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Gas Holdup in Trayed Bubble Column Reactors

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The sectionalization of conventional bubble columns by perforated trays into trayed bubble columns (TBCs) has much potential in biochemical and petroleum refining processes, as an effective way to improve the gas−liquid contacting efficiency and reduce the liquid backmixing. In this work, the effects of tray geometry, superficial gas velocity, and superficial liquid velocity on the overall gas holdup have been investigated. It was detemrined that the trays significantly increase the overall gas holdup. The tray hole diameter and superficial gas velocity were determined to be the most important factors.

Introduction

The sectioning of conventional bubble columns by perforated trays into trayed bubble columns (TBC) has considerable potential in biochemical and petroleum refining processes. It is an effective way to improve the gas−liquid contacting and the intensity of transport and to reduce the axial dispersion back mixing of the gas and liquid phases.[1−4] In bioprocessing, Schugerl et al.[5] analyzed TBC's performance as biological fermentors in aerated slurry systems (e.g., continuous single-cell protein production). Bakopoulos[6] described a very promising application as photobioreactors employing algae to produce fine chemicals. Other chemical processes that have benefited from the unique hydrodynamic and mixing characteristics of TBC include ozonation of drinking and wastewater,[7] oxidation of anthraquinone to produce hydrogen peroxide,[8] and Fischer−Tropsch synthesis of paraffins from syngas by use of a slurry catalyst.[9] The Visbreaking of petroleum residues,[10] a petroleum refining process, is a recent and very important application of trayed bubble columns.

Overall gas holdup (ϵ_g) is one of the key parameters in the design and scale-up of trayed bubble columns. In these columns, the operating mode and conditions, physical properties, flow arrangement, plate internals, and sparger design affect the overall gas holdup. Table 1 summarizes most of the published work in the open literature regarding overall gas holdup in multistage bubble columns.

Hence, the focus of this work is to study the relative effect of trays, superficial gas and liquid velocities, and gas sparger type on the overall gas holdup of cocurrent upward trayed bubble column reactors.

Experimental Section

Figure 1 shows the schematic of the continuous cocurrent trayed bubble column setup and the trays used in this work. The column is made of four intermediate sections plus a top (disengagement) and a bottom (plenum) section, all built of Plexiglas and attached by flanges. The intermediate sections have an inside diameter of 19 cm and a total height of 52 cm each. The upper section has the same diameter as the intermediate ones, but it is only 33 cm tall. There is also a 33 cm tall cone-shaped plenum section, where the gas and liquid phases enter the column and are mixed. The total height of the column, from the base of the plenum to the top of the disengagement section, is 241 cm.

This is a five-stage setup unit with a total of four trays, which are mounted between the flanges. Three different types of trays were studied, whose design and dimensions can be seen in the lower panel of Figure 1. Tray types 1 and 3 both have the same total open area (10.2% OA) but different hole diameters (1.74 and 0.6 cm, respectively). Conversely, tray types 2 and 3 share the same hole diameter (0.6 cm) but have different total open areas (5.2% and 10.2% OA, respectively). The holes in the trays are distributed in a triangular pitch design. Filtered and compressed air and city water were used as the gas and liquid phases, respectively. The liquid phase is pumped from the feed tank to the column at a maximum flow rate of 38 L/min. Once the liquid fills the column, it overflows through the side of the column's upper section, from where it returns to the feed tank. There are two ball valves with fast open−close mechanisms on the air and liquid lines for the measurement of the overall gas holdup. The valves are close to each other so that they can be manually shut off at the same time.

The overall gas holdup (ϵ_g), which is the volumetric gas fraction based on the total volume of the gas−liquid bed, was measured by the ratio between the increase in the bed height and the total height of the gas−liquid bed ($\epsilon_g = \Delta H/(\Delta H + H_l)$).

The overall gas holdup was determined in a range of superficial gas velocities from 1 to 25 cm/s so that both bubbly and churn-turbulent flow regimes could be covered. Three values of superficial liquid velocity were selected: 0.5, 1.0, and 1.5 cm/s. At each set of experimental conditions, the measurement of the overall gas holdup was repeated three times, and the mean value was reported along with the corresponding 95% confidence interval.

In addition, pressure drop measurements from a Valydine differential pressure transducer (DP15) between two points (separated by 20 cm) at four axial locations along the column (Figure 1, top panel) were used to measure average gas holdup at these four sections. For all the experiments, the pressure signals were measured over 4 min at a sampling frequency of 100 Hz. The time-average value corresponding to the chosen sampling period was taken.

The time-averaged gas holdup between these two points within the stage can be approximately determined from the pressure drop measurement (Figure 1, top panel), with the assumptions that the dissipation term between the pressure

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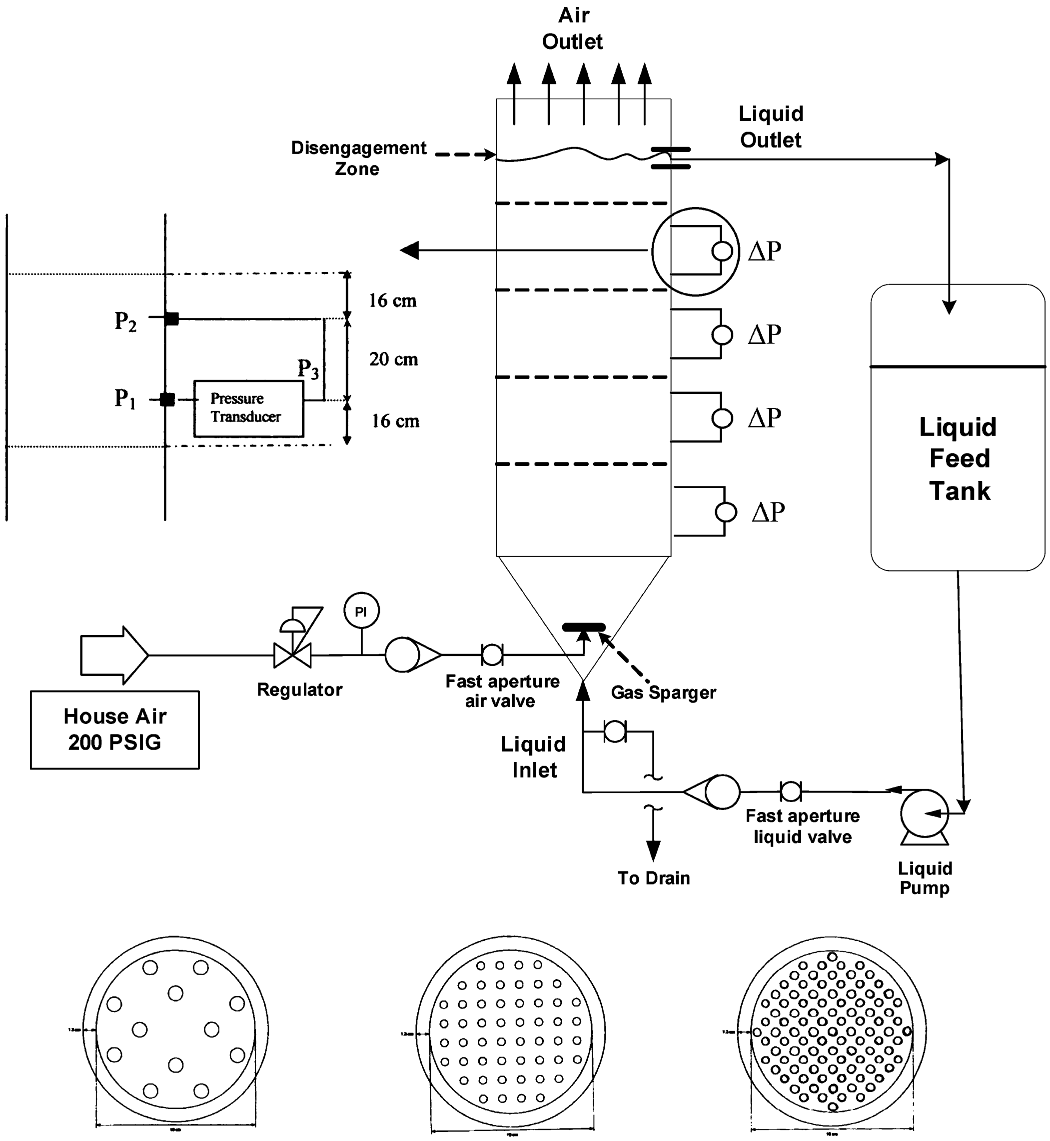
Table 1. Summary of the Correlations and Experimental Data Available for Overall Gas Holdup in Trayed Bubble Columns

gas−liquid system	range of parameters and operating conditions	results
	Schugerl et al.[5]	
air, 0.5−2% methanol, 0.5−1% ethanol, 0.5−2% glucose	countercurrent $U_g = 1-8$ cm/s; $U_l = 2$ cm/s $D_c = 20$ cm; $H_c = 381$ cm 6 stages: $H_s = 10, 50$ cm; $d_0 = 0.2, 0.4$ cm; OA = 12.5%, 28%	experimental data
	Kato et al.[11]	
air−water−kerosene	cocurrent upflow $U_g = 1.5-13$ cm/s; $U_l = 0.1-1.0$ cm/s $H_c = 200, 220$ cm; $D_c = 6.6, 12.2$ cm 4 and 8 stages: $H_s = 25, 50$ cm; $d_0 = 0.65, 0.90, 1.20$ cm; OA = 6%, 12.8%, 28.9% perforated plate of Karr[20]	$\epsilon_g = \dfrac{U_g}{30 + 3.34{U_g}^{0.8}}$
	Nishikawa et al.[12]	
air, N_2, tap water, (carboxymethyl)cellulose, 10−60% sugar solution	countercurrent $U_g = 1-7$ cm/s; $U_l = 0-3$ cm/s $D_c = 15, 10$ cm; $H_c = 140, 115$ cm 3 and 4 stages: $H_S = 40, 35$ cm; $d_0 = 0.5, 0.75, 0.8, 0.5, 0.3$ cm	experimental data
	Chen et al.[13]	
air, water	cocurrent upflow $U_g = 0-8.22$ cm/s; $U_l = 0-6.12$ cm/s $H_c = 300$ cm, $D_c = 7.5$ cm stainless steel wire screen:	stainless steel wire screen mesh: $\epsilon_g = 0.0416{U_g}^{1.23}{U_l}^{-0.092}$
	37 trays, $d_0 = 0.058$ cm OA = 64%, $H_S = 5$ cm Karr type trays: 84 trays, $d_0 = 1.27$ cm, OA = 53%, $H_S = 2.54$ cm	perforated plates (Karr type): $\epsilon_g = 0.0448{U_g}^{0.81}{U_l}^{-0.055}$
	Chen and Yang[14]	
air, tap water, 0.2% (carboxymethyl)cellulose	cocurrent upflow $U_g = 0-7$ cm/s; U_l = not specified $D_c = 5, 7.5, 15$ cm; $H_c = 300$ cm 6-mesh stainless steel sheets: 38 stages; OA = 64%; $H_S = 5$ cm	$V_S = 0.115{\epsilon_g}^{-0.182}$ $V_S = \dfrac{U_g}{\epsilon_g} - \dfrac{U_l}{1 - \epsilon_g}$
	Vinaya[15]	
air−water/air−kerosene	countercurrent $U_g = 0.42-11.6$ cm/s; $U_l = 0.7-1.2$ cm/s; $D_c = 9.8-15.4$ cm; $H_c = 110, 180$ cm	bubbly regime: $\epsilon_g = 2.4\left(\dfrac{{U_g}^2}{gd_0}\right)^{0.54}\left(\dfrac{d_0}{H_s}\right)^{0.26}\mathrm{OA}^{-0.22}\left(\dfrac{\sigma}{\sigma_w}\right)^{-0.3}$
	perforated trays: $d_0 = 0.3, 0.5, 1, 1.2$ cm; OA = 10%, 12.5%, 20%, 30%, 38%, 52%; $H_S = 5, 12, 20, 85$ cm	churn-turbulent regime: $\epsilon_g = 0.59\left(\dfrac{{U_g}^2}{gd_0}\right)^{0.4}\left(\dfrac{{U_l}^2}{gd_0}\right)^{0.03}\mathrm{OA}^{-0.03}{H_S}^{-0.21}\left(\dfrac{\sigma}{\sigma_w}\right)^{-0.41}$
	Yamada and Goto[16]	
hydrogen−water−1,2-dichloroethane−carbon	cocurrent $U_g = 1-3$ cm/s, $U_{l,water} = 5 \times 10^{-2}$ cm/s, $U_{l,organic} = 5 \times 10^{-2}$ cm/s $H_c = 22.5$ cm stainless steel mesh: $d_0 = 0.2$ mm, OA = 73%, $H_S = 1.5, 2.5, 4.5, 7.5, 22.5$ cm	$\epsilon_g = \dfrac{U_g}{0.01 + 1.6{U_g}^{0.8}}$

probes is negligible compared to the hydrostatic term and that the gas head is neglected since the liquid density is much larger than the density of the gas ($\epsilon_{g(sectional)} = \Delta P_{transducer}/(\rho_l gh)$). The average gas holdup in the column can be calculated by integrating the axial gas holdup profile obtained by pressure drop measurement along the column, using the four discrete values of gas holdup measured ($\epsilon_g = (\int_a^b \epsilon_g(z)\,dz)/(b - a)$). The limits of integration a and b in the above equation correspond to the points in the column located above the sparger and below the disengagement region, respectively. In the estimation of the gas holdup within the stage, the contribution of the measurements due to the pressure drop across the trays



Tray # 1: 12 holes, 1.74 cm tray hole diameter, 10.2 % open area

Tray # 2: 52 holes, 0.6 cm tray hole diameter, 5.2 % open area

Tray # 3: 105 holes, 0.6 cm tray hole diameter, 10.2 % open area

Figure 1. (Upper panel) Trayed bubble column experimental setup. (Lower panel) Configurations and dimensions of the trays.

is not considered since the pressure measurements are located within the column stages and not between them.

Results and Discussion

Comparison of Overall Gas Holdup Obtained by Gas−Liquid Disengagement and by Pressure Drop Measurements. Figures 2 and 3 show the comparison for the column without trays and for the column with tray type 1 at $U_l = 1.0$ cm/s, respectively. For the column without trays, it can be seen that the match between the overall gas holdups obtained by the two methods is reasonably good. On the other hand, for the trayed column, the gas holdups estimated by pressure drop measurements are slightly lower than the overall gas holdup data measured by gas−liquid disengagement. This difference is attributed to a small layer of gas that was observed to form under the trays during the experimental runs, particularly in the high superficial gas velocity range. The fraction of the total gas dispersed in the column, due to the gas accumulated under the trays, can be approximately determined by subtracting the pressure drop-estimated gas holdup from the gas−liquid disengagement-estimated gas holdup. Further, if one assumes that the total gas trapped under the trays is equally distributed among the four trays and that this region contains no liquid (e.g., gas holdup is equal to 1), and then the height of the gas cap can be determined. The gas cap steadily increases with superficial gas velocity, ranging from 0.6 cm at $U_g = 1$ cm/s to 1.32 cm at $U_g = 16$ cm/s (approximately 3% of the total height of the stage). An attempt to measure the time-averaged gas holdup in the gas cap region using the differential pressure drop method was conducted. Unfortunately, the pressure drop transducers available were not sensitive enough to accurately resolve the small pressure drop in the region beneath the trays.

In conclusion, the overall gas holdup, as measured by gas−liquid disengagement (which also includes the gas accumulated under the trays, since it displaces the liquid to be measured as part of the gas−liquid bed height at the steady state), provides a fair representation of the total amount of gas dispersed in the continuous trayed bubble column used in this work. It is worth noting that Dreher and Krishna[4] reported that the height of the gas cap decreases noticeably with an increase in column

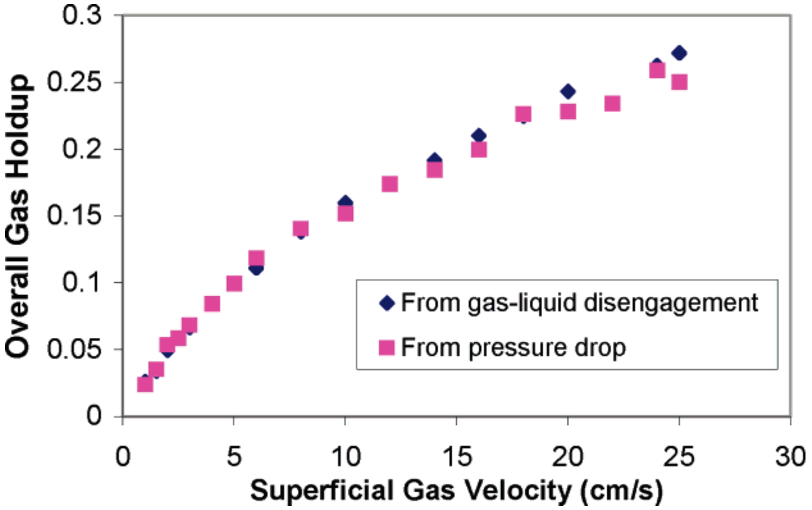


Figure 2. Comparison of the gas−liquid disengagement and the pressure drop measurements in the estimation of the overall gas holdup in the bubble column without trays at $U_l = 0.5$ cm/s.

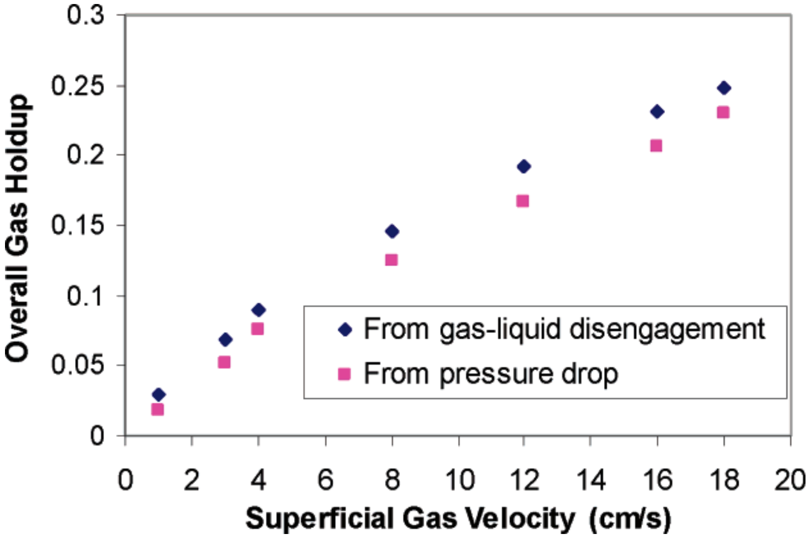


Figure 3. Comparison of the gas−liquid disengagement and the pressure drop measurements in the estimation of the overall gas holdup in the bubble column with tray type 2 ($d_0 = 0.6$ cm and 5.2% OA) at $U_l = 0.5$ cm/s.

diameter, making gas accumulation under the trays insignificant in industrial-size trayed bubble column reactors.

Effect of Ssuperficial Gas Velocity. Figure 4 compares the overall gas holdup between the trayed bubble column and the bubble column without trays at a range of superficial gas velocities ($U_g = 1-25$ cm/s) and at one superficial liquid velocity ($U_l = 1$ cm/s). Two different regions can be clearly distinguished. In the low superficial gas velocity region ($U_g <$ 4−6 cm/s), known as the bubbly flow regime, there is an almost linear increase in the gas holdup with superficial gas velocity. As can be seen, the trays have little or no effect on the gas holdup because, in this regime, the diameter of the holes is bigger than the mean bubble diameter and the gas can easily travel across the trays. As the gas velocity is increased, the gas−liquid flow becomes turbulent and the hydrodynamic properties of the system change noticeably. Visually, the bubbles present a wide distribution of sizes, shapes, and rise velocities. In this churn-turbulent regime, the almost linear relation between gas holdup and superficial gas velocity is no longer seen. It is in this turbulent region where the introduction of perforated trays in the column significantly increases the overall gas holdup as compared to the column without trays.

Effect of Ttrays. As previously mentioned and shown in Figure 4, the effect of the trays on the overall gas holdup is significant in the churn-turbulent flow regime. The average gas holdup increment measured in the trayed bubble column, as compared to the column without trays, varied between 20% and 50% depending upon type of tray, superficial gas and liquid velocities, and the gas sparger used. The gas holdup increase in the presence of trays is due to the reduction of liquid circulation velocity and the increase in gas-phase residence time within the stages. In addition, the redistribution of the gas phase in each of the stages helps to reduce the overall bubble size and the bubble coalescence rate. Therefore, the increase of the gas residence time, smaller average bubble size diameter, and lower rate of bubble coalescence all contribute to the increase of the gas holdup in the column when the trays are present.

Important parameters to consider in the design of the trays are hole diameter (d_0) and total tray open area. The experimental results in Figure 4 show that tray hole diameter has a stronger effect on the overall gas holdup than total tray open area, as can be seen by comparing the results obtained between tray types 1 and 3 (same total open area but different hole diameter) and between tray types 2 and 3 (same hole diameter but different total open area). Tray hole diameter directly controls the size of the bubbles entering each of the stages. On the other hand, an increase in tray open area, for trays with the same hole diameter, translates into an increase in the total amount of energy

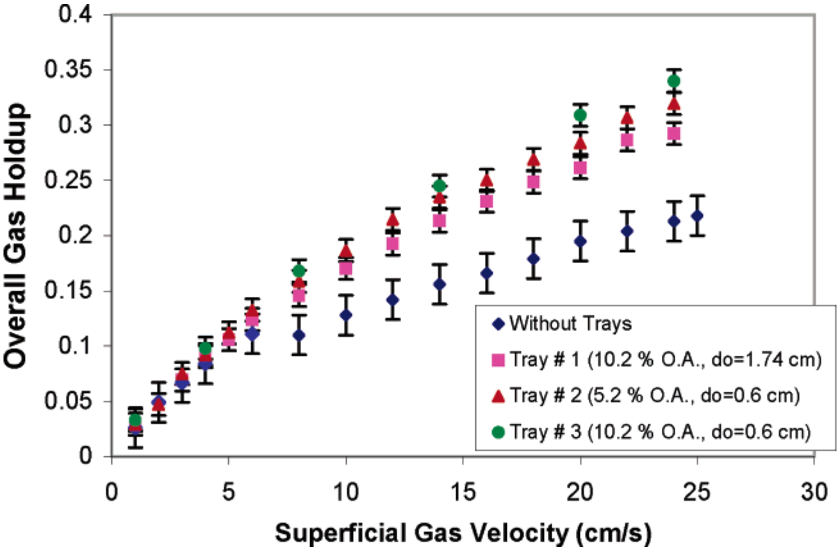


Figure 4. Comparison of the overall gas holdup between the bubble column with and without trays at $U_l = 0.5$ cm/s.

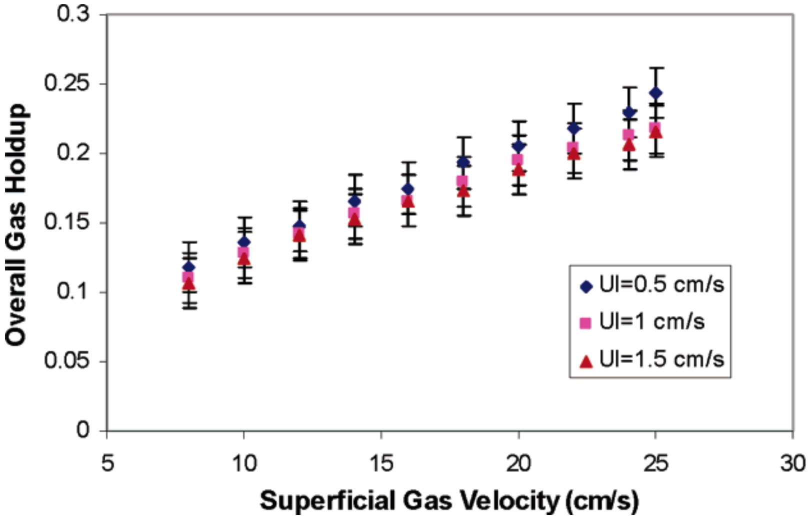


Figure 5. Effect of the superficial liquid velocity on the overall gas holdup in the bubble column with tray type 1 ($d_0 = 1.74$ cm and 10.2% OA).

dissipated due to the larger number of holes in the tray. Hence, the gas holdup measured with tray type 3 is slightly larger than with tray type 2.

Effect of Superficial Liquid Velocity. Figure 5 shows the effect of superficial liquid velocity on the overall gas holdup in the trayed bubble column (tray type 1). There is a slight decrease in the overall gas holdup (from 2% to 10%) for each increment of 0.5 cm/s in the superficial liquid velocity. Joshi et al.[17] reported that in single-stage bubble columns, with the liquid flowing concurrently with the gas phase, an increase of the superficial liquid velocity decreases the extent of liquid downflow near the wall of the column. The overall effect of the net liquid flow is to increase the upward liquid velocity in the riser area and decrease the downward velocity in the downcomer area. The liquid flows upward where the gas holdup is high, which is in the central region of the column. As a result, the overall bubble rise velocity is higher in the central region, where the gas fraction is also high, which in turn reduces the gas residence time. Joshi and Sharma[18] and Patil et al.[19] postulated the formation of a gas−liquid circulation cell in each of the stages of a multistage bubble column. Thus, an increase in the superficial liquid velocity increases the circulation of the liquid in the upward direction, which in turn accelerates the bubble motion. The reduction of the gas residence time results in the observed decrease of the gas holdup in the column.

Comparison of Experimental Data with Published Correlations. In Figure 6, the overall gas holdups obtained in this work are plotted versus the values predicted by four published correlations developed for upflow cocurrent trayed bubble columns (Table 2). Only the correlation from Kato et al.[11] is able to reasonably predict the experimental gas holdup data within 10% mean relative error. The configuration and geometry of the trayed bubble column and the range of the chosen superficial gas and liquid velocities ($U_g = 1.5-13$ cm/s, $U_l = 0.1-1.0$ cm/s) used by Kato et al.[11] are within the range of the values selected in this work. However, Kato et al.[11] found that the overall gas holdup was sensitive only to the variation in the velocity of the gas phase: the type of tray used, the properties of the phases, and the tray spacing did not produce any effect on the overall gas holdup. It can also be seen in Figure 6 that the correlations of Chen et al.[13] and Chen and Yang[14] predict the experimental overall gas holdups obtained in this work better in the low range than in the high range of superficial gas velocity.

Nonetheless, Yamada and Goto[16] fitted their experimental gas holdup data with the same type of empirical expression used by Kato et al.;[11] their correlation overpredicts the experimental data of this work with a mean relative error larger than 160%. This would be due to the use of a very small column ($H_c = 22.5$ cm, $D_c = 2$ cm), which was partitioned into 3, 5, 9, and

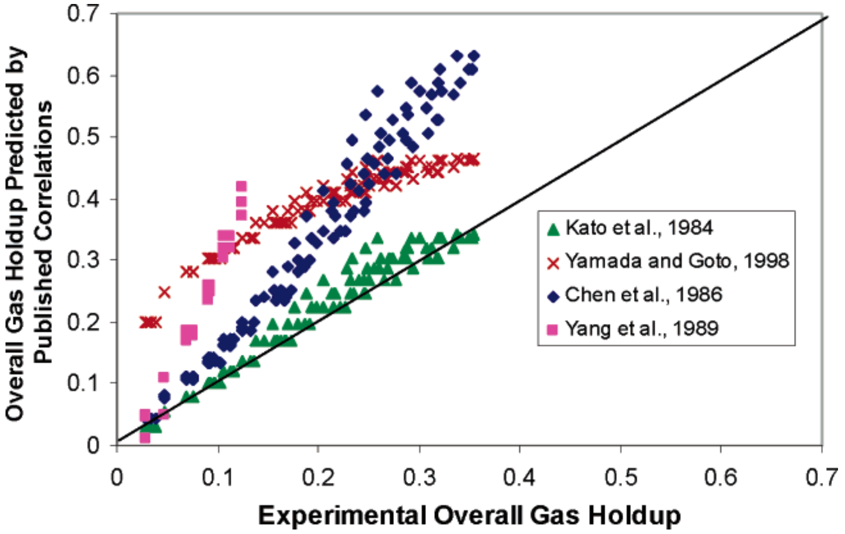


Figure 6. Comparison between the experimental data of this work and the predictions of the reported correlations[11,13,14,16] of the overall gas holdups in the trayed bubble column.

Table 2. Published Correlations for Upflow Cocurrent Trayed Bubble Columns in Air−Water System

ref	correlation	mean relative error (%)
Kato et al.[11]	$\epsilon_g = \dfrac{U_g}{30 + 3.3U_g^{0.8}}$	9.1
Chen et al.[13]	$\epsilon_g = 0.0448U_g^{0.81}U_l^{-0.055}$ $V_S = 0.115\epsilon_g^{-0.182}$	66
Chen and Yang[14]	$V_S = \dfrac{U_g}{\epsilon_g} - \dfrac{U_l}{1 - \epsilon_g}$ V_s = slip velocity (=) m/s	153
Yamada and Goto[16]	$\epsilon_g = \dfrac{U_g}{0.01 + 1.6U_g^{0.8}}$	160

15 stages by plates made of stainless steel screen (d_0 = 0.12 cm and 73% OA).

Although the correlation of Kato et al.[11] reasonably predicts the experimental data of this work, there is a need to develop an empirical correlation that, in addition to the superficial gas velocity, also incorporates the effect of other important parameters such as tray geometry (tray hole diameter, tray open area), stage height, physical properties of the phases, and superficial liquid velocity. However, an extensive database of experimental data, collected at a wide range of the important operating and design parameters, needs to be available. Unfortunately, in this work, the range of the variables studied is not wide enough to develop such a general correlation for the prediction of the overall gas holdup in trayed bubble columns. Therefore, the obtained experimental data is used to formulate an empirical expression for the purpose of assessing the effects of the studied design and operating parameters as discussed in the following section.

Empirical Expression for Overall Gas Holdup. The overall gas holdup in the trayed bubble column is represented as a function of the variables studied in this work [$\epsilon_g = f(U_g, U_l, d_0, \mathrm{OA})$] that can be expressed in the form of Froude dimensionless numbers [$\epsilon_g = f(Fr_g, Fr_l, \mathrm{OA}) = kFr_g^{a}Fr_l^{b}\mathrm{OA}^{c}$).

The magnitude of the coefficients k, a, b, and c represents the extent of the effect of these dimensionless groups on the overall gas holdup. These coefficients were estimated by fitting the experimental data via multilinear regression analysis, and the following expressions were obtained. For the bubbly flow regime, $\epsilon_g = 0.914Fr_g^{0.776}Fr_l^{-0.053}\mathrm{OA}^{0.372}$, $k = 0.914$, $a = 0.776 \pm 0.067$, $b = -0.053 \pm 0.017$, and $c = 0.372 \pm 0.141$. The ranges over which the dimensionless numbers were varied are $Fr_g = 0.02-0.33$, $Fr_l = 0.06-0.12$, and OA = 5.52−10.2%. For the turbulent flow regime, $\epsilon_g = 0.317Fr_g^{0.590}Fr_l^{-0.143}\mathrm{OA}^{0.155}$, $k = 0.317$, $a = 0.590 \pm 0.057$, $b = -0.143 \pm 0.048$, and $c = 0.155 \pm 0.080$. The ranges over which the dimensionless numbers were varied are $Fr_g = 0.04-1.03$, $Fr_l = 0.06-0.12$, and OA = 5.2−10.2%.

Good agreement between the experimental overall gas holdups and the estimated values from the empirical expressions has been obtained. The mean relative errors between the experimental and the predicted overall gas holdups in the bubbly and churn-turbulent flow regimes are 11% and 3.5%, respectively.

The relative effect of the different variables studied on the overall gas holdup then can be inferred by looking at the power of the dimensionless numbers. Below, the effect is classified in order of importance, from more to less important. In the bubbly flow regime, $U_g > d_0 > \mathrm{OA} \gg U_l$; in the turbulent flow regime, $U_g > d_0 > \mathrm{OA} > U_l$.

In general, we can see that superficial gas velocity has the most important effect, whereas tray open area and superficial liquid velocity are less important. It should be noticed that the overall gas holdup is more sensitive to the superficial gas velocity in the bubbly regime ($\epsilon_g \propto U_g^{0.776}$) than in the churn-turbulent regime ($\epsilon_g \propto U_g^{0.590}$). The effect of superficial liquid velocity becomes almost insignificant in the bubbly flow regime ($\epsilon_g \propto U_l^{-0.053}$) as compared to the turbulent flow regime ($\epsilon_g \propto U_l^{-0.143}$).

Summary

The effects of tray open area, tray hole diameter, superficial gas velocity, and superficial liquid velocity on overall gas holdup have been studied in a 20 cm in diameter cocurrent upward trayed bubble column with air and water as the gas−liquid system. The overall gas holdup in the column was determined by gas−liquid disengagement and pressure drop measurements. The underprediction of the pressure drop-estimated gas holdup, as compared to the gas−liquid disengagement method, showed that gas tends to accumulate beneath the trays, forming a gas cap. The gas holdup was found to be a strong function of superficial gas velocity. This study has confirmed that the introduction of perforated trays into conventional bubble columns increases the volumetric fraction of the gas phase, as compared to columns without trays. Tray hole diameter seems to play a more important role than total tray open area in the effect that the trays have on the gas holdup. In the cocurrent gas−liquid arrangement, the effect of superficial liquid velocity on the gas holdup in both trayed and single-stage bubble columns is small. There is a slight reduction of the gas holdup with an increase in the superficial liquid velocity. The experimental data collected in this work, for the trayed bubble column, have been fitted to empirical expressions to assess the effects of the studied parameters on the overall gas holdup in the bubbly and churn-turbulent flow regimes.

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Nomenclature

d_0 = tray hole diameter (cm)
D_c = column diameter (cm)
D_L = turbulent diffusivity of the liquid phase (cm/s^2)
g = acceleration of gravity (cm/s^2)
H_c = column height (cm)
ΔH = change of bed height when gas phase is introduced (cm)
H_l = height of the liquid phase in the column before introduction of the gas phase (cm)
H_s = stage height (cm)
L = length (cm)
U_g = superficial gas velocity (cm/s)
U_l = superficial liquid velocity (cm/s)
V_s = slip velocity (cm/s)
ΔP = pressure drop (g/m/s^2)
ϵ_g = overall gas holdup
ρ_g = density of the gas phase (g/cm^3)
ρ_l = density of the liquid phase (g/cm^3)
μ_l = viscosity of the liquid phase (Pa s)

Dimensionless Numbers

Fr_l = Froude number of the liquid phase; $Fr_l = U_l/\sqrt{gd_0}$
Fr_g = Froude number of the gas phase; $Fr_g = U_g/\sqrt{gd_0}$

Literature Cited

(1) Shah, Y. T.; Kelkar, B. G.; Godbole, S. P.; Deckwer, W. D. Design Parameters Estimations for Bubble Column Reactors. *AIChE J.* **1982**, *28* (3), 353−379.

(2) Van Baten, J. M.; Krishna, R. Scale-up Studies on Partitioned Bubble Column Reactors with the Aid of CFD Simulations. *Catal. Today* **2003**, *79−80*, 219−227.

(3) Alvaré, J. Gas Holdup and Liquid Phase Mixing in Trayed Bubble Column Reactors. M.S. Thesis, Department of Chemical Engineering, Washington University in St. Louis, St. Louis, MO, 2002.

(4) Dreher, A. J.; Krishna, R. Liquid-Phase Backmixing in Bubble Columns Structured by Introduction of Partition Plates. *Catal. Today* **2001**, *69*, 165−170.

(5) Schugerl, K.; Todt, J.; Lucke, J.; Renken, A. Gas Holdup and Longitudinal Dispersion in Different Types of Multiphase Reactors and Their Possible Application for Microbial Processes. *Chem. Eng. Sci.* **1977**, *32*, 369−375.

(6) Bakopoulos, A. Scale-up of Three Phase Photobiochemical Reactors for Algae Microorganisms Production. Paper presented at AIChE Annual Meeting, Scale-up/Scale-down of Multiphase Reactors Session, Miami Beach, FL, Nov 1998.

(7) Munter, R.; Kamenev, S.; Sarv, L. Design of a Staged Downflow Bubble Reactor. *Ozone Sci. Eng.* **1990**, *12* (4), 437−455.

(8) Eickhoff, H.; Schutte, R. Design of Perforated Trays for Use in Bubble Column Reactors, Especially for Use in Manufacture of Hydrogen Peroxide. PCT Int. Appl. 2000.

(9) Maretto, C.; Krishna, R. Design and Optimization of a Multi-Stage Bubble Column Slurry Reactor for Fischer−Tropsch Synthesis. In *Third International Symposium on Catalysis in Multiphase Reactors*, Naples, Italy, May 29−31, 2000; pp 111−122.

(10) Palaskar, S. N.; De, J. K.; Pandit, A. B. Liquid-Phase RTD Studies in Sectionalized Bubble Column. *Chem. Eng. Technol.* **2000**, *23* (1), 61−69.

(11) Kato, Y.; Kago, T.; Morooka, S. Longitudinal Concentration Distribution of Droplets in Multi-Stage Bubble Columns for Gas−Liquid Systems. *J. Chem. Eng. Jpn.* **1984**, *17*, 429−435.

(12) Nishikawa, M.; Shino, K.; Kayama, T.; Nishioka, S.; Hashimoto, K. Gas Absorption in a Multi-Stage Gas−Liquid Spouted Vessel. *J. Chem. Eng. Jpn.* **1985**, *18*, 496−501.

(13) Chen, B. H.; Yang, N. S.; Mc Millan, A. F. Gas Holdup and Pressure Drop for Air−Water Flow through Plate Bubble Columns. *Can. J. Chem. Eng.* **1986**, *64*, 387−392.

(14) Chen, B. H.; Yang, N. S. Characteristics of a Co-current Multistage Bubble Column. *Ind. Eng. Chem. Res.* **1989**, *28*, 1405−1410.

(15) Vinaya, M. *Multistage Bubble Column and Liquid Pulsed Column*. D.Sc. Thesis, Department of Chemical Engineering, Indian Institute of Technology, Madras, India, 1994.

(16) Yamada, H.; Goto, S. Gas and Liquid Holdups in Multi-Stage Bubble Columns for Gas−Liquid−Liquid−Solid Four-Phase System. *J. Chem. Eng. Jpn.* **1998**, *31* (5), 813−817.

(17) Joshi, J. B.; Veera, U. P.; Prasad, C. V.; Phanikumar, D. V.; Deshphande, N. H.; Thakre, S. S.; Thorat, B. N. Gas Holdup Structure in Bubble Column Reactors. *PINSA: Proc. Indian Natl. Sci. Acad.* **1998**, *64* (4), 441−567.

(18) Joshi, J. B.; Sharma, M. M. A Circulation Cell Model for Bubble Columns. *Trans. Inst. Chem. Eng.* **1979**, *57*, 245.

(19) Patil, V. K.; Joshi, J. B.; Sharma, M. Sectionalized Bubble Column: Gas Holdup and Wall Side Solid−Liquid Mass Transfer Coefficient. *Can. J. Chem. Eng.* **1984**, *62*, 228−232.

(20) Karr, A. E. Performance of a Reciprocating-Plate Extraction Column. *AIChE J.* **1952**, *5*, 446−452.

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