FLUID-BED HEAT TRANSFER

Gas-fluidized bed behaviour and its influence on bed thermal properties

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Preface

This book sets out to review some of the background knowledge about gas-fluidized bed behaviour and to draw out its implications for fluidized bed heat transfer. The wide variety of behaviour encountered between different size range and shape of particles is particularly stressed but no pretence is made to being able to prepare the reader for all the surprises to be encountered in the application of the technique of fluidization.

With regard to the basic behaviour of gas fluidized systems, three important mechanisms are now clear. Firstly, there is the way in which rising bubbles induce particle circulation. This was shown in a series of attractive experiments. Secondly, there is the important consequence of particle mixing for fluid-bed heat transfer. Thirdly, there is the understanding of gas flow through the bubble phase which was predicted from a consideration of potential flow in a porous medium and, although this is not of importance in the application of the heat transfer properties of the system, it has very important consequences in gas/solid contacting operations and hence in chemical reactor design.

Some fundamental studies have made considerable contribution to the understanding of the behaviour of gas-fluidized systems. Nevertheless, the basic behaviour is so complicated that there are no fundamental models which can, for example, describe the characteristic development of bubbles and be used for predictive purposes. Inevitably, the field has attracted a few workers with mathematical ability but only passing interest in fluidization who have applied their expertise in the formulation and solution of models whose content had little physical reality. There is still much scope for experimental investigations to explore what really happens and, because of the complexity of gas-fluidized systems, it would seem that a profitable way to proceed in the interpretation of experiments is to formulate the simplest possible hypotheses and to test them against the experimental data. It is then to be expected that fuller understanding will be forthcoming as the assumptions made are questioned and those which are inadequate are refined. For practical applications of the technique of fluidization, what is required is an understanding of the mechanism by which advantageous properties are obtained so that the designer can co-operate with nature in order to achieve his desired result. I would claim that the objectives are understood with regard to the achievement of good bed-to-surface heat transfer; but there remains the problem of finding the ingenuity to engineer the system so that they can be attained.

The worker new to the field of fluidization should take early warning that gas-fluidized beds behave in very varied ways according to the, experimental circumstances. A bed of 80 μm mean diameter particles, for example, will continue to expand stably as the gas flow rate through it is increased beyond the point of fluidization and will not break down into the characteristic bubbling condition encountered with larger bed material until the minimum fluidizing velocity has been exceeded by a factor of 2 or so. The behaviour of small-scale fluidized beds is very different to that of big beds because the solids circulation and bubble growth patterns will be very different between the systems. Nevertheless, results from small-scale tests have often been presented in the form of correlations of general applicability although no attempt had been made in the work to take into account the important effect of change in equipment scale. The outcome of an experimental programme may often only have meaning within the context of the restricted experimental conditions. The worker who wants to apply a correlation should check back to the original paper to see over what range of variables it may be applicable.

A quotation from *The Secret of the Sea* by Longfellow is apposite:—

> 'Wouldst thou'—so the helmsman answered,
> 'Learn the secret of the sea?
> Only those who brave its dangers
> Comprehend its mystery!'

There is no substitute for first-hand experience when each material can display peculiar properties under some conditions. It is always salutary to do the experiment and look at the way a material behaves when fluidized. When much is at stake, tests should be carried out on as large a scale relevant to the proposed design as practicable. There is much advantage in "unit-cell" tests when the full-scale unit incorporates a series of repeated units. That is to test at full-scale, equipment which reproduces all the features of one unit of the proposed design. This would be the area of bed fed from one fuel distribution point, for example, in the design of a combustor. Much can be gleaned from comparatively inexpensive tests on fluidization behaviour at atmospheric pressure and moderate temperature.

Chapters 2 and 3 require a note of explanation. Originally they were intended to tell the reader what he might want to know about fluidization. However, they ended up as a broad descriptive review of gas-fluidized bed behaviour. When the reader looks, for example, at the book "Fluidization" edited by Drs. Davidson and Harrison, which itself makes no claim to give comprehensive coverage, the reason for this will be appreciated. Nevertheless, it is hoped that in providing an "index" to more than 400 papers they will be helpful both to those who are beginning to study fluidization and to those who are familiar with many aspects of the phenomena.

The dedication of this book records my gratitude to all those on whose work I have drawn. It has given me much pleasure to have had opportunity of meeting so many of them personally. I would particularly thank Dr. Jack Broughton for much constructive discussion of the manuscript. It is to be expected that we should still hold different views of many aspects of fluidization. Errors in fact and emphasis belong to my view. They may well be changed in detail before this book is published as some fresh surprises become known. Finally, I would acknowledge my indebtedness to Professor Douglas Elliott of the Department of Mechanical Engineering, the University of Aston in Birmingham. We have worked closely together for a number of years and I have especially valued the imagination, ingenuity and enthusiasm that he brings to his work. Originally it was our intention that this book should have been of wider scope under our joint authorship and I still hope that Professor Elliott will contribute that other part on "Industrial Applications" as originally planned.

J. S. M. Botterill

Nomenclature

The S.I. system of units (kg, m, s) has been used with temperatures generally expressed in °C. Symbols are usually defined in the text as they arise and appropriate units given to avoid uncertainty (e.g. particle diameter, d_p, is usually expressed in μm as being of more convenient scale; certain radiative heat transfer expressions use temperatures in degrees absolute).

A	constant of equations 2.11 and 2.12
A	area, m^2
A_b	cross-sectional area of bed, m^2
A_m	area of packet in contact with transfer surface, m^2
A_p	surface area of particles per unit length of bed, m^2/m
A_t	cross-sectional area of tube, m^2
a	solid surface area per unit volume of bed, m^2/m^3
B	dimensionless parameter depending on the geometry of the bed and thickness of particle film at surface equation 4.60
C	constant of equation 4.18 relating heat transfer coefficient to bed fluidization conditions or film to particle heat transfer relationship of equation 4.61
C_B	drag coefficient round a bubble of diameter, d_B
C_f	heat capacity of film, J/kg °C
C_g	heat capacity of gas, J/kg °C
$C_{H_{1\&2}}$	cumulative heat contents of particles of types 1 and 2 respectively, J/°C
C_p	heat capacity of particle, J/kg °C
C_{p_1} and C_{p_2}	volumetric specific heat of particles of types 1 and 2 respectively, J/m^3 °C
C_R	constant in equation 5.14 to allow for non-axial position of heat transfer surface
D	dispersion coefficient m^2/s or pipe diameter in equations 3.2 and 3.3, m
D_e	equivalent hydraulic diameter; 4 × flow area/wetted perimeter, m

D_o	orifice diameter, m
d_B	bubble diameter, m
d_{Bh}	bubble diameter at height h, m
d_b	bed diameter, m
d_f	particle diameter of sieved fraction, m (μm)
d_p	particle diameter, m (μm)
d_{pk}	packing diameter, m
d_t	diameter of immersed object equation 2.11, m
d_{ti}	internal diameter of tube, m
d_{to}	external diameter of tube, m
E	fluidization efficiency defined by equation 4.16
$\bar{\mathscr{F}}$	mean value of the view factor
F_o	weight of material entrained through unit area in unit time at a given height as given by equation 3.6, kg/m^2 s
f_b	fraction of particles at temperature different to the rest of the bulk of the bed
f_d	drift fraction for particle transport
f_o	fraction of time a surface is shrouded by bubbles
f_{sd}	volume fraction occupied by downward moving solids
f_{su}	volume fraction occupied by upward moving solids
f_w	wake fraction of bubble
G	mass flow rate of gas based on the empty tube, kg/m^2 s
G_{mf}	gas mass flow rate at minimum fluidization, kg/m^2 h, equations 2.23 and 2.24.
g	acceleration due to gravity, m/s^2
H	height of bed, m
H_t	settled bed height, m
H_{mf}	bed height at minimum fluidization, m
H_o	equilibrium bed height stick-slip flow condition of equation 3.1, m
h	height above distributor equations 2.9 and 2.10 or height above bed in the Froude group of equation 3.6, m
h_o	notional depth below distributor from which bubbles would originate equation 2.10, m
h	heat transfer coefficient, W/m^2 °C
h_{av}	heat transfer coefficient averaged over length of heater equation 4.63, W/m^2 °C
h_{gc}	interphase gas connective component of bed to surface heat transfer coefficient, W/m^2 °C
h_{gp}	gas/particle heat transfer coefficient, W/m^2 °C
h_i	instantaneous heat transfer coefficient, W/m^2 °C
h_L	local heat transfer coefficient, W/m^2 °C
h_m	maximum bed to surface heat transfer coefficient, W/m^2 °C

h_{mf}	heat transfer coefficient under minimum fluidization conditions, W/m^2 °C
h_{mpc}	maximum particle convective component of bed to surface heat transfer coefficient, W/m^2 °C
h_{pc}	particle convective component of bed to surface heat transfer coefficient, W/m^2 °C
$(h_p)_{BA}$	particle B to particle A heat transfer coefficient, W/m^2 °C
h_r	radiative component of heat transfer coefficient, W/m^2 °C
$\bar{h}_c$	gas convective heat transfer coefficient in the heat balance of equation 4.73, W/m^2 °C
$\bar{h}_o$	overall heat transfer coefficient in the heat balance of equation 4.73, W/m^2 °C
$\bar{h}_R$	radiative heat transfer coefficient in the heat balance of equation 4.73, W/m^2 °C
K	proportionality constant relating number of particles moved to excess gas flow rate in a packed fluidized bed equation 2.18; proportionality constant in relationship between particle velocity and film thickness equation 4.17
k	constant in bubble rise velocity equation 2.7 and in consistency index of equation 3.5, N/m^2 s^n
k_1 and k_2	constants in equation 2.9 for the prediction of bubble diameter at a given height above the distributor
k_{bmf}	effective bed conductivity at the minimum fluidization condition, W/m °C
k_e	effective conductivity factor equation 4.47 or bed thermal conductivity in equation 4.63, W/m °C
k_{ed}	effective thermal conductivity through the solid phase in the parameter of equation 4.8, W/m °C
k_{et}	effective thermal conductivity of emulsion layer, W/m °C
k_{ep}	effective thermal conductivity inside an aggregate of particles, W/m °C
k_{ew}	effective thermal conductivity in the zone adjacent to a surface, W/m °C
k_f	fluid thermal conductivity, W/m °C
k_g	gas thermal conductivity, W/m °C
k_p	particle thermal conductivity, W/m °C
k_s	solids interchange coefficient, equation 2.14, m^3/m^3 s
L	exposed length of heater or length of pipe, m
l_e	effective thickness of emulsion layer, m
M_p	mass of particles in element equation 4.3, kg
m	mass of particles in bed, kg
N	number of particles/unit area, 1/m^2
N	number of orifices through which gas is flowing equation 2.9

N'	bubble number concentration equation 2.10, $1/m^3$
n	bubble frequency, 1/s
n	index in equation 2.30
n	flow behaviour index in equation 3.5
P	jet penetration depth, m
ΔP	pressure difference between bottom and top of bed or along length of pipe, L, in equation 3.2, N/m^2
ΔP_D	pressure drop across distributor, N/m^2
P_b	pressure at bottom of bed, N/m^2
P_t	pressure at top of bed, N/m^2
p	index in equation 2.9
Q	flow rate of injected gas, m^3/s, or total rate of heat flow, W/m^2, chapter 4
Q_B	bubble volume flow rate, m^3/s
q	heat flow to packet of particles, W/m^2, or index equation 2.9
q_A	heat flux to particle A surrounded by B particles W/m^2
q_f	heat flux via fluid phase, W/m^2
R	Parameter representing bed condition for bubbles of given size equation 2.6 or expansion ratio beyond that of bed at U_{mf}, chapter 4
R_a	instantaneous thermal resistance of aggregate of solid particles, m^2 °C/W
R_r	effective resistance of aggregate of solid particles, m^2 °C/W
R_w	thermal resistance adjacent to the wall, m^2 °C/W
r	radius, m
r_b	radius of bed at the bottom, m
r_h	radius of heater, m
r_p	radius of particle, m (μm)
r_t	radius of bed at top, m
S_L	stirring factor defined by equation 4.24
T	temperature, °C (or °K for radiant heat transfer)
T_b	bed temperature, °C
T_e	emulsion packet temperature, °C
T_g	gas temperature, °C
ΔT_g	gas temperature difference, °C
T_{gi}	inlet gas temperature, °C
T_{go}	outlet gas temperature, °C
T_p	particle temperature, °C
T_s	surface temperature, °C or source temperature with radiant heat transfer (°K)
t	time (for particles to travel from top to bottom of bed equation 4.15), s

U	velocity, m/s
U_{av}	average fluid (or flowing bed velocity equation 3.3) m/s
U_B	bubble rise velocity, m/s
U_f	superficial fluid velocity, m/s
U_g	superficial gas velocity, m/s
U_t	coefficient of equation 2.30
U_m	gas flow rate for maximum bed to surface heat transfer, m/s
U_{MB}	gas velocity at which bed begins to bubble equations 2.1 and 2.2 (mm/s)
U_{mf}	gas velocity for minimum fluidization, m/s (mm/s)
U_{max}	maximum flowing bed velocity, m/s
U_p	particle velocity or characteristic velocity in elutriation test equation 2.27, m/s
$\bar{U}_p$	average downward velocity of emulsion phase, m/s
U_s	linear velocity of solids, m/s
V_B	volume of bubble, m^3
W	dimensionless flow parameter defined by equation 2.5
x	weight fraction of sieved particles
x_g	gas slab thickness, m
Z_s	apparent weight of particles, N
α	thermal diffusivity, m^2/s
α_e	emulsion phase thermal diffusivity, m^2/s
α_f	fluid phase thermal diffusivity, m^2/s
α_g	gas thermal diffusivity, m^2/s
α_p	particle thermal diffusivity, m^2/s
β	conductivity ratio, k_p/k_f equation 4.40, or interparticle friction factor equation 4.15
γ	shear rate equations 3.3 and 3.5, 1/s
γ	dimensionless rate of heat transfer equation 4.45
γ_f	fluid kinematic viscosity, m^2/s
δ	volume fraction of bubbles in fluidized bed
ε	bed voidage
ε_b	effective bed emissivity
ε_{mf}	bed voidage at minimum fluidization
ε_p	particle emissivity
ε_w	voidage at the wall
λ	time scale factor equation 4.38
μ_f	fluid viscosity, Ns/m^2
μ_g	gas viscosity, Ns/m^2
σ	Stefan Boltzman constant, $W/m^2\ {}^\circ K^4$
ρ_b	bed density, kg/m^3
ρ_{bmf}	bed density at minimum fluidization condition, kg/m^3
ρ_f	film density, kg/m^3

ρ_g	gas density, kg/m^3
ρ_{go}	gas density at atmospheric pressure, kg/m^3
ρ_p	particle density, kg/m^3
Ω	Beránek criterion equation 2.27
Θ	dimensionless residence time or angle
τ	packet age or contact time with heat transfer surface, s
$\bar{\tau}$	mean packet age or contact time, s
τ	shear stress chapter 3, N/m^2
τ_w	shear stress at the wall chapter 3, N/m^2
A_r	Archimedes number $gd_p{}^3\rho_g(\rho_p - \rho_g)/\mu_g{}^2$
F_r	Froude group $U_f{}^2/d_pg$ or $U_g{}^2/hg$ in equation 3.6
Nu_{gc}	interphase gas convective Nusselt number $h_{gc}d_p/k_g$
Nu_m	maximum bed to surface Nusselt number h_md_p/k_g
Nu_p	particle/gas Nusselt number $h_{pg}d_p/k_g$
P_r	gas Prandtl number $C_g\mu_g/k_g$
Re_p	particle Reynolds number $d_p\rho_gU_g/\mu_g$
Re_{mf}	Reynolds number for minimum fluidization condition $d_p\rho_gU_{mf}/\mu_g$
Re'_{mf}	modified Reynolds number under minimum fluidization condition $d_p\rho_fU_{mf}/\mu_f(1 - \varepsilon_{mf})$

Contents

CHAPTER 1

The Fluidized State

I. INTRODUCTION

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 which the very large throughputs involved. Subsequently, the chemical and oil industries have concentrated largely on applications of gas-fluidized systems which exploit the advantage 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 flow rate under the action of the centrifugal field.

The objective of this book is to describe something of the very wide range of behaviour encountered in gas-fluidized systems and of its implications for fluidized bed heat transfer. Whilst in principal the technique of fluidization is simple, in practice there have been many disastrous failures because the change in behaviour between different materials and different scale of units was insufficiently appreciated. This book has been written as the first part of a text which is intended should lead to a systematic treatment of those operations where the advantageous heat transfer properties of gas-fluidized systems find application.

II. THE PHENOMENON OF FLUIDIZATION

A bed of loose particles offers resistance to fluid flow through it. As the velocity of flow increases, the drag force exerted on the particles increases. If the fluid is flowing downward through a bed of particles it will tend to compact it. However, if the fluid flow is upwards through the bed, the drag force will tend to cause the particles to rearrange themselves within the bed to offer less resistance to the fluid flow. Unless the bed is composed of large particles (mean diameter $\sim$ 1 mm) the bed will expand (Fig. 1.1). With further increase in the upward fluid velocity, the expansion continues and a stage will be reached where the drag forces exerted on the particles will be sufficient to support the weight of the

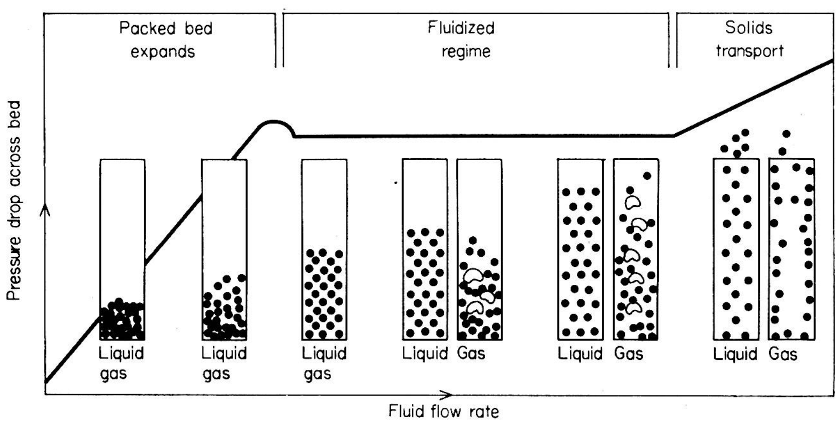

FIG. 1.1. Stylized representation of response of bed to upwards flow of fluid through it.

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 hump in the stylized curve for bed pressure drop as a function of the fluid flow rate, Fig. 1.1). The peak is not present when fluidizing beds of large mean particle diameter with a gas and bed expansion only begins to occur after the point of incipient fluidization then. Whilst vibration 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 few solids 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 type of gas distributor used. Over the flow range between incipient fluidization and the onset of bubbling the bed is in a *quiescent* state. For large, closely sized spherical particles, the ratio between the bubble-point velocity and that for incipient fluidization will be small. For smaller, irregular particles it will be larger, e.g. 2·8 has been reported for 55 μm catalyst particles (Godard and Richardson, 1968). 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 chapters.

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