International Journal of Thermal Sciences 102 (2016) 145–153

Contents lists available at ScienceDirect

International Journal of Thermal Sciences

journal homepage: www.elsevier.com/locate/ijts

Heat and mass transfer in three phase fluidized bed containing high density particles at high gas velocities

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ARTICLE INFO

Article history:
Received 29 March 2015
Received in revised form
8 November 2015
Accepted 18 November 2015
Available online xxx

Keywords:
Bubble column
Fluidized bed
Heat transfer
Mass transfer
Multiphase reactors

ABSTRACT

In a batch gas–liquid contactor, gas holdup in two phase (liquid–gas) and three phase (gas–liquid–solid) fluidized bed has been measured experimentally; bubble size and the corresponding interfacial area of contact between the gas phase and liquid phase were calculated. Mass and heat transfer coefficients were evaluated at different solid contents and variable gas superficial velocities for the cooling and humidification of air bubbles into a water-sand slurry under fluidization conditions. Simplified models were derived to estimate the transfer coefficients.

Within the present range of parameters studied, the presence of solid particles decreased the gas holdup as well as the interfacial area of contact between gas and liquid, while both the heat and mass transfer coefficients increased several times compared to the two phase (liquid–gas) values by a factor ranging from 2 to 5 depending on the solid contents and the gas superficial velocity. Applications of the present results in the design and operation of multiphase reactors were highlighted.

1. Introduction

Gas–liquid–solid fluidized bed systems have been developed since a long time and are used extensively in several applications in the chemical, biochemical, electrochemical, mineral processing, and petrochemical industries because of the high rates of heat and mass transfer of such types of reactors [1,2]. Three-phase fluidized beds normally refer to gas–liquid–solid systems with large particles (particle size in the magnitude of a mm) while bubble columns commonly refer to gas–liquid systems without the presence of particles. Processes in three phase bed (gas–liquid–solid) find increasing applications since the development of fluidized bed catalytic process for the hydrogenation of residual oils, gas purification, extraction, wastewater treatment and crystallization, etc. because three phase processes are distinguished by their high rate and intimate contact between involved phases [2–10]. In a three-phase system, as well as in two phase system transfer operations between phases are affected by the hydrodynamic characteristics of the system, an appropriate design and scale up of gas–liquid–solid system requires the estimation of various transport parameters (heat, mass and momentum), holdups and distribution of various phases [1]. The importance of these parameters depends upon the nature of the reaction and the conditions of the flow of the phases involved. Previous studies on three phase processes such as Fischer–Tropsch reaction and hydrogenation in slurry operations concluded that gas–liquid mass transfer in a three phase process is an important phenomenon and transfer of mass across the gas–liquid interface can be of rate controlling importance [11,12]. Using suspended catalyst and assuming plug flow and complete mixing in the suspension phase Satterfield and Huff [11] concluded that the Fischer–Tropsch reaction becomes controlled by the gas–liquid mass transfer at high synthesis temperatures. In multiphase reaction system a temperature rise increases the intrinsic reaction rate, thus mass transfer resistances may become more relevant. Contrary to the study of Satterfield and Huff [11] Deckwer et al. [13] analyzed Fischer–Tropsch process in slurry phase with eight different catalysts and reactor geometries, the study showed that the process is controlled mainly by the reaction step (resistance). The authors came to the conclusions that at high gas velocity and low catalyst activity the mass transfer resistances

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http://dx.doi.org/10.1016/j.ijthermalsci.2015.11.020
1290-0729/

(limitations) seem to be less important or negligible while if a low gas velocity and a high catalyst activity were applied mass transfer limitations become more significant. The importance of the mass transfer resistance in Fischer–Tropsch process depends on different factors such as type and amount of catalyst, liquid viscosity, gas velocity, interfacial area, reaction temperature, etc. [13].

Gas–liquid–solid system is characterized by a complicated flow behavior; therefore a better understanding of the behavior of heat and mass transfer in this type is necessary. Previous studies on heat transfer in gas–liquid–solid system focused on wall-to-bed [14,15], and on the immersed heater to bed heat transfer [16,17]. The range of gas superficial velocities studied was ranged from 0.02 to 9 m/s and the solid contents were below 20 vol.% [18–23]. While there is no reported data on gas–liquid heat and mass transfer coefficients at a high range of solid contents and high ranges of gas superficial velocities.

The present work aims to study the effect of solid contents on the gas–liquid heat and mass transfer coefficients in a three phase system at values of solid contents and gas superficial velocities higher than that reported in the literature. Also the study aims to investigate to what extent the gas velocity will affect the heat and mass transfer at high solid contents. The study was carried out on the steady state cooling and humidification of hot air. Water and sand were used as the two other phases in the cooling system. Gas holdup was measured at both a wide range of gas velocity (12–29 cm/s) and a wide range of solid content (10–30%). The range of the parameters studied is wider than that reported in the literature [7]. From the measured gas holdup values, the average bubble diameter, the apparent total interfacial area between gas and liquid, and the mass and heat transfer coefficients were determined.

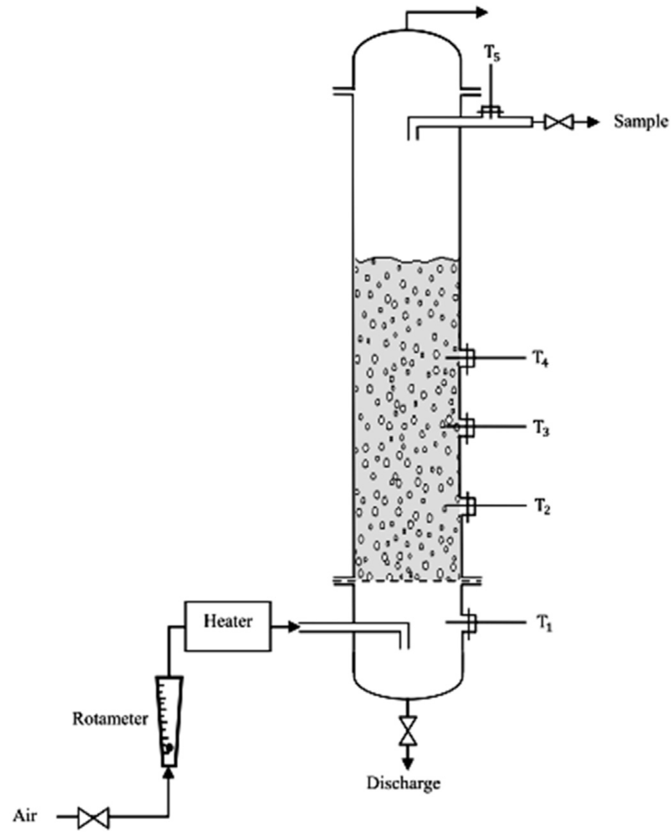

Fig. 1. Apparatus.

2. Experimental section

Experiments were carried out in a column having a single gas distributor. Fig. 1 shows a schematic diagram of the apparatus used in the present work. It consists of a vertical cylindrical column, gas compressor, electrical heater, gas distributor and temperature measuring elements. The column was a thick walled glass cylinder and has an inner diameter of 7 cm and a height of 60 cm. The outer wall of the column was well insulated to prevent heat loss to the surroundings. The gas distributor of the column was a horizontal circular perforated disc with a pore diameter of 1 mm. The height of the expanded bed was measured by a side tube manometer. Digital sensitive thermometers were used to measure the temperature at the inlet and outlet of air as well as in the bulk of the slurry. The heating of the inlet air was carried out in a thermostatically controlled tubular electric heater. Completely washed sand particles of 0.288 mm average diameter and specific gravity 2.6 were used as the solid phase. The height of the static bed in all experiments was kept constant at 27 cm. The gas outlet temperature was measured by an insulated suction probe filled at its inlet with glass wool packing to separate liquid entrainments before temperature measurement. The superficial velocity of air was measured by a Rotameter and ranged from 12 to 29 cm/s. The inlet temperature of the air was kept at 60 °C. The experiments in the present study were conducted at the atmospheric pressure. Kolbel et al. [24] studied the effect of pressure on the bubble size for an air–water system, up to pressures of 1.6 MPa the authors found no effect on the gas holdup and mean bubble diameter if the gas velocity is corrected to take into account the actual pressure of the column. Letzel et al. [25] also defined a ratio of mass transfer coefficients to gas holdup that seems to be constant at a value of 0.5 up to a pressure of 1.0 MPa. Other authors concluded that increasing pressure promotes the formation of smaller bubbles and the ratio of mass transfer coefficients to gas holdup increases with increasing pressure [26].

2.1. Procedure

Air is introduced into the column at the start, followed by the weighted amount of sand and enough hot water is fed into the column from its top until the level reach to the required value. The system was left to cool down until it attains the steady state and the temperature becomes constant for a long time. At this moment the average bulk temperature was recorded as well as the inlet and outlet gas temperature. The gas holdup was measured and the surface area was calculated.

2.2. Theory

To derive the models required to estimate the gas–liquid heat and mass transfer coefficients in the three phase fluidized bed these assumptions are used; (i) the intraparticle temperature gradient is neglected, (ii) a well mixing in the three phase system is achieved by the presence of solids, (iii) the liquid side resistance to heat transfer is neglected. Under these conditions, calculations can be performed for gas heating or cooling without resort to interface transfer coefficient. As in two phase fluidized bed system heat transfer coefficient (h) based on the total surface area and on an overall effective driving force can be applied to the three phase system by means of the conventional rate equation:

$$q = hA\Delta T \tag{1}$$

where: $\Delta T = T_g - T_b$