EFFECT OF GAS VELOCITY AND BED MASS ON LIQUID MINIMUM FLUIDIZATION VELOCITY IN A THREE-PHASE FLUIDIZED BED WITH CYLINDRICAL PARTICLES

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ABSTRACT

Three-phase fluidized beds have been applied successfully in the bio-oxidation process for wastewater treatment in which various low to moderate density solid particles of different shape and size are used as cell support. Ceramic raschig ring possess moderate density and high surface area due to its hollow cylindrical structure. Thus can be used as solid support for microorganisms, where higher mass transfer rate can be achieved. In this study the hydrodynamic characteristics viz. the pressure drop, bed expansion and liquid minimum fluidization velocity of a co-current gas-liquid-solid three-phase fluidized bed have been studied using liquid as the continuous phase and gas as the discontinuous phase. Bed pressure measurements have been made to predict the minimum liquid fluidization velocity and the effect of gas velocity and bed mass on it has been studied.

INTRODUCTION

Gas-liquid-solid fluidization also known as three-phase fluidization is a subject of fundamental research since the last three decades due to its industrial importance. Three-phase fluidized beds have been applied successfully to many industrial processes such as in the H-oil process for hydrogenation and hydro-desulfurization of residual oil, the H-coal process for coal liquefaction, Fischer-Tropsch process, and the bio-oxidation process for wastewater treatment. Three-phase fluidized beds are also often used in physical operations (Muroyama and Fan, 1985). Gas-liquid-solid fluidization is classified mainly into four modes of operation. These modes are co-current three-phase fluidization with liquid as the continuous phase (mode I-a); co-current three-phase fluidization with gas as the continuous phase (mode-I-b); inverse three-phase fluidization (mode II-a); and fluidization represented by a turbulent contact absorber (TCA) (mode II-b). Modes II-a, and II-b are achieved with a countercurrent flow of gas and liquid. The most striking one from the above contacting modesis the co-current three-phase fluidization with the liquid as the continuous phase (Muroyama and Fan, 1985). The co-current gas-liquid-solid fluidization is defined as an operation in which a bed of solid particles is suspended in gas and/or liquid upward flowing media due to the net gravitational force on particles. Such an operation generates considerable intimate contact among the gas, liquid and solid particles in these systems and provides substantial advantages for applications in physical, Chemical or biochemical processing involving gas, liquid and solid phases (Dhanuka and Stepanek, 1978).

The successful design and operation of a gas-liquid-solid fluidized bed system depends on the ability to accurately predict the fundamental characteristics of the system. Viz. the hydrodynamics, the mixing of individual phases, and the heat and mass transfer characteristics (Begovich and Watson, 1978). Knowledge of minimum liquid fluidization velocity is essential for the successful operation of gas-liquid-solid fluidized beds. For gas-liquid-solid fluidized systems the minimum liquid fluidization velocity is the superficial liquid velocity at which the bed becomes fluidized at a given gas superficial velocity (Briens and Briens, 1997).The minimum liquid flow rates required to achieve fluidization are determined by a plot of the pressure drop across the bed vs. the superficial liquid velocity at constant gas flow rate. When fluidized, the pressure drop across the bed will no longer change with increasing liquid flow rate. Thus the flow rates at which a break in curve occurs correspond to the MF velocities (Begovich and Watson, 1978). Visual observations determine the liquid minimum fluidization velocity as either the velocity at which the bed first begins to expand or as the velocity at which any particle with in the bed continuously shifts position with neighboring particles (Briens and Briens, 1997). The bed pressure drop (ΔP) is given by the equation $\Delta P = gH\ (\rho_l \varepsilon_l + \rho_g \varepsilon_g + \rho_s \varepsilon_s)$. Three-phase fluidized beds have been applied successfully in the bio-oxidation process for wastewater treatment in which various low-moderate density solid particles of different shape and size are used as cell support. Ceramic raschig ring possess moderate density and high surface area due to its hollow cylindrical structure and hence can be used as solid support for microorganisms, where higher mass transfer rate can be achieved.

In the present study experiments were conducted to examine the effect of gas velocity and bed mass (initial static bed height) on minimum liquid fluidization velocity of a co-current gas-liquid-solid three-phase fluidized bed using liquid as the continuous phase and gas as the discontinuous phase. Ceramic raschig rings of size 6.6mm are used as solid particles. Pressure drop measurement was used to determine the minimum liquid fluidization velocity.

EXPERIMENTAL

A schematic diagram of the experimental setup is shown in Figure-1. The vertical Plexiglas fluidizer column is of 100 mm ID with a maximum height of 2m.The column consists of three sections, v.i.z., the gas-liquid disengagement section, test section, and gas-liquid distributor section. The gas-liquid distributor is located at the bottom of the test section and is designed in such a manner that uniform distribution of the liquid and gas can be maintained in the column. The distributor section is a conical frustum of 30 cm in height, one end 5.08 cm in diameter and the other end of 10 cm in diameter having liquid inlet at the bottom low cross-section end. The conical section is mounted with a perforated plate made of G.I. sheet of I mm thick, 120 mm diameter, of about 278 numbers of 2, 2.5 and 3mm holes placed at the top of this section. There is a gas distributor is a tube network sparger consists of 50 numbers of 1mm pores placed randomly placed in side the conical section. In this section the gas and liquid streams merged and passed through the perforated grid. The mixing section and grid ensure that the gas and liquid are well mixed and evenly distributed into the bed. Gas-Liquid disengagement section is at the top of the column, which allows gas to escape and liquid to be circulated. Any entrained particles if any are retained on the screen attached to the top of this section. For pressure drop measurement the pressure ports are fitted to the U-tube manometers of 1m & 0.5m long filled with carbon tetrachloride. The scope of the experiment is shown in Table-1. The air-water flow are co-current and upwards. Accurately weighed amount of material is fed into the column and adjusted for a specified initial static bed height. Four initial static bed heights 15.4 cm, 21.4 cm, 26.4 cm and 31.4 cm are used in this experiment. Water is pumped to the fluidizer at a desired flow rate. Then air is injected into the column through the air distributor. Approximately five minutes is allowed to make sure that the steady state has been reached. Then the readings of each manometer are taken.

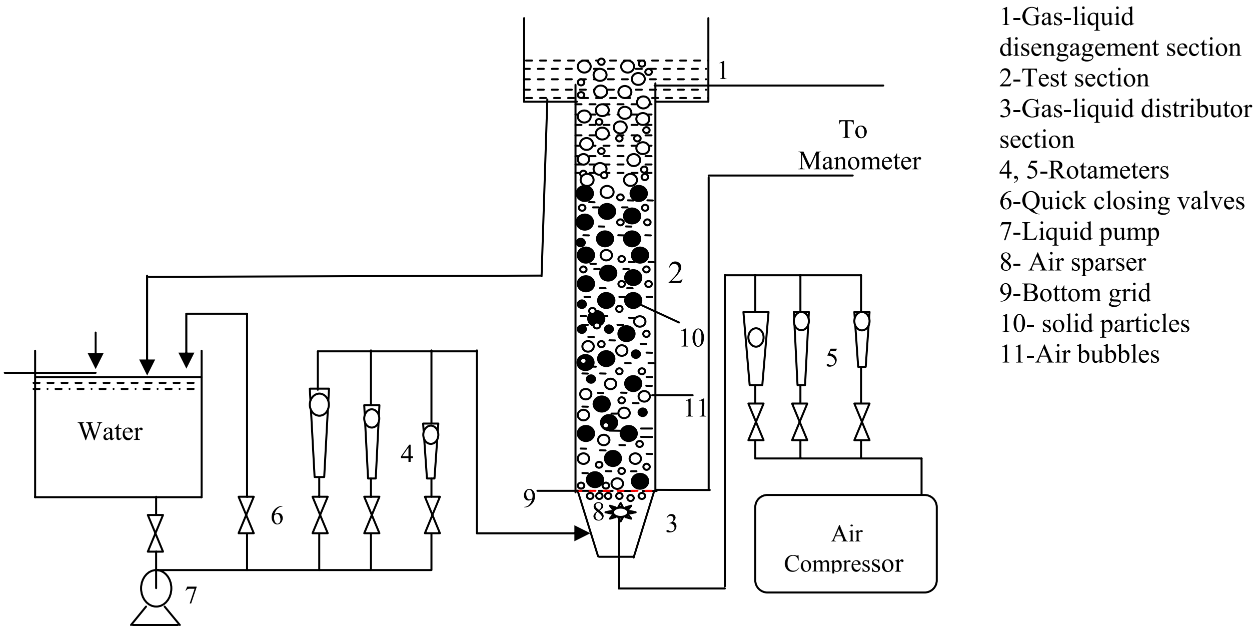

Figure 1: Schematic diagram of gas-liquid-solid three-phase fluidized bed

Table 1: Properties of Bed Materials (A), Fluidizing Medium (B), Manometric Fluid (C)

A. Properties of Bed Materials

Materials	Particle Size, L = OD, mm	ID, mm	ρ_p (kg.m^{-3})
Ceramic raschig ring	6.6	3.3	1670

B.Properties of Fluidizing Medium

Fluidizing Medium	ρ (kg.m^{-3})	μ (Ns/m^2)
Air at 30^{0}C	1.168	0.00187
Water at 30^{0}C	999.4	0.095

C.Proprties of Manometric Fluid

Manometric Fluid	ρ (kg.m^{-3})	μ (Ns/m^2)
Carbon Tetra-Chloride (CCl_4)	1600	0.09

RESULTS AND DISCUSSION

The minimum liquid fluidization velocity (U_{lmf}) in this study has been obtained from the relationship between pressure drop and superficial liquid velocity. Figure-2 shows the variation of pressure drop with superficial liquid velocity for gas-liquid-solid system at various superficial gas velocities. From this it is observed that the minimum fluidization velocity decreases with increase in gas velocity. The variation of minimum liquid fluidization velocity with superficial gas velocity is shown in Figure-3. It indicates sharp decrease in minimum liquid fluidization velocity with gas velocity initially but then the rate of decrease is slow. This indicates the bubble supported fluidization in presence of gas. The values of Ulmf are listed in Table-2a. Figure-4 shows the variation of bed pressure drop with superficial gas velocity for different static bed height. From this pressure drop plot minimum liquid fluidization velocity has been calculated. Observed pressure drop increases with the initial static bed height which indicates higher energy requirement to fluidize higher bed mass. The observed minimum liquid fluidization velocities are plotted against the initial static bed height and are shown in Figure-5. The plot shows that the minimum liquid fluidization velocities are approximately same for all static bed heights (the values are listed in Table-2b).

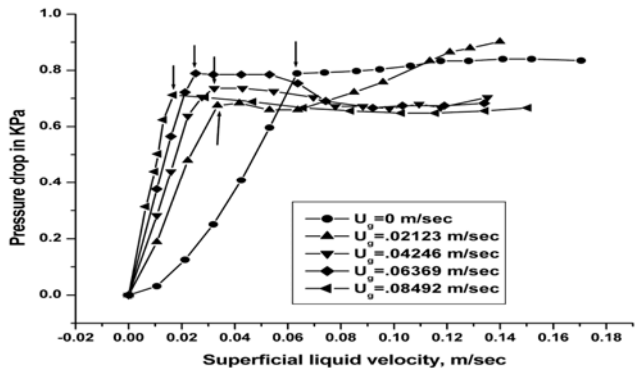

Figure 2: Variation of bed pressure drop with superficial liquid velocity for different superficial gas velocities at static bed height 21.4 cm.

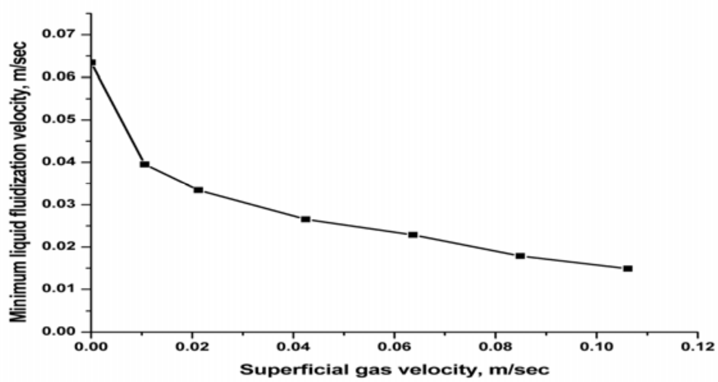

Figure 3: Variation of minimum liquid fluidization velocity with superficial gas velocity at constant static bed height 21.4 cm.

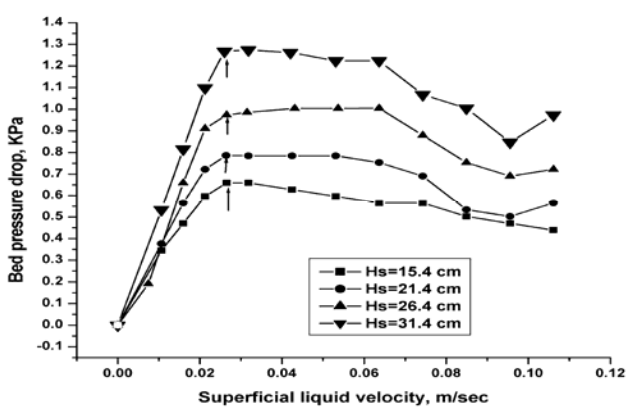

Figure 4: Variation of bed pressure drop with superficial liquid velocity for different static bed height at superficial gas velocity 0.06369 m/sec.

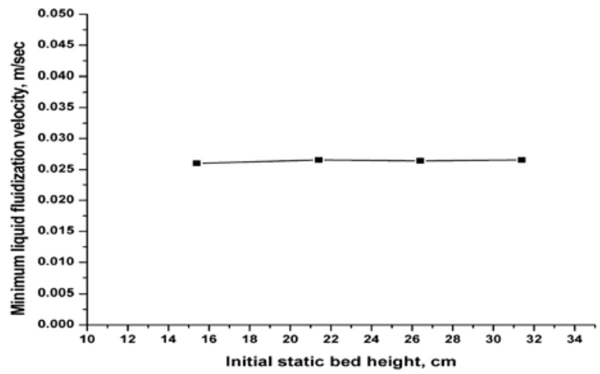

Figure 5: Variation of minimum liquid fluidization velocity with initial static bed height at superficial gas velocity 0.06369 m/sec.

Table 2a: U_{lmf} in m/sec for different gas velocities, m/sec for H_s=21.4 cm.

$U_g = 0$	$U_g = 0.02123$	$U_g = 0.04246$	$U_g = 0.06369$	$U_g = 0.08492$	$U_g = 0.1062$
0.06344	0.03348	0.030247	0.02527	0.01687	0.01485

Table 2b: U_{lmf} in m/sec for different H_s at U_g=0.06369 m/sec.

Hs = 15.4 cm	Hs = 21.4 cm	Hs = 26.4 cm	Hs = 31.4 cm
0.026539	0.02527	0.026539	0.02599

The Correlation for minimum liquid fluidization velocity developed by the authors is either in the form of dimensionless groups or in the form dimensional terms. Here the correlation for minimum liquid fluidization velocity has been developed in dimensional form. The parameters of importance here are gas velocity, particle size, column diameter, static bed height, and density of solid, liquid and gas. In this experiment except in the variation of gas velocity and static bed height, other parameters have been kept constant. As from the above results it is observed that bed mass (static bed height) has no effect on minimum liquid fluidization velocity, only the relation between U_{lmf} and superficial gas velocity U_g will therefore be in the form $U_{lmf} = k\ (U_g)^n$. This was done using the power law correlation option in Microsoft Excel. The correlation obtained as $U_{lmf} = 0.0067(U_g)^{-0.4079}$. By using this correlation the minimum liquid fluidization velocities have been calculated for different gas velocities and have been compared with their respective experimental values (Figure-6). The values calculated from correlation found to agree well with the experimental values with mean and standard deviation of 9.04 and 9.269 respectively.

CONCLUSIONS

The study of the three-phase fluidized bed with cylindrical particles reveals that the minimum liquid fluidization velocity is a strong function of gas velocity and decreases with increase in gas velocity, but not affected by the bed mass (initial static bed height). The developed correlation can be used for calculating the minimum liquid fluidization velocity for hallow cylindrical particles with in the range of study.

NOMENCLATURE

H_s Static bed height, [cm]

U_l Liquid velocity, [m/sec]
U_g Gas velocity, [m/sec]
U_{lmf} Minimum liquid velocity for a three-phase system, [m/sec]
ρ Phase density,[kgm^{-3}]
s Solid phase
l Liquid phase
g Gas phase

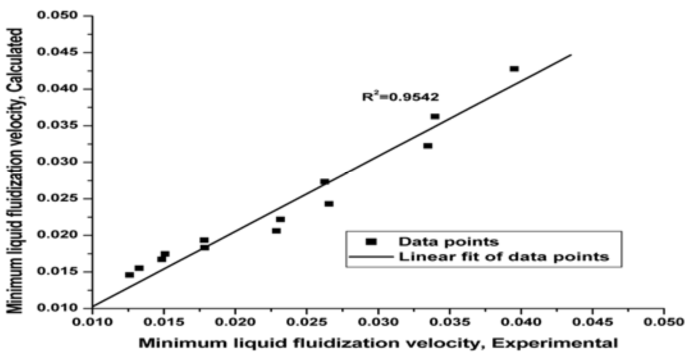

Figure 6: Comparison of experimental and calculated values of U_{lmf}.

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