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The dual effect of viscosity on bubble column hydrodynamics

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ABSTRACT

Some authors, in the last decades, have observed the dual effect of viscosity on gas holdup and flow regime transition in small-diameter and small-scale bubble columns. This work concerns the experimental investigation of the dual effect of viscosity on gas holdup and flow regime transition as well as bubble size distributions in a large-diameter and large-scale bubble column. The bubble column is 5.3 m in height, has an inner diameter of 0.24 m, and can be operated with gas superficial velocities in the range of 0.004–0.20 m/s. Air was used as the dispersed phase, and various water-monoethylene glycol solutions were employed as the liquid phase. The water-monoethylene glycol solutions that were tested correspond to a viscosity between 0.9 mPa s and 7.97 mPa s, a density between 997.086 kg/m^3 and 1094.801 kg/m^3, a surface tension between 0.0715 N/m and 0.0502 N/m, and $\log_{10}(Mo)$ between −10.77 and −6.55 (where *Mo* is the Morton number). Gas holdup measurements were used to investigate the global fluid dynamics and the flow regime transition between the homogeneous flow regime and the transition flow regime. An image analysis method was used to investigate the bubble size distributions and shapes for different gas superficial velocities, for different solutions of water-monoethylene glycol. In addition, based on the experimental data from the image analysis, a correlation between the bubble equivalent diameter and the bubble aspect ratio was proposed. The dual effect of viscosity, previously verified in smaller bubble columns, was confirmed not only with respect to the gas holdup and flow regime transition, but also for the bubble size distributions. Low viscosities stabilize the homogeneous flow regime and increase the gas holdup, and are characterized by a larger number of small bubbles. Conversely, higher viscosities destabilize the homogeneous flow regime and decrease the gas holdup, and the bubble size distribution moves toward large bubbles. The experimental results suggest that the stabilization/destabilization of the homogeneous regime is related to the changes in the bubble size distributions and a simple approach, based on the lift force, was proposed to explain this relationship. Finally, the experimental results were compared to the dual effect of organic compounds and inorganic compounds: future studies should propose a comprehensive theory to explain all the dual effects observed.

1. Introduction

1.1. Bubble column fluid dynamics

Two-phase bubble columns are equipment used for bringing one or several gases in the form of bubbles (or "*coalescence-induced*" bubbles) into contact with a liquid phase (either pure or a mixture). Bubble columns are typically built in several forms, but the simplest configuration consists of a vertical cylinder with no internals, in which the gas enters at the bottom through a gas sparger that may vary in design (Deckwer, 1992) (i.e., porous sparger, perforated plates, ring or spider spargers, …). Eventually, internal devices may be added to control the heat transfer, to limit the liquid phase back-mixing, or to foster the bubble break-up rate. The liquid phase may be supplied in the batch mode or may be led in either co-currently or counter-currently to the upward gas stream. These contacting devices have found many applications in the chemical production industry and biotechnology thanks to their simplicity of construction, the lack of any mechanically operated parts, low energy input requirements (gas dispersion only and liquid recirculation in some cases), reasonable prices and high performance (i.e., a large contact area between the liquid and gas phase and good mixing within the liquid phase throughout the column) (see refs. (Leonard et al., 2015; Rollbusch et al., 2015)). The heat and mass transfer that occurs in these gas-liquid systems does not necessarily involve reactions between gas and liquid phases, even if this occurs in some cases. For example, oxidation, hydrogenation, chlorination, phosgenation, alkylation and other processes are performed in bubble column reactors (Zehner and Kraume, 2005).

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Nomenclature

Non-dimensional numbers

$Eo=\frac{g(\rho_L-\rho_G)d_{eq}^2}{\sigma}$ Eötvös number
$Fr=\frac{U^2}{gd_{eq}}$ Froude number
$Mo=\frac{g(\rho_L-\rho_G)\mu_L^4}{\rho_L^2\sigma^3}$ Morton number
$Re=\frac{\rho_L U d_{eq}}{\mu_L}$ Reynolds number
$We=\frac{d_{eq}U^2\rho_L}{\sigma}$ Weber number
$N_\mu=\frac{\mu_L}{\mu_G}$ Viscosity ratio

Acronyms

BSD Bubble Size Distribution
CMC Carboxymethyl cellulose
MEG Monoethylene glycol

Symbols

a Major axis of the bubble [m]
b Minor axis of the bubble [m]
c_i (i=1, ..., 5) Coefficient in the ellipse equation (Eq. (15)) [Dimensionless]
c_{MEG} Mass concentration of MEG [%]
c_L Lift coefficient in Eq. (27) [Dimensionless]
d_{23} Sauter mean diameter [mm]
$<d_{23}>$ Gas velocity-average mean diameter [mm]
d_o Gas sparger holes diameter [mm]
d_c Diameter of the column [m]
d_{eq} Bubble equivalent diameter [mm]
D_H Hydraulic diameter [m]
D_H^* Non-dimensional diameter [Dimensionless]
$D_{H,\,Cr}^*$ Critical non-dimensional diameter [Dimensionless]
g Acceleration due to gravity [m/s^2]
h Height along the column [m]
H_c Height of the column [m]
H_D Height of the free-surface after aeration [m]
H_O Height of the free-surface before aeration [m]
J Drift-flux [m/s]
k_i (i=1,...,5) Coefficients in the aspect ratio correlation [Dimensionless]
m Exponent in Eq. (19) [Dimensionless]
n Exponent in Eqs. (13) and (14) [Dimensionless]
N Number of classes used in Eq. (26) [Dimensionless]
p_i (i=1,2,3) Coefficients in the parabola equation (Eq. (18)) [Dimensionless]
S_i (i=1,2,3) Parameters in the swarm velocity method (Eq. (8)) [Dimensionless]
T Temperature [K]
t_G Mean residence time of the dispersed phase [s]
U_b Parameter in the drift-flux method (Eq. (11)) [m/s]
U_∞ Terminal velocity of an isolated bubble (Eqs. (11)-(13)) [m/s]
U Superficial velocity [m/s]
u Mean rise velocity [m/s]
v_b Bubble terminal velocity through the Clift diagram [m/s]
z_i (i=1,...,5) Coefficients in the aspect ratio correlation [Dimensionless]
V Volume [m^3]

Greek symbols

α Proportionality coefficient in Eq. (19) [Dimensionless]
α_G Local gas fraction in Eq. (27) [Dimensionless]
β Exponent in Eq. (19) [Dimensionless]
γ Volume fraction contribution (Eq. (33)) [Dimensionless]
ε Holdup [Dimensionless]
λ Ratio between d_{eq} and d_c [Dimensionless]
ν Bubble terminal velocity [m/s]
ρ Density [kg/m^3]
σ Surface tension [N/m]
ϕ Aspect ratio [Dimensionless]

Subscripts

c *Parameter related to the bubble column*
Slug-bubble *Parameter related to slug-bubbles*
L Liquid phase
G Gas phase
T, E Subscripts in the drift-flux formulation (Eqs. (10)-(12))
trans Transition point
swarm Swarm velocity
wallis Wallis plot method

Despite the simple column arrangement, bubble columns are characterized by extremely complex fluid dynamic interactions and coupling between the phases. For this reason, their correct design, operation and scale-up rely on the knowledge of the fluid dynamics at different scales: mainly, the "*bubble-scale*" (i.e., bubble size distributions and shapes, single bubble dynamics, collective bubble dynamics, ...) and the "*reactors-scale*" (i.e., flow patterns, mean residence time of the disperse phase, dynamics of mesoscale clusters, ...). At the "*bubble-scale*", the bubble motion and bubble dynamics characterize and influence the medium-scale circulation (i.e., eddies that transport the dispersed phase) and the large-scale circulation (i.e., the liquid phase flowing upward in the center of the column and downward in the region near the wall). A typical approach is to apply scale-up methods to estimate the fluid dynamics of industrial-scale reactors from laboratory-scale facilities. Subsequently, models for the interfacial heat and mass transfer and, eventually, to take into account the multi-phase reactions, are applied. The knowledge of the fluid dynamics at the different scales can be quantified through the precise estimation of the local (i.e., the bubble size distributions, BSD, and the bubble aspect ratio; "*bubble-scale*") and the global (i.e., the gas holdup, ε_G—a dimensionless parameter defined as the volume of the gas phase divided by the total volume; "*reactor-scale*") fluid dynamic properties. Indeed, at the "*bubble-scale*", the size and shape of the interface of the dispersed phase characterize the heat and mass exchange; conversely, at the "reactor-scale", the gas holdup determines the residence time and, in combination with the BSD, the interfacial area for the rate of interfacial mass transfer (related to the reactor size). The global and local fluid dynamic properties are strictly related to the prevailing flow regime: the homogeneous flow regime, the transition flow regime and the heterogeneous flow regime (if considering large-diameter bubble columns, see refs. (Nedeltchev, 2015; Nedeltchev and Schubert, 2015; Nedeltchev and Shaikh, 2013)). In the following, these flow regimes are described: (i) first, the general definitions are given; (ii) secondly, the literature concerning the flow regimes is summarized; (ii) finally, the definition of the flow regimes within this research is given.

The homogeneous flow regime is, generally, associated with small gas superficial velocities (U_G) and is characterized by limited interactions between the bubbles. The transition flow regime is characterized by large flow macro-structures with large eddies and a widened bubble size distribution due to the onset of bubble coalescence. The hetero-