

Fluidized bed heat transfer

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ABSTRACT

OUT of the three distinct mechanisms of heat transfer in fluidized beds namely (i) fluid-to-particle, (ii) particle-to-fluid and (iii) wall-to-bed, only the first one has been widely studied and detailed review of the several investigations has been reported earlier by J. J. Barker.^{1,2} In the other two cases no such survey has been made and an attempt is made here to give a concise report of the investigations available in literature. The report covers references, majority of which deal with the aspects of study of latter two types of mechanisms of fluidized bed heat transfer. Some important industrial applications of fluidized bed heat transfer have also been discussed.

Introduction

THE varied applications of fluidization technique in chemical and allied industries are recognised and scientific investigations into this field have formed an important and ever-widening proportion of research during the past few years. The pursuit is still on and complete facts about it are yet to be known. The investigations that are already made or yet to be made in this field, can be in the following lines:

- (a) Momentum transfer studies
- (b) Heat transfer studies
- (c) Mass transfer studies

The present article only discusses some salient aspects of heat transfer studies in a fluidized bed.

Heat transfer studies in fluidized bed is a very extensive topic which has drawn the attention of research scientists only during the last two decades. Generally there are three important aspects that are investigated in connection with fluidized bed heat transfer:

- (i) Fluid-to-particle heat transfer
- (ii) Particle-to-fluid heat transfer
- (iii) Wall-to-fluid heat transfer

The discussion here is presented in four sections.

In the first part, the mechanism of heat transfer from fluid to solid particles is discussed. Besides, a brief review of work of a few investigators for predicting fluid-to-particle heat transfer coefficients is also given. The second section includes the physical mechanisms of heat transfer from solid particles to the fluid medium in a fluidized bed and the empirical

correlations obtained for different systems by various investigators. The mode of heat transfer from the wall to the fluid and the generalised equations have been dealt with in the third section. In the fourth part, some important industrial applications of the fluidized bed heat transfer mechanism have been discussed.

FLUID-TO-PARTICLE HEAT TRANSFER

Knowledge of the rates of heat transfer from gas to particle is desirable in order to estimate particle surface temperature from the measured temperature of the gas. Such information is especially useful when considering reactions involving large heat effects that occur on and within fluidized catalyst particles. The particle surface temperature affects the reaction kinetics and small changes in temperature often cause large variations in reaction rate. It is also desirable to know actual particle temperature when dealing with heat sensitive materials.

Single-particle theory

It is well known that high rates of heat transfer between solid and gas occur in fluidized beds. This is due to the large surface available for gas-solid contact rather than high heat transfer coefficients. Coefficients are normally in the range of 1-40 Btu/hr. ft².°F; whereas heat transfer areas vary from 1,000 to 15,000 ft²/ft.³ of the bed.

The major resistance to heat flow between gas and solid in fluidized bed is to conduction through a thin gas film around the particle. Thus the point coefficients, h_1 , is,

$$h_1 = k_1 / x_1 \quad (1)$$

$$\text{where } x_1 = \phi (D_p V \rho / \mu, \psi, \delta, \epsilon)$$

Considering this single particle conduction in the line of Fourier, the final heat transfer equation has been derived as,

$$h D_p / k = D_p / x + 2 \quad (2)$$

which for an infinite fluid, reduces to the limiting value of 2.

Unsteady state conduction

Due to particle movement in fluidized bed, it would be expected that for a given particle the process is

an unsteady one. From dimensional analysis it has been found to be

$$h D_p/k = \phi_1 (D_p V \rho/\mu)^c (C_p \mu/k)^t \times \phi_2 (\epsilon, \psi, \delta) \quad (3)$$

SUMMARY OF WORK

(a) Gas-solid system

Kettenring, Manderfield and Smith¹¹ were the first to report gas-solid transfer coefficients in case of silica gel and alumina particles of narrow size range, heated and fluidized by air. Walton et. al.²³ measured heat transfer coefficients between gas and solids in beds of crushed coal fluidized with air. Wamsley and Johanson²⁴ investigated gas-to-particle heat transfer in fluidized beds of glass beads, washed alumina and Dowex-50 particles under unsteady state heating condition. The next unsteady state heat transfer studies were due to Fritz⁵ followed by Ferron⁴. Heertjes and McKibbins^{7,8} investigated temperature and humidity variations in a drying fluidized bed of wet silica gel fluidized by air. The evaporation of brine solution in fluidized salt bed with high temperature inlet gas was studied by Frantz.⁶ The coefficients were correlated with Re. No. although only one particle size of sand was used. Richardson and Ayers¹⁶ also studied heat transfer under steady state but in rectangular vessel. Many type of particles of different sizes were fluidized with air and CO₂. The coefficients were correlated with particle Reynolds no.

Sunkoori and Kaparthi¹⁹ were the only investigators to report heat transfer coefficients between fluidized particles and a liquid medium, and coefficients have been calculated by a modification of the unsteady state method of Wamsley and Johanson.²⁴ Quartz and granite particles were fluidized with water and equation of Wamsley was used for evaluating heat transfer coefficients.

The results of the above investigators have been summarized by Frantz,⁶ and presented in the form of the following table:

All the investigators except Fritz⁵ and Ferron⁴ have correlated their experimental data using equation containing only the Reynolds number and Nusselt number. These two found appreciable variation of h with the height of the bed and hence correlated their data in terms of velocity of fluid and the length of the bed.

Frantz⁶ observed that for the same Reynolds number the values of Nusselt number prediction according to the various equations proposed varied from 0.001 to 2. From dimensional analysis they have observed that the best form of correlation should be of the type

$$Nu = \phi_1 (Re)^a (pr)^b \times \phi_2 (\psi)$$

From the data of all the above investigators the following two equations — one for gas-solid and the other for liquid-solid fluidization — have been suggested:

Gas-solid system:

$$h D_p/k = 0.015 (D_p G/\mu)^{1.6} (C_p \mu/k)^{0.67} \quad (4)$$

Liquid-solid system:

$$h D_p/k = 0.016 (D_p G/\mu)^{1.3} (C_p \mu/k)^{0.67} \quad (5)$$

PARTICLE-TO-FLUID HEAT TRANSFER

The problem of particle-to-fluid heat transfer is widely encountered. Until comparatively recently, all operations involving particle-fluid heat transfer employed fixed and moving beds. A pertinent example is the pebble heater, the contacting unit of which may be fixed or moving bed type. Another and older example is the blast furnace. The charge descends down the furnace shaft and is contacted by a rising stream of hot gases, which is a complicated type of moving bed.

In recent years, applications of particle-fluid heat transfer in the fluidized state have become more

Investigators	Technique used	Correlations used	h range
(1) Kettenring et al.	Steady state gas fluidized	$h D_p/k = .0135 (D_p G/\mu)^{1.35}$	3-10
(2) Walton et. al.	Steady state gas fluidized	$h D_p/k = .0028 (D_p G/\mu)^{1.7} (D_i/D_p)^{.2}$	5-35
(3) Wamsley et. al.	Unsteady state gas fluidized	$h = 1270 D_p^{1.27}$	0.07-9
(4) Fritz	Unsteady state gas fluidized	$h = V/(116 + 52.4 Lh) + (1 - 0.62)/(274 + 447L + 146L^2)$	0.0005-0.012
(5) Ferron	Unsteady state gas fluidized	"	0.0002-0.06
(6) Heertjes et. al.	Steady state gas fluidized	$h = 1.31 (D_p G/\mu)^{0.76}$	6-28
(7) Frantz	Steady state gas fluidized	$h D_p/k = 0.018 (D_p G/\mu)^{1.2}$	0.8-2.0
(8) Sunkoori et. al.	Unsteady state liquid fluidized	$h D_p/k = 0.00391 (D_p G/\mu)^{2.1}$	110-620

numerous. But most of the studies of fluidized heat transfer, so far reported are only concerned with heat transfer from an exposed surface to a fluidized medium and suitable empirical correlations are available for predicting the heat transfer rates under conditions similar to those reported in the various investigations. Only a few studies consider the heat transfer from the fluidized particles to the fluidizing medium. But this study becomes essential, especially for the rational design of catalytic reactors.

The problem of the proper temperature difference is the source of much of the difficulty with particle-to-fluid heat transfer in fluidized beds. In systems other than fluidized beds, the problem is usually comparatively simple; the concept of a laminar zone near the surface of a particle and a single relatively homogeneous turbulent zone, next to that (or a well defined laminar region in its place) serves to handle the majority of cases found in practice. In a fluidized bed, however, it appears that the situation is rarely simple. There is good evidence that, especially in gas fluidized beds, the fluid in excess of that required to just fluidize the bed passes through the bed as a particle-pour dispersed phase in a matrix of a "dense" phase which is continuous and which contains the bulk of the particles. The amount of exchange of particles and fluid between the two phases is, for the most part not known, difficult to measure, impossible to guess and appreciably affected by numerous common perturbations in fluidized bed parameters. There are, however the data which must be available before a genuine point coefficient can be determined.

Physical mechanism

The heat transfer from fluidized particles to the surrounding medium depends primarily on the flow around the individual particles. As for spherical particles, as a first approximation of heat transfer coefficients might be made by the equation already derived for the previous case. For this, Reynolds number is based on the particle diameter and free stream velocity past the sphere. By assuming uniform distribution of particles, the velocity past the particle (V_p) may be calculated by

$$V_p = V_o \left\{ 1 / [1 - (1 - \epsilon)^{0.86} \cdot \pi^{0.33} (0.75)^{0.86}] \right\} \\ = F_e V_o \quad (6)$$

where, F_e is the velocity correlation factor. The velocity correlation factor F_e can be calculated from the experimental values of bed voidage-at various superficial fluid velocities.

Particle interactions play a large role in determining the heat transfer. The frequency with which the particles strike each other, as well as their velocity, influences the degree to which the boundary layers on the particles are distributed. The interaction velocity

is dependent on particle density in the fluidized (medium) column. This particle density in turn is dependent on density of the solid material and the particle diameter. These three parameters i.e. u , p_p and D_p can be expressed in the dimensionless forms

$$\mu/\mu_o, \rho_p/\rho_f, D_p/D_T$$

which are useful in correlating experimental data, where u_{z0} is some suitable reference value for viscosity.

Investigations

Heat transfer from single spheres has been reported by Kramers¹² Vliet and Lappert²² have extended the data previously available and reported the following relation for flow of water and oil over spheres in Reynolds number from 1 to 50,000.

$$N_u (Pr)^{-0.3} (\mu_w/\mu)^{0.25} = 1.2 + 0.53 Re^{0.54} \quad (7)$$

It is suggested that this relation can be extrapolated to Reynolds number as high as 3×10^5 . Van Heerden et. al.^{20,21} suggested the importance of the particle convective mechanism of heat transfer in gas fluidized beds. This was substantiated experimentally by Ziegler et al.²⁷ who found that 80 to 95% of the heat transfer was by particle convection. They presented a model for particle convective heat transfer in fluidized beds.

Extensive study was made by Holman, Moore and Wong.⁹ Stainless steel and lead spheres were fluidized in water and heated by an induction heating field. Reynolds number based on particle diameter and superficial velocity particles to the water was correlated with:

$$N_u = 1.92 \times 10^{-5} (Re Fe)^{2.0} (Pr)^{0.67} \\ (D_T/D_p)^{0.5} (\rho_f/\rho_p)^2 (\mu/\mu_o)^{0.83} \quad (8)$$

The velocity correlation factor, F_e , was introduced to account for variations in porosity.

Effectiveness factor for heat transfer

Information dealing with the transfer of heat from solid particles in the fluidized state to a gas or a liquid flowing past then varies widely, and correlation of the generalised type for heat transfer coefficients are not very common. Lack of progress may be due to experimental difficulties encountered in the evaluation of the proper temperature difference and transfer area associated with such fluidized systems. In addition heat transfer in fluidized system is complicated by the existence and generation of bubbles which produce a non-ideal flow pattern within the bed. These non-ideal conditions give rise to the introduction of effectiveness factors to the conventional heat transfer equations as applied to packed bed conditions.

Petrovic and Thodos¹⁵ have introduced effectiveness factors for the above and have given the heat transfer equation as:

where ' h_{gb} ' is the heat transfer coefficient at bubble point (bubble point mass velocity has been found to be equal to 30% above the minimum fluidization mass velocity). Subsequently this has been correlated to the heat transfer coefficient in an actual fluidized bed. Also

$$q = h_{gb} \chi a v (\Delta t)_m \quad (10)$$

where, χ is the overall effectiveness factor.

A normalised heat transfer factor is defined as:

$$g = (JH)_f / (JH)_b \quad (11)$$

The ratio defines 'g' as,

$$g = (h_{gf} / h_{gb}) \cdot (G_b / G_f) \quad (12)$$

$$\text{or, } h_{gf} = g \cdot h_{gb} \cdot (G_f / G_b) \quad (13)$$

For fluidized bed,

$$\begin{aligned} q &= h_{gf} a v (\Delta t)_m \\ &= g \cdot h_{gb} G_f / G_b a v (\Delta t)_m \end{aligned} \quad (14)$$

Since,

$$\begin{aligned} (G_f / G_b) &= (Re_f / Re_b) \\ q &= h_{gb} g \cdot (Re_f / Re_b) \cdot a v (\Delta t)_m \end{aligned} \quad (15)$$

From equations (10) and (15)

$$\chi = g (Re_f / Re_b) \quad (16)$$

The values of g and R (Re_f / Re_b) corresponding to fluidized runs have been calculated and correlated as

$$g = 1.0 / R^{1.13} \quad (17)$$

which gives,

$$\chi = g^{0.115}$$

The authors have also correlated heat transfer factor to porosity as:

$$g = 1.7 / (\epsilon_t / \epsilon_p)^{0.73} \quad (19)$$

$$\text{for, } \epsilon_t / \epsilon_p > 1.22$$

$$\text{For, } \epsilon_t / \epsilon_p < 1.22,$$

'g' can be taken as unity combining equations (18) and (19).

$$\chi = 1.063 / (\epsilon_t / \epsilon_p)^{0.314} \quad (20)$$

$$\text{for } \epsilon_t / \epsilon_p > 1.22.$$

From a knowledge of the fluidized bed conditions it is possible to evaluate the effective heat transfer coefficient for the bed, (h_{gt}).

WALL-TO-FLUID HEAT TRANSFER

Physical mechanism

The primary resistance to heat transfer occurs in the thin layer of fluid at the wall of the heated column, and substantially all of the radial temperature drop between wall and fluid occurs in this region. In the central portion of the fluidized column the radial temperature distribution is very uniform. The heat transfer coefficient for fluidized system is substantially higher than for flow systems without the presence of fluidized solids, and the higher values may be explained by the fact that the solids scrub the wall and disturb the laminar sublayer to such an extent that its thickness is decreased or at least there is an injection of turbulence into the layer which decreases the thermal resistance and brings about a high heat transfer rate. The specific heat of solids has been found to affect the heat transfer rate from the heated wall to the fluidized systems. Particles with high specific heats may transport energy from the fluid region near the wall more readily and consequently bring about higher values of heat transfer coefficient.

As the heat transfer characteristics of the fluidized systems are observed over a range of porosity a maximum in the heat transfer coefficient is experienced. After the initial fluidization, the heat transfer coefficient increases with an increase in the mass velocity of the fluid (and corresponding increase in porosity). The coefficient reaches a maximum and then decreases with a further increase in mass flow. The mass velocity corresponding to the maximum heat transfer coefficient has been designated as the dividing line between the so-called dense-phase and dilute-phase fluidization.

An interesting effect in fluidized system has been found out by Frantz⁸ and others. In a heated vertical column containing the fluidized solids, the axial temperature profile of the fluid indicates that practically all the temperature rise in the fluid occurs in a shallow region near the bottom of the column and the temperature remains essentially constant above this point. This shallow region is called the "active section". Lemlich and Caldas¹³ report that this effect is more pronounced for high Reynolds numbers.

A casual inspection of these results could lead to the conclusion that practically all of the heat transfer occurs in the region near the bottom of the column, even though the entire column is heated. This does not account for the secondary flow regions which may be present in such fluidized systems. Since the solid particles are maintained in suspension by the vertical fluid drag forces, the particle motion will vary with radial distance from the wall of the tube. In the central portion of the tube, velocity is largest and hence the fluid exerts a force on the particles sufficiently strong to cause them to flow upward. Near

the tube wall the fluid velocities are smaller as a result of the viscous action at the boundary and are not large enough to maintain the particle suspension; consequently they move downward in the column. Thus there is an upward motion in the centre of the column and downward near the wall.

The particles near the wall receive energy from the wall and transport it downward into the active section, where it is distributed uniformly among the particles which flow upward. Thus the temperature of the fluid measured at the axis of the flow may under certain conditions, show no substantial increase at points above the active section.

Investigations

Wall-to-fluid heat transfer coefficients are generally very high. Two mechanisms have been proposed to account for the increased heat transfer rates. First, it is suggested that the particles stir the fluid in the bed and thereby decrease the thermal resistance of the film at the heat transfer surface, and secondly the particles transfer heat themselves as they move in and out of the thermal boundary layer at the wall.

Heat transfer coefficients between the wall and a liquid fluidized bed have been measured by Lemlich and Caldas¹³, Richardson and Mitson¹⁷, Jagannadharaju¹⁰ and Richardson and Smith¹⁸. These workers found that the heat transfer coefficients increased markedly as the porosity was decreased from unity, and reached a maximum at a porosity of about 0.75. The coefficients then decreased steadily as the porosity was further reduced to fixed bed condition. All of the workers, except Caldas et al.¹³ reported that the heat transfer coefficients increased with increasing particle size. They also observed that the porosity at which maximum value of heat transfer coefficients occur decreases as the particle size is increased.

Richardson and Smith¹⁸ examined the effect of the thermal properties of the solids upon the heat transfer coefficients by using solids of widely varying properties; glass, iron, copper, lead and gravel. Their empirical correlation of h is:

' h ' is wall-to-fluid heat transfer coefficient Btu/hr. ft.² °F. The above equation places considerable emphasis on the specific heat of solids, thus suggesting that heat transport by particle convection is quite important. The assumption that the resistance to heat transfer from the wall to a gas fluidized bed is confined to a region very close to the wall seems to be justified since the experimental measurements of

several investigators show that the radial temperature profile is extremely steep at the wall and virtually flat across the bulk of the bed. Mickley and Fairbanks¹⁴ studied the nature of the resistance controlling heat transfer between the fluidized beds and surfaces in contact with them. They observed that unsteady state diffusion of heat into mobile elements of quiescent bed material to be the controlling factor for heat transfer. Further they noted that the heat transfer coefficients are proportional to the square root of the thermal conductivity of the quiescent bed.

Wasmund and Smith²⁵ developed a model for predicting the rate of heat transfer by particle convection in liquid fluidized beds which was based on radial temperature profiles and particle velocities. They reasoned that a particle entering the thermal boundary layer at the wall absorbed heat from its surroundings proportional to the temperature driving force which varies continuously as the particle approaches and departs from the wall. They obtained solutions for two cases, (i) the thermal conductivity of the particles is large, hence the film surrounding the particles is the controlling resistance (ii) the thermal conductivity of the particles is small, and as a result its resistance must also be considered. Their solution for the particle convective heat transfer coefficient, h_0 for case (i) is:

$$h_0 = \frac{(1-\epsilon) v_r \rho_p C_p (e^{-bc} - 1)^2}{2 bc}$$

where,

$$c = 6 h_p / d_p \rho_p C_p v_r$$

b = thickness of the thermal boundary layer at wall.

Later calculations, based on the temperature profiles of the present investigation, indicate that particle convective mechanism is considerably less important in liquid fluidized beds.

Later Wasmund and Smith²⁶ studied wall-to-fluid heat transfer with an aim to investigate the effect of the thermal conductivity of the solids on transfer coefficients.

All the fluidized bed heat transfer experiments were conducted at a constant bed height of 25-in. The porosity was adjusted by adding known weights of the particles through the spout in the expansion section. Porosities were varied from 0.95 to 0.45, the latter being minimum fluidization porosity. The water temperature at the inlet to the test section WHS maintained at $77 \pm 0.5^\circ\text{F}$ for all runs. The temperature driving force, $(T_W - T_B)$, was kept small for all runs in the order of 5 to 10°F , to minimize changes in the physical properties of the water. Heat transfer coefficients were calculated from the equation—

$$h_w = (q_w / (T_W - T_B)) \quad (22)$$

where, $(T_w - T_B)$ is the temperature difference of the wall and the bulk stream measured above the bottom of the heating section.

Typical results of the heat transfer experiments in glass and aluminium systems were reported, where the coefficients are plotted as functions of porosity. Since the porosity is a unique function of the velocity for any given sphere size and solids density the heat transfer coefficients for each system are represented by curves in a single plot.

The heat transfer curves for each particulate system passed $h_{\max}$ and the porosity at which the maximum occurs decreases with increasing particle size; an effect which was also observed by Richardson and Smith¹⁸ and Ruckenstein et. al. The maximum values of h , for the glass and aluminium systems were correlated by the following equation which fitted the data well for $D_p > 0.02''$.

The observation that $h_{\max}$ increases with particle size is in agreement with all of the previous investigations of liquid fluidized heat transfer except that of Lemlich and Caldas."

Heat transfer coefficients at the wall

Wall coefficients for the fluidized glass system are plotted as Stanton numbers (St_w) vs. porosity on semi-logarithmic co-ordinates. It is observed that Stanton number increases linearly with decreasing porosity for $0.45 < c < 0.90$. For porosities greater than 0.90, Stanton number decreases sharply. It is also noteworthy that at any given porosity, Stanton number increases with decreasing particle size. The following correlations were obtained for the film Stanton number for the glass and aluminium particles.

$$\text{Glass: } St_w = 10^{(-1.1 - 1.75 \epsilon)} Re^{-0.33} \quad (24)$$

$$\text{Aluminium } St_w = 10^{(0.29 - 1.75 \epsilon)} Re^{-0.37} \quad (25)$$

This shows that Stanton number for aluminium is larger than for glass particles. It is clear that solids thermal conductivity also influences the heat transfer mechanism at the wall, particularly at low voidages. This is reasonable because the temperature gradient at the wall is extremely steep, and as a result heat conduction through the particles is enhanced by the large temperature drop across the particles.

Conclusions

(1) Heat transport by particle convection is insignificant in the process of wall-to-fluid heat transfer in liquid fluidized beds, except possibly at low porosities. In this regard fluidization by gases and liquids are distinctively different because particle convection is the predominant mechanism of heat transfer in gas fluidized beds.

(2) Fluid convection is the main mechanism of heat transfer in liquid fluidized beds; particle convection and conduction are of minor importance at low porosities.

(3) The resistance to heat transfer in a liquid fluidized bed shifts progressively from the region near the pipe wall to the bulk of the bed as the porosity is decreased from unity.

(4) Effective thermal diffusivities in the bed, calculated from the radial temperature profile decrease nearly linearly with porosity for $E_r < 0.9$.

(5) A series mechanism of heat transfer suggest a correlation of the form:

$$1/St = 1/St_w + UD_T/8E_r \quad (26)$$

This correlation shows a dependence on column diameter which should be tested experimentally. This correlation also correlates wall-to-fluid heat transfer coefficients obtained in the water fluidized beds of glass and aluminium spheres with an average deviation of less than 10% for particle diameter ranging from 0.112" to 0.02" in the porosities ranging from 0.9 to 0.45.

Industrial applications

One of the novel characteristics of fluidized beds is the uniformity of temperature found throughout the system. Essentially constant temperature is known to exist in both horizontal and vertical directions in beds from 1" to 30' in diameter. Temperature variations occur in some beds near zones where large quantities of relatively hot or cold particles are injected; however, these temperature variations are generally small. The isothermal conditions found in fluid beds make them ideal for many catalytic reactions, since yield and selectivity of the catalyst depend to a large extent on both temperature level and uniformity. Thus an important application of fluidized bed heat transfer is in the design of commercial catalytic reactors. Especially the particle-fluid heat transfer study finds its use for the rational design of catalytic reactors as it has been recognised that, the prediction of particle-fluid heat or mass transfer coefficients is important in the course of the estimation of local or overall conversions. It is also necessary to know these heat transfer rates in order to design a fluidized nuclear reactor. With this end in view experimental study was initiated by Holman et. al⁹ to determine the heat transfer coefficients for lead spheres fluidized in water and it is anticipated that the data will be applicable to natural-uranium water fluidized nuclear reactor.

Another important industrial application of the fluidized bed heat transfer is the fluidized bed drying. Of late, this fluidized bed technique has met with wide and growing acceptance in chemical and allied industries for drying of relatively free flowing granu-

lar materials in a wet state. Conventional dryers like, continuous through-circulation type and pneumatic conveying type were earlier performing similar duties and are still preferred for some specific situations. But, fluidized bed dryers are holding the ground firmly now-a-days because they command several selective advantages over conventional types. Compact unitised design, intimate gas-solid contact resulting in uniform material treatment with a wide range of materials, close temperature control minimum operating attention and maintenance costs and high thermal efficiency are among the proved advantages.

Scope

The influence of particle shape and void fraction have not been extensively studied and incorporated in the general expressions for heat transfer coefficient for gas-solid as well as liquid solid systems. More work is needed in liquid fluidized bed to carry out the studies of the influence of Prandtl numbers.

NOMENCLATURE

a	= Interfacial area of packing, L^{-1}
A	= Area of fluidized bed participating in heat transfer, L^2
C_p	= Heat capacity, QMT^{-1}
D_p	= Diameter of particle, L
D_T	= Diameter of column, L
E_r	= Radial thermal diffusivity, $L^2 \theta^{-1}$
Fe	= Velocity correlation factor, dimensionless
f_a	= Effectiveness factor for area dimensionless
f_b	= Effectiveness factor for heat transfer coefficient, dimensionless
f_t	= Effectiveness factor for temperature coefficient dimensionless
G	= Superficial mass velocity, $M \theta^{-1} L^{-2}$
G_b	= Superficial mass velocity, at bubble point, $M \theta^{-1} L^{-2}$
G_f	= Superficial mass velocity of fluidized bed, $M \theta^{-1} L^{-2}$
g	= Normalised heat transfer factor, dimensionless
h_o	= Particle convection heat transfer coefficient, $Q \theta^{-1} L^{-2} T^{-1}$
h_{gb}	= Heat transfer coefficient at bubble point, $Q \theta^{-1} L^{-2} T^{-1}$
h_{gf}	= Heat transfer coefficient of fluidized bed, $Q \theta^{-1} L^{-2} T^{-1}$
h_{max}	= Maximum value of the heat transfer coefficient in fluidized bed, $Q \theta^{-1} L^{-2} T^{-1}$
h_p	= Particle-to-fluid heat transfer coefficient in fluidized bed, $Q \theta^{-1} L^{-2} T^{-1}$

h_w	= Wall-to-fluid heat transfer coefficient in fluidized bed, $Q \theta^{-1} L^{-2} T^{-1}$
JH	= Heat transfer factor ($h/C_p G$) ($C_p \mu/k$) ^{0.66}
$(JH)_b$	= Heat transfer factor at bubble point
$(JH)_f$	= Heat transfer factor of fluidized bed
K (k_i or k_f)	= Thermal conductivity of the fluid $Q \theta^{-1} L^{-1} T^{-1}$
L	= Length of column, L
L_B	= Height of bed, L
N_u	= Nusselt number, $h D_p/k$, dimensionless
Pr	= Prandtl number, $C_p \mu/k$ dimensionless
q	= Rate of heat transfer in the bed, $Q \theta^{-1}$
q_w	= Rate of heat input to the wall (as calculated from electrical power consumption), $Q \theta^{-1}$
Re	= Reynolds number, $D_p G/\mu$, dimensionless
$(Re)_b$	= Reynolds number, at bubble point, dimensionless
$(Re)_f$	= Reynolds number for fluidized bed, dimensionless
R	= Normalised Reynolds number, $(Re)_f/(Re)_b$
St_w	= Stanton number with respect to column wall, dimensionless
T_w	= Wall temperature, T
T_B	= Bulk stream temperature, T
$(\Delta t)_m$	= Log mean temperature difference, T
U	= Average superficial velocity, $L\theta^{-1}$
U_o	= Velocity obtained by extrapolating Richardson-Kaki. Curves to unity porosity, $L\theta^{-1}$
U_p	= Average particle velocity, $L\theta^{-1}$
V_p	= Free stream velocity past the particles, $L\theta^{-1}$
V_o	= Superficial velocity in the tube (velocity at 100% porosity) $L\theta^{-1}$
V_r	= Average velocity of particles in radial direction, $L\theta^{-1}$
V	= Volume of the bed, L^3
x_i	= Point film thickness of fluid, L

GREEK LETTERS

ϵ	= Void fraction
ϵ_p	= Void fraction of packed bed
ϵ_f	= Void fraction of fluidized bed
δ	= Particle roughness
ψ	= Particle shape factor
ρ_f (ρ)	= Fluid density, ML^{-3}
μ (μ_f)	= Fluid viscosity, $ML^{-1} \theta^{-1}$
χ	= Overall effectiveness factor for heat transfer, $f_a \cdot f_t \cdot f_b$

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