# 1 Introduction

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This introductory chapter follows the contours, and in some cases even the exact wording, of Chapter 1 in *Spouted Beds*,[1] the only book prior to the present publication that deals *exclusively* with this subject. Indeed, the current venture was originally to be a revised version of that 1974 book by the present editors. However, after writing the first draft of this chapter for the revision, we realized that the breadth and variety of work on spouted and spout-fluid beds since 1974 required input from a wide range of authors for coverage to be completed in a finite time. Changes in the subsequent draft were mainly with respect to layout of chapter topics (Section 1.6). Despite advances since 1974, the earlier book of that year remains a repository of useful information not available in this volume or elsewhere.

## 1.1 The spouted bed

Consider a vessel open at the top and filled with relatively coarse particulate solids. Suppose fluid is injected vertically through a centrally located small opening at the base of the vessel. If the fluid injection rate is high enough, the resulting high-velocity jet causes a stream of particles to rise rapidly in a hollowed central core within the bed of solids. These particles, after being carried somewhat beyond the peripheral bed level, rain back onto the annular region between the hollowed core and the column wall, where they slowly travel downward and, to some extent, inward as a loosely packed bed. As the fluid travels upward, it flares out into the annular region. The overall bed thereby becomes a composite of a dilute phase central core with upward-moving solids entrained by a cocurrent flow of fluid, and a dense phase annular region with countercurrent percolation of fluid. A systematic cyclic pattern of solids movement is thus established, giving rise to a unique hydrodynamic system that is more suitable for certain applications than more conventional fluid-solid configurations.

This system is termed a *spouted bed*, the central core a *spout*, the surrounding annular region the *annulus*, and the solids above the bed surface entrained by the spout and then raining down on the annulus are designated as the *fountain*. To eliminate dead spaces at the bottom of the vessel, it is common to use a diverging conical base topped by a

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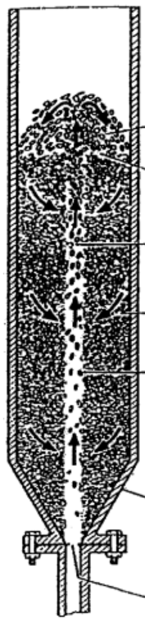


**Figure 1.1.** Schematic diagram of a spouted bed. Arrows indicate direction of solids motion.

cylindrical column, with fluid injection at the truncated apex of the cone (Figure 1.1). The use of an entirely conical vessel is also widely practiced. An example of each is shown in Figure 1.2.

Solids can be added into and withdrawn from spouted beds, so, as for fluidized beds, spouting can be performed both batchwise and continuously with respect to the solids, although batchwise operation is more likely to be adopted in spouting than in fluidization applications. The solids may be fed into the bed at the top near the wall, so they join the downward moving mass of particles in the annulus (Figure 1.3a). Alternatively, the solids may enter by being entrained by the incoming fluid (Figure 1.3b). Because the annular solids are in an aerated state, the solids may be readily discharged through an overflow pipe at the top of the bed, as in Figure 1.3b, or through an outlet at a lower level, such as in the conical base, as in Figure 1.3a. When solids are fed with the inlet fluid and discharged at the top of the bed, as in Figure 1.3b, they are effectively being conveyed vertically, in addition to being contacted with the fluid.

As in fluidization, the fluid in spouting applications, and therefore in research studies, is more likely to be a gas than a liquid. Except where otherwise indicated, most of this book therefore deals with gas spouting, with only one chapter devoted to the mechanics of spouting with a liquid. Although the spouted bed was originally developed as a substitute for a fluidized bed for coarse, uniformly sized particles to overcome the poor quality of gas fluidization obtained with such particles, some of its unique characteristics, including the cyclic recirculation of the solids, have proved valuable, making spouted beds capable of performing certain useful operations more effectively than fluidized beds with their more random solids motion.

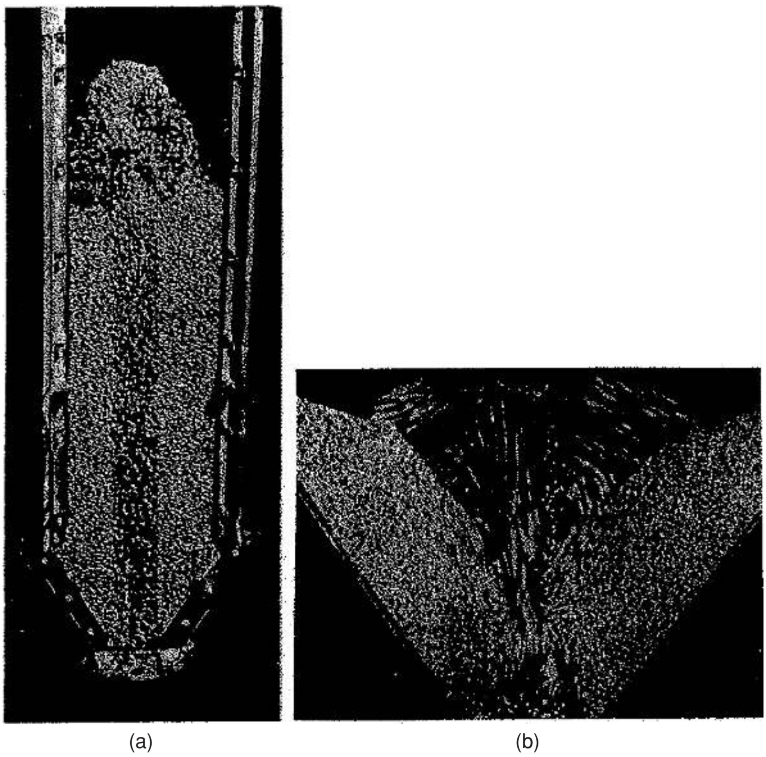

**Figure 1.2.** (a) Spouting in a conical-cylindrical vessel,[1,2] courtesy of National Research Council of Canada. (b) Spouting in a conical vessel.[1,3]

The circular cross-section devices shown in Figure 1.2, now commonly referred to as conventional spouted beds (CSBs), are the principal subject of this book. A wide variety of nonconventional spouting techniques and equipment has also been reported in the literature, and some of these are described in subsequent chapters.

## 1.2 Brief history

Although the terms *spouting* and *spouted bed* were coined at the National Research Council (NRC) of Canada by Gishler and Mathur (1954)[6] to describe the type of dense-phase solids operation (or CSB) depicted by Figures 1.1 through 1.3, a dilute-phase solids operation activated by an air jet for ore roasting (Robinson, 1879),[7] or coal combustion (Syromyatnikov, 1951)[8] predates the CSB. Such dilute solids spouting, referred to as "air-fountaining"[3] and, more frequently, as "jet spouting,"[9] are alluded to in several succeeding chapters.

The original spouted bed was developed in 1954 as an alternative method of drying to a badly slugging fluidized bed of moist wheat particles.[10] Because of the vigorous particle circulation, much hotter air than in conventional wheat driers could thus be used without damaging the grain.[4] Realizing that the technique could have wider application, the

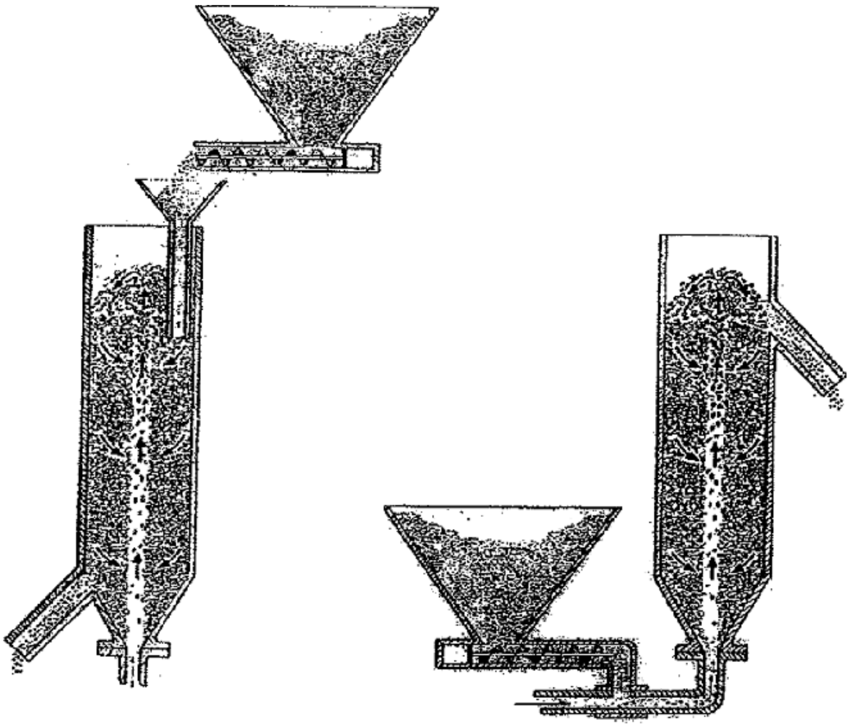


**Figure 1.3.** Continuous spouting operation. (a) Solids fed into annulus from above, after Mathur and Gishler.[4] (b) Solids fed with incoming gas into spout, after Manurung.[5]

NRC group studied the characteristics of spouted beds for a variety of solid materials, with both air and water as spouting media.[2] On the basis of this preliminary study, they asserted that "the mechanism of flow of solids as well as of gas in this technique is different from fluidization, but it appears to achieve the same purpose for coarse particles as fluidization does for fine materials." This assertion is an understatement because, as already noted, spouting has some unique characteristics, different from those of fluidization.

Early research papers continued to originate from Canadian sources – the NRC Laboratories in Ottawa, the Prairie Regional Laboratories of the NRC, and the University of Ottawa. It was only after the publication in 1959 of *Fluidization* by Leva,[11] which summarized the NRC work in a chapter titled "The Spouted Bed," that interest spread to other countries. The translation of Leva's book into Russian in 1961, followed by the publication in 1963 of a book by Zabrodsky (English translation, 1966)[12] appears to have triggered considerable activity on the subject at several research centers in the Soviet Union, with particular emphasis on spouting in conical vessels, rather than in cylindrical columns with conical bases. By the time the book by Mathur and Epstein (1974)[1] was in print, some 250 publications, including patents, on spouted beds had appeared from countries as diverse as Australia, Canada, France, Hungary, India, Italy, Japan, Poland,

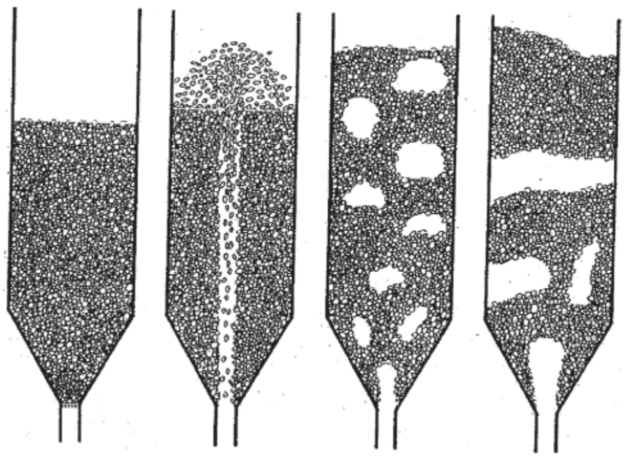


**Figure 1.4.** Regime transitions with increasing gas flow.

Rumania, the UK, the United States, the USSR (most prolifically), and Yugoslavia. At present, the number of publications on the subject exceeds 1300, and to contributors from the above countries (including mainly Belarus, Russia, Ukraine, and Uzbekistan from the former Soviet Union) should now be added those from Brazil, Chile, China, Mexico, the Netherlands, New Zealand, Spain, and no doubt others.

The first commercial spouted bed units in Canada were installed in 1962 – for the drying of peas, lentils, and flax – that is, drying of the granular particles undergoing spouting. Since than, units have been built in other countries for a variety of other drying duties, including evaporative drying of solutions, suspensions, and pastes in a spouted bed of inert particles, as well as for solids blending, cooling, coating, and granulation. Most commercially successful spouted bed installations have involved such physical operations, but a wide variety of chemical processes have also been subjected to laboratory- and bench-scale spouting investigations; some of these, including electrolysis in a liquid-spouted bed,[13] show considerable promise for further development.

## 1.3 Flow regime maps encompassing conventional spouting

Spouting, which is visually observable in a transparent column with a fully circular cross-section by virtue of the rapidly reversing motion of particles in the fountain and the relatively slow particle descent at the wall, occurs over a definite range of gas velocity for a given combination of gas, solids, vessel geometry, and configuration. Figure 1.4 illustrates schematically the transition from a quiescent to a spouted bed, and hence often to a bubbling and a slugging bed, as the superficial gas velocity (gas volumetric flow rate/column cross-sectional area) is increased.

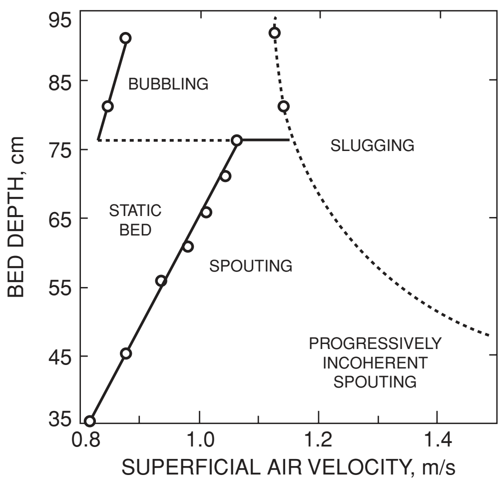


**Figure 1.5.** Flow regime map for wheat particles (prolate spheroids: 3.2 mm × 6.4 mm, $\rho_p =$ 1376 kg/m$^3$). $D_c = 152$ mm, $D_i = 12.5$ mm. Fluid is ambient air. After Mathur and Gishler.[2]

Those transitions can be represented quantitatively as plots of bed depth $H$ versus superficial gas velocity $U$, or regime maps (sometimes referred to as "phase diagrams"), examples of which are given in Figures 1.5 and 1.6. The line representing transition between a static and an agitated (spouted or bubbling-fluidized) bed is more reproducible in the direction of decreasing velocity than vice versa, the resulting static bed then being in the reproducible *random loose packed* condition.[14] Figure 1.5 shows that, for a given solid material contacted by a specific fluid (at a given temperature and pressure) in a vessel of fixed geometry, there exists a maximum spoutable bed depth (or height) $H_m$, beyond which spouting does not occur, being replaced by poor-quality fluidization. In Figure 1.5, $H_m$ is represented by the horizontal lines at a bed depth of 0.76 m. The minimum spouting velocity, $U_{ms}$, is represented in the same figure by the inclined line that terminates at $H_m$, at which $U_{ms}$ can be up to 50 percent greater[15] than the corresponding minimum fluidization velocity, $U_{mf}$, although less difference between these two critical velocities has usually been found.[16,17] Figure 1.6 shows a gas inlet, particle, and column diameter combination for which spouting does not occur. For the same column and particles, but with a smaller gas inlet ($D_i = 12.5$ mm instead of 15.8 mm), coherent spouting could be obtained.[2]

Becker[16] attempted a more generalized regime diagram by plotting upward drag force (as measured by frictional pressure drop, $-\Delta P_f$) normalized with respect to downward gravitational weight of solids against $U/U_m$, with $H/H_m$ as a parameter, whereas Pallai and Németh[15] simply plotted $-\Delta P_f$ against $U/U_{mf}$ with $H$ as a parameter. The amount of information provided by these procedures for any given system of fluid, solids, and column geometry is considerable, but the applicability to other systems is quite limited, given the complexity of the regime transitions.

A typical spouted bed in a cylindrical or conical-cylindrical vessel has a depth, measured from the fluid inlet orifice to the surface of the loose-packed static bed or the

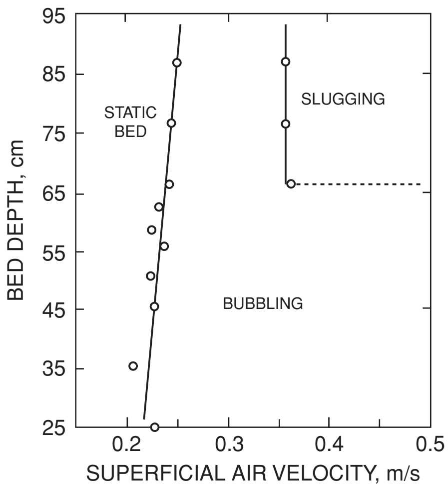


**Figure 1.6.** Regime map for Ottawa sand ($d_p = 0.589$ mm). $D_c = 152$ mm, $D_i = 15.8$ mm. After Mathur and Gishler.[2]

spouted bed annulus, of at least one-half the cylinder diameter. If the bed is much shallower, the system differs hydrodynamically from true spouting, and any generally formulated principles of spouted bed behavior would not be expected to apply. A minimum spoutable bed depth has, however, not been precisely defined or investigated, except in the case of conical beds,[18] nor have any detailed studies been made about the maximum spouting velocity at which transition from coherent spouting to either bubbling fluidization or slugging occurs. For most practical purposes, however, there is usually sufficient latitude between the minimum and maximum spouting velocity that the fluid flow can be amply increased above the minimum without transition to fluidization.

## 1.4 Nonaxisymmetric geometries of spouted beds

The photographs of Figure 1.2 and, in most cases, illustrations such as those of Figure 1.4 are obtained from the transparent flat face of semicircular conical or conical-cylindrical vessels, with aligned semicircular gas inlets. Qualitatively, what is seen and what is measured in such half-sectional columns is comparable to what can be detected by various techniques (e.g., piezoelectricity, stop-flow, laser-Doppler anemometry, optical fiber probes) and what is measured in full columns. Quantitatively, however, although some characteristics of spouted beds are well matched in the fully cylindrical and semicylindrical (labeled "full" and "half," respectively) columns, others show significant differences.

Experimental data required to construct a regime map such as Figure 1.5 are, by and large, quite similar when obtained for either a half or a full column. Thus, measured values of $U_{ms}$ from half-columns have been shown to be equal to,[17,19] or at most only 10 percent greater than,[20] $U_{ms}$ for full columns, whereas $H_m$ from half columns has been measured as approximately equal to,[19] and never more than 32 percent smaller than,[20] $H_m$ for full columns. Bed pressure drops and pressure profiles during spouting have also been found to be quite similar for the two-column geometries, as have pitot tube measurements of radial and longitudinal gas velocity profiles,[21] except in the immediate vicinity of the half-column flat wall.

The situation is different for details of particle motion and spout geometry. All walls tend to retard vertical particle motion relative to what occurs a few particle diameters from the wall (typically by about 20 percent or more for smooth walls and smooth spherical particles), whether the motion is downward in the annulus at flat or circular walls[22] or upward in the spout at the flat wall of a half column (measured by laser-Doppler anemometry[23] or by fiber optic probes[24]). Consequently, cinephotographic measurements at the transparent flat face, which are biased toward particles very near the face, tend to underestimate particle velocities in the spout, especially at or near the axis of a full column. Particle velocities in the fountain are, however, well matched between half and full columns.[23] As for the annulus, although the downward particle velocity at the flat wall has sometimes been reported as being equal to that at the corresponding half-round wall, as well as to that at the round wall of a full column,[2,19,20] Rovero et al.,[22] by a carefully monitored stop-flow technique using tracer particles, showed that particle trajectories at the half-column flat face are not representative of full column behavior at the same plane, and that only the middle 60° sector of a half-column is representative of a full column as far as particle velocity is concerned.

Spout geometry is more contentious. Mikhailik,[25] based on a piezoelectric probe for measuring particle velocity in the spout of a fully circular bed, coupled with visual observation of spouting in a larger diameter transparent semicircular column, concluded that the shape and size of the spout were the same in both types of column. However, the longitudinally uniform cylindrical spout shape reported by Mikhailik is at odds with the recorded observations of most other workers, and the criterion used for size equality, given the inequality of vessel diameters and the absence of geometrical similarity between the two systems, was the adherence of measured spout diameters in both cases to the same empirical equation relating average spout diameter to operating system parameters. A more convincing study by He et al.,[26] using optical fiber probes for voidage and particle velocity in geometrically matched semicylindrical and fully cylindrical cone-based columns, showed cross-sectional modification of the full column circular spout to an approximate semiellipse (with the larger axis perpendicular to the flat wall) in the half column. There was longitudinal distortion from monotonic divergence of spout width in the upward direction, both for the full column and for the half-column at 90° to the flat wall, to an S-shape along the flat wall itself. Average spout diameters in the full column were underestimated by about 35 percent if based on measurements along the flat face of the half-column.

The corners of half columns contribute most to distortion of full-column spouted bed behavior.[22] 90°, 60°, and 30° sector columns, which produce ever narrower corners as the sector angle decreases, although they give rise to spouted bed behavior of sorts, do not in most important respects compare well to measurements obtained in full or even half columns, and show increasing instability as the sector angle decreases.[27,28] Unlike half-sectional beds, which retain their value as qualitative and partly quantitative visual simulators of full beds, sector beds are not recommended for these purposes.

As in the case of fluidized bed research, "two-dimensional" beds – that is, beds of particles encased between parallel transparent vertical planes a few millimeters apart with a fluid inlet that spans this thickness – have also been used in spouted bed research to give qualitative insight into three-dimensional behavior. In the case of spouted beds, it has been claimed that such slot-rectangular devices can be scaled up readily – for instance, for grain drying purposes.[29] This claim is discussed in Chapter 17.

## 1.5 Spouted beds in the gas–solid contacting spectrum

### 1.5.1 The spectrum

Gas–solid contacting systems may be broadly classified as (1) nonagitated, (2) mechanically agitated, and (3) gas-agitated. Categories (1) and (3), unlike (2), have no moving parts.

Fixed and moving packed beds, which belong to the first category, are applicable to processes that do not call for maintaining high rates of heat and mass transfer between the gas and the coarse solids, and in which uniformity of conditions in different parts of the bed is either not critical or not desirable. In a fixed bed, solids cannot be continuously added or withdrawn, and treatment of the gas is usually the main objective – for example, to recover solvent vapors by gas adsorption, to dry a gas, or to carry out a gas-phase reaction catalyzed by the particles, in which the catalyst has an extended life (limited deactivation). A moving bed provides continuous flow of solids through the reaction zone. Its application therefore extends to chemical and physical treatment of solids in such processes as roasting of ores, calcining of limestone, and drying and cooling pellets and briquettes. In both fixed and moving beds, the gas movement is close to *plug flow*, a feature that is almost always advantageous for chemical reactions, where axial dispersion tends to reduce conversions and yields.

Limited agitation can be imparted to the solids by mechanical means, either by movement of the vessel itself, as in rotary driers and kilns, or by internal impellers. In either case, most of the material remains in a packed-bed condition, but the differential movement of particles improves contacting effectiveness because fresh surface is continually exposed to the gas. Also, the blending of solids by agitation levels out interparticle gradients in composition and temperature. Mechanical systems are used mainly for processes involving solids treatment, such as drying, calcining, and cooling, but are unsuitable for processes that require the gas to be treated uniformly.

In gas-agitated systems, such as fluidized beds or solids transport systems, intense agitation is imparted to each solid particle by the action of the gas stream. In fluidization, whether of the dense, noncirculating (externally), particulate, or bubbling bed variety at one extreme, or dilute, circulating, or fast-fluidization at the other, the large surface area of the fine well-mixed solids gives rise to high particle-to-gas heat and mass transfer rates; coefficients of heat transfer to or from the vessel walls or submerged surfaces (e.g., of heat exchangers) are considerably higher than those at the same superficial gas velocity in the absence of solids or in the presence of a fixed bed of solids. Continuous operation of fluidized beds can be readily accomplished because solids are easily added to, and withdrawn from, such beds. In fluidization, the average residence time of the particles in the bed can be controlled by adjusting the solids feed rate and the holdup of the bed. Because of these basic features, fluidized beds are the preferred contacting method for many processes, including chemical reactions (both catalytic and noncatalytic) involving the gas, as well as physical treatment (e.g., drying) of the solids.

Dilute solids transport (e.g., pneumatic conveying) may be more suitable than fluidization for some gas–solid contacting operations. In such systems the contact time between a given particle and a gas is very short – no more than a few seconds – because of very high gas velocities. Intense turbulence facilitates high coefficients of heat and mass transfer, but the extents of heat and mass transfer to or from the particles is limited by the small residence time of the particles in the reaction zone. Dilute solids transport systems are especially suitable for processes that are surface-rate controlled rather than internal-diffusion controlled with respect to the solid particles. Examples of established applications are combustion of pulverized coal, flash roasting or smelting of metallic sulfides, and flash drying of sensitive materials that can tolerate exposure to heat for only a few seconds.

Because the annular region of a spouted bed, which contains most of the solids, is virtually a moving packed bed, with counterflow of gas, spouted beds have sometimes been categorized as moving-bed systems.[30] However, the dilute solids transport that characterizes the spout region is crucial to overall spouted-bed performance and, inasmuch as bed solids recirculate many times, therefore becoming well mixed, the operation comes closer to conventional fluidization in its attributes. Although the early statement that spouting "appears to achieve the same purpose for coarse particles as fluidization does for fine materials"[2] remains true for certain applications, such as solids drying, the differences in the two operations are important for others. For example, the systematic cyclic movement of particles in a spouted bed, in contrast to more random motion in fluidization, is a feature of critical value for granulation and particle coating processes. In the spectrum of contacting systems, therefore, spouted beds occupy a position that overlaps moving and fluidized beds to some extent, but at the same time they have a place of their own by virtue of certain unique characteristics.

### 1.5.2 Spouting versus fluidization

In what has become the well-entrenched C-A-B-D classification scheme for distinguishing between various types of fluidization by air at ambient conditions, Geldart[31]