Refereed Proceedings

The 12th International Conference on Fluidization - New Horizons in Fluidization Engineering

Void Fraction Near Surfaces Immersed in Fluidized Beds by Heat Transfer Measurements

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ABSTRACT

A semi-empirical model is used to calculate the averaged surface void fraction in fluidized beds, ε_w, starting from experimental data on surface-to-bed heat transfer coefficient. The model is able to describe the effect of the main process parameters and shows that ε_w increases with minimum fluidization void fraction and particle Archimedes number.

INTRODUCTION

Assessment of fluid dynamic field around immersed surfaces is of primary concern for the correct design of industrial scale fluidized beds, as for the case of heat exchangers and membrane reactors ([1]), or for the optimization of surface treatments of immersed objects ([2]-[3]). Its crucial importance is mainly related to its influence on surface-to-bed transport phenomena whose high values are among the most relevant characteristics of fluidization technology ([1]).

The bed structure near surfaces immersed in the emulsion phase of a fluidized bed mainly consists of a region characterized by gas velocity and void fraction higher than the bulk of the bed ([4]-[10]), which induces, in some cases, the formation of superficial bubbles (e.g. [7], [8]) and whose characteristics strongly depend on the surface shape ([5]). These features mainly influence the particle motion around the surface, whose properties can be described by means of a particle residence time distribution (e.g. [9]) and a surface bed voidage (e.g. [5]-[10]). Particle motion can be analyzed using direct methods, as optical investigations ([5]) and PEPT techniques ([9]), or indirect ones, such as numerical simulations ([7], [8]), abrasion ([3]) or heat and mass transfer measurements ([10]-[30]). However, while the particle residence time distribution near immersed surface has been analysed in details (e.g. [9], [12], [23]-[29]), the surface void fraction has been estimated in a very limited number of experimental and numerical studies ([5]-[10]). In some cases, the surface void fraction is derived from the application of heat transfer models to experimental results ([6], [9], [10]).

Surface-to-bed heat transfer has been extensively studied in the past (e.g. [10]-[30]) due to its relevance in industrial applications ([1]). This phenomenon is strongly related to the bed struc-ture near the exchange surface because of the so called particle convective heat transfer coefficient (e.g. [6], [11]), h_{pc}, which accounts for the

unsteady contacts between the surface and the fluidized particles. For these reasons heat transfer has been used in the past as an indirect measure of local bed structure (6, 9, 10, 12, 23, 30). Nevertheless, the lack in information on surface void fraction leads to the absence of a definite correlation between heat transfer and bed structure near the immersed surface.

This paper analyzes the possible application of heat transfer coefficient measurements for the determination of the surface void fraction. The correlation between the local structure of the bed (in terms of surface renewal frequency and void fraction) and the particle convective heat transfer coefficient is obtained using a semi-empirical single particle model. The model is validated with several experimental data (11-22) in a wide range of the main process parameters (gas velocity, pressure, temperature and bed material properties).

PARTICLE CONVECTIVE HEAT TRANSFER MODEL

It is generally accepted (e.g. 25) that the overall surface-to-bed heat transfer coefficient can be considered to be made up of three components which are approximately additive:

$$h = h_{gc} + h_{pc} + h_r ; \qquad\qquad (1)$$

where h_{gc}, h_{pc}, h_r are the gas convective, the particle convective and the radiative heat transfer coefficients. In this notation, h_{gc} accounts for both the contributions of the bubbles and gas which percolates in the emulsion phase.

The particle convective heat transfer coefficient can be described starting from the analysis of the thermal transient of the particles in contact with the exchange surface. In this sense, the proposed model has been developed under the following hypotheses:

- Validity of the approach of single particle heat transfer mechanism (25).
- The solid particles are considered as monosized spheres whose diameter is equal to the averaged Sauter diameter of the bed material.
- Negligible thermal gradient within the particle (i.e. uniform particle temperature). At ambient conditions this condition is verified for a ratio of particle and gas thermal conductivity higher than 30 (1).
- The main wall-to-particle heat transfer resistance is represented by a contact resistance due to the gas gap between the surface and the particle. For spherical particles the thick-ness of this gas layer is usually assumed to be equal to $d_p/10$ (e.g. 26).

Under the hypotheses of single particle approach, the surface-to-bed heat flow due to particle convection can be modelled as the sum of the heat flows between the surface and each partic-le, $q(t)$, weighted by means of the particle residence time distribution, $E(t)$ and multiplied by the number of particles simultaneously in contact with the surface, n. Hence, h_{pc} is calculated as:

$$h_{pc} = \frac{\dot{Q}}{A_w(T_w - T_\infty)} = \frac{1}{A_w(T_w - T_\infty)} \int_0^{+\infty} q(t) \cdot E(t) \cdot n \, dt \qquad\qquad (2)$$

Assuming uniform particle temperature *q(t)* is given by:

$$q(t) = \frac{\rho_p c_p V_p}{\tau} \cdot (T_w - T_\infty) \cdot \exp\left(-\frac{t}{\tau}\right) \tag{3}$$

where, according to ([26]), τ is the characteristic particle heating time:

$$\tau = \frac{\rho_p c_p d_p^2}{6} \frac{0.1}{K_g} \tag{4}$$

The residence time distribution for particle contacts with the exchange surface, *E(t)* results in a Gamma function with shape factor α equal to 1 or 0 ([27]-[29]):

$$E(t) = \frac{1}{\alpha! <t>^{\alpha+1}} \cdot t^\alpha \cdot \exp\left(-\frac{t}{<t>}\right); \tag{5}$$

where <t> is the mean residence time and its inverse, *f*, is usually defined as surface renewal frequency. The particle residence time distribution is the result of different scales of motion which include the mixing induced by bubbles motion, the Brownian-like microscopic motion typical of a granular flow and the particle displacement due to the peculiar fluid dynamic field near the surface. As previously reported, this last consists in the formation of a layer of higher gas velocity and bed porosity which may eventually give rise to the formation of superficial bubbles. Typical values of particle motion frequencies induced by this surface layer are around 5-20 Hz ([9], [23], [24]) while its average thickness is of 6-10 mm ([5]-[7]) a value close to Phillips' (25) estimation of the Kolmogorov length scale in a fluidized bed. Hence, its effect on particle motion can be considered more like a rigid high frequency displacement than a mixing phenomenon. For the cases of exchange surfaces with characteristic dimension of the order of a few millimetres, the surface renewal mechanism is mainly related to this phenomenon as its particle displacement is higher than the surface length and thus the contacting particles are completely renewed by this mechanism. This result has been clearly pointed out by the works of Boerefijn et al. ([30]) and Pence et al. ([23]) starting from the measures of transient heat transfer coefficients for probes with very small exchange surface area. For the case of surfaces with characteristic dimensions of the order of some centimetres, as those typically involved in heat exchange processes of industrial interest, this rapid displacement does not give rise to a complete renewal of particulate phase near the surface. This result can be connected to the presence of multiple particle-to-surface contacts which happen at different surface positions ([27], [29]). In these cases, lower values of surface renewal frequency are obtained (typically 0.7-2 Hz) and the leading phenomenon for particle mixing appears to be that induced by the bubbles motion. Accordingly, data on temperature–time signals (e.g. [12]) reveal that the surface temperature mainly changes at the passage of a rising bubble while only a limited oscillation is observed during the contact with the emulsion phase, when only the effect of surface gas layer is recognisable. For this reason, in a first analysis, for a fully bubbling or a slugging regime, the surface renewal frequency almost coincides with the bubble frequency.

Detailed values of surface renewal frequency are available in literature ([12], [13], [27]-

29, 31), and they appears as a function of several parameters as physical properties of solid and gas phases, fluidization velocity and particle diameter. For the case of silica sand and ballotini, a power law regression function for the description of these experimental data gives the following expression for *f:*

$$f = 2.324 \cdot 10^{-3} \cdot \left(U - U_{mf}\right)^{0.2446} \cdot \rho_g^{0.2760} \cdot d_p^{-0.7880} \qquad (R^2 = 0.97) \qquad (6)$$

Finally, the number of particles simultaneously in contact with the surface can be calculated by a simple geometric relation in function of the time and space averaged surface void fraction, ε_w:

$$n = \frac{A_s d_p (1 - \varepsilon_w)}{V_p} \qquad (7)$$

Substituting eqs.(3-7) in eq.(2) and solving the integral, the model equation in explicit form is:

$$h_{pc} = \rho_p c_p d_p (1 - \varepsilon_w) \frac{f}{1 + f\tau}; \qquad (8)$$

CALCULATION OF ε_w BY HEAT TRANSFER COEFFICIENT MEASURES

Starting from experimental values of surface-to-bed heat transfer coefficient in function of fluidization velocity, U, the values of surface void fraction can be derived by eq. (8) whose LHS is calculated by eq. (1). In line of principle, for each experimental run surface void fraction is a function of gas velocity, $\varepsilon_w(U)$ expressed by:

$$\varepsilon_w(U) = 1 - \frac{\left[h_{exp}(U) - h_{gc}(U) - h_r(U)\right] \cdot \left[1 + f(U) \cdot \tau\right]}{\rho_p c_p d_p \cdot f(U)}; \qquad (9)$$

where h_{exp} is the experimental value of the heat transfer coefficient. In first analysis, neglecting the bubble phase contribution to heat transfer, the term h_{gc} is assumed to coincide with the product (1- $\delta(U)$)·h_{mf} (11) where h_{mf} is the overall heat transfer coefficient at minimum fluidiza-tion. The bubble fraction, $\delta(U)$, is calculated by the classical Two Phase Theory using Darton's formula to estimate the bubble diameter (1) or by considering the expression proposed by Cui et al. (32), without appreciable differences. Radiative effects, usually relevant above 600-800 K are currently neglected.

The model has been applied to several experimental data on surface averaged heat transfer coefficient for small vertical surfaces (typically cylinders with thermal insulating ogival heads) in a broad range of experimental conditions, including variation of bed material properties, particle diameter and gas pressure (Table 1). Each experimental run consists in a complete set of heat transfer coefficients given in function of the gas velocity. By applying eq. (9) to the experimental data it is found that $\varepsilon_w(U)$ is almost independent by the gas velocity and for this reason, in the following, a unique, averaged value, denoted ε_w, is considered regardless the gas velocity. The values of ε_w mainly result a function of Archimedes number and void fraction at minimum fluidization, ε_{mf} (Figure 1). The best fitting of experimental results in terms of $\varepsilon_w - \varepsilon_{mf}$ and Ar is obtained by a logarithmic regression function as:

$$\varepsilon_w - \varepsilon_{mf} = 0.045 \cdot Log_{10}(Ar) \qquad (10)$$

with a maximum error of the 7%. This result can be explained considering that the higher void fraction requires higher drag forces exerted by the emulsion gas flow on the particles, i.e. higher gas density and interstitial velocity. In bubbling-slugging regimes this last is weakly dependent on gas flow rate and for different bed materials it is an increasing function of U_{mf} and hence of Ar (1). As a consequence, ε_w increases with particle diameter and gas density (hence for increasing pressure and decreasing temperature), accordingly with precedent studies (4, 6, 10).

Figure 2 reports the comparison of experiments with the proposed model at different temperatures (Figure 2A), pressures (Figure 2B), particle diameters (Figure 2C) and gas velocities (Figure 2D). For the sake of simplicity, in some cases only the averaged value of heat transfer coefficient for a fully bubbling fluidized bed, h_{max}, is considered. The model assures a correct evaluation of heat transfer coefficient in all the investigated conditions. In particular, it gives an acceptable estimation of heat transfer coefficient for gas temperatures as high as 900 K (where radiative effects are relevant) and for particle diameter as fine as 100 μm regardless the theoretical failure of the assumed single particle approach (25).

CONCLUSIONS

This paper presents a model for the description of particle convective heat transfer coefficient in bubbling/slugging fluidized beds. The model has been applied to a wide number of experiment results (Table 1) to give a significant expression for the averaged surface void fraction in a broad range of operating conditions. The estimated values of ε_w mainly increases with ε_{mf} and Archimedes number and are similar to experimental and numerical results available in literature (5-10). The model gives a profitable first order correlation between heat transfer coefficient and averaged surface void fraction which allows the application of heat transfer measures for the estimation of bed structure near immersed surfaces. Nevertheless, further analyses are cur-rently required for the evaluation of local values of the surface void fraction (5, 9) and for complex shaped surfaces (16).

NOTATION

A_s	Exchange surface area	(m^2)	h	Heat transfer coefficient	(W/m^{2}K)
α	Parameter of Gamma function	(-)	K	Thermal conductivity	(W/m K)
c	Specific heat	(J/g K)	n	Number of contacting particles	(-)
d	Diameter	(m)	q	Particle heat flow	(W)
δ	Bubble fraction	(-)	ρ	Density	(kg/m^3)
$E(t)$	Residence time distribution	(Hz)	T	Temperature	(K)
ε	Bed voidage	(-)	t	Time	(s)
f	Renewal frequency	(Hz)	U	Gas velocity	(m/s)
g	Acceleration of gravity	(m/s^2)	V	Volume	(m^3)

Subscripts			
∞	Bulk	mf	Minimum fluidization
exp	Experimental value	p	Particle
g	Gas	pc	Particle convective
gc	Gas convective	r	Radiative
ma	Averaged maximum	w	Wall
x			
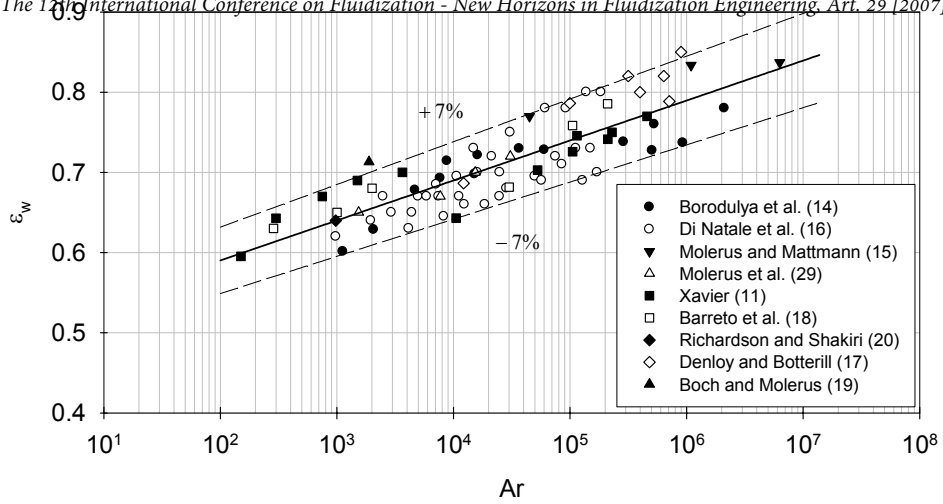
Figure 1 Value of surface void fraction derived from the application of the proposed model to the experimental data reported in Table 1 (symbols) compared with eq.(10) (full line).
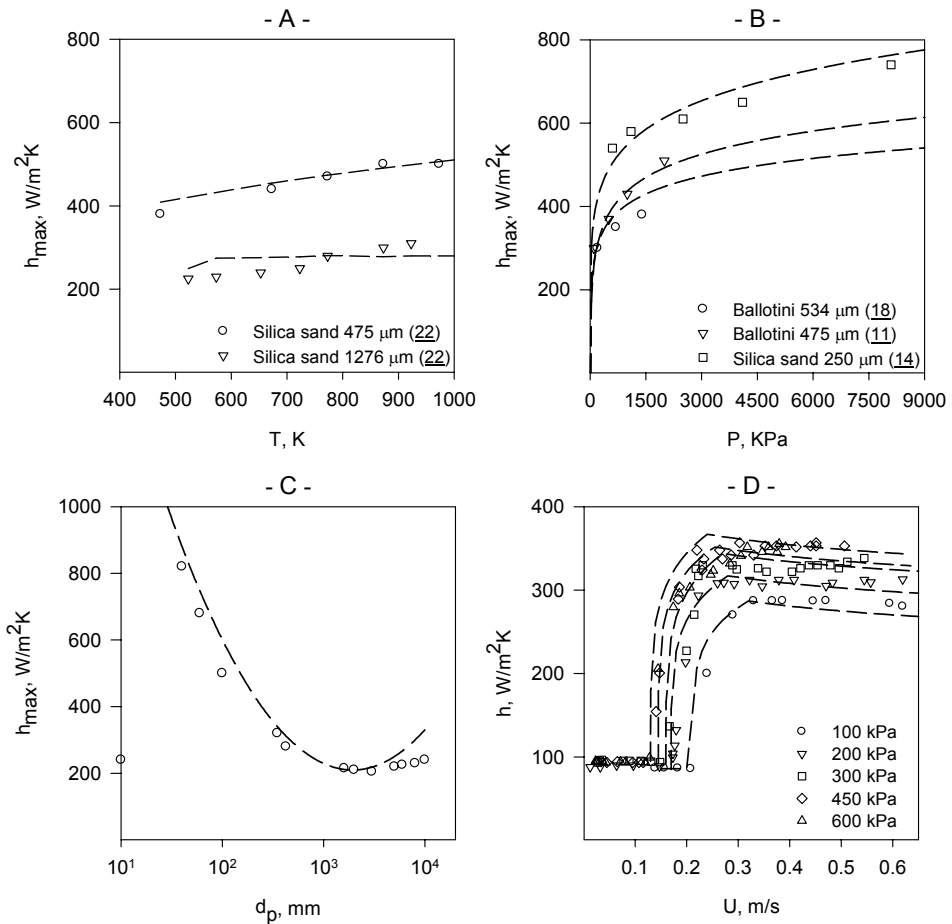
Figure 2 – Comparison between model prediction and experimental data . A – Effect of temperature. B – Effect of pressure. C – Effect of particle diameter (21); D – Effect of gas velocity (corundum 310 μm, 16).

Table 1: Experimental data used for the model calibration

Authors	Solid material	P, kPa	d_p, µm	Ar X 1000
Barreto et al. (18)	Ballotini	200 – 1400	534	30.0 – 209.8
Boch and Molerus (19)	Silica sand	100	272 - 488	1.9 – 10.9
Borodulya et al. (14)	Silica sand	600 – 8100	126 - 1220-	1.13 – 15.2
	Ballotini	600 – 8100	950	512.9 – 6760.8
Denloye and Botterill (17)	Silica sand	100 – 1100	1020	99.6 – 896.4
	Copper shot	400 – 900	160 - 620	315.8 – 710.5
Di Natale et al. (16)	Ballotini	100 – 600	210 - 670	1.0 – 144.0
	Silica sand	100 – 600	290 - 670	2.5 – 184.2
	Corundum	100 – 600	310 - 590	4.1 – 171.0
	Carborundum	100 – 600	300 - 590	3.6 – 147.2
Molerus and Mattmann, (15)	Ballotini	100	770 - 4000	44.7 – 630.9
Molerus et al. (29)	Ballotini	100 – 2000	250	1.5 – 30.7
Richardson and Shakiri (20)	Ballotini	100	215	1.0
Xavier (11)	Ballotini	100 – 2000	475 - 615	10-5 – 457.8

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