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Heat Transfer Between Gas and Liquid in a Bubble Column

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Multiphase contactors, e.g. bubble columns, are often used in operations accompanied by heat transfer between the phases. Literature studies have divided heat transfer measurements in two- and three-phase systems into: (i) estimations of bed-to-wall heat transfer coefficients, and (ii) estimations of immersed object-to-bed heat transfer coefficients. Although wall-to-bed and heat-exchange element-to-bed heat transfer coefficients have been intensively studied and published, no correlations have been found for gas-to-liquid heat transfer for bubble columns.
The aim of this paper is to determine the effect of multi-orifice aerator for the heat transfer between gas and liquid and to compare two different aerator patterns. This study was performed on the bubble column 0.15 m in diameter with various water levels within the range of superficial gas velocity varying from 0.01 m s^{-1} to 0.1 m s^{-1}. The gas-to-liquid heat transfer measurements were performed by a non-steady state method based on measurements of the gas- and liquid- temperature in the time for an evaluation of the heat fluxes and heat transfer coefficients.

1. Introduction

Bubble columns are intensively used as multiphase contactors and reactors in the chemical, biochemical and petrochemical industries. They are used especially in chemical processes involving reactions such as oxidation, chlorination, alkylation, polymerization and hydrogenation, in the manufacture of synthetic fuels by gas conversion processes and in biochemical processes such as fermentation and biological wastewater treatment. Some very well-known chemical applications are in the famous Fischer-Tropsch process, an indirect coal liquefaction process for producing transportation fuels, and also in methanol synthesis and the manufacture of other synthetic fuels which are environmentally much more favourable than petroleum-derived fuels (Kantarci et al., 2005, Shah et al., 1982, Prakas et al., 2001).
The intensive heat transfer rate is one of the most important characteristics in the operation of bubble columns. The heat transfer rate in gas–liquid bubble columns is reported to be generally 100 times greater than in single phase flow (Deckwer, 1980). This rate is influenced by a number of physical parameters and operating conditions: (i) gas-holdup, (ii) superficial gas velocity, (iii) circulation velocity, and (iv) the physical properties of the liquid. All these factors are highly interactive, and control the performance of the bubble column. Thermal control is most important in absorption columns with chemical reactions accompanied by a heat supply operation (endothermic) or a heat removal operation (exothermic). For example, Fazeli et al. (2008) presents that the design of heat transfer equipment for slurry bubble column reactors a limiting factor in sizing and scale up. The proper design of a heat exchange system depends on knowledge of gas-to-liquid heat transfer. It is usually assumed that the gas output temperature is the same as the liquid output temperature. On the basis of our industrial practice, we can conclude that this assumption is not correct.
Literature studies have divided heat transfer measurements in two- and three-phase systems into: (i) estimations of bed-to-wall heat transfer coefficients, and (ii) estimations of immersed object-to-bed heat transfer coefficients (Deckwer, 1980).
Bed-to-wall heat transfer coefficients were investigated by e.g. Kast (1962), Fair et al. (1962), Nishikawa et al. (1977), Steiff and Weinspach (1978), Hikita et al. (1981), Kato et al. (1981), Lewis et al. (1982). Immersed object-to-bed heat transfer coefficients were investigated by e.g. Fair et al. (1962), Konsetov (1966), Burkel (1972), Nishikawa et al. (1977), Steiff and Weinspach (1978), Saxena and Vadivel (1988), Verma (1989),

Avdeev et al. (1992), Saxena et al. (1992), Saxena and Chen (1992), Avdeev and Halme (1993), Fazeli et al (2008). Heat transfer from gas-liquid bed to single tubes was investigated by Avdeev et al. (1992) and Burkel (1972). Heat transfer from bed to vertical and horizontal tube bundles was investigated e.g. by Saxena and Vadivel (1988).
The enthalpy balances in the bubble column require knowledge of the heat transfer coefficients from wall-to-bed, heat-exchange element-to-bed and gas-to-liquid. Although wall-to-bed and heat-exchange element-to-bed heat transfer coefficients have been intensively studied and published, no correlations have been found for gas-to-liquid heat transfer for bubble columns.

2. Theoretical background

The proposed procedure for experimentally determining the heat transfer between gas and liquid in the column is based on transient heat transfer measurements. The principle of the proposed procedure is as follows: The column is filled with a given volume of liquid. A gas at a lower temperature than the liquid is bubbled through the liquid at a constant gas flow rate. Due to the different temperatures of the liquid and the gas, there is heat transfer from the liquid to the gas. The gas is heated and the liquid is cooled. The inlet gas temperature is practically unchanged and remains constant. The temperature of the liquid decreases with time. The exhaust gas temperature changes over time. Thus the process is non-stationary. If the saturated pressure corresponding to the liquid temperature is higher than the partial pressure of the liquid vapours in the gas stream, there is evaporation of the liquid into the gas stream. In addition, if the temperature of the system is higher than the temperature of the surroundings, there are heat losses to the surroundings. The enthalpy balance diagram of the system is schematically depicted in Figure 1.

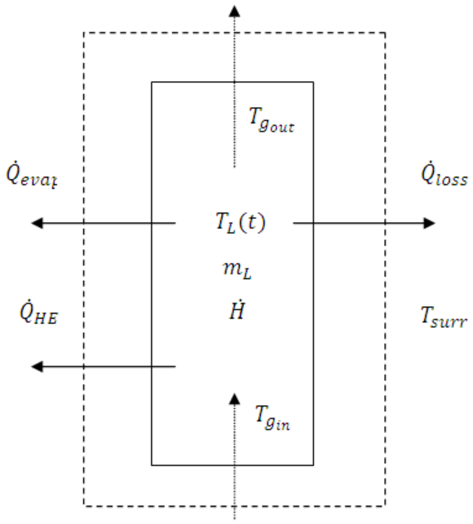

Figure 1: Enthalpy balance diagram of the system.

Assuming ideal mixing of the liquid batch and neglecting heat losses to surroundings the enthalpy balance can be written in the case of a liquid being cooled by a gas, i.e. $T_L > T_{gi}$, as follows:

$$m_L \cdot c_{pL} \cdot \frac{dT_L}{dt} = -\dot{m}_{vap} \cdot \Delta h^{vap} - \alpha \cdot S_b \cdot (T_L - T_g) \tag{1}$$

where dT_L/dt is a derivative of liquid temperature T_L with respect to time t (K s^{-1}), m_L is the weight of the liquid batch (kg), c_{pL} is the specific heat capacity of the liquid (J kg^{-1} K^{-1}), α is the gas-to-liquid heat transfer coefficient (W m^{-2} K^{-1}), S_b is the gas bubble interfacial area (m^2), T_L is the liquid temperature (K), T_g is the gas temperature (K), $m^{\bullet}_{evap}$ is the mass flow of the evaporated liquid (kg s^{-1}) and Δh^{evap} is the heat of vaporization (J kg^{-1}).
In our case, the experimental data were obtained in discrete form. It was found that directly solving the differential equation (1) was not satisfactory in this case, and that an integral formula should therefore be used. The time dependences of $m^{\bullet}_{evap}$ and T_g were expressed indirectly as a polynomial function of instantaneous liquid temperature T_L (t). The following polynomials of the 2nd order were used for substitution:

$$T_g = a_2 \cdot T_L^2 + a_1 \cdot T_L + a_0 \tag{2}$$

$$\dot{m}_{vap} \cdot \Delta h^{vap} = b_2 \cdot T_L^2 + b_1 \cdot T_L + b_0 \tag{3}$$

This trick enables us to integrate Eq. (1). Combining Equations (1), (2) and (3) and integrating the following final formula can be obtained for $\Delta > 0$:

$$\left[\frac{2}{\sqrt{\Delta}} arctg\left(\frac{2 \cdot C_2 \cdot T_L + C_1}{\sqrt{\Delta}}\right)\right]_{TL0}^{TL} = \frac{t}{m_L \cdot c_{pL}} \tag{4}$$

where

$$C_2 = \alpha \cdot S_b \cdot a_2 - b_2 \tag{5a}$$

$$C_1 = \alpha \cdot S_b \cdot a_1 - \alpha \cdot S_b - b_1 \tag{5b}$$

$$C_0 = \alpha \cdot S_b \cdot a_0 - b_0 \tag{5c}$$

$$\Delta = 4 \cdot C_2 \cdot C_0 - C_1^2 \tag{5d}$$

where T_{L0} is the initial temperature of the liquid batch (K) and T_L is the temperature of the liquid batch (K) at time t (s). The Equation (4) can be rewritten in discrete form. Then, applying least-square method, the product $\alpha.S_b$ is obtained by iterative procedure. The Δ value was found to be positive for all runs. The changing gas temperature profile along the height of the liquid was respected by log-mean temperature difference between a liquid batch and a passing gas. The detailed description of the method is preparing for publication.

The gas-to-liquid mass transfer has often been described by volumetric mass transfer coefficient $k_L.a$ due difficulties in determining the bubble interfacial area S_b. By analogy, the gas-to-liquid heat transfer can be characterized by means of volumetric gas-to-liquid heat transfer coefficient $\alpha.a$ (W m^{-3} K^{-1}), which is defined as follows:

$$\alpha \cdot a = \alpha \cdot S_b / V_{L0} \tag{6}$$

where V_{L0} is the equipment/bed volume that corresponds to the volume of a non-gassed liquid batch (m^3).

3. Experimental

Experiments were performed on a PVC tube vertical column 0.15 m in inner diameter in air-water system in the range of superficial gas velocity from 0.01 m s^{-1} to 0.1 m s^{-1}. The liquid (water) height was held at four different levels - 0.5 m, 0.65 m, 0.85 and 0.95 m. The two patterns of gas multi-orifice distributor were tested (Figure 2). The diameter of holes was 1 mm. The sparger was located at the bottom of a bubble column.

The column was filled with warm water ($\propto$ 40 °C). The batch was started bubbled by an air flow having lower temperature ($\propto$ 25 °C). Thus, air passing through liquid batch was warmed up and the liquid was cooled, i.e. liquid temperature decreased with time. The outlet gas temperature changed with time, depending on heat transfer. The heat transfer was also accompanied by simultaneous evaporation of the liquid that was taken into account when heat-transfer coefficient was evaluated. The following experimental data sets were obtained in a discrete form: 1) the liquid batch temperature T_L (t_i), 2) the inlet air temperature $T_{g\ in}$ (t_i), 3) the outlet air temperature at column outlet $T_{g\ out}$ (t_i), 4) the inlet air pressure $p_{g\ in}$ (t_i), 5) the relative humidity of inlet air $\varphi_{g\ in}$ (t_i), and 6) the relative humidity of outlet air $\varphi_{g\ out}$ (t_i). Time of the measurement was about 20 minutes. The inlet air temperature was found to be practically unchanged and constant. As it was assumed the heat losses to the surroundings were found to be negligible.

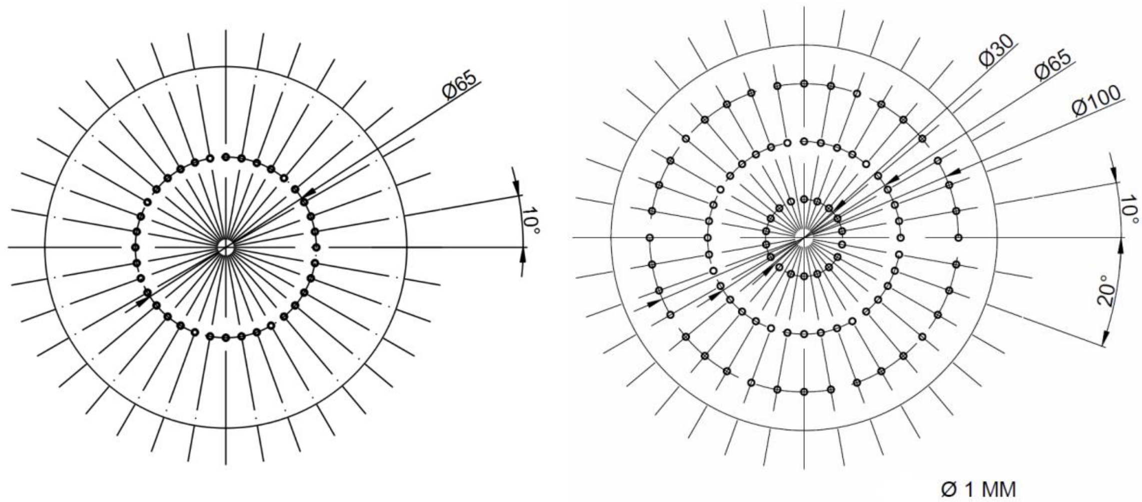

Figure 2: Tested multi-orifice aerator patterns: a) SC (left) and b) CC3 (right).

4. Results and discussion

4.1 Hold up

The hold-up ε (-) was determined visually and one was obtained as follows:

$$\varepsilon = (H_b - H_{L0}) / H_{L0} \tag{7}$$

where H_b is the height of the bubbled liquid batch (m), and H_{L0} is the height of non-gassed liquid batch (m). The dependence of hold up on superficial velocity is depicted for four liquid levels in Figures 3a and 4a for SC and CC3 pattern respectively. This dependence can be expressed by the linear relation regardless of liquid height as follows:

for SC pattern $\quad \varepsilon = 2.005 \cdot u_{g0} + 0.0027 \quad$ (R = 0.973) (8)

for CC3 pattern $\quad \varepsilon = 1.992 \cdot u_{g0} + 0.0027 \quad$ (R = 0.942) (9)

where u_{g0} is the superficial gas velocity (m s^{-1}). The hold up was found to be practically the same regardless of distributor pattern.

4.2 Volumetric gas-to-liquid heat transfer coefficient

It was found that the effect of the liquid height on the volumetric gas-to-liquid heat transfer coefficient α.a can be compensated using aeration VVM as a measure of air flow. The dependence of volumetric heat transfer coefficient α.a on aeration VVM is depicted for four liquid levels in Figures 3b and 4b for SC and CC3 pattern respectively. This dependence can be expressed by the linear relation as follows:

for SC pattern $\quad \alpha \cdot a = 6985 \cdot VVM + 371 \quad$ (R = 0.914) (10)

for CC3 pattern $\quad \alpha \cdot a = 8858 \cdot VVM + 308 \quad$ (R = 0.942) (11)

where α.a is the volumetric gas-to-liquid heat transfer coefficient (W m^{-3} K^{-1}), VVM is the aeration (s^{-1}) defined as a ratio of air flowrate and the volume of a non-gassed liquid batch, i.e. VVM = $\dot{V}_g/V_{L0} = u_{g0}/H_{L0}$. For CC3 pattern the two data items were excluded from regression (one data item for 0.6 m and one data item for 0.95 m for the aeration 0.015 s^{-1} and 0.012 s^{-1} respectively). As it follows from the obtained dependences, for lower values of the aeration up to 0.034 s^{-1} the heat transfer coefficient for SC pattern is a higher comparing with CC3 pattern. Unlike this, the CC3 pattern is more efficient for heat transfer for higher values of the aeration greater then 0.034 s^{-1}. In this case the heat transfer coefficient for CC3 pattern is higher comparing with SC pattern.

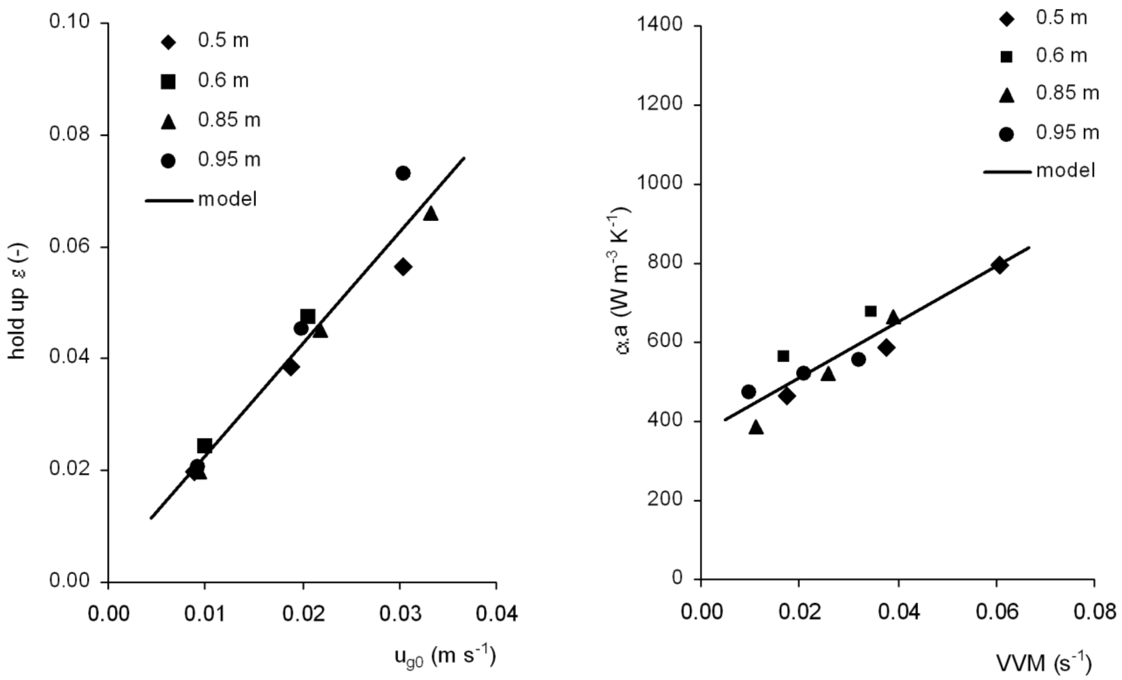

Figure 3: SC pattern: a) the hold up ε (left), b) the volumetric gas-to-liquid heat transfer coefficient α.a (right).

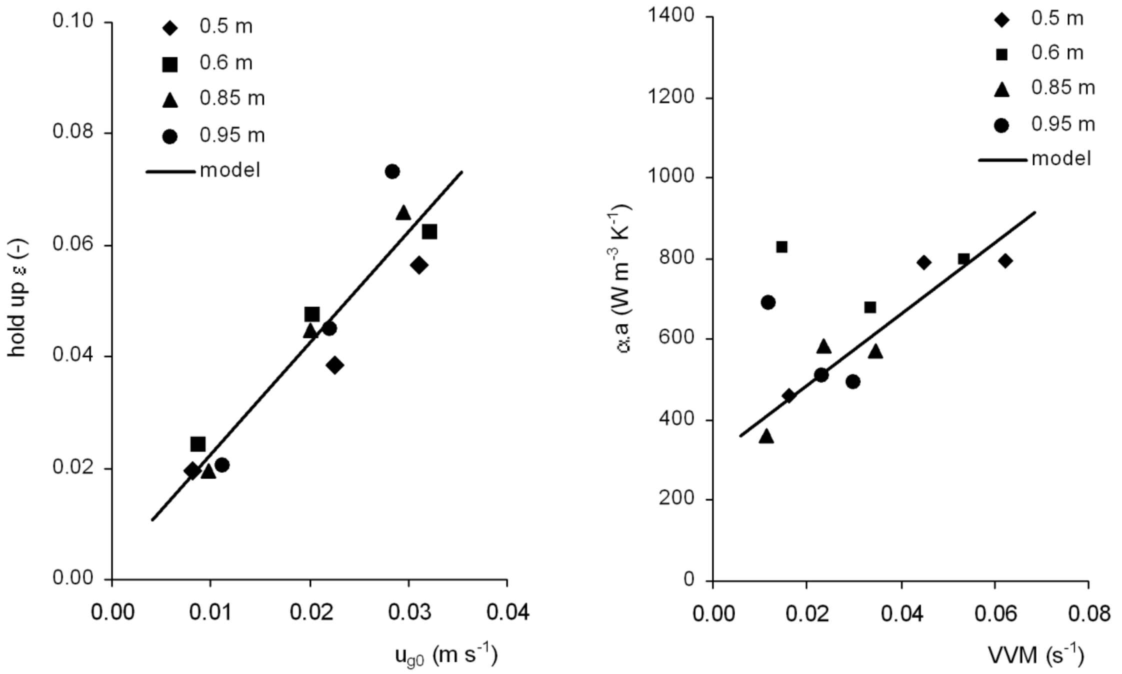

Figure 4: CC3 pattern: a) the hold up ε (left), b) the volumetric gas-to-liquid heat transfer coefficient α.a (right).

5. Conclusions

The effect of two different aerator patterns on the hold up and the heat transfer between gas and liquid was investigated for air-water system in the bubble column in diameter of 0.15 m with various water levels within the range of superficial gas velocity varying from 0.01 m s^{-1} to 0.1 m s^{-1}. The hold up was found to be linearly depending on the superficial gas velocity regardless of liquid height. The volumetric gas-to-liquid heat transfer coefficient was found to be linearly depending on the aeration regardless of liquid height. For lower values of the aeration up to 0.034 s^{-1} the SC patter was to be more efficient for heat transfer. For higher values of the aeration greater then 0.034 s^{-1} the CC3 pattern was found to be more efficient for heat transfer.

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