A small-scale regularly packed circulating fluidized bed

Part I: Hydrodynamics

A.G.J. van der Ham, W. Prins and W.P.M. van Swaaij

Chemical Reaction Engineering Laboratories, Chemical Engineering Department, Twente University, PO Box 217, 7500 AE Enschede (Netherlands)

(Received April 26, 1993; revised November 29, 1993)

Abstract

The present investigation is based on the idea of intensifying the gas–solids contact in a circulating fluidized bed by introducing obstacles into it. Such obstacles may effectively suppress radial inhomogeneities in the solids flux and concentration, increase the dynamic solids hold-up, and break up solids clusters. This article (Part I) deals with the hydrodynamics (pressure drop and solids hold-up) investigated at ambient conditions, for cocurrent upward flow of air and microsize solid particles (FCC, 70 μm diameter) over a regularly structured inert packing introduced into the riser part of a circulating fluidized bed unit. The packed section has a height of 0.48 m, a cross-sectional area of $0.06 \times 0.06 \text{ m}^2$, and contains regularly-stacked 0.01 m diameter Perspex bars as the obstacles meant to enhance the gas–solids contact. Slide-valves mounted above and below the packed section can be used to trap the solids inventory and determine the (dynamic) solids hold-up. Gas and solids mass fluxes have been varied in the range of $0.7 \leq G_g \leq 4.4$ and $0 \leq G_s \leq 15 \text{ kg m}^{-2} \text{ s}^{-1}$, respectively. Part II will report on the results of gas–solids mass transfer measurements, which have been carried out in the same set-up at comparable experimental conditions. Results of this work show that: (i) the pressure gradient over the packed section increases linearly with increasing solids mass flux, but faster than linearly with increasing applied gas mass flux, (ii) the dynamic solids volume fraction can be described quite well by the correlation $\beta_{\text{dyn}} = 0.0084 G_s G_g^{-1.22}$ for almost the entire range of applied gas and solids mass fluxes, (iii) the value for the solids friction factor derived for the gas flux range $0.7 < G_g < 3.7 \text{ kg (m}^{-2} \text{ s}^{-1})$ varies from 1.4 to 2.5 and is linear with the solids volume fraction. These f_s values are about 2 to 3 decades higher than those obtained from f_s correlations derived for dilute-phase pneumatic conveying lines operated under the same experimental conditions.

Introduction

This study is part of a research program concerning high-temperature desulfurization of gasifier product-gas by contacting it with a regenerable sorbent. The specific properties of the applied sorbent, *viz.* a small mean particle diameter and a high reactivity towards hydrogen sulfide, has resulted in the selection of the circulating fluidized bed (CFB) as a proper contactor.

Much research has already been done in the field of CFB technology. Basu [1], Basu and Large [2], Basu *et al.* [3], Yerushalmi and Avidan [4], Yoshida and Mineo [5] and Grace [6] provide a useful overview of possible advantages and disadvantages of this reactor type. One of the disadvantages frequently encountered is the presence of a radial solids flux and solids concentration profile: a dilute core of solids moves upward at high velocity while surrounded by a slowly-falling, relatively dense annulus (Hartge *et al.* [7]; Monceaux *et al.* [8]; Rhodes *et al.* [9]).

The present investigation began from the assumption that the gas–solids contact could be intensified through suppressing these radial solids profiles by introducing a regularly stacked packing into the fast fluid bed column. More specifically, the expected advantages of this packing type are:

- (i) break-up of the unfavorable radial porosity and solids flux profile, as well as of possible solids clusters, resulting in an improved gas–solids contact,
 - (ii) increase in solids hold-up and, consequently, in the conversion capacity per cubic meter reactor,
 - (iii) prevention of gas and solids backmixing on reactor scale, (mixing now takes place on the scale of a packing layer with almost no gas or solids exchange to the packing layers underneath; each layer represents a mixing unit)
 - (iv) simplified scale-up due to regular structure.
- Disadvantages are the increase in pressure drop due to friction with the packing, a more complex construction and, in principle, erosion and attrition of the packing although this was not observed during our experiments

which lasted in this case for more than 500 h. It is not yet clear whether the advantages related to the introduction of packing bars into the fast fluid bed will prevail.

Preliminary results, concerning the hydrodynamics of a cocurrent gas–solids flow in a pilot-plant-scale regularly packed circulating fluidized bed (RPCFB), have already been reported in an earlier paper (van der Ham *et al.* [10]).

The objective of the present study was to determine the hydrodynamics of the smaller set-up which will be needed later for the mass transfer measurements (Part II). These results will also be used for a comparison with the results of hydrodynamic measurements obtained for a similar pilot-plant-scale set-up (van der Ham *et al.* [10, 11]).

Previous work

The use of obstacles to enhance gas–solids contact as such is not new; it has already been applied for countercurrent gas–solids flow. Claus *et al.* [12] investigated the hydrodynamics of countercurrent gas and solids (sand of 235 μm diameter) flow through a column with randomly dumped packing of high-voidage cylindrical screens. The results showed a behaviour comparable to that found for gas–liquid systems. Like in gas–liquid operation, the packing was inserted to reduce axial mixing of gas and particles, to redistribute and carry the solids, and to enhance gas–solids mass and heat transfer. These results were confirmed by Roes and van Swaaij [13–16], Kuczynski [17] and Westerterp and Kuczynski [18] for fine powders (70 μm Fluid Cracking Catalyst (FCC)) flowing downwards (countercurrently to the gas flow) over randomly dumped packings of Raschig rings, Pall rings, cylindrical screens etc. Verver and van Swaaij [19, 20] improved the concept by applying a regularly stacked packing of bars (in a square column) instead of a dumped packing, to further minimize pressure drop and static solids hold-up, to realize a rapid radial solids distribution throughout the column and to prevent wall flow of solids. The solids used varied from FCC to coarse particles such as sand and steel shot (Verver and van Swaaij [19]).

Kiel *et al.* [21, 22] extended the research program concerning the gas–solid trickle-flow reactor, which was begun by Verver *et al.* They investigated the hydrodynamics of countercurrent gas–solids flow over five regularly-structured packing configurations at dilute phase or trickle-flow conditions. Data on average solids hold-up and pressure drop were obtained at ambient temperature in a $0.10 \times 0.10 \times 0.50 \text{ m}^3$ test column. Air was used for the gas phase, while the solids phase consisted of different size fractions of glass beads, the average particle diameter ranging from 200 to 750 μm .

Furthermore, Kiel *et al.* developed a one-dimensional steady-state hydrodynamic model, as an improvement of Verver and Van Swaaij's model. Kiel's model describes the experimental data on average solids hold-up and pressure drop quite well over a wide range of solids and gas mass fluxes.

The same set-up and experimental conditions were also used to study the hydrodynamic properties of cocurrent downflow of gas and solids over regularly stacked packings at dilute phase conditions. The presence of the packing effectively suppressed radial segregation of the solids phase (visual observation) and increased the average dynamic solids hold-up by a factor of about 10. This however was achieved at the expense of a considerable increase in pressure drop. The static solids hold-up in the packed section appeared to be negligible. The experimental results were compared again with computations from the one-dimensional steady-state hydrodynamic model developed for countercurrent gas–solids flow but now applied to the cocurrent situation. With two empirical fit parameters, which appeared to differ considerably from the corresponding values for the countercurrent flow case, it was possible to describe the pressure drop over the packing. In contrast to the countercurrent situation, the model failed to give an accurate description of the average solids hold-up. This failure was attributed to a different gas flow pattern just above the packing elements, which was not described adequately by the model.

As mentioned before, hydrodynamic properties (pressure drop, solids volume fraction) of a cocurrent upward flow of gas and solids in a regularly packed circulating fluidized bed (RPCFB) have already been discussed in an earlier paper (van der Ham *et al.* [10, 11]). The RPCFB used was different from and larger than the present one; it was provided with two square sections of $0.125 \times 0.125 \times 0.67 \text{ m}^3$ and contained regularly stacked packing bars (similar to those used by Verver and Kiel) of 0.02 or 0.03 m diameter. The study suggested that the solids hold-up increases linearly with the solids mass flux and was increased by a factor 1.6 to 5 compared to the column *without* packing. It further appeared that part of the solids inventory had settled on the packing. Finally, the pressure gradient was found to increase linearly with the solids mass flux and faster than linearly with the gas mass flux. A one-dimensional steady-state hydrodynamic model, based on the separate continuity and momentum equations for both phases, described the pressure gradient to consist of contributions due to gravity and friction of gas and solids with the wall and packing. Solids friction appeared to become increasingly dominant with increasing gas and solids mass flux. Its contribution to the pressure drop could be described with a solids friction coefficient, varying from

1.15 for a packing bar diameter of 0.02 m to 1.7 for a packing bar diameter of 0.03 m.

Jiang and coworkers [23] investigated (at ambient conditions) the effect of stator rings on the performance of a catalytic circulating fluidized bed reactor. More specific, the change in solids hold-up, pressure drop and ozone conversion has been measured as a function of the gas velocity (1.5, 2.0, 2.5 m s⁻¹) and the solids mass flux (<30 kg m⁻² s⁻¹). Ozone was catalytically decomposed using FCC particles with a mean diameter of 89 μ m and impregnated with ferric oxide as catalytic material. The CFB used was 0.10 m in diameter, 6.3 m in height and equipped with four stator rings at 1.1, 2.4, 3.6 and 4.8 m above the distributor. The open area of the ring was 56% of that of the riser. Ozone concentrations were measured in both axial and radial directions under various operating conditions in a riser with and without stator rings. Experimental results showed that, except for the lowest gas velocity, the riser equipped with stator rings yielded (i) an increase in solids hold-up (up to 50%), (ii) a higher pressure drop mainly due to an increase in solids hold-up, (iii) a more uniform radial ozone concentration distribution and (iv) a higher ozone conversion in the gas phase than without baffles. The more uniform concentration profile indicates an increase in radial gas and solids mixing brought about by the rings. A mathematical model, developed to describe the gas-phase ozone conversion, showed that the increased ozone conversion in the system equipped with stator rings was partly due to a higher solids hold-up but also caused by an improved gas–solids contact.

Although the results of Jiang and coworkers [23] showed an interesting improvement in the gas–solids contact by using ring-shaped baffles (stator rings), we expect that a packing geometry of grids (regularly stacked bars) will be just as successful but less sensitive to scale-up than the stator rings.

Recently Zheng *et al.* [24] published a paper concerning characteristics of fast fluidized beds with internals. Their study focusses on an appropriate design and installation of internals to minimize solids backmixing in both the dilute-phase and dense-phase regions while preserving the overall solids inventory. They therefore investigated the effects of rings, mounted as sets of four with an interval of 0.2 m, on the axial voidage profile and the solids backmixing. Unfortunately more information about the shape of the rings, *e.g.* thickness, internal diameter, was not available. The CFB used was 0.09 m in diameter, 10 m in height and was equipped with 2 units of four rings each, at 2.25 and 4.75 m. All experiments were carried out with air as the fluidizing gas and with FCC (47 μ m) as the solids. Voidage profiles were calculated from the axial pressure profiles and solids backmixing was determined from residence-

time distribution curves obtained with FCC tracer particles saturated with NaCl. The experimental results showed that (i) it is possible to cut a CFB into multiple segments with alternating dense and dilute zones, (ii) internals can be introduced without any notable decrease in the total amount of solids inventory and (iii) segmentation reduces backmixing significantly.

The results of both Jiang *et al.* [23] and Zheng *et al.* [24] show that the CFB performance is increased due to the rings installed. This conclusion is in agreement with our results. However, a more quantitative comparison of their and our data is not possible because of the large difference in shape of the internals applied.

Experimental

Apparatus and experimental procedure

All the experiments were carried out at room temperature and atmospheric pressure. Other conditions and properties of the solids material are given in Table 1.

The packed circulating fluidized bed set-up used is shown in Fig. 1. Except for the two cyclones (2) and the fluidized-bed (5) storage vessel (glass), all parts of the construction are made of Perspex or PVC. The Perspex riser unit (1) is cylindrical (0.04 m internal diameter) and 2.8 m high. The fluidized bed (0.1 m internal diameter; 1.5 m height) stores approximately 3 kg MZ-3S cracking catalyst. Solids flux measurement is carried out by recording the increase in solids volume of a small vessel (3) after blocking the solids return flow to the storage vessel by closing valve (4). A 45° inclined feed-line, with a vibrated slide valve (6) for a constant solids feed rate, connects the gently fluidized storage vessel with the lower part of the riser section. The riser unit contains: (i) a single, regularly packed section (A) of 0.48 m height and 0.06 × 0.06 m² cross-sectional area, the packing being built up of 0.010 m

TABLE 1. Experimental conditions and properties of the solids material applied

Gas flux range kg m ⁻² s ⁻¹	0.7 up to 4.4
Solids flux range kg m ⁻² s ⁻¹	up to 15
Solids type	FCC (MZ-3S)
Particle size distribution:	
diameter (μ m)	cumulative wt. %
< 149	99
< 105	79
< 80	45
< 40	8
< 20	1
Particle density (kg m ⁻³)	880
Mean particle diameter (μ m)	70
Terminal velocity (m s ⁻¹)	0.11

Fig. 1. Schematic diagram of the packed CFB set-up used for hydrodynamic measurements. The small-scale CFB consists of a riser (1) with a regularly-packed square section (A) and two slide-valves (B) for solids hold-up measurements, two cyclones (2), a solids flux measuring device (3, 4), a fluid bed (5) for solids storage, a 45° inclined feedline (6) with a slide-valve for solids flow control, a water saturator (7) and a screwcap (8) used for solids removal.

diameter Perspex rods, and (ii) two slide valves (B), operated by hand and positioned just above and below the packed section for the solids hold-up measurement. Before entering the vessel, the gas flow for the fluidized bed (air, stream II) is humidified in a water-saturator (7) to avoid undesired effects of static electricity in the installation. A series of six pressure taps (P1 to P6) was mounted in the wall of the packed section at equal vertical distances (8 cm).

The packing type studied is composed of layers of 0.01 m diameter Perspex bars. In each layer, these bars are positioned in parallel and at a distance of one rod diameter apart. Bars of successive layers are orientated perpendicularly, while alternating ones are parallel

again, but shifted over a distance of one diameter. The regular packing structure is repeated every four bar layers. Such a packing arrangement yields curved, alternately widening and narrowing channels in the axial direction. The solids flowing upwards through these channels are alternately slowed down by the packing and accelerated again by the gas. Furthermore, we expect that the packing ensures a continuous redistribution of gas and solids in the lateral directions, so that channeling of gas and solids is suppressed. This favorable quality of the packing has yet to be proved experimentally, nevertheless indications based on visual observations are already positive.

Backflow of solids to a lower packing layer is visually observed to be almost absent. This is due to the restriction imposed by the local high velocity of gas and solids at the center of each layer where the cross-sectional area is reduced by a factor of 1/2. Each packing layer can therefore be represented by a mixing unit. However, axial dispersion in the solids phase is not totally absent due to the fact that part of the solids inventory *may deposit* on the packing and will stay there some time. Only solids residence-time distribution experiments can give further information on the amount of axial dispersion.

After adjusting the gas and solids flux, the system was operated until steady state was achieved. Then, the overpressure profile (with respect to the atmosphere) along the packed section was measured. Secondly, the solids mass flux was determined as described before. Finally, the gas flow was diverted (by opening magnetic valve M2 and closing M1) simultaneously with closure of the slide valves (B) and blocking of the solids flow with valve (6). Consequently, the solids present in the packed section are trapped between two slide valves. To collect the solids, the bottom of the riser (8) was disconnected and the solids inventory of the lower part removed before the slide valve below the packed section was opened. Then the trapped solids fall into a sample flask placed underneath the riser and can be weighed. Solids resting on the packing bars were removed as much as possible by knocking on the walls of the column. After the axial pressure profile for the packed section was demonstrated to be linear, only the pressure difference between pressure tap P1 and P6 (see Fig. 1) was measured.

Various relevant dimensions of the square column section (A) with the packing are summarized in Table 2. These data are needed for processing the experimental results.

Interpretation of results

The results of the hydrodynamic measurements have been interpreted according to the model formulated in the previous work (van der Ham *et al.* [11]). With this model, values of a solids friction coefficient f_s can be derived, characterizing the pressure-drop contribution of the friction between the solids and the wall-plus-packing ($=\bar{F}_{t,s}$) relative to the total pressure gradient. Hereafter, the model used is briefly summarized.

The total pressure gradient across the packing is, according to the model, given by:

$$-\frac{\Delta P}{L} = (\bar{F}_{g,g} + \bar{F}_{g,s} + \bar{F}_{t,g} + \bar{F}_{t,s}) / (1 - \epsilon_p) \quad (1)$$

with $\bar{F}_{g,g}$ and $\bar{F}_{g,s}$ the gravity force per unit reactor volume for gas and solids, and $\bar{F}_{t,g}$ and $\bar{F}_{t,s}$ the friction force per unit reactor volume for gas and solids respectively. Equation (1) has been worked out further by introducing $-(\Delta P/L)_{G_s=0}$, the pressure gradient at zero solids flow ($=\bar{F}_{g,g}$ and $\bar{F}_{t,g}$), and constitutive equations for $\bar{F}_{g,s}$ and $\bar{F}_{t,s}$, which are respectively the contribution of the solids gravity and the solids friction with wall-and-packing.

$$-\frac{\Delta P}{L} = \left[-\frac{\Delta P}{L} \right]_{G_s=0} + \frac{\rho_s g}{(1 - \epsilon_p)} \bar{\beta}_{dyn} + \frac{4}{d_{h,c}} \bar{\tau}_s \quad (2)$$

where $\bar{\beta}_{dyn}$ = the dynamic solids volume fraction

$$\bar{\tau}_s = f_s \frac{1}{2} \frac{\bar{\beta}_{dyn}}{(1 - \epsilon_p)} \rho_s u_s^2$$

= the mean solids shear stress (3)

$$u_s = \frac{G_s}{\rho_s \bar{\beta}_{dyn}} = \text{the local solids velocity} \quad (4)$$

The solids friction coefficient f_s , defined according to the Fanning equation on the basis of solid particle

TABLE 2. Characteristic dimensions of the square column section

Column width (b); m	0.060
Cross sectional area ($A = b \times b$); m ²	0.0036
Column length (L); m	0.480
Number of bars per layer (n_a)	3
Number of layers per column B (n_b)	48
Pitch ($S_q = b/n_a$); m	0.020
Bar diameter (d_b); m	0.010
Volume of square column section B ($V_c = b^2 L \approx (n_a S_q)^2 n_b d_b$); m ³	1.73×10^{-3}
Volume of packing in one section ($V_p = n_a n_b b (\pi d_b^2 / 4)$); m ³	0.68×10^{-3}
Volume of column available for gas-solid flow: $[V_t = V_c - V_p = (n_a S_q)^2 n_b d_b [1 - (\pi d_b) / (4 S_q)]]$; m ³	1.05×10^{-3}
Volume fraction of packing [$\epsilon_p = V_p / V_c = (\pi d_b) / (4 S_q)$]	0.39
Volume of slide valve C (V); m ³	0.11×10^{-3}
Area of column wall in one section [$A_w = 4bL - n_b d_b b - 2n_a n_b (\pi d_b^2 / 4)$]; m ²	0.064
Area of packing in one section [$A_p = n_a n_b b (\pi d_b)$]; m ²	0.271
Hydraulic column diameter [$d_{h,c} = 4V_t / (A_w + A_p)$]; m	0.0125
Hydraulic packing diameter ($d_{h,p} = 4V_t / A_p = 4S_q / \pi - d_b$); m	0.0155

velocity and the dispersed solids density, is the only remaining unknown parameter in these equations.

In the present work f_s values have been calculated from pressure-drop experiments, and compared with those obtained for the pilot-plant scale set-up of the earlier study (van der Ham *et al.* [10, 11]). Unfortunately, other literature data obtained in a similar set-up are not available.

The use of solid friction factors (f_s), to describe the amount of friction loss, is a very common method in the field of dilute-phase pneumatic conveying of solids. Konno and Saito [25], Kato *et al.* [26] and Yang [27, 28] published f_s correlations for vertical pneumatic conveying lines operated at experimental conditions (i.e. gas and solids mass fluxes, solid material) comparable to ours. The f_s values calculated from these correlations for a tube diameter equal to the hydraulic diameter of the RPCFB ($d_{h,c}=0.0125$ m), range from 0.003 up to 0.03.

Results and discussion

The axial pressure profile along the packed sections has been measured as a function of the gas and solids mass fluxes. Some typical profiles are given in Fig. 2(a) and 2(b). The observed profiles are linear, the pressure gradient being constant, for all combinations of gas and solids mass fluxes realized. Figure 3 shows that this pressure gradient increases linearly with the solids mass flux. With respect to the influence of the gas mass flux, no simple relationship can be derived from the variation in the slope of the reported straight lines. At low gas mass fluxes ($G_g=0.7, 1.1$ and 1.5) there is no notable influence on the pressure gradient, whereas at larger fluxes ($G_g \geq 2$) the influence becomes quite significant. A second observation, also reported in our earlier publications (van der Ham *et al.* [10, 11]), is that the influence of the gas mass flux is reversed if compared to that found for circulating fluidized beds without packing elements, where the pressure gradient decreases with increasing gas mass flux for a certain fixed solids mass flux. This difference in behaviour is caused by the additional pressure drop due to friction losses with the packing which obviously dominates at increased gas mass fluxes.

From the mass of the solids trapped in the square packed section (i.e. the solids hold-up M), the mean solids volume fraction $\bar{\beta}$ can easily be calculated according to:

$$\bar{\beta} = \frac{M}{\rho_s(V_c + V_v)} \quad (5)$$

where M =the solids hold-up measured (kg), ρ_s =solids density (kg m^{-3}), V_c =volume of square section, in-

Fig. 2. The axial overpressure ($P(z)$) profile measured in the packed section for (a) a gas mass flux of $2.0 \text{ kg m}^{-2} \text{ s}^{-1}$ and a bar diameter of 0.01 m and (b) a solid mass flux of $5 \text{ kg m}^{-2} \text{ s}^{-1}$ and a bar diameter of 0.01 m.

Fig. 3. The axial pressure gradient along the packed section as a function of the solids mass flux for seven gas mass fluxes. Packing-bar diameter 0.01 m.

cluding packing (m^3) and V_v =volume of one valve (m^3). Values of V_c and V_v are given in Table 2. Because $\bar{\beta}$ is defined according to eqn. (5), the sum of the solids, gas and packing volume fraction is equal to 1, i.e.:

$$\bar{\beta} + \bar{\epsilon}_g + \epsilon_p = 1 \quad (6)$$

In Fig. 4 the solids hold-up and $\bar{\beta}$ are presented as functions of the solids mass flux. It is seen that (i) $\bar{\beta}$ increases linearly with the solids mass flux and (ii) the traced straight lines intersect, except for $G_g = 0.7$, at $\bar{\beta} = 0$ which means that the entire solids inventory is transported by the gas phase. The difference in behaviour at $G_g = 0.7$ could be related to the fact that the average gas velocity $\{u_g = G_g / [(1 - \bar{\beta} - \epsilon_p)\rho_g] \approx G_g / [(1 - \epsilon_p)\rho_g]\}$ of 1.0 m s^{-1} is then below the transport velocity of $1.2\text{--}1.5 \text{ m s}^{-1}$. This value was reported by Yerushalmi and Avidan [4] for a similar FCC material ($d_p = 49 \text{ }\mu\text{m}$, $\rho_s = 1070 \text{ kg m}^{-3}$). The transport velocity may be regarded as the boundary between the turbulent fluidization and the gas-solid transport regime. Consequently, at $G_g = 0.7$ a part of the solids inventory can settle on the packing.

A power form equation for $\bar{\beta}$ has been applied to correlate all the results:

$$\bar{\beta} = \alpha(G_g)^\gamma(G_s)^\delta \quad (7)$$

Values of α , and of the exponents γ and δ , have been calculated by the method of least squares to be:

$$\alpha = 0.0084,$$

$$\gamma = -1.22,$$

$$\delta = 1.0 \quad (8)$$

The above correlation is valid for $1.1 \leq G_g \leq 3.7$ and $0 \leq G_s \leq 15 \text{ kg m}^{-2} \text{ s}^{-1}$. Figure 5 shows all experimental data points together with eqn. (7), its 20% error lines, and the experimental results; 90% of the data lies within this indicated error interval.

Unfortunately, experiments in this column without packing have not been carried out. Therefore the effect of the packing on $\bar{\beta}$ cannot be quantified. However, the effect of the packing on $\bar{\beta}$ has already been reported

Fig. 4. Mean solids volume fraction and solids hold-up as a function of the solids mass flux for six values of gas mass flux. Bar diameter 0.01 m .

Fig. 5. The mean solids volume fraction as a function of $G_s G_g^{-1.22}$. Quantification of correlation coefficient for $1.1 \leq G_g \leq 3.7 \text{ kg m}^{-2} \text{ s}^{-1}$.

for the pilot-plant scale set-up (van der Ham *et al.* [10, 11]) as mentioned earlier.

Equation (2) shows that the total pressure gradient across the packing is caused by gravity and friction with wall and packing, of gas as well as solids. These contributions vary with changing gas and solids mass flux. Rearranging eqn. (2) and introducing the ratio ω yields:

$$\begin{aligned} \omega &= \frac{\frac{\rho_s g}{(1 - \epsilon_p)} \bar{\beta}_{\text{dyn}}}{\left[-\frac{\Delta P}{L} \right] - \left[-\frac{\Delta P}{L} \right]_{G_s=0}} \\ &= 1 - \frac{\frac{4}{d_{h,c}} \bar{\tau}_s}{\left[-\frac{\Delta P}{L} \right] - \left[-\frac{\Delta P}{L} \right]_{G_s=0}} \quad (9) \end{aligned}$$

ω represents the ratio of the hydrostatic pressure gradient over the measured pressure gradient minus the pressure gradient at zero solids flow (=only gravity and friction contribution of the gas). To calculate this ratio, the value of the dynamic solids volume fraction, $\bar{\beta}_{\text{dyn}}$, must be known. In the first instance, $\bar{\beta}_{\text{dyn}}$ is assumed to be equal to the measured mean solids volume fraction, $\bar{\beta}$. Calculated values of ω are shown in Fig. 6 as a function of gas and solids mass fluxes used. With the exception of $G_g = 0.7 \text{ kg m}^{-2} \text{ s}^{-1}$ all values are below 1. They decrease with increasing gas mass flux, which means that the solids friction with the packing and wall is becoming more and more dominant (see right-hand term of eqn. (9)). At $G_g = 0.7 \text{ kg m}^{-2} \text{ s}^{-1}$, ω ranges up to 1.3, which means that $\bar{\beta}_{\text{dyn}}$ is overestimated. Obviously, part of the solids inventory is carried by the packing. This conclusion is in conformity with the results

Fig. 6. The ratio of $\rho_s g \bar{\beta} / (1 - \epsilon_p)$ over the extra pressure drop due to the presence of solids, as a function of the solids mass flux for six gas mass fluxes.

(a)

(b)

Fig. 7. The total pressure gradient subdivided into the individual contributions, as a function of the solids mass flux for a gas mass flux of (a) 0.7 kg m⁻² s⁻¹; (b) 3.7 kg m⁻² s⁻¹.

of Fig. 4, which show that at $G_g = 0.7$ kg m⁻² s⁻¹ the ordinate at zero solids flow is 0.013 > 0. The total measured pressure gradient and the pressure gradient corrected for the contributions due to gravity and gas friction are presented in Fig. 7(a) and 7(b) for $G_g = 0.7$ and 3.7 kg m⁻² s⁻¹ respectively. Figure 7(a) shows a

negative value for the solids friction contribution below approximately $G_s = 4$ kg m⁻² s⁻¹. Here the calculated gravity contribution is too large because of an overestimated dynamic solids hold-up $\bar{\beta}_{dyn}$. In the discussion of Fig. 6 it has already been explained that part of the solids inventory is carried by the packing at $G_g = 0.7$ kg m⁻² s⁻¹. By comparing Fig. 7(a) and Fig. 7(b) it appears that gravity dominates at low gas mass fluxes, whereas friction dominates at high fluxes.

According to Fig. 4 (all fitted lines intersect with zero) and Fig. 6 ($\omega < 1$), it seems that the assumption $\bar{\beta}_{dyn} = \bar{\beta}$ is valid for $G_g \geq 1.1$ kg m⁻² s⁻¹. Visual observations are also in agreement with this conclusion. For $G_g = 0.7$ kg m⁻² s⁻¹, the fit in Fig. 4 shows an ordinate intersection at zero solids flow of $\bar{\beta}_0 = 0.013$ ($M_0 = 0.022$ kg) which is assumed to correspond with the amount of supported solids, independent of the solids mass flux applied (van der Ham *et al.*, [11]). In that case $\bar{\beta}_{dyn}$ is given by:

$$\bar{\beta}_{dyn} = \bar{\beta} - \bar{\beta}_0 \quad (10)$$

The supported solids hold-up ($\bar{\beta}_0$) built up at $G_g = 0.7$ kg m⁻² s⁻¹ is blown-out at higher gas mass fluxes ($G_g \geq 1.1$ kg m⁻² s⁻¹).

The solids friction is described by a solids shear stress $\bar{\tau}_s$ according to eqns. (2) and (3), with the solids friction coefficient f_s as the only remaining unknown parameter. In order to determine f_s , the mean solids shear stress should be calculated first according to eqn. (11) (eqn. (2) rewritten).

$$\bar{\tau}_s = \frac{d_{h,c}}{4} \left\{ \left[-\frac{\Delta P}{L} \right] - \left[-\frac{\Delta P}{L} \right]_{G_s=0} - \frac{\rho_s g}{(1 - \epsilon_p)} \bar{\beta}_{dyn} \right\} \quad (11)$$

At the same time, $\bar{\tau}_s$ is also given by eqn. (3):

$$\bar{\tau}_s = f_s \frac{1}{2} \frac{\bar{\beta}_{dyn}}{(1 - \epsilon_p)} \rho_s u_s^2 \quad (3)$$

The dynamic solids volume fraction ($\bar{\beta}_{dyn}$) is equal to the measured solids volume fraction $\bar{\beta}$, except for low gas mass fluxes when $\bar{\beta}_{dyn}$ should be derived by applying eqn. (10) ($G_g < 1.1$ kg m⁻² s⁻¹). Assuming that f_s is a constant, which is in agreement with the results of the pilot plant-scale set-up (van der Ham *et al.* [10, 11]), its value can be calculated by plotting $\bar{\tau}_s$ as a function of $(\bar{\beta}_{dyn} \rho_s u_s^2) / (2(1 - \epsilon_p))$ (see Fig. 8); the slope of the line which best fits the data while passing through zero, is equal to f_s . However, it can be seen in Fig. 8 that in this case f_s is not constant, but ranges from 2.3 to 1.4. Further analysis of the f_s values, calculated according to eqn. (3), shows that f_s is a linear function of the

Fig. 8. The mean solids shear stress as a function of $\frac{1}{2} \frac{\bar{\beta}_{dyn} \rho_s u_s^2}{(1-\epsilon_p)}$. Quantification of the solids friction coefficient. Packing-bar diameter 0.01 m.

Fig. 9. The solids friction coefficient as a function of the mean solids volume fraction of the gas-solid suspension.

solids volume fraction, as can be seen in Fig. 9. The best fit has been calculated by the method of the least squares and is given by:

$$f_s = 1.4 + 8.3 \left[\frac{\bar{\beta}_{dyn}}{(1-\epsilon_p)} \right] \quad (12)$$

The above correlation is valid for $1.1 \leq G_g \leq 3.7$ and $0 \leq G_s \leq 15 \text{ kg m}^{-2} \text{ s}^{-1}$. Figure 10 (parity plot) shows the total pressure gradient calculated according to the model (see eqns. (1) to (4)) with f_s defined by eqn. (12), as a function of the measured pressure gradient. All the data (except two) lie within 20% of the calculated value.

It has already been shown that the pressure gradient increases linearly with the solids mass flux, while with respect to the gas mass flux no simple relation could be derived. The reason for this will be shown below.

Fig. 10. Parity plot of the predicted [eqns. (2) to (4) and eqn. (12)] and the measured value of the pressure gradient for all data points obtained.

Substitution of eqns. (3) and (4) into eqn. (2) gives:

$$\left[-\frac{\Delta P}{L} \right] - \left[-\frac{\Delta P}{L} \right]_{G_s=0} = \frac{\rho_s g}{(1-\epsilon_p)} \bar{\beta}_{dyn} + \frac{2f_s}{d_{h,c} \rho_s} \frac{G_s^2}{(1-\epsilon_p) \beta_{dyn}} \quad (13)$$

Elimination of $\bar{\beta}_{dyn}$ through combination of eqns. (7), (8) and (13) yields:

$$\begin{aligned} \left[-\frac{\Delta P}{L} \right] - \left[-\frac{\Delta P}{L} \right]_{G_s=0} &= \left[\frac{\rho_s g}{(1-\epsilon_p)} (0.0084 G_g^{-1.22}) \right. \\ &\quad \left. + \frac{2f_s G_g^{1.22}}{0.0084 (1-\epsilon_p) d_{h,c} \rho_s} \right] G_s \end{aligned} \quad (14)$$

Equation (14) shows that the pressure gradient is linear in G_s but rather complex in G_g . Substituting the values for the constants ρ_s , g , ϵ_p , $d_{h,c}$ and $f_s (=2)$ yields:

$$\begin{aligned} \left[-\frac{\Delta P}{L} \right] - \left[-\frac{\Delta P}{L} \right]_{G_s=0} &= [119 G_g^{-1.22} + 71 G_g^{1.22}] G_s \end{aligned} \quad (15)$$

The pressure gradient has a minimum around $G_g = 1.2 \text{ kg m}^{-2} \text{ s}^{-1}$ and increases sharply with increasing G_g due to the second term in eqn. (15). This can be explained as follows. At low gas mass fluxes the solids hold-up, and consequently the hydrostatic pressure drop, is relatively high. The pressure drop caused by friction is then almost negligible. At increasing gas mass flux the hydrostatic pressure drop will decrease, while the

contribution due to friction will increase and finally become dominant. The minimum in total pressure corresponds with a mean gas velocity of 1.6 m s^{-1} which is approximately the transport velocity of FCC.

Comparison with previous work

Comparison of the presently observed results of the small-scale unit (meant for mass transfer measurements) with those of the pilot-plant-scale unit reported previously (van der Ham *et al.* [10, 11]), shows a large number of similarities but also differences. Both set-ups exhibit (i) a linear axial pressure profile, (ii) a pressure drop over the packing which is linear in the solids mass flux and greater than linear in the gas mass flux, (iii) a reversed influence of the gas mass flux on the pressure drop if compared to that found for circulating fluidized beds without packing elements (Yerushalmi and Avidan [4]), and (iv) a solids volume fraction linear in the solids mass flux.

The pressure drop is in both cases mainly due to hydrostatics and solids friction, with solids friction dominant at high gas and solids mass fluxes. The solids friction contribution is described with a solids friction coefficient f_s . Values are shown in Table 3 for both set-ups but different solids materials. The results obtained with the pilot-plant-scale unit show a constant value for the solids friction coefficient which exhibits an almost proportional increase with increasing bar diameter. On the other hand, the f_s values obtained with the small-scale set-up vary from 1.4 to 2.5 with increasing solids volume fraction (for one bar diameter). Furthermore, it seems that a decrease in the solids

density, and therefore the terminal velocity, increases the value of f_s .

However, the presence of a static solids hold-up is different. In the pilot-plant scale set-up a large amount of static solids hold-up was detected, in contrast to the results of the present small scale set-up. The difference is probably related to the bar diameter used, which is 0.01 m in the small set-up, and 0.02 and 0.03 m in the large one. Comparison of the β_{dyn} values found in both studies shows that the present dynamic solids volume fractions (representing solids transported by the gas) are a factor of 2 to 3 higher than those of the previous pilot-plant study. This factor decreases with increasing gas mass flux and is independent of the solids mass flux. The solids volume fraction for the small set-up is probably higher because of the smaller bar diameter applied, yielding a higher specific packing area, which increases the resistance to solids flow.

Friction coefficients calculated from correlations published by Konno and Saito [25], Kato *et al.* [26] and Yang [27, 28] for vertical pneumatic conveying lines, with a diameter equal to the hydraulic diameter of the RPCFB, range from 0.003 to 0.03. These values are a factor 100 to 1000 smaller than our results. Probably, this very large difference in f_s value is due to the perpendicular orientation of the packing bars towards the gas and solids flow, which is very obstructive to solids.

The applicability of the results (e.g. correlation for the solids volume fraction, friction coefficients) obtained so far is limited to the present configuration of the packing and the experimental conditions applied. Scale-

TABLE 3. The value of the friction coefficient f_s obtained in different set-ups and for different packing bar diameters

Set-up	Bar diameter	f_s	Solids material (FCC)
Small-scale (present study)	0.01 m	1.4–2.5	$d_p = 70 \text{ } \mu\text{m}$, $\rho_s = 880 \text{ kg m}^{-3}$, $u_t = 0.11 \text{ m s}^{-1}$
Pilot-plant-scale	0.02 m	1.15	$d_p = 75 \text{ } \mu\text{m}$, $\rho_s = 1660 \text{ kg m}^{-3}$, $u_t = 0.22 \text{ m s}^{-1}$
(van der Ham <i>et al.</i> [11])	0.03 m	1.7	$d_p = 75 \text{ } \mu\text{m}$, $\rho_s = 1660 \text{ kg m}^{-3}$, $u_t = 0.22 \text{ m s}^{-1}$

bar diameter 0.01 m bar diameter 0.02 m bar diameter 0.03 m
Packing configuration for lab-scale and pilot-plant-scale set-ups

up or extrapolation based on these results only will therefore be unsatisfactory. Even comparison with the results obtained for the pilot-plant set-up yields, thus far, mainly qualitative relations due to the limited number of packing configurations (e.g. bar diameters) and solid materials investigated.

Conclusions

The hydrodynamic properties of a gas–solids flow in a regularly packed circulating fluidized bed were investigated. The average solids hold-up and the pressure drop were measured at ambient temperature for a solids phase consisting of FCC (70 μm) and air as the gas phase. Furthermore, a one-dimensional hydrodynamic model was applied which describes the pressure drop as consisting of contributions due to gravity and to gas and solids friction with the wall and packing. The contribution of the solids friction is described with a solids friction coefficient.

The results of the hydrodynamic measurements show that:

- The pressure gradient, which is observed to be constant over the packed column height, increases linearly with increasing solids mass flux but faster than linearly with increasing gas mass flux.
- The influence of the gas mass flux on the pressure gradient is reversed compared to that found in a CFB without packing.
- The solids volume fraction is described fairly well by the following correlation:

$$\bar{\beta} = \bar{\beta}_{\text{dyn}} = 0.0084 G_s G_g^{-1.22} \quad (16)$$

which is valid for $1.1 \leq G_g \leq 3.7 \text{ kg m}^{-2} \text{ s}^{-1}$ and $0 \leq G_s \leq 15 \text{ kg m}^{-2} \text{ s}^{-1}$. In this range all the solids seem to be carried by the gas, i.e. no solids have settled upon the packing.

- The calculated solids friction coefficient f_s is a linear function of $\bar{\beta}_{\text{dyn}}$ according to:

$$f_s = 1.4 + 8.3 \left[\frac{\bar{\beta}_{\text{dyn}}}{(1 - \epsilon_p)} \right] \quad (12)$$

- Present f_s values are about 2 to 3 decades higher than those obtained from correlations for dilute-phase pneumatic conveying lines operated at the same experimental conditions.

– The hydrodynamic results obtained in the small-scale set-up are similar to those obtained in a previous study (van der Ham *et al.* [10, 11]) for a pilot-plant scale unit, except for the static solids volume fraction, which was only significant in the latter.

Acknowledgement

We wish to thank F. Horvath and A. den Hertog for their valuable contribution to the design of the packed CFB set-up and for performing the experimental work. The financial support of the Netherlands Agency for Energy and the Environment (NOVEM) is gratefully acknowledged.

List of symbols

A	area, further specified by subscript (m^2)
b	column width (m)
d	diameter (m)
d_h	hydraulic diameter (m)
d_p	mean particle diameter (m)
f_s	solids friction coefficient
$\bar{F}_t$	friction force per unit reactor volume (N m^{-3})
$\bar{F}_g$	gravity force per unit reactor volume (N m^{-3})
g	acceleration due to gravity (m s^{-2})
G	mass flux ($\text{kg m}^{-2} \text{ s}^{-1}$)
M	solids inventory of packed section (kg)
M_0	solids inventory carried by packing (kg)
L	column length (m)
n_a	number of bars per layer
n_b	number of layers per column A
P	pressure (Pa)
S_q	pitch (m)
U	superficial velocity (m s^{-1})
u	local velocity (m s^{-1})
u_t	terminal velocity (m s^{-1})
V	volume (m^3)
V_c	total volume of a square column section A (m^3)
V_t	reactor volume corrected for packing volume (m^3)
z	coordinate in vertical direction (m)

Greek letters

$\bar{\beta}$	mean solids volume fraction
$\bar{\beta}_0$	ordinate at zero solids flow, of the straight line fitting the mean solids volume fraction
$\bar{\beta}_{\text{dyn}}$	average dynamic solids volume fraction
$\bar{\epsilon}_g = 1 - \bar{\beta} - \epsilon_p$	average volume fraction of gas phase
ϵ_p	volume fraction of packing
ρ	density (kg m^{-3})
$\bar{\tau}$	average shear stress (N m^{-2})

ω ratio of the hydrostatic pressure gradient over the total pressure gradient minus the pressure gradient at zero solids flow [see eqn. (9)]

Subscripts

b	bar
c	column
g	gas phase
p	packing
s	solids phase
t	total
v	valve
w	wall

References

- 1 P. Basu (ed.), *Circulating Fluidized Bed Technology I*, Pergamon, Canada, 1986.
- 2 P. Basu and J.F. Large (eds.), *Circulating Fluidized Bed Technology II*, Pergamon, Canada, 1988.
- 3 P. Basu, M. Horio and M. Hasatani (eds.), *Circulating Fluidized Bed Technology III*, Pergamon, Canada, 1991.
- 4 J. Yerushalmi and A. Avidan, in J.F. Davidson, R. Clift and D. Harrison (eds.), *Fluidization*, 2nd edn., Academic Press, London, 1985, pp. 225–291.
- 5 K. Yoshida and H. Mineo, in L.K. Doraiswamy and A.S. Mujumdar (eds.), *Transport in Fluidized Particle Systems*, Vol. 1, Elsevier, Amsterdam, 1989, pp. 241–285.
- 6 J.R. Grace, *Chem. Eng. Sci.*, **45** (1990) 1953.
- 7 E.U. Hartge, D. Resner and J. Werther, in P. Basu and J.F. Large (eds.), *Circulating Fluidized Bed Technology II*, Pergamon, Canada, 1988, pp. 165–180.
- 8 L. Monceaux, M. Azzi, Y. Molodtsov and J.F. Large, in P. Basu (ed.), *Circulating Fluidized Bed Technology I*, Pergamon, Canada, 1986, pp. 185–192.
- 9 M.J. Rhodes, P. Laussmann, F. Villain and D. Geldart, in P. Basu and J.F. Large (eds.), *Circulating Fluidized Bed Technology II*, Pergamon, Canada, 1988, pp. 155–164.
- 10 A.G.J. van der Ham, W. Prins and W.P.M. van Swaaij, in P. Basu, M. Horio and M. Hasatani (eds.), *Circulating Fluidized Bed Technology III*, Pergamon, Canada, 1991, pp. 605–611.
- 11 A.G.J. van der Ham, W. Prins and W.P.M. van Swaaij, *AIChE Symp. Ser.*, **89**, No. 296 (1993) 53.
- 12 G. Claus, F. Vergnes and P. Le Goff, *Can. J. Chem. Eng.*, **54** (1976) 143.
- 13 A.W.M. Roes, *Ph.D. Thesis*, University of Twente, The Netherlands, 1978.
- 14 A.W.M. Roes and W.P.M. van Swaaij, *Chem. Eng. J.*, **17** (1979) 81.
- 15 A.W.M. Roes and W.P.M. van Swaaij, *Chem. Eng. J.*, **18** (1979) 13.
- 16 A.W.M. Roes and W.P.M. van Swaaij, *Chem. Eng. J.*, **18** (1979) 29.
- 17 M. Kuczynski, *Ph.D. Thesis*, University of Twente, The Netherlands, 1986.
- 18 K.R. Westerterp and M. Kuczynski, *Chem. Eng. Sci.*, **42**(7) (1987) 1539.
- 19 A.B. Verver and W.P.M. van Swaaij, *Powder Technol.*, **45** (1986) 119.
- 20 A.B. Verver and W.P.M. van Swaaij, *Powder Technol.*, **45** (1986) 133.
- 21 J.H.A. Kiel, W.P.M. van Swaaij, *AIChE Symp. Ser.*, **85**, No. 270 (1989) 11.
- 22 J.H.A. Kiel, *Ph.D. Thesis*, University of Twente, The Netherlands, 1990.
- 23 P. Jiang, H. Bi, R.H. Jean and L.S. Fan, *AIChE Journal*, **37** (1991) 1392.
- 24 C. Zheng, Y. Tung, H. Li and M. Kwauk, in O.E. Potter and D.J. Nicklin (eds.), *Fluidization VII*, Engineering Foundation, New York, 1992, pp. 275–283.
- 25 H. Konno and S. Saito, *J. Chem. Eng. Jpn.*, **2** (1969) 211.
- 26 K. Kato, Y. Ozawa, H. Endo, M. Hiroyasu and T. Hanzawa, in K. Ostergaard and A. Sorensen (eds.), *Fluidization V*, Proc. 5th Eng. Foundation Conf. Fluidization, Pergamon, Canada, 1986, pp. 265–272.
- 27 W.C. Yang, *AIChE Journal*, **24** (1978) 548.
- 28 W.C. Yang, in P. Basu and J.F. Large (eds.), *Circulating Fluidized Bed Technology II*, Pergamon, Canada, 1988, pp. 181–191.