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Theoretical and Experimental Investigation of Heat Transfer in Packed Beds

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Abstract: The review of heat transfer in packed beds has been carried out with the emphasis on both experimental and theoretical techniques. It is observed that the two major modes of heat transfer, namely, conduction between the particles in the bed and convection between the flowing fluid and the particles, interact with each other. Also, in the designing of packed bed for thermal energy storage application the Biot number should be kept as low as possible so that the thermal resistance within the solid does not become dominant.

Key words: Heat transfer, packed-bed, thermal energy storage, conduction, convection, biot number

INTRODUCTION

The phrase, packed bed heat transfer, is currently used to describe a variety of phenomena, namely: the convective heat transfer from the walls of the packed bed to the fluid, the convective heat transfer from the particles to the fluid flowing through the bed, sometimes referred to as the fluid-particle mode, the conduction heat transfer from the walls of the bed to the particles constituting the bed, the conduction heat transfer between the individual particles in the bed; this is sometimes referred to as the particle-particle mode, radiant heat transfer and heat transfer by mixing of the fluid. These modes are illustrated schematically in Fig. 1. (1): Wall to fluid convection; (2): particle to fluid convection; (3): Wall to particle conduction; (4a): Radial particle to particle conduction; (4b): Axial particle to particle conduction; (5a): Radiant heat transfer between particles; (5b): Radiant heat transfer between wall and particles; (5c): Radiant heat transfer between fluid and particle; (6): Heat transfer by mixing of the fluid. The fourth mode, namely the conduction between the particles, can be further subdivided into the axial and radial directions. Moreover, at elevated temperatures heat transfer by radiation will also be an important mode. In many industrial applications, it is found that two or more of the modes cited above take place simultaneously.

Moreover, the modes may interact with one another. For example, the conduction between the particles may be affected by the convection between the particles and the fluid. This interaction among the different modes is one of the main reasons for the difficulty in correlating the total heat transfer and analyzing the experimental data in this field.

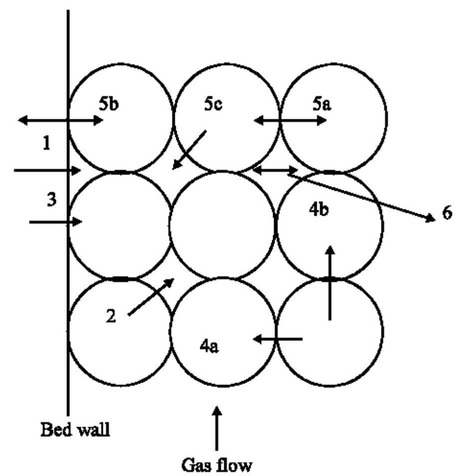

Fig. 1: Modes of heat transfer in packed beds

THEORETICAL INVESTIGATION

The first research on heat transfer in packed beds was published by Anzelius (1926), although, Schumann (1929) is usually the first reference cited in most study. Each of them made a number of simplifying assumptions and solved the heat transfer equations for an incompressible fluids passing uniformly through a bed of solid particles with perfect conductivity. The derived heat transfer equations were:

$$(T_s - T_{s,0})/(T_{f,0} - T_{s,0}) = 1 - e^{-Y-Z} \Sigma Y^n M^n (yz) = e^{-Y-Z} \Sigma Z^n M_n \quad (1)$$

$$(T_f - T_s)/(T_{f,0} - T_{s,0}) = 1 - e^{-Y-Z} \Sigma Y^n M^n (yz) = e^{-Y-Z} \Sigma Z^n M_n (yz) \quad (2)$$

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Where,

T = The temperature (of fluid and solid)
Y and Z = Dimensionless quantities

The solutions of these equations were presented in graphical form, called Schumann curves. Thus, to evaluate volumetric coefficients of heat transfer using these curves, it is only necessary to measure exit air temperature and the bed temperature. These curves could be used to evaluate the heat transfer coefficients for a given packed bed undergoing heat exchange with a fluid provided these conditions, which were the simplifying assumptions made by Schumann are satisfied:

- The solid particles are so small or have such a high thermal conductivity that no temperature gradients exist within the solid particles. This means that the solids offer a negligible resistance to heat transfer
- The resistance to heat transfer by conduction in the fluid is also negligible
- The rate of heat transfer from fluid to solid or vise versa at any point in the bed is directly proportional to the average temperature differential between them at that point
- The densities of solid and fluid and other transport properties are independent of temperature

Upholding the above conditions, Furnas (1930) extended the Schumann curves to wider coverage temperatures. He also, postulated an empirical relation for the evaluation of the heat transfer coefficient thus:

$$h_v = AG^{0.7}T^{0.3}10^{(1.68\varepsilon - 3.56\,\varepsilon 2)}/d_p^{0.9} \quad (3)$$

Where,

h_v = The volumetric heat transfer coefficient (KW/m^{3}K)
A = A constant dependent on the bed material is the mass velocity of the fluid is the average air temperature
d_p = The particle
ε = The voidage

Saunders and Ford (1940) used dimensional analysis to derive correlations to calculate heat transfer coefficient. The research was, however, limited to spheres and can not directly be applied to other geometries of soli particles.

Kays and London (1964) presented another correlation for evaluating heat transfer coefficient between gas and randomly packed solid spheres. Using the Colburn j-factor, the correlation was given as:

$$j_H = 0.23/Re_p^{\,0.3} \quad (4)$$

Where,

$$j_H = St.Pr^{\,2/3}$$

$$200 < Re_p < 50{,}000 \text{ and } 0.37 < \varepsilon < 0.39$$

Lof and Hawley (1948) investigated heat transfer between air and packed bed of granitic gravel. Unsteady state heat transfer coefficients were correlated with the air mass velocity and particle diameter to obtain the equation:

$$h = 0.652\,(G/d_p)^{0.7}\ (KW/m^3\,K) \quad (5)$$

This was evaluated for 8 mm$<d_p<$33 m; $50<Re_p<500$ and temperature range of 311-394 K.They also concluded that the temperature of the entering air had no appreciable effect on the coefficient.

Leva (1948) determined heat transfer coefficient between smooth spheres of low thermal conductivities and fluids (air and carbon dioxide) in packed beds and tubes of 2 and ¾ inches (50.8 and 6.4 mm, respectively) diameters. The ratio of particles to tube diameters was varied from 0.08-0.27. Gas flow rate was of Reynolds number range 250-3,000. Correlation of film coefficient was found to be:

$$h = 3.50\,(k/D_t)\,e^{-4.6Dp/Dt\,(DpG/\mu)\,0.7} \quad (6a)$$

By approximation, this reduces to:

$$h = 0.40\,(k/D_t)\,(DpG/\mu)\,0.7 \quad (6b)$$

or,

$$N = 0.4\,Re^{0.7} \quad (6c)$$

Maximum film coefficient was predicted and verified at a value of D_p/D_t of 0.153. Riaz (1977) and Jefferson (1972) studied the dynamic behavior of beds undergoing heat exchange with air using single and two phased modes. By incorporating factors of axial bed conduction and intra-particle resistance, which Schumann ignored, the heat transfer coefficients were evaluated and found to be (1 + Bi/5) times smaller than those predicted using Schumann curves.

Ball (1958), Norton (1946), Meek (1961), Bradshaw and Meyers (1963), Harker and Martyn (1985) and Bouguettaia and Harker (1991) also have all worked on various packed beds using air and other gases as fluids and have developed correlations involving the heat transfer coefficient. Barker (1965) and Chechetkin (1963) have given extensive reviews of these works and were appropriated compared results. Few workers studied solid-water systems. Furnas (1930) reported some study of a packed bed, in which water was used as heat transfer fluid but he thought the data to be too unreliable to draw any good conclusions. The little interest shown in these systems is probable due to their limited industrial applications.

The boiling point of water (373 K at normal atmospheric pressure) limits its use, while in liquid state and most moderate to high temperature packed bed systems use air as heat transfer fluid as found in blast furnaces, regenerators, coke ovens and so on. However, there is growing potential use of water as a heat transfer fluid for example in energy conservation in industry, such as in the recovery of waste heat from industrial effluent and the use of water (and other fluids) in solar energy devices. It would therefore, be necessary to develop appropriate correlations for evaluating the heat transfer coefficients in the design of such systems. Baldwin (1961) has developed correlations for heat transfer between water and oriented spheres in the range 3,000<Re<70,000: For regular cubic arrangement ($\varepsilon = 0.478$):

$$Jh = 0.992/Re^{0.33} \quad (7)$$

For dense cubic arrangement ($\varepsilon = 0.261$):

$$Jh = 0.940/Re^{0.30} \quad (8)$$

With regards to heat transfer between air and packed solids, Bouguettaia (1989) assumed the Schumann conditions outlined in the above subsection to hold and subsequently used the Schumann curves to calculate the volumetric heat transfer coefficient for spheres of glass, water, air and palm oil (a phase change material). In doing this he adopted a procedure used by Lof and Hawley (1948) whereby, the curves drawn are compared with the standard Schumann curves. The results were correlated against air mass velocity to obtain an equation of the general form:

$$h_v = KG^n \quad (9)$$

The summary of the results is given in Table 1. One of the fundamental assumptions made in the derivation of the Schumann curves is a negligible resistance to heat transfer within the solids. This can only hold if, according to McAdams (1954), the particle Biot number is <0.1. Kreith (1966) also, submitted that the error introduced in assuming negligible particle resistance is <5% when, the Biot number is <0.1, while Davidson and Harrison (1971) stated that the error is negligible for Biot numbers of up to 0.25. The Biot number is defined as the ratio of resistances due to particle conduction to the one due to convection and is given by the Eq. 10:

$$Bi = h_s L_c / k_p \quad (10)$$

Where,

h_s = Convective heat transfer coefficient (W/m^{2}k)
k_p = The thermal conductivity of the particle (W/mk)
L_c = The characteristic length corresponding to the maximum spatial temperature differential, which for spheres is $d_p/2$

Equation 10 can be written in terms of volumetric heat transfer coefficient for a bed of packed spheres, thus:

$$Bi = h_v d_p / 2a_s k_p \quad (11)$$

Where,

a_s = The surface to volume ratio of the packing material (m^{-1})
d_p = Particle diameter (m)

To verify the validity of the assumption of the negligible solid resistance, it is necessary to evaluate the Biot numbers. This was therefore carried out for some of the published works. Typical results were used and these are shown in Table 2. The particle diameter used for Furnas (1930), Saunders and Ford (1940), Lof and Hawley (1948) were 8.6 and 20 mm for Bouguettaia (1989), which was his smallest change material. Typical results from the research of Bouguettaia are corrected as shown in Table 3.

Table 1: Correlation of heat transfer coefficients

Bed material (in spheres)	Operation	Air mass velocity (G)	Inlet air temp. (K)	Values of K	K and N
Glass balls	Heating	1.40-2.48	326-333	9.24	0.90
	Cooling	1.18-2.96	296-298	10.09	0.90
Air	Heating	0.74-1.90	308-315	3.32	0.77
	Cooling	0.65-1.71	292-297	3.38	0.77
Water	Heating	0.76-2.90	309-317	9.98	0.92
	Cooling	0.66-2.57	292-301	10.22	0.92
Palm oil	Heating	0.82-3.29	319-328	10.64	0.86
	Cooling	0.72-2.90	290-301	11.10	0.86

Table 2: Comparison of biot numbers of heated packed beds

Material	d_p	G ($kg/m^2/s$)	Air inlet temp. (k)	h_v (KW/m^3)	a ($^{m-1}$)	K (w/mk)	Bi
Limestone	8.6	0.245	367	8.57	360	10.0	0.0102
Iron balls	8.6	0.245	367	12.44	360	172.0	0.0009
Pb/steel	8.6	0.245	367	20.88	360	140.0	0.0018
Gravel	8.6	0.245	367	7.43	360	19.7	0.0045
Glass	20.0	1.696	328	14.95	207	11.7	0.0315
Palm oil	20.0	2.460	321	22.14	207	0.2	6.2660

Table 3: Corrected volumetric heat transfer coefficients

Material	Operation	h_v	k_p	K_p'	Bi	h_{ve}
Palm oil	Cooling	19.30	0.171	-	5.4665	9.244
Palm oil	Heating	2.18	0.171	1.320	0.8116	19.082

The two coefficients h_v and h_{ve} in the cooling run show significant difference. This is due to the large Biot number exhibited by the palm oil, which is a poorly conducting material. During heating (melting), natural convection significantly improves the heat transfer as indicated by the low Biot number and the h_{ve} is not significantly attenuated. To ensure high rates of heat transfer it is necessary to have as low Biot numbers in the solids as possible. Equation 10 can be written as follows:

$$Bi = h_v\, d_p/2k_p - St.Re_p.\, Pr.\, (k_f/2k_p) \qquad (12)$$

From the equation, it is evident that lower values of Biot numbers can be obtained by either of the following particle size. Each of these workers reported normal voidage in the packing.

EXPERIMENTAL INVESTIGATION OF PACKED BED HEAT TRANSFER

Table 4 summarizes, the experimental conditions, proposed correlations, range of parameters, etc. of the many investigations, which are reviewed here. Two parameters, the effective or apparent conductivity, k and the total heat transfer coefficient, h, are commonly used to express the heat transfer rates in packed beds. The effective thermal conductivity is an averaged parameter that describes the total thermal performance of the granular medium that constitutes the bed. In other words, the effective conductivity is the conductivity the medium would have were it homogeneous. It should be noted that the effective conductivity is not the same as the conductivity of the material that constitutes the bed. The effective conductivity is dependent both on the thermal properties of the bed and on the flow rate, but in most experimental correlations, it is generally expressed as a function of the Reynolds number only. This implies that these correlations are valid only for the particular bed materials, which were used in developing them. The other preferred parameter to describe the thermal performance of a packed bed-the total heat transfer coefficient, h, -is also, an averaged parameter. This total heat transfer coefficient generally incorporates the conduction mode between the bed particles, the convective mode between the bed particles and the flowing medium, the wall to bed conduction and the wall to fluid convection. The last two modes can be expected to occur when, the walls of the bed are maintained at isothermal conditions.

Therefore, as in the case of the effective conductivity, the total heat transfer coefficient is also dependent both on the thermal properties of the bed and on the flow rate, but most experimental correlations express it (in dimensionless form) as a function of the Reynolds number only. Therefore, the applicability of these correlations is also limited to the particular bed materials used in developing them.

The commonly used dimensionless numbers for expressing it are the Colburn-J factor and the Nusselt number. They are defined as:

$$jh_t = h_t/C_{pf}G_f\, .\, (Pr_f)^{2/3} \qquad (13)$$

and

$$Nu_t = h_tD_p/k_f \qquad (14)$$

It is expected that depending on the experimental technique, different modes of heat transfer will contribute to the total heat transfer. For example, in transient experiments performed in beds with adiabatic walls, the modes that contribute will be axial conduction and convection between the bed and the fluid. On the other hand, in beds that have heated walls the significant modes are wall to bed and wall to fluid heat transfer in addition to the radial conduction and the bed to fluid convection. Therefore, depending on the modes that have contributed, the experimentally obtained correlations relating the total heat transfer rates to the Reynolds number can be expected to show wide variation. Moreover, the thermal properties of the bed have a significant effect on the conduction mode and therefore, studies using similar experimental techniques but different bed materials can also be expected to yield different correlations relating the heat transfer rates to the Reynolds numbers.

Table 4: Experimental review on packed bed heat transfer

Authors	System	Reynolds no. range, (Re_p)	Type of heating	Correlation proposed	Remarks
Balakrishnan and Pei (1974)	Commercial catalyst formulations; (alumina based spheres and cylinders)	300-400	Micro wave heating	$(jh)f_p = 0.018 (Fr_{pm})^{-0.5} \times \phi_s^{3.76}$	Use of microwave heating results in uniform temperature throughout the bed eliminating thermal gradients in solid phase and therefore particle conduction. Convective effects truly determined
Baumeister and Bennet (1958)	Air-steel	200-10400	Electrical induction (steady state)	$Jh = 1.09\ Re_p$	The bed consisted of 4in. internal diameter. Transite cylinder and the packing 0.155, 0.2495 and 0.3745 in diameter high carbon-chrome alloy steel. The heat transfer, therefore, consists of axial conduction as well as fluid-particle modes.
Bradshaw and Myers (1963)	Air-celite cylinders, etc.	300-2500	Water soaked packing's dried by air at room temp.	$Jh = 2.52 (Re_{pm})^{-0.5}$	Data are based on drying experiments. Assumption of constant surface temperature of all the particles is debatable.
Bunnel *et al.* (1949)	Air-alumina	30-150	Heated airbed wall maintained at boiling point of water	$K_{er}/k_{ef} = 5.0+0.061Re_p$	Entering air at 400°C cooled by boiling water-jacket at the walls. K_{eff} rad .is therefore, effective radial conductivity. Solid and gas temperatures are claimed to be identical, which is a questionable statement. Bed consisted of 0.125in. diameter alumina cylinders
Coberly and Marshall (1951)	Air-celite	70-300	Steam-heated wall	$K_{er} = 0.018+ 0.00098Re_p$ (Dimensional equation in British units. Multiply k_{er} obtained from equation by 1.730735 to convert to J/msK)	The gas inlet and outlet and radial and axial temp. have been measured with high response thermo couples. The experimental technique would seem to indicate considerable axial conduction, which has been ignored in the analysis.
Colburn (1931)	Air pebbles, granules ,pellets, porcelain balls,etc	G:1-9	Steam-jacketed packed tube	$h = 8aC_pv^{0.2}G^{0.83}$ $a = f(D_p/D_t)$ (Multiply h obtained from equation by 5.678264 to get SI units.)	This study relates to the rate of heat transfer from the wall to the fluid flowing in the packed bed. One of the earliest (1931) studies, it does not consider the gradients within the bed.
De Acetis and Thodos (1960)	Air-catalyst spheres	6.71-1000	Water soaked particle dried by air at rooms temp.	$Jh = 1.1/(Re_p^{0.41} -0.15)$	The usual assumption of uniform temp.at the surface of the bed particles, being the wet bulb temp.is used. This could be wrong and the source for the breakdown of the heat and mass transfer analogy.
Eichorn and White (1952)	Air + Co^2 and divinyl benzene	58-374	r.f heating	-	Transient temp.measurements are made. While, a constant temperature may be achieved during steady- state measurements resulting in pure fluid-particle mode, during the cooling cycle particle-particle heat transfer will be significant.
Furnas (1930)	Air-iron spheres	120-1200	Heated air	$h = 6.91 \times 10^{-3} \times G_f^{0.75} \times t_f \times G\text{-}1.56$ (Multiply by 5.678264 to get h in SI units)	One of the earlier investigations (1930),temp.of both solids and gases were measured. Test section was 15 cm diameter and 105 cm high, coated with Sil-O- Cel for insulation. Heat transfer therefore consisted of fluid-particle mode and particle-particle mode in axial direction.
Gamson *et al.* (1943)	air-celite spheres and cylinders	40-4000	Water-soaked packings dried by air	$Jh = 1.064 \times Re_p^{-0.41}$ Re >350 $Jh = 18.1 \times Re_p^{-1.0}; Re_p<40$	The assumption of wet bulb temperature as the constant surface temperature of the bed is debatable. If true the heat transfer consists of only fluid-particle mode.
Glasser and Thodos (1958)	Air-ceramic/ brass spheres	330-1500 300-4000	Transient heating by regeneration technique; steady-state heating by single electrically heated sphere	$Nu = 1.25\ Re_p^{0.56}$ $300<Re_p<4000$	In the regenerative technique the surface temp.is quite different from the values at the center. The electrical heating method is perhaps much more accurate.
Gupta and Thodos (1962a)	Existing data	33-6500	Electrical heating of single sphere	$\varepsilon jh = 0.0108 + 0.929/Re_p^{0.58}-0.483$ Re >20	Used existing data in the literature to obtain the correlation
Gupta and Thodos (1962b)	Air-celite spheres	95-2500	Water soaked particles dried by air	$\varepsilon jh=2.06/Re_p^{0.58}$	In this study, mass transfer rates are minimized so as to achieve the direct heat and mass transfer analogy.

Table 4: Continue

Authors	System	Reynolds no. range, (Re_p)	Type of heating	Correlation proposed	Remarks
Houghen and Piret (1951)	Air-celite spheres	193-2824	Heated air entering the bed	$K_e/C_p\ \mu = 3.7/\varepsilon \times (Re_p)^{1/3}$	The bed is cooled by water at 60°F. Therefore, there is a radial transport of heat in both phases. Air is flowing downward.k_e values obtained are 20-100 times that of the fluid and 2-15 times that of solid.
Leva (1947)	Air-smooth spheres	500-3408	Heated wall	$h = 0.813k_f/D_t$. $e^{-6\,(Dp/Dt)}$, $Re_p^{0.9}$ (Multiply h by 5.678264 to get SI unit values)	Heat is transferred from wall to gases flowing in a packed tube.D_p/D_t. Values ranged from 0.05-0.3. Maximum heat transfer occurred at $D_p/D_t = 0.15$
Leva and Grummer (1948)	Air-smooth spheres	500-3408	Heated wall	$h = 0.813k_f/D_t$. $e^{-6\,(Dp/Dt)}$ $Re_p^{0.9}$ (Multiply h by 5.678264 to get SI unit values)	Extension of above work for non uniform beds arithmetic average of D_p is used. Nominal D_p for Raschig rings and D_p of equal volume sphere for cylinders.
Leva *et al.* (1948)	Air-smooth spheres	250-3000	Heat air and Co_2; bed wall is cooled	$h = 3.5k_f/D_{t\,x}$ $e^{-4.6\,(Dp/Dt)}$, $Re_p^{0.7}$ (Multiply h by 5.678264 to get SI unit values)	Extension of earlier research by leva.different equation is obtained for cooling operations. Maximum heat transfer is at $D_p/D_t = 0.153$
Lindauer (1967)	Air-steel and tungsten	23-18200	Electrical resistance heating to produce sinusoidal gas temperature variations.		Cyclic temperature variations are used in the gas phase. This could cause temperature gradients in the solid phase leading up to the question, whether the measured values pertain to the gas-solid mode alone.
Plautz and Johnstone (1955)	Air-glass spheres	100-200	Steam heated wall; entering air at 85°C maintained by heating and cooling coils	$K_{er} = 0.439 + 0.00129xRe_p$ $hw = 0.09G_f^{0.75}$ [Multiply k_{er} and h obtained from correlation by 1.730735 and 5.678264, respectively, to get SI unit values.]	Reports both radial effective conductivity and wall transfer coefficient.
Wakao and Kato (1969)	Air-glass		Concentric carborundum electric heater	$K_e/k_f = (k_e/k_f)_{Nur} = 0 + 0.707$ $Nu_r^{0.96}(k_s/k_f)^{1.11}$	The effect of radiative heat transfer in conduction has been evaluated.
Yagi and Kunii (1957)	Air-iron,porcelain, cement clinker, fire brick and Rasc-hig rings	-	Coaxial electric heater at the center; outside wall is insulated with fire brick.	$K_e/k_f = B + Re_pPr_f/D$	Model agrees well with data at $Re_p \rightarrow 0$

Consequently, in the detailed review of the experimental studies the various investigations are grouped under the heating technique employed. By this classification a grouping of the studies by the modes or combination of modes that contribute to the total heat transfer is achieved. This provides a convenient basis for the review.

Heated bed walls: Leva (1947) studied the heat transfer from the wall to a gas flowing in a packed bed. The mode that may be expected to occur in this configuration is wall to particle conduction, wall to fluid convection, particle to particle conduction radially and finally particle to fluid convection. The study involved the use of different sizes of spheres and found the maximum heat transfer occurred when, the particle to the diameter ratio, D, /D, was 0.15. In a subsequent study, Leva and Grummer (1948) extended the research to non uniform spheres and found the use of a simple arithmetic average of the D, value in their correlation fit the data satisfactorily. Similarly, the use of the nominal diameter for Raschig rings and the diameter of a sphere of equal volume for cylinders was suggested. It may be noted that the researchers did not use the well known shape factor for the beds of non uniform shapes. In an extension of this research, Leva *et al.* (1948) found that if the gas were cooled instead of being heated, the constants in their correlation had to be altered. The differences in the constants for heating and cooling probably arose due to the fact that the thermal gradients within the solid phase of the bed, especially in the radial direction near the wall, were different in the two cases. The electrically heated wall would produce a constant flux and the water-cooled wall in the second case would produce approximately, an isothermal surface at the wall.

Similar studies were performed by Colburn (1931). He also studied, the rate of heat transfer from the tube surface to the air flowing in a packed tube. Steam-jacketed pipes were used and these can be expected to produce a constant temperature wall surface. It is therefore, not surprising that the numerical values estimated by Colburn's (1931) correlation are closer to the correlation of Leva *et al.* (1948) for the cooling case.

Bunnell *et al.* (1949) used hot air at 400°C in a bed, which was cooled by a steam jacket. Using chromel alumel thermocouples, they measured both solid and gas temperatures and found them to be identical at any given radial position. This would normally imply that convective effects are absent as there is no temperature driving force and the only heat transfer mode is conduction through the solid phase in the radial direction. However, at steady state as heat is conducted toward the wall, the packing at the center of the bed would tend to cool unless heat is supplied by the air. Furthermore, the correlation proposed for the effective conductivity shows a strong dependence on Reynolds number proving convective effects is significant. Therefore, it seems reasonable to conclude that the gas temperature measurements were probably in error.

Similar studies were performed by Coberly and Marshall (1951), Houghen and Piret (1951) and Plautz and Johnstone (1955). All of these authors assumed the packed bed to be a homogeneous medium and used the heat diffusion equation together with the experimentally obtained values of the temperature profile to calculate the effective conductivity of the bed in the radial direction.

A significant feature of these studies is that the effective conductivity was obtained experimentally as a function of the Reynolds number. In other words, the convective effec on the conductive mode has been studied. However, the fundamental properties of the bed materials have not been included as a parameter in the correlations and therefore, each correlation is restricted to a specific bed material only.

Heat transfer rates from drying experiments: Glasser and Thodos (1958), DeAcetis and Thodos (1960), Gupta and Thodos (1962a, b, 1963) and Bradshaw and Myers (1963) used a drying technique to determine heat and mass transfer rates in packed beds. This technique involved soaking the bed material, which generally consisted of catalyst carriers. The packing was then dried with air. The several correlations based on this research are listed in Table. The heat and mass transfer rates obtained from these experiments were expressed in terms of the Colburn J factor: jh for heat transfer and jd for mass transfer. In every case, the value of jh was greater than jd (The two values would have to be equal if the heat transfer consisted of particle-fluid convection alone, this follows directly from the heat transfer-mass transfer analogy). The reasons for the heat transfer phenomena in the drying experiments consisting of both convection and conduction are as follows.

- The bed surface temperature, which was generally assumed to be constant and uniform at the wet bulb temperature of the air has been shown by Gupta and Thodos (1964) to be incorrect
- To avoid wall effects the bed walls were maintained at the wet-bulb temperature of the air, but the bed itself, being at a different temperature, leads to even greater temperature gradients

From the above considerations, it may be concluded that the heat transfer in the bed consisted of both conduction through the solid phase and convection between the bed and the drying air.

Other heating methods: Furnas (1930) studied heat transfer in a bed of iron spheres. In this study, a step change in inlet temperature in the air was used. The column was insulated ensuring adiabatic conditions. When, the bed attained steady state, heated air was introduced in an upward flow through the vertical column. Here, the modes of heat transfer that can be expected to contribute to the total heat transfer are convection from the fluid to the spheres and conduction between the particles in the axial direction (walls are insulated to minimize radial conduction). Lindauer (1967) used electrical resistance heating to produce cyclical temperature variations. These temperature variations cause similar variations in the gas but with a change in phase. Therefore, there is always temperature driving forces between the particles and fluids and between the particles. Since, the walls were insulated, the contributing heat transfer modes are axial conduction and convection between the fluid and the particles.

Electrical induction was used by Baumeister and Bennett (1958). Steady-state measurements were made in a bed with insulated walls and therefore, here again axial conduction and fluid to particle convection are the contributing modes of heat transfer.

Eichorn and White (1952) used heating in their transient measurements. While, a constant temperature may be achieved in steady-state measurements during the heating cycle resulting in pure convective transfer between the bed and the fluid, during the cooling cycle particle to particle conduction will also, be significant. This results in both the conduction mode and convective mode contributing to the total heat transfer. Although, the modes contributing to the total heat transfer in the studies of Lindauer (1967), Baumeister and Bennett (1958) and Eichorn and White (1952) are the same, their proposed correlations show wide differences. This is believed to be due to the fact that the bed materials used in these three investigations were each different and their physical and thermal properties can be expected to affect the correlations.

The research of Balakrishnan and Pei (1974) and Bhattacharyya and Pei (1975) involved the use of microwaves as the heating medium. When, a bed of metallic oxides is subject to microwave radiation, the entire bed attains a uniform temperature instantaneously. This gives rise to a situation where, thermal gradients within the bed are eliminated and thereby, the particle to particle conduction mode is absent and the fluid to particle convection heat transfer can be determined.

Using this technique, Bhattacharyya (1973) proposed the following correlation

$$(jh)_{fp} = (0.018\ Ar_m/Re_p)^2 \qquad (15)$$

Subsequently, Balakrishnan (1973) working with non spherical packings, extended the correlation to

$$(jh)_{fp} = 0.018[Fr_{pm}]^{-0.5}[\phi_s]^{3.76} \qquad (16)$$

where, ϕ_s is the shape factor of the pellets. Equation 16 may also be written in terms of the Nusselt number as:

$$Nu_{fp} = 0.O16\,[Ar_m]0.25\,[Re_p]\,[\phi_s]^{3.76} \qquad (17)$$

Obviously as Rep→0, Nu_{fp}→0; i.e., at no flow there is no forced convection heat transfer.

THEORETICAL AND EMPIRICAL MODELS

In the experimental studies on heat transfer in packed beds, it was seen that most of the studies were directed at obtaining either the effective conductivity or the total heat transfer coefficient as functions of Reynolds number. However, it is obvious that the conduction mode, apart from being dependent on the Reynolds number, is also strongly dependent on the physical, thermal and transport properties of the bed material.

Furthermore, the conduction mode can be subdivided into the radial and axial directions. Radiation will also be a significant contributor to the total heat transfer at elevated temperatures. In view of all this, a more fundamental approach is to analyze the contributions of each mechanism or mode in terms of the properties and parameters, which are likely to affect the heat transfer.

The implicit assumption, in many of the early models that used this approach, is that the different modes are essentially additive. This type of analysis was first used by Schumann and Voss (1934) and Damkohler (1937). They summed up the contributions of each mechanism, by which heat is transferred radially in the bed. Later, Wilhelm *et al.* (1948) modified this work for systems, where the gas is not flowing.

Verschoor and Schuit (1950) divided, the effective conductivity of packed beds into two parts, one which is independent of fluid flow and the other, which is dependent on the lateral mixing of the fluid in the packed bed. Singer and Wilhelm (1950) suggested separate mechanisms, by which heat is conducted radially within the bed, namely:

- Thermal conduction through the fluid phase
- Thermal solid-fluid-solid convection
- Thermal solid-solid conduction through contact surfaces
- Thermal conduction within the solid
- Thermal conduction through the fluid film near the contact surface of the packings

Ranz (1952) extended this model by adding contributions due to:

- Radiant heat transfer between packings
- Radiant heat transfer between neighboring voids
- Heat transfer by lateral mixing of the fluid

Agro and Smith (1953) used the model by Ranz (1952) and their experimental results indicate good agreement with their predictions. The research by Ranz (1952) also laid the foundation for the model suggested by Yagi and Kunii (1957). They also divided, the conductive heat transfer into two parts, namely those dependent and independent of fluid flow. Heat transfer mechanisms independent of fluid flow, according to Yagi and Kunii (1957) are:

- Thermal conduction within the solid
- Thermal solid-solid conduction through the contact surfaces
- Radiant heat transfer between packings
- Radiant heat transfer between adjacent voids

The mechanisms dependent on fluid flow are:

- Thermal conduction through the fluid film near the contact surface of the two solids
- Solid- fluid-solid heat transfer by convection
- Heat transfer by lateral mixing of fluid

Their generalized of fluid flow is:

$$k_e/k_f = k_e^o/k_f + (\alpha\beta)\, Pr_f\, Re_p \tag{18}$$

where, k_e^o, the effective conductivity of the bed under no flow conditions, is given by

$$k_e^o/k_f = \frac{\beta(1-\varepsilon)}{\dfrac{(1^{\cdot})}{\gamma(k_f/k_s) + (1/\phi) + (D_p h_{rp}/k_t)}} + \varepsilon\beta(D_p h_{rp}/k_f) \tag{19}$$

On comparison with experimental data, the Yagi and Kunii model Eq. 19 is found to underestimate the effective conductivity of the packed bed. They attributed the discrepancies, which were appreciable in some cases, to the unavailability of the physical parameters of the system. However, Bhattacharyya and Pei (1975) suggest that these discrepancies probably resulted from not incorporating the effects of fluid flow on the conductive heat transfer mode. Yagi *et al.* (1960) were among the first to measure the effective thermal conductivity of packed beds in the axial direction. Their purpose was apparently to test the validity of the Yagi and Kunii (1957) model with reference to the axial conduction. Using a bed with adiabatic walls to eliminate radial gradients and an infrared lamp applied continuously to the fluid flow as the heat source, they were able to correlate their results with Eq. 18. ($\varepsilon\beta$) took on values between 0.7 and 0.8 in this case.

Wakao and Kato (1969) and Wakao and Vortmeyer (1971) developed a lattice model to predict the effective thermal conductivity (of beds with stagnant fluids) as compared to the simplified geometric model of Yagi and Kunii (1957). The heat transferred through a unit cell was calculated using a relaxation procedure, where the solid and gas were replaced by a network of rods (A point contact was assumed between the spheres and the conductance at the contact point in the grid was evaluated by assuming unidirectional heat transfer). Numerical solutions were obtained for combined conduction and radiation effects in cubic (ε = 0.476) and orthorhombic (ε = 0.395) lattice models of spheres. The generalized correlation is:

$$k_e/k_f = [k_e/k_f]_{Nurp=0} + 0.707\, Nu_{rp}^{\ 0.96}\, [k_e/k_f]^{1.11} \tag{20}$$

for

k_e/k_f = 20-1000, Nu_{rp} = 0-0.3 and $Nu_{rp} = h_{rp}D_p/k_s$

Bhattacharyya and Pei (1975) also used the approach of correlating the contributions of each mode separately. They evaluated the total heat transfer by:

$$Nu_t = k_e^o/k_f + Nu_{fc} + Nu_{fp} \tag{21}$$

where, k_e^o/k_f is the effective conductivity with motionless fluid evaluated by the Yagi and Kunii (1957) model Eq. 19 for Nu_{fc}, the effect of fluid flow on conduction, they proposed:

$$Nu_{fc}/Re_p Pr_f^{1/3} = jh_{fc} = 2.05\times10^{-5}(P_p C_{pp}/P_f C_{pf}) \tag{22}$$

Nu_{fc} the convective heat transfer was evaluated by their earlier experimental correlation (Eq. 17). The model by Bhattacharyya and Pei (1975) uses the Yagi and Kunii (1957) model, which is empirical in nature. Furthermore, the effect of fluid flow, Nu_{fc},, was obtained empirically from experimental data in the literature on total heat transfer rates. This was done by evaluating k_e^o/k_f and Nu_{fc}, respectively and subtracting them from Nu_t values available in the literature and subsequently correlating them as shown in Eq. 22. Therefore, while the convective part of the Bhattacharyya and Pei (1975) model is general and valid for all bed materials, the applicability of the conductive mode, particularly the effect of fluid flow on conduction, Nu_{fc}, is limited to bed materials and ranges used in obtaining the Nu_t, values in the study.

While, the models presented up to now treat the problem of heat transfer in packed beds by looking at the conduction mode theoretically and using experimental studies to evaluate the convective transfer, there are several investigations, which treat the bed as a homogenous medium and make no distinction between the contributions of the different modes.

Kunii and Suzuki (1967) used the familiar two-dimensional diffusion equation for packed beds in the range of low Peclet number, i.e., Pe <10. However, this model over predicted the heat and mass transfer coefficients and this was attributed to channeling or local uneven contacting of fluids with the solids in the real situation.

Galloway and Sage (1970), using an instrumented copper sphere in a bed of packed spheres, obtained data on the local heat transfer rates. They discussed their results in terms of boundary layer theory with particular reference to single spheres and cylinders in turbulent flow. Wakao *et al.* (1973) theoretically showed that the Nusselt number between a spherical particle in a packed bed and stagnant fluid has a limited value of 2, which is the same as a single sphere in a stagnant fluid environment. They suggest that this should be considered as the lower limiting Nusselt number. Sorenson and Stewart (1974) assumed a constant bed temperature and solved the continuity and conservation equations and obtained numerical solutions for heat and mass transfer rates for creeping flow at low Peclet numbers. Nelson and Galloway (1975), working on a model for mass transfer, solved the unsteady-state diffusion equation with boundary conditions from the Penetration Theory and obtained an expression for the Sherwood number, which is valid for a Reynolds number range of 0.01-100. Rowe (1975) modified Nelson and Galloway's model slightly to make it applicable to liquid fluidized beds. Votruba *et al.* (1975) modeled and obtained heat and mass transfer rates experimentally in monolithic honeycomb catalysts, whose main application is in pollution control in the automobile industry. Gunn and Khalid (1975) presented an analysis showing the effect of axial and radial thermal dispersion and wall thermal resistance heat transfer to fixed beds of solids. Application of this analysis to experiments yielded axial, thermal and wall transfer coefficients. The discrepancies on comparison with experimental values of other workers were attributed to the fact that axial dispersion was neglected by some workers and also the thermal characteristics of fixed beds of metallic and nonmetallic particles were neglected.

Recently, there has been considerable interest in the use of packed beds of spheres for insulation purposes, especially in cryogenic applications. Here, of course, the convective mode is absent as there is no fluid flow through the bed. Yovanovich (1975) used the concept of thermal constriction resistance within the spheres and conduction resistance of an effective gas gap thickness to predict the apparent conductivity of glass microspheres from atmospheric pressure vacuum. Chan and Tien (1973) have presented an analytical study of the heat transfer through the solid phase of a packed bed of spheres. Both these models use the concept of a finite contact spot between the spheres in the bed. The dimensions of the finite contact spot dimensions may be estimated by the Hertz relation from elasticity theory:

$$r_c = [3/4(1-v^2/E)Fa]^{1/3} \quad (23)$$

CONCLUSION

Majority of the experimental studies were directed toward correlating the total heat transfer rates (generally in dimensionless form) with Reynolds number. The total heat transfer consisted, in most cases, of both the conduction mode and the convective mode. Since the conduction mode depends on the physical and transport properties of the bed materials, these correlations although, often reliable are applicable to the particular bed materials, for which they were developed only. Packed bed actually offers moderate heat transfer rates and pressure drops and due to this passive solar systems are recommended to operate with it for thermal energy storage.

Nomenclature

a	:	Radius of bed particle (m)
A_{rm}	:	Archimedes number
L	:	Height of bed
a_s	:	Specific surface area
C	:	Specific heat (J kg^{-1} K)
d	:	Particle diameter (m)
D	:	Tube diameter (m)
e	:	Emissivity, dimensionless
E	:	Young's modulus (N m^{-2})
F	:	Force acting on contact spot (N)
Fr_{pm}	:	Modified Froude number-dimensionless
g	:	Acceleration due to gravity (m sec^{-2})
G_f	:	Fluid mass velocity (kg m^{-2}sec)
G	:	Air mass flow rate (kg/m^{2}s)
h_v	:	Heat transfer coefficient (J m^{-2}sec K)
hr	:	Radiation heat transfer coefficient between solids
h	:	Thermal conductivity (J ms^{-1} K)
Nu	:	Nusselt number, dimensionless
BI	:	Biot number, dimensionless
Pe	:	Peclet number, dimensionless
Pr_f	:	Prandtl number, dimensionless
St	:	Stanton number, dimensionless
r	:	Contact radius (m)
Re	:	Particle Reynolds number, dimensionless
T	:	Temperature (°C)
u_f	:	Fluid velocity (m sec^{-1})

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