หอสมุดกลาง จุฬาลงกรณ์มหาวิทยาลัย

CHAPTER 4

FLUIDIZATION

The technique of fluidization is related to operations first used commercially in the fields of mining and metallurgical engineering such as liquid settling, sedimentation and density classification. Agricola (1556) refers to what is virtually an application of fluidization in mineral dressing. Perhaps the first major successful application of gas-fluidized bed techniques, however, was to the engineering of the catalytic cracking process. During the 1939-1945 war, there was a great demand for high octane aviation spirit as a fuel for the piston driven aircraft of the time. In the process by which oil is cracked over a catalyst to produce it, carbon is deposited on the catalyst which soon becomes fouled and has to be regenerated. The conventional fixed bed reactors could not be developed to cope with the very high throughputs required. The techniques of fluidization provided the means whereby the cracking and regeneration reactions could be carried out and the problem of transferring the fouled catalyst from the reactor to the regenerator and back could be solved on a scale in which very large throughputs were involved. Subsequently, the chemical and oil industries have concentrated largely on applications of gas fluidized systems which exploit the advantages of the technique for handling solids in solid/fluid contacting operations. Although, as with the cracking process, the heat transfer properties of the fluidized systems have often been essential to successful operation, this aspect was usually taken for granted and less explicitly exploited. Thus it is that insufficient prominence has been given to the advantageous heat transfer properties of gas-fluidized systems which, nevertheless, find application in fields as diverse as the heat treatment of metals and power station boilers. Nevertheless, some with greater imagination have considered the possibility of a 25,000 MW nuclear powered rocket motor for which the fissile fuel would be contained within a small centrifugal cylindrical bed; the helium heat transfer medium being the fluidizing gas. The particles would be retained within the bed against the necessary high gas flowrate under the action of the centrifugal field.

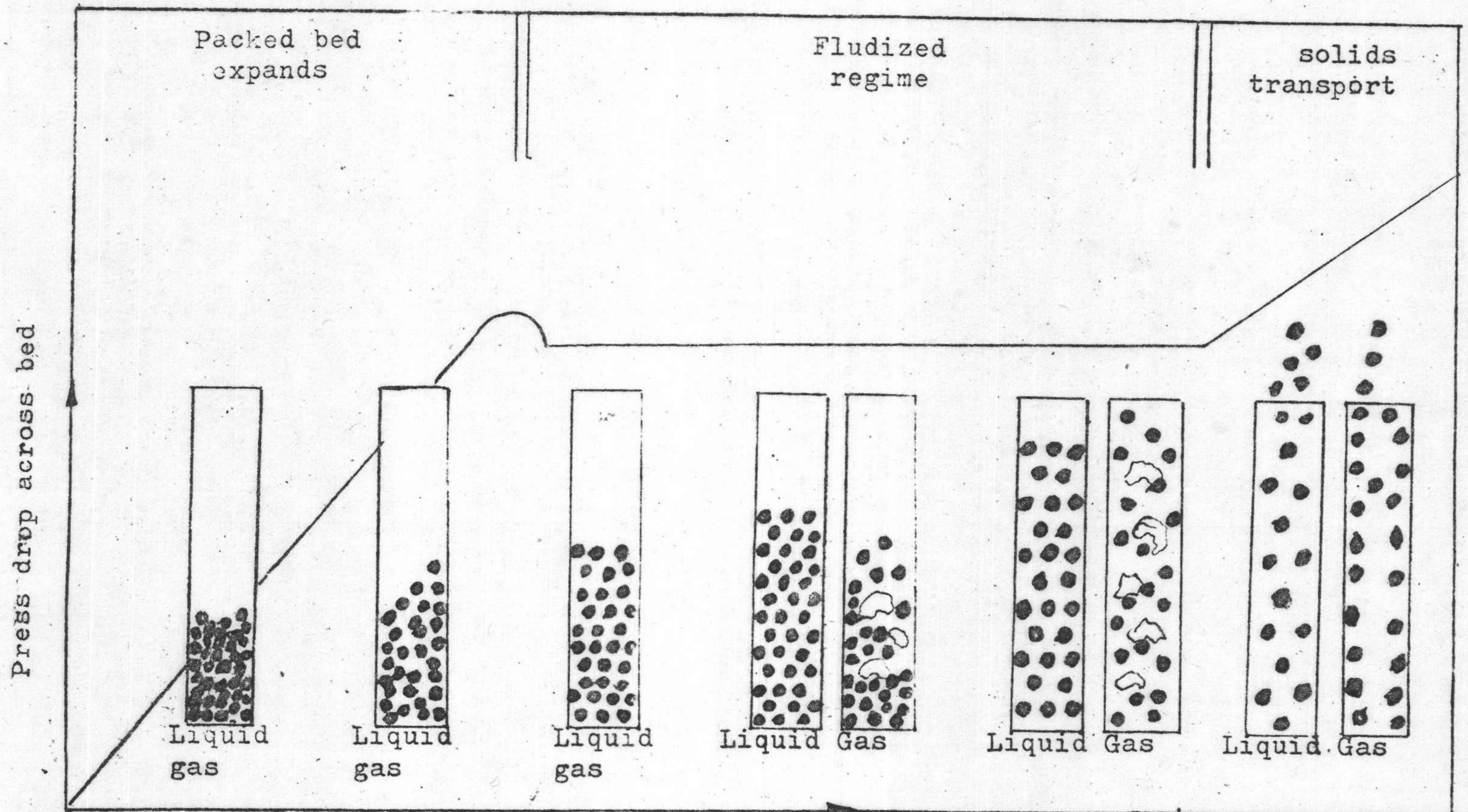
The objective of this is to describe the basis principles of fluidization and heat tranfer of gas-solid in fluidized bed.

4.1 THE PHENOMENON OF FLUIDIZATION

A bed of loose particles offers resistance to fluid flow throught it. As the velocity of flow increases fluid flowing downward through a bed of particles will tend to compact it. However, if the fluid flow is upward through the bed, the drag force will tend to cause the particles to rearrange themselves within the bed to offer less resistance to fluid flow. Unless the bed is composed of large particles (meam diameter 1 mm) the bed will expand(fig 4.1). With further increase in the upward fluid velocity, the expansion continues and a stage will be reached where the drag force exerted on the particles will be sufficient to support the weight of the particles. In this state, the fluid particle system begins to behave like a fluid and it will flow under a hydrostatic head. This is the point of "incipient fluidization". The pressure drop across the bed will be equal to the weight of the bed although it is likely that this pressure drop will be exceeded just prior to the achievement of fluidization with gas-fluidized systems because the residual packing and interlocking of particles within the bed must first be broken down (this is indicated by the lump in the stylized curve for bed pressure drop as a function of the fluid flow rate,(fig 4.1). The peak is not present for gas fluidized beds of large mean particle diameter and bed expansion only begins to occur after the point of incipient fluidization . Whilst variation or gentle tapping of the container assists the particles to compact if the fluid is flowing downward through the bed of particles it will also usually assist the uniform expansion of the fixed bed when the fluid is flowing upwards through the bed and produce a pressure loss/velocity curve without a "hump".

At the onset of fluidization the bed is more or less uniformly expanded and, up to this point, it makes little difference whether the fluid is a liquid or a gas apart from the fact that the velocity at which the bed becomes fluidized is less for a given bed particle size and density when the fluid is a liquid. Beyond this point, however, the bed behaviour is markedly different. If the fluid is a liquid, the bed continues to expand uniformly with increase in the liquid velocity. If the fluid is a gas, the uniform expansion behaviour is soon lost except with fine particles, the system becomes unstable and cavities containing little solid are formed, these look like bubbles of vapour in a boiling liquid. The value of the gas flow rate at which this happens depends on the properties of the fluidized solids, the design of the bed and particularly on the incipient fluidization and the onset of bubbling the bed is in a "quiescent" state. For smaller, irregular particles it will be larger, e.g. 2.8 has been reported for 55 micron catalyst particles. The bubbles are responsible for inducing particle circulation within the gas-fluidized bed and it is this circulation which has a most important bearing on the advantageous heat transfer properties of gas-fluidized systems. This is described in more detail in subsequent sections.

At high fluid velocities, whether the fluid is a gas or a liquid a point is reached where the drag forces are such that the particles become entrained within the fluid stream and are carried from the bed. Smaller particles tend to become entrained at lower fluid velocities than larger ones and the way that bubbles "BURST" at the surface of a gas-fluidized bed and throw a spray of particles into the space



<u>Fig</u> 4.1 Stylized representation of response of bed to upwards flow of fluid throgh it.

above the bed and can considerably affect the rate of loss of particles from a gas-fluidized bed. Between the extreme conditions of the packed bed and that of solids transport-hydraulic or pneumatic conveying, there lies the regime of fluidization. Throughout this regime the pressure drop across the bed remains approximately constant and sufficient to support the weight of the bed(fig 4.1).

4.2 INITIATION OF FLUIDIZATION AND MINIMUM FLUIDIZING VELOCITY

4.2.1 FACTORS GIVING RISE TO WELL-FLUIDIZED SYSTEMS

Despite the large amount of research which has been carried out on fluidized systems, it is not possible to predict precisely the behaviour of a system completely in terms of the physical properties of the solid particles and the fluid and the operating conditions. Furthermore, it is frequently found that those materials which are capable of giving a uniform well-fluidized bed are those for which the initiation of fluidization can represent difficult problems.

With liquid-solid systems, there are not normally any serious difficulties in initiating fludization, and once the system has been fluidized uniform conditions are usually obtained; the condition of the bed will be discussed in details (see reference No.36,37). In general, the following particles of solid and fluid are conductive to well-fluidized systems:

a) low particle density,
b) small particle size,
c) small particle range,
d) particle shape approaching spherical,
e) high fluid density.

Unfornately, those properties of the solid which make it capable of giving a well-fluidized system(small particle size and low density) are also those which tend to make the initiation of fluidization difficult. If surface forces between the particles are significant, they will be greater with small particles because of their high specific surface. If the particles are of low density, the gravitational forces tending to pull them apart will be small. Thus, small low density particles can give rise to serious channelling even though they may be fluidized well once they have been brought into suspension.

4.2.2 THE TRANSITION FROM A FIXED BED TO A FLUIDIZED BED

Seldom is there a sharp transition between a fixed bed and a fluidized bed and this is shown by the fact that the pressure drop-velocity relationship for a particulate system does not normally follow the idealized pattern. At low velocities of flow through a bed, the particles maintain a fixed orientation with respect to one another. However, there is evidence that at a velocity considerably below the minimum fluidizing velocity, some movement of the loosely held particles takes place, and the flow of the fluid is no longer simple plug flow with superimposed axial dispersion. As the fluid velocity is increased, fluidization of small localized pockets probably occurs. Thus, over a fairly small velocity range, most of the bed becomes fluidized, but it may often contain small pockets of unfluidized solids at velocities considerably in excess of those at which the bed gives the appearance of being well fluidized.

In an idealized system, the minimum fluidizing velocity is that velocity at which the bed suddenly changes from a fixed to a fluidized state. In practice,however, there may be a large transitional region and the minimum fluidizing velocity will then have no absolute significance. The problem of definition generally becomes greater for particles of wide size range.

4.2.3 DEFINITION OF MINIMUM FLUIDIZING VELOCITY

As the minimum fluidizing velocity has no alsolute siginificance, it is desirable to standardize a method of determination so that the characteristics of different systems can be compared. This is most conveniently done by using the graph of frictional pressure drop against velocity(as done in this work). If separate lines are drawn through the points for the fixed(flow decreasing) and fluidized regions, and the intermediate region are ignored(as shown in fig 4.2) the point of intersection of these two lines will give a reproducible value for the minimum fluidizing velocity, The whole of the bed is not fluidized until all the particles are fully supported in the fluid and the pressure drop becomes exactly equal to the buoyant weight per unit area ΔP_{eq}. The minimum velocity at which this occurs will be called the "full supporting velocity", U_{fs}. This is a difficult quantity to determine accurately because the pressure drop reachs its limiting value very gradully. Furthermore, the value of U_{fs} may well be influenced by the initial packing of the solids and the bed support, and may not therefore have a consistent result as shown in fig.4.2, U_{fs} is more than 50% greater than U_{mf}. The whole problem of measurement can be complicated especially in beds of small diameter, by frictional effects between the circulating solids and the walls of the container.

4.2.4 CALCULATION OF MINIMUM FLUIDIZING VELOCITY

In the absence of facilities to carry out an experimental determination, it is useful to be able to calculate approximately the value of the minimum. This may be done by using an expression for the relation between pressure drop and superficial velocity for a fixed bed and putting the pressure drop equal to the buoyant weight of the particles. This does, of course, neccessitate a knowledge of the voidage of the bed at the minimum fluidizing velocity(ϵ_{mf}). This will depend on the shape and size range of the particles, but a value of about 0.4 would be appropriate for approximately spherical particles. Various attempts have been made to relate the value of ϵ_{mf} to a shape factor for the particle, but these have not been entirely satisfactory(Narsimham, 1965; Wen and Yu, 1966).

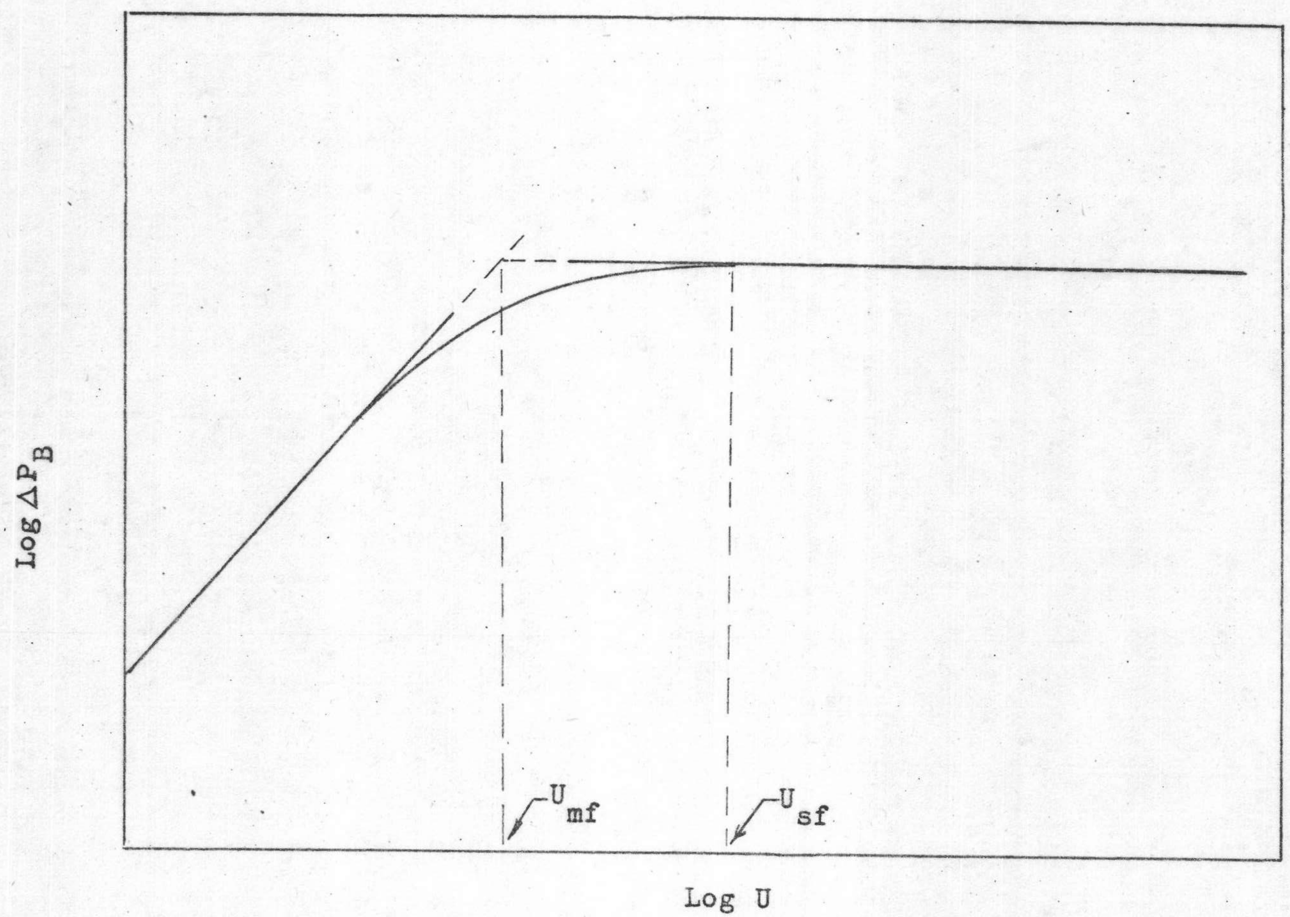


Fig. 4.2 Effect of presence of non-fluidized regions in bed

The pressure drop over the fluidized bed may be given by;

$$\Delta P_B = (\rho_s - \rho_f)(1-\epsilon)Hg \qquad (4.2.1)$$

If the bed expands, the product$(1-\epsilon)H$ will remain c nstant. Using the values of each term appropriate to the condition of incipient fluidization:

$$\Delta P_B = (\rho_s - \rho_f)(1-\epsilon_{mf})H_{mf}g \qquad (4.2.2)$$

For fine particles the pressure drop-velocity relation will be given by the Carman-Kozeny equation which will have the following form at the incipient fluidizing point:

$$U_{mf} = \frac{\epsilon^3_{mf}}{5(1-\epsilon_{mf})^2} \cdot \frac{\Delta P_B}{S^2 \mu H_{mf}} \qquad (4.2.3)$$

Hence, from equations(4.2.2) and (4.2.3)

$$U_{mf} = \frac{\epsilon^3_{mf}}{5(1-\epsilon_{mf})} \cdot \frac{(\rho_s - \rho_f)g}{S^2 \mu} \qquad (4.2.4)$$

For uniform spherical particles $S = 6/d$ and taking $\epsilon_{mf} = 0.4$

$$U_{mf} = 0.00059 \frac{d^2(\rho_s - \rho_f)g}{\mu} \qquad (4.2.5)$$

For larger particles, a more general equation such as the Blake(1922) Carman(1937), or Ergun(1952) equation must be used for the pressure drop through the fixed bed. Using the Ergun equation:

$$\frac{\Delta P_B}{H_{mf}} = 150 \frac{(1-\epsilon_{mf})^2}{\epsilon^3_{mf}} \cdot \frac{\mu U_{mf}}{d^2} + 1.75 \frac{(1-\epsilon_{mf})\rho U^2_{mf}}{\epsilon^3_{mf} d} \qquad (4.2.6)$$

Substituting for$\Delta P/H_{mf}$ from equation(4.2.2) and multiplying both sides by $\rho_f d^3/\mu^2(1-\epsilon_{mf})$:

$$\rho_f(\rho_s - \rho_f)gd^3 = 150 \frac{(1-\epsilon_{mf})}{\epsilon^3_{mf}} \frac{U_{mf} d\rho_f}{\mu} + 1.75 \frac{(U_{mf} d\rho_f)^2}{\epsilon^3_{mf} \mu^2}$$

writing

$$\frac{U_{mf}\rho_f d}{\mu} = Re_{mf} \quad \text{and} \quad \frac{\rho_f(\rho_s - \rho_f)gd^3}{\mu^2} = Ga$$

(where Ga is the Ga lileo number):

$$Ga = 150\frac{(1-\epsilon_{mf})Re_{mf}}{\epsilon^3_{mf}} + 1.75\frac{Re^2_{mf}}{\epsilon^3_{mf}} \quad (4.2.7)$$

Taking ϵ_{mf} as 0.4

$$Re^2_{mf} + 51.4Re_{mf} - 0.0366Ga = 0$$

Thus

$$Re_{mf} = 25.7(1+\sqrt{5.53*10^{-5}Ga} -1) \quad (4.2.8)$$

Equation(4.2.8) can be used for non-spherical particles if the diameter of the sphere with the same specific surface as the particles is used.

If the pressure drop through the bed is a significant proportion of the total pressure and the fluid is compressible, the velocity of the fluid will increase as it passes through the bed. The top of the bed will thus tend to fluidize at a lower flowrate of fluid than the bottom. Some workers have suggested that it is advantageous to taper the bed outwards towards the top so that the whole bed becomes fluidized at the same flowrate.

4.3 HEAT TRANSFER IN FLUIDIZED BED

As emphasized by Juveland et al.(1966), although average particle/fluid heat transfer cofficients based on the total particle surface area are often not large(of the order of 6 to 23 $W/m^{2\circ}C$) a fluidized bed of particles is capable of the very large surface area exposed by the particles(3,000-45,000 m^2/m^3). It would appear that heat transfer primarily occurs by-conduction through a gas film surrounding the individual particles. The presence of **adjacent** particles affects the thickness of the film and gas by-passing zones of the bed will adversely affect rates of heat transfer between the fluidizing gas and the particles. The degree of by-passing is dependent on the bed material, degree of fluidization, design of the apparatus and the consequent gross mixing patterns which become established so that rigorous analysis is not feasible. Nevertheless, Zabrodsky(1963) gives examples showing that the fraction of the temperature difference retained after the gas has flowed past the first "row" of particles in the bed is often very small. For particles 500 micron there is practically complete cooling(or heating) of the gas by the aggregate to the temperature of the material. Negligible temperature gradients are set up within the small particles as Dow and Jakob(1951) gas that follows the temperature of the particles rather than the other way round because of the high heat capacity of the particles. Largely, because of the high particle surface area exposed to the fluid, fluid particle heat exchange is not the limiting factor in most fluidized bed operations apart from some solid endothermic reactions and coating processes.

In this section is distinct in interphase heat transfer, between solid particles and fluidizing agent(fluid). Because of vigorous mixing of the solid phase, heat transfer between different points in the fluidized bed itself is usually very rapid; therefore, the bed is practically isothermal.

HEAT TRANSFER BETWEEN SOLID PARTICLES AND THE FLUID

4.3.1 TEMPERATURE IN A FLUIDIZED BED

The temperature of solid particles are practically constant throughout a fluidized bed but the temperature of the fluid T_f changes rapidly in the "active section" Ha (close to the distribution grid) and very slightly beyond this section(fig 4.3(a)). This typical temperature profile sometimes does not hold because the temperature of the solid particles is not constant in the immediate neighbourhood of the grid(fig 4.3(b), (c)), or due to by-passing of some of the gas(fig 4.3(c)), or because of other effects (fig.4.3(d)) etc.

4.3.2 CLASSIFICATION OF HEAT TRANSFER PROCESSES

a) Stages in the process: Heat transfer in a fluidized bed includes the following stages,

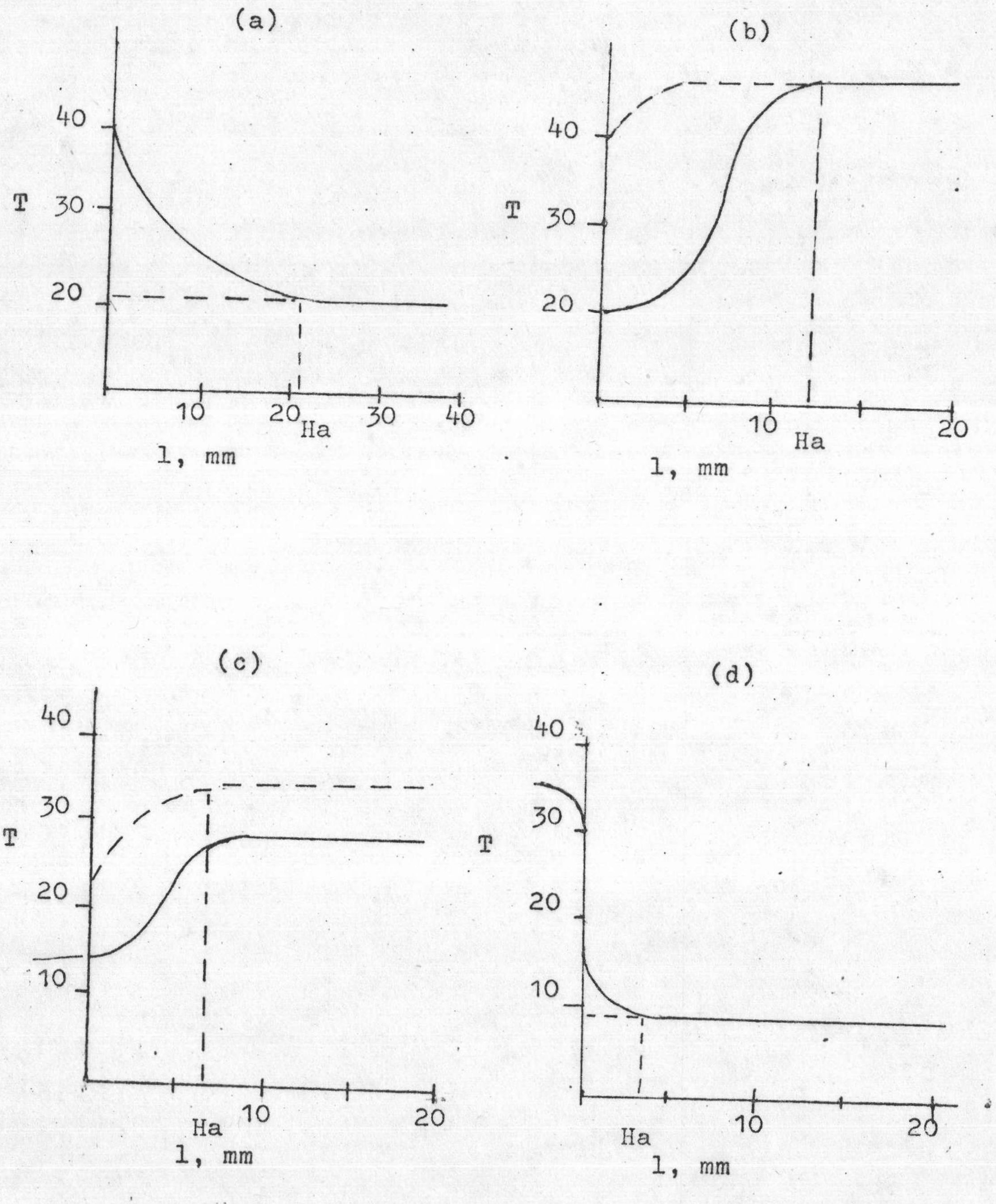


Fig. 4.3 Change of temperature vs height of fluidized bed

(i) supply of heat to(or take-off from) the system by fluid flows or solid particles(the "thermal balance problem"),

(ii) heat transfer from the fluid flow to the surface of solid particles, or conversely(the "external problem"),

(iii) propagation of heat within the solid particle(the "internal problem").

Of course, the rate of heat transfer as a whole depends on the slowest stage. The problem is clearly complex if the rates of two or all three stages are comparable.

(b) Boundaries between the individual stages of heat transfer:

A characteristic of the first stage is that when the gas leaves the fluidized bed its temperature is the same as that of the solid particles, and it is true if the bed is high enough, i.e. H Ha. When Ha is the height of active section, in which the change in fluid temperature is practically completed, besides depending on the fluid velocity and particle size and density also depends on the geometrical characteristics of the bed, on the presence of solid objects in the active zone, and on the construction of distributor,

Let us consider heat transfer under external problem conditions. Postulating that the solid particles are completely mixed(and therefore that their temperature is constant) and that the fluid flow is piston flow type, we readily obtain a relationship for the gas temperature rise T_f over the height of the bed l.

$$h_p(T_f-\Theta)\, dAs = -AUC_f\rho_f\, dT \qquad (4.3.1)$$

as $$As = A6(1-\epsilon)l/d$$

$$\frac{\Theta-T_f}{\Theta-T_f} \quad \exp\left(\frac{-6h_p(1-\epsilon)l}{UC_f\rho_f d}\right) \qquad (4.3.2)$$

where $T_f^!$ is the fluid temperature at entry to the bed.

Assuming that after a twenty fold reduction in the initial temperature difference($\Theta-T_f^!$) heat transfer is practically completed, the height of the active section, l=Ha is obtained from equation(4.3.1)

$$\frac{Ha}{d} = 0.5\frac{St^{-1}}{1-\epsilon} = 0.5\frac{RePr}{(1-\epsilon)Nu_p} \qquad (4.3.3)$$

Using the value $(Nu_p)_{min} = 2$, for air and diatomic gases (Pr=0.72), with a certain safety factor we obtain

$$\frac{Ha}{d} = 0.18\frac{Re}{1-\epsilon} \qquad (4.3.4)$$

In real system, because of incomplete particle mixing (fig 4.3(b),(c)) of some longitudinal mixing and by-passing of the gas, the value of Ha may exceed that calculated by equation(4.3.3) and (4.3.4). In any case, since small values of Re are typical of a fluidized bed, Ha rarely exceeds some tens of particle diameters. In practice Ha usually ranges from a millimetres to several centimetres, which is usually much less than the working height of the bed, so that we may put $T_f'' = \Theta$

Demarcation between the second and third stages consists in assessing the quantity $Bi = h_p Re_p / k_s$. When $Bi < 0.25$ the internal thermal resistance of the particle may be neglected as taking place with no internal temperature gradient. On other hand, when $Bi > 20$ the problem is treated as purely internal the rate of heat tranofer from the fluid to the particle does not limit the process.

In general the processes of interphase heat transfer (external problem) in fluidixed systems may be considered when the particles are sufficiently large or the bed are very shallow ($H < Ha$), and the processes may be considered as the internal problem when $B \gtrsim 20$. If h is unknown, the condition $B \gtrsim 20$ may be replaced by condition $B \simeq k_s/k_f \gtrsim 20$, which is obtained(with a certain safty factor) by proceeding from$(Nu_{pe})_{min} = 2$. This boundary may be somewhat displaced for non-spherical particles.

The mathematical description of heat transfer processes under the conditions of the internal and external problem has been developed in detail(36,37) and is not considered here.

c) Determine average particle/fluid heat transfer coefficient in fluidized bed based on the total particle surface area

Basically, two types of experimental determination of gas/particle heat transfer coefficient are possible. In one type of experiment, the temperature of the entering gas(which is usually above that of the bed) is controlled and measured. The gas exchanges heat with the fluidized bed which is cooled by internal heat exchange surfaces or by transfer of heat through the wall. The gas temperature changes rapidly in the entry zone of the bed and a simple heat balance can be written if there is plug flow of the gas and the particles within the bed are well mixed:

$$C_f G dT_f = h_p a(T_f - \Theta) dH \qquad (4.3.5)$$

Integrating this directly gives:-

$$\ln \frac{T_f - \Theta}{T'_f - \Theta} = \frac{-h_p a H}{G C_f} \qquad (4.3.6)$$

so that the slope of the semilog plot of the temperatrue variation with height through the bed will give a measure of the particle gas heat transfer coefficient.

In the steady state measurement, the inlet gas temperature is measured and maintained constant. The temperature change of the bed or of the outlet gas is then followed as it changes with time, there being no other source of heat supply or removal. In this case, the heat transferred from gas to the solids accumulates over time so that:-

$$-C_f G dT_f = h_p a(T - \Theta) dH \qquad (4.3.7)$$

$$= C_p \frac{d\Theta}{dt} dM_p$$

where dM_p is the mass of solids in the element.

If complet mixing of the gas within the bed is assumed as it was in the two references cited above, the overall heat balance can then be written as:

$$C_f G(T_f - T_f'') = h_p a(T_f'' - \theta)H \qquad (4.3.8)$$

so that

$$\ln\frac{\Delta T_f}{\Delta T_{f(initial)}} = \frac{-h_p aHC_f GA}{M_p C_p(h_p aH + C_f G)} \cdot t \qquad (4.3.9)$$

where

$$\Delta T_f = T_f' - T_f$$

and A = cross-section area of the tube

Thus, for both of these models, assumption have to made about the gas flow through the bed in order to interpret the implications of the temperature measurements and in neither is any accout taken of heat transport by solids because the heat transfer between the gas and particles is so rapid on entering the bed, the uniform gas temperature expected with a perfectly mixed system is indistinguishable from the flat tail of the exponential decay to be expected for plug flow. Accordingly, beyond the entry zone where it is particularly difficult to make measurements in any case. It is impossible to distinquish between the two extreme flow patterns. However, the type of flow occuring can be expected to approximate more closely to plug flow than to complete mixing and particularly at the higher gas flowrates, eq(4.3.5) gives the temperature gradient through the bed. It approximates to a steady state problem because the temperature of the solids changes much more slowly than that of the gas because of the solids very high thermal capacity compared to that of the gas. Then;

$$\ln\frac{\Delta T_f}{\Delta T_{f(initial)}} = \frac{C_f GA}{M_p C_p}\left(1 - \exp\left(-\frac{h_p aH}{C_f G}\right)\right) t \qquad (4.3.10)$$

The earlier work has been extensively reviewed by Barker(1965). Frantz(1961) deduced two correlations according as to whether temperature were measured with a suction thermocouple or with a bare thermocouple. He found a dependence on the 1.6th and 1.3rd powers of the Reynolds Number($d_p G/\mu_f$) respectively and also attempted to make some allowance for the

Prandtl number effect by using the liquid fluidized bed results of Sunkoori and Kaparthi(1960). Holman et al.(1965) also report results for water-fluidized bed results.

In their analysis of the published work, Kunii and Levenspiel(1969) have shown that whereas measurements based on a perfect gas mixing model are widely and exhibit a trend with bed height(broken lines in fig 4.4), those following a plug flow model present a more consistent pattern which they corelate by the equation:

$$Nu_p = 0.03\, Re_p^{1.3} \qquad (4.3.11)$$

with corresponds closely with that given by Frantz(1961)

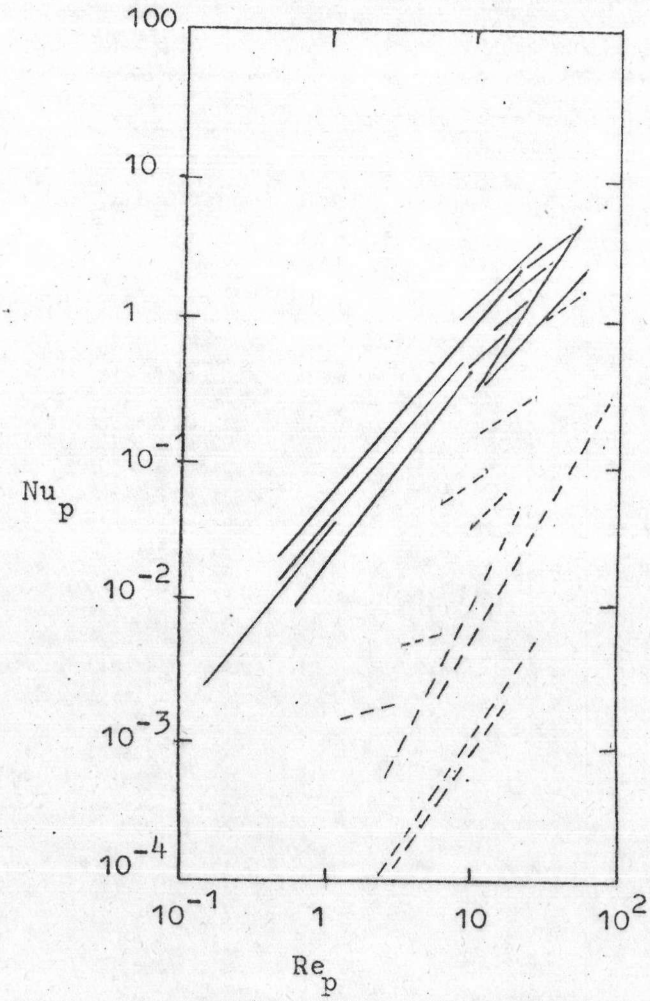


Fig 4.4 Range of reported gas to solid heat transfer coefficients Experiments interpreted using a perfect gas mixing model indicated by a broken line. (After Kunii and Levenspiel, 1969)