總 說
REVIEW

기체유동상과 수평관 사이의 열전달

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Heat Transfer between Gas Fluidized Bed and Vertical Tubes

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要 約

유동층과 유동층내에 삽입된 수직 열전달관 사이의 발표된 총괄열전달계수식들을 종합하였다. 유체의 속도, 입자의 크기, 입자의 농도, 입자와 유체의 열물리적 특성, 수직 열전달관의 위치, 유동층의 압력, 충진물의 특성 등이 총괄 열전달계수에 미치는 영향을 검토하였다. 열전달의 기본과정을 이해하기 위하여 유동층과 유동층 벽사이의 열전달에 관하여 조사하였다. 이제까지 보고된 유동층과 유동층내의 수직관 사이의 총괄 열전달계수식들을 조사 비교한 결과 Tamarin과 Khasanov의 실험식이 가장 유용하게 사용될 수 있음을 알았다.

Abstract

Overall heat transfer coefficient correlations are given for fluidized bed in which vertical tubes are immersed. Correlations available in the literature are surveyed and compared on a consistant basis. Effects of mass flow rate, particle size, solid concentration, solid and fluid thermophysical properties, immersed tube location, bed pressure and packing materials on heat transfer coefficients were summarized. In order to understand the fundamental mechanisms of heat transfer, wall to fluidized bed heat transfer is also reviewed. Those correlations on overall heat transfer coefficients are evaluated and find that the most appropriate one for predicting overall heat transfer coefficients between fluidized bed and vertically immersed tubes is one of Tamarin and Khasanov.

Introduction

A large fluidized bed boilers for the efficient production of energy is currently considerable renewed interest. This necessitates not only an optimum economic design but also a design which incorporates the essential features necessary for gaseous and solid pollutants control. The development of fluidized bed boilers for the combustion of sulphur containing coals has particular promise. The sulphur dioxide produced during combustion is retained by an additive such as dolomite and limestone. Steam is generated within the boiler tubes immersed into the fluidized bed. The steam is used to drive turbines for electric generation. Consideration has also been given to using the hot flue gases along with the steam to drive a combined cycle of steam and gas turbines.

The heat transfer mechanism is very complicated because of the so many fluidized bed variables (i.e. particle shape, solids and gas thermal properties, particle size distribution, reactor geometry and type of gas distributor) and the variation in heat transfer tube design such as shape, spacing, pitch and material. Correlations were empirically developed on the basis of the observed data with limited attention to the mechanisms of heat transfer process in larboratory scale fluidized beds. Therefore, the scope of these correlations are restricted to the data for the specific test conditions. In this review, the information that does exist will be examined and evaluated.

I. Wall-to-bed heat transfer

However, before this work is appraised it is best to review what has been learned about the fundamental mechanisms of wall-to-bed heat transfer and to establish the criteria for evaluation of correlations for full scale application. There are a good number of reviews of basic studies on fluidized bed heat transfer [1)~9)].

The significant mechanism for modeling wall-to-bed heat transfer for vertical walls are essentially:

1. A fluid film is the principal resistance to heat transfer, and the moving fluidized particles scour the film to reduce the resistance to heat transfer [9)~12)]. This model is not supported by the experimental work of Van Heerden et al.[13)] and of Ziegler and Brazelton [14)]. They observed that absorption of heat by the fluidized particles rather than reduction of the film resistance was the principal mechanism for heat removal at a surface.

2. Heat is absorbed by packets of particles having a finite residence time on the heater surface[15)]. The rate of heat transfer is evaluated by using an effective bed thermal conductivity obtained from steady-state measurements. However, data reported by Harakas and Beatty[16)], by Ennst[17)], by Botterill et al. [18)], and by Gabor[19)] show that measured heat transfer coefficients for short solids residence times are much less than those predicted by this model. Since the thermal response is different in the gas and solid phases, transient heat transfer in a gas-solids bed can not be treated as heat transfer in a homogeneous medium as implied by using an effective thermal conductivity.

The Mickley-Fairbanks model has been modified by the introduction of a contact resistance [20)~23),25)]. The contact resistance at the bed-wall interface is attributed to various effects, e.g. Kubie and Broughton[25)] modeled voidage variations in the bed near the wall, and Wunschmann and Schlunden[24)] considered the reduced free path of the gas molecules.

3. Heat transfer is modelled by unsteady-state conduction from the heater surface to a single particle [26)~28)], and to depts of two particles [18)], four particles [29)], and a particle chain of essentially unlimited length[30),31)]. This type of model analysis is made of the heat as it transferred to distinct particles rather than by treating the bulk of the bed as a homogeneous medium using average bed properties and an effective thermal conductivity. Both types of models are based on heat adsorption by the particles.

The chain-of-particles modelwas simplified by approximating the gas particle bed by a series of alternate gas and solid slabs. In this model the primary resistance to heat flow is in the gas phase and the solid phase absorbs heat. Therefore, the heat transfer coefficient is primarily dependent on the gas thermal conductivity and the particle heat capacity.

The model based upon a chain of particles of unlimited length compared well with the data for both short and long particle residence times obtained by Harakas and Beatty[16)], Botterill et al.[18)] and Gabor.[19)] The data included a wide range of particle sizes, particle heat transfer properties, and gas heat transfer properties. This chain of particles model coincides with the Mickley-Fairbanks model at long particle residence times.

These basic studies on wall-to-bed heat transfer provide criteria for evaluating empirically derived heat transfer correlations for tubes in fluidized beds. The significant points are as follow;

a) The principal mode of heat removal is by fluidized solids heat adsorption. Particle volumetric heat capacities are of the order of 1,000 times those of gases [32)]. In that sense, a generalized heat transfer correlation must contain a $\rho_s C_{ps}$ term(Volumetric heat capacity of the particle). The $\rho_f C_{pf}$ term can be neglected in the analysis of gas-solids systems.

b) The rate of heat transfer depends on the particle residence time. The factors which affect the particle residence time must be considered. For gas fluidized beds the residence times are dependent on the bubble mechanics [33),34)] as well as the particle drag forces.[35)] The particle movement is caused by the rising bubbles, and during the period in which a gas bubble resides on the heat transfer surface heat transfer to the bed is negligible. The two-phase theory for gas fluidized beds assumes that all gas in excess of that required for minimum fluidization forms bubbles.[36)] Of course the particle residence time will depend upon internal geometry such as tube size, shape and orientation.

c) Heat is conducted from the wall and between the adsorbing particles through the interstitial gas phase. There is negligible conduction through the particle contact points [26)]. Therefore, a generalized correlation must contain the gas thermal conductivity, k_f.

With the above three significant points, a generalized correlation should account for at least a) bubble characteristics, b) the system geomerry c) particle drag forces.

It must be additionally noted that the presence of internal heat transfer surfaces will affect the quality of fluidization. The effect of the heat transfer tubes on the fluidization properties must carefully considered in design optimization. Tube bundles, both vertical and horizontal, play an important roll to break up bubbles and limit bubble size. The reaction efficiency of fluidized bed is very much dependent upon the amount of gas by-passing associated with the rising bubbles. The gas flow patterns for isolated undisturbed bubbles in a fluidized bed have been very nicely analysed both empirically and theoretically.[37] However, the baffling caused by reactor internals greatly changes this analysis.

Properties such as bed expansion, particle mixing, elutriation, and entrainment are directly related to the bed bubbling characteristics. It has been known as small bubbles will have a lower rise velocity. Therefore, heat transfer tubes within the bed will causes a greater bed expansion and lower particle mixing rates than for a fluidized bed without internals. Fine particles leave the bed when the bubbles burst at the surface. Smaller bubbles will burst less violently and result in lower elutriation rates.[38)~39]

It is therefore important in the design of heat transfer tube bundles to optimize the overall reacton performance. For instance, wide spacing of the tubes will promote increased solids mixing and higher heat transfer coefficients at each tube surface but conversely gas-solids contacting will decrease because of decreased bubble dispersion and elutriation losses will increase.

Ⅱ. Vertical tube-to-bed heat transfer

Heta transfer coefficients at immersed horizontal surfaces are lower than for vertical-surfaces. Also, vertical tube inserts in reactors of large height-to-diameter ratios not only increase the heat transfer surfaces but also reduces slugging. For these reasons, there is a preference for vertical heat exchanging surfaces. Therefore, a knowledge of the heat transfer characteristics of fluidized bed with vertically immersed surfaces is of practical importance to the design of industrial chemical reactors.

A) Local Heat Transfer Coefficients

A knowledge of local heat transfer coefficients is needed to design the detailed configuration of heating or coolirg tubes in a fluidizer bed reactor.

The distribution of local heat transfer coefficient along the length of a vertical cylinder on a copper ball immersed in a fluidized bed was determined by Berg et al.[40] The difference between maximum and minimum value of h_{wL} was in general, at least 50%. In many cases this difference was two on three times as large. There was no appreciable variation in h_{wL} at various heights from the distributor. At the same fluidizing conditions, the maximum of h_{wL} were more pronounced in the bed of coarse particles than that of fine particles.

Local heat transfer coefficients for a single tube in the presence of other tubes were obtained in four tube positions by Noe and Knudsen.[41] The pattern of the local heat transfer coefficients on the tube was similar to that indicated by the bed density variation.

Genetti and Knudsen[42] investigated local heat transfer from an internal vertical tube in a bundle of tubes to a dilute phase fluidized bed for both batch and solids recycle systems. From measurement of axial and radial temperature profiles, the thermal gradients in the bed were found to be small. Therefore, the coefficients for a fluidized heat exchanger with all the tubes in the bundle heated would not differ substantially from the coefficients for an exchanger with only one tube in the bundle heated.

Mickley et al.[43] determined the local time average heat transfer coefficient, from a vertical tube immersed in a fluidized bed.

Mickley and Fairbanks[15] proposed a packet model on the assumption that the heat transfer process in controlled by unsteady-state conduction from the surface to a packet in contact with it, it following instantaneous local heat transfer coefficient was proposed:

$$h_{wiL}=\frac{1}{\sqrt{\pi}}\sqrt{k_m\,\rho_m\,C_{pm}}\quad\tau^{-\frac{1}{2}}$$

The local time average heat transfer coefficient, h_{wL}, will be

$$h_{wL}=\frac{1}{\sqrt{\pi}}\sqrt{k_m\,\rho_m\,C_{pm}}\int_0^{\infty}\tau^{-\frac{1}{2}}\psi(\tau)\;d\tau$$

The heat transfer rate between the surface and the gas when no packet is in contact with the surface is negligible.

On the basis of the above few investigations, the following summaries can be made on local heat transfer for vertical tubes:

1) For a single vertical tube, the maximum of heat transfer coefficients are more pronounced in the bed of coarse particles than that of fine particles at the same fluidizing conditions.

2) There is no appreciable variation in heat transfer coefficient for a small thermal probe immersed in the bed at various heights from the distributor, therefore, the length of the heated tube affects the values of local heat transfer coefficients.

3) For a bundle of vertical tubes, lowest local coefficients in the dense phase are, generally, observed at the center location and highest values are observed at the tube locations near the wall.

B) Overall Heat Transfer Coefficients

Most of the experimental studies[15,44~49,61,68] have dealt with single heated on cooled tube immersed in a fluidized bed in order to determine overall heat transfer coefficients. In contrast, not much experimental works[41,42,65~67] performed for determing overall heat transfer coefficients between bundle of vertical tubes and a fluidized bed.

The published experimental correlations and variables for overall heat transfer coefficients were summarized in *Table* 1. From which the effect of each experimental and physical properties on the coefficients could be analysed and evaluated.

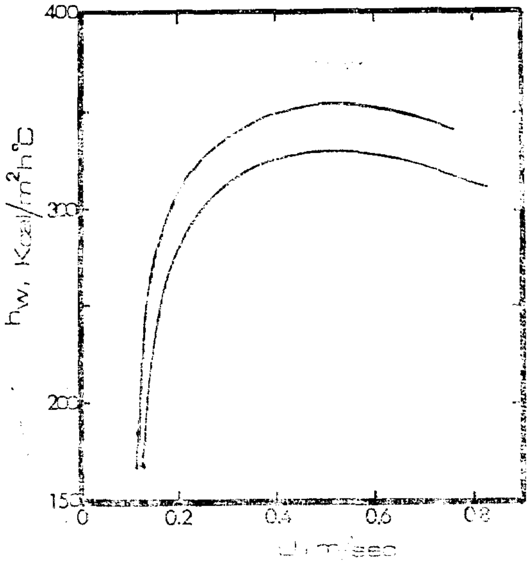

Figure 1. Heat Transfer from Bundles of Vertical Tubes

a) Effect of Mass Flow Rate, Particle Size and Solid Concentration

In general, an increase in fluid velocity above u_{mf} resulted in an initial rapid rise in overall heat transfer coefficient, h_w, followed by a more gradual rate of increase until a maximum value for the coefficient was reached. Further increase in fluid velocities, the coefficients start to decrease due to the decrease of solid concentration in fluidized bed (*Figure* 1). The rise is one to two order of magniture with fluidization by gas and by factor of two to four with fluidization by liquid.

The presence of particles, the coefficients were from three to seventy times greater than that at the same gas flow rates in the absence of particles. This increased heat

Table 1. Summary of Correlations of h_w and Variables

REFERENCE	CORRELATION	D_b, mm	d_p, mm	V_s, m/sec	$D_t \times L_H$ (mm×mm)	REMARKS
Mickley and Trilling (44)	$h_w = 0.0282(\rho_s^2/d_p^3)^{0.238}$	73	40−450 (glass beads)	0.24−4.57 (air)	12.5×876	
Baerg et al. (45)	$h_w = 278.210\,g(7.05\times10^{-3}\rho_s/d_p) - 312.3\ e^{-0.00246(V_g - 0.216\rho_s)}$	140	60−880 (glass beads)	0.0−0.683 (air)	32×254	
Gamson (46)	$h_w = (2.8(6G/A_v u\phi)^{-0.84}(1-\varepsilon)^{-0.3} \times (C_{pf}\mu/k_f)^{-\frac{2}{3}} C_{pf}G) + h_e$	73	40−450 (glass beads)	0.24−4.57 (air)	12.5×876	use the data of (44)
Miller and Logwinuk (47)	$h_w = \dfrac{1.56\,G^{0.5}k_s^{0.072}k_f^{2.4}}{d_p^{0.96}C_{pf}^{1.6}\mu^{0.5}}$	51	96−250 (silicon carbide, Al_2O_3)	0.445−13.6 (air, CO_2, He)	9.5	
Vreedenberg (59)	$h_w = 1.05(GV/G_{mf}V_{mf})^{0.35}(k_f/d_p)$	565	230−597	4.51−11.97 (air)		
Mickley and Fairbanks (15)	$h_w = \sqrt{k_m \rho_m C_{pm} S}$	101	68−81 (glass beads)	0.06−0.55	6.4×609	$S = d_p^{-0.6-1.2}\rho_m^{-1.1}\ u^{small},\ u^{wide}\rho_f^{-0.2-0.2}$ $L_H^{-1.0}\left[\frac{S_2L_H}{V},\ \text{small}\right]$ to $0\left[\frac{S_2L_H}{V},\ \text{large}\right]\cdot L_s^0$ (lowu) −1.0(u>0.3 m/s) other factors
Toomey and Johnstone (49)	$h_w = -420\log d_p - 3065$	120	55−848 (gla s beads)	1.61−104.8	13.0×−	
Wender and Cooper (50)	$h_w = \left[0.033\,C_r\left(\frac{C_{pf}\rho_f}{k_f}\right)^{0.43}\left(\frac{Gd_p}{\mu}\right)^{0.23}\left(\frac{C_{ps}}{C_{pf}}\right)^{0.8}\left(\frac{\rho_s}{\rho_f}\right)^{0.66}\left(\frac{k_f(1-\varepsilon)}{d_p}\right)\right]$	73.1−1929	40−878	6.54−264.5	−X100.5− 8107.7	use data of (15, 44, 45, 49, 51) and of Kellog Hydro-former regenerator

Vreedenberg (56)	*For fine and light particles ($Gd_p\times\rho_s/\rho_f\mu$ <2050) $h_w=0.27\times10^{-5}\left[\frac{G(D_b-D_t)\rho_s}{\rho_f\mu}\right]^{3.4}\left(\frac{D_t}{D_b}\right)^{-\frac{1}{3}}$ $\times\left(\frac{k_f}{C_{ps}\mu}\right)^{-\frac{1}{2}}\left(\frac{k_f}{D_b-D_t}\right)$ for $G(D_b-D_t)$ $\rho_s/\rho_f\mu\leq0.237\times10^6$ $h_w=2.2\left[\frac{G(D_b-D_t)\rho_s}{\rho_f\mu}\right]^{0.44}\left(\frac{D_t}{D_b}\right)^{-\frac{1}{3}}\left(\frac{k_f}{C_{ps}\mu}\right)$ $^{-\frac{1}{2}}\left(\frac{k_f}{D_b-D_t}\right)$ for $G(D_b-D_t)/\rho_f\mu>0.237$ $\times10^6$ *For coarse and large particles($Gd_p\rho_s/\rho_f\mu$ >2550) : $h_w=0.105\times10^{-3}\left(\frac{G(D_b-D_t)}{\rho_f d_p^{\frac{3}{2}}g^{\frac{1}{2}}}\right)^{2\cdot}\left(\frac{D_t d_p k_f}{D_b(D_b-D_t)}\right.$ $\left.C_{ps}\mu\right)^{-\frac{1}{3}}\left(\frac{k_f}{D_b-D_t}\right)$ for $G(D_b-D_t)/\rho_f$ $d_p^{\frac{3}{2}}g^{\frac{1}{2}}<1020$ $h_w=240\left[\frac{G(D_b-D_t)}{\rho_f d_p^{\frac{3}{2}}g^{\frac{1}{2}}}\right]^{-0.1}\left[\frac{D_t d_p k_f}{D_b(D_b-D_t)C_{ps}}\right]$ $^{-\frac{1}{3}}\times\left(\frac{k_f}{D_b-D_t}\right)$ for $G(D_b-D_t)/\rho_f d_p^{\frac{3}{2}}g$ $^{\frac{1}{2}}\geq1070$	51−565	40−900	—	6.4−33.8 × 33−876	used data of (15, 44, 45, 47, 49, 51, 59, 68-71)
Shlapkova (57)	$h_{w,\max}=2\rho_s^{0.2}D_t^{-0.11}d_p^{0.4}$ $h_{wp,\max}=h_{w,\max}/(1+100K_n)^{0.45}$	48	100−772 (glass beads & sand)		6, 8, 3.4X−	*Effect of bed pressure

Gabor (60)	$h_w = \frac{4[k_e^0 + 0.1(C_{ps}d_pG)C_{ps}G]}{L_H} + \frac{1/2k_e^0 + 0.1(C_{ps}d_pG)}{r_2}$	101.6	3.2−34.4 (glass, Cu, Al, cellulose acetate, Al_2O_3)	0.076−1.106 (air)	28.6×50.8, 101.6, 203.2	$k_e^0 = 0.9065\left[\frac{2}{\left(\frac{1}{k_f}\right)-\left(\frac{1}{k_s}\right)} \frac{k_s}{k_s-k_f}\ln\left(\frac{k_s}{k_f}\right)-1\right]+$ $0.0935\,k_f$
Chernov et al. (68)	$h_w = \frac{(1-\varepsilon)/(1-\varepsilon_{mf})}{R_k + 0.5\sqrt{\frac{\pi}{k_mC_{pm}\rho_m}}\sqrt{\frac{1-\varepsilon}{1-\varepsilon_{mf}}(u-u_{mf})}} \times \frac{1-\varepsilon}{1-\varepsilon_{mf}}M$	500.0	60−800	0.04−0.32 (air)	9×200	*Effect of packing R_k; thermal resistance
Noe and Knudsen (41)	$h_w = 0.066\left(\frac{C_{pf}\rho_f}{k_f}\right)^{0.43}\left(\frac{Gd_p}{\mu}\right)^{0.23}\left(\frac{C_{ps}}{C_{pf}}\right)^{0.8}\left(\frac{\rho_s}{\rho_f}\right)^{0.66}$	152 & 1118	145−477	54.38−203.95		Bundle of tubes
Genetti and Knudsen (42)	$h_w = \frac{5(1-\varepsilon)^{0.48}\phi k_f}{\left[1+\frac{580}{Re_p}\left(\frac{k_f}{d_p^{\frac{3}{2}}C_{ps}\rho_s g^{\frac{1}{3}}}\right)\left(\frac{\rho_s}{\rho_g}\right)^{1.1}\left(\frac{G_{mf}}{G}\right)^{\frac{7}{3}}\right]^2 d_p}$	152 & 1118	132−780 (glass beads & aluminum)	46.22−583.31		
Maskaev and Baskakov (67)	$h_w = 0.21\left[\frac{gd_p^3}{\nu^2}\ \frac{\rho_s-\rho_f}{\rho_f}\right]^{0.32}\left(\frac{k_f}{d_p}\right)$		2000−12900			
Tamarin and Khasanov (65)	$h_w = 0.32\left[\frac{gd_p^3}{\nu^2}\ \frac{\rho_s-\rho_f}{\rho_f}\right]^{0.25}\left(\frac{d_p}{l}\right)^{0.07}\left(\frac{k_f}{d_p}\right)$	150 & 300	90−760 (sand. silicagel, Corundum.)	0.05−1.25	20x−	

transfer rate may be due to the heat transport by the solids from hot to the cold region in the gas phase.

From the published investigations [15),42),44),45),47),49),50),65),67)] and the *Table* 1, it may universally concluded that the coefficients decreased with particle size at a given fluidized bed and heater and the same gas and solid characteristics. For instance, Mickley and Trilling [44)] found that $h_w \propto d_p^{-0.6}$, Baerg et al.[45)] found that $h_w \propto d_p^{-0.3}$ and Miller and Logwinuk [47)] found that $h_w \propto d_p^{-0.96}$. The usual explanation for this effects is due to the decrease of gas film thickness which offers appreciable resistance to the conduction of heat between solids and heater. Another explanation has been given by Mickley and Fairbanks[15)] is that the particle size effect is one of bed motion, i.e. beds of smaller particles circulate fresh solid to contacting walls more rapidely at given gas flow rate. The rounded and smoother particles pronounced somewhat higher h_w values than that of coarser particles.[45)]

Since solid concentration could be changed with mass flowrate, it may not be an independent effect on overall heat transfer coefficient. However, Mickley and Trilling [44)] found that $h_w \propto \rho_m^{0.48}$, and Mickley and Fairbanks[15)] found that $h_w \propto \rho_m^{0.5}$. In contrast, Wender and Cooper[50)] found that the heat transfer coefficient varied as the 1.0 power of solid concentration. Same trend was also found by Chernov et al.[68)] It may imply that dense phase material in contact with the average fraction of the heated surface is directly proportional to the solid concentration.

b) Effect of the Thermophysical Properties of Solids and Fluid.

The thermal fluid conductivity, k_f, has the greatest influence on heat transfer [47),50),55),56),59),60)]. The heat transfer coefficients varied as the 0.5 to 2.4 power of k_f. The increase of h_w with fluidized bed temperature largely resulted from the increase of k_f with bed temperature and not just from increased thermal radiation.

The specific heat of solids. C_{ps}, has positive effect on heat transfer. The heat transfer coefficients increased with C_{ps} as the 0.5 to 1.0 power of C_{ps}[11),15),50),68)].

The thermal conductivity of solids, k_s, has practically very little effect on h_w[42),47)]. The thermal conductivity of aluminum was 500 times larger than that of glass yet no increase in heat transfer was observed in the beds of aluminum particles.[42)]

The effect of fluid specific heat, C_{pf}, on heat transfer are more contradictory. The volumetric heat capacity of solids are three orders of magnitude greated than that of fulid at moderate pressures. Therefore, the volumetric heat capacity of fluid, $C_{pf}\rho_f$, can not play an important role for heat transfer in gas-solids systems.

c) Effect of the tube location in the bed

The effect of tube location on h_w using a vertical tube at three locations (0~200 mm) has been studied by Vreedenberg[59)]. It has been found that there is a significant increase in h_w on nearing the center line of the bed and some reduction in h_w at the center line. His data is shown in *Figure* 2. The ratio of $h_w/h_{w,\text{axial}}$ was designated as a correction factor C_r for nonaxial tube location. The low value of C_r in the axis position may be due to the highest number of air bubbles in the center of the bed. The decrease of the coefficient as the distance increased from 100 to 200 mm may be caused by the wall effect. Wender and Cooper[50)] used the correction factor for nonaxial tube location in their correlation for h_w. In contrast, Genetti and Knudsen[42] claimed that h_w did not vary appreciably with tube location. The effect of tube location on h_w was a function of mass fluidizing air velocity. At low mass velocity ($G<3418$ kg/hr m^2), it appeared that the tube location would significantly affected h_w. However, at higher mass velocities ($G>4883$ kg/hr m^2) the tube location affected h_w only slightly.

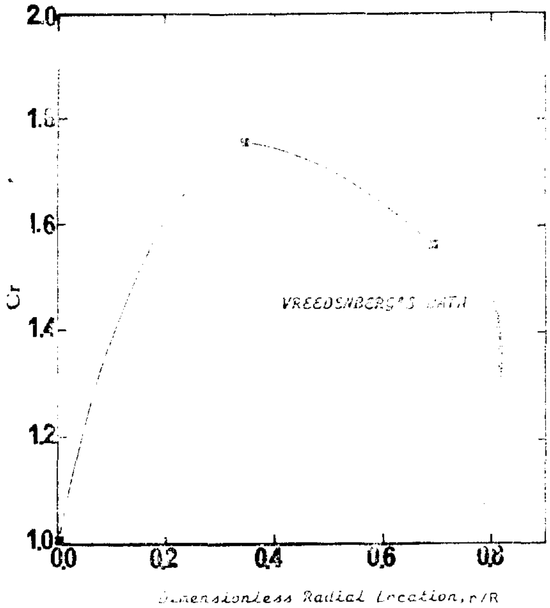

Figure 2. Correction for Nonaxial Location

d) Effect of Bed Pressure

The effect of pressure on heat transfer was investigated by Shlapkova[58]. The maximum heat transfer coefficient increased with increse from low to atomospheric pressure. Starting at a pressure of 4000 N/m^2 and up to atomospheric pressure, the heat transfer coefficient remained practically constant. The same effect was also observed by Mickley and Fairbanks[15]. Increasing gas pressure from 101.3 to 13117 KN/m^2 had no noticeable effect on heat transfer when the gas velocity was maintained constant. Vreedenberg[59] also observed same effect of bed pressureon h_w.

e) Effect of Packing Materials

Chernov et al.[68] measured heat transfer coefficient from a vertical heated tube to a fluidized bed containing cylindrical helical packing. The heat transfer coefficient increased smoothly with increase in air fluidizing velocity ($u/u_{mf}=0$ to 8) for fluidized bed with packing. For air velocities, $u/u_{mf}>4$, the value of h_w in the beds with packing were larger than the $h_{w\,max}$ for the bed without packing. Thus the results indicated that the presence of the wire packing did not impair the intensification of the heat transfer rate over a certain range of velocities. It was found that packing size should not exceed the particle size by at least two order of magnitude so that the solid particles are mobile. Packing caused complete or partial disruption of the densification zone and wake and prevented coalescence.

The effect of four kinds of small packing on heat transfer has been studied by Tamarin et al.[66] Tamarin and Khasanov[65] further studied the effect of eight different kinds of packing on heat transfer. They found that maximum heat transfer in a packed bed was attained at higher velocities than that in a bed without packing. The $h_{w\,max}$ differed from one kind of packing to another. The $h_{w\,max}$ decreased with decreasing the bed volume per unit of packing surface area.

Evalaution of correlations

In order to evaluate what correlation would be the most suitable one to predict overall heat transfer coefficients of vertically immersed tubes in a fluidized bed, eight published correlations[44),45),47),49),50),58),60),65] were examined with same experimental variables [64]. The correlations of Baerg et al.[45] and Toomey and Johnstone[49] were excluded since their correlations account only particle size effect. The mean value of the heat transfer coefficient was taken from the published correlations with the same experimental variables and computed standard deviation from the mean of h_w.

In general, the correlations of several investigators[44),50),65] predict h_w values with reasonable accuracy at the given conditions. However, the correlation of Gabor[60] was deviated about 86% from the mean in the h_w range of 50~383 g/m²sec°C. It is not a surprising finding since the operating gas velocities less than requal to that required for minimum fluidization.

The correlation of Mickley and Trilling[44] consists of solid concentration and particle size terms but did not account bed geometric effect. However, their correlation predict h_w values with standard deviation of 21% in the h_w range of 14~150 g/m²sec°C.

Werder and Cooper[50] correlated data for vertical tubes and similar transfer surfaces reported by Mickley and Fairbanks[15], Mickley and Trilling[44], Baerg et al.[45], Toomey and Johnstone[49] and Olin and Dean[51] and Kellog built commercial fluid-hydroformer-catalyst-regenerator bed coolers. Their correlation showed rather high deviation in the h_w range of 56.8~700 g/m²sec°C. In contrast, the standard deviation was only 20% in the h_w range of 700~1100 g/m²sec°C. It may imply that their correlation can be utilized in the higher range of heat transfer coefficients.

As shown in *Table* 1, Vreedenberg[56] proposed four correlations which were derived from the data of 11 investigators [15),44),45),47),48),51)~55)]. Two were for fine and light particles where the viscous force on the particle are predominant in which an analogy between the fluidized bed and a flowing fluid was assumed, and the other two were for coarse and heavier particles where inertia effects prevail. In his correlations, the important term, $\rho_s C_{ps}$, which represents the principal mode of heat removal is neglected. As can be seen the correlations accounts many experimental and operating variables with bed geometric properties. However, the correlations produced rather high standard deviation (−85~150%). The deviation increased with h_w values below 500 g/m²sec°C and higher than 3,300 g/m²sec°C. Therefore, correlations can not be recommended for general use to calculate heat transfer coefficients.

The heat transfer coefficient data between vertical tubes and a fluidized bed with fixed packing were reported by Tamarin and Khasanov[65], Tamarin et al.[66] and Khasanov et al.[67]. As shown in *Table* 1 the correlation of Tamarin and Khasanov[65] consists of Archimedes number, particle size and l is free bed volume to immersed tube or packing surface ratio terms. Therefore, this correlation account bed geometric effect inside the bed. The correlation predict h_w values with comparatively small standard deviation namely 20% at given test conditions(d_p=23−880 μm, ρ_s= 1800−8906 kg/m³, ρ_g=1.30−7.77 kg/m³ and μ_g=0.0000105−0.000038 kg/m sec). Since this correlation has a smallest standard deviation among the tested correlations in this study, the correlation and Khasanov[65] can be recommeded for general use for calculating h_w values for vertical tubes immersed in a fluidized bed reactor.

Conclusions

From the reviews on overall heat transfer coefficients of vertically immersed tubes in a fluidized bed, the following conclusions can be drawn:

1) In general, overall heat transfer coefficients increased with solid concentration, thermal conductivity of fluid and solid, specific heat of solid, density of solid, mass flow rate and equivalent diameter of annulus of fluidized bed and decreased with particle size, specific heat of fluid, viscosity of fluid and density of fluid. However, the effect of mass flow rate was mainly dependent on the range of mass flow rate and of thermal conductivity of solid had minor effects on overall heat transfer coefficients.

2) The most appropriate design correlation for vertical tubes immersed in a fluidized bed can be recommended by this study is one of Tamarin and Khasanov[65] as shown;

$$h_w=0.32\left[\frac{gd_p^3}{\nu^2}\ \frac{\rho_s-\rho_f}{\rho_f}\right]^{0.25}\left(\frac{d_p}{l}\right)^{-0.07}\left(\frac{k_f}{d_p}\right)$$

3) In the bundle of heat transfer tubes, the tube location was a function of mass velocity. The tube location would significantly affecte heat transfer coefficients at low mass velocity. However, at high mass velocities, the tube location affecte h_w only slightly.

4) The bed pressure had minor effect on overall heat transfer coefficients at given mass flow rate.

Nomenclatures

C_{pf} specific heat of fluid at constant pressure
C_{pm} heat capacity of fluidized solid
C_{ps} specific heat of solid
C_r correction for non-axial tube location for cases of internal heat transfer surface
d_p average diameter of particle
D_b diameter of fluidized bed
D_t diameter of the immersed tube
g accerleration of gravity
G mass flow rate at fluidizing condition
G_{mf} mass flow rate at minimum fluidization condition
h_e heat transfer coefficient in the empty tube
h_w overall heat transfer coefficient
$h_{w,\text{axial}}$ overall heat transfer coefficient at axial position in the bed
h_{wiL} instantaneous local heat transfer coefficient
h_{wL} local heat transfer coefficient
$h_{w,\max}$ maximum overall heat transfer coefficient
k_e effective thermal conductivity for zero gas flow
k_f thermal conductivity of fluid
k_m thermal conductivity of quiescent bed
k_s thermal conductivity of solid
K_n Knudson number
l free bed volume to immersed tube or packing surface ratio
L_H heater length
r_2 radial coordirate
R_k contact thermal resistance
s stirring factor
u superficial gas velocity at fluidizing condition
u_{mf} superficial velocity at minimum fluidizing condition
V_g average linear superficial velocity

Greek Symbols

ε void fraction at fluidizing condition
ε_{mf} void fraction of minimum fluidizing condition

μ viscosity of fluidizing gas
ρ_f fluid density
ρ_m solid concentration in fluidizing mixture
ρ_s solid density
τ average residence time of solid particles
ϕ sphericity of particle
ψ a coefficient for bubble pass through a given point
ν kinematic viscosity
ν_{mf} kinematic viscosity at minimum fluidizing condition

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