integral values and comparing the PSTV method with PSIV methods in a simplified measurement setup (the flow was seeded with 2 g of 50 µm Polyamide particles - Dantec Dynamics A/S). In order to obtain a quasi-twodimensional measuring plane, the depth of field was adjusted to 2 mm (calibrated as described in ref. [13]) by using a Samyang lens with an aperture of 2.0, a focal length of 135, and a focus distance of 0.13 m. The optical setup, was placed on a linear unit to move it stepwise along the column radius, in the range of 0.119 - 0 m (Figure 3a). With this approach, 11 measuring planes of 6x25.3 mm (Width x Height) were obtained, having a resolution of 397x1686 pixels along the radial coordinate (11.5 µm/Pixel). This method of moving the camera in radial direction is executed for the 15 flow configurations listed in the next sub-section (viz. 330 total measurements). The camera was stepwise moved until representative test prints were detected as blurred from the edge detector algorithm. It is worth noting that, despite the bubble column is operated in the homogeneous regime, a bubble plume with very long time scales exists (owing to the pseudo-homogeneous flow regime properties). Even after 1000 seconds, the averaged velocities show some changes over time. Nevertheless, the changes are rather small so that a 1000-second measuring time was a good compromise between accuracy and effort; additionally, a 1000-second measuring time has been also used for the optical probe measurements performed by Besagni and Inzoli [7]. A set of 10 images with 1000 frames per second was recorded every 2 seconds for 1000 seconds for all 11 measuring planes so that in total 55000 images in 3 hours measuring time were recorded for each operation point.



Figure 3. *PVT* method: measurement method.

2.3. Experimental matrix

Taking into account the previous experimental investigations [7], this paper focuses on the description of the pseudo-homogeneous flow regime. In particular, the camera setup was installed for the particle tracking measurements at the two measuring heights of the optical probes (1.9 m and 0.7 m above the gas sparger). The flow conditions tested are listed in Table 1; in addition, Table 1 lists the flow regime transition points for the different superficial liquid velocities; it can be observed that the tested conditions falls within the pseudo-homogeneous flow regime (as also displayed in **Errore. L'origine riferimento non è stata trovata.**Figure 5) in dense bubby flows conditions (ε_G is in the range of 1.02 – 7.55 %).



Figure 4. Local liquid velocity: example of outcome of the method.

Table 1. Operating conditions experimentally evaluated in the present study (marked as \checkmark) - in bracket, the corresponding gas holdup [%], measured by Besagni and Inzoli [7], is specified.

U_L [cm/s]	$U_{G,trans}$ [cm/s]	$U_G [m cm/s]$				
		0.37	0.74	1.11	1.48	1.85
0.0	2.63	\checkmark	\checkmark	\checkmark	\checkmark	\checkmark
		(1.02 %)	(1.64 %)	(3.72 %)	(5.30 %)	(6.60 %)
-3.3	2.45	\checkmark	\checkmark	\checkmark	\checkmark	
		(1.32 %)	(2.91 %)	(4.61 %)	(6.63 %)	
-6.6	2.08	\checkmark	\checkmark	\checkmark		
		(1.70 %)	(3.62 %)	(5.99 %)		
-9.2	1.73	\checkmark	\checkmark	\checkmark		
		(2.44 %)	(5.00 %)	(7.55 %)		



Figure 5. Operating conditions experimentally evaluated in the present study compared with the regime transition points (as evaluated by Besagni and Inzoli [7]).

3. Experimental results

Herein, the experimental results are presented and discussed in terms of both upward and angular velocities. First, the single-phase flow situations are presented; second, the two-phase flow conditions are discussed. Such presentation of the experimental results is supposed to better stress the influence of the dispersed phase on the fluid dynamic properties imposed by the liquid phase.

3.1. Single phase flow conditions

Measurements of the single phase flow situation (viz. $U_G = 0$ cm/s, changing U_L) were performed, to provide preliminary observations concerning the influence of the inlet and outlet boundary conditions on the flow field: the averaged liquid velocity profile for the single-phase flow is flat in the center with a steep slope towards the wall (Figure 6), owing to the boundary conditions imposed by the wall itself. Such velocity profiles are typical of turbulent single-phase flow conditions. Increasing U_L , it is observed that the profiles are just shifted to higher absolute velocities (figure 6a). Due to the asymmetrically inflow in the basin on top of the column (see Figure 7a), the water is rotating in the basin (despite a porous plate was placed at the bottom of the gas sparger, Figure 7b). This rotation can be clearly observed in the angular velocity (Figure 6b). Indeed, in the angular velocity, an asymmetry can be seen since the sign change is not in the center of the column for superficial liquid velocities above $U_L = -3.3$ cm/s. Also from visual observation, this circular current is preserved towards the bottom of the bubble column.



Figure 6. Upward and angular liquid velocity velocities profiles without gas ($U_G = 0$ cm/s) at different superficial liquid velocities.



Figure 7. Outlet conditions for the liquid phase

3.2. Two phase flow conditions

Figure 8 and Figure 9 display the time-averaged upward liquid velocities for the operating conditions listed in Table 1. In the different conditions, the liquid flows upward in the centre of the column, becomes zero at an intermediate radial position and then flows downward near the wall, which is in agreement with the liquid velocity profiles displayed in Figure 1 and discussed in ref. [5]. For all the conditions, the minimum velocity is located at approximately 7.5 mm away from the wall. Increasing U_G , at fixed U_L , the gradients in the liquid profiles get steeper, which can be nicely seen for the batch mode (Figure 8a). These findings are in qualitative agreement with the modeling approach proposed by Trividi et al. [5], despite some differences concerning the magnitude of the velocity. The upward velocity at wall increases with U_G (in agreement with ref. [5]); conversely, the influence of U_L on the upward velocity at wall is limited (which is different from the predictions of ref. [5]).



Figure 8. Averaged upward liquid velocities for different superficial gas velocities at different counter-current flow setups 0.7 m above the gas sparger.



Figure 9. Averaged upward liquid velocities for different counter-current flow setups at different superficial gas velocities 0.7 m above the gas sparger.

It is observed that the upward liquid velocity profiles are shifted downwards with no further modification, while increasing U_L at fixed U_G (Figure 8 and Figure 9). This observation is very interesting and suggests that the prevailing effect of the counter-current operation mode is to decelerate the bubbles raising the column. These experimental findings confirm the previous discussions proposed by Besagni and Inzoli [7] and provide a rational explanation regarding the prevailing effects of the liquid velocity on the bubble column fluid dynamics: the counter-current mode increases the gas holdup and the local void fraction; conversely, it decreases the local velocity. In addition, Figure 10 displays the influence of the counter-current mode on the angular velocity: the bubbly flow generates a clear circulating flow, which is changing its direction towards the center. This counter-current angular flow is slightly hindered when the liquid velocity is increased to $U_L = -3.3$ cm/s; conversely, owing to the asymmetrically boundary conditions, the circulating flow of the liquid inflow dominates the bubbly flow $U_L = -6.6$ and -9.2 cm/s



Figure 10. Averaged angular liquid velocities for different superficial gas velocities at different counter-current flow setups 0.7 m above the gas sparger.

3.3. Description of the "local-scale" bubble column fluid dynamics

As stated above, three parameters determine the bubble column fluid dynamics at the "local-scale": (a) the liquid velocity, (b) the local gas fraction, and (c) the bubble size distributions (BSDs). To this end, Figure 11 couples the local liquid velocity profiles obtained in this study (Figure 8a), with the local void fraction profiles measured in ref. [7] and the BSDs reported in ref. [17] (which is a followup of the experimental investigation proposed in ref. [7]). The liquid flows upward in the center of the bubble column, it becomes zero at an intermediate radial position and, finally, it flows downward near the wall. The liquid velocity profiles is strictly related to the BSDs by the lift force. It is observed that near the wall, there is a higher number of "small bubbles". Indeed, it is well-known that the direction of the transversal lift force depends upon the bubble size and shape: for "small bubbles", the lift force acts in the direction of decreasing liquid velocity (i.e. the lift force pushes the small bubbles toward the wall); conversely, for "large bubbles" the lift force changes its direction (a force that can be assimilated to the lift force tends to push large and deformed bubbles towards the center of the column [18-20]). The distribution of the gas phase, imposed by the prevailing BSDs, determines the local void fraction profiles; owing to the poly-dispersed nature of the BSDs and given the local liquid velocity field, the local void fraction profile is center-pecked owing to the "large bubbles" moving towards the center of the bubble columns. Future studies should couple all these data by using a lift force approach and extend the present comparison to the counter-current mode.



O UG = 0.37 cm/s O UG = 0.74 cm/s O UG = 1.11 cm/s O UG = 1.49 cm/s O UG = 1.88 cm/s

Figure 11. The "*local-scale*" bubble column fluid dynamics: liquid velocity measurements (Figure 8), local void fraction [7] and bubble side distributions [17] – batch mode.

4. Conclusions

A complete knowledge of bubble column fluid dynamics rely on understanding the global and the local fluid dynamic properties. Indeed, datasets encompassing the different scales are important to validate the capability of existing CFD models for upscaling. Unfortunately, most of the previous literature focused on the "global-scale" fluid dynamics, whereas a limited attention was devoted to the "local-scale". It is known that three parameters characterize the fluid dynamics at the "local-scale": liquid velocity, local gas fraction, and bubble size distributions (BSDs). The present study experimentally obtains local liquid velocity and turbulence data in a large-scale bubble column in the batch and in the counter-current flow modes. Such data complete a previous dataset concerning both global-scale (i.e., gas holdup) and local-scale (i.e., bubble size and shape, local volume fraction, ...) measurements, under the same operating conditions. It is found that, increasing the superficial liquid velocity in the counter-current mode flow only shifts the liquid velocity profiles downward. From the angular velocity, a distinct counter current circulating flow in batch mode is observed, which is dominated by the rotation of the counter-current liquid flow. The rotating counter-current liquid flow is very likely an inlet effect. The liquid velocity have been coupled with our previous measurements, to completely understand the "local-scale" fluid dynamics. In conclusion, the present study completes our previous studies and provides a comprehensive validation database for Computational Fluid-Dynamic (CFD) model validation. It has not escaped out notice that these data, coupled with the radial measurements may be used-in future studies-to provide detailed information on the local void fraction profiles and for developing a revised Drift Flux model [21] of large-scale bubble columns.

Acknowledgement

This work was partially funded by the Deutsche Forschungsgemeinschaft (DFG) under the project number 335448239.

References

- [1] Rollbusch P, Bothe M, Becker M, Ludwig M, Grünewald M, Schlüter M and Franke R 2015 *Chem Eng Sci* **126** 660-678
- [2] Besagni G, Inzoli F and Ziegenhein T 2018 *ChemEngineering* **2** 13
- [3] Bartrand T A, Farouk B and Haas C N 2009 Int. J. Multiphas Flow 35(2) 171-184
- [4] Leonard C, Ferrasse J H, Boutin O, Lefevre S, and Viand A 2015 *Chem Eng Res Des* **76** 391-421
- [5] Ueyama K and Terukatsu M 1979 *AIChE J* **25**(2) 258-266.
- [6] Trivedi R, Renganathan T and Krishnaiah K 2018 Chem Eng J 338 636-650,
- [7] Besagni G and Inzoli F 2016 *Chem Eng Sci* **146** 259-290
- [8] Roustan M and Wang R Y 1996 Ozone Sci Eng 18 99-115
- [9] Hidaka N, Mizuguchi K I and Matsumoto T 1998 J Chem Eng Jpn 31 1016-1019
- [10] Jin H, Yang S, He G, Wang M and Williams R A 2010 J Chem Technol Biotechnol 85 1278-1283
- [11] Deen N G, Westerweel J and Delnoij E 2002 Chem. Eng. Technol 25
- [12] Ziegenhein T, Garcon M and Lucas D 2016 Chem Eng Sci 153 155-164
- [13] Hessenkemper H and Ziegenhein T 2018 Int J Multiph Flow Accepted
- [14] Wilkinson P M, Spek A P and van Dierendonck L L 1992 AIChE J. 38 544-54
- [15] Lucas D, Rzehak R, Krepper E, Ziegenhein T, Liao Y, Kriebitzsch S and Apanasevich P 2016 Nucl Eng Des 299 2-11
- [16] Ziegenhein T and Lucas D 2016 Flow Meas Instrum 48 36-41
- [17] Besagni G, Inzoli F, Ziegenhein T and Lucas D 2017 Petroleum In press
- [18] Lucas D, Prasser H M and Manera A 2005 *Chem Eng Sci* 60 3609-3619
- [19] Tomiyama A, Tamai H, Zun I and Hosokawa S 2002 Chem Eng Sci 57 1849-1858
- [20] Santarelli C and Fröhlich J (2016) Int J Multiph Flow 81 27-45
- [21] Zuber N and Findlay J A 1965 J. Heat Transf 87 453-68