FLUIDIZED BEDS
Combustion and Applications

Edited by

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FOREWORD

It can be justifiably claimed that fluidized bed combustion is the only really new approach to combustion of fossil fuel that has emerged for at least six decades. The potential merits of burning coal in a fluidized bed of mineral matter, in which boiler tube surface is immersed, were recognized by the late Douglas Elliott as long ago as 1962 when he was a member of staff of one of the Central Electricity Generating Board's research laboratories. Although he was not the first to have thought of fluidized bed combustion (and it is by no means certain who did) the initiation of the current phase of activity on FBC in the Western world can most certainly be attributed to him. The climate in the early 1960s was not, however, favourable for application of the technology to power generation; there was a growing momentum towards nuclear power, there were ample supplies of low cost oil, legislation overseas on sulphur dioxide and NO_x emissions had not begun to bite, and (before its problems were adequately recognized) the more glamorous magnetohydrodynamic system of power generation absorbed the major part of the effort available in the UK for research and development on new routes based on coal. His experimental programme at the CEGB became a casualty of this situation but he continued to have an influence on work elsewhere.

I first met Douglas in 1952 when we both were involved in the UK Ministry of Fuel and Power's programme on combustion of coal for gas turbines. Subsequently our activities in the fields of development in coal utilization frequently brought us together on topics such as gasification of coal for combined cycle power plant, high intensity combustion for MHD power generation and in particular during the later involvement at Leatherhead, of the British Coal Utilisation Research Association in fluidised bed combustion.

The one major exception to this was in the work on fluidized bed combustion for industrial steam generation. This began at Leatherhead in 1964 but we were unaware at that time of Douglas's already significant contributions to the technology.

Our main association was a consequence of a conversation in the car park at the CEGB Marchwood Laboratories between Douglas, Alan Roberts and myself towards the end of 1967. We there and then became convinced that, for power generation applications FBC would offer the maximum benefit if carried out under pressure in a combined gas and steam cycle. Douglas Elliott inspired the ensuing pilot-plant programme at Leatherhead and chaired the informal group comprising representatives of the CEGB, National Coal Board and BCURA that directed the work during the early stages and laid the foundation for the many subsequent research and development programmes in this field.

In the fertile environment of the University of Aston in Birmingham he worked on expanding the range of applications for FBC technology, laid the foundations for the commercial range of boilers and heat treatment equipment manufactured by Fluidfire, and provided education and training in FBC technology which will be of benefit to future generations.

Douglas Elliott was a remarkable individual who inspired all those privileged to claim his acquaintance. His advanced thinking and his enthusiasm to achieve an objective provided a spur for all those associated with him. I am not aware of any fluidized bed combustion system in vogue today that was not at one time or another commented upon by him.

His untimely death in 1976 prevented him from seeing the substantial application and commercialization of the technology. Although he would have been gratified at the progress so far made, he would probably be exasperated at the time that it has taken before some of the potentially more beneficial applications have been followed through; some indeed have yet to reach the demonstration stage.

It is to be hoped that this book, prepared as a tribute by some of his friends, will inspire more advances towards the achievement of his objectives.

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PREFACE

The untimely death of Professor Douglas Elliott on 16 June, 1976, was grievous to his wife and children, and his countless friends and colleagues; but it was grievous too for the engineering profession, just at a time when the challenges facing it were, and still are, formidable. Engineers like Douglas are a rare breed; he was one of the most perceptive and creative of his generation.

Production of this book was first thought about shortly after his death, when Douglas Probert, who succeeded him as Editor of the journal *Applied Energy*, suggested that five to six years hence might be a good time to launch a book to commemorate Douglas Elliott, because the technology to which he had contributed so much was on the threshold of industrial scale demonstration. Such an undertaking clearly required the help of many experts. This volume is the result of their labours. All of the authors knew him as a personal friend, colleague and collaborator. I am deeply grateful to all of them for giving their time and professional skill so generously at a time when, such is the interest in fluidized systems, their services are in great demand. Their eagerness is testament to the high esteem in which Douglas Elliott was held by his peers. Any views expressed in their chapters are those of the authors and not necessarily those of their employing organizations. Nonetheless, I am grateful to all these organizations for their cooperation in helping to bring this work to fruition.

The story of Douglas's own involvement in fluidized beds begins in the late 1950s and early 1960s when he was working at the Marchwood Engineering Laboratories of the Central Electricity Generating Board. He began some experiments which led him to re-discover fluidized bed combustion for himself and he saw its potential immediately. The objective

of such work was to develop more economic systems for utilization of coal for electricity generation and for industrial boilers. All factors relating to the installation were meant to be covered by the adjective 'economic', including amortization of capital, maintenance and, very important, the environmental regulations.

Douglas's hopes for increased utilization of coal found renewed expression in 1969 in his inaugural lecture after he was appointed Professor of Mechanical Engineering at the University of Aston in Birmingham. The lecture, which examined some of the technical arguments which were at the root of electricity generation strategy at that time, was entitled 'Can Coal Compete?—The Struggle for Power'. Events and changes in economic and political conditions over a span of 13 years often confounds the predictions and policy recommendations of experts. Douglas's basic thesis remains pertinent today, for increased utilization of coal has since become an urgent priority, backed by governments and energy users. Fluidized bed technologies are an important element toward achievement of this objective particularly for the poorer qualities of coal, and low grade fuels. These are in abundance world-wide, are cheap, but are difficult to burn by other means particularly when clean, non-polluting combustion is essential.

However, the technology, as some chapters in this book show, has not been confined to combustion of coal; liquid fuels and gases can be burned and the technology exploited for incineration, drying, metallurgical furnaces and heat recovery.

The literature on application of fluidized beds to combustion, gasification and chemical engineering processes is vast. It is hoped however that this volume will be valuable to engineers, students and indeed anyone wanting to obtain a rapid grasp of the fundamental principles of fluidized beds, how they are applied to energy using processes and to conservation, while at the same time meeting the required environmental standards.

J. R. HOWARD

The selection and presentation of material and the opinions expressed in
this publication are the sole responsibility of the authors concerned.

CONTENTS

Chapter 1

FLUIDIZED BED BEHAVIOUR

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NOMENCLATURE 木语表

A_b	Cross-sectional area of bed (m^2)
C_g	Heat capacity of gas (J/kg K)
d_B	Diameter of bubble (m)
d_f	Arithmetic average of adjacent sieve apertures (m)
d_p	Particle diameter (m) 粒径 径
D_b	Diameter of bed (m)
F 粒程	Fraction of material in bed $< 45 \, \mu m$
g	Acceleration due to gravity (m/s^2)
h_{gc}	Interphase gas convective component of bed-to-surface heat transfer coefficient (W/m^2 K)
h_{gp}	Gas-to-particle heat transfer coefficient (W/m^2 K)
h_{max}	Maximum bed-to-surface heat transfer coefficient (W/m^2 K)
h_{mf}	Bed-to-surface heat transfer coefficient at minimum fluidization (W/m^2 K)
$h_{overall}$	Overall bed-to-surface heat transfer coefficient (W/m^2 K)
h_{pc}	Particle convective component of bed-to-surface heat transfer coefficient (W/m^2 K)
h_{rad}	Radiative component of bed-to-surface heat transfer coefficient (W/m^2 K)
k	Constant
k_g	Gas thermal conductivity (W/m K)
m	Mass of particles in bed (kg)

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ΔP_b Pressure drop across bed (N/m^2)
ΔP_D Pressure drop across distributor (N/m^2)
T_b Temperature of bed (K)
T_s Temperature of heat transfer surface (K)
U Superficial gas velocity (m/s)
U_B Rise velocity of single bubble (m/s)
U_m Gas velocity for maximum bed-to-surface heat transfer (m/s)
U_max Maximum operating velocity to avoid slugging (m/s)
U_mB Gas velocity at which bubbling first occurs (m/s)
U_mf Gas velocity at which bed becomes fluidized (m/s)
U_t Particle terminal velocity (m/s)
X Mass fraction
ε_b Emissivity of bed
ε_m Modified emissivity allowing for effect of immersed surface on adjacent bed temperature
ε_mf Voidage at minimum fluidization
ε_r Reduced emissivity
ε_s Emissivity of surface
ϕ Particle shape factor in Ergun equation (ref. 20)
ρ_f Density of fluid (kg/m^3)
ρ_g Density of gas (kg/m^3)
ρ_p Density of particle (kg/m^3)
μ_g Viscosity of gas (kg/m s)
Ar   Archimedes number, $d_\text{p}^3 \rho_\text{g}(\rho_\text{p} - \rho_\text{g})g/\mu_\text{g}^2$
Nu   Nusselt number, hd_p/k_g
Nu_gc Nusselt number based on h_gc
Nu_gp Nusselt number based on h_gp
Nu_max Nusselt number based on h_max
Pr   Prandtl number, $C_\text{g}\mu_\text{g}/k_\text{g}$
Re   Reynolds number, $d_\text{p}U\rho_\text{g}/\mu_\text{g}$
Re_mf Reynolds number based on U_mf
Re_opt Reynolds number based on U_m

1   INTRODUCTION

1.1 The phenomenon of fluidization

The advantages of the technique of fluidization have often been listed (for example, see Botterill[1]). Principally they stem from the very large particle surface area exposed to the fluid, the ease with which the solids can be

handled in the fluidized state and, in the heat transfer applications which are the concern of this book, the good heat transfer properties of a bubbling gas-fluidized bed. Inevitably there are also disadvantages. Operating rates are limited to within the range over which the bed can be fluidized. The cost of the pumping power to fluidize the bed may be excessive and particularly so with deep beds. There is a limit to the size and type of particle that can be handled by this technique. As will be stressed in this chapter, it is difficult to
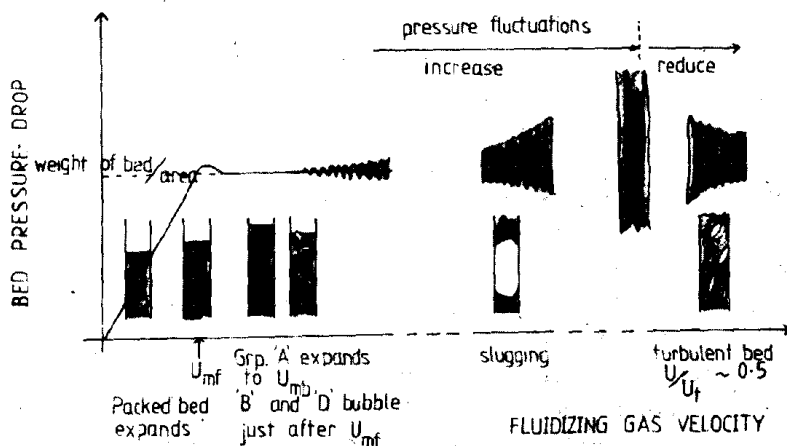

FIG. 1. Schematic diagram of variation in bed pressure drop with gas flow rate.

characterize the particles themselves and there can be a very wide range of behaviour according to the conditions under which a fluidized bed is being operated. For a fuller consideration of bed behaviour and design see Geldart,[2] 'Gas Fluidization'.

A bed of particles offers resistance to fluid flow through it. As the velocity of flow increases, the drag force exerted on the particles increases. With upward flow through an unrestrained bed, the particles rearrange themselves to offer less resistance to the fluid flow and the bed will tend to expand unless it is composed of large particles (mean diameter >1 mm). With further increase in the upward fluid velocity, the expansion continues until a stage is reached where the drag force exerted on the particles will be sufficient to support the weight of the particles in the bed (Fig. 1). The fluid/particle system then begins to exhibit fluid-like properties and it will flow under the influence of a hydrostatic head. This is the point of *incipient fluidization* and the gas velocity needed to achieve this is referred to as the *minimum fluidizing velocity*, U_{mf}. Beyond this velocity, the pressure drop

across the bed will be approximately equal to the weight of the bed per unit area. Thus:

$$\Delta P_b A_b = \frac{m}{\rho_p}(\rho_p - \rho_f)g \qquad (1)$$

where ρ_p is the effective density of the particle which may be porous, and ρ_f is that of the fluid. The effective ΔP excludes the hydrostatic pressure drop across the bed which can be neglected in gas-fluidized systems operating at atmospheric pressure. It is likely, however, that this pressure drop will be exceeded just prior to fluidization with gas-fluidized systems in order to overcome cohesive forces between the particles and break down the residual packing and interlocking of particles within the bed. It is also likely to be exceeded when operating under more extreme conditions at high gas flow rates when rising slugs of solids (see below) can be formed and energy is required in order to accelerate them. When the bed is composed of a powder with a wide size distribution, the pressure drop may never quite equal that of the weight of the bed per unit area thus indicating that not all of the bed is freely supported by the gas flow.

At the onset of fluidization the bed is more or less uniformly expanded and, up to this point, it makes little difference to the general bed behaviour whether the fluid is a gas or a liquid. However, whereas the liquid fluidized bed tends to continue to expand stably with further increase in upwards flow velocity until the particles are carried from the bed, the uniform expansion behaviour is soon lost as the gas velocity is increased except with fine powders. Thus, with the gas-fluidized system, instabilities develop and cavities containing few solids are formed. These look like bubbles of vapour in a boiling liquid and, as they rise through the bed, they are responsible for generating the solids mixing which is such an important feature of gas-fluidized bed behaviour.[3] Over the flow range between incipient fluidization and the onset of bubbling, the bed is said to be in a quiescent state.

At high velocities conditions are such that a lot of material is lost from the bed by elutriation even before the regime of pneumatic conveying is reached. Bubbles bursting at the surface generate pressure fluctuations and throw a spray of particles into the freeboard space above the bed. Many are carried away in the gas stream according to the particle size distribution and the gas velocity in the freeboard region above the bed. There is increasing interest in the possibility of operating beds at higher gas velocities in order to increase the throughput and to obtain more advantageous reaction conditions according to circumstances; the regimes of the *turbulent* and *fast* fluidized beds[4] (Fig. 1). Changes in bed behaviour

on entering this regime are outlined in Section 1.4, below. Under these more extreme operating conditions it is obviously necessary to make provision for the constant return or replacement of the material carried from the bed.

1.2 Range of behaviour of gas-fluidized beds

Much of the range of behaviour encountered with gas-fluidized systems is a consequence of the different bubbling behaviour that can occur.[3] The rise rate of a single bubble of diameter d_B is given by:

$$U_B = k\sqrt{(gd_B/2)} \tag{2}$$

where the constant, k, depends on particle shape and size distribution, although its value is generally about 0·9. This velocity is usually greater than the interstitial gas velocity except for beds of Group 'D' materials (see below this section). A bubble within a chain or cloud of bubbles rises faster than one in isolation, except for those in beds of large mean particle diameter (~ 1 mm) when the rise rate tends to be slower. As bubbles rise through the bed they grow by collection of gas from the surrounding continuous phase and, most importantly, by coalescence. In this latter process, a smaller bubble comes within the influence of a faster rising one which overtakes it and draws it into its wake except, again, for the case of beds of large particles when cross-wise coalescence between bubbles predominates. On occasion, bubbles are also observed to shrink and this tendency is apparently stronger for bubbles below a critical size. Bubbles may divide spontaneously or on contact with an obstruction. General bubbling behaviour is affected by the design of the distributor by which the gas is introduced to the bed and by surfaces immersed within the bed. Strong, bubble-induced solids convection streams can become established which will tend to sustain themselves until something occurs to disturb the flow pattern and the system will switch to a new form.[5] As a bubble rises it displaces particles within its path and draws a streak after it giving rise to a drift profile. Also, apart from the case of bubbles in beds of large mean particle diameter, particles are carried within the bubble's wake and material is constantly collected by and shed from the wake as the bubble rises through the bed. To replace the material carried upwards through the bed by these processes there is the downwards return flow of solids through other regions of the bed.

With deep beds of high aspect ratio and at higher operational velocities, bubbles may grow until they occupy the whole cross-sectional area of the bed. These then carry a slug of particles ahead of them until instability

occurs and the solids collapse back to the bed. This is the so-called *slugging* bed and, as mentioned above, energy has to be supplied to accelerate the slug of particles upwards so the overall pressure drop across the bed will then exceed the weight of the bed per unit area. Pressure fluctuations across the bed will increase[6] (Fig. 1). With a wide size distribution, there will be the tendency to lose a lot of the fines under the operating conditions required to keep the larger fractions within the fluidized state. It is also possible, of course, to fluidize larger particles within the dense fluid obtained by fluidizing the finer fractions. However, in either instance, on reduction of the gas throughput, there will be a tendency for segregation to occur with the deposition of the larger particles on to the distributor. This is generally to be avoided but it may be advantageous to encourage such deposition when it is desired to protect the distributor from the full rigour of exposure to a higher temperature bed.

It has often been suggested that some breadth of size distribution is desirable to promote stable fluidization. Geldart[7] has argued that it is the mean size on a surface to volume basis which is the important factor and, as he has pointed out, that the addition of a comparatively small amount of fines by weight can have a considerable influence on the average particle diameter expressed in this way and on bed behaviour. Later work shows that the actual proportion of fines (classed as material $<45\,\mu$m) is of particular importance.[8] The significance of this effect can be better appreciated in relation to the classification[9] that Geldart suggested earlier which is primarily based on particle density and size from tests under ambient conditions (Fig. 2). Materials categorized within his *Group 'A'* generally have densities less than $1400\,\text{kg/m}^3$ and fall within the size range of 20 to 100 μm. These powders exhibit a pronounced degree of stable bed expansion when the minimum fluidization velocity is first exceeded, and it may be possible to sustain such uniform or *particulate* fluidization until the minimum fluidizing velocity has been exceeded by a factor of two to three times. However, with further increase in gas velocity a point will be reached when the bed will collapse back to a less expanded state approximating more closely to the degree of expansion under the minimum fluidizing conditions and most of the excess gas will flow through the bed as the bubble phase, i.e. the so-called *aggregative* fluidization. The gas velocity at which this bed collapse occurs is referred to as the minimum bubbling velocity, U_{mB}. Whilst the bed is in the quiescent state before it begins to bubble, only very limited particle mixing by a diffusive mechanism can take place.

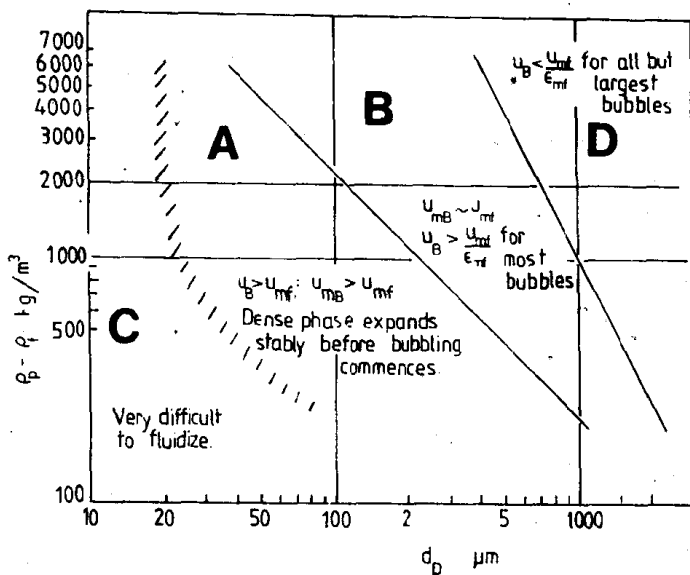Geldart's *Group 'B'* materials tend to have a mean size within the range

FIG. 2.   Powder classification diagram. (After Geldart.[9])

40 to 500 μm and a density in the range 1400 to 4000 kg/m^3. These exhibit much less stable bed expansion; free bubbling (aggregative fluidization) occurring at or a little above the minimum fluidization velocity. Abrahamsen and Geldart[8] gave the following relationship for prediction of the minimum bubbling velocity, U_{mB}:

$$U_{mB} = 2 \cdot 07 e^{0 \cdot 716 F} \left(\frac{d_p \rho_g^{0 \cdot 06}}{\mu_g^{0 \cdot 347}} \right) \tag{3}$$

where F is the fraction of material less than 45 μm and the dimensional constant is for SI units. This was established from an extensive series of tests under ambient conditions. Earlier, Baeyens and Geldart[10] suggested that beds of particles would fall within Group 'B' rather than 'A' when their minimum fluidizing velocity exceeded the minimum bubbling velocity. From tests with air at ambient pressure and temperature the condition was approximately

$$(\rho_p - \rho_g)^{1 \cdot 17} d_p > 906\,000 \tag{4}$$

Whereas bubbles tended to grow until their diameter was finally limited by the dimensions of the bed with materials of Group 'B', there seemed to be a definite maximum stable bubble size with beds of Group 'A' material and this size was influenced by the fines fraction, F. Thus, an equilibrium could

apparently be reached between bubble growth by coalescence and spontaneous splitting to give a stable bubble size distribution within the upper levels of deep beds of Group 'A' materials.

Group 'C' materials are those of smaller mean size ($< 30 \,\mu m$) and/or of lower density so that interparticle forces have an effect greater than that of gravity. Such materials are very difficult to fluidize. It is unlikely here that the pressure drop across the bed will equal its weight per unit area. This shows that part of the weight, even if the bed is displaying quasi fluid-like properties, is supported by the interparticle forces and surface contacts. With these materials, *channelling* is very prone to occur, i.e. the flow of fluid opens up low resistance channels through the bed. Once such a path has been created, it tends to enlarge with further increase in gas velocity so that the gas is not properly distributed into the bed which never becomes truly fluidized. This can be overcome to some limited extent by using a mechanical stirring element moving close to the distributor which will break down the channels and so cause the gas flow to redistribute itself through the bed in the lower regions.

Group 'D' materials are usually of mean diameter greater than $600 \,\mu m$ and/or dense particles. Although a bubbling fluidized bed looks very turbulent and may be described as being turbulently fluidized when operated at higher fluidizing velocities, the gas flow condition within the interstices tends to be laminar or, at the most, transitional until one is dealing with materials of this Group D. With these materials, the interstitial flow rate of the gas through the continuous phase necessary to fluidize the bed is greater than the rise velocity of the bubbles. This gives rise to a through flow of gas without the gas circulation cloud round the bubble, as obtained with the other materials for which the bubble rise velocity is higher than the interstitial gas velocity.[3] Gas flow conditions tend now to be in the turbulent regime and the bubble coalescence pattern is different. Instead of a smaller bubble being caught up in the wake of a faster rising, larger bubble, cross-wise coalescence takes place between bubbles at a similar level in the bed. The ensuing solids mixing generated by this process is less effective than that by vertical coalescence.

Because of the difference in bubbling behaviour between different classes of material there can be very different changes when scaling up from small experimental beds to full-scale operational beds. This particularly comes about because bubbles are constrained in size by the scale of the equipment in beds of small diameter but have the opportunity to grow very much larger in larger scale equipment. With finer, less dense powders (Group A) the maximum stable bubble size is much smaller than with

coarser, denser particles (Group B) so a stable bubble size distribution can be reached with smaller diameter beds with A materials than with B materials. Because larger bubbles rise faster than smaller ones, the gas associated with the bubbling process will escape from the bed more rapidly when the average bubble size is larger, so there can be much variation in the overall bed expansion brought about by change in the total bubble volume hold-up within the bed according to the bubble size distribution. This is
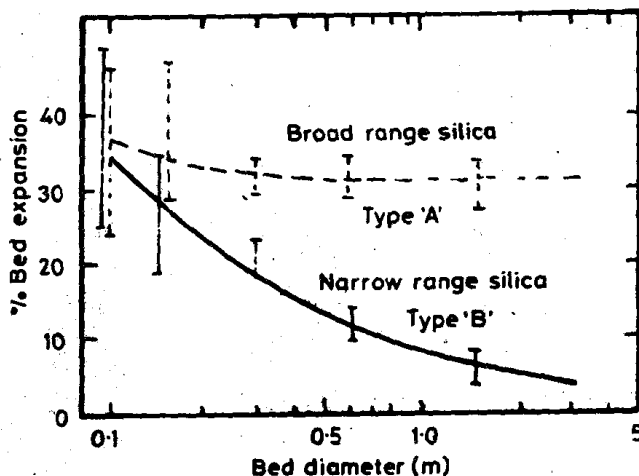

FIG. 3. Bed expansion as a function of bed diameter, only difference being size distribution of catalyst; expansion reflects bubble hold-up.[11]

well illustrated by some tests reported by de Groot[11] (Fig. 3). Here the only difference between the two series of tests on beds of a catalyst is the quantity of fines in the material and hence the mean particle diameter, a difference which is sufficient to alter the fluidization characteristics very markedly as Geldart noted;[7] the broad-range falling within the 'A' group and the narrow-range in the 'B' group of his classification. Associated with this change in behaviour will be changes in solids mixing within the bed and also in the efficiency of gas/solids contacting with important implications for reactor conversion efficiency.[2] The situation is further complicated by changes in behaviour when other operating variables, such as temperature and pressure, are changed (see Section 2 below). Much caution should therefore be used when applying the results of small-scale ambient temperature tests. Again, the immersion of heat transfer surfaces within the bed will affect bubbling behaviour but the disturbances caused by this are less likely to be scale dependent because the inserts will have the effect of breaking up the bed into smaller, repeated sections.