Hydrodynamics of a Gas-Solid Fluidized Bed at Elevated Temperatures Using the Radioactive Particle Tracking Technique

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ABSTRACT: *Effect of temperature on hydrodynamics of bubbling gas-solid fluidized beds was investigated. Experiments were carried out in the range of 25-600 ºC and different superficial gas velocities in the range of 0.17-0.78 m/s with sand particles. Time-position trajectory of particles was obtained by radioactive particle tracking technique. These data were used for determination of mean velocities of upward-moving (including bubble wake and ascending clusters) and downward-moving (descending clusters) particles. It was found that the upward velocity increases by increasing temperature up to 300 ºC, however, it decreases by further increase in temperature. Due to the wall effect, there is no significant change in the mean velocity of downward-moving clusters. The change in hydrodynamic parameters with temperature can be a consequence of changing physical properties of the bed which have been represented by Reynolds number in this study.*

KEY WORDS: *Fluidized bed, High temperature, Hydrodynamics, Radioactive particle tracking***.**

INTRODUCTION

The widespread application of fluidized beds in chemical industries and the demand for improvements in fluidization efficiency have increased the need for a better understanding of the fluidization phenomena.

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Most of the researches on fluidization have been carried out at the ambient temperature. However, industrial fluidized bed reactors operate at high temperatures. Thus, there is a lack of fundamental studies at high temperatures due to difficulties associated with measuring techniques. Few experimental studies reported in literature have mostly focused on the effect of temperature on minimum fluidization velocity and voidage [1-5]. *Geldart & Kapoor* [6] studied the effect of temperature on the minimum bubbling velocity and bubble diameter for Group-A particles. They found that the bubble size and minimum bubbling velocity are reduced slightly with increasing the temperature from ambient to 200 ºC. No significant changes, however, was reported at 300 ºC. *Kia & Furusoki* [7] found the same trend for the bubble size in the fluidization of the FCC and alumina particles in a temperature range of 280-400 K. Unlike group-A particles, *Hatate et al.* [8] reported a rise in the bubble size by increasing temperature from ambient to 600 K in fluidization of Group-B particles. It is worth noting that the correlations predict an increase in bubble diameter and a decrease in minimum fluidization velocity by an increase in temperature [2, 9]. This leads more fluidity of particles which is a result of the increase in gas viscosity. If for any reason the interparticle forces become dominant, the deviation between correlations and actual bubble diameter or velocity increases. *Lettieri et al.* [10, 11] reported a case, where interparticle forces can be dominant. They studied the fluidization of fresh and used FCC catalysts at temperatures up to 650 ºC. The large deviation between calculated and measured pressure drops for FCC and doped silica catalysts at 200 ºC shows that the interparticle forces become important at this temperature. Effect of temperature on solids mixing and phase dynamics of Group-A and B particles have been previously studied in the temperature range of 25-400 ºC [12-14]. The objective of this work was to investigate the behavior of solids in fluidized beds using Radioactive Particle Ttracking (RPT) at high temperature and obtain more information about the flow structure of fluidized solids.

EXPERIMENTAL SECTION

The fluidized bed column is made of stainless steel that withstands high temperatures. It consisted of a bed and freeboard of 78 ID mm and 750 mm in height, a windbox, which also serves as a pre-heater and a disengagement zone of 150 ID mm and 900 mm in height. The gas flow rate to the reactor was measured by a rotameter at low gas velocities and an orifice plate at high gas velocities. Air at ambient temperature first passed through the windbox or preheating zone and then entered the bed through a distributor. The distributor was a perforated plate made up of 1-mm holes and a total free area of 0.5%. The temperature in the bed and at various locations in the freeboard was measured using k-type thermocouples. A controller was used to maintain the temperature in the bed at the desired set point value. Silica sand particles with an average particle size of 250 μm and density of 2650 kg/m^3 was used in the experiments. The settled bed height above the distributor was 20 cm for all the experiments.

The radioactive tracer was made of scandium oxide with a density and size close to those of the bed material. It was activated to 300 μCi in the SLOWPOKE nuclear reactor of École Polytechnique de Montréal. The long half life of radioactive scandium made it possible to run the experiment continuously for a long period. Eight detectors were located around the fluidized bed to track the tracer in the bed. The sampling time in the experiments was 20 ms. Each experiment lasted at least 4 hours. Details of the RPT experiments and the tracer position reconstruction can be found elsewhere [15].

RESULTS AND DISCUSSION

The effect of temperature and superficial gas velocity on evaluated mean upward and downward velocities is presented and discussed in the followings.

Mean Velocity

The obtained time-position data from RPT is used to determine the mean velocity of upward and downward-moving particles. The particles in the gas-solid fluidized bed usually do not move as single and isolated particles [16-18] but they do as clusters. This idea forms the basis of the algorithm for recognizing the bubbles and clusters among the trajectories in the RPT experiments, as fully reported by *Mostoufi & Chaouki* [19]. Fig. 1 shows velocity distribution of upward and downward particles at 25 ºC and superficial gas velocity of 0.52 m/s. As can be seen in this figure there are two distinguishable segments in this distribution, corresponding to velocities of

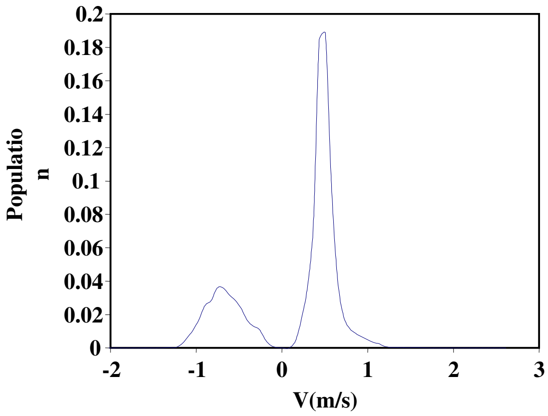

Fig. 1: Velocity Distribution at Superficial Gas Velocity of 0.52 m/s and T=25 °C.

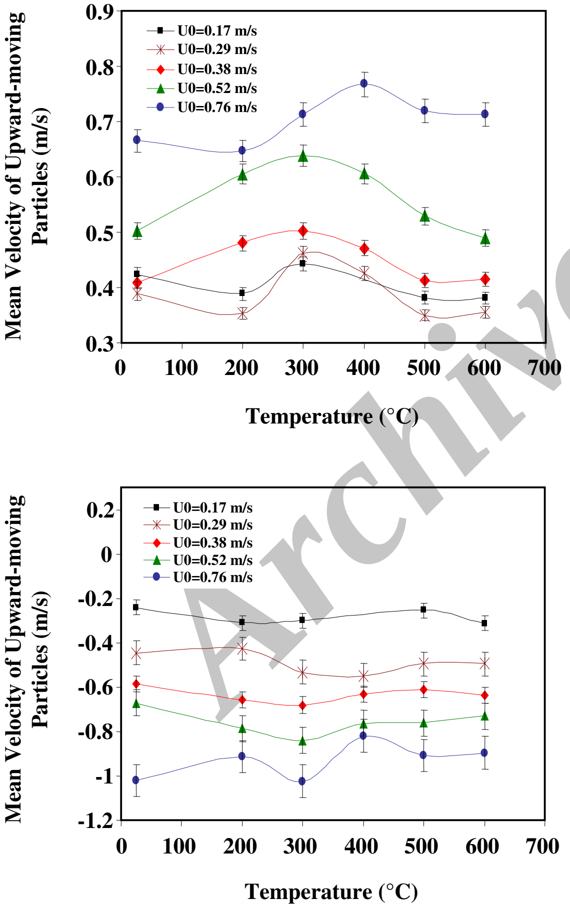

Fig. 2: Mean velocity of particles as a function of temperature (a) upward-moving (b) downward-moving

downward and upward-moving particles. Evidently, the left peak with negative velocity corresponds to the downward-moving clusters and the right peak with positive velocity belongs to the upward-moving particles. Therefore, ascending bubbles/clusters and descending clusters in the fluidized beds can be recognized by plotting the velocity distribution of these species.

Effect of temperature on the mean velocities of upward and downward moving particles was first investigated. These mean velocities are plotted as a function of temperature at different superficial gas velocities in Fig. 2a-b. As expected, by an increase in the superficial gas velocity, mean velocities are increased due to the fact that more particles are being picked up by the bubbles and ascending clusters as the superficial gas velocity is increased. As shown in Fig. 2a, at a given superficial gas velocity, the mean velocities of upward-moving particles are increased with increasing temperature from ambient temperature up to about 300 ºC. The mean velocities then are decreased by increasing the temperature over 300 ºC. This is directly related to the diameter of the bubbles in the bed. In fact, the higher bubble velocity corresponds to larger bubbles and vice versa. Therefore, according to Fig. 2a, larger bubbles are made by increasing the temperature up to 300 ºC. However, further increase in the temperature results in occurring smaller bubbles in the bed. In spite of the upward velocity, Fig. 2b indicates that there is no significant change in the mean velocity of downward-moving clusters.

Bubble Diameter

It can be assumed that in a fluidized bed, the emulsion phase acts as a liquid through which the bubbles move up. The momentum balance for a moving bubble would be then as follows:

$$F_g - F_b = -F_D \tag{1}$$

In other words,

$$g - \rho_{emulsion} V_b g = -\left(\frac{\pi D_b^{\,2} \rho_{emulsion} v_b^{\,2}}{4}\right) C_D \tag{2}$$

where

$$V_b = \frac{\pi D_b^{\,3}}{6} \tag{3}$$

Thus,

$$D_b = \frac{3}{2g}\left(\frac{\rho_{emulsion}}{\rho_{emulsion} - \rho_g}\right) {v_b}^2 C_D \qquad (4)$$

In order to calculate the bubble diameter from Eq. (4), estimation of the bubble velocity is required. The following correlations were used to estimate this value [9]:

$$v_b = U_\circ - U_{mf} + U_{br} \qquad (5)$$

$$U_{br} = 0.711\sqrt{gD_b} \qquad \frac{D_b}{D_t} \le 0.125 \qquad (6)$$

$$dU_{br} = \left(0.711\sqrt{gD_b}\right)\left[1.2\exp\left(-1.49\frac{D_b}{D_t}\right)\right] \qquad (7)$$

$$0.125 \langle \frac{D_b}{D_t} \langle 0.6$$

$$U_{mf} = \frac{\mu_g}{d_p\rho_g}\left[(28.7)^2 + 0.0494\left(\frac{{d_p}^3\rho_g(\rho_s - \rho_g)g}{{\mu_g}^2}\right)\right]^{\frac{1}{2}} - \frac{28.7\mu_g}{d_p\rho_g} \qquad (8)$$

The drag coefficient shown in Eq. (4) may be estimated prior to calculating the bubble diameter. Among several correlations for drag coefficient [20], the more adaptive correlation of *Bai & Jin* [21] (Eqs. (9)-(11)) was used.

$$\frac{C_D}{C_{D0}} = 1.68{\varepsilon_{emulsion}}^{-0.25}\left(\frac{Re_t}{Re}\right)^{-1.21}\left(\frac{D_b}{D_t}\right)^{0.11} \qquad (9)$$

$$C_{D0} = \left(\frac{24}{Re}\right) + \left(\frac{3.6}{Re^{0.313}}\right) \qquad Re \le 2000 \qquad (10)$$

$$C_{D0} = 0.44 \qquad Re > 2000 \qquad (11)$$

in which:

$$Re = \frac{\rho_{emulsion} v_b D_b}{\mu_{emulsion}} \qquad (12)$$

where the emulsion density is:

$$\rho_{emulsion} = \rho_p(1 - \varepsilon_{emulsion}) + \rho_g\varepsilon_{emulsion} \qquad (13)$$

and emulsion viscosity could be evaluated from [22]:

$$\mu_{emulsion} = \frac{4}{3}(1 - \varepsilon_{emulsion})^2\rho_s d_p g_0(1+e)\left(\frac{\theta}{\pi}\right)^{\frac{1}{2}} \qquad (14)$$

in which the radial distribution function is [22]

$$g_0 = \left[1 - \left[\frac{(1 - \varepsilon_{emulsion})}{\varepsilon_{s\max}}\right]^{\frac{1}{3}}\right]^{-1} \qquad (15)$$

$$\varepsilon_{s\max} = 0.7405, e = 0.99 \qquad (16)$$

The voidage of the emulsion can be estimated from [23]:

$$\varepsilon_{emulsion} = (\varepsilon_{mf} + 0.2) - 0.059\exp\left(\frac{U_{mf} - U_0}{0.429}\right) \qquad (17)$$

In order to calculate the emulsion viscosity from Eq. (14), the knowledge of granular temperature is required. The granular temperature is defined as:

$$\theta = \frac{1}{3}({\sigma^2}_\theta + {\sigma^2}_r + {\sigma^2}_z) \qquad (18)$$

which was evaluated from the experimental data.

The mean diameter of the bubble evaluated by the procedure described earlier is plotted against the temperature at a given superficial gas velocity in Fig. 3. This figure shows that the bubble diameter is increased by an increase in superficial gas velocity. This is consistent with correlation reported by *Werther* [9] for bubble size in a bed of Geldart B solids. In addition, as seen in Fig. 3, increasing temperature up to 300 ºC causes the growing of the bubbles, however, by further increase of temperature over 300 ºC, diameter of the bubbles is decreased. This can be attributed to drag force between emulsion and bubble phases. The increase in temperature makes the gas viscosity increase and gas density decrease. At lower temperatures well below 300 ºC, an increase in the gas viscosity is dominant in comparison with gas density decrease. However, at higher temperatures the decrease in gas density is more effective. These changes make the drag force decrease after increase initially. Based on the explained changes of drag force, it can be concluded that increasing temperature up to 300 ºC can facilitate growing of bubbles while further increase acts in opposite way. These results confirm the validity of the assumption in bubble diameter evaluation, in which emulsion was considered as a liquid through which the bubbles move up.

By implementing the obtained bubble diameter and calculated emulsion viscosity and density (Eqs. (13) and (14)), Reynolds number, which represents the turbulence

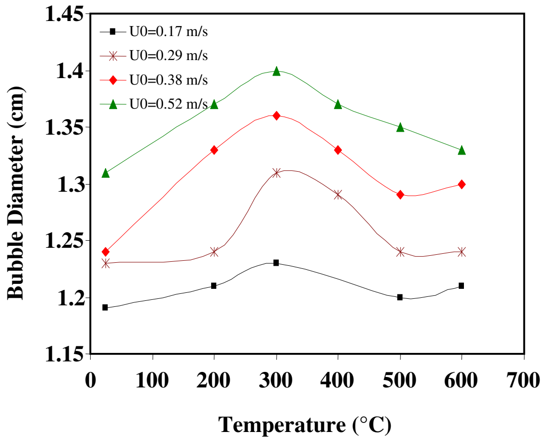

Fig. 3: Temperature effect on estimated mean diameter of bubble.

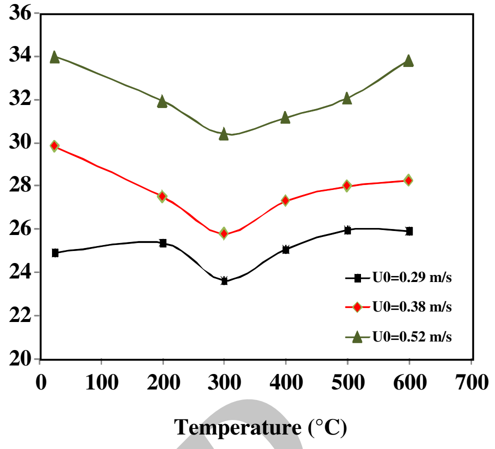

Fig. 4: Temperature effect on Reynolds number.

in the bed, was calculated. As can be seen in Fig. 4, Reynolds number shows the opposite trend of bubble diameter, i.e., it is decreases by increasing temperature up to 300 ºC and then increases by further increase in temperature afterwards. Increasing temperature up to 300 ºC leads to increase in emulsion viscosity, bubble diameter and bubble velocity. However, all these parameters have reverse trend by increasing temperature above 300 ºC. Emulsion density is almost constant and does not change significantly by increasing temperature. The reason of the opposite trend in Reynolds number is that by making any change in temperature, the change in emulsion viscosity is dominant and more significant compared to the change of bubble diameter and bubble velocity. Changes in Reynolds number confirms that physical properties of the bed (e.g., emulsion viscosity) vary with temperature which leads to different hydrodynamic behavior of the bed at various temperatures.

CONCLUSIONS

Time-position trajectory of particles was obtained by the RPT technique at elevated temperatures. The time-position data were used to determine mean velocities of upward-moving (including bubble wake and ascending clusters) and downward-moving (descending clusters) particles. The results showed that the upward velocity increases by increasing temperature up to 300 ºC and decreases by further increase in temperature. Bubble diameter was evaluated by assuming that in a fluidized bed the emulsion phase can act as a liquid through which the bubbles move up. Results indicated that the bubbles grow up to a maximum diameter by increase the temperature up to 300 ºC. Results show that, due to the wall effect, there is no significant change in the mean velocity of downward-moving clusters. The change in hydrodynamic behavior of the bed by making any change in the temperature can be attributed to the functionality of physical properties in the bed (e.g., emulsion viscosity) with temperature which has been represented by Reynolds number.

Nomenclature

U_{mf}	Minimum fluidization velocity, m/s
U_{br}	Rise velocity of bubble with respect to emulsion phase, m/s
U_0	Superficial gas velocity, m/s
v_b	Bubble velocity, m/s
D_b	Bubble diameter, m
F_g	Gravity force, N
F_b	Buoyancy force, N
F_D	Drag force, N
g_0	Radial Distribution Function
e	Restitution Coefficient
C_D	Drag Coefficient
V_b	Bubble volume, m^3
D_t	Bed diameter, m
Re	Reynolds number
d_p	Particle diameter, m

Greek Letters

ε_{mf}	Voidage at minimum fluidization
$\varepsilon_{emulsion}$	Emulsion voidage
$\varepsilon_{s\ max}$	Maximum solid fraction in emulsion
ρ_g	Gas density, kg/m^3
$\rho_{emulsion}$	Emulsion density, kg/m^3
μ_g	Gas viscosity, kg/m.s
$\mu_{emulsion}$	Emulsion viscosity, kg/m.s
ρ_s	Solid density, kg/m^3
θ	Granular temperature, m^2/s^2
σ_k	Standard deviation of k-component of velocity

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