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Effect of Horizontal Tube Position on Wall-to-Bed Heat Transfer in Air-Solid Fluidized Bed of Large Particles

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Abstract

Fluidized beds are extensively used in a variety of industrial applications, such as heat recovery systems, coal combustion, and solid particle drying. The effect of the position of the heat transfer tube on heat transfer from a horizontal bare tube in a 150 mm square cross-section fluidized bed containing large particles, ragi, and mustard is investigated. The influence of varying fluidizing velocities on heat transport is also examined for a fixed bed height of the large particles. When the position of the tube from the distributor plate was increased in 0.028m increments from 0.52 m to 0.080 m and then to 0.108 m, the heat transfer coefficient was reduced. Due to enhanced fluidization, all bed heights yielded almost comparable heat transfer coefficient values at velocities exceeding 1 m/s, 170 and 160 W/m^{2}K for ragi and mustard, respectively. The dependency of the heat transfer coefficient on fluidizing velocity is qualitatively like that of small particles. An error deviation of around 8.8% was found when comparing experimental heat transfer coefficient values to those predicted from empirical correlations in the published literature.

Keywords: *Fluidized Bed Heat Transfer, Large Particles, Tube Position*

1.0 Introduction

A fluidized bed is typically a vertical column of particles with a distributor at the bottom that injects a fluidizing medium upwards. A body submerged in a fluidized bed at a temperature that is different from its own would experience a rate of heat transfer that is many times higher than it would with just the gas. The essential characteristic of a fluidized bed that draws researchers' interest is its isothermal nature coupled with a rapid heat transfer rate[1]. Fluidized beds are extensively used in a variety of industrial applications, such as heat recovery systems, coal combustion, and solid particle drying. A particle having a mean particle size of less than one millimetre is considered small, and greater than one mm is considered large[2]. The heat transfer mechanism from a submerged surface differs significantly between large and small particle systems. The heat transfer behaviour of these particles is often conflicting, and the conclusions obtained for small particles cannot be readily generalized to large particles.

Grewal and Saxena[3] evaluated heat transfer coefficients in a horizontal tube submerged in a small particle bed with average particle diameters between 178 and 504 µm reporting no significant changes in heat transfer with a change in bed height. Kim *et al.*[4] used the packet renewal model[5] and the emulsion properties surrounding the tube to describe the heat transfer coefficient variation as flow

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velocity increases. Using annular fins on a horizontal tube, Rasouli *et al.*[6] found a reduction in heat transfer coefficient in a 200 and 307 μm silica sand bed. The most recent experimental research on horizontally submerged tubes and fluidized beds examined the effect of axial tube position[7], heater inclinations[8], tube diameter, moisture[9], and velocity[10] on small particles. Vogtenhuber *et al.*[11] found identical values of heat transfer coefficients for single tubes as well as inline and staggered arrangements in amine bulk material. Nonetheless, Lechner *et al.*[12] studied heat flow dependency of horizontal tube diameter in large particle fluidized beds. However, Masoumifard *et al.*[7] reported a weak dependency of the heat transfer coefficient on the heat transfer tube position for sand particles of three different particle sizes (280, 490 and 750 μm). However, none of these studies examined the heat transfer characteristics of large particle systems by varying the position of the heat transfer tube.

The present work is aimed at investigating the influence of heat transfer tube position above the distributor plate, at varying fluidizing velocities on the heat transfer from a single horizontal bare tube submerged in a large particle bed. The experimental findings are compared with empirical correlations that have been published in the literature.

2.0 Experimental Set Up

The experimental setup depicted in Figure 1 consisted of a 150 mm square cross-section fluidized column. The height of the test section was 0.5 m extended with a freeboard region of 0.5 m. The initial static bed height was maintained at 0.125 m for both ragi and mustard. A horizontal copper tube with an outer diameter of 0.016 m and length of 0.15 m is used as a heat transfer surface. The experiments were conducted for the three different horizontal positions of the heat transfer tube i.e., 0.052 m, 0.08 m, and 0.108 m, above the distributor plate. A cartridge heater of outer diameter 0.012 m was inserted right into the copper tube. Nylon plugs were used at both ends of the heat transfer tube to avoid axial heat loss. K-type thermocouples were used to measure the temperatures of the tube wall and the bed. The electrical energy input to the cartridge-type heater fitted inside the tube was regulated with a wattmeter with a variac. Air from a centrifugal blower was used as fluidizing gas.

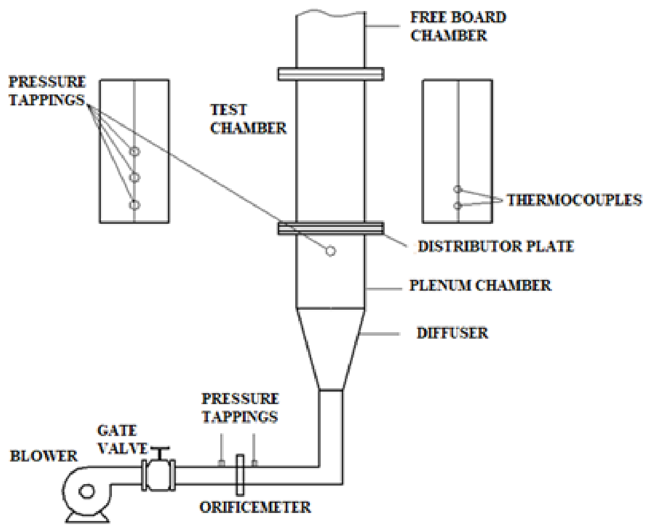

Figure 1. Schematic diagram of experimental set up.

The particles used in the present work were ragi and mustard with average particle diameters of 1.14 and 1.4 mm, respectively. Sieve analysis was used to find the average particle diameter. Particle densities of ragi and mustard are determined using the kerosene test. Particle densities of ragi and mustard were 1300 and 1100 kg/m^3, respectively. The bed was supported by a multi-orifice plate with 111 holes of diameter 3 mm. A stainless-steel mesh was fitted on the distributor plate to accomplish uniform air distribution. The velocity of air varied between 0.4 and

Figure 2. Experimental setup.

Figure 3. Test section.

1.1 m/s. Fluidizing velocity was measured using an orifice meter attached to a differential pressure manometer.

A 3-kW centrifugal blower was used to provide the air for fluidization. Throughout the experiment, a steady heat input of 50 watts was maintained. Heat transfer coefficient h_w in W/m^{2}K was calculated from;

$$h_w = \frac{V\ I}{A_t(T_s - T_b)} \tag{1}$$

In all the experiments, the temperatures are measured using K-type thermocouples. For these specifications of K-type thermocouples used, the uncertainty level in the measurement of temperature is ± 2 %. The uncertainty in obtaining heat transfer coefficient h_w was found by employing the Kline and McClintock[13] Method, Equation (2) and by Equation (1):

$$U_{h_w} = \sqrt{\left(\frac{\partial h_w}{\partial T_s}\omega_s\right)^2 + \left(\frac{\partial h_w}{\partial T_b}\omega_b\right)^2 + \left(\frac{\partial h_w}{\partial V}\omega_V\right)^2 + \left(\frac{\partial h_w}{\partial I}\omega_I\right)^2} \tag{2}$$

The percentage uncertainty of the heat transfer coefficient $\frac{U_{h_w}}{h_w} x\ 100$ is 6 per cent.

3.0 Results and Discussion

Variation of heat transfer coefficient (h_w) with different tube positions, at different velocities is depicted for ragi and mustard in Figures 4 and 5 respectively. All other conditions are unchanged, the change in tube position from 0.52 m to 0.80 m, then to 0.108 m, resulted in a slight reduction in h_w at low velocities. This reduction was caused by the increased airflow resistance across the fluidized bed with increased height of the tube from the distributor plate, which lowered the gas convective share to the total heat transfer coefficient at low velocities.

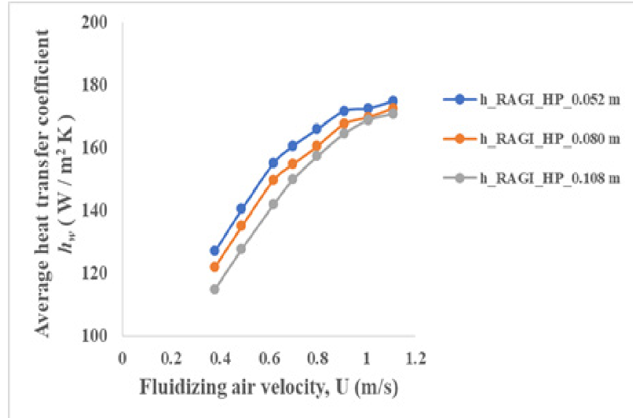

Figure 4. Variation of h_w coefficient with the heat transfer tube position (Ragi).

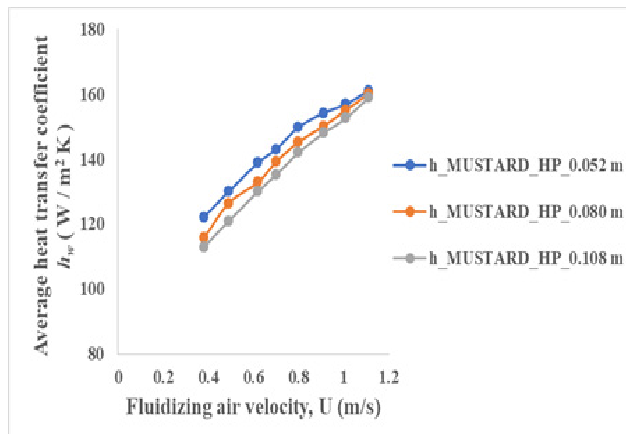

Figure 5. Variation of h_w with the heat transfer tube position (Mustard).

Higher tube positions restricted the air penetration through the bed material to reach the heat transfer surface, causing an increasing gas conduction path. At higher velocities, the heat transfer coefficient is reduced for higher positions, due to a reduction in solid volume fraction at the heat transfer surface. Interesting that all bed heights resulted in similar values of heat transfer coefficient at velocities greater than 1 m/s, about 170 W/

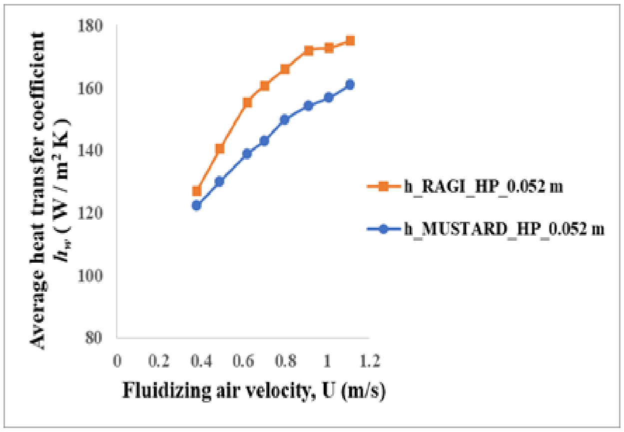

Figure 6. Variation of h_w with fluidizing air velocity (at H_s = 0.125 m).

m²K for ragi and 160 W/m²K for mustard. This resulted from the rapid movement of particles under bubbling fluidized conditions at greater velocities. The vigorous motion of the particles observed at velocities greater than 1 m/s led to the same values of heat transfer coefficient regardless of the tube position.

The trend of variations in h_w with fluidizing velocity almost remained the same for both ragi and mustard. Except for velocities greater than 1 m/s, it was evident that the heat transfer coefficients for mustard were consistently lower than those for ragi for all tube positions evaluated. This is due to an increase in the gas conduction path between the tube and the initial row of particles in the case of large-sized mustard compared to ragi.

The variation of the wall-to-bed heat transfer coefficient with fluidizing velocity for large particles, ragi, and mustard at the same bed height of 0.125 m and tube position 0.052 m above the distributor plate are compared in Figure 6. Initially, a rapid increase in the values of h_w with an increase in air velocity was noted in both situations. This is attributed to greater particle contact with the tube surface at lower velocities as the particle residence time is longer.

Any additional increase in velocity reduces the particle residence time on the tube surface due to fluidized conditions. However, a flatter curve was observed for mustard compared to ragi. Heat transfer coefficients were lower for mustard compared to ragi as the gas conduction path between the tube and the initial row of particles increased.

A maximum error deviation of around 8.8 per cent was estimated when the experimental heat transfer coefficient values were compared to those predicted by empirical relationships in the available literature, as presented in Table 1.

$$\bar{\delta}_1 = \frac{100}{N}\sum_{i=1}^{N}\left|\frac{prediction - experiment}{experiment}\right| \tag{3}$$

$$\bar{\delta}_2 = 100\sqrt{\frac{1}{N}\sum_{i=1}^{N}\left(\frac{prediction - experiment}{experiment}\right)^2} \tag{4}$$

4.0 Conclusion

The influence of the position of the heat transfer tube above the distributor plate on heat transfer from a single horizontal bare tube was investigated for an air-solid fluidized bed of 150 mm square cross-section containing large particles, ragi, and mustard at fluidizing velocities between 0.4 and 1.1 m/s. When all other parameters were constant, changing the position of the heat transfer

Table 1. Mean percentage deviation of h_w from predicted heat transfer coefficient.

Author	Correlation	δ_1	δ_2
Andeen and Glicksman[14]	$Nu_T = \frac{hD}{K_g} = 900(1-\epsilon)\left[\left(\frac{GD\rho_s}{\rho_g H}\right)\left(\frac{\mu^2}{d_p^3\rho_s^2 g}\right)\right]^{0.326} Pr^{0.3}$	6.8	8.8
Rasouli *et al*[6]	$Nu_p = \frac{hd_p}{K_g} = 1.475\, Re_p^{0.404} Pr^{0.3}$	4.0	5.5

tube above the distributor plate resulted in a drop in the heat transfer coefficient at low velocities. Heat transfer coefficients were observed to be slightly higher for the tube placed at a height of 0.052 m above the distributor plate as compared to higher positions at low velocities. A maximum error deviation of around 8.8% was estimated by comparing the experimental heat transfer coefficient values to those predicted by empirical relationships in the available literature.

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Nomenclature

d_p	Particle diameter (m)	T_s	Average temperature of the tube surface (K)
μ	Absolute viscosity of gas, (Ns/m^2)	T_b	Average temperature of the bed (K)
ρ_g	Density of gas (kg/m^3)	V	Voltage (V)
D	Tube diameter (m)	h_w	Wall-to-bed heat transfer coefficient (W/m^{2}K)
ϵ	Void fraction	Re_p	Particle Reynolds number
ρ_s	Particle density (kg/m^3)	Pr	Prandtl number
H_s	Static bed height (m)	Nu	Nusselt number
A_t	Tube surface area (m^2)	Nu_p	hd_p/K_g
U	Fluidizing air velocity (m/s)	Nu_T	hD/K_g
G	Fluidizing air mass flow rate (kg/m^{2}s)	I	Current (A)
$\bar{\delta}_1$	Absolute average percentage deviation	$\bar{\delta}_2$	Root mean square percentage deviation
N	Sample size		