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Heat Transfer Around a Horizontal Tube in Freeboard Region of Fluidized Beds

An experimental investigation was carried out to measure local and average heat transfer coefficients for horizontal tubes located in freeboard region of air fluidized beds. Tests were carried out at room temperature and atmospheric pressure in a rectangular fluidized bed, with mean particle diameters of 275 to 850 μm.

Both local and average heat transfer coefficients were found to vary with particle diameter, flow rate, static bed depth, and elevation in the freeboard region.

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SCOPE

In fluidized-bed combustors, the heat transfer tubes are located with a portion of the tubes submerged within the fluidized bed and the remaining portion placed in the freeboard space above the fluidized bed. The effective heat transfer coefficients for tubes located within the bed region can be an order of magnitude greater than the heat transfer coefficients for tubes in the freeboard space. Designers hope to utilize this different behavior to adjust operating power level of fluidized-bed combustors. Current plans call for "turning-down" the thermal output by dropping the fluidized-bed level to expose greater portions of the coolant tubes in the freeboard.

The rate of heat transfer between a fluidized bed and a submerged tube has been measured by a number of investigators for uniformly sized particles (Vreedenberg, 1958; Ainshtein, 1966; Andeen and Glicksman, 1976; Chen, 1976; Ozkaynak and Chen, 1978, 1980; Saxena et al., 1978; Staub, 1979; Chandran and Chen, 1980; Grewal and Saxena, 1980) and for mixed particle sizes (Biyikli and Chen, 1982). Heat transfer to tubes in freeboard region of fluidized beds depends on a great number of factors, including the properties of the bed material and the fluid, bed and tube geometries, tube elevation and fluidization state. To our knowledge, only a few data are available for circumferentially averaged heat transfer coefficients to tubes in freeboard region of fluidized beds (George and Grace, 1979; Wood et al., 1980; Byam et al., 1981), and there is no detailed information about the local heat transfer coefficients.

The objective of the present investigation was to measure the local and circumferentially averaged heat transfer coefficients for a horizontal tube placed in the freeboard region of fluidized beds, for different test particles and over a range of gas velocities and at different tube elevations.

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CONCLUSIONS AND SIGNIFICANCE

Heat transfer from a horizontal tube, placed in the freeboard region of air fluidized beds, was investigated experimentally. A rectangular fluidized bed (0.2 × 0.3 × 3.0 m) and a 32-mm O.D. by 150-mm-long instrumented tube were used to measure local heat transfer coefficients around the tube. Average heat transfer coefficients were calculated by area averaging of the measured local coefficients. Experiments were carried out for different particles, air flow rates, and tube elevations. Glass beads with 275 and 850 μm mean diameters, and sand with 285 and 465 μm mean diameters were used as fluidizing particles. Air velocity was varied up to 50 times the minimum fluidization velocity. Tube elevation was varied from the static bed surface to a height of 2.25 m above the surface.

For any given gas flow rate, the local heat transfer coefficients varied with circumferential position around the tube. Maximum heat transfer coefficients tended to occur at the top of the tube. Heat transfer coefficients at all positions around the tube decreased with increasing height in the freeboard region.

The variation of average heat transfer coefficient with elevation in the freeboard was as much as an order of magnitude, decreasing to gas convection heat transfer with increasing elevation. This decline was found to be moderated by increasing flow rates. At a constant flow rate and constant elevation in the freeboard, average heat transfer coefficients increased with decreasing particle diameters.

As a first-order correlation of the experimental data, nondimensional groups of heat transfer coefficients and air velocities were used to consolidate the results for different particle diameters into single curves for a constant elevation. Experimental results in these nondimensional forms were also compared with the few existing data. It is suggested that these dimensionless parameters may be a useful method for generalizing heat transfer data in freeboard region of fluidized beds.

EXPERIMENTAL APPARATUS AND PROCEDURE

A rectangular fluidized bed (0.20 × 0.30 × 3.0 m) was designed and fabricated for freeboard experiments at room temperature. Three sides of the bed were made from aluminum plates to decrease the static charge in the bed. The fourth side was constructed from glass plates, to permit visual observations of the fluidization conditions. The distributor was a sandwich assembly of a perforated steel plate, porous plastic plate, and a stainless-steel screen. Acrylic plastic doors on the aluminum walls were used for the attachment of heat transfer tubes. A cyclone separator was used to collect and recycle the entrained particles during operations at high fluidization velocites. The flow diagram of the fluidized bed test unit is shown in Figure 1.

A specially instrumented tube was built for the measurement of local heat transfer coefficients. As shown in Figure 2, the test tube was fabricated from a 3.175-cm diameter Lexan rod, around which were embedded 8 Inconel foil strips to generate the required heat flux. The power was supplied to each strip separately with individual voltage control. Thermocouples, located in grooves under the strips, were used to determine rod surface temperatures. In operation, power to the individual strips was adjusted to obtain an isothermal temperature around the tube.

The fluidizing air was supplied by a pair of reciprocating compressors, and flow rate was measured with annubar sensors (pitot tube rakes) with a range of 0.0004 to 0.57 m^3/s. High gas velocites were reached, up to 3.5 m/s. Steam was used to humidify the inlet air to reduce problems of electrostatic charges in the bed.

For each test run, many data samples were recorded and processed to obtain time-averaged information at steady-state conditions. Local heat transfer coefficients were calculated from measured power input to each heating strip and the temperatures of the tube surface and the bulk bed.

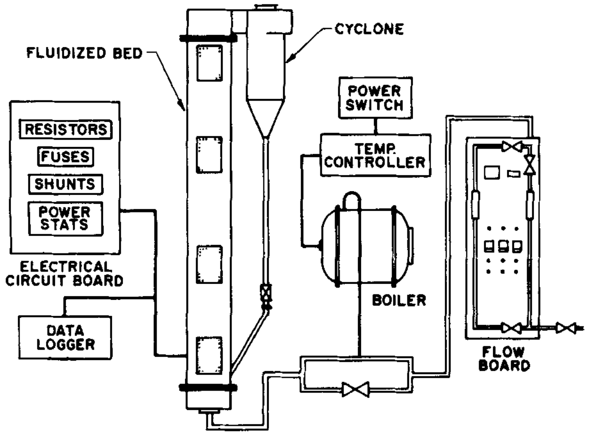

Figure 1. Fluidized-bed test unit.

Average heat transfer coefficients were calculated by area averaging of the local coefficients.

Tests were carried out to determine the variation of heat transfer coefficients with tube elevation in the freeboard of the fluidized bed. For a selected static bed height, the heat transfer tube was placed at 1.6, 19, 58, 147 and 225 cm elevations above the static bed. Tube elevation is defined as the height between static bed upper surface and the centerline of the test tube. The local and average heat transfer coefficients were measured for different particles, air flow rates, and tube elvations. Glass beads with 275 and 850 μm mean diameters and sand with 285 and 465 μm mean diameters were used as fluidizing particles. Properties of the test particles are given in Table 1. Entrainment and minimum fluidization velocities were calculated using the mean diameter of the particles. Sphericities of the test particles were approximately unity.

EXPERIMENTAL RESULTS

These experiments gave information on: (a) circumferential distributions of the local heat transfer coefficients around the horizontal tube; (b) the variation of average heat transfer coefficients with tube elevation; and (c) effect of static bed height on heat transfer coefficients in the freeboard region. The estimated uncertainty in measured heat transfer coefficients was ∓4% due to instrumentation precision and ∓2% due to air humidification effects, for a total uncertainty of ∓6%.

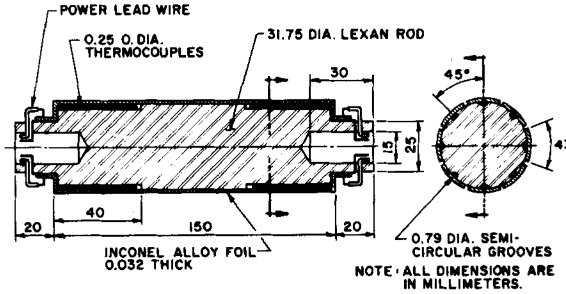

Figure 2. Test tube of local heat transfer coefficients.

TABLE 1. PROPERTIES OF TEST PARTICLES

Designation	Size Range (μm)	Mean Size d_p (μm)	U_{mf} (m/s)	U_t (m/s)
Glass Beads	200–300	275	0.0615	2.15
Glass Beads	680–1100	850	0.434	6.66
Silica Sand	115–350	285	0.065	2.24
Silica Sand	250–700	465	0.173	3.65

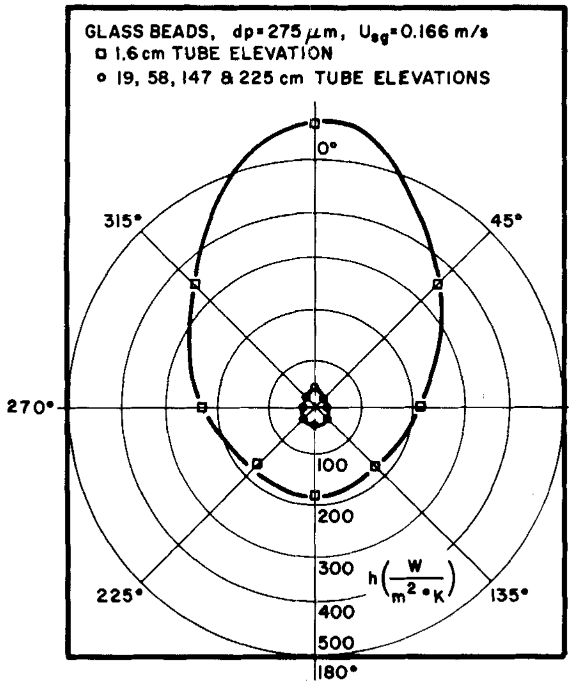

Figure 3. Local heat transfer coefficient around tube in freeboard region for three times minimum fluidization velocity.

Local Heat Transfer Coefficients

Local heat transfer coefficients were measured around the circumference of the horizontally placed test tube in freeboard region of fluidized beds of different particles. Air flow velocities varied from 0.1 to 3.5 m/s.

Some experimental results showing the variation of heat transfer coefficients around the tube for various heights in freeboard region are plotted in Figures 3 and 4. The particles were glass beads with 275 μm mean diameter. Local heat transfer coefficients are given at low and high gas flow rates. In these polar plots of the local heat transfer coefficient, zero degree is defined as the top of the horizontal tube. Figure 3 shows the local heat transfer coefficients for a low gas flow rate, corresponding to three times minimum fluidization velocity. At 1.6-cm elevation, the tube is essentially immersed into the bed, while at higher elevations the tube is exposed to

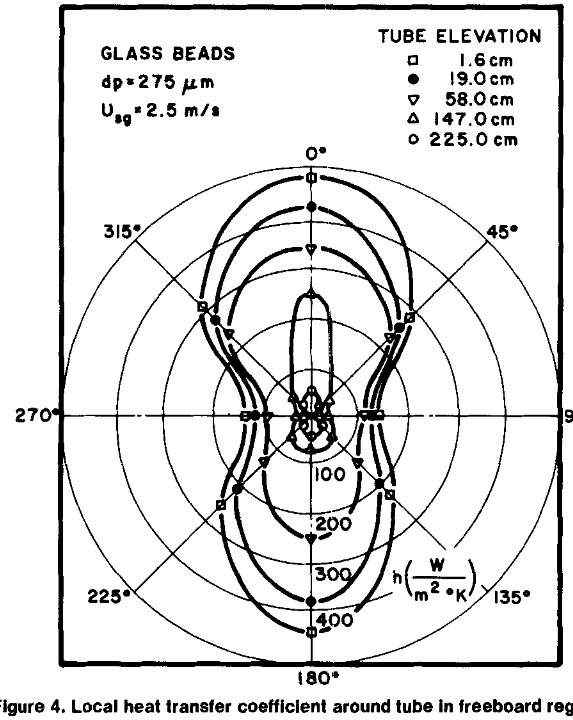

Figure 4. Local heat transfer coefficient around tube in freeboard region for 40 times minimum fluidization velocity.

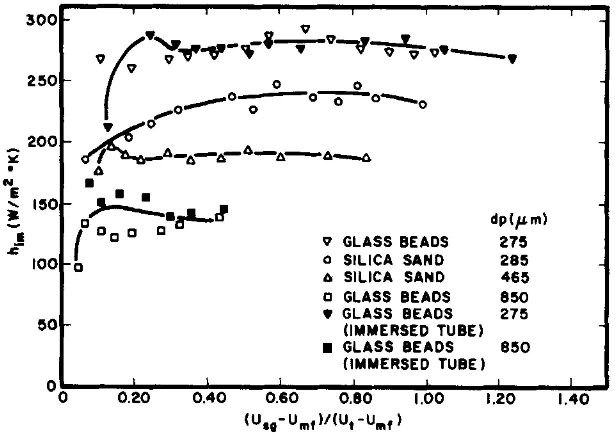

Figure 5. Average heat transfer coefficients at static bed surface and immersed tubes.

forced convection of air only. At this low gas flow rate, the distinctly reduced heat transfer coefficients for the tubes in the freeboard region are quite obvious. Figure 4 shows the heat transfer coefficient at high gas flow rate, corresponding to 40 times minimum fluidization velocity. At this higher gas flow rate, some particle entrainment occurs into the freeboard region so that significantly enhanced heat transfer (over air convection) was observed at tube elevations up to 147 cm.

For a given gas flow rate and a given particle, the local coefficient varied with circumferential position around the tube. Maximum heat transfer coefficients tended to occur at the top of the tube. It is seen that the local heat transfer coefficient at all positions around the tube decreased with increasing height in the freeboard region. This trend was consistent for all particles and at all gas flow rates. At higher gas flow rates, as indicated in Figure 4, the local heat transfer coefficients are nonuniform around the tube. The coefficients at the top and bottom segments of the tube were found to be as much as three times greater than the coefficients at the sides of the tube. It is hypothesized that this trend is due to particle motion which results in more particle contact at bottom and top of the tube, and less contact at the sides of the tube. Conceivably, the forward stagnation point (lower surface) is impacted by upward entrained particles, which upon deentrainment then "rains down" onto the top surface. This sporatic process could lead to efficient particle-surface renewal and provide relatively high heat transfer coefficients at the top and bottom portions of the tube. The depressed heat transfer coefficients at the sides of the tube would indicate a gas-convective mechanism with relatively small particle surface transfer. These mechanisms are speculative at this time. It would be necessary to obtain direct measurements of solid contact over various portions of the tube surface to assess the validity of these hypotheses.

Average Heat Transfer Coefficients

Whereas the local heat transfer coefficients are of fundamental interest, design applications require averaged values of heat transfer coefficients. Area-averaged heat transfer coefficients were calculated from the measured local heat transfer coefficients in this investigation. The results give the effect of freeboard height on such average heat transfer coefficients as a function of superficial gas velocity.

Figure 5 plots the average heat transfer coefficients vs. nondimensional gas velocity for different test particles, for the case where the bottom of

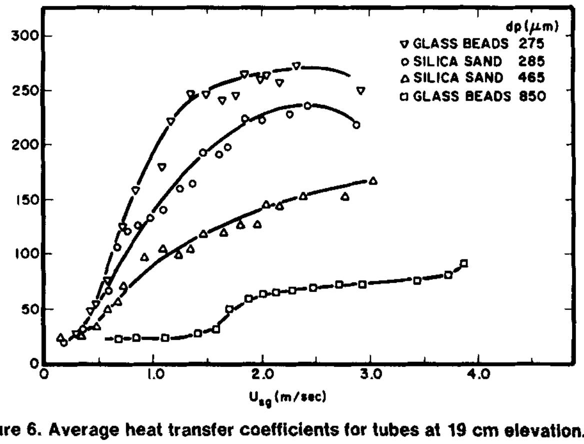

Figure 6. Average heat transfer coefficients for tubes at 19 cm elevation.

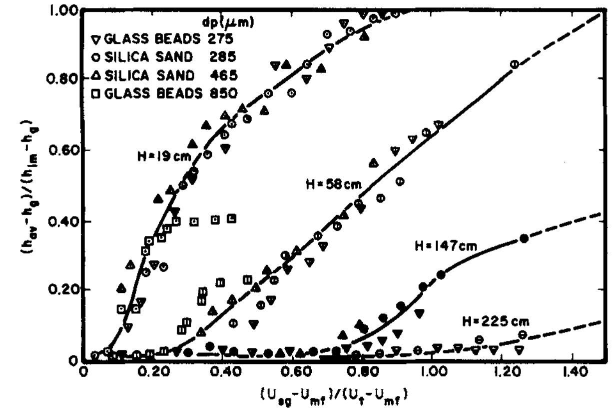

Figure 7. Average heat transfer coefficients as a function of gas velocity in freeboard region for different test particles.

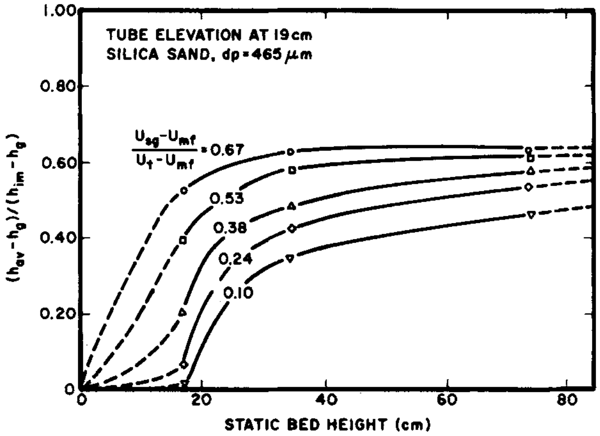

Figure 9. Effect of static bed height on heat transfer coefficients.

the test tube was just at the same height as the static bed surface. The heat transfer coefficients, for the case of the tube immersed in the static bed (tube centerline at midelevation of static bed height), are also shown for two particles for comparison. The immersed-tube heat transfer coefficients were found to be in close agreement with the heat transfer coefficients for the tube at the static bed surface.

The heat transfer data vs. superficial gas velocity for 19-cm tube elevation are shown in Figure 6 for all four particles. At a constant gas velocity, heat transfer coefficients increased with decreasing particle diameters. Similar trends were observed for other elevations tested. This effect of particle diameter is due to different particle entrainment into the freeboard region. At a constant gas velocity, particle entrainment is higher at smaller diameters. To consider the effect of particle diameter, it is suggested here that a nondimensional form of heat transfer coefficients and velocities can be used, Figure 7.

The heat transfer coefficients in freeboard region were normalized by using immersed-tube heat transfer coefficients from Figure 5 and single-phase heat transfer coefficients calculated for gas alone in cross flow at corresponding velocities. The superficial gas velocities were nondimensionalized by using entrainment and minimum fluidization velocities which were calculated based on mean particle diameters. Sphericities of the test particles were assumed to be unity. It can be seen, in these dimensionless coordinates, that heat transfer data for different test particles form a band at a given tube elevation. The solid lines indicate the approximate mean of these bands. It is suggested that these dimensionless parameters may be a useful method for generalizing heat transfer data in freeboard region of fluidized beds, if future accumulation of data agrees with these results. Figure 7 shows that the increase of heat transfer coefficients at a given tube elevation is as much as an order of magnitude with increasing gas velocity in freeboard region of fluidized beds. For a given tube elevation, heat transfer coefficient increases with increasing gas velocity and approaches the immersed tube value. This approach to the immersed tube heat transfer coefficients occurs at different gas velocities for different elevations. At lower elevations, heat transfer coefficients approach their immersed values at lower gas velocities.

The variation of heat transfer coefficients along the freeboard height are shown in Figure 8 for different gas velocities. These cross plots were obtained by using mean values of heat transfer data which are shown as solid lines in Figure 7. The variations of average heat transfer coefficients with elevation in the freeboard was as much as an order of magnitude, decreasing to gas convection heat transfer with increasing elevation. The decline of the average heat transfer coefficients with increase in elevation was found to be moderated by increasing flow rates. This decrease was sharpest for low flow rates, reflecting decreased particle entrainment into the freeboard region with lower flow rates.

Effect of Static Bed Height on Heat Transfer

The effect of static bed height on the average heat transfer coefficient was studied for some cases. Experiments were carried out for silica sand with 465 μm mean diameter by keeping tube elevation at 19 cm above the static bed for different static bed heights. The results are shown in Figure 9 for different gas velocities. For a given gas velocity, increasing the static bed depth increases the freeboard heat transfer coefficient for small static bed depths. For static bed depths greater than 30 cm, this effect was negligible.

COMPARISON WITH OTHER EXISTING DATA

To our knowledge very few data are available for heat transfer to tubes in freeboard region of fluidized beds (George and Grace, 1979; Wood et al., 1980; Byam et al., 1981). The available data are for tube bundles and are often complicated by effects of changing static bed heights. In spite of the different operating conditions, to see the trend, the available existing data on average heat transfer coefficients were compared with heat transfer coefficients measured in this study. Table 2 summarizes the test conditions of the various investigations for data used in this comparison. The comparison is shown in Figure 10 for two different gas velocities over a range of freeboard heights. For a given gas velocity, the heat transfer data of this work agrees well with the data of Byam et al. (1981) and shows similar trend with the data of George and Grace

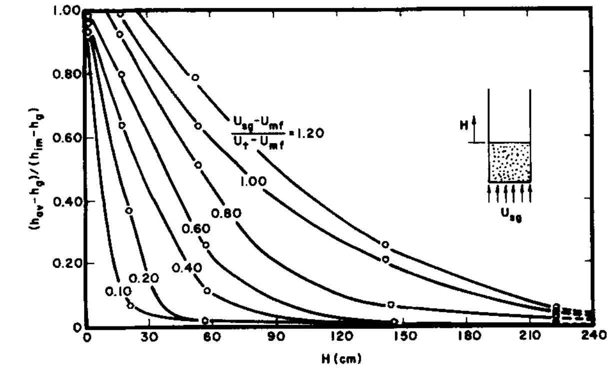

Figure 8. Variation of normalized heat transfer coefficient along freeboard for different gas velocities.

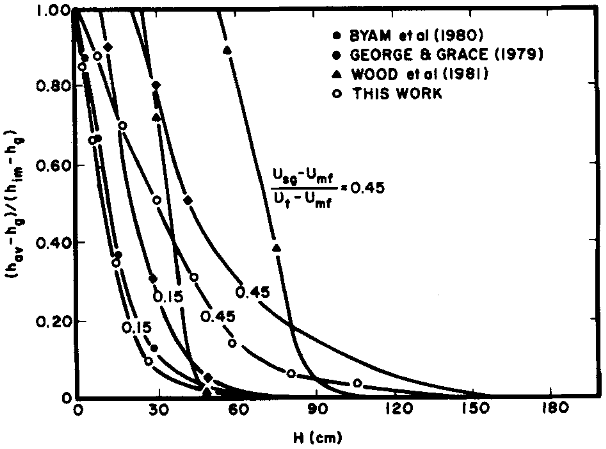

Figure 10. Heat transfer coefficients in freeboard region of fluidized beds.

TABLE 2. TEST CONDITIONS OF VARIOUS INVESTIGATORS

Reference	Particles	Mean Particle Size (μm)	Heat Transfer Tube	Static Bed Hgt. (m)	Bed Temp. (°C)	Bed Pres. (atm)
George and Grace (1979)	Silica Sand	102, 470, 890	Tube Bundle	0.22–0.75	120–145	1
Byam et al. (1980)	Coal-Dolomite	1100	Tube Bundle	1.2–1.5	780–890	6
Wood et al. (1981)	Silica Sand	930	Tube Bundle	0.15–0.70	Room Temp.	1
This Work	Glass Beads Silica Sand	275, 850 285,465	Single Tube	0.36	Room Temp.	1

(1979). However, the results of Wood et al. (1980) show higher heat transfer coefficients for the same height and steeper decline to gas convection coefficients. Only this qualitative comparison is possible in view of the different test conditions.

ACKNOWLEDGEMENT

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SUMMARY

The heat transfer characteristics for horizontal tubes in freeboard region of fluidized beds were investigated experimentally. Both local and tube-averaged heat transfer coefficients were measured for different operating conditions (gas velocity, particle size, tube elevation). The conclusions drawn from these results may be summarized as follows.

1. It was found that heat transfer coefficients decrease with increasing elevation along the freeboard, and finally approach the coefficient for gas forced convection heat transfer at the given gas velocity.
2. At a given tube elevation, increasing the gas velocity increases the heat transfer coefficients and for lower regions of the freeboard, the coefficients can approach values corresponding to tubes submerged in the fluidized bed.
3. Heat transfer coefficients can be affected by static bed height in freeboard region of fluidized beds for a given tube elevation and gas velocity.

NOTATION

d_p = mean particle diameter, μm
h = local heat transfer coefficient, w/m^2·K
h_{av} = average heat transfer coefficient, w/m^2·K
h_g = heat transfer coefficient for gas alone, w/m^2·K
h_{im} = immersed tube heat transfer coefficient, w/m^2·K
H = freeboard height, cm
U_{mf} = minimum fluidization velocity, m/s
U_{sg} = superficial gas velocity, m/s
U_t = terminal velocity of particles, m/s

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