CHAPTER 1

THE FLUIDISED STATE

1.1. Introduction

The phenomenon of fluidisation can best be visualised in terms of a simple experiment in which a bed of solid particles is supported on a horizontal gauze in a vertical tube. Gas or liquid is then forced to flow upwards through the gauze, and so through the particle bed. This flow causes a pressure drop across the bed, and when this pressure drop is sufficient to support the weight of the particles the bed is said to be *incipiently fluidised*. Any further increase in flow causes the bed to expand to accommodate the increase. The fluidised bed thus formed has many of the properties of a liquid; its upper surface remains horizontal when the containing apparatus is tilted, and it hardly impedes the movement of objects floated on the surface. When the flow of gas or liquid through the bed is increased still further, to the point at which the flow velocity becomes greater than the free-falling velocity of the particles, then, clearly, the particles are carried out of the apparatus.

1.2. The uses of fluidised systems

A fluidised system has a number of highly useful properties, the more important being concerned with temperature control and heat transfer, continuity of operation, and catalytic reactions.

Temperature control and heat transfer

The same temperature is quickly established throughout a fluidised system because the general agitation of the particles disperses local regions of hot or cold. A fluidised bed is therefore very suitable for catalytic reactions requiring close temperature control. There is also a high rate of heat transfer to a solid object placed in the bed, so that it is a very convenient heat transfer medium; for this reason a gas-fluidised bed can be used as a constant temperature bath in which to immerse a reaction vessel that has to be at a high temperature (Adams, Gernand and Kimberlin, 1954).

Continuity of operation

A fluidised system enables solid particles to be handled essentially as a liquid, and this can be a very considerable asset in the design of a continuous process. The addition and withdrawal of solid particles from the process equipment is also facilitated.

Catalytic reactions

Fluidisation is an excellent way of bringing a gas into contact with a solid, and therefore catalytic reactions are often well suited to the technique. Fluidised beds first became of major importance through the development of a fluidised process for cracking heavy hydrocarbons into petroleum spirit (Murphree, Brown, Fischer, Gohr and Sweeney, 1943).

The applications of fluidisation fall into two broad categories: (i) chemical reactions and catalysis, and (ii) physical and mechanical processes. The drying of solid particles (Jobes, 1954) is an example of (ii). Apart from fluid catalytic cracking, the chemical uses of fluidisation have also included the Fischer–Tropsch process (Hall and Crumley, 1952), the roasting of pyrites (Thompson and MacAskill, 1955), and the reduction and fluorination of uranium (Hawthorn, Shortis and Lloyd, 1960). This brief list of applications is only intended to be illustrative; it is by no means exhaustive. Zenz and Othmer (1960) describe some 20 uses of fluidisation, and they refer to 50 more.

The disadvantages of fluidised systems

A fluidised bed is not suitable for all fluid–solids processes, and considerable difficulties have arisen in the past when some of its disadvantages have not been clearly recognised, for instance:

(i) The quick equilibration of temperature in a fluidised system means that it is unsuitable for a reaction which is best carried out in a reactor giving a temperature gradient along the reaction path.

(ii) The ease with which particles can be fluidised can vary enormously, and thus a fluidised process is usually precluded for particles which do not flow freely or which agglomerate (e.g. waxes).

(iii) The bubbles of gas characteristic of most gas–solids beds can cause both chemical and mechanical difficulties. For instance,

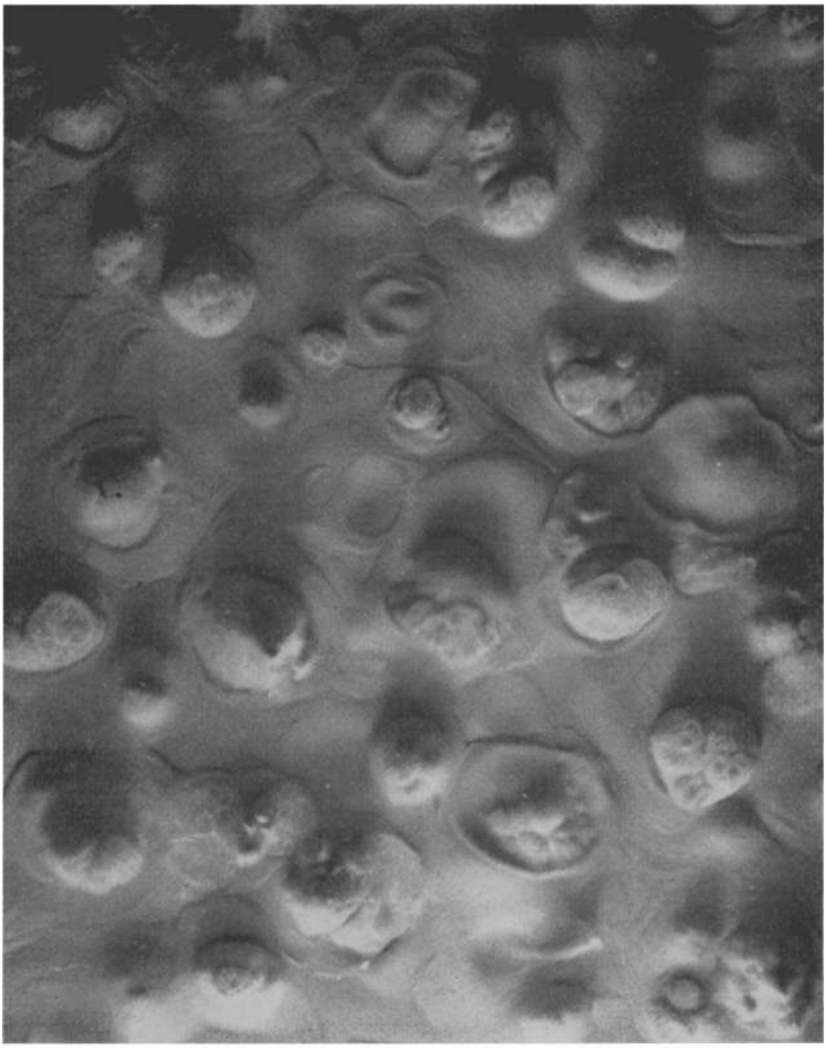

Plate I. Aggregative fluidisation showing bubbles breaking surface at the top of a bed of sand fluidised by air.

(*Facing p.* 3)

in a gas–solids reaction, it is possible for some of the gas in the bubbles to by-pass the particles altogether, and so the overall efficiency of contact is lowered. The bubbles can also cause mechanical buffeting sufficient to cause breakage when, for example, weak fabric or paper is dried in a hot fluidised bed.

1.3. Definitions

(*a*) *Incipient fluidisation*

Typical results for a gas-fluidised bed of particles are shown in fig. 1; the pressure drop across the bed, and its height, are plotted as functions of the superficial gas velocity. There is marked hysteresis, so that for slowly increasing flow the curves *A* are generated, while the curves *D* are for slowly decreasing flow; the amount of hysteresis depends upon the degree of consolidation of the original bed. At point *B*, the overall pressure drop is slightly more than enough to support the weight of the particles, owing to the wedging action within the bed. A slight increase in flow above point *B* frees the particles, the pressure drop becomes just enough to support their weight, and consequently point *C* is usually defined as the point of *incipient fluidisation*, the superficial gas velocity being U_0 and the voidage fraction, ϵ_0. If the flow rate is now slowly decreased from point *C*, the particles are more loosely packed, and consequently the bed height is greater and the pressure drop smaller, as shown by curves *D*. By starting with a bed of loosely packed particles (one that has just been fluidised) it is possible to get the curves for increasing flow-rate to coincide with the curves for decreasing flow-rate; the hysteresis is then eliminated.

Below incipient fluidisation, results similar to those in fig. 1 are obtained for liquid-fluidised beds.

(*b*) *Aggregative fluidisation*

For a gas-fluidised bed in which the gas velocity is greater than U_0, some of the gas may pass through the bed as bubbles, and these can be seen to burst when they reach the top surface of the bed as shown in Plate I. This is commonly known as *aggregative fluidisation*, and usually occurs when solids are fluidised by gases. The bubbles agitate the bed, and consequently its height fluctuates as indicated in region *E* of fig. 1.

(*c*) *Particulate fluidisation*

For a liquid-fluidised bed in which the velocity is greater than U_0, the bed height increases with velocity, but there are usually no marked fluctuations in level, the particles spacing themselves evenly so that the liquid passes smoothly through the interstices without the formation of bubbles. This is known as *particulate fluidisation.*

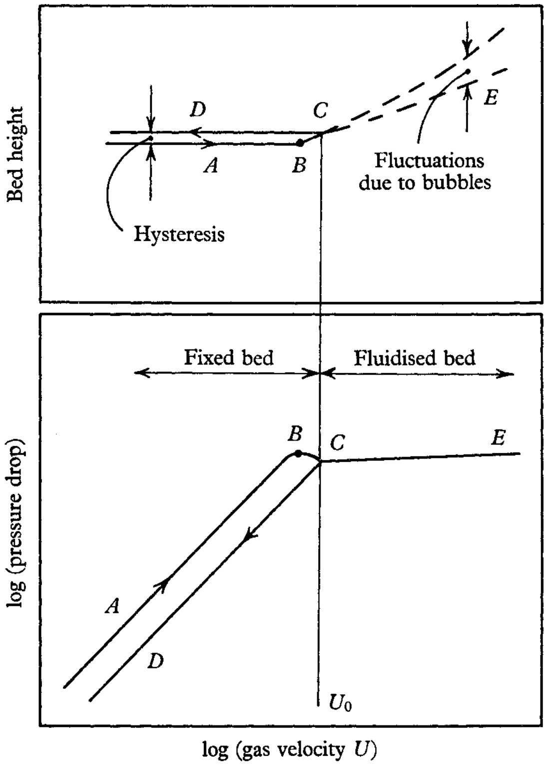

Fig. 1. Typical curves for a gas-fluidised bed of particles of approximately uniform size.

The difference in behaviour between gas- and liquid-fluidised beds may be readily observed by fluidising glass beads of about 0·25 mm diameter in turn with air and water; experimental data on both types of fluidisation are given by Wilhelm and Kwauk (1948).

Although aggregative fluidisation is normally found with gas–

solids systems and particulate behaviour with liquid–solids systems, there are exceptions to this generalisation. For example, Wilhelm and Kwauk (1948) found that a bed of lead shot fluidised by water behaved aggregatively (i.e. water bubbles were present); and Leung (1961) has observed particulate behaviour when fluidising light resin particles with carbon dioxide under pressure. The detailed relationship between aggregative and particulate fluidisation is considered in Chapter 5.

1.4. The range of the fluidised state

The general appearance of an aggregative bed as the flow of fluid is increased is shown diagrammatically in fig. 2. First, the bed

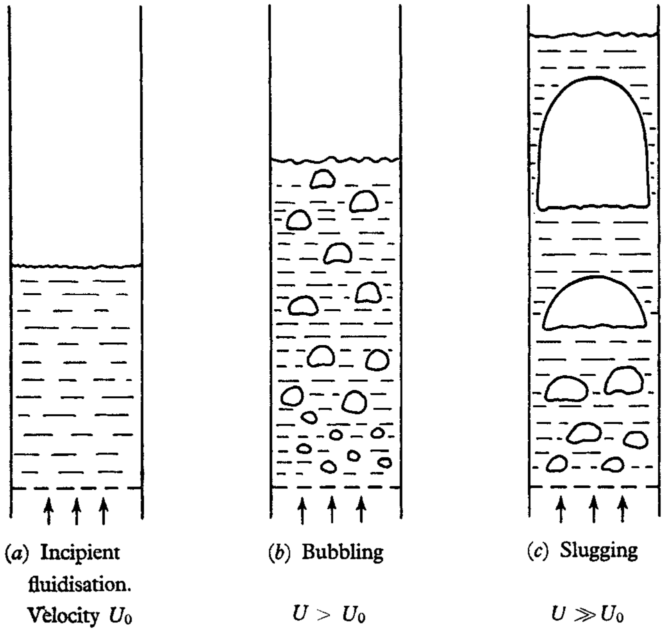

Fig. 2. Flow régimes in an aggregative fluidised system.

expands from a fixed bed to the point of incipient fluidisation and then, as the flow-rate is further increased, bubbling begins. With still greater flows the bubbles grow and appear more frequently, until their frontal diameters are equal to the diameter of the containing apparatus. The bed, shown in fig. 2 (*c*), is then said to be *slugging*. A further increase of flow to the bed carries the particles out of the apparatus. The voidage fraction is then high (over 0·80), and the phenomenon is that of pneumatic transport, or *dilute-phase fluidisation.*

The general appearance of particulate systems is shown in fig. 3. In this case the bed is always homogeneous, with uniform expansion to take up the increased flow, dilute-phase fluidisation being reached without the formation of bubbles. The expansion of a particulate bed is considered in detail in § 1.6, p. 15.

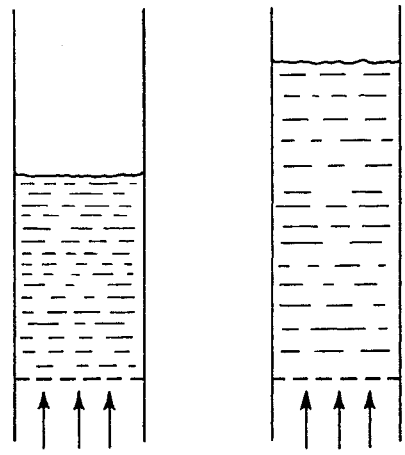

Fig. 3. Flow régimes in a particulate fluidised system.

Most applications of fluidisation at present are concerned with gas–solids systems rather than with liquid–solids systems; and a quantitative consideration of the bubbling and slugging regions of gas-fluidisation takes up the major part of this book. However, there are other important aspects of gas-fluidised behaviour which still await analytical attention, for example:

(*a*) *Channelling and spouted beds*

Channelling is the phenomenon observed when a disproportionately large amount of the fluidising fluid follows one or two particular paths through the bed. This is often a marked characteristic of a bed of very fine particles, or of sticky or waxy particles which tend to agglomerate.

A spouted bed may be seen as an extreme form of channelling, in which the fluidising fluid takes only one (usually central) path up the bed. An account of spouted beds has been given by Leva (1959).

(*b*) *Distributor design*

Little systematic work has been reported on the influence of the bed support on fluidised behaviour; although many data have been collected by industry on the design of distributors for specific purposes. Grohse (1955) discovered that the point of incipient fluidisation was more reproducible with a porous plate distributor than if the bed was supported on either a 300 BS mesh ($\simeq$ 0·005 cm spacing) screen or a multi-orifice plate. Rowe and Stapleton (1961) observed the behaviour of a gas-fluidised bed of 12 in diameter fitted in turn with a 'bubble-cap' distributor, a conical distributor, and a porous plate. They too found that the porous plate distributor allowed a more even expansion of the bed than the other distributors, and that it gave rise to more—and smaller—bubbles. They also found that the distributor design affected the behaviour of the bed over most of its height.

(*c*) *Baffles*

The use of 'baffles' to promote even fluidisation is another subject on which published data are sparse. It was discovered early in the study of fluidised beds that internal baffles tend to break up bubbles and, as a result, provide smoother operation. A strikingly successful example of this was the smooth fluidisation of a bed 80 in deep and 1 in diameter by Hall and Crumley (1952). The baffle arrangement consisted of discs of 10 BS mesh ($\simeq$ 0·15 cm) steel gauze, dished slightly convex to the gas flow, attached at 2 in intervals to a vertical rod in the centre of the bed. This type of baffle impeded bubble coalescence over the full depth of the bed, and thus the slugging of the bed, which occurred without the baffles, was avoided.

Beck (1949), Massimilla and Westwater (1960), and Volk, Johnson and Stotler (1962), have also investigated baffled beds and, aside from the effect on bubbling, two other general points emerge from their work:

(i) The average linear velocity of the particles in the bed is reduced, in some cases to 10–20 % of the velocity in an unbaffled bed; and

(ii) heat transfer between the bed and the containing walls is adversely affected, possibly as a consequence of (i).

At the present time the precise design of baffles for a particular system allows considerable scope for trial-and-error ingenuity, and so the patent literature on this subject is more extensive than the academic (Hassett, 1963).

(*d*) *The fluidisation of very large particles*

Squires (1962) distinguishes between beds with particles of a size smaller than about 20 BS mesh ($\simeq$ 0·08 cm) ('fluid beds') from those with particles larger than about 10 BS mesh ($\simeq$ 0·15 cm) ('Teeter beds'). The work described in this book is concerned in the main with gas- and liquid-fluidised systems containing smaller particles of less than 10 BS mesh.

1.5. The incipient fluidising velocity U_0

There is no doubt that the best way to determine U_0 is to measure it. The method is to measure the pressure drop through the bed of particles as a function of the flow-rate for slowly increasing and then slowly decreasing flow-rates. The results give curves of the kind shown in fig. 1, and U_0 is the velocity at point C, though this point is not always well defined. An experiment of this kind can be done on the laboratory scale and the results can be applied with reasonable confidence to a large-scale plant, provided the pressure and temperature are the same.

Nevertheless, it is useful to be able to estimate U_0 from first principles, both for design purposes, and because the plant conditions may be difficult to simulate in the laboratory if they involve high pressure or temperature. Since the individual particles derive mutual support from one another for $U < U_0$, the problem of predicting U_0 is essentially the problem of finding the flow which will produce a pressure drop through the fixed particle bed equal to its weight per unit cross-section. For this purpose a short account will be given of the theory of flow through fixed beds of particles.

(*a*) *Pressure drop through a fixed bed of particles*

In this section we shall consider the flow of fluid through a fixed bed of particles under the influence of a uniform pressure gradient.

The following theory was originated by Kozeny (1927) and Carman (1937), and to take the simplest view of it, the idea is to obtain an equivalence between the tortuous passages through the packing and a single passage having the same volume and surface

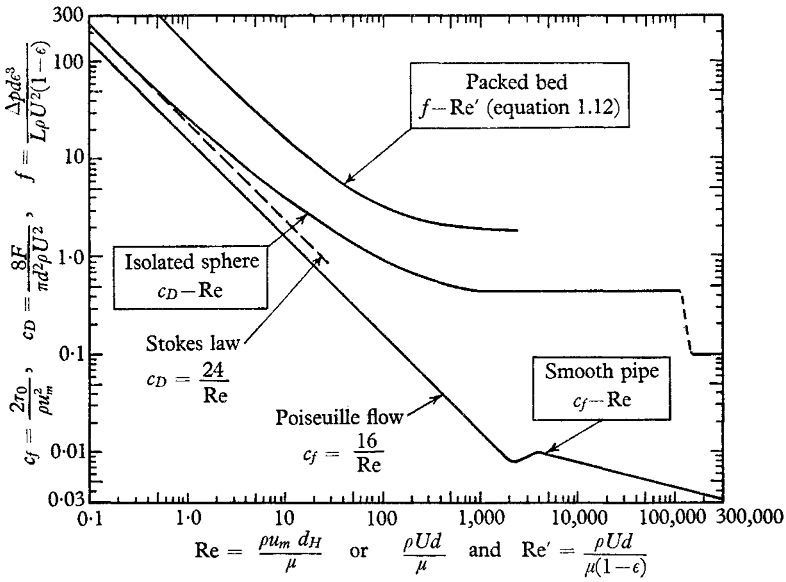

Fig. 4. Relations between pressure drop and flow for a smooth pipe, an isolated sphere, and a packed bed.

area. When fluid flows through a parallel-sided passage, the friction coefficient, c_f, is uniquely related to the Reynolds number Re, where

$$c_f = 2\tau_0/\rho u_m^2, \tag{1.1}$$

and

$$\text{Re} = \rho u_m d_H/\mu, \tag{1.2}$$

τ_0 being the wall shear stress, ρ and μ the fluid density and viscosity, u_m the mean velocity in the passage and d_H its hydraulic mean diameter. Fig. 4 shows the relation between c_f and Re for a circular pipe (e.g. Prandtl, 1952, p. 165) and a similar relation between