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Cold model characterisation of a fluidised bed catalytic reactor by means of instantaneous pressure measurements

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Abstract

The analysis of instantaneous pressure fluctuations is utilised in a cold modelling study of a pilot scale fluidised bed catalytic reactor for ethane oxychlorination. It is demonstrated that the fluid dynamic behaviour of a cold model of this fine particle (Geldart group A) system is well characterised by the frequency spectrum, the standard deviation and the cross-correlation of pressure signals detected at different locations within the bed and in the freeboard.

The influence of operating parameters, such as gas flow rate, initial bed height, reactor configuration (internals) and average value of the local pressure, is investigated, and a comparison is made with the features exhibited by Geldart group B systems (fluidised bed combustors) for which the same diagnostic techniques have been applied.

Finally, the applicability to fine particle suspensions of literature correlations for dominant frequency and propagation velocity of pressure waves is discussed with reference to the combined influence of the eruption of many small bubbles at the bed surface, and the compressibility of the gas contained within the particle-bed interstices. © 2002 Elsevier Science B.V. All rights reserved.

Keywords: Fluidised bed reactors; Pressure fluctuations; Modelling

1. Introduction

Pressure measurements have long been used to monitor operating conditions in industrial fluidised-bed chemical reactors. Applications include the measurement of overall bed expansion, which is the factor primarily responsible for maintaining reactor conversion, and the control of elutriated particle-fines recovery (and return to the bed through the cyclone-dipleg circuit) from the gas product stream.

More recently, in addition to static pressure monitoring, instantaneous pressure fluctuations around their average value have been recorded and analysed. These data are strongly related to the intrinsically heterogeneous state of fluidised suspension. The detection and analysis of pressure fluctuations has become a powerful inspection tool in cold model studies and pilot scale reactor evaluation. There are also proposals in hand for direct applications of the technique to industrial reactors [1].

Following on from the first experimental study of pressure fluctuations in fluidised beds [2] different dynamic pressure measurement techniques have been developed: their application to the characterisation of fluidised bed hydrodynamics has been reviewed by Yates and Simons [3]. Several methodologies have also been proposed for relating the time series of the experimentally recorded pressure fluctuations to the fluid dynamic state of the bed [4,5]. In most cases the fluidised suspension may be regarded as being in a quasi-stationary state, with the mean values of its physical properties being time independent over a finite period of time. The transduced, oscillating signals are then subjected to statistical and spectral analysis to yield standard deviations, power spectrum density functions, and auto- and cross-correlation functions.

Numerous, practical applications of the instantaneous pressure fluctuation measurements are reported in the fluidisation literature. Jin et al. [6] associated the transition from bubbling to turbulent fluidisation with a maximum in the standard deviation of the pressure fluctuations, thereby defining, in terms of dynamic pressure, the boundary between these two fluidisation regimes. Kai and Furusaki [7] found that the standard deviation of the differential pressure fluctuations (measured between two pressure probes located a few centimetres apart in a lab scale reactor)

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Nomenclature

Ar	Archimedes number
d_p	average particle size
De	density ratio
D_t	bed diameter
f_d	dominant frequency
f_f	fundamental frequency
f_n	natural frequency
Fr	Froude number
g	gravitational acceleration
H_0	initial bed height
Le	length scale ratio
P	pressure
R	gas-law constant
T	temperature
U	gas-fluidising velocity
U_{mb}	minimum bubbling velocity
U_{mf}	minimum fluidization velocity
U_s	sound velocity

Greek letters

ε	voidage
ε_{mb}	minimum bubbling voidage
ε_{mf}	minimum fluidization voidage
μ_g	gas viscosity
ρ_b	density of the particle suspension at minimum bubbling voidage
ρ_g	gas density
ρ_p	particle density
σ	standard deviation of pressure fluctuations

was well correlated with the conversion obtained in the methanation reaction of carbon monoxide. Chong et al. [8] utilised the differential pressure fluctuation variance to implement a control system for maintaining conditions close to the minimum fluidisation point all along a deep fluidised bed, drawing-off gas at different levels to counteract its expansion. Wilkinson [9], in order to evaluate the minimum fluidisation velocity without having to de-fluidise the bed, employed a method based on the measurement of the standard deviation of the pressure fluctuations at different fluidising velocities; the method enables the constancy of the average particle size of the bed inventory to be assessed, in particular under high temperature conditions, and in the presence of sticky particles. Schouten and van den Bleek [1] suggested a method that uses the short-term predictability of local pressure fluctuations to provide a warning signal of incipient agglomeration of the particles, and possible de-fluidisation, in high temperature fluidised bed reactors. They also compared experimental pressure fluctuations with corresponding ones calculated from simulations carried out using an Eulerian model of a gas–solid fluidised bed [10].

1.1. "Cold model" studies and pressure fluctuation analysis

The most noteworthy application of dynamic pressure measurements is probably in the validation of *cold modelling* studies based on the fluid dynamic similarity rules. In process design, in the absence of direct empirical knowledge of the likely behaviour of the fluidised suspension of interest, some means of estimating the fluid dynamic characteristics becomes essential, as these enter into the process model by means of which the conversion in an envisaged commercial reactor is evaluated. One method for obtaining the necessary information is by means of scaling relations, which enable laboratory experiments to be performed on relatively simple cold models that relate, with regard to fluidisation characteristics, to the proposed reactor. In practice there are few practical alternatives to this procedure. Ideally, one would wish to have available a reliable model for the fluidisation dynamics that could be solved numerically using boundary conditions that relate to a proposed design, but approaches of this kind are only at a very early stage of development [11,12].

Fitzgerald et al. [13] and Glicksman [14] first applied dimensional analysis to the equations of change for a fluidised suspension, identifying the dimensionless groups characterising the fluid dynamic behaviour. As long as the equations themselves represent an adequate description of the controlling mechanisms, it becomes possible to construct and operate a *cold model* in such a way that its scaled performance is identical to that of an envisaged fluidised bed reactor, the fluid and particle properties and fluidising velocity for the model being chosen so as to match the defining dimensionless groups. In this way such things as bed expansion and the minimum fluidisation and bubbling velocities, together with less easily measurable properties relating to bubble characteristics, which strongly influence the extent of heat and mass transfer in a bed, can be studied in a simply constructed prototype system. As indicated above, pressure fluctuations are intrinsically related to the heterogeneous nature of the system, and are therefore able to provide a readily available measure of the quality of the fluidisation.

Only recently such investigations have been extended to fluidised bed chemical reactors, which are generally characterised by smaller particle sizes, belonging to the group A of the Geldart classification [15]. In this area, the approach has been largely influenced by the concept of particle–particle interactions [16], directing the study of fine particle systems away from dimensional analysis and toward experimental investigations leading to empirical and semi-empirical correlations [17]. The increasing reliability of physical models for particulate fluidisation [18] has progressively encouraged a more comprehensive approach, taking into consideration the fluid dynamic aspects inherent in the behaviour of such systems. Pressure fluctuation measurements have confirmed the validity of this approach, both by comparison of fine and coarse particle system behaviour at high temperatures

[19] and by investigation of the behaviour of different cold models of fluidised bed reactors [20].

This paper describes an experimental, cold modelling study of the fluid dynamic behaviour of a pilot scale, fluidised bed reactor for ethane oxychlorination. This has involved the analysis of instantaneous pressure fluctuations, both in the frequency and pressure intensity domains, as functions of the following variables: gas flux, initial bed height, reactor configuration (for which a vertical tube bundle was used to simulate the proposed heat transfer system), average value of the local pressure, and the axial and radial location of the pressure probe within the bed. The results show that dynamic pressure measurements represent an effective way of monitoring fluidisation quality in the bed, and are, at the same time, quite sensitive to changes in the operating parameters. The study also reveals that statistical analysis of the pressure signal provides a powerful tool in the development of a reliable control strategy for a fluidised bed reactor.

2. Experimental

The experimental apparatus, shown schematically in Fig. 1, consists essentially of a 376 mm i.d. and 9 m high Plexiglas column containing the fluidised bed, the gas feeding system, the dynamic pressure measuring and recording system, measurement devices and ancillaries.

2.1. Cold model design and operation

The dynamic similarity of the experimental rig with the oxychlorination fluidised bed reactor has been obtained by imposing geometrical similarity, together with equality of the following dimensionless parameters:

$$Ar = \frac{d_p^3 \rho_p (\rho_p - \rho_g) g}{\mu_g^2}, \qquad De = \frac{\rho_g}{\rho_p},$$
$$Le = \frac{d_p}{D_t}, \qquad Fr = \frac{\mu_g}{(gD_t)^{0.5}} \tag{1}$$

In addition, the dimensionless particle size distribution (PSD) and particle shape were also approximately matched. The catalyst particle size distribution is shown in Fig. 2.

Once the characteristics and operating conditions of the pilot reactor are known, and for obvious reasons air is chosen as the cold model fluidising gas, the only additional variable which can be freely fixed is the operating pressure. At this point the dynamic similarity conditions imposed above are sufficient to define the test-rig size, particle properties and gas flow rates. The higher the pressure permitted in the cold model, the smaller is its corresponding size and the average particle diameter. In this work it was decided to keep the geometric scale ratio close to 1. This implies that the operating pressure level should be such as to render the kinematic viscosity of air equal to that of the reacting gases (about

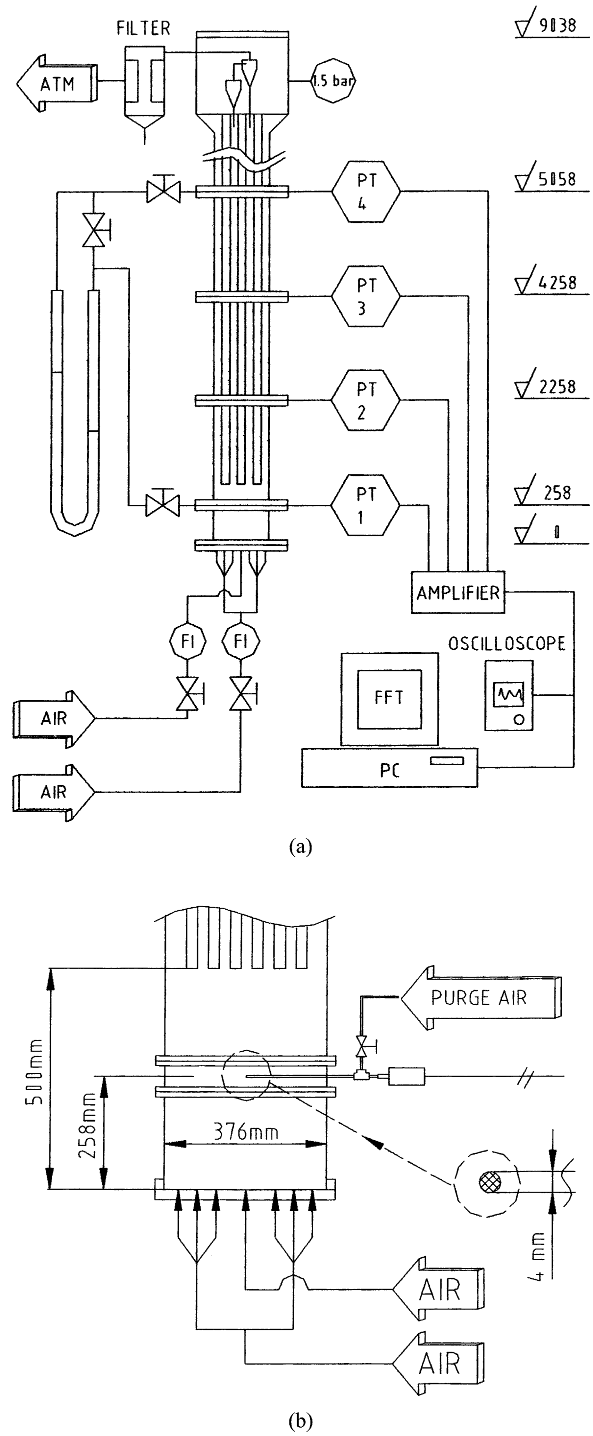

Fig. 1. (a) The overall experimental rig; (b) a schematic view of the bottom part of the fluidising column.

250 kPa in the practical case examined here). Moreover, the condition of keeping equal values for the Froude number implies that the cold model should be operated at the same volumetric gas flow rate as the reactor.

The Plexiglas column contains a bundle of 19, 37 mm o.d., vertical tubes, positioned, with a 70.5 mm triangular pitch, 500 mm above the bottom plate and extending to the top of

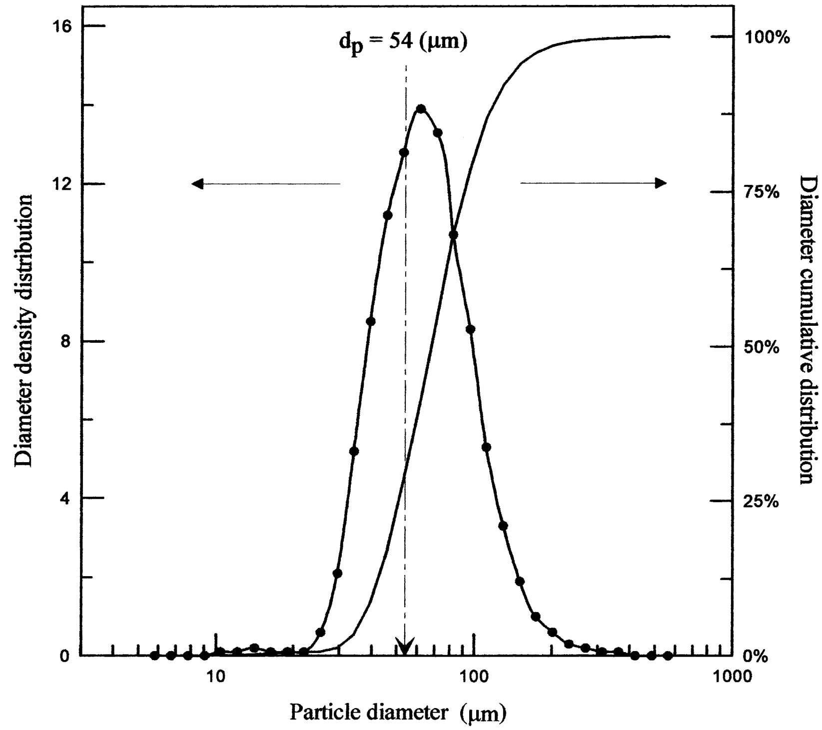

Fig. 2. Catalyst particle size distribution.

the column; this arrangement simulates the heat exchange system in the reactor. Tests have been carried out with and without this tube bundle in order to check its influence on the fluid dynamic conditions. The gas distributor consists of seven nozzles, equally and symmetrically spaced over the bottom plate; the flow rate through the central nozzle can be fixed independently, whereas the overall air stream fed to the lateral nozzles is controlled and measured before being divided among them. In the experimental tests, the ratio of the volumetric gas flow through the central nozzle to the total feed rate was fixed at 18.5%; this value is dictated by the reactor operating conditions.

In the design of the distributor plate, attention has been paid to the preservation of geometric as well as dynamic similarity: the ratio pressure drop through the distributor to pressure drop through the bed being kept the same in the model as in the reactor. A rapid change in gas density and velocity occurs in the reactor, just above the distributor, as a result of the difference in temperature of the entering gas and the reactor itself; this is not the case in the cold model, where the gas density change is small, due only to compressibility effects. The cold model distributor design is therefore subjected to the additional constraint that the ratio of the gas momentum in the nozzles to that in the bed is preserved, so that the momentum imparted to the particles by the gas jets in the model is matched to that in the reactor.

To avoid accumulation of electrostatic charges, which could affect the model performance, metal components of the model (distributor plate, vertical tube bundle) were all earthed.

Table 1
Main properties of the fine catalyst and the sand batch utilised in the experimental tests; values for U_{mf}, U_{mb}, ε_{mf} and ε_{mb} are those measured at ambient pressure and temperature in the laboratory test rig

	Fine catalyst Geldart group A	Sand Geldart group B
d_p (mm)	54	348
ρ_p (kg/m^3)	1439	2640
U_{mf} (m/s)	0.0016	0.12
ε_{mf}	0.40	0.40
U_{mb} (m/s)	0.0059	0.12
ε_{mb}	0.50	0.40

The main physical properties of cold model bed inventory are listed in Tables 1 and 2. According to the Geldart [15] classification, this is a typical group A powder. In order to establish some comparison with typical Geldart group B

Table 2
Minimum fluidisation and minimum bubbling data measured with the cold model ($P = 220\,\mathrm{kPa}$)[a]

	Measured	Calculated
ε_{mf}	0.43	0.43 (assumed)
U_{mf} (m/s)	0.0016	0.0016 [28]
	0.0016	0.0014 [29]
	0.0016	0.0019 [30]
ε_{mb}	0.47	0.51 [31]
U_{mb} (m/s)	0.0059	0.0051 [32]

[a] Comparison with calculated data.

behaviour, a sample of sand particles was also tested: its main properties are also reported in Table 1. Bed heights were examined over the range 0.5–4.5 m.

2.2. Dynamic pressure measurement and recording system

As shown in Fig. 1a, four piezo-electric pressure transducers (Kistler 7261), connected by means of 4 mm i.d. horizontal pressure probes, were installed along the bed at distances of 258, 2258, 4258 and 5058 mm above the distributor plate; the lowest position corresponds to the middle of the bottom region of the bed, which is free of internals, whereas all the others were inserted in the tube bundle. Each probe was 250 mm long, and its tip could be moved across the bed, from the axial to lateral positions. The probe tips were protected by a wire net to inhibit entry of powder; an air purging flux was also utilised for the same purpose, as shown in Fig. 1b. It was verified that the experimental set-up did not distort the amplitude and the phase of the measured signal in the range of interest.

The transducers outputted to a multi-channel charge amplifier, Kistler 5019A, equipped with a low pass (100 Hz) filter, and were tuned to obtain maximum signal amplification without overload. The amplified voltage signal was then fed to a data acquisition system, processed by means of the LabVIEW® software package, thereby enabling discrete values of the pressure fluctuations to be stored in a PC.

A preliminary check of the data acquisition/processing system was performed by recording the pressure fluctuations induced by a low viscous liquid (*n*-hexane) oscillating freely in a U-tube. The transducers were calibrated using regression analysis of the data acquired when measuring known values of static pressure.

The data acquisition frequency for each channel, was maintained in the range 100–130 Hz, much higher than that typically observed in the gas-fluidised bed under study (less than 10 Hz); this choice is mainly dictated by the precision requirements for the discrete cross-correlation function evaluations. The time interval for data acquisition was fixed by imposing the condition that the standard deviation of the pressure fluctuations for time series obtained under identical operating conditions should differ by no more than 5%. This condition was usually satisfied for acquisition times in the range of 2–3 min. It was also verified that noise levels did not significantly affect the measured values: blank signals were always negligible in comparison with the signal of interest over the whole range of operating conditions.

Stored data were processed by means of statistical algorithms to obtain standard deviations, power spectral density functions (PSDF), discrete auto- and cross-correlation functions, and frequency distributions; commercial software routines, designed to optimise data processing times, were used for these purposes. For the data files collected in this work, the frequency and time domain calculation times never exceeded 6 s (using a PII processor—excluding the time needed to load the data in the PC memory). Both data acquisition and processing times could certainly be significantly reduced for on-line control applications.

3. Results and discussion

3.1. Preliminary tests

The minimum fluidisation velocity was obtained by vigorously fluidising the bed, and then measuring the pressure drop while progressively decreasing the air flux. This test was performed at two initial bed heights 0.99 and 2.54 m. The results of preliminary tests are reported in Table 2, together with estimated values. The same table reports experimental and calculated values of the bed voidage and corresponding values of air flux at the minimum bubbling point. This latter condition was detected by examining the pressure fluctuations while gradually decreasing the gas flow rate: this shows quite clearly the extinction of bubbles and the transition between stable and unstable states; the average bed voidage was then obtained from the visual observation of the bed height. As is clear from Table 2, the experimental findings are in good agreement with both the minimum bubbling voidages predicted by the *particle-bed model* [31] and calculated minimum bubbling velocities [32].

3.2. Analysis of pressure fluctuations in the frequency domain

The aim here has been to obtain sufficient experimental evidence to relate the power spectrum of the pressure fluctuations to the hydrodynamic state of the cold model. For this purpose, dynamic pressure measurements were performed varying the powder inventory so as to obtain initial bed heights varying from 0.5 to 4.5 m. Air fluxes were also varied over a range of 4–30 times the minimum bubbling velocity. All pressure transducers were used routinely in the experimental tests, so that for each set of operating conditions, pressure measurements were logged at three levels within the bed and in the freeboard.

Under bubbling fluidisation conditions, the pressure fluctuations comprise a vigorously oscillating signal displaying a dominant range of frequencies. Fig. 3 shows a typical frequency spectrum obtained in a laboratory scale fluidised bed (160 mm i.d. Plexiglas column; initial bed height of 53 cm) using the same catalyst particles as in the cold model, and fluidised by air at a flux of three times the minimum bubbling value. In order to provide a meaningful comparison, the corresponding spectrum obtained with sand particles (a typical Geldart group B system) at similar operating conditions is reported in the same figure. It has long been recognised [2] that lower amplitude/higher frequency pressure fluctuations correspond to smoother fluidised states, involving small bubbles and better gas solid contact. The comparison shown in Fig. 3, together with the corresponding standard deviations

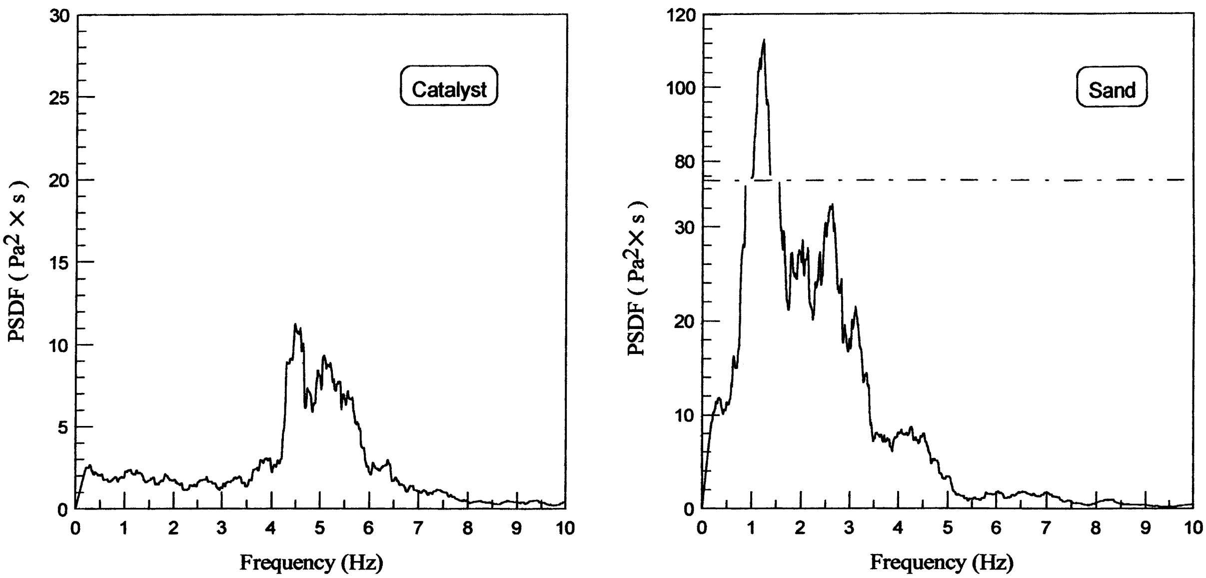

Fig. 3. Typical PSDF curves obtained in the laboratory test rig with Geldart group A and group B systems, respectively.

of the two signals reported in the following section, provides a clear indication as to why group A systems are often to be preferred as chemical reactors. It is also clear from the figure that the Geldart group A system displays a comparatively broader power spectrum band-width, as previously noted by Dhodapkar and Klinzing [21], corresponding to the presence of a number of smaller bubbles, even in the upper region of the bed, which result in significant side frequencies.

The power spectrum delivers the dominant frequency, f_d, of the pressure fluctuations, defined as the location of the maximum peak of the spectral density function. The values obtained in this way, for different initial bed heights and at different bed levels, are reported in Fig. 4 as functions of gas flux. The figure shows the dominant frequency to be a strong function of initial bed height, only marginally affected by the fluidising velocity, the axial and radial position of the probe, and the presence of a tube bundle (which significantly reduces the bed mass at comparable initial bed heights). This result is in full agreement with previous experimental findings reported for Geldart group B powders [4]. Moreover, the fact that for a given initial bed height the dominant frequency of the pressure fluctuations, at all levels along the column, remains essentially constant implies that the pressure fluctuations themselves are not attributable to the local bubble flux near the tip of the probe, but relate to the overall heterogeneous behaviour of the fluidised system—with bubble eruption at the bed surface and bubble formation at the distributor plate playing the most important roles [22].

The dependency of dominant frequency on the initial bed height is reported in Fig. 5, on logarithmic scales. The dots and the stars represent the experimental values found in this work. The solid line shows the predictions of the model developed by Baskakov et al. [23] for the fundamental frequency, f_f, of bed surface oscillations associated with the eruption of bubbles:

$$f_f = \frac{1}{\pi}\sqrt{\frac{g}{H_0}} \tag{2}$$

and the dash–dot line gives the values of the natural frequency, f_n, of pressure fluctuations in a homogeneously

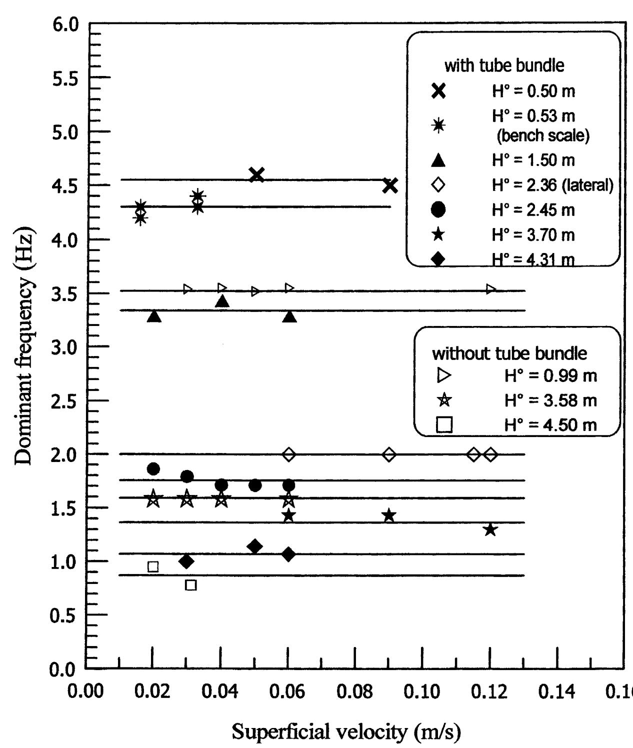

Fig. 4. Dominant frequency of the pressure signal as a function of the gas superficial velocity and initial bed height.

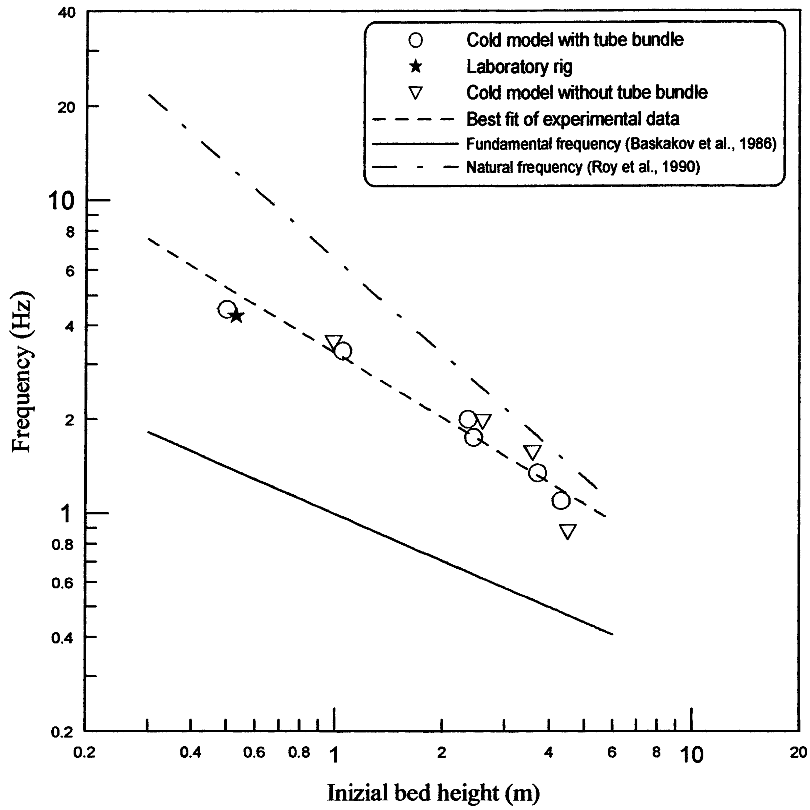

Fig. 5. Experimental and calculated values of the dominant frequency of the pressure fluctuations.

fluidised bed excited by a vertical impulse, according to the model of Roy et al. [24]:

$$f_n = \frac{1}{4H_0}\sqrt{\frac{P}{\rho_b \varepsilon_{mb}}} \tag{3}$$

where ρ_b is the density of the particle suspension at the minimum bubbling void-fraction, ε_{mb}: $\rho_b = \rho_p(1 - \varepsilon_{mb})$.

In the approach followed by Baskakov et al. [23], the behaviour of the bubbling fluidised bed is related to that of an incompressible fluid oscillating in a U-tube: the dominant frequency of pressure fluctuations is simply evaluated as the fundamental frequency of a harmonic oscillator. It is clear from Fig. 5, however, that this model considerably underestimates the experimental results over the whole range of experimental conditions, although the trend is in reasonable agreement with the empirical findings (the straight line best-fit has a gradient of −0.65 as opposed to the prediction of −0.5). Eq. (2) has been found to relate well to experimental evidence obtained with Geldart group B fluidised systems; the much higher values of dominant frequency obtained in this study provide a clear confirmation that fine particle suspensions are characterised by a larger number of small bubbles, which do not coalesce into a few, much-larger ones in the upper part of the bed.

On the other hand, comparison with the model of Roy et al. [24] shows predictions of natural frequency to be greater than the dominant frequency observed under bubbling conditions in the present study, and to become increasingly closer to the experimental data as the initial bed height is increased beyond 2 m.

Under these conditions it appears reasonable to describe the overall pressure fluctuation behaviour in terms of the eruption of many small bubbles at the bed surface, together with a significant contribution that is due to compressibility of the gas contained within the interstices of the particle suspension, this latter mechanism corresponding to that postulated by Roy et al. [24]. This dual dependency is discussed further below, following an examination of the propagation velocity of the pressure signal.

Even though some of the experimental tests were carried out with initial bed heights of several meters, giving rise to significant differences in static pressure at the sensor locations along the bed, dominant frequencies obtained at all measurement positions remained essentially the same for a given initial bed height. This point is of particular relevance to industrial chemical reactor applications, which often involve considerably taller units than do other fluidised bed systems.

Fig. 6 shows a complete PSDF obtained for the case of maximum initial bed height (4.31 m) and the inclusion of vertical tubes, which simulate a heat transfer system in the industrial unit. In spite of this considerable height, the spectrum still keeps the general shape typical of freely bubbling

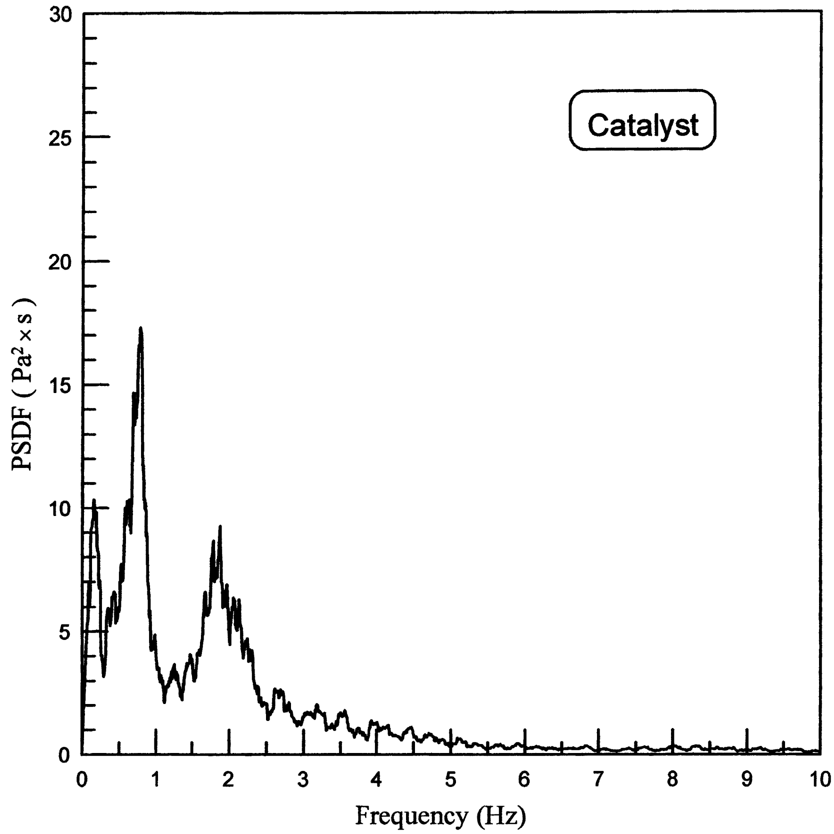

Fig. 6. Power spectrum density function with initial catalyst bed height $H_0 = 4.31$ m.

systems, with no evidence at all of the development of slugging conditions, which would result in a substantial reduction in the range of characteristic frequencies. Bubble evolution along the bed follows a quite different pattern to that of Geldart group B systems in which coalescence predominates, leading to a progressive increase in bubble size with elevation along the bed. This observation becomes particularly striking in the presence of the simulated heat exchange system: although this reduces considerably the free cross-section of the bed, bubbles remain small enough to rise almost undisturbed between the tubes.

3.3. Analysis of pressure fluctuations amplitude

The concluding remarks of the previous section provide an appropriate introduction to a discussion on the influence of operating parameters on the amplitude of pressure fluctuations, which is certainly related to the size of the bubbles bursting through the bed surface. Fig. 7 shows experimental values of the standard deviation of the pressure fluctuations, σ, expressed as a function of the dimensionless excess gas flux. These results were obtained in laboratory tests performed both with the catalyst powder used in the cold model simulation experiments, and with the sand particles used to obtain the PSDF curve reported in Fig. 3. All laboratory tests were performed in the same test rig, referred to in the previous section. The standard deviation values have been normalised by dividing them by the average pressure drop through an unit height of the corresponding bed; so that the ordinates in Fig. 7 directly provide a measure of the bubble equivalent diameter at the bed surface (when it is considered that for a normal distribution, the standard

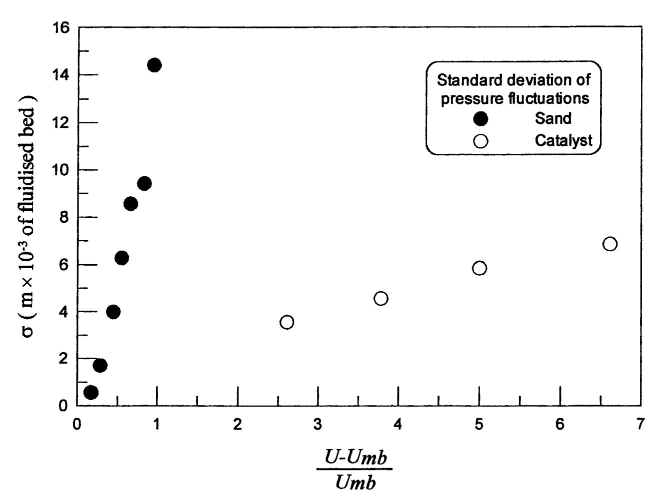

Fig. 7. Experimental values of standard deviation of pressure fluctuations obtained in the laboratory test rig with Geldart group A and group B systems, respectively.

deviation is about 15% of the maximum range of the pressure fluctuations [22]). This shows that the maximum bubble size is confined to about 4 cm in the case of the catalyst powder, whereas for sand particles it rapidly reaches 10 cm—hence the tendency of the standard deviation to increase much more rapidly in this latter case.

The data obtained with the cold model provide a more complete and quantitative picture of the influence of gas flux on the standard deviation of the pressure signal. Fig. 8 shows values of σ obtained for the bottom region of the bed, both on the axis and close to the wall, for an initial bed height of 2.36 m. Similar trends were observed for all values of initial bed height investigated in this study. There is a clear tendency for the pressure fluctuation amplitude to level-off with increasing fluidising velocity, due to the more intense interaction among the bubbles, leading finally to a transition from the bubbling to the turbulent regime. This effect has been widely discussed in the literature [25]. A good correlation has also been reported [26] between the approach to a maximum in standard deviation of the pressure fluctuations and increasing gas mixing in the bed.

The experimental data reported in Fig. 8 were obtained with the simulated tube bundle in the bed. Removal of these tubes was found to have little effect on the values shown in this figure. On this basis, and as a result of the discussion in the previous section, we are led to conclude that bubble size and frequency are not significantly affected by the presence of vertical tube internals, at least with regard to the configuration adopted in this study. They may have an influence on the bubble travelling velocity, by providing preferential rising paths along the smooth surfaces of the tubes. Indirect evidence of this phenomenon is provided by visual observations of bed expansion under similar operating conditions with and without the tube bundle: these show that the presence of the bundle leads to the bed being somewhat less expanded, and to the expansion gradients with gas flux being somewhat smaller.

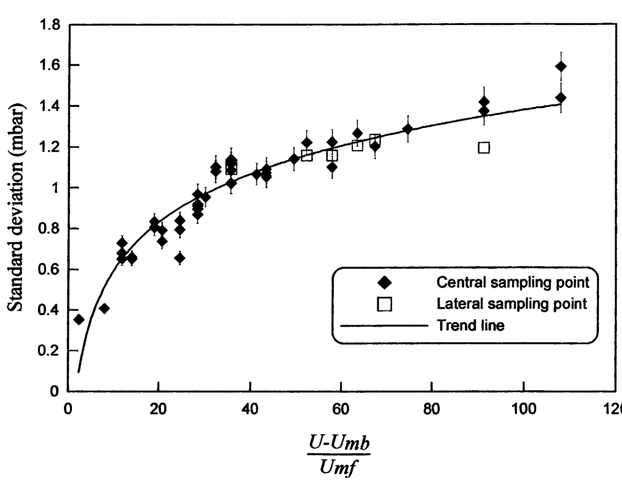

Fig. 8. Standard deviation of the pressure signals in the bottom region of the cold model, as a function of fluidising velocity.

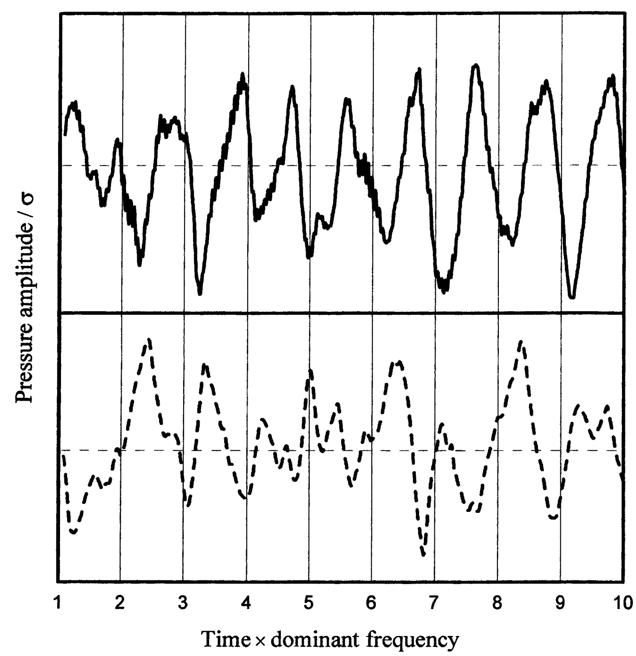

Fig. 9. Pressure fluctuations recorded inside the bed (- - -) and in the freeboard (—): $H_0 = 3.58$ m, $U = 0.04$ m/s.

3.4. Propagation velocity of pressure fluctuations along the fluidised bed

The cross-correlation of pressure signals detected simultaneously at two different bed levels enables an estimate to be made of the propagation velocity of a gas pressure wave through bed. Dynamic pressure measurements, obtained with two probes, placed axially in the bed or, alternatively, one in the bed and the other in the freeboard, were used for this purpose. Results were obtained for different values of the initial bed height.

Before presenting the results obtained by means of the cross-correlation analysis, it is worth pointing out that the freeboard pressure signal exhibits an intrinsic 180° phase shift with respect to that within the bed itself: a maximum in the one corresponding to a minimum in the other (Fig. 9). This finding clearly indicates that the main origin of pressure fluctuations both within and above the bed relate to bed surface phenomena. This surface oscillates between minimum and maximum levels as a result of bubbles bursting through it: as the bed expands, the gas in the freeboard is compressed, and vice versa when the bed contracts, following bubble emergence from the bed. These simple observations illustrate the correspondence between pressure disturbances propagation and gas compression and expansion phenomena, which will be expanded upon later in this section on quantitative grounds. In addition, they highlight the significance and convenience of monitoring fluidised bed behaviour by means of a pressure probe located in the freeboard: this represents a useful, non-intrusive diagnostic method, avoiding the erosion and abrasion problems

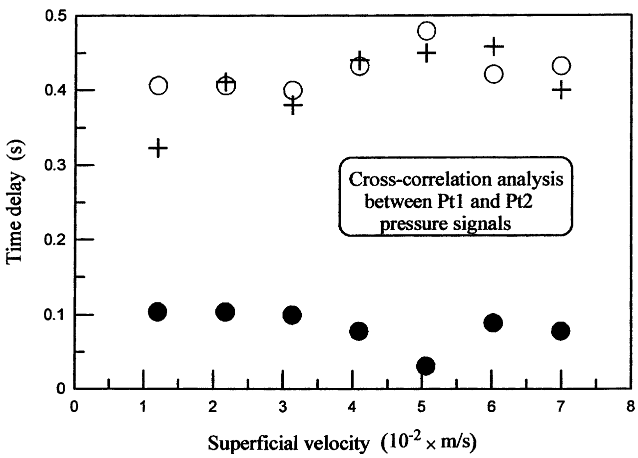

Fig. 10. Results of the cross-correlation analysis: (●) time delay of signal Pt1 with respect to Pt2; (+) time lead of signal Pt2 with respect to Pt1; (○) estimated time lead of signal Pt2 with respect to Pt1 (from the knowledge of the dominant frequency and the time delay).

associated with internal bed monitoring under severe conditions of operating. It appears to offer a more attractive alternative to pressure monitoring within the plenum chamber [27], which can be hindered by the high pressure drop through the distributor, and influenced by the volume of the windbox itself.

The results of the cross-correlation are reported in Fig. 10, for the signals recorded by the transducers Pt2 and Pt1, and for different values of gas flux: the black dots show the time lag of the pressure fluctuations detected at the bottom of the bed with reference to the corresponding fluctuations measured higher up in the column; the crosses show the time lead calculated by exchanging between themselves the pressure signals in the cross-correlation algorithm; finally, the open circles show the time lead obtained from the knowledge of the signal dominant frequency and of the time lag. The good agreement between corresponding values of the time lead calculated with different methods assures about the consistency and reliability of the results of the cross-correlation. It is clear from the figure that the delay is affected very little by the gas flux, i.e. by the amplitude of the pressure fluctuations. The bed height is about 2.60 m, in this case, so that the probe linked to the transducers are both located inside the bed (see Fig. 1a).

Knowing the vertical distances between the respective pressure probes enables the propagation velocity of gas pressure disturbances to be evaluated. The measured velocity was found to be approximately 20 m/s. This value may be compared with the estimate of the travelling velocity of sound waves in fluidised bed proposed by Roy et al. [24]:

$$U_s = \sqrt{\frac{RT\rho_g}{\varepsilon(\rho_s(1-\varepsilon)+\rho_g\varepsilon)}} \tag{4}$$

This model assumes that the fluidised bed can be described as a continuous, single-phase system, in which the particles provide most of its mass, and gas compression is ideal and isothermal (the presence of particles allows for very rapid heat transfer). Eq. (4) predicts a sound wave velocity of 26 m/s at a void-fraction of 0.5, in quite reasonable agreement with the experimental findings, as was also the case for all conditions tested.

4. Conclusions

This study shows that instantaneous pressure measurements and data analysis enable important features of the fluid dynamic behaviour of fine particle fluidised systems (Geldart group A) operating in the heterogeneous regime to be monitored, and the influence of operating parameters such as initial bed height and reactor configuration (internals) to be assessed.

The experimental technique adopted has been widely applied to the study of fluidised bed combustors, where the particles are larger and more dense (Geldart group B systems). The present work demonstrates the usefulness of the technique in the important field of catalytic chemical reactors, and the applicability of cold modelling studies to such systems for estimating the fluidisation quality to be expected in a proposed industrial unit.

At high reaction operating temperatures, changes in the solid particle properties may occur, which could affect the fluidisation quality. In such cases the significance of cold model studies is certainly reduced. However, dynamic pressure measurements still provide a useful technique for monitoring the behaviour of the reactor itself. For such applications, the feasibility, investigated in this study, of locating a pressure probe in the freeboard would appear to offer a number of diagnostic advantages, circumventing, at least to mitigating, obstruction and erosion problems.

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